2020 Compendium of Technical Papers
IsaMill’s horizontal configuration means it’s completely different from other mills. IsaMill™ gives me an increased recovery that outweighs the cost. I never have to worry about the mill.”

– Amandelbult Operation, Anglo American

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→ World’s only horizontal fine grinding mill, it avoids short-circuits and gives the highest availability
→ Most efficient fine grinding mill in the world
→ Strongest performance guarantee in the world
→ Most consistent product size
→ Delivers better results to downstream flotation and leaching

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isamill@glencore.com.au
Tel +61 7 3833 8500
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A HISTORY OF ISAMILL PROGRESS AT THE TECK RED DOG MINE

*Michael Larson¹, Brigitte Lacouture², & Greg Anderson³

¹Molycop USA
8116 Wilson Road, Kansas City, MO 64125, USA

²Teck, Red Dog Mine
3105 Lakeshore Drive, Suite A101, Anchorage, AK 99517, USA

³Glencore Technology
Level 10, 160 Ann St, Brisbane, QLD 4000, Australia

(*Corresponding author: mike.larson@molycop.com)

Abstract

In December 2011, Teck’s Red Dog Operation commissioned two 1.5 MW M3000 IsaMills as part of a project to improve their zinc metallurgy. This paper examines the history, including initial performance, characterization of a feed that has managed to be both abrasive and viscous at the same time, reviews improvements to the mill flexibility through an operating vessel size upgrade and the optimization of the internal component configuration for improved wear life. Red Dog has also completed a program for grinding media optimization. Recently, Red Dog finalized testwork and design on a value improvement project (VIP#2) that will install the world’s first M15000 IsaMill in 2019 into their grinding circuit to ensure throughput and grind size targets are maintained as harder ores are processed in the near future.

Keywords

IsaMill, regrind
Introduction

The Teck Red Dog Operation in Alaska has been running two IsaMills in the zinc circuit since 2011. These two IsaMills replaced seven smaller vertical tower mills. The circuit is shown in Figure 1. One IsaMill treats the zinc rougher concentrate while the other treats the feed to the zinc retreat circuit.

![Figure 1 – Teck Red Dog flotation and regrind flowsheet (Lacouture, 2013)](image)

The following will attempt to explain the methods involved in optimizing the wear and reliability of the two Red Dog IsaMills. The focus of this was not to optimize the grinding performance, but to improve the runtime, stability, and internal component wear of the two mills. This was a more demanding project, in that in a typical IsaMill operational challenge, either abrasive wear or viscosity will be an issue. Red Dog has both. The general operating conditions of the two Red Dog IsaMills are shown in Table 1.

This optimization program turned out to be a complex process involving flow, mixing, viscosity, grinding media, internal components, and the entire shell liner. Thermal imaging and regular viscosity measurements would prove to be critical tools in this process.

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Zn Rougher</th>
<th>Zn Retreat</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cyclone U/F (µm – Malvern)</td>
<td>90 – 120</td>
<td>50 – 70</td>
</tr>
<tr>
<td>Discharge (µm Malvern)</td>
<td>45 – 65</td>
<td>3 – 50</td>
</tr>
<tr>
<td>Recombine (µm Malvern)</td>
<td>28 – 37</td>
<td>20 – 30</td>
</tr>
<tr>
<td>Power (kW)</td>
<td>1000 – 1200</td>
<td>900 – 1100</td>
</tr>
</tbody>
</table>
**Parameter** | **Zn Rougher** | **Zn Retreat**
--- | --- | ---
Power (kWh/t) | 11 – 13 | 14 – 17
Feed Flowrate (gpm/m³/h) | 500/115 | 400/90
Feed Density (%) | 53 – 60 | 47 – 55
Feed Viscosity (cp) | 20 – 40 | 20 – 40
Feed Silica (%) | 17 – 27 | 35 – 55
Feed Barite (%) | 7 – 12 | 8 – 15

**Discussion**

**DISC SIZE AND THERMAL IMAGING**

From January 2012 (one month after startup) to June 2012 Glencore Technology and Red Dog experimented with different disc configurations in the rougher IsaMill to minimize disc and shell liner wear. This was aided with the use of a thermal imaging camera to track media movement and compression in the mill. The thermal imagery has proved to be a useful indication of the initiation and development of wear areas within the mill, particularly on the shell lining.

Initially the rougher mill was operated at around 1250 kW with excessive wear found towards the feed end of the mill on both the discs and shell liner. The thermal imaging in Figure 2 (Anderson, 2012) shows the concentration of media towards the feed end of the mill resulting in higher heat generation and wear on the shell liner.

As an initial step to address this, smaller diameter discs (SDD) were installed into the mill. Initially the maximum of seven were installed from the feed end, leaving a single normal-diameter disc in front of the rotor. The SDDs rotate at the same shaft speed but at a lower tip speed due to the reduced diameter. In addition, the smaller diameter creates a larger gap between the disc tip and the shell that allows for reduced compression and a more fluidized media pattern at the shell liner, as well as a reduced media impact speed on the shell liner. Upon startup under this configuration, the mill was limited to a power draw of 650 kW. Thermal imaging of the shell in Figure 3 (Anderson, 2012) showed the media had shifted towards the discharge end of the mill. The reduction in power was a direct result of the smaller disc diameter drawing less power per disc and also drawing the media further down the mill overcoming some of the pumping action of the rotor.

To improve the pumping action and distribution of the media within the mill, the rotor was reconfigured with a rhomboidal finger design in place of the standard square design. Adjustment of the rotor finger configuration allows for changes in the pumping volume of the rotor, the volume pumped by the rotor when operated at a fixed rotational speed. The disc configuration was left as seven SDDs. The resultant thermographic image is shown in Figure 4 (Anderson, 2012), clearly illustrating that the change in the rotor configuration had allowed the media to be pushed back towards the feed end of the mill. In addition, the overall power drawn by the mill was able to be increased to a maximum of 900 kW under these conditions. This configuration stabilized the wear on both the discs and shell liner in the mill. The disc wear can be seen from both the feed and discharge end in Figure 5.
Figure 2 – IsaMill thermal image – gouging

Figure 3 – IsaMill thermal image – media at discharge (left side)

Figure 4 – IsaMill thermal image – media at feed end (right side)

Figure 5 – IsaMill small diameter disc installation from discharge (left picture) and feed (right picture) ends
SHELL LINER

In order to increase the power draw on the mill under stable wear conditions it became apparent that the Red Dog IsaMills could benefit from a retrofit in shell sizing. Though the original IsaMills in operation at Mount Isa and McArthur River Mine were 3,000 litre (L) shells, a 5,000 L shell had recently been designed for non ultra-fine grinding duties with larger ceramic media. This shell would allow for a larger gap between the shell wall and the grinding discs, thereby reducing wear while allowing for a higher power draw. The disc-to-shell ratio of the M5000 with the normal diameter discs would replicate the use of the SDDs in the M3000. This extra space allows the media to properly fluidize and mix between the discs and the shell wall, rather than packing and gouging the rubber. In September of 2012 Glencore Technology converted both mill shells from 3,000 L to 5,000 L. This allowed the use of full-size discs and a return to +1000 kW operation without the previous wear issues.

GRINDING MEDIA

When the Red Dog IsaMills were first started, the retreat mill utilized 2 mm ceramic media and the rougher mill 3.5 mm ceramic media. The original IsaMill grinding theory was to use the smallest media possible. This was valid when the mills were first introduced in the industry given the approximately 7-micron (µm) target regrind sizes from an already fine feed, and limited selection of grinding media at the time. However, with improvements in grinding media and a better understanding of the IsaMill grinding process it became clear that the original theory was flawed, and that in many cases larger media would be beneficial to both the grind and internal wear properties of the process. By utilizing larger media, the coarser particles present in the flotation regrind feeds can be broken down quicker, so they do not accumulate at the front of the mill and prematurely wear those rubber components at the feed end of the mill.

To prove this point, Red Dog contracted ALS Metallurgical in Kamloops, BC, to run a media comparison trial on the zinc retreat feed in March of 2012. ALS tested graded charges of three different media top sizes, 2, 2.8, and 3.5 mm. These energy signature plot results are shown in Figure 6.

Figure 6 – IsaMill laboratory media sizing comparisons, retreat feed (Mehrfert, 2012)
The 2 mm results would indicate better performance, but what Figure 6 does not show is that the 2 mm test was holding in the coarse material, indicated by a higher motor power-draw — that insight is in the full test report.

As a result of this testwork, grinding media size was adjusted from 2 to 3 mm in the retreat. At the same time the rougher mill was switched from 3.5 to 4 – 5 mm. These changes also coincided with a change in supplier. This was brought on by availability of sizes, wear performance, and shape of worn media. The newer media was found to maintain a round shape longer than the original media, which would also contribute to improved component wear. As an added benefit it is thought that the relatively larger media with bigger gaps between packed pieces is impacted less by viscosity.

**SEPARATOR**

The discharge end of an IsaMill contains a centripetal separator that provides internal classification for the mill. This also provides the backflow necessary to keep grinding media and coarse particles in the mill. Typically, the fingers on this separator are a square or hexagonal shape. The Retreat IsaMill was experiencing higher than normal wear at the feed end flange, indicating a concentration of material in this area. To remedy the problem, some of the separator fingers were rounded (Figure 7) to make it less aggressive and reduce back flow.

![IsaMill separator retrofitted with round fingers (Lacouture, 2013)](image)

**VISCOSITY**

If viscosity were regularly measured at all IsaMill installations, it would be likely that Red Dog would rank at the top of challenging rheology. In this case, over the normal operating density range of 50% to 55% solids for the rougher mill, the viscosity increases by over 50% from the low to high point (Figure 8).
The mineralogy of barite, sphalerite, and silicates ground fine has resulted in some unique challenges to mill operations. The general power trend of this variable viscosity feed is shown in Figure 9. This is a good example that had been previously theorized for IsaMill operations but rarely demonstrated with actual plant data. As the feed solids and viscosity increase, the power draw increases, eventually past a point of optimum operation. As the viscosity continues to rise, the power draw will eventually drop off as the mixing of the mill decreases and the discs begin to rotate freely without agitating the charge adequately.
One option to reduce viscosity was to reduce the amount of recycle directed back to the mill feed. The recycle is primarily used to maintain constant flow into the mill, but there was room to reduce this. This was preferred over reducing density, as it has historically been shown that operation at dilute density also increases the component wear in the mill. All else being normal, it is usually recommended to operate an IsaMill at about twenty percent solids by volume. Across sites this has been a good starting guide to regulate surface area present.

One interesting product of this work is the beginnings of a rheological model based on surface area of the solids in slurry. A series of samples were taken in the Red Dog plant and from laboratory grinding. The plant samples included feed and product from both IsaMills and the laboratory samples were ground for varying lengths of time in a laboratory rod mill. These were analyzed in a Malvern laser sizer for a full-size distribution. Each individual particle size was assumed to be a sphere for simplicity, and then assigned its given surface area. Given the solid specific gravity (SG) and slurry density the total number of each particle size and then the total surface area per unit of volume was calculated. This was plotted against the viscosity of each sample as measured in the Red Dog laboratory (Figure 10). In general, there will be some differences based on method of grind (and resulting particle shape) and slurry stream mineralogy, but overall this looks like it could potentially be a start to a guide to viscosity characteristics if ever necessary.

![Figure 10 – Surface area vs. viscosity (Larson, 2012)](image)

IsaMill Grind

A series of surveys was completed in June of 2013 to review the two IsaMills’ performance. Figure 11 and Figure 12 show typical examples of mill breakage performance for the zinc retreat and rougher mills. In this case the rougher IsaMill ground from an F₈₀ of 105 µm to a P₈₀ of 42 µm at 10 kWh/t. The retreat IsaMill ground from an F₈₀ of 45 µm to a P₈₀ of 30 µm at 12 kWh/t.
In 2019 Red Dog will commission a new M15000 IsaMill as part of Red Dog’s latest mill upgrade. The mill will treat the prefloat tailings, grinding the material from a feed of 150 µm down to a product of 65 µm to feed the lead flotation circuit. The M15000 was chosen over the M10000 due to the larger gap between the discs and shell liner, along with a capability to process a higher feed flow than the M10000.
Conclusions

Given the complexity of the issues around the Red Dog Zinc IsaMills and the relative quickness with which they were addressed, this program should be considered a success. Less than eight months after commissioning, the mill had been adjusted both in internal components and media sizing to reduce wear and internal instability. Within one-and-a-half years Red Dog gained an understanding of the effect of viscosity on their mills that should be the envy of any number of researchers. This has resulted in more than seven years of reliable operation of the first two IsaMills at Red Dog and has given the team the confidence to install the first ever M15000 IsaMill to maintain mill throughput with harder ores.

Acknowledgements

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References


Abstract

The IsaMill™ was specifically developed to address the energy efficiency issues associated with grinding to fine and ultrafine sizes. It has been doing this successfully for more than 20 years in over 120 installations.

Recent efforts by Glencore Technology to further improve the energy efficiency of the IsaMill™ have resulted in the development of a new IsaMill™ shaft spacer design. The spacer was developed through laboratory, pilot and full scale testing prior to being commercialised. Full-scale results indicate that the new configuration was able to reduce the specific energy consumption by 13-17% in an M1000 gold application grinding to a P80 of 18 microns. The spacers were also shown to have no adverse impact on the shape of the product particle size distribution.

This paper discusses the development process of the new spacers through lab, pilot and full scale trials, operating results and processing implications.

Key Words

IsaMill™, fine grinding, energy efficiency, inert media

Introduction

It is well known that comminution energy consumption is a major concentrator operating cost. As a result there has always been a strong focus at mine sites to reduce energy consumption where possible to minimise operating costs – this has become increasingly important in the current economic environment. Increasingly, there are also political and social pressures to reduce overall energy consumption and associated carbon footprint. However, as orebodies become more complex, finer liberation sizes are required in order to produce saleable concentrates, driving up the specific energy required to produce each tonne of saleable concentrate.

The IsaMill™ was developed in the 1990’s by Mount Isa Mines (MIM, now Glencore) and Netzsch Feinmahltechnik, to specifically address the requirement for fine grinding down to a P80 of 7µm for the
further development of MIM’s McArthur River deposit and Mount Isa Pb/Zn ore bodies. The IsaMill™ addressed the two major issues of energy efficiency and the resultant downstream metallurgy, primarily through the use of small stirred, inert grinding media. The first 1.1 MW IsaMill™ using inert media was commissioned at Mt Isa in 1994 and four 1.1 MW IsaMills™ became the enabling technology for the McArthur River project in 1995. The early development and implementation of the IsaMill™ is well described by a number of authors including (Enderle, Woodall, Duffy, & Johnson, 1997), (Harbort, Murphy, Vargas, & Young Michael, 1999), (Johnson, Gao, Young, & Cronin, 1998) and (Pease, Curry, Barns, Young, & Rule, 2006).

Since its inception, the IsaMill™ has been successfully used to grind metalliferous concentrates to P₈₀ product sizes from 7µm to 60µm for over 20 years in over 120 installations and proven to be significantly more energy and metallurgically efficient than conventional ball mills and tower mills (Larson, Young, & Morrison, 2008) and (Larson M., Anderson, Morrison, & Young, 2011).

Ongoing efforts by Glencore Technology to further improve the energy efficiency of the IsaMill™ have resulted in a new IsaMill™ shaft spacer design. The new spacer was developed through laboratory, pilot and full scale testing prior to being commercialised and has shown operational specific energy reductions of 13-17% in a production scale gold concentrate regrind application. This paper discusses the development, plant performance and processing implications of the new spacer design.

**Grinding Mechanism and Conventional Spacer**

A conventional IsaMill™ configuration uses a cylindrical spacer to separate the grinding discs at the required distance. Figure 1 shows the conventional spacer on the shaft of an M1000 IsaMill™

![Figure 1 - Conventional Shaft Spacer between two Grinding Discs for an M1000 IsaMill™](image)

Figure 2 illustrates the grinding mechanism within the IsaMill™. The IsaMill™ typically operates around 70% media filling volume. As the shaft rotates, the grinding discs agitate the media such that it is drawn out along the face of the discs towards the shell liner. As it reaches the shell liner, the media is turned around and directed back towards the mill shaft area. This happens on the face of each disc, where there is sufficient media present and sets up a chamber of agitated grinding media between each of the grinding discs. Slurry to be ground enters opposite the shaft end cap at one end of the IsaMill™ and must pass through each of the agitated grinding chambers in series before it can exit, making it virtually impossible for any material to short circuit the IsaMill™. At the discharge end of the IsaMill™ is the patented product separator which makes use of a closer spacing between the final disc and the rotor to centrifuge any coarse particles and media towards the shell. The rotor, which acts like a pump, then
returns this material to the grinding zones. This mechanism allows ground product to flow through and exit the IsaMill™ whilst retaining the grinding media inside, all without the use of fine screens.

![IsaMill™ Grinding Mechanism](image)

**Figure 2 - Simplified IsaMill™ Grinding Mechanism**

**Laboratory Testwork**

In 2010, a large IsaMill™ development testwork program was undertaken at the Netzsch Feinmahltechnik laboratory in Germany. The program utilised an M20 IsaMill™ (20 litre volume), with a clear shell, to investigate and understand the impact on IsaMill™ power draw and axial media distribution of different designs and configurations of the discs, shaft spacers and rotor in conjunction with different operating shaft speeds and media loadings. The use of conical shaft spacers was investigated as part of this work.

The clear shell M20 is the same as a standard M20 IsaMill™ used for pilot scale test work other than being fitted with a clear plastic shell to allow the inner workings of the IsaMill™ to be observed under different operating conditions. It was built specifically for investigative work and demonstration purposes. To allow observation of the inner workings and preserve the integrity of the clear shell, only water and glass beads are used in this mill.

Prior to any data recording, the M20 was operated for 20-30 minutes to allow the bearings to warm up, so as not to impact on the power draw readings. Two shaft configurations were tested to isolate the impact of changing from standard cylindrical to conical spacers. Firstly, standard grinding discs and standard cylindrical spacers were used to generate the baseline data. Following this, the standard cylindrical spacers were replaced with conical spacers – the disc spacing and disc position along the shaft remained the same. In each of the two cases, the M20 was filled with 13.6 litres of 2mm glass beads and operated at a steady 16.3 litres/min of water throughput while the shaft speed was varied in stages from 600 to 1400 rpm. Typical operating speed for the M20 is 1200-1400rpm. The media distribution inside the mill was visually observed and the overall mill power draw recorded for each case. Figure 3 shows the conical spacer arrangement inside the glass shell M20 IsaMill™, with the feed end on the left and the product separator on the right.
As a note, the laboratory M20 IsaMill™ used was configured with a 9-disc arrangement and therefore a narrower shaft spacing compared to the standard 7-disc setup used for pilot scale metalliferous grinding. This could not be altered as all shaft parts were sized to fit the 9 disc arrangement. Other than that, the configuration was the same as that conventionally used with the 7-disc M20 arrangement.

Figure 4 illustrates the power draw as a function of the M20 IsaMill™ shaft speed and the shaft spacer type – which was the only parameter difference between the two curves. For a given shaft speed, there was a clear reduction in overall mill power draw when the conical spacers were used. A reduction of 20%-30% was evident at the typical M20 operating speeds of 1200-1400rpm. This was a direct result of using the conical spacers. As the shaft speed decreased, the difference between the two spacer designs decreased to zero at 800rpm and at 600rpm the conical spacer actually drew more power than the standard cylindrical spacer.

The conical spacers allowed the same volume of media to be agitated at the same mill shaft speed, but at a lower drawn power, than when the standard cylindrical spacers were used. Given that the agitated
media does the grinding, the impact of the reduced power draw on grinding efficiency – which could not be determined in the current testing program – was questioned. A separate pilot plant testwork program was designed and executed to answer this question.

Pilot Scale Testwork

An opportunity was identified during an onsite pilot campaign in January 2012, utilising an M20 IsaMill™, to conduct a grinding efficiency comparison between the standard cylindrical spacer and conical spacer configurations. This was a standard M20 pilot plant mill configured for metalliferous testing, with 7 discs rather than the 9 discs of the Netzsch laboratory M20. The spacing between each disc was therefore wider, which resulted in an increased included angle at the peak of the conical spacers. The spacers used in this testwork were specifically designed and built for the 7-disc arrangement and all tests conducted at the pilot plant utilised the 7-disc configuration.

Stage 1 of the pilot tests was designed to replicate the water and media testing carried out at Netzsch. Again, the only changed parameter between the two tests was replacement of the standard cylindrical shaft spacers with the conical spacers. The disc spacing and position along the shaft remained the same for both cases. At the selected pilot plant operating speed of 1390rpm and using 13 litres of 2.5mm ceramic grinding media at a water flowrate of 17.6 litres/min, there was a 15% reduction in the drawn power when the new conical spacer design was used. The overall power reduction attributed to the change in spacer design was less than that achieved at Netzsch but was still significant. Differences between the pilot tests and the Netzsch tests could likely be attributed to some of the key parameter differences including the number of discs, resultant changes in the conical spacer geometry, the shell material (rubber vs perspex), the media type (ceramic vs glass) and media volume. Figure 5 shows the cylindrical and conical spacer shaft configurations.

Stage 2 of the pilot tests involved conducting standard IsaMill™ signature plots using each configuration. The feed material was a coal and was managed such that the feed type and size distribution was as similar as possible between the two tests. Figure 6 illustrates the signature plot results for the two sets of spacer configurations.
There was a clear distinction between the two sets of curves. Average net power draw treating the slurry for the conical spacers of 7.5kW was 6% lower than the standard cylindrical spacers at 8.0kW. At the grind target \( P_{80} \) of 15 microns, the conical spacer configuration used 15% less specific energy than the standard cylindrical spacer configuration. This was the first evidence to support the contention that the conical spacers not only resulted in a reduced power draw for the same volume of agitated media, but that the resultant grinding action was more energy efficient.

Unfortunately due to the pilot plant requirements and schedule there was no further opportunity to complete any duplicate or further investigative work. However, it was considered that the result obtained in the single comparison, and the potential benefits it offered, warranted progression to full scale testing.

**IsaMill™ M1000 Spacer Design**

The initial production tests were carried out on an M1000 (500kW) IsaMill™. A series of design parameters for the M1000 spacer and for other full scale IsaMills™ were agreed upon based on the IsaMill™ M20 spacer design. In order to conduct the testwork as efficiently as possible, the test spacers were manufactured from polyurethane rather than the standard rubber lined method.

**IsaMill™ M1000 Stage 1 Testwork**

An initial, basic testwork program was designed to determine whether the observations from the pilot scale M20 IsaMill™ translated to the production scale M1000 IsaMill™. A site with a large surge tank ahead of the IsaMill™ circuit was identified to minimise variations in feed mineralogy, flow and size distribution between the tests as much as possible. The trial plan consisted of three stages with the aim of having the spacers as the only changed variable. The tests were conducted over a 4-day period in July 2013 and the results summarised in Table 1 below. All particle size analysis was done on a Malvern laser sizer.
Table 1 - M1000 Stage 1 testwork results

<table>
<thead>
<tr>
<th>Survey</th>
<th>Test Comment</th>
<th>Spacer Type</th>
<th>Power (kW)</th>
<th>Feed (tph)</th>
<th>kWh/t</th>
<th>P80 (µm)</th>
<th>P50 (µm)</th>
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<tbody>
<tr>
<td>1</td>
<td>Std conditions</td>
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<td>3</td>
<td>Std conditions</td>
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</tbody>
</table>

The first stage involved surveying the IsaMill™ circuit under the existing standard IsaMill™ configuration using the cylindrical spacers (survey 1, 2). Under standard operating conditions in survey 1, the IsaMill™ drew a gross power of 430kW for a specific energy of 54.7kWh/t, producing a product P80 of 21.1µm. In survey 2, the IsaMill™ drew a gross power of 427kW for a specific energy of 55kWh/t producing a product P80 of 21.4µm.

For the second stage (survey 3, 4), the IsaMill™ was reconfigured by swapping the standard cylindrical spacers with the conical spacers. As far as possible, the same volume of grinding media that was removed from the IsaMill™ after the completion of the first stage was returned to it. Figure 7 shows the conical spacers installed in the M1000 IsaMill™.

After the initial startup of each stage and stabilisation, top up grinding media was added in the usual way to maintain the initial power draw for each stage. This ensured, as much as possible, that the grinding media volume in the mill remained constant throughout the testwork period.

In survey 3, the IsaMill™ drew a gross power draw of 333kW - a reduction of 22% compared to the standard cylindrical spacer configuration. The IsaMill™ operated at 46.8 kWh/t – a 15% reduction compared to the cylindrical spacer configuration – and produced a product P80 of 21.5µm, similar to that produced from the cylindrical spacer configuration survey 2 and slightly higher than survey 1. This first comparative result suggested that the introduction of the conical spacer design had the potential for a 15% reduction in the required specific energy to produce the same product size, similar to what was observed in the pilot scale work.

The aim of survey 4 was to increase the specific energy, using the conical spacers, towards that of survey 1 and 2. By reducing the tonnage through the mill, the specific energy was increased to
51.4 kWh/t. The resultant product size decreased to a P80 of 20.4 µm. Note that survey 4 onwards was subject to a slightly coarser feed size due to an upstream ore change, which coarsened the feed into the surge tank ahead of the IsaMill™. Although the specific energy was increased in survey 4, it was still less than that of survey 1 and 2. Despite this, and the coarser feed size of survey 4, a finer product size was produced from survey 4. This again suggested a grinding efficiency benefit from the conical spacers.

At the completion of survey 4, the IsaMill™ was returned to the original configuration using standard cylindrical spacers for the third stage of testwork. Again, as far as possible, the same volume of grinding media that was removed from the IsaMill™ at the completion of the second stage was returned to it. Two more surveys (surveys 5, 6) were completed to compare against the original standard configuration surveys and the conical spacer surveys.

In survey 5, the IsaMill™ gross power draw returned to 429 kW, essentially the same power that was drawn under the same spacer configuration in the first two surveys. This confirmed that essentially the same volume of media was present in the IsaMill™ as had been present under the original conditions in stage 1. The IsaMill™ operated at 52.1 kWh/t for a product size P80 of 21.3 µm.

The specific energy in survey 5 with the cylindrical spacers had increased slightly from 51.4 to 52.1 kWh/t (in comparison to Survey 4, which had similar feed size distribution but with the conical spacers installed) but the product sizing had also increased from P80 of 20.4 to 21.3 µm.

Survey 6 confirmed a similar result, consuming 50.6 kWh/t to a P80 of 21.3 µm. When compared to surveys 3 and 4, surveys 5 and 6 both suggested a grinding energy efficiency benefit for the conical spacers over the standard cylindrical spacers.

Overall, the results of the initial on site testwork were very encouraging and supported the findings from the pilot plant work. The M1000 data suggested that the same product size could be produced about 15% more efficiently by using the conical spacer configuration. Given the exponential increase in specific energy requirements when grinding to finer sizes, this was a significant finding. Based on this success, a further, more detailed technical program was proposed to better quantify and confirm the advantages of the conical spacers.
IsaMill™ M1000 Stage 2 Testwork – Standard Mill Operating Speed / Varied Throughput

The second stage of testwork was conducted at the same site in September 2013 and again involved trials of the two different spacer configurations. A standard IsaMill™ signature plot, shown in Figure 8, was generated for each configuration by varying the throughput, within the operational limitations of the site. As per the initial stage of M1000 site testwork, the IsaMill™ was operated with nominally the same media load in each case to isolate the impact of the change in spacer design.

The signature plot equations from Figure 8 were used to construct Figure 9, which illustrates the reduction in specific energy requirement to a given target product \( P_{80} \) and \( P_{98} \) sizing. Table 2 summarises the data (within the range of product sizes produced during the testwork, i.e. \( 18\mu m < P_{80} < 22\mu m \)). The data indicated that for a \( P_{80} \) target of 22\( \mu m \), the conical spacers require 36.8kWh/t. This was 12.7% more efficient than the cylindrical spacers which require 42.1kWh/t. The \( P_{98} \) predicted for the conical spacer configuration was 55.5\( \mu m \) at 36.8kWh/t, 17.4% more efficient than the cylindrical spacers, which would require 44.5kWh/t to produce the same \( P_{98} \). Similarly, at a \( P_{80} \) target of 18\( \mu m \), the conical spacers were predicted to consume 53.3kWh/t. This was 17.4% more efficient than the cylindrical spacers which would consume 64.5kWh/t. The \( P_{98} \) predicted for the conical spacer configuration was 45.3\( \mu m \) at 53.3kWh/t, 21.1% more efficient than the cylindrical spacers, which would require 67.4kWh/t to produce the same \( P_{98} \).

An additional comparison to a target \( P_{80} \) of 15\( \mu m \) was made by using the signature plot equations (signature plots have been extensively proven to exhibit a linear relationship on a log-log plot (Larson M., Anderson, Barns, & Villalolid, 2012)). Analysis of Figure 9 shows that the predicted efficiency improvement of the conical spacers increased further (21.5% in \( P_{80} \) terms and 24.1% in terms of the energy equivalent \( P_{98} \) of 37.8\( \mu m \)) as the desired grind size was reduced.
Figure 9 - Reduction in specific energy to target product P₈₀ and P₉₈ sizing (varied throughput)

Table 2 - Energy Efficiency Improvement Presented by Conical Spacers during Varied Throughput Trial

<table>
<thead>
<tr>
<th>Cylindrical (kWh/t)</th>
<th>P₈₀ = 22µm</th>
<th>P₈₀ = 18µm</th>
<th>P₈₀ = 15µm</th>
<th>P₉₈ = 55.5µm</th>
<th>P₉₈ = 45.3µm</th>
<th>P₉₈ = 37.8µm</th>
</tr>
</thead>
<tbody>
<tr>
<td>Conical (kWh/t)</td>
<td>42.1</td>
<td>64.5</td>
<td>95.1</td>
<td>44.5</td>
<td>67.4</td>
<td>98.3</td>
</tr>
<tr>
<td>Efficiency Improvement</td>
<td>12.7%</td>
<td>17.4%</td>
<td>21.5%</td>
<td>17.4%</td>
<td>21.1%</td>
<td>24.1%</td>
</tr>
</tbody>
</table>

This data suggested that the impact of the conical spacers on grinding efficiency became more significant at finer grind targets and diminished at coarser targets. Although it was not proven whether the impact at coarser targets was neutral, or even negative.

It is well known that the shape of product size distributions from different grinding devices or circuits can vary such that the P₈₀ values may be the same for two very differently shaped curves (Gao, Reemeyer, Obeng, & Holmes, 2007) and (Larson M., Anderson, Morrison, & Young, 2011). Further, the downstream metallurgical performance can be influenced by the shape of the coarse end of the size distribution curve. In this case, the P₉₈ of each product size distribution was also measured and compared to the corresponding P₈₀ value as a method of quantifying changes in the shape of the coarse end of the size distribution – both between the two spacer designs and also as the specific energy input changed. The smaller the P₉₈/P₈₀ ratio, the tighter the distribution and therefore a lower proportion of particles at coarser sizes for a given P₈₀.

Figure 10 illustrates the P₉₈/P₈₀ ratios generated from the Figure 8 signature plot equations. It highlights the reduction in P₈₀ and P₉₈ values for a given specific energy as a result of using the conical spacers compared to the cylindrical spacers. For example, at a specific energy of 40kWh/t, the P₈₀ was reduced by around 7% and the P₉₈ by 9%. This corresponded to a reduction in the P₉₈/P₈₀ ratio from 2.59 to 2.52 – indicating a slightly tighter size distribution produced by the conical spacers, at the same specific energy input. Figure 10 indicates that larger percentage reductions in P₈₀ and P₉₈ product sizings
occurred at higher specific energies but the improvement in the \( P_{98}/P_{80} \) ratio between the spacer designs also diminished but was still in favour of the conical design.

![Figure 10 - Effect of Conical Spacers on \( P_{98}, P_{80} \) and \( P_{98}/P_{80} \) Ratio](image)

In reality, any efficiency gains will likely be realised as reduced specific energy consumption to the target product size, rather than reduced product sizing at the same specific energy target. Table 3 summarises the impact on the \( P_{98}/P_{80} \) ratio at target sizes of 22, 18 and 15µm for the M1000 trials. Clearly, there was only a nil to small improvement to the \( P_{98}/P_{80} \) ratio once the same target \( P_{80} \) sizing was considered for both spacer designs; however, the important fact was to confirm that it had not increased. The \( P_{80}/P_{50} \) and \( P_{80}/P_{20} \) ratios are also included for completeness. The ratios were largely consistent between the two spacer types, indicating that the conical spacers did not adversely impact the shape of the size distribution curves (at equivalent \( P_{80} \) target sizing).

<table>
<thead>
<tr>
<th>Target ( P_{80} )</th>
<th>Specific Energy (kWh/t)</th>
<th>( P_{98}/P_{80} )</th>
<th>( P_{80}/P_{50} )</th>
<th>( P_{80}/P_{20} )</th>
</tr>
</thead>
<tbody>
<tr>
<td>15µm</td>
<td>95.1</td>
<td>2.56</td>
<td>2.17</td>
<td>4.77</td>
</tr>
<tr>
<td>22µm</td>
<td>64.5</td>
<td>2.57</td>
<td>2.22</td>
<td>5.16</td>
</tr>
<tr>
<td>22µm</td>
<td>42.1</td>
<td>2.60</td>
<td>2.28</td>
<td>5.62</td>
</tr>
</tbody>
</table>

An additional set of tests to construct signature plots for both spacer designs was conducted where the mill speed was varied, rather than the mill throughput, to adjust the drawn mill power. Figure 11 illustrates the signature plots and indicates similar relationships to those developed when the mill throughput was varied.

IsaMill™ M1000 Stage 2 Testwork – Varied Mill Speed

<table>
<thead>
<tr>
<th>Target ( P_{80} )</th>
<th>Specific Energy (kWh/t)</th>
<th>( P_{98}/P_{80} )</th>
<th>( P_{80}/P_{50} )</th>
<th>( P_{80}/P_{20} )</th>
</tr>
</thead>
<tbody>
<tr>
<td>15µm</td>
<td>95.1</td>
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</tr>
<tr>
<td>22µm</td>
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<td>2.57</td>
<td>2.22</td>
<td>5.16</td>
</tr>
<tr>
<td>22µm</td>
<td>42.1</td>
<td>2.60</td>
<td>2.28</td>
<td>5.62</td>
</tr>
</tbody>
</table>
Figure 11 - M1000 IsaMill™ Signature Plot for Conical and Cylindrical Spacers – by varying Mill Speed

Figure 12 illustrates the reduction in specific energy requirement to a given target product $P_{80}$ and $P_{98}$, based on the signature plot data equations from Figure 11, and shows similar to trends to those observed for Figure 9 where the IsaMill™ throughput was varied. Table 4 summarises some of the data.

To achieve a $P_{80}$ target of 22µm, the conical spacers required 36.4kWh/t. This was 13.4% more efficient than the cylindrical spacers, which would consume 42kWh/t. The $P_{98}$ predicted for the conical spacer configuration was 55.6µm at 36.4kWh/t, 20.3% more efficient than the cylindrical spacers, which would require 45.6kWh/t to produce the same $P_{98}$. Similarly, at a $P_{80}$ target of 18µm, the conical spacers would require 56kWh/t. This was 15% more efficient than the cylindrical spacers, which would require 65.9kWh/t. The $P_{98}$ predicted for the conical spacer configuration was 45.4µm at 56kWh/t, 18% more efficient than the cylindrical spacers, which would require 68.2kWh/t to produce the same $P_{98}$.

An additional comparison to a target $P_{80}$ of 15µm was made by using the signature plot equations above. Analysis of Figure 12 shows that the conical spacers required 82.9kWh/t for a $P_{80}$ of 15µm, 16.4% more efficient than the cylindrical spacers, which required 99.1kWh/t.

These results are consistent with the magnitude of reductions observed by varying the mill throughput.
Continuous Operation

Based on the results achieved during the testwork campaigns, the test site opted to install the spacers for continuous operation. A set of rubber-lined spacers was designed and manufactured for this purpose. The site was able to operate the IsaMill™ at reduced power draw and produce their required grind size, in line with the testwork results. The spacers were expected to become more of a wear item in the mill due to their contribution to the grinding process. Figure 13 shows the spacers after 375 hours operation.
Process Implications and Considerations

There are a number of potential benefits and considerations for IsaMills™ operating with the new conical spacers due to the improved grinding efficiency.

- Reduced specific energy consumption for the same product size target or a finer product at the same specific energy consumption.
- Small improvement in the top end of the size distribution (P_{90}/P_{80} ratio) and associated downstream benefits, particularly if the same specific energy input is maintained.
- Reduced operating temperature for the same product size target or improved product size at the current operating temperature.
- If the mill power draw can be increased (by further media addition), potential for increased mill throughput at the same product size target.
- Operation at lower power draw will result in reduced component wear rates in the IsaMill™, although the conical spacers are expected to become more of a wear item.

The conical spacers will occupy volumetric capacity within the mill and based on the work here will result in a lower drawn power. Depending on the typical media loading in the mill, it may not be possible to draw full power if no more media can physically fit into the mill.

If the percentage grinding efficiency improvement is the same as the percentage amount that the power draw decreases then the mill can continue operating at the reduced power without any effect on throughput. If the grinding efficiency improvement is greater than the percentage amount that the power draw decreases then the mill can operate with a reduced media load and the same throughput. If, however, the grinding efficiency improvement is less than the percentage amount that the power decreases, then an increased media volume will be required to allow the IsaMill™ to process material at the same throughput. This may be an issue if the spacers have taken up the remainder of the available operating media volume.

In the M1000 work discussed in this paper, the power decreased by approximately 22% for the given operating media volume. At an 18µm P_{90} target, the grinding efficiency improved by ~17%, meaning
that if the IsaMill™ was at media capacity due to installing the conical spacers, the throughput would need to be reduced to maintain the same grind, as the media load could not be further increased.

**Optimal Sites**

Based on this work it appears that the sites which would benefit the most from installation of the new spacers are those with:

- Relatively fine grind targets
- Throughput constraints and a need to process more through the mill
- High specific energy applications running at close to the temperature limitations

**Summary**

The new conical spacer design was successfully tested from laboratory to pilot and then onto a full scale IsaMill™. At full scale, a reduction of approximately 13-17% in specific energy was observed over the range of throughput conditions tested (at standard operating mill speed) with an accompanying decrease of about 22% in drawn mill power. Based on the signature plots produced, the benefit was shown to increase to 21% at P<sub>80</sub> of 15µm target. In fine grinding, where energy requirements increase exponentially as the target size decreases, this is a significant finding. The new design offers a number of potential benefits and options to existing and future IsaMill™ installations, the most significant of which is a substantial improvement in grinding efficiency.

**Acknowledgements**

The authors acknowledge the M1000 operating site for allowing the testwork to proceed, the assistance of Voltaire Villadolid and Katie Barns in conducting and organising some of the associated testwork and Glencore Technology for permission to publish this work.

**References**


MOUNT ISA MINES NECESSITY DRIVING INNOVATION

*V. Lawson, H. DeWaal, G. Heferen, N. Ašlin, P. Voigt, and M. Hourn

Glencore Technology
Level 10, 169 Ann St
Brisbane, Australia, 4000

(*Corresponding author: virginia.lawson@glencore.com.au)

ABSTRACT

Mount Isa Mines (MIM) acquired a reputation for the successful application of R&D to develop break-through technologies for the mining industry starting in the 1978's through until the early 2000's. The ISAPROCESS™ tank-house technology has been licensed to copper refiners throughout the world, and a significant per cent of the world’s copper is refined using this technology. Since development in the late 1980’s more than 20 ISASMELT™ copper and lead smelting furnaces are now installed in countries around the world. Jameson Cell flotation technology developed jointly by Mount Isa Mines and Professor Graeme Jameson is widely used in the Australian coal mining industry and increasingly in the base-metal and gold industry. The IsaMill™'s developed at Mount Isa and McArthur River made it possible to develop the McArthur River and George Fisher orebodies and has been successfully implemented into base metal fine grinding applications around the world. The most recent commercialised innovation is the atmospheric leach Albion Process™ with its supersonic HyperSparge™ gas sparger, is being adopted as a solution to the increasing complexity of orebodies.

MIM’s contribution to the industry was significant given the size and the remote location of its operations with Townsville Copper Refineries more than 1350 km and Mount Isa 1800 km from the nearest state capital of Brisbane. This paper will briefly discuss the development of each of these technologies and why MIM – now owned by Glencore - was so successful innovating and developing such technologies over a period of nearly 40 years.

KEYWORDS

INTRODUCTION

Mount Isa is located in the Gulf Country region of Queensland about 1800 kilometers North West of Brisbane (see Figure 1). It came into existence because of the world class mineral deposits found in the area. In 1923 the orebody containing lead, zinc and silver was discovered by the miner John Campbell Miles. Mount Isa Mines Limited (MIM) was founded in 1924 to develop the minerals discovered by Miles, but production did not begin until May 1931. It paid its first dividend in 1947 after 16 years of troubled production. In 1954 the 1100 copper orebody was discovered and with rapidly rising reserves during the 1950’s and 1960’s led to the construction of new concentrators to treat lead/zinc/silver ores in 1966 (#2 concentrator) and copper ore’s in 1973 (#4 concentrator). The difficult nature of the Mount Isa lead-zinc orebodies has meant that the company had always needed to be at the forefront of mining technology. In the 1970’s through to the 1990’s, it became a world leader in developing new mining techniques and processing technologies as a response to declining metal prices and rising costs. Mount Isa has been smelting copper since 1953 and lead since the early 1930’s. Copper Refining at Mount Isa’s fully owned subsidiary of Copper Refineries Proprietary Limited (CRL) had commenced operations in 1959.

Figure 1 – Location of Mount Isa and Townsville relative to Brisbane the nearest Capital City

Technologies to come out of Mount Isa include the ISAPROCESS™ copper refining technology, the ISASMELT™, The Jamieson Cell, the IsaMill™, the Albion Process™ and the Hypersparger™. Mount Isa Mines Ltd was acquired by Xstrata in 2003 and Xstrata was then merged with Glencore in 2013. The level of innovation achieved at Mount Isa Mines is unsurpassed and was the result of the difficult nature of the Mount Isa ore bodies and its response to declining metal prices and rising operational costs in the 1970’s and 1980’s. By the 1990’s, Mount Isa had become a world leader in innovative mining techniques and state of the art processing technologies. The processing technologies are discussed below.

INNOVATIONS

Each of the innovations developed at Mount Isa Mines had a driver but the overarching desire was to make technology more efficient and cost effective. Each of these process developments will be discussed separately.
ISAPROCESS™

The development of the ISAPROCESS™ tank house technology had its beginning in the zinc industry. During the mid-1970s, MIM was considering building a zinc refinery in Townsville to treat the zinc concentrate produced by its Mount Isa operations. As a result, MIM staff visited the zinc smelters using the best-practice technology and found that modern electrolytic zinc smelters had adopted permanent cathode plate and mechanised stripping technology. MIM realised that the copper refineries performance was constrained by the conventional practice of copper starter sheets. The preparation of these copper starter sheets was labour intensive and the overall cycle was several weeks in duration.

MIM initiated a research program aimed at developing similar permanent cathode technology for copper refining. CRL, a subsidiary of MIM, had been operating in Townsville since 1959, using conventional starter-sheet technology and treating blister copper produced in the copper smelter at Mount Isa. Permanent cathode technology was developed and adapted over many years of in-plant experimental work and successfully introduced to the Townsville refinery in 1978. The fundamental difference between the new ISAPROCESS™ and the conventional starter sheet technology is the use of a permanent reusable cathode blank instead of a non-reusable copper starter sheet and the introduction of mechanised and automated electrode handling machines replacing labour-intensive manual operations. The vertical edges had plastic strips and the bottom cased in wax to prevent copper cathode from growing around edges of the cathode plate during stripping and allowing two separate copper sheets from each cathode plate. This technology led to major advances in the electrode handling systems and automation in copper tank houses. The improved geometry of the cathode plates and the significantly shorter cathode cycle times allowed for increased intensity and efficiency of the refining process. Introduction of permanent cathode technology resulted in higher capacity, better copper cathode quality with less defects, safer operation and a four-fold improvement in productivity. Considerable development work was required to modify original stripping machines from their zinc cathode origins due to the heavier cathodes. The stripping capacity of the machines has increased from 250 plates per hour to 600 plates per hour in the latest designs. More recent developments include the elimination of wax masking from the cathode plate, robotic electrode handling machines, and the introduction Duplex Stainless Steel cathode plates giving greater durability and corrosion resistance. Through the use of ISAPROCESS™ user forums, to exchange ideas and developments in the technology and to share operational experiences, the technology has enjoyed continued improvement with higher productivity and improved quality at low cost.

Figure 2 – The IsaKidd process
In mid 1981 Falconbridge Limited commissioned a copper smelter near Timmins to treat concentrate from its Kidd Mine. The original copper cathode produced at Kidd suffered from the presence of higher concentrations of lead and selenium and could not meet customer specifications. It was determined that the use of copper cathode sheets was preventing the Kidd refinery from meeting its cathode quality targets. Testwork began with the use of permanent stainless steel cathodes after preliminary tests showed a significant reduction in deleterious elements. The Kidd Process cathode used a solid copper lead bar welded onto stainless steel resulting in a lower voltage drop than the ISAPROCESS™. Falconbridge began marketing the Kidd Process technology in 1992 providing competition between the two suppliers of permanent cathode technology. Between 1992 and 2006, 25 Kidd technology licenses were sold and 52 ISAPROCESS™ licenses.

The development of the ISAPROCESS™ and Kidd Process set the scene for a run of technology developments that continued until the mid 2000’s. Xstrata took over MM in 2003 and then Falconbridge in 2006. The Kidd Process technology consequently became part of the tank house package and together they have since been marketed as IsaKidd™ representing the dual heritage of the technology. The current robotic stripping machine (Figure 2) is based on over 30 years of copper refining and winning technology. Today over 100 licenses are using IsaKidd™ technology.

ISASMELT™

The sister plant/blast furnace combination was the dominant technology for lead smelting throughout the 20th century. In the early 1970’s companies using this technology came under sustained political and economic pressure as tighter environmental regulations were introduced, and energy costs increased, leading to higher capital and operating costs (Fewings 1988). It was in this environment that Mount Isa Mines sought a process that would improve the performance of the operations at their lead smelter in Mount Isa. After investigating the various processes under development, researchers turned their attention to the Sirosmelt lance. It had recently been developed on a laboratory scale at the Commonwealth Scientific and Industrial Research Organisation (CSIRO) in Melbourne. Following initial investigations Mount Isa Mines recognised the potential of the novel concept for smelting of lead concentrates and embarked on an extensive development program.

In 1978 a joint project was initiated between Mount Isa Mines and CSIRO to investigate the application of the Sirosmelt submerged-combustion technology to the smelting of Mount Isa lead concentrates. The ISASMELT™ process, as it became known, was developed to maturity for smelting copper, nickel, lead and zinc feeds by Mount Isa Mines through the 1980’s and 1990’s using incremental scale up. Commercialisation only occurred once the process had been proven on laboratory, pilot and demonstration scale over many years. Approximately ten years were required for development of the lead and copper ISASMELT™ from crucible to demonstration scale (refer to Figure 4). During this decade the core know-how that was accumulated enabled the development team to reach the point where they were much better equipped to design and construct a full scale commercial plant – the final stage of the scale up process. Key aspects in this process were the selection of the scale up factors and the systematic design, development and re-engineering of several components of the technology. Figure 3 shows a comparison for the scale up stages for the lead and copper ISASMELT™ processes. Pilot scale was defined as unity for scale up comparison.
During the scale up process, refer to Table 1, several aspects of the technology were developed to a high standard that allowed the ISASMELT™ technology to become a commercial success. As a result, ISASMELT™ technology now operates successfully at numerous plants around the world. The methodical approach to development of the technology has allowed owners to modernise their existing operations or create new businesses with significantly reduced technical risk.

An important parameter in the evolution of the ISASMELT™ technology has been the refractory campaign life. Figure 5 shows the history of the refractory campaigns at the commercial copper ISASMELT™ plant at Mount Isa since commissioning. At the time Mount Isa Mines management considered the installation of water cooling on the furnace refractories undesirable because of the potential for fatal incidents and increased operating costs. As a result the commercial scale furnaces were constructed with minimal water cooling. Although this led to shorter campaign lives initially, a development program was begun that focussed on optimising refractory materials selection and installation methodology. When coupled with process control strategies and continuous on-line monitoring of the bath temperature using systems developed over more than 10 years of operation, it allowed Mount Isa Mines to achieve campaign lives of more than 3 years without using any water cooling of the furnace refractories.
Table 1 – Key Indicators of ISASMELT™ Plants from pilot to commercial scale

<table>
<thead>
<tr>
<th>Topic</th>
<th>Unit</th>
<th>Pilot Scale</th>
<th>Demo Scale</th>
<th>First Full Scale</th>
<th>Current Design</th>
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<tbody>
<tr>
<td></td>
<td>Pb</td>
<td>Cu</td>
<td>Pb</td>
<td>Cu</td>
<td>Pb</td>
</tr>
<tr>
<td>Furnace ID</td>
<td>m</td>
<td>0.4</td>
<td>0.4</td>
<td>1.8</td>
<td>2.3</td>
</tr>
<tr>
<td>Lance Diameter</td>
<td>mm</td>
<td>38</td>
<td>38</td>
<td>150</td>
<td>250</td>
</tr>
<tr>
<td>Lance Control</td>
<td></td>
<td>Manual</td>
<td>Semi</td>
<td>Manual</td>
<td>Automatic</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>Automatic</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Oxygen Enrichment</td>
<td>%</td>
<td>21</td>
<td>21</td>
<td>21</td>
<td>28</td>
</tr>
<tr>
<td>Nominal Feed Rate</td>
<td>qph</td>
<td>0.12</td>
<td>0.25</td>
<td>5</td>
<td>15</td>
</tr>
<tr>
<td>Offgas Treatment</td>
<td></td>
<td>Flue System / Baghouse</td>
<td>Gas cooler/ Baghouse</td>
<td>WHB</td>
<td>WHB²</td>
</tr>
</tbody>
</table>

Notes:
- ID: Internal Diameter; WHB: Waste Heat Boiler
- Refer to maximum throughput
- Some of the plants use a combination of radiation section and evaporative cooler for offgas treatment
- Refer to the smelting furnace from the two stage lead ISASMELT™ process

Figure 5 – Mount Isa copper ISASMELT™ plant campaigns (as of 2013)

Jameson Cell

The Jameson Cell (Figure 6) was jointly developed by Mount Isa Mines and Laureate Professor Graeme J Jameson (AO) of the University of Newcastle. Mt Isa had commenced operations with conventional flotation cells but was installing columns in cleaning duties in the mid 1980’s. The columns had the benefit of froth washing that was likely to allow significant grade benefits in the very fine lead zinc circuit. The first observations of the columns was that the collection process was slow necessitating long residence times and large volumes which remains a limitation of columns even today. In 1985 Professor Jameson was commissioned to undertake a project to improve the column sparger design.
Following initial work to provide an alternate method to bring together bubbles and particles, the downcomer was created. In the downcomer the air and the slurry are co-current with the air being entrained into the plunging jet under vacuum. Investigation showed that all of the bubble particle contact took place in the downcomer and thus the flotation tank could be much smaller. The first application at an industrial scale was in the lead zinc concentrator on the heavy media plant (HMP) lead slimes circuit. The initial improvement in performance were attributed to the very short residence time that allowed the minimisation of oxidation of galena fines. The cells were significantly smaller than the columns and there is no doubt the performance was superior as shown in Figure 3.

The testwork and trials in the early applications showed improved metallurgical performance when operated correctly. The challenge was operating them correctly. The technology hadn’t been sufficiently developed to be successfully adopted into plant operations. The cell fell out of favour in base metals and in the 1990’s was adopted into the Australia Coal industry and into niche SXEW applications where the main design challenges were resolved. The operability was improved by the introduction of a partial recycle to maintain constant flow and the maintainability of the cell was improved through various design modifications in operating plants. It was a period of continuous improvement. The result was a robust, low maintenance, easy to operate cell with the original features of excellent bubble particle contact.

The final obstacle was overcome when its adaption into the flowsheet was recognised to enable successful installations at the head of cleaner circuits and as low cost brownfield expansions. It is clear that the fast failures have had a significant effect on the success of the cell limiting its adoption into the industry. It is interesting that a significant proportion of sales are to return customers. Once you get over the hurdle of getting a Jameson Cell into your plant then seeing is believing. 2016 was the best year for Jameson cells into base metals and include the first sales back into South America where the cell had been abandoned after the difficulties of operations and maintenance of the Alumbrera installation. The metallurgical performance in Alumbrera was never the issue but the operators and maintainers hated the cells and they failed fast and hard.
The Jameson Cell celebrates its 30\textsuperscript{th} birthday this year and has finally been adopted into mainstream base metals concentrators mainly as cleaner scalper at the head of the cleaner circuit. The cells generally recover up to 80\% of the cleaner feed at high grades enabling much lower capital expenditure on the entire circuit. Process performance can be predicted from laboratory and pilot plant testing with demonstrated direct scale-up. It may have taken 30 years but the Jameson Cell is finally a success story. There are many lessons that can be learned from the implementation of innovation into industry from this case study.

\textbf{IsaMill\textsuperscript{TM}}

Unlike the developments of some of the other technologies at Mount Isa where efficiency was the main driver, the IsaMill was developed based on necessity. Figure 7 shows photomicrographs with the same scale of 40 micron demonstrating the increased complexity of Mount Isa ore over Broken Hill ore and the very difficult McArthur River ore. Although McArthur River was discovered in 1955 it was not able to be economically processed until the successful development of ultrafine grinding. McArthur River processing began in 1995 – 40 years after discovery when the IsaMill\textsuperscript{TM} made it technically and economically feasible to grind all of the rougher concentrate to 7 micron to facilitate the rejection of non-sulphide gangue. Even at 7 micron galena liberation is not possible and a bulk zinc-lead concentrate is produced.

![Photomicrograph](image)

\textbf{Figure 7 – Photomicrograph of a) Broken Hill ore b) Mount Isa ore c) McArthur River ore}

Investigations into fine-grinding started at Mount Isa started in the 1970s using conventional grinding technology to increase mineral liberation by grinding to fine sizes. These technologies were not only found to have high power consumption but also proved to be detrimental to flotation performance as a result of pulp chemistry and iron contamination from steel media. These poor results were revisited during pilot plant and tower mill testwork in the 1980s which also showed an inability of tower mills to economically achieve the required sizes. When it became clear that the solution to efficient fine-grinding did not exist in the minerals industry, MIM looked for ideas to “crossover” from other industries that also ground fine particles – pigments, pharmaceuticals, foodstuffs (e.g. chocolate). While these mills operated at a much lower scale and treated high value products they demonstrated the principle that stirring fine media at high speed was highly efficient. The challenge was transferring this concept to continuous, high tonnage and lower-value streams in the minerals industry.

In 1991 the introduction of a Netzsch laboratory stirred mill to the Mount Isa site was a turning point in fine grinding and ultrafine grinding. The $\frac{1}{2}$ litre bench scale mill resembled a milk shake maker and used fine copper smelter slag as grinding media. Testwork on McArthur River ore started in 1991, and by January 1992, a small pilot scale mill, LME100, had been designed and installed at the Mount Isa pilot plant. The testwork showed that high speed, inert, horizontal mills could efficiently grind to 7 microns at laboratory scale providing major improvements in metallurgical performance. To make ultrafine grinding applicable to full-scale production a program of development was undertaken between Mount Isa Mines Limited and NETZSCH Feinmahltechnik GmbH.
After 7 years of development and testing of prototypes in the Mount Isa operations, the IsaMill™ evolved. It was large scale, continuous, and most importantly robust because it was developed by operators. The crucial breakthrough was the perfection of the internal product separator — this allowed the mill to use cheap natural media (sand, smelter slag, ore particles) and to operate in open circuit. These are significant advantages for operating cost and circuit simplicity. Scale-up was tested using trial installations at the Hilton and Mount Isa lead/zinc concentrators. By the end of 1994, the first full scale IsaMill™ (1.1MW) was installed in the Mount Isa concentrator. Improvements to the technology were continually made by the operators, maintainers and engineers working with the technology.

In 1998 the rights for commercialisation of the IsaMill™ were transferred from Mount Isa Mines Limited to MIM Process Technologies (now Glencore Technology) and under an exclusive agreement with Netzsch. In December 1998, the IsaMill™ technology was launched to the metalliferous industry as a cost effective means of grinding down to and below 10 microns. The IsaMill™ is now a mainstream fine grinding machine with over 130 installations around the world.

**The Albion Process™**

In the 1990’s, MIM were studying options for the development of the large Frieda River/Nena project in PNG through its subsidiary Highlands Pacific. The Nena ores were not amenable to smelting, due to the elevated arsenic content, and several hydrometallurgical options were examined. Out of this work, MIM developed the Albion Process™, named after the suburb in Brisbane where MIM’s development laboratory was located. The Albion Process™ is a combination of ultrafine grinding using Glencore Technology’s IsaMill™, followed by oxidative leaching at atmospheric pressure in a series of reactors designed to achieve high oxygen mass transfer efficiency. The HyperSparge™ was also developed to deliver oxygen to the reactors efficiently.

Various small scale continuous pilot plant campaigns were conducted in 1994 and 1995. A larger pilot plant (120kg zinc cathode/day) was constructed in 1997 to conduct testwork as part of a feasibility study on the zinc/gold resources of Pueblo Viejo in the Dominican Republic. Extensive piloting was also conducted on lower grade chalcocerite concentrates for Cyprus Amax in 1998, and for Mount Isa Mines in 2000. Pre-feasibility and feasibility pilot testing was conducted on the zinc/lead bulk concentrates from McArthur River and Mount Isa in Australia between 2001 and 2005. During this time the Albion Process™ was successfully tested on over 70 different ores and concentrates. The process is designed to recover gold and base metals from refractory ores. The key to the process is the ultrafine grinding stage followed by a hot oxidative leach at atmospheric pressure.

In the period from 1994 until 2004, the Albion Process™ (see Figure 8) was seen as strategic to the MIM/Xstrata group, and was not marketed externally. In 2005, a decision was made to offer the technology to external clients under licence, and a marketing agent – Core Resources, was appointed to market the technology globally. Interest in the technology has been very strong in the subsequent period, with early licences signed in 2005 for the Las Lagunas Project, and 2006 for the Cerrej Project. The technology moved into commercial production in 2010 with the commissioning of Glencore’s Albion Process™ plant in Spain (4,000 tpa zinc metal), followed in 2011 by the commissioning by Glencore of a second plant in Germany (10,000 tpa zinc metal). The Las Lagunas refractory gold project commissioned in 2012, and the GPM Gold refractory gold project commissioned in 2013.
Figure 8 – The Albion Process oxidative leach plant in Armenia

The major scale up risk with any oxidative leaching technology is oxygen mass transfer. High agitator power demands are common to achieve the shear rates in the vessel required for effective mass transfer at a commercial scale. A different approach was taken in the design of the Albion Leach Reactor to lower the agitator power demand. Glencore developed the HyperSparger supersonic gas injection lance to provide gas injection velocities of the order of 500 m.s⁻¹ within the leaching vessel, compared to the 4 – 8 m.s⁻¹ achieved with a typical agitator. Supersonic oxygen injection is a far more efficient method of generating shear than conventional agitation, allowing the total power input into the vessel to be significantly reduced, and greatly reducing the scale up risk for the oxidative leach.

The Albion Process™ was enabled by the fine grinding of the IsaMill™ and the process was designed to deliver a lower cost processing option for treating refractory mineral resources. There are now six operating Albion Process™ plants and the process has now an extensive database of potential applications.

CONCLUSIONS

MIM developed a significant number of processing innovations that are technical and economic successes. The ability to innovate at MIM was enabled by very challenging preconditions and the need to process efficiently to remain economically viable. The success has been attributed to the development of these technologies on an operating site with the R&D group solving the technical issues on small scale. Each subsequent scale up was completed in the operating plants where the operators, maintainers, engineers and metallurgists were required to achieve production goals at each step of the scale up to ensure funding for the next step.

The number of innovations, at MIM, was disproportionate to the scale of operations and may have been enabled by the remoteness of the site and the researchers and operators working collaboratively to solve economic and technical problems. The research group were not capital city based but worked on the same site and were required to assist with installation, commissioning and operation of the various stages. This co-operation led to adoption into the plant and a fast feedback loop for improvements. The ultimate success of the innovations has been their widespread adoption into the mainstream industry where feedback from operating sites based on a user group model has enabled continuous improvement of each of the technologies.

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ECONOMIC RECOVERY AND UPGRADE OF METALS FROM MIDDLING AND TAILING STREAMS

*P. Voigt¹, M. Hourn¹, V. Lawson¹, G. Anderson¹, and D. Mallah².

¹Glencore Technology
160 Ann St
Brisbane, Australia 4000
(*Corresponding author: paul.voigt@glencore.com.au)

²Glencore Zinc
PMB 6, Mount Isa
Queensland, Australia, 4825

ABSTRACT

As mine head grades decline and orebodies become more complex, traditional mineral processing techniques and flowsheets to achieve saleable concentrate become more difficult to design and construct. Mines with lower quality concentrates or concentrates with penalty elements are particular under threat. The economics of these operations are far more susceptible to metal price, concentrate treatment terms and the availability of other, cleaner concentrates. Additional value may be realised for these orebodies through improved recovery by producing a low grade middling concentrate for further processing, in conjunction with a saleable concentrate.

The most cost effective way to reduce impurity levels is to do so as early as possible in the mining value chain. Technologies such as fine grinding and fine particle flotation are well established as effective methods for impurity rejection in mineral processing. What is normally overlooked is how a hydrometallurgical process could also be integrated in the overall flowsheet to achieve higher overall recovery at the mill. In the base metals environment, this is mainly because hydrometallurgical processes are associated with production of metal or use of expensive and toxic precipitating agents once the minerals of interest are solubilised. These processes can be very expensive, particularly with rising power costs and poor economies of scale in capital costs associated with low production rates from middling streams.

Glencore Technology (GT) has recent experience in the treatment of middling and low grade concentrate streams as well as tailings streams to compliment a concentrator flowsheet in a refractory gold and base metals setting. The value proposition is the isolation of a low grade middlings concentrate from the primary circuit or the tailings stream for upgrading to an intermediate product with an equal or higher grade than the primary concentrate to allow blending for sale. This allows plants to operate on a more favourable part grade-recovery curve while avoiding the expense of metal production. For existing operations this is particularly attractive since it can be added on with no process interruptions.

Two case studies are examined showing flowsheets and costings to arrive at the value proposition of the GT low grade treatment flowsheet.

KEYWORDS

Albion, IsaMill, Jameson cell, Glencore
INTRODUCTION

One significant challenge facing the gold and base metals mining industries is the globally observed trend of reducing mined head grades. This challenge is faced with existing operating assets and presents a significant hurdle in the justification for new projects. Figure 1 shows global trends in mined head grade from the mid 1800’s to 2010 (CSIRO 2015).

New projects or orebodies within existing operations, where reasonably good head grades can be maintained, tend to defy this globally observed trend through a corresponding increase in geological and metallurgical complexity. This is important for miners because metal input can be maintained without increasing milling rates however for a given flowsheet the quantity and quality of metal outputs will be compromised. While throughput of high-grade, complex orebodies can maintain input metal units they may not achieve target grade or recovery, and may even introduce penalty elements to the concentrate (Munro 2015), effecting economic viability.

By way of example, challenges faced with complex ore treatment in a flotation concentrator where the concentrate would be directed to a smelter include:

- Inability to find an economic operating point on the grade recovery curve such that both grade or recovery is not achieved for economic sale to a smelter
- Sacrificing the final product concentrate grade by inclusion of a middling concentrate that serves to increase overall recovery but negatively impacts grade and introducing penalty elements
- Inability to separate the economic minerals in an orebody rendering the production of a bulk concentrate with poor terms of sale to a smelter

Figure 2 illustrates the general trend of how grade versus recovery curves have become less defined as recovery increases.
In general, the gradient of the grade recovery curve has decreased resulting in a non-optimal operating position on the curve. This is a reflection of the general increasing complexity of ore to maintain mined head grade. The reason the ore complexity is reflected in Figure 2 is that minerals become more difficult to separate from one another and are recovered together, (Young 1997).

Certain ore-types cannot be upgraded with mineral processing techniques to produce a concentrate for downstream treatment in smelters. One example of these ores are the highly weathered or oxidised ores that are treated via heap leaching or whole ore leaching for copper and cobalt recovery. This processing method suits certain ore types and where sufficient infrastructure can be established at low cost. For example, heap leachable ore, requires ore with minimum levels of competency and permeability when it is stacked. The mineralisation must be such, that it is readily acid leached, with sufficiently low net acid consumption to be economic. A separate plant is then required to recover the copper once leached into solution, and then solvent extraction and electro-winning plants are required to produce saleable copper cathode. This flowsheet is conditional on the availability of reliable and relatively cheap power.

Ore types that don’t fit neatly into conventional flotation or heap leaching flowsheets have traditionally been relegated to waste. These ores are generally referred to as Complex Ores. Increasingly, these ore types can no longer be viewed as waste due to the contained metal content and high costs of pre-stripping this material where it overlies more economic deposits. Often these pre-stripping costs can make a project uneconomic unless metal can be recovered from this waste material.

Treatment of Complex Ores Through the Concentrator

During the mining boom of 2004 to 2009 the challenges of increasing metal prices coupled with the increasingly long lead times to bring a project to production led to mining companies adopting a strategy of a standard concentrator design (Combes 2011). The standard concentrator would allow significant improvement in the design and procurement phases of a project, and allow projects to be implemented faster to take advantage of rising metal prices. A common circuit used in a standard concentrator design is reflected in a simplified flowsheet in Figure 3.
The standard concentrator is a valid concept where a number of concentrators are envisaged to be built across projects with identical geological and metallurgical characteristics. The standard concentrator is still an excellent concept where there will be some variation between projects where some slight modification to the standard concentrator can be tolerated. When processing complex ores, however, where a single deposit may have multiple complex metallurgical domains, it becomes very difficult to design a single flowsheet that can treat all ore types while maximising economic performance.

Figure 2 shows that treatment of ore with increasing complexity in a set flowsheet will result in a compromise between recovery and grade or contamination of the final concentrate with deleterious metals and gangue. This can significantly impact the economics of a mining operation, (Munro 2015).

Modification of the Concentrator for Complex Ores

Modifications made from the conventional flowsheet for the treatment of complex ores generally comprises two approaches. The first is increasing the extent of grinding. This is based on the premise that when grade or recovery of an economic mineral is not obtained it is not liberated from gangue and does not have exposed surfaces to float. While there is some focus on the primary grind, usually when ore complexity increases, concentrate re-grinding, using a fine grinding mill such as the IsaMill™ is included in the circuit (Burford 2007). This style of flowsheet is reflected in Figure 4.
Figure 4 – Modification of the conventional flowsheet for fine grinding

The addition of concentrate regrinding to a flotation circuit as shown in Figure 4 is based entirely on increasing liberation. Once the minerals are liberated they must then be floated which can create further issues as finer particles have slower flotation kinetics compared to courser particles, and require more residence time to achieve the same recovery.

The second modification to the standard flowsheet commonly used is an increase of residence time through the installation of more flotation capacity or modification of the circuit configuration as reflected in Figure 5.

For fine particle flotation, a properly designed circuit may have the benefits of reducing recirculating loads, reducing reagent demand, as well as improved metallurgy. Examples such as McArthur River Mines routinely produce average zinc concentrate less than 10um in size, (Pease 2004). Alternatively, different flotation equipment can be added to the flowsheet such as the Jameson Cell which is well documented in the literature and proven in the field for improved fine particle flotation compared to a mechanical cell (Young 2006).

The basic modifications to the conventional flotation flowsheet are all validated when they result in improvements in the grade and recovery of the valuable metals to economic levels. Some extremely
complex ore types, however, will still not respond fully to such modifications and there is a need for a more encompassing approach.

A further complication occurs when designing a new project or modifying an existing operation for the treatment of a more complex ore, where the circuit design is based on the treatment of the most complex metallurgical domain in the orebody. This leads to circuit complexity that is not needed for a large percentage of the ore treated. Due to the mining sequence and certain ore types not stockpiling well, the more complex ores will not be treated in discrete campaigns. The result can be the installation of excess flotation and re-grinding capacity that is not utilised all of the time, resulting in an inefficient use of capital.

**GT Process for Treatment of Complex Ores**

Over the past 20 years there have been significant advances in technology and equipment in the fields of mineral processing and hydrometallurgy in the mining industry. GT has been at the forefront of these advances with the following technologies:

- **IsaMill™** – A high efficiency fine grinding technology in a horizontally stirred mill utilising inert ceramic media
- **Jameson Cell™** – A high intensity pneumatic flotation machine with no moving parts generating fine bubbles
- **Albion Process™** – Fine grinding followed by atmospheric leaching technology for refractory and base metals concentrates, including low cost recovery of base metals from solution to high grade concentrates with low grade reagents

The Albion Process™ is a patented technology developed by Glencore in 1994. The Albion Process™ consists of two key steps. The first step is ultrafine grinding of a sulphide concentrate, using Glencore Technology’s IsaMill™, to particle sizes in the range 80% passing 10 – 15 microns. The second step is an oxidative leach of the finely ground sulphides at atmospheric pressure to breakdown the sulphide matrix and liberate base and precious metals prior to metal recovery. There are currently five Albion Process™ plants operating globally in base and precious metals duties.

As a response to the increasing ore complexity, GT proposes a flowsheet that is a combination of these recent advances in mineral processing and hydrometallurgy processes. A conventional flotation flowsheet is still adopted when designing for new projects or existing operations encountering increasing ore complexity but with the addition of a hydrometallurgical processing step to deal with the low grade concentrates bled from the flotation circuit to smooth our variations in plant operation. One version of the concept is illustrated in Figure 6.

![Figure 6 – GT concept for complex ore treatment](image-url)
Figure 6 shows how the conventional flowsheet may be modified in an example where the concentrate reporting to the cleaner circuit is treated through an IsaMill™ and then a Jameson Cell prior to the Cleaner bank, obtaining grade, however the recovery is not at target levels meaning that further cleaning must be employed to achieve recovery. The Cleaner and Cleaner Scavenger banks provide further recovery, however the combined concentrate is below target, resulting in an overall dilution of the concentrate grade. In the modified flowsheet, a Jameson cell treats the Cleaner Scav tailings to recover a low grade concentrate. This low grade concentrate is bled from the circuit and processed in a dedicated hydrometallurgical plant. Complex middling particles recovered in the Jameson cell are removed from the circuit and are not recirculated through the flotation plant.

The fine grinding stage prior to the cleaning circuit allows for high pull rates from the Rougher and Scavenger, improving primary circuit recovery. The use of the Jameson Cell allows for good quality concentrate to be produced after fine grinding, with the wash water on the cell reducing the recovery of non-gangue particles.

The operation of the Cleaner and the Cleaner Scav enables the operation to balance the grade and recovery to be achieved from the circuit. Too high a recovery from this circuit recovers not only wanted liberated valuable mineral, but also the locked and complex particles towards the end of the circuit, unnecessarily diluting the concentrate. There is also the possibility of penalty elements that could be recovered in the concentrate with too high recovery rates. Therefore recovery needs to be controlled to prevent these particles from being recovered and left in the Cleaner Scav tails.

The Cleaner Scav tails are treated with a Jameson Cell, further increasing circuit recovery, but targeting complex particles that cannot be collected to concentrate in their current state due to the low concentration of valuable minerals.

The concentrate collected from the Jameson Cell doesn’t need to be high grade to be economically treated through the Albion Process™. Grades down to 5% copper in concentrate have found to be economic. One issue with including the Albion Process™ in the flowsheet is how to recover the metals that are leached into solution at low cost. This is achieved through a process developed by GT for zinc, copper, nickel and cobalt where either lime or limestone is used to continuously precipitate the base metals at controlled pH. A common problem with this type of precipitation process is the co-precipitation of gypsum. GT has developed procedures to overcome this issue by carefully controlling the process conditions such that the gypsum grows to particle sizes significantly coarser than the base metal oxides, and can be separated by hydrocyclone, as illustrated in Figure 7.

![Figure 7 – Integration of the Albion Process™ into the mineral processing flowsheet](image)

When applied to a copper circuit, the feed material to the Albion Process™ would be a low grade 5% copper middling concentrate which is then leached to solution and precipitated as a 45% copper oxide concentrate. This high grade intermediate can then be sold for use in a range of industries, or alternatively blended with the final concentrate product for sale.
CASE STUDIES FOR THE GT PROCESS

Two case studies are presented to illustrate the concept and high level information has been provided on the incremental improvement that can be achieved by incorporating the Albion Process™ into the concentrator flowsheet for processing a low grade middling stream.

Case 1 – Copper NW Queensland

Case 1 relates to an opportunity for brownfields expansion of mining and concentrator operations at a mine in North Queensland. The ore complexity at the operation will increase for a short duration due to the need to mine through mainly transitional/weathered ore zones with a variable base of oxidation, resulting in some primary zones intermixed with the transitional ores. The variation in the contact zone between the transitional and primary ores is such that the ores cannot be separated and must be treated together. They are in a quantity such that the contained metal units from both transitional and primary ore must be recovered to justify the overall project.

The main copper bearing minerals comprise native copper, chalcopyrite and chalcocite with minor chrysocolla and malachite. The sulphide gangue comprises mainly pyrite with minor pyrrhotite, galena and sphalerite.

The ore presented to the process typically grades 1 to 2% copper with varying mineralogy. Within the feed some of the ore types can be recovered to a concentrate grading greater than 25% copper, however for the pure transitional ores, the maximum copper concentrate grade is only 5% copper. The transitional concentrates, however, contain predominantly leachable minerals, and lend themselves to be separated to a middlings stream for separate hydrometallurgical processing. A simplified flowsheet is illustrated in Figure 8.

Figure 8 – Copper treatment flowsheet

The simplified flowsheet in Figure 8 indicates a pathway for liberated copper sulphide mineralogy to report to a final cleaner concentrate to achieve on specification concentrate at +26% copper grade for sale. The material that is not amenable to upgrading, diluted by both sulphide and non-sulphide gangue is recovered in both the scavenger concentrate and the cleaner tailings as a 5% Cu middling concentrate that is treated through the Albion Process™. Even at low feed grades the hydrometallurgy treatment option is economical since the final intermediate copper oxide produced grades approximately 50% copper and overall circuit grade and recovery are maximised.
The mineral processing and hydrometallurgy flowsheet adds between 4% to 30% copper recovery at target grade depending on what material is treated. The project allows access to a further 6.0 Mt of high grade primary ore. The treatment of the transitional cap alone has an IRR of approximately 25%.

**Case 2 – Zinc NW Queensland**

This application of the GT flowsheet to zinc processing is treatment of historical zinc tailings. It was acknowledged that significant zinc was contained in the tailings but when the tailings were re-floated a zinc grade of around 10% had to be accepted for any economic recovery levels due to both sulphide and non-sulphide gangue. A number of hydrometallurgy flowsheet options were considered to treat the low grade concentrate, however these options were marginal economically due to the high cost of installing expensive processing equipment to recover the final zinc metal product. The GT approach overcame these economic and technical limitations. A simplified flowsheet is illustrated in Figure 9.

![Figure 9 – Zinc treatment flowsheet](image)

The flotation tailings can be re-floated in a conventional flowsheet to produce a 10% zinc concentrate at 90% recovery. The resulting concentrate is fed to an IsaMill™ for grinding to 80% passing 20 micron or below and then to the Albion Process™ oxidative leach for the extraction of zinc. The oxidative leach achieves zinc recoveries of up to 99.5%. Glencore has installed two Albion Process™ plants to recover zinc from a bulk concentrate, in Spain and Germany, and has experience in the design of these plants (Hourn 2012).

Once in solution, rather than producing metal, the GT precipitation process was used with either lime or limestone to produce a zinc oxide at a grade between 50 to 60% zinc. Strategically, since most zinc roast-leach-electrowinning plants are limited at the roaster, such a zinc product can be treated by conventional zinc refineries to maximise cellhouse capacity or operate during roaster downtime.

The plant was sized to produce 100ktpa contained zinc metal but could be easily scaled down if required. The project IRR was 30%.

**CONCLUSION**

GT has developed novel flowsheet configurations for the treatment of complex ores through minor modification to conventional flowsheets with minimal process disruption and integration with hydrometallurgy unit operations. The flowsheets presented are just examples and many other variations are possible.
In the mining value chain, value can be most easily added when complexity in ores can be overcome at the earliest part of the chain as practicable. This starts in the mine with grade control, understanding metallurgical domain definition and optimising blast patterns. In the mineral processing sphere this can start with screening and dense medium separation through to grind size and reagent use. In extractive metallurgy this can mean blending with different feeds, additional plant and equipment through to by-product waste disposal. The final product is sold in the market relative to the product quality of other producers. The authors recognise that most value can be added through addressing complex ore treatment as early as possible in the value chain and this paper focuses on blending hydrometallurgical techniques with minerals processing to address ore complexity in the concentrator.

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IMPROVING CONCENTRATE GRADE THROUGH SMART DESIGN AND PILOTING

*V. Lawson1, G. Anderson1, T. Strong2

1Glencore Technology
Level 10, 160 Ann St
Brisbane, Australia, 4000
(*Corresponding author: virginia.lawson@glencore.com.au)

2Chief Operating Officer
Panoramic Resources Limited
PO Box Z3487
Perth, WA, 6831

ABSTRACT

Selectively recovering pentlandite from high levels of pyrrhotite presents a technical and operational challenge. The liberation characteristics and also the chemical environment is critical to performing a successful separation. Reviewing the mineralogy to understand the characteristics of the ore and the plant flotation behavior is fundamental to designing a successful improvement program.

At Savannah Nickel the challenges presented by a falling nickel price required a rethink of the flowsheet to improve the project economics through improving the concentrate grade and recoveries. Glencore Technology reviewed the plant performance and proposed and tested solutions to improve the grade performance of the concentrator with the support of the plant team. A pilot plant comprising a M20 IsaMill™ followed by an L500/1 Jameson Cell was operated on various plant streams. In addition to the pilot plant, flotation surveys and a plant based laboratory program was used to supplement the plant understanding and within two weeks significant improvements were identified and demonstrated. The combination of the pilot operation and the plant based laboratory program was able to quickly demonstrate the improvements and test with equipment that is fully scaleable from pilot plant to full scale.

KEYWORDS

Pentlandite, Pyrrhotite, IsaMill™, Jameson Cell, Cu/Ni separation, Mineralogy
INTRODUCTION

Panoramic Resources Limited is a Western Australian mining company formed in 2001 for the purpose of developing the Savannah Nickel Project in the East Kimberley region of Western Australia. Panoramic successfully commissioned the $65 million Savannah Project in late 2004. The process plant at Savannah comprises a single stage crusher, SAG mill, flotation, thickening and filtering stages to produce a bulk nickel, copper, cobalt concentrate. Metallurgical recoveries average 86-89% for nickel, 94-97% for copper and 89-92% for cobalt. The plant was originally designed for a throughput of 750,000 tonnes per annum, but has consistently outperformed the design specifications.

Due to low nickel prices the Savannah Nickel mine and processing plant were placed into care and maintenance in May 2016. Prior to suspending production Glencore Technology was engaged to investigate options to improve the concentrate grade. With the discovery of Savannah North in 2014 and the Savannah North Scoping Study demonstrating LoM ~8 years and Mining Inventory of 6.07Mt @ 1.26% Ni, 0.64% Cu and 0.09% Co significant future potential exists. By completing pilot plant studies prior to shutting down, the plant engineering studies can be completed and the plant readied for restart by the time the nickel price improves.

Background

The Savannah Nickel deposit is located in the Kimberley region of Western Australia south west of Kununurra (Figure 1). The project operates as a fly-in fly-out operation with the majority of its workforce sourced from Perth.
The current processing plant is shown in Figure 2 and Figure 3. The main features include the flash flotation cell and the rougher 1 concentrate that are in open circuit while the rougher 2 concentrate and scavenger concentrate are cleaned through a single stage rougher cleaner (RCC) and scavenger cleaner (SCC). The plant produces a low grade Cu-Ni concentrate at high nickel recoveries.

**Figure 2 – Savannah Nickel Processing Plant**

Historical Data Review

To understand the requirements for improving the concentrate grade at Savannah Ni, an historical review of plant data was undertaken by Glencore Technologies. This review included previous survey work, daily and monthly balances and where available mineralogy of selected streams. The plant had been successfully operating in a high recovery, low concentrate grade scenario. The incentive for the review was to investigate whether the concentrate grade could be improved to improve project economics through reduction in transportation and smelting costs at the historically low nickel prices.

The composition of the nickel concentrate was reviewed by size to determine the source and composition of the diluents. These data are given in Figure 4. The valuable minerals are distributed across...
all sizes while the non sulphide gangue (NSG) is in the finest fractions and pyrrhotite (Po), the most dominant diluent, is in the intermediate fractions although it is diluting the concentrate across all fractions. The source of the pyrrhotite needed to be identified across the four streams that combine to make the final concentrate.

By reviewing the diluents by size and by stream (Figure 4) it was determined that the non-sulphide gangue dilution was distributed across the various concentrate streams with the rougher 1 concentrate and rougher 2 concentrate being the major sources. The pyrrhotite was also distributed across all streams however the major source was the flash concentrate in the particle sizes greater than 38 micron. The remaining pyrrhotite was in the floatable fractions between C4 (approximately 15 micron) and 38 micron mainly in the rougher 1 and 2 concentrates.

Limited liberation data existed for the individual concentrates however the point counting data that was available indicated that the pentlandite (Pn) appeared to be adequately liberated as defined by Johnson (1988) with values greater than 80% while the chalcopyrite (Cp), pyrrhotite and non-sulphide gangue were only moderately liberated. The main association of chalcopyrite was with pyrrhotite while the pyrrhotite was associated with pentlandite. The major association for the non-sulphide gangue in the concentrate was with chalcopyrite.

These data when reviewed collectively suggested that to remove the diluents from the concentrate would require a combination of a moderate regrind and improvements in flotation separation to improve the concentrate grade. A proposal for an onsite pilot campaign using both an M20 Isamill™ in combination with an L500/1 Jameson cell was provided. As the separation of pentlandite from pyrrhotite is challenging and often requires an optimisation of chemistry, an accompanying laboratory campaign was proposed to be conducted on the samples from both the plant and the pilot plant to confirm the correct chemistry of separation.

EXPERIMENTAL

The pilot plants were shipped to site in April 2016. The onsite personnel arranged for set up and connection to services. In addition a laboratory proposal had sourced the required chemicals and readied all the required laboratory equipment in advance to make best use of the two week campaign. Planning also included sourcing mine supply that would best represent the future operation as any flowsheet changes would take time to implement and current ore sources were in decline. The installed pilot plants are shown in Figure 6 near the flotation feed conditioning tank.
Figure 5 – Diluent by size and stream

Figure 6 – Pilot plants on site
The pilot plant consisted of an M20 Isamill™ and an L500/1 Jameson Cell. Both are highly instrumented allowing for minimal operator intervention. This enabled a single operator to run both pilot plants while a second was able to conduct flotation tests. A total of 120 pilot plant surveys and 57 flotation tests were conducted on various streams over the two weeks on site. To complement the pilot plant work a full plant survey was conducted and a series of mini circuit block surveys to allow a comparison between the current operation and the proposed circuit changes.

**Rougher 1 Concentrate**

Initial testing was conducted on the rougher 1 concentrate while the remaining tie-ins were completed for the flash concentrate stream. The rougher 1 concentrate was tested with and without regrind in the pilot plant and through various pH and reagent conditions in the laboratory (Figure 7).

![Figure 7 – Pilot Plant performance rougher 1 concentrate](image)

The combination of the pilot plant data the laboratory flotation tests indicated that regrind had a small but significant effect on the grade recovery curve of copper with a less significant effect on nickel. Variables such as collector addition rate were evaluated and its addition was significant on improving recovery at the expense of grade in the regrind case. Diethylenediamine (DETA) and sodium sulphite (SS) were also evaluated and determined to have a significant impact on grade although testing was only conducted on the laboratory samples. Excess DETA additions resulted in almost complete depression of pentlandite and indicated it may be useful if copper nickel separation is ever considered. Optimisation of DETA dosage was conducted as the two weeks progressed to restore the recovery lost after its addition. The recovery of pyrrhotite to the rougher 1 concentrate is very high and the easier targets for pyrrhotite removal are the streams where regrinding is more effective for its removal. Neither pH nor DETA resolved the high pyrrhotite flotation in the rougher 1 concentrate at the levels tested.

![Figure 8 – Laboratory flotation of rougher 1 concentrate](image)
Flash Concentrate

The next stream to be tested was the flash concentrate. The sample was found to be very coarse and the lines sanded on occasion. Due to this materials handling challenge, all the pilot tests were conducted with regrinding while some laboratory tests were conducted without regrind. Analysis of the laboratory tests determined that regrinding had a significant impact on the cleaner grade recovery curve of the flash concentrate and the work continued with regrinding only tests in the pilot plant. Three stage laboratory cleaning of the flash concentrate at low solids density (referred to as dilution cleaning) resulted in no separation from pyrrhotite nor did modification of the pulp pH in the case with no regrind. Optimum conditions for the flash concentrate were regrinding with the addition of DETA/SS at a pH of 9.5. It is expected that regrinding with DETA/SS and Jameson Cell flotation will provide a superior grade recovery curve. In all cases the grade of the flash concentrate could be significantly improved and as this stream represents over a third of the final concentrate this will result in a significant shift in the grade recovery curve for Savannah Nickel.

Rougher 2 Concentrate

Laboratory tests on the rougher 2 concentrate with and without regrind clearly showed the improved grade recovery curve position following regrind. For this reason this stream was tested with regrind for all pilot tests. In conjunction with the pilot plant operation plant block surveys were conducted to measure the performance of the existing plant cleaner.
The demonstration in the pilot plant of the improvement in grade recovery position with regrind is shown in Figure 10. The pilot plant produced a superior grade recovery curve for both copper and nickel compared to the laboratory. The difference is likely the result of the superior cleaning efficiency of the Jameson Cell compared to the laboratory single stage cleaner flotation. DETA/SS was able to achieve a similar response to regrind alone however the regrinding in combination with the Jameson Cell without DETA/SS produced the highest grade.

Scavenger Concentrate

The final stream tested was the scavenger concentrate. As the scavenger concentrate consisted of the highest pyrrhotite content additional laboratory tests were conducted to try to understand the behaviour of pyrrhotite. The pilot plant was only operated for one day on this stream and a single plant survey was also conducted. The impact of various parameters investigated in the laboratory are given in Figure 11 and are compared to the performance of the pilot plant.

The laboratory testing showed that the grade of the scavenger concentrate could be improved by regrinding. The best performance was indicated by the dilution cleaning test after regrinding suggesting that the small number of pilot tests hadn’t achieved the optimum performance in the Jameson Cell that is indicated by the laboratory simulation. A significant level of entrainment after regrinding for the scavenger concentrate requires optimisation of wash water addition and air rates in the pilot tests. Increases in concentrate grade by a factor of 2 are possible by Isamill™ regrinding and cleaning.

To determine the impact of conditioning on the selectivity of pentlandite from pyrrhotite, Isamilled scavenger concentrate was conditioned with nitrogen or air or combinations of the two. The closed nature of the Isamill™ means that the discharge slurry is often devoid of oxygen when highly reactive sulphides such as pyrrhotite are ground as any oxygen present in the pulp is consumed as new surfaces are created. The Isamilled sample conditioned with nitrogen only remains at the low Eh level of -200 mV. The oxygen level was not measured but the rise in Eh is assumed to be associated with an increase of oxygen in the pulp as the pulp is conditioned with air. As the air is introduced the recovery of pyrrhotite decreases relative to the pentlandite improving the grade of the concentrate. This has implications for plant and pilot plant practice suggesting that conditioning will be required after and optimisation of air rates to achieve optimised concentrate grades.
ENGINEERING DESIGN

To enable engineering design on the preferred flowsheet a full plant survey was conducted in order to be able to accurately predict the required mass flows for sizing the Isamill™ and Jameson Cells. The full plant survey confirmed the distribution of masses from earlier work supporting the high mass from the flash concentrate. Any improvements in concentrate grade or recovery would rely on being able to shift the position of the flash concentrate grade recovery curve. The mass splits are provided in Table 1.

<table>
<thead>
<tr>
<th>Stream</th>
<th>tph</th>
<th>%Solids</th>
<th>Ni</th>
<th>Cu</th>
<th>Co</th>
<th>Mass</th>
<th>Ni</th>
<th>Cu</th>
<th>Co</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flash</td>
<td>5.6</td>
<td>33.4</td>
<td>12.0</td>
<td>5.3</td>
<td>0.7</td>
<td>4.0</td>
<td>30.8</td>
<td>29.1</td>
<td>35.1</td>
</tr>
<tr>
<td>Ro 1 Conc</td>
<td>4.9</td>
<td>44.5</td>
<td>11.4</td>
<td>7.4</td>
<td>0.6</td>
<td>3.5</td>
<td>25.9</td>
<td>35.9</td>
<td>29.0</td>
</tr>
<tr>
<td>Ro 2 Conc</td>
<td>8.8</td>
<td>34.1</td>
<td>6.45</td>
<td>3.31</td>
<td>0.31</td>
<td>6.3</td>
<td>26.2</td>
<td>28.7</td>
<td>25.4</td>
</tr>
<tr>
<td>Scav Conc</td>
<td>9.8</td>
<td>27.5</td>
<td>1.57</td>
<td>0.52</td>
<td>0.07</td>
<td>7.0</td>
<td>7.1</td>
<td>5.0</td>
<td>6.1</td>
</tr>
<tr>
<td>Ro Clnr Conc</td>
<td>6.5</td>
<td>37.4</td>
<td>8.1</td>
<td>4.3</td>
<td>0.4</td>
<td>4.7</td>
<td>24.6</td>
<td>27.6</td>
<td>24.1</td>
</tr>
<tr>
<td>Scav Clnr Conc</td>
<td>2.5</td>
<td>31.8</td>
<td>4.3</td>
<td>2.0</td>
<td>0.2</td>
<td>1.8</td>
<td>4.9</td>
<td>5.0</td>
<td>5.3</td>
</tr>
<tr>
<td>Final Conc</td>
<td>19.6</td>
<td>28.0</td>
<td>9.5</td>
<td>5.1</td>
<td>0.5</td>
<td>14.0</td>
<td>86.3</td>
<td>97.6</td>
<td>93.5</td>
</tr>
</tbody>
</table>

Based on the results of the piloting a flowsheet was proposed to improve the grade and recovery at Savannah Nickel. Using the mass splits from the full plant survey in conjunction with the laboratory results a simulation of the expected plant benefit was completed. It is proposed that the installation of the Isamill™ and Jameson Cell modified circuit will improve the concentrate grade to 10.5% Ni at the same recovery. This represents a 1% grade improvement compared to the full plant survey and a 2% grade improvement compared to the operating data for the remainder of the time on site.
The proposed Isamill™ was based on the design throughput and an energy consumption determined from both onsite testing and from an independent signature plot. The proposed Isamill™ was an M1000 with 500 kW of power. Following regrinding the functions of grade and recovery are separated by including a Jameson Cell in open circuit followed by a Jameson Cell and the existing cleaner circuit in closed circuit. This cleaner flowsheet has been published previously by Huyhn et al (2014). The benefit of the flowsheet is the immediate removal of over half the final concentrate at high grade while the remainder is floated to a discard tailings grade to reject liberated sulphide and non sulphide gangue. The final step is also to achieve final concentrate grade in a second Jameson at moderate stage recovery in closed circuit. This allows the separation of the grade and recovery to different flotation cells making the circuit easier to operate. This circuit has been employed in several base metal concentrators with great success. The Jameson Cells would be an E2532/6 and an E2514/3 cell representing rectangular cells of dimensions 2.5 x 3.2 m with 6 downcomers for the cleaner scalper and 2.5 x 1.4 m with 3 downcomers for the recleaner cell.

Pilot and laboratory plant testing in an intensive two week campaign enabled identification of opportunities to increase the grade and recovery of the Savannah Nickel operation in Western Australia. In combination with plant surveys a complete understanding of the behaviour of each stream when subjected to regrinding and laboratory flotation cleaning was achieved. The flowsheet could then be redesigned with the identified improvements. These improvements have been presented to develop a business case for the flowsheet to position the plant for higher grades and recoveries when the nickel price improves.

Regrinding was able to significantly shift the grade recovery performance of the flash, rougher 2 concentrate and scavenger concentrate. The rougher 1 concentrate showed limited benefit from regrinding however pyrrhotite flotation levels were high in this stream. The addition of the Jameson Cell on the Isamill™ discharge increased the grade compared to single stage laboratory cleaning demonstrating that entrainment remained a factor in the recovery of pyrrhotite. The combination of Isamilling and Jameson Cell flotation provided the best improvement and a flowsheet has been suggested that allows easy control of grade and recovery. A concentrate grade of 10.5 % Ni at the same recovery can be achieved.
ACKNOWLEDGMENTS

The authors wish to acknowledge the plant personnel at Savannah Nickel for their support in the execution of the plant trials in such a short time frame and acknowledge Glencore Technology and Panoramic Resources for permission to publish.

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IsaMills™ at Kalgoorlie Consolidated Gold Mines – from the M3000 to the M10000 and Replacement of the Roasters at Gidji Processing Plant

G S Anderson¹ and N W McDonald²

ABSTRACT

The IsaMill™ was originally developed to address liberation issues at Mount Isa Mines (now Glencore) operations, McArthur River and Mount Isa in the early 1990s. Following on from the industrial success at those operations, the IsaMill™ was commercialised. In the year 2000, Kalgoorlie Consolidated Gold Mines (KCGM) became the first external user of the technology with a 1.1 MW M3000 IsaMill™ installed to supplement the roster capacity at its Gidji Processing Plant. The IsaMill™ was used to ultra-fine grind about 10 t/h of pyrite flotation concentrate to 10–12 μm prior to gold cyanide leaching. Based on the success of the Gidji installation, another was installed at the processing plant at KCGM’s Fimiston operation in 2002 in a similar duty.

In 2014, a 3 MW M10000 IsaMill™ was selected to replace the roasters at Gidji as part of the A98 M emissions reduction project. The mill was commissioned in April 2015 at the design 30 t/h, allowing the immediate shutdown of the two roasters. Replacement of roasting with ultra-fine grinding completely eliminated air emissions at Gidji while maintaining overall plant production rates.

This paper details the history of the IsaMill™ at KCGM from the original M3000 to the latest M10 000. It covers the basis of the process selection, circuit design, operating and maintenance data.

INTRODUCTION

The IsaMill™ was originally developed by Mount Isa Mines (MIM, now a Glencore Company), in conjunction with Netzsch Feinmahltechnik, to address key liberation issues at the McArthur River project and within MIM’s lead/zinc Mount Isa concentrator, which both required reground sizes as low as 80 per cent passing seven microns. The first 1.1 MW IsaMill™ using inert media was commissioned at Mt Isa in 1994 and four 1.1 MW IsaMills™ became the enabling technology for the McArthur River project in 1995. The early development and implementation of the IsaMill™ is well described by a number of authors including Enderle et al (1997), Johnson et al (1998) and Harbort et al (1999). Modern internal operation of the IsaMill™ has been described by numerous authors including Anderson and Burford (2006).

By 1998 there were five IsaMills™ in operation at McArthur River, two in the lead/zinc concentrator at Mt Isa and a further six mills on order as part of the George Fisher project. The IsaMill™ program had been an outstanding success, allowing both the development of the McArthur River project and significant improvements in the grade-recovery performance of the Mt Isa lead/zinc concentrator, including a ten per cent gain in recovery and two per cent increase in zinc grade after the commissioning of the six mills. This was achieved through improved liberation and clean mineral surfaces due to the inert grinding environment (Pease et al, 2006).

The IsaMill™ technology was commercialised in December 1998 and made available to the general mining industry as a means to economically grind with inert media to ultra-fine sizes.

Kalgoorlie Consolidated Gold Mines (KCGM) is a 50/50 joint venture between Barrick and Newmont. It currently produces 700 000 up to 800 000 oz of gold annually and was formed in 1989 through the amalgamation of many individual leases and underground mines in an area known as The Golden Mile. A large-scale open cut mine, the Fimiston open pit, now popularly known as ‘The Super Pit’, was developed in place of the original underground operations, with the ore treated through the new Fimiston processing plant to produce a pyritic gold concentrate. The concentrator tailings were subject to cyanide leach at Fimiston while the concentrate was roasted to calcine and cyanide leached at the Gidji Processing Plant, 20 km north of Kalgoorlie-Boulder.

Due to a change in the mined areas in 1999, sulfur grades in the mined ore increased, resulting in a significant increase in the tonnage of flotation concentrate produced which exceeded Gidji Plant’s roster capacity (Ellis and Gao, 2002). Alternative treatment methods were investigated and evaluated, resulting in KCGM becoming the first external user of the IsaMill™ and

1. ISAuSIMM, IsaMill™ Technology Manager, Glencore Technology, Brisbane QLD 4000, Email: greg.anderson@glencore.com.au
2. Senior Metallurgist, Gidji Processing, Kalgoorlie Consolidated Gold Mines, Kalgoorlie WA 6433, Email: rmcdonald@kcgold.com.au

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the first gold application through the installation of an ultra-fine grinding and leach circuit at Gidji in 2001.

An additional 1.1 MW IsaMill™ was installed at Fimiston in 2002. The IsaMill™ was successfully scaled up to 3 MW in 2003 (Curry, Clark and Rule, 2005) and to date 115 IsaMills™ ranging from 75 kW to 3 MW have been installed across the world in lead/zinc, gold, platinum, copper, nickel, magnetite and molybdenum grinding duties.

This paper discusses the history and performance of the IsaMills™ at KCGM - from the first M3000 installation at Gidji Processing Plant to the latest M10000 installation, which has allowed the two roasters at Gidji to be decommissioned.

THE GOLD PROCESS AT KALGOORLIE CONSOLIDATED GOLD MINES

The Gidji Roaster’s flow sheet, shown in Figure 1, illustrates the process path for the Fimiston pyritic concentrates prior to leaching. The roasters were constructed in 1989 to process the sulfide concentrate from the Fimiston mill and was the first commercial application of a circulating fluid bed roaster in the gold industry. Filtered flotation concentrate at 33–38 per cent sulfur grade was trucked to the Gidji Plant where it was repulsed to 60–65 per cent solids by weight and transferred into holding tanks. The slurred concentrate was pumped at around 18–27 t/h into one of two Lurgi circulating fluidized bed roasters operating at around 650°C maintained through the following exothermic roasting reactions:

\[
4 \text{FeS}_2 + 11 \text{O}_2 \rightarrow 2 \text{Fe}_2\text{O}_3 + 8 \text{SO}_2
\]

\[
3 \text{FeS}_2 + 8 \text{O}_2 \rightarrow \text{Fe}_3\text{O}_4 + 6 \text{SO}_2
\]

\[
\text{FeS}_2 + 3 \text{O}_2 \rightarrow \text{Fe}_2\text{O}_3 + \text{SO}_2
\]

The roasting process oxidised the pyritic concentrate particles to iron oxides (calcine), breaking down the particles and allowing the encapsulated gold to be accessed in the subsequent cyanide leaching process. The leached gold was processed through an eight-stage carbon adsorption circuit and the loaded carbon trucked back to the Fimiston Plant for elution and carbon regeneration. Gold recoveries generally achieved via the roasting, cyanide leaching route were 93–95 per cent.

As part of the roasting reaction, the sulfide sulfur was released as sulfur dioxide along with quantities of mercury and arsenic. The off-gas was processed through the electrostatic precipitator and then the gas released via the 180 m high stack.

The most dominant constraint for the roasters was the strict requirements for air quality control (AQC) to meet enforced environmental limits. This necessitated shutting down of the roasters whenever the prevailing atmospheric conditions required. AQC constraints included surrounding atmospheric sulfur dioxide levels – these limits were originally as high as 2000 µg/m³, reduced down to 1400 in 1997, 700 in 2005 and just prior to closure was down to 400 µg/m³.

In a typical week each roaster might be down for 50+ hours or over 30 per cent of available time. The stopping and starting created both operational and production issues. If the AQC was prolonged there was often a chance of having to reheat with the auxiliary diesel burner to regain adequate temperature. The time required to restore temperature was as much as 116 hours from cold and as much as 44 hours from 375°C. This led to a considerable loss of production. Table 1 shows a typical period (quarter) in 2014 clearly demonstrating the level of downtime experienced by the roasters. Corresponding data from the IsaMill™ M3000 ultra-fine grinding (UPG) circuit (UPGA), which was not impacted by AQC requirements, is included for comparison.

THE NEED FOR ULTRA-FINE GRINDING

During 1999, there was a significant shift in the areas being mined within the open pit operation such that the sulfur head grade of the ore increased by around 75 per cent from 0.8 per cent to 1.4 per cent (Ellis and Gao, 2002). As a result there was a significant increase in the amount of concentrate produced from the Fimiston concentrator, which exceeded the treatment capacity and availability of the roasters at Gidji. It was identified that additional processing capacity was required to supplement the roasters. A number of alternative treatment routes were technically and economically reviewed, including

---

**TABLE 1**

Quarterly data from 2014 – Gidji Plant Roasters (R1, R2) and ultra-fine grinding circuit IsaMill™ operating.

<table>
<thead>
<tr>
<th>Month</th>
<th>Run hours R1</th>
<th>Run hours R2</th>
<th>Run hours ultra-fine grinding circuit</th>
<th>Utilisation R1 (%)</th>
<th>Utilisation R2 (%)</th>
<th>Utilisation ultra-fine grinding circuit (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>506</td>
<td>380</td>
<td>715</td>
<td>68.0</td>
<td>51.5</td>
<td>96.1</td>
</tr>
<tr>
<td>2</td>
<td>400</td>
<td>551</td>
<td>705</td>
<td>55.5</td>
<td>76.6</td>
<td>98.0</td>
</tr>
<tr>
<td>3</td>
<td>608</td>
<td>595</td>
<td>730</td>
<td>81.7</td>
<td>80.0</td>
<td>97.8</td>
</tr>
<tr>
<td>Averages</td>
<td>505</td>
<td>509</td>
<td>717</td>
<td>68.4</td>
<td>69.4</td>
<td>97.3</td>
</tr>
</tbody>
</table>

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FIG 1 – Gidji Roaster flow sheet for treating Fimiston concentrates.
addition of an acid plant, pressure oxidation, bacterial oxidation and UFG. Based on net present value calculations, UFG followed by direct cyanide leaching was identified as the best alternative to the roasting process, warranting further investigation. A large body of work had been completed on this option over preceding years demonstrating that gold recoveries of up to 92 per cent could be achieved after grinding to a P80 of 10 microns (Ellis, 2003). Further laboratory and then pilot test work was carried out by KCGM and is summarised by Ellis (2003).

TECHNOLOGY SELECTION

During the laboratory test work phase, KCGM concentrate samples had been sent to a number of stirred mill providers in order for them to provide estimates of the specific energy requirements to grind to a ten micron product. As reported by Ellis and Gao (2002), due to a wide and differing variety of operating parameters employed by the different manufacturers, it was not possible to make any conclusive comparisons other than that all samples tested indicated that the energy consumption was high to grind below 20 microns.

Following the pilot scale test work there was still uncertainty within KCGM as to which fine grinding technology to select for the process, partly due to the fact that stirred milling to ultra-fine sizes was still a relatively new concept in the minerals industry at that time. A site team made a number of visits to operating sites examining all aspects of the technologies. As a result, the technology choice was narrowed down to the IsaMill™ and the Svedala Detritor (now the Metso SMD (Stirred Media Detritor)). To try and distinguish between the two mills, a head to head on-site pilot mill demonstration was conducted. After working through different media types, grind targets and other performance parameters, KCGM’s conclusion was that there was no significant difference in the metallurgical performance of the alternative mills (Ellis and Gao, 2002).

One of the key final selection requirements for KCGM was that only a single mill with proven operational and maintenance performance could be installed for the duty as part of a risk reduction and capital cost minimisation process. As a result of the IsaMill™ having a proven operational and maintenance record at 1.1 MW scale, whereas the SMD was only available at 355 kW, the IsaMill™ became the selected technology.

GIDJI INSTALLATION AND PERFORMANCE

The Gidji 1.1 MW M3000 IsaMill™ circuit, was designed on the back of the laboratory, pilot and demonstration plant data to treat 12 t/h at a P80 of 120 microns to a product size of ten microns. It was commissioned in early 2001. Commissioning performance of the Gidji IsaMill™ circuit has been described by Ellis and Gao (2002).

Commissioned circuit

Figure 2 shows the feed preparation and commissioned UFG flow sheet at Gidji. The new IsaMill™ UFGA operated in parallel with the existing roaster circuit. The ground concentrate product from the IsaMill™ was thickened and processed through a three-stage cyanide leach circuit before joining with the leached calcine from the roaster circuit for combined processing through the eight-stage carbon absorption circuit.

Due to the internal product separator arrangement (described by Anderson and Burford, 2000), most current IsaMill™ operate in open circuit configuration with feed generally being prepared by a pre-cyclone to both increase operating density and allow bypass of material already at the target product size. The KCGM IsaMill™ was designed in closed circuit arrangement. Due to the high specific energy input required, open circuit arrangement would have created temperature issues within the IsaMill™, which is limited to a maximum 70°C to protect the internal rubber components. An additional benefit was the preference for high specific gravity (SG) pyrite (containing the gold) to report to the cyclone underflow at equivalent sizes where the lower SG gangue minerals reported to the cyclone overflow (Ellis, 2005). This essentially resulted in the pyrite being preferentially ground to a finer size than the gangue minerals, leading to a more efficient application of the grinding power to the target minerals rather than grinding gangue finer than necessary.

Grinding media

Grinding media selection for optimised energy efficiency must consider both the slurry feed size distribution and the required product size target. Primarily, the selected media must be large enough to ensure adequate top size breakage and prevent mill bogging. Build-up of unbroken coarse material in the IsaMill™ is characterised by a drop in drawn power due to locking of the charge, allowing one or more discs to rotate freely without engaging the media/ slurry mixture. This was observed several times in the KCGM pilot plant operation and has also been observed in other full scale IsaMill™ installations where inadequate media size has been used, one example being during the coarse grinding development work at McArthur River (Anderson, Smith and Strohmayer, 2011).

The coarse media required to deal with the top size was not the optimal size to grind efficiently to the fine target product size. Therefore, the desire to make the size reduction at KCGM in a single step ensured that a penalty in grinding efficiency would be incurred due to the need for a compromised media size selection. At the time of the original installation, sand media was the only identified viable economic option (Ellis, 2003) and a 6 mm top size sand media was employed.

In the years since the KCGM installation, there has been a rapid uptake of both IsaMill™ and alternative stirred mill technology – IsaMill™, for example, grew from an installed base of 17 MW at the time of the Gidji installation to over 200 MW by 2013. This growth resulted in ceramic grinding media becoming a much more cost-effective option with many suppliers now vying for a share of the market, driving prices down. Over 90 per cent of all IsaMills™ installed now use ceramic media. Ceramic media provides significant advantages over sand due to its higher density, high hardness, smoother surface characteristics and lower consumption rates, leading to benefits in grinding efficiency and mill component wear (Curry and Clermont, 2006). Critically, it is
manufactured and so can provide a much more stable and consistent product compared to sand, which is subject to natural variations that can impact on quality, consumption and therefore on mill performance in terms of both efficiency and wear characteristics.

KCGM started to investigate ceramic media use in 2006 and eventually changed both mills to operate on ceramic media by 2011. The higher density of the ceramic at SG 3.7 allows a smaller size media to be used and still generate the same breakage stress intensity inside the IsaMill™. The ceramic media also has a much smoother surface profile compared to the sand. These two factors assist to reduce the wear rate of the internal mill components, which lead to improved availability and reduced overall maintenance costs. Test work reported from a two month trial at the Finistston M3000 IsaMill™ (Blake; Gianattì and Clermont, 2012) identified a 20 per cent reduction in specific energy, a 50 per cent increase in maintenance interval and a 25 per cent improvement in plant throughput for the same target grind size. Historical consumption rates of the sand at Gidji were around 17 kg/t of concentrate, whereas the current 3.5 mm Magotteaux Keramex ceramic media has a measured consumption rate of 1.30 kg/t, which is about 15 g/kWh. It is noted that the Finiston UFG mill, also an M3000, uses a 2.5 mm Keramex ceramic media.

**Wear performance**

One of the main issues post commissioning was the initially high wear rates of discs at the feed end of the IsaMill™, which required mill shutdowns approximately every ten days for disc maintenance. The performance was significantly worse compared to the Mt Isa or McArthur River mills and was a direct outcome of the coarse and hard pyrite feed material coupled with the coarse sand media in use (Ellis and Gao, 2002). Shell liner wear was also higher than the other sites. Through a joint effort between KCGM and Glencore Technology (GT), significant improvements were made in the wear performance during the initial operating period.

There have been a number of changes to mill configuration, material development and media use since then. Currently, the M3000 IsaMill™ operates on a four to five weekly shutdown cycle, consuming approximately 14 discs and two shell liners per annum.

**Smaller diameter disc configuration**

In late 2014, the Gidji mill was configured with smaller diameter discs (SDD), which had been developed by GT to address issues at other sites experiencing higher than average wear rates as a result of processing coarse, hard feed material (Rule and deVaal, 2011). A total of two SDD were ultimately installed at the first two disc positions in the Gidji mill leading to further improvements in the mill operational stability. Long-term benefits from the SDD included measured reductions in shell liner wear rates of 50 per cent and a ten per cent increase in mill throughput at the same grind size target. Shutdown frequencies were increased from a typical 18 days to 33 days.

**FINISTON INSTALLATION**

A second M3000 IsaMill™ in a similar duty was installed and commissioned at the Finiston concentrator in 2002 to process additional excess concentrate as result of higher concentrator throughput rates and increased sulfur grades. The IsaMill™ was selected for the UFG duty on the basis of the successful Gidji installation. Based on learnings from the Gidji circuit, several changes were made to the Finiston circuit including cyclones and pumping discussed by Ellis and Gao (2002).

**EMISSIONS REDUCTION PROJECT**

The emissions reduction project (ERP) commenced in 2012 with the aim of eliminating sulfur dioxide and mercury emissions from the Gidji Processing Plant. The A986 M project included the installation of the new M10000 IsaMill™ at Gidji with associated flow sheet equipment, additional leaching tanks, decommissioning of the roasters and installation of an additional carbon regeneration kiln and new mercury scrubbing and capture system at Finiston for the carbon regeneration process.

**SELECTION, DESIGN AND COMMISSIONING OF THE GIDJI M10000 ISAMIL™**

A number of signature plot tests were completed by ALS Ammtac in Perth under various conditions resulting in a 3 MW M10000 IsaMill™ being confirmed as the selected mill for the project to treat 29.5 t/h of fresh feed to a product size of P80 12 microns. GT was contracted to supply its standard IsaMill™ package, which included the basic IsaMill™ supply (including drivetrain and associated lubrication/cooling systems) together with feed and discharge pump boxes and pumps, media collection bin and media charging system, instrumentation, associated piping, platforms and structure.

The mill circuit design was based on a two-stage closed circuit arrangement. Repulsed feed from the existing concentrate holding tanks is fed into the IsaMill™ feed hopper from where it is pumped into the IsaMill™ with grinding media, as required. Discharge from the IsaMill™ enters the primary cyclone feed hopper from where it is pumped to the primary cyclones (32 Warman CVX 100 mm) with the underflow stream gravitating to the secondary cyclone feed hopper. Underflow from the secondary cyclones (also 32 Warman CVX 100 mm) is returned by gravity to the IsaMill™ feed hopper to combine with the fresh feed.

Both cyclone overflow streams report to a new 18 m diameter thickener. The thickener underflow is pumped to the existing cyanide leach circuit with thickener overflow reporting to a dedicated process water tank for use in the new circuit. The overall Gidji flow sheet utilising the two IsaMills™ and no roasters is shown in Figure 3.

A late change to the project was the inclusion of a variable speed drive (VSD) and Toshiba motor in order to prevent potential penalties associated with power flicker issues on the local reticulated power supply. Typically the large-scale IsaMills™ operate with a fixed speed motor to reduce the capital costs – a wound rotor motor and liquid resistance starter (LRS) combination can be less than half the cost of a squirrel cage motor with VSD. However, inclusion of a VSD can provide the operator with additional control variables with which to optimise grinding and component wear performance. To date, the IsaMill™ has only been operated at its design speed but there does exist an option to investigate the impact of mill speed in the future.

Figure 4 shows an overview of the new M10000 IsaMill™ installation. Installation of the IsaMill™ and surrounding equipment was largely completed by mid-March 2015 and initial commissioning of the IsaMill™ commenced from 9 March. IsaMill™ commissioning follows a basic standard plan and checklist which ensures all items are correctly commissioned in the correct order. Within the main plan there is flexibility with the order, depending on the daily site conditions and priorities.
The initial priority was to prepare the necessary equipment to enable the uncoupled run of the IsaMill™ motor when it was available. This involved standard instrumentation/electrical and interlock checks through to the plant distributed control system (DCS) and flushing/commissioning of the mill motor bearing lubrication system. Availability of the motor for commissioning was delayed several days due to technical issues with the VSD. These were resolved and uncoupled motor testing was able to commence on 21 March. In parallel, commissioning of the IsaMill™ and surrounding equipment progressed. This included:

- standard instrumentation/electrical/interlock checks on each system through to the DCS
- operational testing of all valves
- flushing of all lines and tanks
- flushing and commissioning of the IsaMill™ bearing and gearbox lubrication systems
- flushing and commissioning of the IsaMill™ gland system
- water commissioning of the IsaMill™ media charging system (IsaCharger)
- water commissioning of IsaMill™ feed tank and pumps
- water commissioning of IsaMill™ discharge tank, pumps and cyclones
- control sequence logic (start-up, shutdown, media addition) simulation testing at the DCS.

The final step involved pumping water through the mill and actual testing of the start/stop sequences, without the motor coupled. Upon satisfactory completion of those tests, the motor was coupled to the gearbox and IsaMill™ for the final stage of water commissioning, which involved simulating slurry conditions, adding an initial charge of media into the mill and running final checks on the sequencing logic. In this testing the IsaMill™ was operated with both sets of cyclones in operation. Water testing was completed by 27 of March; however, due to other circuit issues, slurry remained unavailable to the IsaMill™ circuit and GT departed site, returning on 8 April to commence slurry commissioning.

The roaster suffered a significant failure on 5 April and the decision was made to not repair, which hastened the slurry commissioning in order to maintain site gold production. The IsaMill™ commenced treating slurry on 8 April and was quickly ramped up to the design target of 30 t/h at 2200 kW with measured thickener underflow circuit product distributions in the 11–13 micron range. As per standard commissioning practice, the IsaMill™ was shut down for an initial inspection on 12 April after 83 operating hours. The
inspection highlighted no abnormal wear conditions in the mill, so the mill was closed and operation continued.

The IsaMill™ internal configuration was changed from three small diameter discs to two small diameter discs on the second inspection, which occurred on 21 April, to permit better distribution of the grinding media within the IsaMill™.

**GIDJI M10 000 PERFORMANCE**

**Maintenance**

Since commissioning, the M10 000 has operated at 96.3 per cent availability with shutdowns occurring approximately once every four to five weeks. The current data suggests about 11 discs and less than two shell liner change outs will be required per annum. The most recent shell liner lasted 285 days.

For the M10 000, the most significant maintenance issue since commissioning has been the drive end bearing failure, which was traced to an issue related to an extended period between the mill installation and the commissioning process.

**Grinding media**

The same 3.5 mm Magotteaux Keramax media is used in the M10 000 as the M3000. Consumption rate is now approximately 1.25 kg/t of dry concentrate processed.

**Current grinding circuit performance – Comparing the M3000 and M10 000 IsaMills™**

Table 2 compares the operational parameters between the two IsaMills™ at Gidji. The M10 000 is currently operating at more than ten per cent above the original design throughput of 29.5 t/h and achieving the grind targets set by the plant metallurgists.

**Leaching performance**

When feed grade variation is taken into account, overall gold recovery at Gidji is assessed as being currently around 0.5 per cent lower than the modelled recovery and four per cent lower than the previous recovery obtained via the roasting process. The roasting process produces very porous particles once the sulfides are oxidised – essentially destroying the pyrite host. The high degree of porosity enhances the dissolution of gold by allowing access to a high percentage of the gold particles by the cyanide solution. The UFG route breaks the particles down to a fine size and induces stress into the mineral lattice allowing a high percentage of the gold to be accessed by the cyanide. Finer grinds promote further exposure; however, eventually this can introduce the formation of some surface coatings, which limit the extraction. A comprehensive study is underway to re-examine the full gold deportment and determine methods to improve the gold recovery by a minimum of two per cent to approach the same recoveries achieved under the roaster operation.

**Ultra-fine grinding versus roasting costs**

Table 3 summarises the maintenance and operating costs for the Gidji Processing Plant comparing the last 12 months of roaster operation with the M3000 circuit to the ten months up until March 2016 of the all grinding/leaching circuit (roaster decommissioned) comprising the M3000 and the M10 000. Operating cost has reduced by 8.5 per cent and maintenance cost by 23.4 per cent for an overall combined reduction of 11 per cent or nearly $13/t. On an annual basis this represents a A$4.5 M saving in operating and maintenance costs.

**Emissions reduction**

The reduction of atmospheric stack emissions associated with the operation of the roasters at Gidji has been a long-term focus of KCGM and its joint venture owners, Newman and Barrick. Following the formation of KCGM in 1989, the establishment of the Gidji Processing Plant resulted in the decommissioning of the remaining three in-town roasters. This resulted in a significant reduction in sulfur dioxide (SO2) levels in the City of Kalgoorlie-Boulder. With the decommissioning of the roasters at Gidji Processing Plant, KCGM has eliminated 170 000 t of sulfur dioxide from entering the atmosphere annually. Gidji was previously listed in the National Pollutant Inventory (NPI) top two highest mercury and sulfur dioxide sources. Replacement of the roasters with the UFG circuit means that air emissions have now been totally eliminated from the site.

**GIDJI M10 000 ACOUSTIC MONITORING**

Typically, a skilled IsaMill™ operator can determine changes and diagnose the mills performance based on listening to the sound on the IsaMill™ shell with their ear pressed directly onto the shell. However, this is a difficult skill to learn and is subject to individual differences and interpretations. In conjunction with the CSIRO, GT has developed an acoustic analyser for the IsaMill™ to allow quantitative measurement and optimisation of the media charge position within the mill. Stress waves, generated by the impact of grinding media with the internal mill shell liner, are detected by broadband accelerometers and processed into an acoustic emission mean signal power to allow comparison at various points along the length of the mill shell including the product separator compartment. The unit itself has been well described by Jackson et al (2014). Two of these units were originally installed as part of the acoustic development program. One of the acoustically equipped IsaMills™ was idled during 2014, so the opportunity was taken to shift the analyser to the KCGM M10 000 site for the commissioning process. It has proved very useful in both assisting the

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**TABLE 2**

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Units</th>
<th>M3000</th>
<th>M10 000</th>
</tr>
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<tbody>
<tr>
<td>Solids new feed</td>
<td>t/h</td>
<td>9.9</td>
<td>32.9</td>
</tr>
<tr>
<td>Recirculating load</td>
<td>wt%</td>
<td>497</td>
<td>435</td>
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<tr>
<td>New feed pulp density</td>
<td>wt%</td>
<td>52.5</td>
<td>50.0</td>
</tr>
<tr>
<td>Specific energy consumption</td>
<td>kWh/t</td>
<td>85</td>
<td>71.1</td>
</tr>
<tr>
<td>Media consumption</td>
<td>g/kWh</td>
<td>15</td>
<td>18</td>
</tr>
<tr>
<td>IsaMill™ circuit feed Fodem</td>
<td>µm</td>
<td>125</td>
<td>125</td>
</tr>
<tr>
<td>IsaMill™ circuit product Fd</td>
<td>µm</td>
<td>12.1</td>
<td>12.8</td>
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<tr>
<td>Cyclone feed pulp density</td>
<td>wt%</td>
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<td>22.0</td>
</tr>
<tr>
<td>Solids split to cyclone overflow</td>
<td>wt%</td>
<td>16.8</td>
<td>11.9</td>
</tr>
<tr>
<td>Water split to cyclone overflow</td>
<td>wt%</td>
<td>80.0</td>
<td>74.2</td>
</tr>
<tr>
<td>Cyclone underflow pulp density</td>
<td>wt%</td>
<td>51.7</td>
<td>62.9</td>
</tr>
</tbody>
</table>

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**TABLE 3**

Ultra-fine grinding and roasting cost comparison.

<table>
<thead>
<tr>
<th></th>
<th>Operating cost saving</th>
<th>Maintenance cost saving</th>
<th>Total cost saving</th>
</tr>
</thead>
<tbody>
<tr>
<td>Change</td>
<td>-8.5%</td>
<td>-23.4%</td>
<td>-11.0%</td>
</tr>
</tbody>
</table>
operators to understand the condition and operation of the mill and in diagnosing several issues.

Media in the product separator
The product separator is located at the discharge end of the IsaMill™ and essentially allows ground slurry to exit the mill while retaining the grinding media in the mill through its centrifugal pumping action (Anderson and Burford, 2006). Under ideal operating conditions, the product separator area is free of media; however, certain conditions can cause media to enter the product separator. This can reduce the performance of the product separator, eventually leading to more media in the area, loss of media from the mill, wear of the rotor and unstable power draw.

After a mill shutdown on 5 May 2015, the mill was charged to a higher power draw set point than intended due to an issue with the control system around the media charger. This immediately resulted in an unstable power draw situation with a corresponding distinct change in the acoustic reading in the product separator area. This reading indicated the presence of media in the product separator, which could be confirmed by direct listening to the mill shell and suggested that the mill had been overcharged with media. Due to the closed circuit configuration, any media lost from the mill was ultimately returned back via the cyclone underflow so the actual media load could not be reduced without draining media from the mill back into the media hopper.

Higher flow rates can push media into the product separator area as the pumping action of the rotor is overcome. Reducing the flow rate can change the hydraulic balance back in favour of the product separator, allowing media to clear the area and compress more towards the feed end of the mill. Reductions were made to the flow rate in this case with minimal effect suggesting that the media was in the rotor because the mill was full of media rather than as a direct result of any specific process condition. Another observation was that the mill density was operating around 46–48 per cent solids by weight. The density/viscosity can have a significant impact on the overall mill power draw and grinding efficiency.

The mill was shut down for inspection on 26 May, the only changes being an end-to-end rotation of the shell liner. Some additional wear was observed on the product separator related to the presence of media. Prior to the shutdown, the media bin was run as empty as possible so that media volume dumped from the mill could be inspected and monitored during the recharging procedure.

It was recommended to restart the mill at a 50 per cent solids target as this should allow a higher power draw for a given volume of grinding media in the mill. Upon restart, the only process change was the increase in density and the mill was able to reach the target power draw set point of 2430 kW with no significant acoustic measurement detected in the rotor area. The before and after trends are shown in Figure 5. From observation of the charge level in the media bin, it was estimated that once the mill had reached the power draw set point, there were at least 4 t of media still remaining in the bin that had previously been in the mill (an M10000 operating at full capacity would normally hold around 20 t of media). This confirmed that the mill had been overcharged with media. By remaining at the 50 per cent solids target, the mill was able to be operated successfully without any noise in the rotor area. Figure 5 also demonstrates the response of the acoustics to the variations in the mill power draw.

Shell liner wear
A disposable rubber shell liner supported in a 6 mm steel backer is mounted between the two outer shell halves and the end flanges. During an internal maintenance inspection,
the shell liner is rotatable about its own axis and end-to-end to allow the overall life of the liner to be maximised. The liner wears at different rates along its length and diameter dependant on the slurry/media characteristics and process conditions within the mill.

Due to the over-charging incident in May (referred to previously) there was an abnormal wear pattern on the shell liner. On 26 May shutdown directly after the over-charging period, the shell liner was rotated end-to-end and 180 degrees about its axis. Unfortunately this placed a more worn area of the liner into a section of the mill that was subject to higher wear under the normal operating conditions. As the run hours progressed during June, it was apparent that the acoustic reading at the disc four area steadily increased, in relation to the other acoustic channels, up until the afternoon of 9 June when it started to increase rapidly as shown in Figure 6. It is believed that the rapid increase corresponded with the last of the rubber wearing away and the media then starting to wear into firstly the shell liner steel backer and then into the actual mill shell, culminating in the leakage of slurry from the shell.

The acoustic system was not set up with any alarms at the time; however, on reviewing the acoustic data after the event, a series of alarms were implemented to warn of such an issue in the future. These alarms are now linked to the mobile phones of key plant personnel. This event demonstrated the ability of the acoustic system to pick up abnormal conditions within the mill and act as an early warning system, allowing potential issues to be addressed before they become problems.

SERIES GRINDING

The KCGM grind is a very arduous duty with a significant size reduction in a single step. Typical size reductions for the IsaMill™ are in the order of five to six times, whereas KCGM is around ten times. Previous test work by GT has demonstrated significant energy efficiency improvements through the use of series grinding, particularly where reduction ratios are greater than eight times and the product is ultra-fine.

Series grinding initially utilises a larger media to reduce the top size of the incoming feed down to a point where a smaller, more efficient media can complete the energy intensive grind to the final product target. At ultra-fine product targets, energy efficiency is dependent on media size. However, the media size must still be adequate to break the coarsest particles in the feed to prevent build-up and potential bagging of the mill.

The 3.5 mm ceramic media, employed in each of the parallel IsaMill™ circuits at Gidji, was a compromise between being large enough to break the coarsest particles in the feed and having enough fine media present to enable production of sufficient sub 12 micron material. A 4-5 mm media was considered more ideal for breaking the top size of the incoming feed material but would simply not be efficient to grind all the way to 12 microns. A 2 mm media is much more suited to efficiently grinding to the 12 micron target, however, it would not provide adequate breakage of the coarse end of the incoming feed.

Considering that two IsaMills™ (the new M10000 and the existing M3000) would be on the Gidji site, investigations were conducted into possible energy efficiency improvements through series grinding. Based on the power available in each of the mills it was established that the M3000 would be used to do the initial grind of the coarse feed using larger media to produce a suitable transfer size for the more energy intensive finer grind in the M10000 using the smaller media. This represented about a 25/75 power split between the two mills.

![Graph showing acoustic readings](image-url)
Samples of KCGM regrind circuit feed were sent to ALS in Perth where they were prepared for standard IsaMill™ signature plot test work – ensuring that the samples for each test were identical. Analysis of the feed indicated a \( P_{80} \) of 117 microns. A single-stage test was conducted using the standard graded 3.5 mm media and resulted in a specific energy of 60 kWh/t to grind to a \( P_{80} \) of 12 microns. A two-stage grinding test was then conducted whereby a graded 4–5 mm media was used as the first stage consuming 13.8 kWh/t to produce a \( P_{80} \) of 42 microns. The first-stage product became the feed to the second-stage grind using a graded 2 mm media, which consumed a further 35.7 kWh/t to reach the \( P_{80} \) 12 micron final grind target. The two stages combined consumed a total specific energy of 49.5 kWh/t to the \( P_{80} \) 12 micron target. This represents a 17 per cent reduction in specific energy compared to the single-stage grind using the 3.5 mm media and a potential substantial energy saving for KCGM. The \( P_{80} \) was maintained at 24.3 microns for both the single-stage and two-stage grinds where the \( P_{80} \) of 12 microns was achieved. Figure 7 shows the signature plots of both the \( P_{80} \) and \( P_{85} \) data for the single-stage and series grinds. The equations are shown for the \( P_{85} \) curves and the decrease in exponent (line slope) indicates the change in efficiency. The difference in total energy requirement at the 12 micron \( P_{80} \) target, and in the corresponding \( P_{85} \) curves, can be clearly seen and highlights the improved breakage rates of the two-stage grind.

To maintain the existing combined plant tonnage, overall throughput would be 45 t/h. The M3000 IsaMill™ would operate with 4–5 mm media and be configured with SDD to increase the disc-shell gap. Power draw would be limited to around 750 kW or about 16 kWh/t. The M10000 would operate with 2 mm media in conventional configuration and be able to input up to 60 kWh/t. If the predicted efficiency gains are realised, then there is also opportunity for overall plant tonnage to be increased.

In addition to improved efficiency, maintenance benefits may also arise. Observations by GT over many years have established that coarser feed distributions result in higher IsaMill™ internal component wear rates. The coarsest feed particles would be broken down more rapidly in the M3000 by the use of the 4–5 mm media (compared to the 3.5 mm media) in the M3000. The M10000 will not see any coarse particles due to it receiving the product from the M3000 and will only use fine 2 mm media. Both of these configuration changes will likely result in reduced wear compared to the current configurations using 3.5 mm media. Operationally, the 4–5 mm media will have a wider operating window to handle coarse material compared to the 3.5 mm media. This should result in the two-stage grind being better positioned to handle variations in feed size and hardness into the plant.

Ultimately the new M10000 circuit was installed in parallel to the existing M3000 circuit in order to simplify the commissioning process and develop operational and maintenance understanding of the new IsaMill™ at the plant. Plans are currently in place to investigate the options to transition to the series circuit.

**CONCLUSION**

Gold was traditionally produced at KCGM through roasting of the refractory concentrate followed by cyanide leaching. Since 2001, KCGM has operated the M3000 IsaMill™ technology, initially to provide additional gold output alongside the roaster at Gidji followed by a second unit at Fimiston. In 2015, with the installation and successful commissioning of an M10000 IsaMill™, the roasters were shut down with all gold production via the UFG route. This enabled atmospheric contamination by sulfur dioxide (170 000 t/a) to be eliminated. As a result of the operation no longer being constrained by AQc, March 2016 saw the highest monthly production tonnage at Gidji since 2002.

![Figure 7](image-url)  
**FIG 7** - Kalgoorlie Consolidated Gold Mines’ signature plots for single-stage (3.5 mm) and two stage (5 mm and 2 mm).
Site operating and maintenance costs were reduced by 11 per cent or A$4.5 M/au. Projects are currently underway to investigate increasing gold recovery.

A new acoustic monitoring system was added to the M10 000 IsaMill™, which has proved invaluable in assisting the plant operators to better diagnose issues within the mill. There now exists opportunities at Gidji to further improve the efficiency of the UFG circuit by combining both mills into a series grinding configuration.

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Commissioning and Ramp-up of the Albion Process at the GPM Gold Project*

P Voigt, M Hourn, D Mallah and D Turner

1. Manager - Hydrometallurgy, Glencore Technology, Level 10, 160 Ann St Brisbane QLD 4000, paul.voigt@glencore.com.au
2. General Manager - Technology, Glencore Technology, Level 10, 160 Ann St Brisbane QLD 4000, mike.hourn@glencore.com.au
3. Metallurgist - Hydrometallurgy, Glencore Technology, Level 10, 160 Ann St Brisbane QLD 4000, daniel.mallah@glencore.com.au
4. General Manager Technology, Core Resources, 44 Corunna St Albion QLD 4010

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ABSTRACT

The GPM Gold Project is located in Armenia, and consists of an open cut mine at Zod, near the Azerbaijan border, and a CIL processing plant at Ararat near the Turkish border. Mining at Zod commenced in 1976, and focused on near surface oxide ores, which overlay deeper refractory sulphides. Historical mining has now almost depleted the oxide ores, and the sulphide content of ore delivered to the processing plant at Ararat is increasing. Gold and silver recoveries through the Ararat plant were declining steadily.

GeoProMining, the owners of the project, have expanded the Ararat facility to deal with the increasing sulphide content in the ore. In 2014, GeoProMining refurbished an existing concentrator on site to recover a sulphide concentrate from the ore, and have constructed an Albion Process™ plant to oxidize the refractory concentrate. Glencore Technology (GT) provided the Albion Process™ plant as a technology package.

In July 2014, the progress of the GPM Gold Project Albion Process™ Plant was reported (Hourn, Voigt and Turner, 2014). At the time of writing, the construction of the plant was nearly complete. This paper presents an update of project progress, covering commissioning and ramp up of the GPM Gold Project.

The commissioning phase occurred over June to August 2014, which included an M3,000 IsaMill™ fine grinding plant, a 6 tph limestone milling plant, a 60 tpd vacuum pressure swing adsorption oxygen plant, a 10 m residue thickener and a 12 tph Albion Process™ oxidative leach plant. Since commissioning was completed, ramp up occurred over the following three months with downstream gold recoveries from cyanide leaching reaching over 98%.

INTRODUCTION

The GPM Gold Project

The GPM Gold Project is owned by GeoProMining Gold LLC and is located in Armenia. The project consists of an open cut mine at Zod, near the Azerbaijan border, and a processing plant at Ararat near the Turkish border. The Ararat plant has a milling and flotation facility built during the soviet era, with a capacity of 1 million tonnes annually, and a CIL plant built in 1997, with a capacity of 1.5 million tonnes annually. The gold bearing ore, mined at the Zod Mine, is transported to the process plant at Ararat via a state owned rail link.
The Zod deposit originally consisted of weathered oxide ores overlying deeper sulphides. Arsenopyrite and pyrite are the major sulphide minerals. Historical mining has depleted the oxide ores, and the processing plant at Ararat currently treats significant quantities of sulphide ore with increasing amounts of gold locked within refractory sulphides.

The mineral reserves for the Zod mine, at August 2011, were estimated to be 14.2 million tonnes at a gold grade of 4.3 g/t. The mineral resources were estimated to contain 28 million tonnes of ore in the indicated category, at a gold grade of 4.2 g/t and 16 million tonnes of ore in the inferred category, at a gold grade of 4.2 g/t.

GeoProMining are expanding the Ararat plant by re-commissioning an existing flotation concentrator to recover a sulphide concentrate from the ore, and constructing an Albion Process (Hourn & Turner, 2012) plant to oxidize the refractory concentrates. The Albion Process plant will convert the sulphides to oxides, breaking down the sulphide matrix and liberating gold and silver for recovery. Tailings from the concentrator and the Albion Process plant will be combined and transferred to an existing CIL plant to recover the gold and silver as bullion.

Refurbishment of the concentrator and construction of the Albion Process plant commenced in 2013, with commissioning completed in August 2014. The Albion Process plant will has a design capacity to process up to 110,000 tonnes per annum of refractory concentrate from the concentrator.

Deposit Geology

The Zod deposit is located in the Vardenis District of Western Armenia within a setting of volcanogenic and volcanogenic-carbonate sequences, with gabbro-peridotite intrusions that have metamorphosed to serpentinite (Konstantinov & Grushin, 1970).

Gold mineralization is associated with carbonate alteration of ultramafic rocks and is commonly hosted within hydrothermal alteration zones, represented by talc carbonate and quartz-carbonate assemblages. The ore is moderately hard with a medium level abrasion index.

Gold occurs as native free gold, finely dispersed gold in arsenical sulphides, gold tellurides and secondary native gold remaining after oxidation of sulphides and tellurides. Silver occurs in its native form in quartz, chalcopyrite and pyrite, and as silver tellurides.

The deposit has an average sulfur grade of 1.4 % w/w, with an average gold and silver grade of 4.54 g/t and 4.65 g/t, respectively. The arsenic grade across the deposit is 0.3% w/w. The majority of the sulphides occur as relatively coarse mineral grains. The dominant gangue minerals are quartz, talc and chlorite, with minor magnesite, dolomite and calcite.

Development Testwork

Development testwork for the GPM Gold Project began in 2009 initially with batch testwork, and culminated in a continuous flotation and Albion Process pilot plant run over the months May and June 2010. Approximately 4,600 kg of sulphide ore samples were collected from across the Zod ore body to support the testwork program. The samples were classified by ore type, spatial location and sample type and blended into 163 composites. The composites were then grouped into the four major orebodies identified in the primary sulphide resource – orebodies 1, 4, 16 and 23.

Diagnostic leaching and ore characterization testwork (Rohner & Andreatidis, 2010) confirmed that an average gold recovery of only 48% w/w was possible from the blended Life of Mine ore adopting conventional carbon in leach (CIL) leaching methods, and that the majority of the refractory gold was present in arsenical minerals, such as arsenopyrite and arsenical pyrite. Laser ablation work showed that the majority of the pyrite had levels of arsenic in the lattice, averaging 0.9% w/w.
Comminution testwork focused on generating comminution modeling parameters to determine the capacity of the existing crushing and grinding circuit at the Ararat plant. The ore displayed an average Bond Crushing Index of 10 kWh/t, an abrasion index of 0.085 and an unconfined compressive strength of 59 kN. The Bond Ball Mill work Index was 16.5 kWh/t and the Bond Rod Mill work index was 15.8 kWh/t. Modeling work by SMMC (Morrell, 2010) confirmed the milling circuit at the Ararat plant would be capable of processing between 0.9 and 1 Mt/a of ore from the Zod deposit, with minor refurbishment.

Batch and locked cycle flotation campaigns were completed on the testwork samples and a flow sheet consisting of a bulk roughing and single cleaning stage was developed and taken forward into a continuous pilot run. The continuous pilot plant testwork proved a sulphide recovery of 93% could be achieved from the Zod ores, at a mass recovery of 9 – 10%. Gold recovery to the sulphide concentrate was 87%, at a silver recovery of 91%. The sulphide grade of the concentrate was in the range 16 – 18%.

The analysis of the blended pilot plant cleaner concentrate is presented in Table 1.

<table>
<thead>
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The flotation tailings contained 13% of the gold, and CIL testwork indicated a gold recovery of 60% could be achieved from the flotation tailings at modest reagent demand. The CIL plant at Ararat has a capacity well in excess of the 1 million tonnes per annum treatment rate for the project, and so co-treatment of both the oxidized flotation concentrate and the flotation tailings was incorporated in the design.

IsaMill signature plot testwork on the pilot plant composite concentrate sample returned a specific grinding energy of 59 kWh/t to grind the concentrate to the target 80 % passing size of 10 μm. The current plant is operating at a coarser grind of around 12μm with a specific energy of 45 kWh/t. To date the coarser grind has not impacted precious metals recovery.

Extensive testwork was carried out to determine the best oxidative leach pH for the finely ground concentrates. The testwork examined oxidative leaching under mildly acidic conditions for selective oxidation of the arsenopyrite minerals, and leaching at a more neutral pH. All tests were carried out under atmospheric pressure, with oxygen gas as the oxidant.

Leaching at near neutral pH was ultimately chosen for the oxidative leach. Leaching at near neutral pH allowed lower cost materials of construction to be used in the leaching circuit, and resulted in a
final residue with more stable arsenic phases when tested in accordance with the USEPA TCLP protocol. Cyanide and lime demands were lowest for the residue generated under near neutral pH, and the gold and silver recoveries were higher.

The two major oxidative leach reactions observed under the near neutral oxidative leaching conditions were:

Pyrite: \( \text{FeS}_2 + \frac{15}{4}\text{O}_2 + \frac{9}{2}\text{H}_2\text{O} + 2\text{CaCO}_3 \rightarrow \text{FeO.OH} + 2\text{CaSO}_4.2\text{H}_2\text{O} + 2\text{CO}_2 \)

Arsenopyrite: \( \text{FeAsS} + \frac{7}{2}\text{O}_2 + \text{2H}_2\text{O} + \text{CaCO}_3 \rightarrow \text{FeAsO}_4 + \text{CaSO}_4.2\text{H}_2\text{O} + \text{CO}_2 \)

Confirmatory bench scale oxidative leaching testwork was then carried out on flotation concentrates from the four main orebodies at the Zod deposit. Economic modeling work that compared the capital and operating costs for the Albion Process plant at varying levels of sulphide oxidation was carried out using the batch leach test results. The modeling work showed that a sulphide oxidation of 70% returned the highest Net Present Value for the project, at a hurdle rate of 10%. The gold recovery at this level of oxidation was 93%.

Continuous pilot plant oxidative leaching testwork was then carried out on a blended concentrate. The continuous pilot testwork confirmed a sulphide oxidation of 70% was required to achieve an average gold recovery from the blended feed of 93%. The average silver recovery was 80%. A design oxidation target of 75% was taken forward into detailed design of the oxidative leach circuit. The oxygen demand for the concentrate to achieve the design oxidation of 75% was 336 kg/tonne, and the limestone demand was 326 kg/tonne. Mass and heat balance modeling indicated an average operating temperature in the oxidative leach circuit of 96°C.

**PROCESS PLANT DESCRIPTION**

**Plant General**

The Ararat Process Plant experiences a hot, dry summers and cold winters, with an absolute maximum temperature of +42°C; and a minimum of -30°C. The region is classified as semi-desert with an average rainfall of 238 mm. The maximum 10 days depth of snow mantle is 35 cm, at a design snow load pressure of 70 kg/m². The area is seismically active, and the plant is designed to survive a magnitude 7.2 earthquake. A flow sheet for the GPM Gold Project is shown in Figure 1.
Comminution and Flotation

Ore is mined at the Zod Mine by open cut methods and delivered to a run of mine stockpile for blending ahead of crushing. Crushed ore is trucked to a stockpile at a rail siding and loaded into rail cars for transport to the Ararat Plant. Ore is recovered at the plant by tippler, into ore storage bins. Ore is recovered from the bins by apron feeder and conveyed to the comminution circuit.

The comminution circuit consists of two parallel milling lines. Each line has a primary 1600 kW Autogenous Grinding mill, operating in closed circuit with spiral classifiers, followed by two secondary 630 kW Ball Mills, each operating in closed circuit with cyclones. The Ball Mills are configured in parallel. Cyclone overflow from the Ball Mills are directed to the flotation feed tank.

The flotation circuit had not been operated for over 10 years and refurbishment works were part of the overall project. The existing flotation plant equipment consisted of two banks with five cells in each, and two banks with four cells in each. Each cell is fitted with dual 45 kW agitators, with a cell volume of approximately 32 m$^3$. The detailed design, procurement and installation was conducted by GPM and their nominated engineering company separate from Glencore Technology.

Slurry is fed to an agitated 25 m$^3$ flotation feed tank which overflows to the first cell for conditioning with frother. Conditioned slurry then gravitates to Pre-flotation cells, which will be used for pre-flotation of talc and carbonaceous slimes. The pre-flotation concentrate is gravitated to the tailings pumpbox and then to the 15 m diameter Flotation Tailings Thickener.

Tailings slurry from the pre-flotation stage gravitates to a Rougher Conditioner cell, which is used for conditioning of the slurry with copper sulfate and collector prior to rougher flotation.Conditioned slurry flows to the Rougher/Scavenger cells to produce a rougher concentrate for cleaning. The tailings slurry from Rougher/Scavenger flotation is transferred to the Flotation Tailings Thickener, with the thickened underflow slurry pumped to the CIL circuit.

Rougher concentrate will be transferred to the Cleaner Flotation Bank. The two cleaner banks operate in series with both cleaner concentrates combined as the final concentrate. The final Cleaner Concentrate is pumped to the 10 m diameter Flotation Concentrate Thickener, with the thickened underflow concentrate pumped to the IsaMill circuit for fine grinding. Cleaner tails are recycled to the scavenger circuit.
Albion Process

The Albion Process is a combination of ultrafine grinding and oxidative leaching at atmospheric pressure. The Albion Process™ technology was developed in 1994 by Glencore Technology and is patented worldwide. There are five Albion Process plants currently in operation.

The first stage of the Albion Process is fine grinding of the concentrate. Most sulphide minerals cannot be efficiently leached under normal atmospheric pressure conditions. The process of ultrafine grinding results in a high degree of strain being introduced into the sulphide mineral lattice. As a result, the number of grain boundary fractures and lattice defects in the mineral increases by several orders of magnitude, relative to un-ground minerals. The introduction of strain lowers the activation energy for the oxidation of the sulphides, and enables leaching under atmospheric conditions. The rate of leaching is also enhanced, due to the increased mineral surface area.

Fine grinding also prevents passivation of the leaching mineral by products of the leach reaction. Passivation occurs when leach products, such as iron oxides and/or elemental sulfur, precipitate on the surface of the leaching mineral. These precipitates passivate the mineral by preventing the access of oxidants to the mineral surface.

After the concentrate has been finely ground, the slurry is then leached in agitated vessels with oxygen to oxidize the sulphide minerals. Leaching is carried out under atmospheric pressure, and autothermally. Excess heat generated from the oxidation process is removed through humidification of the vessel off gases.

The average nominal throughput for the Albion Process Plant is 94,007 t/a of cleaner concentrate, with a design factor of 15 % applied to the average rate to achieve a design rate of 108,108 t/a. The average gold and silver throughput is 127,000 and 131,000 ounces per annum, respectively.

The feed rate to the IsaMill Fine Grinding circuit is 12.1 t/h, with a design feed rate of 13.9 t/h of concentrate and a final 80% passing size of 11 μm. Since commissioning, the IsaMill energy demand to achieve this grind size is typically 800 kW. The available drawn power for an M3000 IsaMill is 1,120 kW and this mill was chosen for the ultrafine grinding circuit. The completed IsaMill circuit at Ararat is shown in Figure 2. The current grind size is typically 80% passing size of 11 – 12 μm with a specific energy of 45 kWh/t which results is above target downstream gold recovery.

Finely ground slurry is then pumped to an agitated ground concentrate storage tank. The oxidative leach circuit consists of nine 240 m³ Albion Leach Reactors, each with a live height of 9.4 meters and a diameter of 5.4 meters. Each reactor is agitated by a 160 kW dual impeller agitator, with oxygen delivered by a bank six HyperSparge oxygen injection lances in each reactor. The HyperSparge units are shown in Figure 3.

The slurry pH is maintained at 5.0 – 5.5 in each reactor by limestone slurry dosing.
The design rate of sulphide oxidation within the oxidative leach is 1800 kg/h. Under the near neutral pH conditions employed in the oxidative leach, sulfate is the reaction product of sulphide oxidation, with a design oxygen requirement of 3750 kg/h. The Albion Leach Reactors have all been designed to achieve an oxygen transfer rate of 4700 kg/h. The design oxygen capture efficiency in the leach train was 80%. Site survey data collected to date suggests that the oxygen capture efficiency currently being achieved exceeds 90%.

The oxygen mass transfer rate for the oxidation of the sulphide minerals is defined by the following equation (Shuler and Kargi, 2002):

\[
\text{Oxygen Transfer Rate} = K_L \cdot a \cdot (C_{\text{sat}} - C)
\]

where:

- \( K_L \) = liquid film mass transfer coefficient for oxygen into the slurry, in units of m.s\(^{-1}\)
- \( a \) = the specific gas surface area, in units of m\(^2\).m\(^{-3}\) = m\(^{-1}\)
- \( C_{\text{sat}} \) = the solubility of oxygen in the slurry at saturation, in units of g.m\(^{-3}\)
- \( C \) = the steady state oxygen level in the slurry, in units of g.m\(^{-3}\)

The “\( K_L \)” and “\( a \)” terms are typically combined in the form of a mass transfer coefficient for the system. The design \( K_L \cdot a \) for the Albion Leach Reactors is 0.12 s\(^{-1}\). Oxygen gas has poor solubility in water, and so mechanical devices such as agitators and spargers are required to assist the mass transfer.
transfer. In the Albion Leach Reactor, oxygen gas is sparged into the vessel using the HyperSparge supersonic gas injection lances. The HyperSparge oxygen injection system achieves very high oxygen mass transfer rates at the interface between the supersonic gas jet and the impinging slurry, reducing the amount of power required from the agitation system.

The agitator drawn power required to achieve the design mass transfer coefficient was determined using an empirical correlation of the form (Nielsen and Villadsen, 1994):

$$K_{La} = A \times \frac{U_s^a \times (P_g / (p_{SL} \times V))^{\beta}}{}$$

(2)

where:

- $A$ = a constant specific to the ionic strength of the leach solution
- $U_s$ = the gas superficial velocity in the reactor, in units of m.s$^{-1}$
- $P_g$ = the agitator drawn power under gassed conditions, in units of Watts
- $p_{SL}$ = the density of the slurry, in units of kg.m$^{-3}$
- $V$ = the volume of the slurry, in units of m$^3$
- $a, \beta$ = dimensionless empirical constants

The $A, a$ and $\beta$ parameters used for sizing the agitator were determined based on over 900 laboratory and pilot mass transfer tests. This correlation has been used successfully in the scale up of all operating Albion Process plants to date. A drawn power requirement of 120 kW per Albion Leach Reactor was determined using the correlation.

The residence time for the oxidative leaching circuit was designed based on the specific rate constant for pyrite leaching measured in the batch and continuous leaching testwork. Pyrite oxidation under near neutral pH conditions is first order (Singer and Stumm, 1970), allowing a simple scale up. The residence time scale up was based on the method of Henein and Beigler (Henein & Beigler, 1988). A design residence time of 40 hours was calculated for the oxidative leach circuit.

Each Albion Leach Reactor was fabricated from lean duplex alloy steel having a diameter of 5460 mm and a live height in the range 9100 – 8100 mm. The Albion Leach Reactors were supplied in modular sections for rapid assembly on site. Each Reactor was constructed from 15 panels, each with a height of approximately 2.0 m and an arc length of 5.9 m. These panels were all fabricated off site and imported to the plant site in shipping containers. Baffles, slurry risers, leach tank lids, agitator support platforms and off gas stacks were all provided as part of the modular Glencore Technology equipment supply. Assembly of the oxidative leach train was rapid, with all nine leach reactors and two slurry storage tanks complete within 8 weeks. The final two tanks were erected in approximately three days each. The completed oxidative leach train is shown in Figure 4.

Overflow slurry from the oxidative leaching circuit will gravitate via a slurry sampler to a 10 m diameter thickener and be thickened to 45 %w/w solids prior to transfer to the CIL circuit. Thickener overflow is returned to the leach circuit to compensate for evaporative losses in a density control loop.
A limestone plant with a capacity of 6 t/h was installed to generate limestone slurry for neutralizing duty. Limestone for the oxidative leach will be milled to an 80% passing size of 75 microns in a 132 kW overflow ball mill operating in closed circuit with cyclones. Cyclone overflow will report to a 150 m³ agitated distribution tank and be circulated through the oxidative leach train by a ring main. Individual dosing lines will add limestone slurry to each Albion Leach Reactor. The limestone distribution tank was a 150 m³ ZipaTank zip join tank – the first of its kind the in the world. The tank was erected in approximately 5 days and was internally sealed with specially selected paint. The joins sealed on the first filling. The limestone distribution tank is shown in Figure 5.

Two 60 t/d VPSA oxygen plants will operate in parallel to provide oxygen to the Albion Process Plant. Oxygen will be delivered from each plant at a maximum flowrate of 1,745 Nm³/h, at a purity of 93% v/v.

The thickened oxidative leach residue and thickened flotation tailings will report to a 100 m³ mixing tank and be blended prior to feed to the CIL plant. The CIL Plant will process 137.5 t/h of feed comprised of oxidized residue and flotation tailings. All six existing CIL tanks will be utilized, providing a total residence time in the CIL circuit of 41 hours. The CIL Plant is expected to consume
5.3 kg/t of sodium cyanide and 10 kg/t of lime. Carbon levels in the CIL Plant will be 10 – 15 kg/m³, with a design carbon loading of 2,500 g/t. Carbon movements will total 7.5 t/d, and the existing dual AARL elution circuits will be used for carbon processing.

CIL Plant tailing gravitates to a cyanide destruction plant prior to being pumped to tailings. The tailings will be deposited within the existing tailings impoundment, approximately 6 km from the Ararat plant site.

Plant control is achieved through a Distributed Control System (DCS) located in a centralized control room between the concentrator and Albion Process plants. Training for field and control room operators was provided by GT and sub-contractors as part of the commissioning process. The control room is shown in Figure 6.

![Figure 6 – Central control room for concentrator and Albion Process™ plant.](image)

### PROJECT STATUS AND PLANT PERFORMANCE

The Albion Process plant was provided to the GPM Gold Project as a Lump Sum technology package by GT. The package included all detailed design, mechanical equipment, electrical, instrumentation and control equipment, structural steel, flooring, handrails, piping and valves. The scope of supply includes the fine grinding plant, oxidative leaching and thickening plant and the supporting limestone, oxygen, flocculent and caustic reagent plants.

Mechanical design was completed in December 2012, with the majority of mechanical equipment and fabricated components delivered to site by May, 2013. Site civil works were completed in March 2013. Construction was completed in April 2014.

The pre-commissioning phase was conducted during April/May 2014 with wet commissioning commencing in May/June 2014 and completed in August 2014. The commissioning was managed and coordinated by Glencore Technology. The commissioning team comprised three permanent Glencore Technology personnel (manager/process engineer, mechanical engineer and instrumentation engineer) supported by equipment specialists brought to site during crucial commissioning events. The GPM site team provided all other support.

The main setback during commissioning was the failure of an oxygen plant blower which had a lead time of 12 months to replace. Dual oxygen plants were supplied to the project, each with the capacity to oxidise 70% of the design sulphide feed, and so the blower failure has not impacted on plant throughput to date. The blower will be repaired and the second oxygen plant will be in service by May 2015.
The main setback for ramp-up has been the lack of feed quantity and quality from the refurbished concentrator. A project is in place to improve concentrator performance with expected results by October 2015.

As of March 2015 the refurbished concentrator was running at around 60% capacity with recent assistance from GT during March increasing throughput by 30%. The Albion Process plant performance has not been impacted by the slow ramp up of the concentrator, with the plant regularly achieving 95% gold recovery with around 50% sulfur oxidation.

A survey of the nine Albion Leach Reactors was collected to determine tank-by-tank sulfur oxidation and resulting gold extraction. The sulfur oxidation was determined using a Leco sulfur analyzer. Gold extraction was determined by subjecting each collected sample to a bottle roll test at the GPM laboratory and cross checked with an agitated CIL test at hltesting laboratory in Brisbane. The sulfur oxidation against gold recovery from the plant and compared to pilot plant results is shown in Figure 7.

![Figure 7 – Sulfide oxidation against gold recovery comparing pilot plant data and actual operating data.](image)

Testwork done on individual components of the pilot plant feed are consistent with this level of oxidation and corresponding gold recovery.

A profile of sulfur oxidation and gold recovery was collected down the nine leach reactors. The data are presented in Figure 8.
Figure 8 – Profile of sulfide oxidation and gold recovery down the leaching train. Plant is on reduced rates.

Figure 8 shows the profile of the sulphide oxidation and gold recovery down the leach train. Although one oxygen plant is operating, there is sufficient oxygen available to increase the throughput rate if feed was available from the concentrator.

The current oxygen consumption is operating at around 215 kg / t concentrate which is below the design value of 336 kg / t owing to higher than design oxygen utilization and lower oxidation levels for this feed material compared to the design case.

Current limestone consumption is very low owing to higher entrainment of acid consuming gangue with the flotation concentrate. Steady feed composition at full production rates will allow limestone consumption to be better analysed.

Cyanide consumption in the CIL plant is within the range expected during pilot testwork. The plant is currently operating at 1.8 to 2.2 kg sodium cyanide per tonne of feed to the CIL plant which is the combination of leach residue and flotation feed.

**CONCLUSIONS**

The GeoProMining Albion Process™ plant was commissioned successfully over a 14 week period. The plant is achieving greater than 95% gold recovery in the cyanide leaching plant, consuming 1.8 – 2.2 kg cyanide / t combined leach residue and flotation tails.

The plant is running on reduced rates due to concentrate feed availability, as mine development has been slower than planned. GT continues to work with GPM to improve the concentrator performance. At the end of April 2015, GPM achieved a 30% increase in concentrator throughput with the assistance of GT. The second oxygen plant commissioning is scheduled for June 2015, and full plant throughput should be achieved by July 2015. An update will be provided after the plant is at full capacity with more comprehensive performance data including oxygen and limestone consumption as well as sulfur oxidation, oxygen utilization and resulting gold recovery.
ACKNOWLEDGEMENTS

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REFERENCES


Premium coal fuels with advanced coal beneficiation

Louis J Wibberley¹ and Dave Osborne²
¹Leader DICE Development Program
CSIRO Energy Technology, Mayfield West 2304, Australia
Email: louis.wibberley@csiro.au
² Coal Technology Consultant and Managing Director,
Somerset International Australia, Brisbane, 4000, Australia
Email: dosborne@somersetpty.com

Abstract
Research and pilot plant trials over the last 5 years have shown that a range of premium quality coal fuels, with ash contents as low as 1%, can be produced economically from a wide range of coals (including coal tailings). This capability can improve both the economic and environmental performance of coal through increased grade recovery, recovery of saleable coal from existing tailings emplacements, and new higher value products such as a fuel oil replacement in boilers, enhanced coking blends, and for higher efficiency power generation using the direct injection carbon engine (DICE). While a conventional coal preparation plant can be used to produce suitable ultra-low ash coal for these fuels, changes to both the philosophy of coal preparation and operating strategy are needed for achieving the best results. Further improvements are possible via the application of the latest milling and dewatering technologies. These fuels also require a rethink of the supply logistics. The paper discusses findings from both laboratory and pilot scale trials in Australia, in the context of new export products for DICE and boiler fuels.

Key words: ultra-low ash coal, coal grain analysis, coal slurry fuel, advanced beneficiation, DICE, high efficiency engine

INTRODUCTION

Producing very low ash coal products has for many years been a challenge for coal treatment specialists and researchers. Much of the early work in the 1970s and 1980s used chemical cleaning processes involving leaching with acidic or caustic solutions, most of which proved uneconomic when scaled up to commercial application. The most successful of these include the AMAX 2-stage leach process developed with US-DOE funds in the mid-1980s¹ and the Australian UCC process².

The use of physical cleaning to ultra-low ash levels, using more conventional coal processing technologies, has proven more challenging. Old school thinking typically regards ash contents below 2-3%, as both technically and economically unviable. This is due to several factors/misnomers:

- The so-called “inherent ash” content of coal is usually regarded as the lowest ash possible
- Lower ash contents require finer grinding to increase liberation which is costly and produces a product for which there are few markets
- Flotation of ultra-fine coal can be problematic, and requires high reagent consumption

- Fine coal products are inevitably high moisture (> 35%) which means costly dewatering and/or drying to produce saleable products
- Ultra low ash coal products are uneconomic to produce

All of these factors are incorrect, or at best very misleading, as recent research/pilot plant tests have shown.

Fine coal cleaning is generally regarded as being the treatment of coal sized below 0.5mm (500µm) but the challenges have tended to descend further down the size distribution to 0.1mm (100µm) and lower. Most recently, with the widespread use of micro-flotation using the Jameson and Microcell technologies, the elusive size fraction is now <0.05mm (50µm) and this is probably where the current research and development focus lies. Such a size range usually means that coal particles are extensively liberated from the mineral content and are recoverable, providing that the concentrate can be recovered and dewatered to a commercially acceptable level.

Not surprisingly, much of the research and development that has occurred in ultra-fine coal cleaning has taken place for higher value products, often metallurgical, and almost all techniques for both cleaning and dewatering have undergone some form of improvement or development during the past decade in this context. These improvements have been in all areas - optimised design, improved materials of construction and improved wear resistance, process control and monitoring, integrated automation and sampling/analysis.

This paper describes recent work carried out at laboratory and pilot scale in Australia, that has successfully produced coal concentrates with <2% ash content from a wide range of coals and tailings, using a combination of fine grinding and froth flotation. The potential to achieve even lower ash levels is also discussed. The main objective of this work has been to produce a stable, coal-slurry fuel (i.e. micronised refined coal or MRC) for use in DICE, an application which demands the lowest possible ash content in the slurry fuel (preferably <1%) and very fine particle size distribution (typically d80 of 20-40µm). This is much finer than the conventional coal water slurry fuel (CWSF) for boilers or gasifiers.

**CHEMICALLY CLEANED COAL**

Ultra-clean coal can be produced by chemically cleaning coal, and a 2- and 3-stage caustic leach process has been successfully demonstrated in Australia by UCC Energy Pty Ltd. A significant feature of the process is that the coal needs only to be coarsely milled (typically to -2mm) to provide sufficient contacting: all other processes for producing ultra-low ash coals require a higher degree of milling. The main process consumables are sulphuric acid, and lime; caustic soda is regenerated in the process. The waste materials are environmentally benign (gypsum and calcium alumina silicates) and the final ultra-clean coal product has 0.5-0.7% ash (with less than 0.2% possible for some coals). UCC Energy Pty Ltd (a wholly owned subsidiary Yancoal Australia Pty Ltd) owns and has operated a pilot plant at Cessnock in the Hunter Valley, New South Wales. The nominal 350kg/h facility (see Figure 1) was commissioned in 2002.

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Fuel from this plant has been used for longer duration tests in an adapted diesel engine (ie DICE) at the CSIRO, and scale up of the process is now under consideration by Yancoal and partners.

PHYSICAL CLEANING

Although chemical cleaning has been proven to produce very low ash coal products for the highest value applications (e.g. electrode carbon and MRC for DICE – especially smaller engines), the process is more costly than physical cleaning, is less tolerant of coal rank, and requires a coarser lower ash feed coal for optimum results. Physical cleaning is therefore preferred for the production of coals with ash contents <2% from a variety of coals and tailings, and is the focus of the present paper.

Liberation – removing the “inherent ash” constraint

Efficient analysis of the occurrence of mineral matter within the feed coal is key to economically producing premium coal products by physical cleaning. CSIRO in Australia has developed an optical reflected light microscopy system for assessing coal petrography samples. This system collects and creates mosaic images so that quantitative information can be obtained on individual coal/mineral grains. Size and compositional information (the amount of vitrinite, inertinite, liptinite and mineral type) is obtained for each particle, and this information can then also be used to estimate the density and ash value of each particle (both macerals and minerals) and that of the sample.

Although coal grain analysis (CGA) is generally conducted on a small representative sub-sample of -1 mm material, the technique has also been successfully used to characterise particles up to 4 mm in size (a size which perhaps report to a middlings circuit), and for samples which have been micronised to give particle size of less than 20 µm (e.g. for producing ultra-low ash concentrate by flotation).

For the ultra-fine material, CGA is particularly useful to assess the degree of liberation and therefore assist in optimising the beneficiation process to produce the lowest possible ash products (for uses such as DICE, or even direct carbon fuel cells®).

For advanced beneficiation, CGA enables liberation to be assessed and the expected yield at different target ash levels to be estimated. As the CGA information also provides detail on the intrinsic and entrained minerals in a sample, it benchmarks the lowest ash value which could be obtained if all entrained minerals are removed from the sample at that top-size.

An example of its application in the production of MRC for DICE is given in Figure 2 below. The characterised CGA images shown illustrate the extent of liberation for raw tailings (-250µm) and the corresponding micronised concentrate with a d80 of 15µm. The results show that milling has enhanced significantly the particle liberation, and has produced essentially single component grains. This indicates that it is physically possible to reduce the intrinsic mineral being carried over to the product. This implies that an appropriate circuit of grinding and flotation operations should be capable of producing a very low ash product concentrate. In this case sufficient liberation has occurred to enable a product coal with less than 1% ash to be obtained, provided entrainment within the froth can be eliminated. Although it is anticipated that the fineness of the entrained mineral particles will make separation more difficult, there are a number of process optimisations available to achieve this (e.g. use of reagents which enhance the hydrophobicity of the coal and hence allow more aggressive application of wash water to be used, optimisation of the pull between flotation stages). Both have been shown to be effective in practice at the pilot scale.

The results obtained from several CGA studies with a pilot plant, and other minerals tracking studies by the CSIRO with 9 Australian bituminous coals/tailings, have provided a valuable insight into the separation and recovery performance of the current beneficiation work, by tracking the selective recovery of individual grain types. In particular, this has enabled optimising the milling and flotation steps for processing raw tailings samples.

Overall, CGA has proved a valuable method for applying research in a commercial environment.

**Millling/flotation process approach**

In order to achieve lower ash levels, a liberation step needs to be introduced by fine grinding. Typically fine grinding for coal water slurry fuels is by ball mills; for example, ball mills are commonly used in China where over 30Mt of slurry fuel is currently produced annually with ash levels below 4%. However, to enable even lower ash contents and maximum liberation (and with a reduced energy consumption), the present work has used bead mills (IsaMills) for both laboratory and pilot scale tests.
Using this mill, a 3 tonne/day pilot-plant, see Figure 3, owned and operated by Glencore Technology (formerly Xstrata) has been used for proving the feasibility of producing ultra-low ash coal from freshly generated raw coal tailings. The plant is sited at Bulga mine, a large thermal coal operation in the Hunter Valley. The pilot plant comprises two Jameson flotation cells, a small IsaMill and a membrane filter press, which can be configured to simulate a number of circuit designs.

![Fig. 3 The 3 t/day coal tailings treatment pilot-plant (Courtesy Glencore)](image)

A range of different coal feed types have been tested with raw coal ash contents ranging from ~30% feed ash (ad) from easily treated coal seams, to much higher ash coal seams containing difficult-to-separate colloidal clays. The plant is usually configured in a rougher-scavenger circuit where the tailings from the rougher (primary cell) feeds to the scavenger (secondary cell).

The cleaning efficiency of this circuit is clearly shown by the difference in colour of the two product streams in Figure 4 below; the black concentrate is high in vitrinite and the white tailings stream is predominantly kaolinite clay.
A secondary effect of cleaning is a marked improvement in dewatering: Fine mineral matter and clay in a coal concentrate significantly reduces filtration rate and increases product moisture. By removing much of this material, a significant reduction in product moisture can be achieved – essential for metallurgical and briquette products.

Cost effective and efficient dewatering is thus essential to producing premium coal products, for treating both the concentrate and barren tailings streams. Dewatering options available are listed in Table 1. Recent advances in “by zero” fines dewatering has enabled total moistures of flotation product to be reduced to a target whereby it has become commercially viable to include flotation concentrates into the final thermal coal product at a greater number of mine sites. Tailings dewatering still has some challenges, but high-G solid-bowl centrifuges offer promise for on-line dewatering.

Table 1 Equipment Used for Dewatering Coal (adapted from Table 13-1, *The coal handbook*6)

<table>
<thead>
<tr>
<th>Equipment</th>
<th>Footprint size</th>
<th>Throughput (t/h dry solids)</th>
<th>Product moisture (% w/w)</th>
<th>Feed preparation</th>
<th>Application</th>
</tr>
</thead>
<tbody>
<tr>
<td>High frequency screen</td>
<td>0.6-2.4 x 3m</td>
<td>10-100 t/h</td>
<td>15-25</td>
<td>cyclone underflow</td>
<td>fine coal</td>
</tr>
<tr>
<td>Screen scroll centrifuge</td>
<td>0.5-1.5m dia.</td>
<td>45-100 t/h</td>
<td>11-18</td>
<td>cyclone underflow</td>
<td>fine coal</td>
</tr>
<tr>
<td>Horizontal vacuum belt filter</td>
<td>75-150m²</td>
<td>50-130 t/h</td>
<td>20-30</td>
<td>flocculation</td>
<td>ultrafine coal</td>
</tr>
<tr>
<td>Screen bowl centrifuge</td>
<td>1.1m dia. x 3.3 m long</td>
<td>20-60 t/h</td>
<td>16-27</td>
<td>thickening</td>
<td>ultrafine coal</td>
</tr>
<tr>
<td>Solid bowl centrifuge</td>
<td>1.1 m dia. x 3.3 m long</td>
<td>15-20 t/h</td>
<td>15-20</td>
<td>thickening</td>
<td>ultrafine coal slimes</td>
</tr>
</tbody>
</table>

The economic argument for pursuing fines recovery in this application is now very compelling. Technology advances, particularly in dewatering, allow ultrafine product to be included in final product streams without penalising product quality (or introducing handling difficulties) as shown by the simplistic revenue scenario below.

Based on the following assumptions:

- Raw plant feed containing 10% passing 0.1 mm
- Thermal coal operation of 16 Mt/y ROM (1.6 Mt/y of raw feed currently sent to waste)
- Assuming a nominal 50% yield which equates to 0.8 Mt/y of potential saleable product
- At a benchmark price of US$85/product tonne (equivalent to US$69M revenue loss per annum, excluding freight, port, tonnage adjustments, etc.)

Including conservative capital and operating costs produces a reasonably attractive investment opportunity as given in Table 2 below.

Table 2  Economic evaluation of brownfields flotation installation

<table>
<thead>
<tr>
<th>Capex</th>
<th>Operating</th>
<th>Coal rate</th>
<th>Direct costs</th>
<th>Tax rate</th>
<th>Discount rate</th>
<th>NPV</th>
<th>IRR</th>
<th>Payback</th>
</tr>
</thead>
<tbody>
<tr>
<td>$M</td>
<td>$/t feed</td>
<td>t/h</td>
<td>$/t</td>
<td>%</td>
<td>%</td>
<td>$M</td>
<td>%</td>
<td>years</td>
</tr>
<tr>
<td>50</td>
<td>15</td>
<td>230</td>
<td>30</td>
<td>30%</td>
<td>10%</td>
<td>78</td>
<td>51</td>
<td>2.5</td>
</tr>
</tbody>
</table>

Future direction – reducing product ash and moisture

The challenge of cost-effectively recovering a saleable fines component from tailings has been with us for many years, and periodically an apparent solution emerges. Glencore Technology has been operating the pilot plant described above for over 4 years, and this plant incorporates the combination of ultra-fine
grinding by an Isamill and Jameson Cell flotation technology\(^7\). The inclusion of a fine grinding stage enables slurries to be prepared whereby mineral components are almost completely liberated from the carbonaceous material, thereby facilitating recovery of a highly concentrated ultrafine, low ash coal product, with ash levels well below traditional inherent ash limits—even from tailings.

This combination has already been proven capable of achieving very good combustible recoveries (material dependent, but normally over 90\%) for coal derived from the raw tailings stream. The milling step produces a feed with a \(d_{80}\) of 15-30\(\mu\)m, enabling enhanced flotation recovery by a combination of increased liberation and the formation of fresh surfaces on the ultra-fine coal particles. This also reduces reagent consumption to levels significantly below traditional fine coal flotation.

Further enhancement by the addition of improved ultrafine dewatering of the flotation concentrate using a membrane filter press, or equivalent, and appropriate mixing system has resulted in the preparation of ultra-low ash, highly stable slurries with solids concentrations over 60\% (w/w). This product should be very suitable for DICE, a potential large new market for a range of power generation markets:

- To replace fuel oil and natural gas,
- To provide highly efficient, highly flexible and modular coal-based power to backup increased renewables (giving an extremely low CO\(_2\) and secure electricity system), and
- For incremental, low CO\(_2\), coal-based generation capacity\(^8\).

In the shorter term, slurries with a slightly coarser particle size distribution (typically with a \(d_{80}\) of 75\(\mu\)m) can be prepared in a similar way to create conventional coal-water slurry fuels, at over 70\% solids, for direct firing into boilers as a replacement for heavy fuel oil (HFO) and for gasifiers.

Figure 5 below shows a nominal flowsheet based around an Isamill for micronising, and the Jameson Cell for separation. Adding the milling step can be optional for boiler grade product, but is essential for producing MRC for DICE. Various dewatering options, as described earlier, need to be carefully evaluated to achieve the desired solid-liquid outcome for each coal source/product combination. Alternatively, a briquetted product can be included in the normal product stream, thus avoiding slurry or fines related problems in product handling and transportation.

The integrated plant design thus has the functionality of the dual product offering, i.e. coal briquettes that can be added to the conventional product and/or coal-water slurry fuel for heavy fuel oil replacement for boilers, or for more innovative, value added applications such as DICE.

To facilitate logistics, and to enable early adoption of the technology, the concept of coal water slurry supply chains is being promoted to industry, whereby slurry fuels employ existing heavy fuel oil infrastructure to transport and store the fuel at the customer’s facility. The use of these systems was been extensively demonstrated in Japan in the 1990s, and more recently in China.

\(^{7}\) Mercuri, F., Osborne, D.G. and Young, M; 2014. The Future of Thermal Coal Flotation. Australian Coal Preparation Society conference 2014, Gold Coast, Qld.

Conclusions

For more than two decades the appeal of so-called “deep cleaning” of coal via liberation and subsequent beneficiation has been recognised in terms of the significant downstream improvements that would result - maximised resource recovery, minimised transport and handling costs, numerous end-user process improvements, reduced maintenance and wear, lower environmental impacts and sustainable improvements.

While the capability of ultrafine particle separation has matured via progressive improvements in froth flotation, the capability of dewatering the concentrate has continued to be the major barrier until recently. The emergence of larger capacity membrane filter presses, hyperbaric dewatering via decanter centrifuges or large disc filters now offer commercial solutions.

Binderless briquetting has also progressed to machines with capacities of up to ~40t/h for fine coal applications, and providing an alternative product for conventional transportation infrastructure.

The manufacture of stable coal-water slurries with over 65% solids content, and MRC slurries with over 60% solids content, have also reached commercial adoption; such products can also be produced from tailings.

With these technical barriers now overcome, the scene is set for these new applications to progress to become the new generation of clean coal technologies for a wide range of applications, with key economic drivers being a much higher fuel cycle efficiency (i.e. lower carbon intensity), highest grade recovery, and lowest solid waste disposal.

References


**Acknowledgements**

The authors acknowledge the contributions provided by Glencore Technology, Glencore Coal, Yancoal and the CSIRO towards the preparation of this paper and for permission to include figures and data from other recently published papers and articles on this subject.

Dr Osborne thanks Somerset Coal International for encouragement and support in attending and participating in the 2015 Clearwater Clean Coal conference.
Premium coal fuels with advanced coal beneficiation

Louis Wibberley - CSIRO
Dave Osborne – Somerset International

Contents

1. Introduction
2. Ultra-fine Clean Coal (UFCC) Defined
3. Chemically Cleaned Coal
4. Physically Cleaned Coal: Liberation
5. Flotation/Milling process approach
7. Conclusions
8. Acknowledgements
Introduction

• Greater resource recoveries are being sought by mine operators to maximize investment returns.
• Current industry trend - “by zero” recovery of fine coal to maximize resource yield and minimize environmental footprint.
• Fine coal size fraction faces the greatest barriers towards qualification as a product component.
• However, advances in liberation, flotation and dewatering create new opportunities for thermal coal operations. So............

What are the historical barriers and how are they being overcome?

Why have advanced processing options now become viable?

Ultra-fine Coal Beneficiation

Two distinct approaches:

1. Chemical cleaning – coal structure is changed via chemical decomposition – potential is <0.2% ash residue.
2. Physical cleaning – coal structure not changed, but comminution may be applied for liberation – potential is <1% ash residue.

“Old school” thinking typically regards ash contents below 2-3%, as both technically and economically unviable because:

• “Inherent ash "of coal is usually regarded as the lowest achievable ash content
• Lower ash requires milling to finer particle size to increase liberation
• Flotation of ultra-fine coal can be problematic often requiring higher reagent dosages
• Fine coal concentrates are inevitably high in moisture ( > 35%) which means costly dewatering and/or drying to produce saleable products.
Chemically Cleaned Coal

- Caustic leach process has been successfully demonstrated in Australia.
- Similar to the well-proven Bayer alumina process and also the AMAX 2-stage leach process developed with US-DOE funds in the mid-1980s.
- Ultra-low ash residue <0.2%
- Uses include slurry fuel or briquettes
- Costly option difficult to justify in current climate.

Ultra-clean Coal pilot-plant, Cessnock, NSW (Courtesy UCC Energy Pty Ltd.)

Liberation - Inherent ash constraint

CSIRO has developed an optical reflected light microscopy system for assessing coal petrography samples.

System collects and creates mosaic images so that quantitative information can be obtained on individual coal grains, i.e. Coal Grain Analysis (CGA).

CGA generally requires only small representative sub-samples of <1mm material.

Size and compositional information, i.e., macerals vitrinite, inertinite, liptinite and minerals can be determined for each particle.

Information can then also be used to estimate % mass, density & “ash” value of each particle.
Liberation – CGA Images Confirm Status

Characterised images for Raw Coal Tailings Feed compared with Final Concentrate. (Courtesy QCAT-CSIRO)

Flotation/Milling process approach

A 1tonne/h pilot-plant, owned and operated by Glencore Technology (formerly Xstrata) comprising

• two Jameson flotation cells,
• a small IsaMill, and a
• membrane filter press, etc.,
• Located at a large thermal coal operation in the Hunter Valley.

Currently testing freshly generated raw coal tailings to produce coal water slurry fuels
Milling and Sub-50μ Coal Flotation

NSW Coal Water Slurry Fuel (CWSF) Pilot Plant
- Successfully produces stable Coal Water Slurry Fuel (CWSF) from coal tailings
- CWSF can then be further refined to produce very low ash (<1% ash) Micronized Refined Coal (MRC)
- MRC produced from 2011 - 2015 for diesel engine tests
- Process information obtained also used for design of CWSF modules including a package plant and fuel handling rig.

Pilot Plant Fuel Production
## Dewatering Technology

<table>
<thead>
<tr>
<th>Equipment</th>
<th>Throughput (dry solids)</th>
<th>Product Moisture (% w/w)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Horizontal belt filter</td>
<td>50-130 t/h</td>
<td>20-30</td>
</tr>
<tr>
<td>Screen bowl centrifuge</td>
<td>20-60 t/h</td>
<td>16-27</td>
</tr>
<tr>
<td>Centribaric centrifuge</td>
<td>15-20 t/h</td>
<td>15-20</td>
</tr>
<tr>
<td>Vacuum disc filter</td>
<td>50-150 t/h</td>
<td>20-32</td>
</tr>
<tr>
<td>Hyperbaric disc filter</td>
<td>30-150 t/h</td>
<td>17-25</td>
</tr>
<tr>
<td>Solid bowl centrifuge</td>
<td>10-18 t/h</td>
<td>18-25</td>
</tr>
<tr>
<td>Membrane Filter press</td>
<td>15-30 t/h</td>
<td>14-32</td>
</tr>
</tbody>
</table>

Key objective is to achieve a cake moisture of 20% or lower for ultrafines.

## Emerging Coal Fines Treatment Circuit

![Emerging Coal Fines Treatment Circuit Diagram](Diagram)
Wider Fines Treatment Options

Coal Supply Chain

- Current Coal Supply Chain (CSC) has been hampered by an inability to dewater and efficiently transport fine coal.
- Innovative approach - recover and use all the ultrafines via coal-water slurry thereby recovering potential “lost coal” creating higher yield and lower cost/tonne.
Innovative Coal Supply Chain

New Low-cost Fuel

- Coal-water slurry fuel (CWSF) at ~70% solids prepared from coarser (bi-modal) particle size distribution (p80 of 0.075 mm)
- Use for direct firing to boilers as a potential replacement for heavy fuel oil (HFO), or partial replacement for Pulverized Fuel (PF)
- Transport as slurry fuel - avoids sticky, wet or dusting coal problems
- Lower tailings disposal cost - paste-thickening, further dewatering for co-disposal with coarse plant discards and mining waste.

User Benefits

- No further grinding needed, significantly lowering cost
- Major O&M savings and lower ash disposal cost
- Reduced thermal efficiency offset by cost reductions from recovering lost coal from tailings.
- Potential to replace > 30% of the pulverised coal capacity.
- Value Proposition: a 1.0 to 1.5 c/kWh saving once the boiler has been converted for CWSF.

Micronized Coal Water Slurry

Optimization of the fuel cycle (DICE)

- Coal tailings sources
- Including tailings impoundments
- Increased grade recovery
- Recovery of ultra fines
- Minimal dewatering
- Road/rail/ship – cake or slurry
- Pipeline coal water fuels
- Higher solids paste for longer distance
- Mine-mouth or centralized
- Distributed generation
- Support of renewables
Conclusions

• “Deep cleaning” via liberation and subsequent beneficiation has offered significant potential downstream improvements, i.e.,
  – maximised resource recovery,
  – minimised transport and handling costs,
  – numerous end-user process improvements,
  – reduced maintenance and wear,
  – lower environmental impacts and
  – other sustainable improvements.

• Ultrafine coal beneficiation has matured via progressive froth flotation improvements

• Dewatering the concentrate was a barrier, but emergence of membrane filter presses, hyperbaric disc filters or high-g decanter centrifuges now offers commercial solutions.

• Briquetting and agglomeration has progressed to machine capacities of up to ~40 ton/h for fine coal applications to improve product handling.

• Manufacture of stable coal-water slurries with > 65% solids and stable micronized slurries with > 60% solids have reached commercial adoption.

• Scene is now set for new generation clean coal technologies with minimal wasted energy, lowest ash disposal costs and reduced SO₂, NOₓ and CO₂ emission costs.

Acknowledgements

The authors acknowledge the contributions provided by

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• Glencore Coal,
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• CSIRO

towards the preparation of this paper and for permission to include figures and data from other recently published papers and articles on this subject.

Dave Osborne thanks Somerset Coal International for encouragement and support in attending and participating in the 2015 Clearwater Clean Coal conference.
The Jameson Cell - Downcomer

Pressurized slurry enters downcomer through a nozzle at high velocity (typically 15-17 m/s). Typical feed pressure: 130-170 kPa (19-25 psi). Jet plunges into slurry surface causing the entrained air to shear into fine bubbles. High intensity mixing leads to high probability of bubble-particle collision and contact. Residence time in each downcomer is only several seconds. Slurry and the collected particles exit downcomer into tank where particle laden bubbles are separated from the pulp.

Thank you for your time
Questions?
Product Coal Moisture Relationships

- Soil Water Characteristic Curve (SWCC), related to pore size distribution, in turn related to Particle Size Distribution\(^1\).
- Matric Suction is related to Applied Pressure

\(^1\) Source: Prof David Williams; Univ Queensland - D.Williams@uq.edu.au

Coal Water Slurry Fuel

Comparison of Coal Supply Chain Costs for Electricity Generation
Key messages

1. Combining the low cost and availability of coal with the superior thermal efficiency, flexibility and lower capital cost of the diesel engine provides a step change technology for coal
   – this requires ultra low ash coal

2. Ultra low ash (and other premium coals) can be economically produced from a wide range of coal sources, including tailings - using conventional equipment
   … but this requires a change in philosophy
Why? ... provides an alternative LE pathway
Philosophy: higher efficiency + underpinning renewables + niche CCS

MRC-DICE fuel cycle

CO₂ for delivered electricity

Current approach

DICE

renewables

DCFC

renewables

Niche CCS

Major CCS

2010 2015 2020 2025 2030

2010 2015 2020 2025 2030

Carbons

Premium water-based slurry fuel

Ultra efficient diesel engine generation

micronised refined carbons (MRC)
direct injection carbon engine (DICE)
MRC – the most efficient way of converting carbons into liquid fuels

Micronised refined carbon (MRC) has been produced from a range of sources
   – desanded and hydrothermally treated low rank coals
   – deashed black coals (including tailings)
   – chars and algal matter (blended)
Fuel choice determines carbon footprint
Process has very high energy conversion efficiency >97% (LCA basis)

Diesel engine – efficient, flexible and fuel tolerant (but some adaptation required for coal)

Gas turbine
Continuous combustion - hot section is exposed continuously to 1450-1500°C gas at 2-4 MPa
Exotic alloys, hot strength, oxidation
Fouling issues for impure fuels

Turbocharged diesel engine
Higher “Carnot” efficiency
Cyclic hot space allows dirtier fuel without fouling or the need for exotic alloys
Large expansion ratio = smaller waste heat recovery
Batch combustion - hot section hot for <10% of the time, cyclic @ 1-5 Hz (larger engines)
Higher T & P possible and without fouling, >1500°C, 15-25 MPa
Cycle comparisons – too much water?

<table>
<thead>
<tr>
<th>Cycle-fuel P/T$_1$/T$_2$</th>
<th>Fuel (dry t)</th>
<th>Water (t)</th>
<th>Air (t)</th>
<th>$\eta$ (%HHV)</th>
<th>Fuel (dry t)</th>
<th>Water (t)</th>
<th>Air (t)</th>
<th>$\eta$ (%HHV)</th>
<th>$\eta$ sent out (%HHV)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Steam-black 250/650/650</td>
<td>1</td>
<td>0.1</td>
<td>11</td>
<td>88%</td>
<td>12</td>
<td></td>
<td>48%</td>
<td></td>
<td>42%</td>
</tr>
<tr>
<td>Steam-brown 250/650/650</td>
<td>1</td>
<td>2.2</td>
<td>10</td>
<td>73%</td>
<td>11</td>
<td></td>
<td>48%</td>
<td></td>
<td>35%</td>
</tr>
<tr>
<td>Steam-CWM 250/650/650</td>
<td>1</td>
<td>0.7</td>
<td>11</td>
<td>84%</td>
<td>11.5</td>
<td></td>
<td>48%</td>
<td></td>
<td>40%</td>
</tr>
<tr>
<td>Diesel-HFO 200/1500</td>
<td>1</td>
<td>0-1</td>
<td>15</td>
<td>54%</td>
<td></td>
<td></td>
<td>52%</td>
<td></td>
<td></td>
</tr>
<tr>
<td>DICE-MRC 200/1450</td>
<td>1</td>
<td>1</td>
<td>15</td>
<td>51%</td>
<td></td>
<td></td>
<td>49%</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

1 bar/°C/°C

DICE offers game-changing attributes in 5-6 years

1. Match and compete with natural gas; rapid start/stop and load following capability
   - excellent match to electricity grid with high intermittent renewables
2. Step reduction in carbon emissions for electricity generation without CCS
   - 20-35% reduction in carbon emissions versus current black coal
   - 35-50% reduction in carbon emissions versus current brown coal (in Victoria)
3. Cost competitive with new conventional coal
... not possible with other coal technology

4. Small capital investment steps
   – can achieve large power plant size incrementally using 20-100 MW units
   – shorter construction time
5. No cooling water
6. Can be used for various biomasses
7. Capture ready/capture efficient
   – 30-40% lower cost of CO₂ abatement over conventional coal
8. Short path to commercialisation
   – adaptation of current large engines, short cycle time to implement changes, relatively low development cost

What about USC? ... development opportunities restricted and costly

• Thermodynamic efficiency of pf generation is severely limited by the availability of materials that can operate at these conditions for practical service lifetimes
  – EU, USA, Japan, India and China all have extensive material research programs aiming for steam temperatures of 700°C (advanced ultra-supercritical)
  – development cost of billions of $ and long lead times (creep testing)
  – anticipated that a commercial unit could be brought on-line in 2031 (IEACCC/229)
  – high capital cost of advanced ultra-supercritical is of particular concern (high pressure steam pipes currently 80% of the boiler cost)
• While the combustion conditions in the diesel engine are more extreme, the diesel cycle is a batch process
  – high temperature conditions are present for less than 10% of the time, which avoids the need for major exotic alloys
A pathway to net negative CO$_2$ emissions?
(DICE efficiency first, then high penetration renewables with bio-CCS and lastly partial CCS)

Example: Victorian generation

Carbon management sequence

** landscape & soil carbon sequestration credits

If successful DICE could address many aspects of the coal dilemma

Including:

• that large centralised plants are needed for efficiency
• the nexus between water/CO$_2$ and cost (dry cooling)
• technology development by a fragment of generation industry
• inefficient (even if cheap) is no longer acceptable
• very poor image of low rank coals
• poor project economics from long development times
• the higher flexibility needed for current and future electricity markets

Could DICE become the benchmark coal generation technology?
Not new … commercialisation in G3?

By adapting existing technology (engineering)

Generated 3
1-5 MW/cyl
minimal adaption
slurry firing
low cost MRC

Generation 4
1-20 MW/cyl
optimised/intelligent
paste/slurry firing
biofuel co-firing
CO₂ capture

… with a changed philosophy

Coal to replace oil

Clean Coal Technology

By adapting existing technology (engineering)
Coal water fuel in China – stepping stone to MRC?

~40 Mtpa in China
5-8% ash, d90 ~150µm
typically 70% coal
2000 mPa.s @ 100/s

nominal specifications
1-2% ash, d90 ~40µm
typically 55% coal
<300 mPa.s @ >200,000/s

Micronise
and float

MRC for DICE

JGC
MRC is based on ultra-fine coal beneficiation

- Physical cleaning – coal structure not changed, but comminution is needed for optimal liberation/improved flotation response
  - potential product <1% ash, including from tailings
- “Old school” thinking typically regards ash contents below 2-3%, as both technically and economically unviable because
  - “inherent ash” of coal is usually regarded as the lowest achievable ash content
  - lower ash requires milling to impractical ultra-fine sizes for liberation
  - … but flotation of ultra-fine coal is problematic requiring higher reagent dosages
  - fine coal concentrates are inevitably high in moisture (> 35%) which means costly dewatering and/or drying to produce saleable products
- All of these factors are incorrect, or at best very misleading, as recent research/pilot plant tests have shown

Excellent liberation by micronising

Feed

After liberation by milling

Characterised images for raw coal tailings feed compared with final concentrate (QCAT-CSIRO)
Micronising and sub-50µm coal flotation – excellent separation and recovery

Pilot scale production of MRC for MAN Diesel & Turbo

1. MRC cake production at Bulga Pilot Plant (Glencore)
2. Formulation & rheology trim (CSIRO)
3. Certification (ALS)
MRC ... strategic part of a bigger picture of “Premium Coal Products”

Context

<table>
<thead>
<tr>
<th>Coal products</th>
<th>Conventional</th>
<th>Steaming</th>
<th>Coking</th>
</tr>
</thead>
<tbody>
<tr>
<td>“Premium” (&lt;3% ash)</td>
<td>Coal water fuel</td>
<td>Electrode carbon</td>
<td>Micronised refined carbons for DICE (MRC)</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Electricity generation</th>
<th>Conventional</th>
<th>Pf</th>
</tr>
</thead>
<tbody>
<tr>
<td>Alternative</td>
<td>IGCC</td>
<td></td>
</tr>
</tbody>
</table>

Emerging

Direct injection carbon engine (DICE)

Direct carbon fuel cell (DCFC)
... and an alternative Coal Supply Chain

- Current Coal Supply Chain hampered by an inability to dewater and efficiently transport fine coal
- Innovative approach - recover and use all the ultra-fines as coal water slurry thereby recovering potential “lost coal” creating higher yield and lower cost/tonne

The engine side ... towards commercial DICE
Recent developments

CSIRO DICE program since 2008
   - **de-risking based R&D** program (Yancoal, Exergen, Newcrest/JGC, BCIA, Ignite Energy Resources, Xstrata)

**MAN Diesel & Turbo** have taken a lead position in DICE development

Umbrella organisation established to facilitate DICE development internationally
   - 17 participants includes MAN, RWE (Germany), JGC (Japan), Sinarmas (Indonesia), Exergen, Ignite Energy Resources, BCIA, Energy Aust, AGL, Newcrest, Yancoal, Worley Parsons, GHD, ACALET and CSIRO

Recent interest from groups in Korea and China

---

Stage-gated development

**2014-16**  Small scale demonstration, initial demonstration/validation DICE, 1MW single cylinder (brown and black coals)

**2016-17***  Development/design of components for prototype engine

**2017-19***  Full scale demonstration MRC production with a 12-30 MW prototype engine for 8000h campaign

**2020***  First commercial DICE power plant [$1.4-2 M/MW] possible given appropriate funding support

* based on MAN D&T estimates of 3-5 years for engine dev
DICE deployment strategy

DICE favoured when natural gas price >$6-7/GJ forecast to occur by 2020 for most countries
- Australia; $5-6/GJ; forecast >$8.50/GJ in 2020
- China (import); $13.70/GJ; forecast >$10/GJ in 2020
- Europe; $10.80-12.20/GJ; forecast >$8.80/GJ in 2020
- UK; $10-14.60/GJ; forecast >$10/GJ in 2020
- Japan/Korea; $14.20-16/GJ; forecast >$13/GJ in 2020
- Limited incentives in USA (low cost gas,1100 lb CO₂/MWh regulations)

DICE suitable for new coal capacity, and to replace old capacity nearing the end of its economic life (or as it becomes socially unacceptable)

Final comments

1. DICE could provide coals with a innovative step technology to increase its cost competitiveness and environmental acceptance
2. Barriers to commercialisation are mostly engineering
   - adaptation of commercial process & engine technologies
3. Rapid development possible - can be demonstrated at commercial scale at a relatively small cost
   - short lead time between technology development & implement
4. Logistical barriers to commercialisation of the fuel cycle needs broad intra-industry support
   - as part of premium coal products for maximum benefit
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Louis Wibberley
Leader DICE Development Program

t  +61 2 4960 6050
e  louis.wibberley@csiro.au
w  www.csiro.au/lorem
The Arrium IsaMill from Design through Commissioning and Optimisation

M Larson¹, G Anderson², M Mativenga³ and C Stanton⁴

1. Senior Metallurgical Engineer, Glencore Technology, 160 Ann Street Level 10, Brisbane QLD michael.larson@glencore.com.au
2. IsaMill Technology Manager, Glencore Technology, 160 Ann Street Level 10, Brisbane QLD, greg.anderson@glencore.com.au
3. Principle Development Manager, Arrium Mining, PO Box 21, Whyalla, SA. Email: martin.mativenga@arium.com
4. Process Engineer, Arrium Mining, PO Box 21, Whyalla, SA. Email: chris.stanton@arium.com

ABSTRACT

Arrium’s magnetite concentrator has recently completed a modification program to improve the process that has been in operation in Whyalla, South Australia for the previous seven years. The magnetite concentrate is pumped via pipeline to the coast and serves as pellet plant feed, eventually being turned into steel in the Whyalla blast furnace. With the addition of a new 3MW IsaMill at the concentrator, the plant has seen increased production from a reduced feed grade, enabling Arrium to treat ores previously stockpiled as waste. This paper will focus on the design work and commissioning of the fine grinding circuit along with the year of optimisation that followed, taking the IsaMill up to 350+ tph of throughput.

INTRODUCTION

The original Arrium magnetite flow sheet that began operation in 2007 consisted of two stages of grinding, utilizing two HPGRs with a total installed power of 1.8 MW and a 7.5 MW ball mill. The complete flow sheet is shown in Figure 1.

![Figure 1 – Original Arrium magnetite flow sheet.](image)

In 2009 Arrium began a review of the plant with the aim of debottlenecking and optimisation. Much of this early work is covered in the AUSIMM Iron Ore 2015 paper *Unlocking Plant Capability through Targeted Debottlenecking of Arrium’s Magnetite Concentrator* (Mativenga, et al 2015). The main outcome of this work was the conclusion that if the plant was upgraded to improve liberation a large portion of stockpiled waste could be converted to plant feed while also increasing mill
throughput. Arrium then began investigating different options to add grinding power to the plant through both conventional and stirred milling options.

**BALL MILL VS ISAMILL COMPARISON**

In 2011 Arrium Mining ran their first IsaMill lab test at ALS AMMTEC in Perth WA. This was a straight comparison of the existing ball mill circuit to the IsaMill. With an F80 of +350 microns it is coarser than most operating IsaMills. With 5-6 mm media the IsaMill was able to break the coarsest feed and still efficiently reach the final product size in significantly less energy than the ball mill. The ball mill reached a P80 of 32 microns in 24 kWh/t, not including any cyclone pump energy. The M4 IsaMill with the same feed from the ball mill survey reached a 32 micron P80 in 17 kWh/t as shown in Figure 2 (Steele, 2011). This results in a 29% improvement over the ball energy to 32 microns.

![Figure 2 – Arrium ball mill versus M4 IsaMill.](image)

The basic principle of the IsaMill is that it fluidizes a charge of small ceramic grinding media. Media ranging in size from 1.5 to 6.5 mm and coarse particles from the mill feed are centrifuged to the outside of each disc where backflow compression from the product separator keeps them in the mill until the particles are ground fine enough by attrition and compressive forces to pass down the middle of the mill and discharge. Due to this process external classification is not necessary.

As long as the IsaMill can break down the coarsest particles as quickly as they enter the mill it can compete with a ball mill. The M4 IsaMill testwork showed this was possible when grinding a 300 micron Arrium ore feed to P80 values below 100 microns.

However, the Arrium flowsheet has a functional ball mill and the goal of this exercise was not to replace it, but to modify the flowsheet to increase plant capacity and improve grinding energy efficiency.

Through previous Australian magnetite testwork it was known that running the IsaMill with the same feed (i.e. in parallel) as the ball mill is not the most efficient flow sheet. From pilot and lab work presented at MetPlant 2011 in the paper *Optimising Western Australian Magnetite Circuit Design* it is shown that 3 stages of grinding efficiently utilizes each grinding mill to the best of its abilities (David *et al.*, 2011). As shown in Figure 3 the ball mill is perfectly suited for a coarser grind which then reduces the feed size to the IsaMill. From the graph it can be seen the IsaMill is only more efficient at grinding to product sizes below 100 microns in this particular case. With smaller reduction ratios both mills can be optimised with the correct media sizes. As an added benefit the separation step between the two milling stages reduces the amount of gangue material needing to be ground down to 32 microns.
Large reduction ratios and lack of intermediate gangue rejection work against the two stage flow sheet shown in Figure 3. The ball mill, though not ideal for grinding fine, is well suited for an intermediate grind, which can then be followed by an extra step of magnetic separation. This produces a reduced mass of feed which is ideally sized for the IsaMill. This idea was incorporated into the next phase of the Arrium M4 IsaMill work at ALS AMMTEC completed in 2012.

**ISAMILL VARIABILITY SIGNATURE PLOTS**

Test work was next undertaken at ALS AMMTEC with Arrium personnel present to understand how different sections of the ore body grind in the M4 IsaMill. Figure 4 shows three of these tests which display a considerable variation in the energy required to reach 32 microns, with values of 14.2, 12.3 and 18 kWh/t respectively (Ladhams, 2012). These three samples were each taken from different areas of the ore body. It should be noted these three samples were a harder, higher waste material than the first material tested in Figure 2. The test feed was ground down to an F_{80} of 90 microns with no intermediate magnetic separation prior to being run through the IsaMill. Removing liberated gangue material would have the effect of reducing this variability somewhat. It does show the importance of understanding the ore body and not relying on just one sample to design a regrind circuit.
After numerous signature plots and some pilot testing it was determined that one 3 MW M10000 IsaMill would be suitable to grind a maximum design 300 tph of 60 micron feed to a desired 32 µm P80 and Davis Tube silica grade of 2-3%. The mill would be in a tertiary grinding step rather than run in parallel with the ball mill. The design concept is to allow the HPGRs to produce a coarser product at higher tonnage, optimise the ball mill to treat this coarse material and then provide a suitably fine feed for IsaMilling. The IsaMill product would be fed to an installation of Derrick Stack Sizer screens with 63 micron panels to ensure steady silica grade. The role of the Derrick Stack Sizers is primarily to eliminate any remaining oversize material both for silica grade control and pipeline protection. This flowsheet is shown in Figure 5. There are additional 1 mm screens ahead of the IsaMill serving to remove trash from the feed rather than providing actual classification.

**FINAL MILL FLOW SHEET**

Figure 4 – Arrium variability signature plots.

Figure 5 – Arrium modified flow sheet.
ISAMILL START UP

The M10000 IsaMill at Arrium is shown in Figure 6. Feed enters the mill at the non-drive end and discharges from the drive end already classified. Grinding is achieved by discs on an internal rotating shaft, and the discs are designed to keep a ceramic grinding media load of about 70% of the mill volume agitated.

Figure 6 – Arrium M10000 IsaMill installation.

The IsaMill was commissioned in November of 2013. Performance Acceptance Testing (PAT) was conducted on a dedicated ore stockpile shortly after start up. Comparison test work had been run on an M4 IsaMill at ALS AMMTEC from samples of this same stockpile. Minor differences from these feed samples included the media size used and lower solids SG due to there being no DIMs upgrading applied in the case of the M4 feed. The M4 sample was also slightly coarser, resulting in more residence time and a sharper size distribution once it reaches the final P<sub>80</sub>. A comparison of the average results from both the lab and full scale is given in Table 1 (Bandarian, 2014).

Table 1 – Average PAT plant sizing data versus average AMMTEC signature plot test work data.

<table>
<thead>
<tr>
<th></th>
<th>Specific energy input (kWh/t)</th>
<th>Feed F80 (µm)</th>
<th>Discharge % passing 32 microns</th>
</tr>
</thead>
<tbody>
<tr>
<td>Average Plant Data (PAT)</td>
<td>8.0</td>
<td>41.5</td>
<td>86.1%</td>
</tr>
<tr>
<td>Average AMMTEC Data</td>
<td>8.6</td>
<td>41.9</td>
<td>85.6%</td>
</tr>
</tbody>
</table>
Considering the differences in operating conditions and feed, the agreement between the two data sets is acceptable. For all of the M10000 product size distributions measured during the PAT period the average percent passing 32 microns was 86.1%. The average percent passing 63 microns was 98.1%. With an extrapolated P_80 of 27 or 28 microns that gives a P_80/P_80 ratio of about 2.33. The ratio for the three AMMTEC PAT comparative tests averaged around 2.3. The percent passing 75 microns averaged 99.1% in full scale surveys. ALS AMMTEC test C at pass 3 had a P_80 of 26 microns and 99.5% passing 75 microns. Test B was a P_80 of 28 microns and 99.6% passing 75 microns. The slight discrepancy between full scale and the M4 is likely due to the larger media and longer residence time in the M4. Consequently, the M4 produced a sharper size distribution once it reduced the feed to the ~50 micron P_80 size that was feeding the M10000. It should be noted that test A is not included in these comparisons simply because none of the passes were at sizes directly comparable to individual survey results.

Average comparisons of the full feed and product size distributions between the M4 and M10000 for this test are shown in Figure 7 (Bandarian, 2014).

![Particle Size Distribution (Plant PAT Test vs. ALS AMMTEC)](image)

**Figure 7** – M4 and M10000 complete size distribution comparisons.

**MILL CONFIGURATION OPTIMIZATION**

As is standard commissioning and operating practice, temperature checks are regularly taken along the IsaMill shell. These are tracked to monitor for compression and build-up of heat in the mill. Through the first two months of commissioning it was clear that there was more compression at the shell than is recommended for optimum mill liner life. Different mill configurations were implemented to improve the fluidization inside the mill and reduce shell liner and disc wear. Through small incremental changes the ideal mill setup was found. This spread out the temperature profile in the mill and also acoustically gave a more fluid sound inside the mill. Full mill operation with minimal internal wear was realized in early February 2014, about 3 months into commissioning and well ahead of the June deadline to reach full run time and throughput.

The main indicator of this success was the condition of the mill when opened for inspection, shown in Figure 8 (Ziki, 2014). After the optimisation phase wear of the discs and shell liner was found to be
more consistent across all surfaces leading to an increased maintenance interval with better predictability of spares use.

![Figure 8 – IsaMill maintenance opening February 2014.](image)

Currently the IsaMill is inspected every 6 weeks in the line with the rest of the plant. Having the IsaMill maintenance line up with the rest of the plant was critical to ensuring a minimization of disturbances to the pellet plant feed, both in terms of grade and throughput.

**ISAMILL SCALE-UP AND OPTIMISATION**

Once it was confirmed that the mill was achieving the design grind duty and required maintenance interval, work began to fully optimize the operation. An M4 IsaMill was once again used, this time at the University of Queensland. A short survey of the IsaMill was performed to collect sufficient feed sample for multiple M4 tests. A graded charge of media from the M10000 IsaMill was shipped to Brisbane to perform test work. Product was sampled to determine the discharge sizing by sieve screens. The average results of the full scale survey are shown in Table 2.

Table 2 – IsaMill feed and product sizes, February 2014 survey for M4 scaleup (Villadolid, 2014).

<table>
<thead>
<tr>
<th>Size µm</th>
<th>% Passing Feed</th>
<th>% Passing Discharge</th>
</tr>
</thead>
<tbody>
<tr>
<td>106</td>
<td>98.3</td>
<td>99.9</td>
</tr>
<tr>
<td>75</td>
<td>94.3</td>
<td>99.5</td>
</tr>
<tr>
<td>53</td>
<td>87.3</td>
<td>97.6</td>
</tr>
<tr>
<td>38</td>
<td>61.3</td>
<td>84.9</td>
</tr>
<tr>
<td>25</td>
<td>52.7</td>
<td>74.6</td>
</tr>
</tbody>
</table>

The media size of 4-5 mm was initially decided upon by IsaMill rules of thumb as it was thought to give the best product size distribution for the 32 micron $P_{80}$ of product and ~50-60 micron $F_{50}$ design.
feed at a higher than average expected feed rate. At the University of Queensland a GT engineer tested this media against a laboratory 3.5 mm charge and a 5-6 mm charge that had been used at Ernest Henry Mine in their much coarser M10000 magnetite tails operation (Villadolid, 2014).

The M4 at UQ accurately predicted the required Arrium M10000 energy with 4-5 mm media, although the signature plot line shown in Figure 9 needed to be extrapolated due to the fine first pass produced in the M4 (Villadolid, 2014).

![Figure 9 – IsaMill M4-M10000 scale-up.](image)

When comparing the different media size options it would appear that the current 4-5 mm is correct for the full scale application. To a P_{80} of 32 microns the 4-5 mm media required 7.2 kWh/t, the 5-6 mm 11.2 kWh/t and the 3.5 mm 7.6 kWh/t. Due to the superior product size distribution inherently produced by the larger 4-5 mm media compared to any smaller media sizes, and its ability to cope with potentially coarser feed, it was recommended that Arrium continue to use the current size of ceramic media.

Satisfied that the media size was appropriate, a program to push the IsaMill well past its design maximums was implemented in April/May of 2014 just over two months after commissioning was officially ended. Mill feed density was slowly increased while monitoring the mill for proper fluidization and compression. It was found the mill slurry density could be increased from 1.64 to 1.68 kg/l in this period, increasing the solids feed rate by about 20 t/h at a constant flow rate. Next the feed flow rate was increased while monitoring the mill media load to be sure rotor pumping capacity was not exceeded. The flow was gradually increased until it reached 400 m^3/hr. The end result was an increase in tonnage from about 300 tph to 350 tph. Through this period the P_{80} increased slightly along with the feed size but the product top size and Davis Tube silica were barely changed, increasing by less than 0.5% silica.

One realization during this trial was that the Derrick Screen oversize typically contained less than 4 t/h of material that was actually coarser than 63 microns (the oversize mass was much more than 4 t/h but the majority of it was misreporting fines). In essence, the Derrick screens were underutilised because the IsaMill does not leave much +63 µm in the product. It was possible to use the excess capacity in the Derrick Screens and final grade to process additional material equivalent to IsaMill feed, while still processing the full flow if IsaMill product. The Derrick screens control the top size reporting to the cleaner magnetic separators and all oversize is returned to the ball mill cyclone feed rather than the IsaMill feed. In this way total tonnage for the plant could be increased further without exceeding the silica grade limit and without coarsening the product P_{80} significantly. Typically the liberation from the IsaMill even when flow constrained was above what was necessary. With a bypass line installed to the Derrick screen the plant could increase total new feed to the HPGR and
ball mill. The coarsest fraction containing the most silica that bypassed the IsaMill would be removed by the Derrick screens, providing the necessary control over silica grade.

**CURRENT OPERATING CONDITIONS**

From May of 2014 until March 2015 the IsaMill was treating an average of 350 tph while the design maximum tonnage was 300 tph. Mill power draw has run consistently at 2,600 kW. Design disc consumption was predicted to be 26 per year. It is currently under 14 discs per year and continues to drop as the IsaMill Information System (IMIS) program provided by GT receives more information to better predict and utilize the full life of a disc. Though accelerated wear was experienced at the start of operations, current indications predict a liner life of 8-12 months, which is about on budget. Arrium has taken on a methodical approach to moving the shell liner every second mill opening to spread wear to all available rubber areas. Availability for the IsaMill has been excellent, with almost all downtime related to issues outside of the IsaMill circuit. Mill stability has been steady due to consistent upstream operation. The yearlong average for the Cenotec grinding media wear has been 6 g/kWh, versus a budget of 10 g/kWh. The media has worn smooth and round as shown in Figure 10, also contributing to improved disc and shell liner wear.

![Figure 10 – Arrium IsaMill graded media charge.](image)

**Future work**

Beginning in March of 2015 the IsaMill internal configuration was changed again to allow more flow through the mill. This brought the IsaMill to 485 m$^3$/hr of flow and about 425-435 t/h of solids. As of the writing of this paper analysis continues into the sustainability of this production increase. Publications on progress will be made available in the future.

**CONCLUSIONS**

This project has seen the magnetite operation undergo an overhaul, allowing it to produce magnetite concentrate using a lower grade feed material at an increase in production capacity of greater than 400,000 tonnes per year. (Arrium news webpage, 2014) This throughput has continued to be expanded as the maximum capacity of the IsaMill has been pushed well beyond the original design. After a challenging start, handover of the plant was completed on budget and ahead of schedule. Through methodical implementation procedures, step by step improvements in throughput and mill
wear have been realized allowing the plant to process tonnage well beyond anything for the original design.

**ACKNOWLEDGEMENTS**

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**REFERENCES**


Unlocking Plant Capability through Targeted Debottlenecking of Arrium’s Magnetite Concentrator

C. Stanton¹, M. Mativenga², J Gerber³, C Jenkins⁴

1. Process Engineer, Arrium Mining, PO Box 21, Whyalla, SA. Email: chris.stanton@arrium.com

2. Process Development Manager, Arrium Mining, PO Box 21, Whyalla, SA. Email: martin.mativenga@arrium.com

3. Principle Process Engineer, Arrium Mining, PO Box 21, Whyalla, SA. Email: johan.gerber@arrium.com

4. Senior Process Engineer, Arrium Mining, PO Box 21, Whyalla, SA. Email: Clinton.jenkins@arrium.com

ABSTRACT
In 2014, Arrium completed an optimisation program on its magnetite concentrator, enabling a reduction in the overall cost base of producing concentrate through an increased ability to treat a wider range of ore types, previously considered waste. The development was the culmination of four years of work, conducted in house by Arrium’s mining development group. The overall intent of the plant design focused on improving the concentrator’s ability to treat a wider range of magnetite ore types. Metallurgical analysis of these feed ore groups outlined a need for more grinding power if production quality was to be maintained and throughput increased. As such, a tertiary grinding circuit was designed which, alongside the rectification of key operational bottlenecks, would allow the plant to continue to produce concentrate at the required specification and rate. The technology chosen for the tertiary grinding application was the XT IsaMill™, which is known for efficient fine grinding when compared to conventional ball milling. The transfer sizes between the existing comminution circuits were coarsened, allowed for the treatment of different ores at the required rates. The optimisation project was successfully completed in February 2014 after a three month commissioning phase. The initial metallurgical testing and engineering of new equipment proved highly successful with all equipment meeting or exceeding design parameters.

INTRODUCTION
Arrium operates a magnetite concentrator at its Iron Duke mine, located 67km south west of the township of Whyalla, South Australia. The concentrator was commissioned in 2007 as part of Project Magnet, a successful initiative to convert the Whyalla pellet plant from hematite feed to magnetite. Operation of the concentrator in subsequent years highlighted...
several areas of improvement in the original plant design. A project was commenced in 2009 to investigate, design, and construct solutions to optimize the plant. In addition, stockpiles of high silica material previously disregarded presented a business opportunity if made treatable by the upgraded flowsheet. The Magnetite Optimization Project (MOP) represented a body of work consisting of four years’ work in plant design, working with an array of highly experienced subject matter experts, to develop, test and construct an upgraded flowsheet for the processing of Arrium’s magnetite reserve. This paper will discuss the identification of problem areas, development of key process solutions and finally, the implementation measures that occurred.

**ORIGINAL FLOWSHEET**

The original Project Magnet flowsheet consisted of two stages of grinding with low intensity magnetic separation following each comminution circuit. Crushed ore (-32mm) was fed to two parallel 1.8MW High Pressure Grinding Rollers (HPGR) in closed circuit with 2.2mm banana deck screens. Undersize from the screens was slurried and pumped to the Rougher Magnetic Separators (RMS). Rougher concentrate was then screened at 0.5mm, with all oversize going back to the HPGR, in an attempt to liberate and reject as much non-magnetic material as possible. Minus 0.5mm was ground down to an 80% passing size ($P_{80}$) of 38µm within the 7.5MW ball mill. The ball mill was closed out with hydrocyclones, with the overflow gravity discharging to a bank of De-slime Intermediate Magnetic Separators (DIMS). The DIMS concentrate was fed to the Cleaner Magnetic Separator (CMS) before passing through a protection screen prior to the slurry storage tanks.

**INITIAL CIRCUIT ANALYSIS**

In 2009, Arrium began an optimization program on the concentrator plant to improve production rate and reduce cost. An initial review was conducted on plant performance in conjunction with Process Technology and Innovation (Baguley, Jankovic and Valery, 2010). It comprised of metallurgical surveys, assay mass balancing, and historical data analysis. This review provided the following main conclusions:

- The HPGR circuit was power limited at harder ore types; restricting fresh feed due to the recycling load rising beyond conveyor limitations.
- At 0.5mm, the HPGR close out screen size was causing excessive front end recycling load, limiting throughput.
- The ball mill was shown to be overgrinding the heavier magnetics, reducing efficiency.

The existing cyclone feed pumps were operating at the limit of their duty due to the preferred cyclone configuration at the time.

**CHANGES TO THE FEED: BUSINESS OPPORTUNITY**

During the initial years of processing magnetite, a significant amount of high silica magnetite was stockpiled. This material was not treated through the concentrator due to final product quality concerns that existed at the time.

The Geometallurgy group developed a modified Davis Tube Recovery (DTR) process, with the results able to be translated into expected plant performance utilising a regression model. Laboratory test work conducted on these high silica magnetite stockpiles in 2011-2012 indicated that the material achieved good metallurgical performance when ground to the target 80% passing size of 38µm ($P_{80}$) of the concentrate. The test work also indicated
that due to the increased quartz content, the material was harder than the traditionally
dominant carbonate magnetite usually treated by the plant. This material also had a lower
mass recovery when compared to the average magnetite feed blend. These two factors
together would require an increased capacity in the concentrator grinding circuit to
maintain the required final product production rates.

Full scale plant trials were conducted, progressively increasing the ratio of high silica feed
to standard feed, ranging from 10% through to 30%. Initial results on the trial date blend
showed encouraging results in the concentrate quality that was produced. During
subsequent plant trials feeding up to 100% high silica material, it was shown that while the
concentrator could achieve the target grind and an acceptable concentrate quality,
production was significantly reduced due to lower mass recovery. It was also found that
the final product quality was sensitive towards the grind achieved, with any variation from
the target grind causing the product quality to fall outside the required levels.

This added impetus to the notion of conducting a major debottlenecking exercise as the
most applicable means of increasing production.

EARLY BOTTLENECK IDENTIFICATION
Prior to commencing a design study with an engineering company, Arrium conducted an
internal review of the plant bottlenecks identified previously by PTI and other subject
matter experts. In order to increase the feed rate beyond the design these key bottlenecks
would need to be identified first to save time later on.

**HPGR Circuit**
At the time of review, the feed conveyors around the HPGR units were running at near
maximum capacity. Each HPGR was tasked with reducing 1000tph of -32mm feed
material down to below 0.5mm, taking into account the fine recycle from the RMS screens.
The plant operators could not physically increase the circuit raw feed rate without
exceeding the capacity of the HPGR feed conveyor belts. Prior optimisation work to
increase the recycling load was recommended by PTI as a means of increasing the
utilisation of installed power (Baguley, Jankovic and Valery, 2010), which should also
increase the proportion of fine material being generated by the circuit (Dundar et al, 2009).
However, a site review by the HPGR vendor, Koeppern, noted that the amount of fines in
the HPGR feed was already well above the original design specification (Garduła, 2010). It
was also noted that there was up to 40% -2mm in the HPGR feed, as opposed to the 15%
catered for in the original design calculations. The inability to process more tonnage
because of the belt limitations and recycling load was a major conclusion to this stage of
the review.

**Ball Mill Circuit**
From metallurgical surveys and reviews, it was established that the recycling load around
the ball mill was industry average at 250%, but that the cyclone feed pumps were
operating at their limits. This data led to the conclusion that the mill could be optimized for
higher throughput or coarser grind size, and that attention should be instead paid to
cyclone pump capacity.

An in house review of the cyclone pump capacities confirmed that both cyclone feed
pumps were not adequate for the volumetric duty required to achieve the 38µm grind. The
actual static head on the system at the time was 17 metres, as opposed to the original
design of 13.8m (Phillips, Westbrook, 2010). This was further compounded by a back
pressure of 166kPa, as opposed to the original design of 100kPa. Because of these
disparities, the overall static head increase was 9.2m over the design, and the pump could
not be expected to reach its design duty rate (Phillips, Westbrook, 2010). From the
review, the calculated sustainable volumetric flow rate was 712m$^3$/hr, far less than the
original mass balance rate of 1114.5m$^3$/hr.

For the plant operators, this translated simply into a lack of capacity to control product
quality if the feed ore was harder than normal, or an inability to increase feed rate when
the opposite was true.

OPTIMISATION APPLIED TO INITIAL SCOPING DESIGN

While optimization of the existing concentrator concluded with PTI at the end of 2011, work
had already began within Arrium creating multiple base flowsheets upon which a feasibility
study could be developed. These flowsheets would target the bottlenecks identified and
include, where possible, proposed remedies based on the best available information at the
time. Major changes put forward for engineering design were:

Removal of the RMS screens
This would coarsen the feed 80% passing size ($F_{80}$) to the ball mill from approximately
300µm to 1400µm. Fresh feed tonnes into the HPGR circuit would also be increased to
make up the recycling load.

To check the ball mill capacity, a simulation of the ball mill performance at the optimised
conditions was carried out, and only through significant variance in the design parameters
did the model predicted power requirement exceed the installed capacity. An internal
report outlined that with the original design 80% particle passing size ($F_{80}$) of 416µm and
product 80% passing size of ($P_{80}$) of 38µm, the calculated installed grinding power ranges
from 5800kW to 7250kW. In essence, there was more than enough installed power to
achieve the original grinding requirements, yet as the resultant mill grinding efficiency was
lower than anticipated the mill represented a bottleneck on the plant. A paper written by
Partyka outlined the limitations of ball milling when fine grinding, noting that below 20 to
30µm, they become inefficient (Partyka, Yan, 2007). Of particular interest, the paper also
showed a worldwide trend against using ball mills in a regrind application with similar
parameters to Arrium’s operation (Table 1).

This hypothesis was later re-iterated with ongoing design work with Amec. The ongoing
feasibility study also pointed out the inefficiency within the mill could be attributed to its
very fine grinding duty and wide reduction ratio (Lilford, Nofal, 2012). As such, the tonnage
and grind reduction ratio required at Arrium’s concentrator was not conducive for any
increase in throughput, and the design focused on coarsening the mill and increasing
grinding efficiency.

Installation of a new grinding mill
To cater for the throughput increase and coarsened product from the ball mill, an
additional grinding circuit was required. For the size range being considered (-100µm),
stirred milling technology was deemed appropriate for consideration. The two possible mill
configurations considered suitable for the duty were the Metso VertiMill and XT IsaMill™. The early steps involved in the data collection for the decision making were:

1. Visit to Arrium by Metso and Xstrata engineers,
2. Bench-scale testwork on each technology – samples were sent to independent laboratories for grinding testwork with both technologies. This included multiple tests with the XT M4 lab scale IsaMill,
3. Communication and visits to reference sites – a number of reference sites were visited, including Ernest Henry Mine in Queensland and McArthur River Mine in the Northern Territory.
4. Advice sought from independent subject matter experts,
5. Literature review of laboratory mill technology, focusing on the grinding performance and successful scale-up.
6. Construction lead time of equipment, and what auxiliary capital is required (for example, cyclones).

Criteria from the data collection, such as capital and operational expenditure, lead times, and risk were weighed in a selection workshop. From this, the XT IsaMill™ was chosen as the preferred technology to move into feasibility design. For the duty, a single 3MW M10,000 IsaMill™ was required. By the start of 2011, Arrium commenced a feasibility study into the new plant design, with Amec selected as engineering partner.

FEASIBILITY – CONFIRMING ASSUMPTIONS

With the scoping study concluded, the feasibility phase of design aimed to provide a revised flowsheet and capital estimate to +/-25% accuracy.

Flowsheet Simulation Model

In order to mitigate the risk of the flowsheet changes and new equipment, a JKSimMet model was developed simulating the proposed changes on the plant. In order to fulfil this task an extensive metallurgical survey of the concentrator was conducted, including obtaining a bulk sample of the RMS feed. The bulk sample would be processed by the selected lab on bench scale equipment as per the proposed flowsheet steps, including the IsaMill stage. Bond work indices were also conducted on the nominated size range to verify the mill power required. As a final measure for risk mitigation, three separate plant surveys were conducted on different dates to ensure some variance in feed mineralogy. Three bulk samples were in turn submitted for analysis to the laboratory to run through the testwork program independently.

To ensure adequate model comparison against the actual industrial sized concentrator, the metallurgical surveys sampled every major unit of operations within the plant. All samples generated from the lab scale and full scale surveys were subject to identical analytical tests to ensure comparative data. As the M4 IsaMill was situated at ALS Perth, this laboratory was selected to conduct the majority of the analytical work, with the aid of vendors for specific equipment tests.

Results Analysis

The flowsheet simulation program successfully verified many of the assumptions carried through the feasibility study. The model and the final flowsheet design were in turn updated ensuring the rigour of third party review was continued. It was concluded from the testwork program that the results substantiated the flow sheet selection, successfully
mitigating a significant portion of the process risks. In turn, an independent third party review conducted by PTI (Baguley, Jankovic, 2012) outlined no major issues or problems with the flowsheet.

A high level comparison of the grinding transfer point changes between the original plant and that at the end of feasibility is shown below in figure 1.

**FINAL DESIGN AND CONSTRUCTION**

Leighton Contractors (LCPL) and GR Engineering Services (GRES) were selected as final design and construction engineering partners. A final review of flowsheet and testwork results were conducted by GRES, with a few key changes made. These included modifications to the final screening plant, including reducing the complexity from two stages to one, and installing Derrick StackSizer™ screens for greater efficiency. Another notable change to the design was the recycle of the oversize from the screening plant back into the process, as opposed to sending it to tail. Construction commenced in 2013, with LCPL working in conjunction with Arrium’s Major Capital Development Group. All brownfield construction was completed within the monthly 24 hour production shutdowns leading into November, upon which a 10 day production halt enabled the completion of all tie-ins and pre-commissioning activities.

**THE FINAL FLOWSHEET**

The magnetite optimization project targeted multiple areas of the existing plant to install additional capacity, and while this paper focused primarily on the grinding circuits, there were numerous other upgrades that were included and not discussed. Figure 2 is a depiction of the improved flowsheet as constructed and commissioned in 2013.

**COMMISSIONING**

Transitioning the project from design to construction, commissioning, and finally to plant handover was managed through the use of risk management and coordination tools. During the risk management process, all risks associated with the project were classified, ranked and prioritized with timed action plans and performance monitoring processes put in place (figure 3).

The transition and coordination steps involved bringing all stakeholders together to review construction progress and map out when the different phases of commissioning would commence. The overall goal was to ensure the plant was handed over to the operational owner as soon as possible.

A Management Operating System (MOS) was put in place to ensure all stakeholders were able to attend and give input into the required progress review meetings. The project progressed into process commissioning by mid November 2013, with performance acceptance tests successfully concluded in mid-December, 2013. By the beginning of January, the plant was formally handed over to operations and the ramp up continued.

During the first two months from commissioning, process optimization was undertaken to bring the plant to design production rates. The work included optimisation of the IsaMill™ wear components to maximise throughput and grind. Plant surveys and sampling campaigns were conducted to validate the mass balance and design assumptions. Classification (ball mill cyclones and screening) optimisation was also performed to
achieve the required concentrate grade. The optimisation work undertaken resulted in the process operating at above 10% of the design throughput prior to concluding the commissioning phase.

The performance monitoring system put in place to transition from construction to project handover was done successfully, with seamless transition between the different phases of the project delivery model. This success also led into an above target ramp up process after commissioning, shown in figure 4.

RESULTS
Measurement of plant capability post commissioning was demonstrated by the newfound ability to process lower feed magnetic content ores while maintaining production rate. The following parameters have been assessed since commissioning:

- The removal of the RMS screens resulted in a 98% reduction in HPGR recycling load, resulted in an increase of up to 25% in fresh feed rates, depending on the feed magnetic content,
- The de-constraining of the ball mill circuit and installation of the IsaMill has allowed for the processing of the coarser HPGR product, while maintaining design throughput rate and maintaining final grind of 38µm.
- Average concentrate pumping rate has been increased by 6% on the previous financial year, however 15% additional throughput has been demonstrated on higher mass recovery material.
- Finally, the plant has been successfully treating feed blended with 30% silica material successfully since commissioning. This was not achievable prior to the project completion.

CONCLUSIONS
The targeted de-bottlenecking of the magnetite concentrator has achieved its overall goals. The desired feed changes have been incorporated into the production budget and the plant is more than able to maintain grade and throughput.

ACKNOWLEDGEMENTS
The authors would like to acknowledge firstly Arrium Mining Pty Ltd for the approval to publish this paper. Secondly, to all the subject matter experts over the life of the project who have lent their expertise to allow us to achieve our goals.

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FIGURE CAPTIONS

FIG 1 Grinding transfer point comparison between old and new plants

FIG 2 – Revised flowsheet, as constructed

FIG 3 – Risk Management Process

FIG 4 – Ramp Up Curve

TABLE CAPTIONS

TABLE 1: Examples of fine grinding ball mills (Partyka, Yan, 2007).
FIGURES

FIG 1 – Grinding transfer point comparison between old and new plants

FIG 2 – Revised flowsheet, as constructed.

**Risk Management Process**

- **Assess**
- **Risk Classification**
  - Review of the available risk lists
  - Transition Risk Review Workshop
  - Consolidate risk lists into risk records
- **Monitor**
- **Risk Management**
- **Evaluate**
- **Prepare Risk Management Plan**
  - Risk workshops
  - Business Risk Management Guide
    - Risk categorisation, key risk indicators & consequences, weakness & strength, allocation of action plans accountabilities and due dates
- **Execute**
  - Execute the mitigation action plans
  - Assess the residual risk
- **Close out action plans**

**Measure & monitor performance**
- Weekly reporting
- Transition Progress Report

FIG 4 – Ramp Up Curve.

**Cumulative Concentrate Pumped from Startup**

- Cumulative Concentrate Pumped
- Cumulative Ramp Up Target

Contents
TABLE 1: Examples of fine grinding ball mills (Partyka, Yan, 2007).

<table>
<thead>
<tr>
<th>Site</th>
<th>Feed Size μm</th>
<th>Product Size μm</th>
<th>Dia (m)</th>
<th>Length (m)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pajingo Gold</td>
<td>200</td>
<td>38</td>
<td>3.66</td>
<td>4.18</td>
</tr>
<tr>
<td>Germano Iron Ore</td>
<td>120</td>
<td>32</td>
<td>5.18</td>
<td>10.36</td>
</tr>
<tr>
<td>Savage River</td>
<td>140</td>
<td>43</td>
<td>3.9</td>
<td>8.8</td>
</tr>
<tr>
<td>Macraes</td>
<td>20</td>
<td>3</td>
<td>3</td>
<td>8.2</td>
</tr>
<tr>
<td>Pena Colorado</td>
<td>125</td>
<td>38</td>
<td>5</td>
<td>10.67</td>
</tr>
<tr>
<td>Beaconsfield</td>
<td>20</td>
<td>1.83</td>
<td>2.44</td>
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<td>Tritton Copper</td>
<td>45</td>
<td>30</td>
<td>2</td>
<td>3.4</td>
</tr>
<tr>
<td>Brunswick Mining</td>
<td>30</td>
<td>25</td>
<td>3.2</td>
<td>4</td>
</tr>
<tr>
<td>Porgera Gold</td>
<td>106</td>
<td>30</td>
<td>3.05</td>
<td>4.27</td>
</tr>
<tr>
<td>Arrium</td>
<td>500</td>
<td>38</td>
<td>6.1</td>
<td>11</td>
</tr>
</tbody>
</table>
Using Geometallurgy during Process Optimisation Activities at the Southern Middleback Ranges Magnetite Concentrator

J Gerber¹, C Stanton² and D Olwagen³

ABSTRACT
From December 2013 to February 2014 the Southern Middleback Ranges magnetite concentrator went through a successful optimisation process that included the installation of extra grinding power to treat a wider range of ore types. After the initial improvements, sampling surveys were conducted to identify the mineral deportment through the concentrator. Samples collected during the survey were analysed using X-ray fluorescence, quantitative X-ray diffraction and QEMSCAN™ to cross-validate the analysis. Certain streams were also put through a laboratory-scale Davis Tube Recovery washing process to produce clean concentrates for analysis. The QEMSCAN™ analysis identified various metallurgical characteristics (mineral association, particle sizes, grain sizes, particle densities, etc) and these were evaluated for each separation and comminution stage to both understand the processes involved and identify opportunities for improvement. Key findings from the investigation indicated the importance of understanding the elemental deportment of target elements throughout the various minerals. The standard practice of tracking iron deportment through assay provided misleading results on losses to tail through the plant. The investigation provided insight into both wanted and unwanted minerals and the efficiency of each separation stage. The efficiencies of separation are critical due to their impact on downstream capacity, and the investigation provided insight into where opportunities exist to extract more benefit. The findings of the geometallurgical analysis have been implemented on the concentrator, and improvements in the rejection of non-magnetic minerals have been seen.

INTRODUCTION
The Southern Middleback Ranges (SMR) magnetite concentrator was commissioned in 2007 to produce feed for the OneSteel blast furnace in Whyalla, South Australia. The concentrator went through an optimisation project in December 2013, which included the installation of a M10000 IsaMill™ and Derrick Stack Sizer™ screens to give it the ability to treat ore with higher silica content. Magnetite ore is mined from the Iron Magnet pit and then crushed and stockpiled for feed to the concentrator. The stockpiled feed is fed to the high-pressure grinding rolls circuit that produces a coarse grind feed for the first magnetic cobbing stage. The magnetic concentrate then reports to the ball mill for another stage of grinding. The ball mill product goes through a classification stage before reporting to the second stage of magnetic separation. The magnetic concentrate from the second stage reports to the newly installed IsaMill™. The finely ground product is passed through the newly installed Derrick Stack Sizer™ screen system before reporting to the final magnetic cleaning stage. This stage produces the final concentrate that is thickened and pumped down a 63 km pipeline to the filter and pellet plant located in Whyalla.

During the optimisation process, it was identified that the concentrator had increased levels of unwanted elements (predominantly silica) in the final concentrate. While metallurgical characterisation of the ore indicated that marginal increases in the concentration of silica could be expected, the levels seen in the concentrator were significantly higher.

Initial investigations started with a snapshot survey of the major processing streams during stable operation. The analysis of the snapshot surveys included X-ray fluorescence (XRF), quantitative X-ray diffraction (QXRD), QEMSCAN™ and Davis Tube Recovery (DTR) washing. The findings from the snapshot survey indicated high levels of entrainment of non-magnetic minerals in the magnetic concentrates. Discussions with subject-matter experts confirmed that the efficiencies around the magnetic separation stages were lower than expected. The snapshot
survey also found that a significant quantity of non-magnetic iron was being rejected at the first stage magnetic cobbing.

The surveys indicated that the IsaMill™ improved the liberation of the silicates from the magnetite. This led to a more than 30 per cent reduction in the magnesium oxide (MgO) levels in the final concentrate.

These findings led to more detailed focus points around:
- first stage magnetic cobbing, with an emphasis on both the magnetic concentrate and non-magnetic rejects
- the IsaMill™ feed and discharge streams, with an emphasis on silicates liberation.

MINERALOGY

Mineralogical analysis helps to identify and understand the mineral compartment through a minerals processing plant. Understanding which minerals are reporting to the product stream and the reject stream can assist with optimising the efficiency of separation. It also provides understanding of quality constraints on the product stream and the losses of valuable metals to the reject/tail stream.

The current analysis method employed by Arrim Minings Geometallurgy department uses a combination of measuring techniques to validate results across analysis methods. The analysis techniques that are currently employed are:
- XRF
- XRD
- QEMSCAN™
- optical microscopy.

It was found that the combination of XRF, XRD and QEMSCAN™ works well for samples that have a top particle size of 1–3 mm. Samples that contain particles larger than 3 mm are analysed with XRF, XRD and optical microscopy. As discussed by Donskoi et al. (2011), the QEMSCAN™ has trouble differentiating between minerals with similar chemical composition (such as hematite and magnetite). The use of other techniques improves the mineralogists’ ability to differentiate these minerals when setting up the species identification profile.

Arrim has a good working relationship with the mineralogists at the Bureau Veritas laboratory in Adelaide. This has led to Arrim using the QEMSCAN™ iExplorer viewing software on-site to comprehensively investigate the analysis results. The use of the software has assisted greatly in reviewing metallurgical characteristics such as:
- grade-density profiles
- grade-particle size profiles
- particle shape analysis
- elemental deportment.

The biggest benefit associated with using the iExplorer software on-site is the ability to modify graphs as the investigation progresses and new avenues of investigation are pursued.

MAGNETIC SEPARATORS

Process improvement

Discussions held with Wennen (April 2014, personal communication) indicated that the weight per cent solids of the feed slurry to the magnetic separators was too high. Plant trials were completed and the feed slurry density was decreased in multiple stages to identify any improvements in the entrainment levels of non-magnetic minerals. The lowest density achievable was limited by the capacity of the water supply system. The samples taken during the trial were washed using a DJR method. The amount of non-magnetic material measured by this test was assumed to be the entrained material that should have been rejected. The results were reported as mass per cent magnetic recovery. Figure 1 shows the magnetic recoveries for the various trials. A lower magnetic recovery indicates a higher level of entrainment. The test work confirmed that at lower slurry densities, the entrainment decreases. The concept of lower feed density was embraced by the production personnel, who immediately reduced the density to the lowest possible level with the current infrastructure. A review is currently underway to optimise the water distribution system to allow more flexibility in controlling the slurry densities to the magnetic separation stages.

![Graph showing magnetic mass recovery](image1.png)

**FIG 1** - The magnetic mass recovery for various per cent feed solids by weight for the first magnetic cobbing stage.

Mineral deportment

The samples taken during the trials were sent for mineralogical analysis to understand which minerals were being entrained, recovered or rejected at the magnetic separation stages. The feed to the various magnetic separation stages showed that the main mineral groups contributing to iron content were:
- iron oxides (magnetite, hematite) – 78 per cent
- iron silicates (chlorite, minnesotaite) – 14 per cent
- carbonates (siderite, ankerite) – seven per cent
- iron sulfides (pyrite, pyrrhotite) – one per cent.

The magnetic separation recovers any particle that contains magnetic minerals as long as the magnetic force is larger than any of the other forces being applied to the particle (drag force, gravity, particle interaction forces etc). Depending on the size and weight of a particle, only a small amount of magnetic material is needed for the particle to be recovered (Figure 2). From the minerals listed, it was found that only magnetite and pyrrhotite responded positively in the presence of the magnetic field strength used in the concentrator. This limited the iron recovered to those magnetic minerals plus any of the other minerals being entrained during the separation stage.

The magnetite-hematite association is such that a large portion of the hematite is recovered in the concentrate. The XQRD analysis (Table 1) showed hematite and magnetite levels as high as nine per cent and 61 per cent respectively. Other main minerals identified were quartz, talc (in the form of minnesotaite), dolomite/ankerite, siderite and chlorite.

Analysis of the tails indicated that the hematite lost was not associated with the magnetite but rather with iron silicates or as liberated hematite particles. The pyrrhotite was present...
in very small quantities and, from the QEMSCAN™ particle view data, was typically rejected due to the particle sizes it was found in. The particle density and particle size for hematite present in the rejects from the magnetic cobbing stage is shown in Figure 3. Sixty-four per cent of the particles in the cobbing rejects were larger than 45 μm. These particles were rejected due to low magnetic content, but still contained significant amounts of iron (12 per cent by mass) at grades of around 60 per cent. Recovery of these particles with a gravity-based concentration stage is possible.

**ISAMILL™**

**Process improvement**

The original Project Magnet concentrator flow sheet consisted of two stages of grinding to produce 1.8 Mt/a of slurry concentrate at a grind of 38 μm (P80). This grind was necessary to ensure that the concentrate quality was within the blast furnace requirements; however, to meet ongoing throughput and grind requirements and increase the range of feed materials that can be processed, the optimisation

**TABLE 1**

Quantitative X-ray diffraction results from the surveys for the concentrate from the first magnetic cobbing stage.

<table>
<thead>
<tr>
<th>Mineral</th>
<th>Composition</th>
<th>Survey 1</th>
<th>Survey 2</th>
<th>Survey 3</th>
<th>Survey 4</th>
<th>Survey 5</th>
</tr>
</thead>
<tbody>
<tr>
<td>Quartz</td>
<td>SiO₂</td>
<td>10</td>
<td>16</td>
<td>11</td>
<td>11</td>
<td>12</td>
</tr>
<tr>
<td>Magnetite</td>
<td>Fe₃O₄</td>
<td>61</td>
<td>53</td>
<td>56</td>
<td>60</td>
<td>59</td>
</tr>
<tr>
<td>Hematite</td>
<td>Fe₂O₃</td>
<td>9</td>
<td>9</td>
<td>7</td>
<td>7</td>
<td>7</td>
</tr>
<tr>
<td>Chlorite</td>
<td>(X,Al)(Al₂(Si₂O₅)₃(OH)₆</td>
<td>5</td>
<td>5</td>
<td>4</td>
<td>3</td>
<td>2</td>
</tr>
<tr>
<td>Talc (magnesite)</td>
<td>Fe₃O₄</td>
<td>4</td>
<td>3</td>
<td>10</td>
<td>8</td>
<td>11</td>
</tr>
<tr>
<td>Dolomite/ankerite</td>
<td>CaFe₂⁺, Mg, Mn(CO₃)₂</td>
<td>5</td>
<td>8</td>
<td>7</td>
<td>7</td>
<td>6</td>
</tr>
<tr>
<td>Siderite</td>
<td>FeO</td>
<td>5</td>
<td>4</td>
<td>5</td>
<td>4</td>
<td>4</td>
</tr>
<tr>
<td>Sepiolite</td>
<td>Mg₂Si₂O₅(OH)₂(H₂O)</td>
<td>2</td>
<td>2</td>
<td>-</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td>Total</td>
<td></td>
<td>100</td>
<td>100</td>
<td>100</td>
<td>100</td>
<td>100</td>
</tr>
</tbody>
</table>

Chlorite where X = Mg, Fe, Ni and Mn.

FIG 2 – Three magnetite concentrate particles from the first magnetic cobbing stage, each containing the three dominant non-magnetic minerals (quartz, carbonates and iron silicates) together with magnetite.
study concluded that the plant required a third grinding stage to ensure ongoing compliance with concentrate specification. For the upgraded concentrator flow sheet, the installed MI1000 IsaMill™ sits downstream of the existing 7.5 MW ball mill. With the installation of the IsaMill™, the throughput capability of the concentrator was increased by 20–30 per cent while maintaining the final grind and quality specification.

Mineral deportment

The IsaMill™ was installed as an energy-efficient tertiary grind stage to improve the liberation of magnetite from

**Fig 3** – Three-dimensional figures showing the relationships between particle density, particle size and both the mass distribution and iron assays for each grouping of the magnetic cobbles rejects.

**Fig 4** – Particle view report for the IsaMill™ feed and discharge streams arranged by the shape factor of the particles using the QEMSCAN™ Discover software package.
the other non-magnetic minerals, predominantly quartz, ankerite and talc. The samples analysed to identify the improved liberation from the tertiary grinding stage were the IsaMill™ feed and discharge. A DTR wash was completed on both samples as well as mineralogical analysis with the QEMSCAN™. The DTR wash in the feed showed a magnetic recovery of 84 per cent and the discharge was 78 per cent. The drop of 6 per cent in magnetic recovery indicates that more non-magnetic material was liberated from the magnetic particles in the IsaMill™ discharge.

The grade of the concentrate between the feed and discharge showed an improvement from 7.2 per cent to 4.9 per cent for silica and 0.9 per cent to 0.07 per cent for MgO. The improvement in MgO liberation is a significant improvement when compared to historical performance. Figure 4 shows how the elongated shapes of the IsaMill™ feed particles have been ground smaller and the amount of elongated particles present in the material reduced. The majority of the elongated particles are dominated by the ‘silicates’ mineral grouping.

The improvement has led to adjustments in the fluxing philosophy at the pellet plant, which is delivering cost benefits by using cheaper dolomite, rather than limestone, to achieve the required flux levels.

USING GEOMETALLURGY DURING PROCESS OPTIMISATION ACTIVITIES

CONCLUSIONS

Mineralogical investigations have proven quite useful for Arrium Mining. The use of analysis tools is expanding towards not just fault finding, but also to optimisation exercises and opportunity investigations. It is important to use multiple techniques to ensure that measurements and interpretations are representative. Mineral deportment analysis on the concentrator has identified potential iron recovery gains. These gains are currently being investigated for feasibility.

ACKNOWLEDGEMENTS

The authors would like to thank Mike Bannear for allowing this paper to be submitted and the groups involved during the optimisation process. A special thank you to Wade Hodgson and Barry Whittington from Bureau Veritas for taking the time to train one of the authors in using iExplorer.

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Comminution 14
*Larson M, **Morrison R, **Xie W, *Young M
Development of the Larson/Morrison IsaMill JKSimMet Model

*Glencore-XT
**JKMRC, SMI, the University of Queensland

Abstract

The IsaMill is a high intensity horizontal stirred mill utilizing small 2-6 mm ceramic grinding media for attrition grinding. Grinding duties range from feeds of up to 300 microns being ground to 40 microns and UFG grinds as fine as 5 microns. Being completely dissimilar to normal ball mill breakage it was desired to produce a model for this process in JKSimMet. The result of this study was the discovery that IsaMill breakage can be reliably predicted on a basis of energy versus the squared value of the percent passing given sizes. This relationship can be used to analyse circuit efficiency for varying feed sizes. By using this new relationship the product size distributions for new feed sizes and energies can be reliably predicted with simple math, something not previously possible with only a signature plot. This paper will provide details on the model development, validation and implementation.

Introduction

The IsaMill is a horizontal high speed attrition mill developed by Mt Isa Mines and Netzsch for the treatment of fine grained minerals at McArthur River Mine and Mt Isa where ultrafine grinds of sub 7 microns are necessary. (Johnson et al, 1998) As the development of the mill and ancillary items has progressed the mill has expanded its range of duties into relatively coarser concentrate and mainstream grinding duties with F80’s of up to 300 microns. As acceptance of the mill has increased so too has the need to better understand the grinding action taking place. In 2006 Glencore-XT funded the JKMRC to develop a simulation model of the mill breakage. While the mill scaleup based on the proven signature plot is widely accepted, the signature plot has one major drawback in its comparison of P80 versus energy. The plot is only valid for the feed size and conditions tested. If a coarser feed size became part of the design criteria no adjustment was available as the signature plot line does not intersect zero energy at the feed size and cannot be reliably adjusted. A method is required to accurately predict the results of not only changing feed sizes but also different media size and rheological constraints.

IsaMill Grinding Variables

There are many variables within the IsaMill or any stirred mill; however, only a handful of these actually impact the grinding efficiency. Residence time and energy intensity may change but the actual energy input per unit of feed (kWh/t) to a given grind target is a quite robust number. Variation in factors such as disc shape, spacing, and size; mill speed, feed slurry density within a range, separator configuration, flowrate and media type as long as relatively round and an SG of 3.5-4 do not have much impact on the energy per tonne required for the same feed size distribution to achieve a target product size. This is well supported by experience as
the original scaleup as performed by MIM went from lab units of 1.5 to 4 litre capacity to pilot and industrial scale units of 100 to 3000 litres per mill.

If the different mill configuration variables made a difference in grinding efficiency the team at MIM could be considered some of the luckiest engineers in the world in that each step scaled along the signature plot. Further, past work by Ming Wei Gao at CSIRO has shown the lab four litre IsaMill scaled from energy versus P80 to the full size 355 kW Century Zinc SMD’s (Gao et al, 2007). This is despite the fact that the vertical SMD uses pegs to stir the media and slurry and the horizontal M4 and all IsaMills use discs at a higher speed. Similarly the lab scale IsaMills use round pegs with a large open area for a separator while full scale mills use square or rhomboidal fingers with far less relative open area. In all cases the mills are just stirring grinding media and slurry. With regards to mill speed the M4 has a tip speed of about 8.5 m/s, the M20 about 12-14 m/s and full scale mills about 20 m/s, yet they all scale from one another. The relative mill configurations also vary from size to size with regards to disc spacing and size compared to the shell volume. This suggests that the energy being absorbed by shearing the mixture of media and slurry may be the controlling factor. The sensitivity to changes in the shape of the feed size distribution also suggests that coarser fed particles may be contributing to grinding of the finer ones. Over complicating any of this serves no real purpose other than to distract from issues which actually do make a difference.

These can be summarized as media shape, media size, ore type, slurry viscosity (above an optimum point but not below that point) and feed size distribution.

Media shape is generally not considered as all but three IsaMill sites use modern ceramics which in most cases are generally round i.e. close to spherical. Full scale plant test work and pilot test work has shown that the round ceramic is generally 15-20% more efficient than competent sand or oblong shaped ceramics. Different ceramic manufacturers may claim energy efficiency advantages over competitors but the reality is that this media has become much more of a commodity where purchase price and wear rate (g/kWh) is a distinguishing factor, not grinding efficiency.

Media size has only been examined in limited detail but it is known that different size media will produce different energy versus grind slopes for each size fraction. Optimum media size will depend on the incoming ore feed size, desired product size, ore hardness and a myriad of other factors to lesser extents.

Slurry viscosity and its effect on stirred milling efficiency at fine sizes is poorly understood at this time. In general Glencore-XT will try to operate mills at about 20% solids by volume or 20-24 centipoise on the mill discharge. Above these values energy efficiency is lost. Below these values the efficiency will flatline before decreasing once the slurry is dilute enough that energy is spent contacting media against media with no ore particles present to be ground. Ideally in the future there will be a reliable model to predict viscosity based on the surface area present in a given volume calculated from the solid SG, slurry SG and laser sizer product size distribution.
**Signature Plot and Scaleup**

The ability to predict energy for varying feed size distributions is the main goal of the work described in this paper. The IsaMill signature plot with energy (kWh/t) versus P$_{80}$ on a log-log plot is widely accepted as scaling to full scale IsaMills. This has been published over the past two decades showing accurate 1:1 scaleup from M4 and M20 test mills to the M1000, M3000 and M10000 IsaMill. Glencore-XT maintains relationships with a dozen independent laboratories around the world. The robustness and repeatability of the signature plot test was shown in IsaMill 1:1 Direct Scaleup from Ultrafine to Coarse Grinding presented at Comminution 12. However, it has one large limitation in that it does not cross zero energy at the feed size. Thus it can only be used for the feed size tested. If the same P$_{80}$ is to be fed to the mill and the rest of the size distribution is different - for example far fewer fines or a sharper size distribution - then the signature plot will not be valid. A new method is necessary to provide the flexibility required to be useful for a JKSimMet model which must to able to handle reasonable variation in feed rate and feed size distribution in a reasonable manner. The proven robustness of the signature plot though means that any model will still use the signature plot. Given the fact that the entire size distribution, not just the P$_{80}$, can be determined from the signature plot test it can be assumed that any new model method based on the same information will be successful.

**Test Mill Coarse Material Retention**

Based on previous work it cannot be stressed enough that proper test conditions and data analysis are vital to the successful development and implementation of any IsaMill model. Without careful planning and observation results can deceive the user into thinking the mill performed better than reality. The combination of too small media and not enough sample volume can easily result in coarse material being retained in the lab mill. In these cases the too small media will only grind the fines and not have enough energy to break the coarsest material. The coarse material will be held in the mill by the centripetal forces as there is not enough material passing through the mill for those particles to fully build up where they would be discharged. This is usually recognized by either an unusually high power draw or a low density reading on the discharge.

**Flaws in Other Models**

It is well accepted that the Bond equation becomes much less reliable at product sizes finer than about 70 microns. As ore is ground finer, the key Bond assumption that the next increment of input energy will produce a similar increment of fine material (i.e. g/rev of the test mill) becomes invalid and the required energy for each increment of fines generated increases at finer product sizes. This means that the relationship between grind size and energy is not linear at fine sizes.

As shown in Figure 1, grinding from 50 to 33 μm requires significantly more power than the BWI suggests, with a much higher slope (-1.0345). Grinding finer still needs ever more energy. All three points on the graph come from the same test but the finer points look to be less energy efficient in terms of a Bond comparison.
Further, most Bond Work Index data is developed by testing run of mine ore. Most IsaMill feed will be a stream with quite different characteristics. The Bond test uses steel balls, while the IsaMill uses ceramic grinding media. In almost all cases the ceramic media will be harder than the ore it is grinding and of roughly similar stiffness. The steel balls used in a Bond test can vary widely in hardness compared with the ore being ground and will always be much more elastic. Stiffer media transfer energy more efficiently than elastic media.

The commonly used appearance function will also not be considered in this case as it assumes every particle breaks with the same progeny. The breakage rates are modelled but not changed for different energy, feed sizes or media size.

Other researchers (Mannheim, 2011) have attempted to utilize the existing signature plot line to predict results at different feed sizes. This is shown in Figure 2.
Forcing the signature plot lines through the origin in Figure 2 moves the lines from the original plots using only the mill products. While it may appear accurate at first glance, further inspection reveals that the energy predicted by the resulting line is in error by 200-300% compared to where the points actually sit. The use of the log-log plot with inclusions from 0 to 1 kWh/t and large symbols have effectively hidden this from casual observation. If samples had been taken below 100 kWh/t they would better demonstrate this inaccuracy. This approach cannot be considered sufficiently reliable for a model that will be used to size industrial machinery.

Reduction ratio type comparisons have also proven to be popular. In the example from Tati Nickel contained in Table 1, energy requirements were measured to differentiate between three types of stirred mills.

Table 1. Summary of mill performances (Nel et al, 2006)

<table>
<thead>
<tr>
<th>Metric</th>
<th>Mill A</th>
<th>Mill B</th>
<th>Mill C</th>
</tr>
</thead>
<tbody>
<tr>
<td>Specific cumulative breakage rate at 10μm</td>
<td>0.15</td>
<td>0.02</td>
<td>0.055</td>
</tr>
<tr>
<td>Power at Reduction ratio of 4 with Ceramic</td>
<td>kW/h/t</td>
<td>*55</td>
<td>48</td>
</tr>
<tr>
<td>Power at Reduction ratio of 4 with Sand</td>
<td>kW/h/t</td>
<td>*110</td>
<td>97</td>
</tr>
<tr>
<td>Temperature increase</td>
<td>°C/(kW/h/t)</td>
<td>N/A</td>
<td>0.71</td>
</tr>
<tr>
<td>Temperature for 60 kW/h/t</td>
<td>°C</td>
<td>N/A</td>
<td>42.6</td>
</tr>
</tbody>
</table>

*NAt low density (30% solids).

Nel et al concluded “For the horizontal ultra fine mill with ceramic grinding media, 2.4% of the particles bigger than 10 microns were reduced to less than 10 microns using 1 kWh/t specific energy input, thus 33.3 kWh/t specific energy was required to mill Phoenix concentrate to 80% passing 10 microns.”
The math used is basically that it took 1 kWh/t to reduce 2.4% of the feed to under 10 microns. Therefore it must take 33.3 kWh/t to reduce 80% of the feed to below 10 microns. 80%/2.4%=33.3. This approach while being convenient is not necessarily appropriate.

When examining grinding efficiency in this way the examples with the coarsest feed and or product will always appear to be the more efficient option. For reduction ratios to be a valid option in modelling of fine grinding, the relationship between grind size and energy would have to be linear. Unfortunately this is not the case and each step finer requires relatively more energy for smaller and smaller reductions in size.

Figure 3 is another example of a reduction ratio model used to justify a process decision, in this case comparing media type and size. However if the extreme point from the test work was used, the entire graph could be populated by columns of dots. For example, grinds of 60 microns to 30 microns and 40 microns to 20 microns both have reduction ratios of 2, but the finer example will require more energy.

![Figure 3. Comminution energy versus reduction ratio (Farber et al, 2010)](image)

For these reasons reduction ratios are of limited use for comparison or prediction tools when dealing with fine grinding.

**The Squared Function Dependence**

Using the Finch McIvor method for particle production in ball mills as a guide, the first attempt at modelling the IsaMill on a size by size particle basis was made. While the relationship of energy versus percent passing of a certain size is linear for the ball mill, for the IsaMill the relationship was not linear. However it was observed from the curve produced that the relationship approximated a squared function.
Plotting these first attempts on energy versus percent passing 10 µm squared resulted in Figure 4. Not only does the squared function produce a straight line but that line passes through the feed size at zero energy.

![Copper Energy vs 10um Production](image)

Figure 4. Fines production of various copper ores (Larson, 2013)

This shows promise that the mill is creating surface area through fines in a predictable reliable manner.

It was initially thought that by switching the axes, a simple easy to understand model could be used with the squared function as the basis. This is shown in Figure 5. A plot is developed using an original feed and energy values. Then for a new feed the new size is plotted and the line from the original feed is just moved up or down the y-axis while keeping the squared function lines parallel. The new energy to a given percent passing the size of interest can then be calculated.

![Fines Production vs Energy Prediction](image)

Figure 5. Graphical form of IsaMill squared function for fines production model (Larson, 2013)
Validation of Squared Dependence in MS Excel

To validate the basic idea of the squared function, the JKMRC ran a test program with multiple ores ground to varying size distributions in a rod mill. Different feed sizes were then ground in the M4 IsaMill using the same media size distribution to determine if one test could be used to predict the results of the other.

In the first case shown in Table 2, MIM run of mine copper ore was ground down to P80’s of 72 and 47 microns. Both samples were processed through the M4 IsaMill using the same 2.5 mm top size graded ceramic media charge. The raw data for these two tests is shown below, with energy per pass, percent passing a given size and the percent passing that size squared.

Table 2. MIM Copper ROM sizing and energy per pass data (Larson, 2013)

<table>
<thead>
<tr>
<th>Energy</th>
<th>P80=72 μm</th>
<th>2.4 μm</th>
<th>5 μm</th>
<th>9.6 μm</th>
<th>13.5 μm</th>
<th>19 μm</th>
<th>27 μm</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
<td>15.00</td>
<td>225.00</td>
<td>25.28</td>
<td>639.08</td>
<td>38.13</td>
<td>1453.90</td>
<td>45.00</td>
</tr>
<tr>
<td>9.3</td>
<td>18.33</td>
<td>335.99</td>
<td>31.46</td>
<td>989.73</td>
<td>48.41</td>
<td>2343.53</td>
<td>57.57</td>
</tr>
<tr>
<td>19.2</td>
<td>20.62</td>
<td>425.18</td>
<td>35.72</td>
<td>1275.92</td>
<td>55.56</td>
<td>3089.14</td>
<td>66.40</td>
</tr>
<tr>
<td>28.9</td>
<td>23.31</td>
<td>543.36</td>
<td>40.10</td>
<td>1608.01</td>
<td>62.33</td>
<td>3885.03</td>
<td>74.40</td>
</tr>
<tr>
<td>37.7</td>
<td>24.31</td>
<td>590.98</td>
<td>42.16</td>
<td>1777.47</td>
<td>65.77</td>
<td>4325.69</td>
<td>78.17</td>
</tr>
<tr>
<td>44.10</td>
<td>1944.81</td>
<td>69.31</td>
<td>4803.88</td>
<td>81.93</td>
<td>6712.52</td>
<td>96.91</td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Energy</th>
<th>P80=47 μm</th>
<th>2.4 μm</th>
<th>5 μm</th>
<th>9.6 μm</th>
<th>13.5 μm</th>
<th>19 μm</th>
<th>27 μm</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
<td>16.22</td>
<td>263.41</td>
<td>28.11</td>
<td>790.17</td>
<td>43.49</td>
<td>1891.38</td>
<td>51.70</td>
</tr>
<tr>
<td>8.3</td>
<td>19.26</td>
<td>370.95</td>
<td>33.63</td>
<td>1130.98</td>
<td>52.71</td>
<td>2778.34</td>
<td>62.09</td>
</tr>
<tr>
<td>17.3</td>
<td>21.40</td>
<td>457.96</td>
<td>37.34</td>
<td>1394.28</td>
<td>58.51</td>
<td>3423.42</td>
<td>69.94</td>
</tr>
<tr>
<td>24.9</td>
<td>22.73</td>
<td>516.65</td>
<td>40.17</td>
<td>1613.63</td>
<td>63.30</td>
<td>4006.89</td>
<td>75.75</td>
</tr>
<tr>
<td>32.2</td>
<td>24.67</td>
<td>608.61</td>
<td>43.09</td>
<td>1856.75</td>
<td>67.48</td>
<td>4553.55</td>
<td>79.92</td>
</tr>
<tr>
<td>39.5</td>
<td>26.45</td>
<td>699.60</td>
<td>45.99</td>
<td>2115.08</td>
<td>71.83</td>
<td>5159.55</td>
<td>84.05</td>
</tr>
</tbody>
</table>

The main basis for this model is the assumption that the squared function lines for varying size fractions will be parallel for different feed size distributions and that the predictive process only has to move those lines up and down the y-axis depending on the amount of a given size present in the new feed to predict the new energy. The results of doing this are shown in Figure 6 with individual lines for 27, 19, 13.5, 9.6, 5 and 2.4 microns plotted for both feeds tested.
Figure 6. Complete comparison of different feed size squared function lines

It is assumed that as media size is changed to larger media, the lines for the coarsest size fraction will steepen and the finer size fraction will flatten. For smaller media, the coarsest fraction will flatten and the finer fraction will steepen as fine particles are preferentially broken compared to larger particles by the smaller media.

The actual slopes and error associated with Figure 6 are detailed in Table 3 below. The results of the 47 micron feed were plotted and used to predict the product size distribution shown in Figure 7. Given the +/- 5% error associated with the M4 signature plot test this can be considered a successful demonstration of the squared function model.

Table 3. MIM Copper ROM squared function predicted versus actual (Larson, 2013)

<table>
<thead>
<tr>
<th>Size fraction (microns)</th>
<th>Slope 47</th>
<th>Slope 72</th>
<th>Error</th>
<th>New feed intercept</th>
<th>New feed predicted %passing</th>
<th>New Feed actual %passing</th>
</tr>
</thead>
<tbody>
<tr>
<td>2.4</td>
<td>10.658</td>
<td>10.013</td>
<td>6.05%</td>
<td>225</td>
<td>24.93</td>
<td>24.31</td>
</tr>
<tr>
<td>5</td>
<td>32.6</td>
<td>30.85</td>
<td>5.37%</td>
<td>639.1</td>
<td>43.03</td>
<td>42.2</td>
</tr>
<tr>
<td>9.6</td>
<td>80.47</td>
<td>77.63</td>
<td>3.53%</td>
<td>1453.9</td>
<td>66.69</td>
<td>65.77</td>
</tr>
<tr>
<td>13.5</td>
<td>108.81</td>
<td>110.79</td>
<td>1.82%</td>
<td>2025</td>
<td>77.93</td>
<td>78.17</td>
</tr>
<tr>
<td>19</td>
<td>136.65</td>
<td>138.2</td>
<td>1.13%</td>
<td>2650.2</td>
<td>87.94</td>
<td>88.05</td>
</tr>
<tr>
<td>27</td>
<td>165.72</td>
<td>175.56</td>
<td>5.94%</td>
<td>3376.8</td>
<td>97.68</td>
<td>95.65</td>
</tr>
</tbody>
</table>
A second validation example is detailed in Table 4. In this case, MIM lead/zinc run of mine ore was ground in a pilot rod mill to $P_{80}$’s of 131 and 68 microns. Both samples were then processed with the same 3.5 mm top size graded charge of media.

**Table 4. MIM Lead Zinc ROM sizing and energy per pass data (Larson 2013)**

<table>
<thead>
<tr>
<th>$P_{80}$</th>
<th>Energy</th>
<th>$%e$</th>
<th>$%e^{2}$</th>
<th>$5 \mu m$</th>
<th>$9.6 \mu m$</th>
<th>$13.5 \mu m$</th>
<th>$19 \mu m$</th>
<th>$27 \mu m$</th>
</tr>
</thead>
<tbody>
<tr>
<td>131 $\mu m$</td>
<td>0</td>
<td>17.95</td>
<td>322.32</td>
<td>25.92</td>
<td>672.02</td>
<td>34.11</td>
<td>1163.72</td>
<td>38.02</td>
</tr>
<tr>
<td>7.9</td>
<td>27.14</td>
<td>736.58</td>
<td>40</td>
<td>1600.00</td>
<td>53.88</td>
<td>2903.05</td>
<td>60.64</td>
<td>3677.82</td>
</tr>
<tr>
<td>15.0</td>
<td>31.14</td>
<td>969.70</td>
<td>46.61</td>
<td>2172.49</td>
<td>61.93</td>
<td>4087.68</td>
<td>72.9</td>
<td>5144.41</td>
</tr>
<tr>
<td>22.0</td>
<td>34.22</td>
<td>1171.01</td>
<td>51.81</td>
<td>2684.28</td>
<td>72.42</td>
<td>5244.66</td>
<td>82.42</td>
<td>6793.06</td>
</tr>
<tr>
<td>28.7</td>
<td>36.75</td>
<td>1301.41</td>
<td>55.30</td>
<td>3065.28</td>
<td>77.89</td>
<td>6066.85</td>
<td>87.03</td>
<td>7679.89</td>
</tr>
<tr>
<td>35.6</td>
<td>38.04</td>
<td>58.68</td>
<td>82.1</td>
<td>91.11</td>
<td>96.45</td>
<td>99.01</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>$P_{80}$</th>
<th>Energy</th>
<th>$%e$</th>
<th>$%e^{2}$</th>
<th>$5 \mu m$</th>
<th>$9.6 \mu m$</th>
<th>$13.5 \mu m$</th>
<th>$19 \mu m$</th>
<th>$27 \mu m$</th>
</tr>
</thead>
<tbody>
<tr>
<td>68 $\mu m$</td>
<td>0</td>
<td>23.45</td>
<td>550.06</td>
<td>33.63</td>
<td>1131.20</td>
<td>44.29</td>
<td>1961.90</td>
<td>49.43</td>
</tr>
<tr>
<td>6.9</td>
<td>28.6</td>
<td>817.96</td>
<td>42.32</td>
<td>1790.98</td>
<td>57.56</td>
<td>3331.15</td>
<td>65.44</td>
<td>4283.05</td>
</tr>
<tr>
<td>13.8</td>
<td>32.09</td>
<td>1029.45</td>
<td>48.26</td>
<td>2329.51</td>
<td>66.94</td>
<td>4481.63</td>
<td>76.03</td>
<td>5872.92</td>
</tr>
<tr>
<td>20.0</td>
<td>34.61</td>
<td>1197.85</td>
<td>52.77</td>
<td>2784.67</td>
<td>74.36</td>
<td>5529.41</td>
<td>84.44</td>
<td>7130.11</td>
</tr>
<tr>
<td>26.0</td>
<td>36.95</td>
<td>1365.30</td>
<td>56.44</td>
<td>3185.47</td>
<td>79.11</td>
<td>6259.18</td>
<td>88.72</td>
<td>7872.13</td>
</tr>
<tr>
<td>32.4</td>
<td>37.94</td>
<td>58.94</td>
<td>82.83</td>
<td>91.77</td>
<td>96.86</td>
<td>99.17</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

In this test, the slopes matched up very well, with most error at the coarsest size fraction. This is shown in Table 5 and in Figure 8.
Table 5. Lead/Zinc ROM squared function prediction versus actual (Larson, 2013)

<table>
<thead>
<tr>
<th>Size fraction (microns)</th>
<th>Slope 68 micron feed</th>
<th>Slope 131 micron feed</th>
<th>Error</th>
<th>New feed intercept</th>
<th>New feed predicted %passing</th>
<th>New feed actual %passing</th>
</tr>
</thead>
<tbody>
<tr>
<td>2.4</td>
<td>30.88</td>
<td>33.63</td>
<td>8.18%</td>
<td>322.3</td>
<td>34.76</td>
<td>36.075</td>
</tr>
<tr>
<td>5</td>
<td>78.36</td>
<td>82.41</td>
<td>4.91%</td>
<td>672.02</td>
<td>54.05</td>
<td>55.37</td>
</tr>
<tr>
<td>9.6</td>
<td>166.08</td>
<td>170.33</td>
<td>2.50%</td>
<td>1163.72</td>
<td>77.01</td>
<td>77.89</td>
</tr>
<tr>
<td>13.5</td>
<td>210.87</td>
<td>218.63</td>
<td>3.55%</td>
<td>1445.52</td>
<td>86.59</td>
<td>87.64</td>
</tr>
<tr>
<td>19</td>
<td>276.89</td>
<td>294.01</td>
<td>5.82%</td>
<td>1742.78</td>
<td>98.31</td>
<td>94.31</td>
</tr>
<tr>
<td>27</td>
<td>367.1</td>
<td>393.84</td>
<td>6.79%</td>
<td>2089.4</td>
<td>112.35</td>
<td>98.14</td>
</tr>
</tbody>
</table>

Figure 8. Predicted versus actual product size distribution MIM Lead/Zinc ROM (Larson, 2013)

The coarsest sizes tend to be more difficult to predict. As the sizes get closer to the $P_{100}$, there is much less of the material to be created and the curve starts to flatten and give a worse prediction if the data is not cut off correctly. It is also generally a size fraction that is more likely to be held in the mill and is difficult both to sample and to measure correctly in the Malvern.

**JKSimMet Model Implementation and Testing**

Despite the apparent success of validating the model in MS Excel, it was still necessary to prove that JKSimMet programming could be implemented to duplicate these results without manual adjustments to the data. The original signature plot data was provided to the JKMRC along with the feed size to a second test. Besides the new feed size distribution two energies were provided with the test from which the JKSimMet model was to predict the resulting product size distribution curves. The results comparing the model values to what was
actually ground in the M4 IsaMill are shown in Figures 9 and 10 for two MIM copper examples and one MIM lead/zinc sample.

**Figure 9.** JKSimMet IsaMill model predictions versus actual for MIM Copper ROM

**Figure 10.** JKSimMet IsaMill model predictions versus actual for MIM Lead/Zinc ROM

These cases can be considered a success as they accurately match the results generated without the need for manually adjusting the settings. In this case all original data over a $P_{95}$ is ignored automatically by the program.
Implementation into JKSimMet

A brief summary of the model implementation is included. The simplicity of the squared function allows for a relatively easy to understand model. The IsaMill spreadsheet model was first converted into a MatLab program to allow ease of parallel testing. Once this testing was complete, the ability to predict the full product size distribution was added using spline functions. The model was then converted into Fortran within the JKSimMet Model Developers kit. At this point, the full calculation of the signature plot from measured data was also added to ensure a reproducible approach to interpolation and calculation.

The squared function is applied as follows:
The first step is to determine the linear relationship of energy versus the square of percentage passing 9.6um (or other size fraction that is valid for use) from the original feed. Any data from an individual pass that is coarser than the P_{95} for that pass is disregarded;
The second step is to calculate the point of new feed with zero energy input in the map of energy vs. the square of percentage passing 9.6um;
The third step is to plot a line parallel to the line in the first step from the point of new feed of the second step;
The final step is to find the value of the percentage passing 9.6um at the new input energy.
After the above method applied for a few size classes, the product size distribution can be interpolated.
The model was then tested against a range of reliable test work data.

Using the Model as an Evaluation Tool

While reduction ratios cannot be used to evaluate the efficiency of different fine grinding options, the squared function can be used to a certain extent. With just the IsaMill model, different variables including feed sizes can be tested for efficiency. In the case of only comparing IsaMills, the slope of the squared function line at a size or sizes of interest can be compared. The steeper the line the more energy efficient the grind is going to be. This is not proven to work when comparing different technologies. When comparing against a ball mill or tower mill or other high speed stirred mill, the comparison would only be valid if both received the same feed size. In those cases the squared function line could be developed for the IsaMill and just the product vs. energy point plotted for the other technology. If that point was to the right of the IsaMill squared function line, that option would be less energy efficient and if it were to fall to the left of the IsaMill line it would be more efficient. If the feed sizes were different, the analysis would be invalid as the shape of the line is yet to be established for other technologies.

Future Work

Future work to fully develop this model will focus on how a change in media size affects the individual particle size slopes along with how mineralogy and ore SG will impact on energy efficiency. It is thought that as media size increases the spread between slopes will increase. That is the finer particle lines will flatten and the
coarser particle lines will steepen. These need to be able to be predicted to widen the size ranges that can be predicted where the change in feed size necessitates a change in media size.

Predicting the effects of mineralogy with regards to energy will be more difficult. There are some observations in XT test work database that a lower solid specific gravity in magnetite feed will result in a higher required energy to grind to a given target size. This may be due to the lower density material having more particles per tonne needing attrition breakage than a higher density magnetite feed along with the extra silica generally being harder. This explanation is more difficult when it comes to most copper and gold ores where the gangue material can vary more widely than just being silica. A pyrite gangue can increase the solid density but depending on the pyrite can either be very hard or quite brittle.

It is also accepted that the relationship between grinding achieved and incremental energy input becomes even more inefficient at still finer sizes. Hence, an increased exponent may be required for products very much finer than the “normal” IsaMill range.

Acknowledgements

The authors wish to thank Glencore-XT and the JKMRC for allowing this work to be published and Ming Zhao He and Paul Kay for performing the test work that went into the validation examples and improving the understanding of these breakage mechanisms.

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Farber B, Durant B, Bedesi N, 2010. Effect of Media Size and Mechanical Properties on Milling Efficiency and Media Consumption; Comminution 10; Cape Town, South Africa


Recent Process Developments at the Phu Kham Copper-Gold Concentrator, Laos
D Bennett¹, I Crnkovic² and P Walker³

ABSTRACT
The Phu Kham deposit represents a copper-gold porphyry system, with mineralisation present in skarn, stockwork and disseminated styles. Significant folding and alteration events have created a complex heterogeneous mineralogy horizon. Weathering and water table contact have created a leached zone, overlying transition zones with supergene chalcocite- dominant secondary copper mineralisation and clay-rich gangue. Primary ore copper mineralisation is mainly chalcopyrite with minor bornite. The major challenges to the copper-gold flotation process are a wide size distribution of chalcopyrite mineralisation and poor primary grind liberation, a high pyrite content in skarn ore requiring aggressive pyrite depression conditions, clay-rich gangue and non-sulfide copper mineralisation in weathered zones, and a significant association of gold with pyrite.

The Phu Kham concentrator has been developed as a conventional semi-autogenous grinding (SAG) and ball milling circuit followed by selective rougher flotation, regrinding and cleaner flotation to produce a copper concentrate containing payable gold and silver values. The concentrator flow sheet design offered a capital efficient compromise between high copper recovery bulk sulfide flotation with large cleaning capacity, and lower recovery copper selective rougher flotation to ensure concentrate specification of 24 per cent copper grade could be achieved. This paper will examine and discuss concentrator flow sheet development, including projects implemented since commissioning to improve copper recovery, and future projects designed to maintain and enhance copper in concentrate production with decreasing copper grade and increasing pyrite content of ore feed, and increasing hardness of primary ore.

INTRODUCTION
The Phu Kham operation consists of a copper-gold mine using conventional shovel mining and truck haulage to a 12Mt/a concentrator. The project is owned and operated by Phu Bia Mining Limited. PanAust Ltd based in Brisbane Australia holds a 90 per cent interest in Phu Bia Mining through its wholly owned subsidiary Pan Mekong Exploration Pty Ltd, with the remaining ten per cent held by the Government of Laos PDR.

The Phu Kham Copper-Gold deposit is located in Xaisoumboun province as shown in Figure 1, approximately 120 km north of the Lao capital Vientiane. Access to the mine is approximately four hours by road from Vientiane.

The Phu Kham 12 Mt/a concentrator was designed and built to treat a high pyrite copper-gold skarn ore with significant clay content, as described by Meka and Lane (2010). The plant was commissioned in 2008 for a capital cost of approximately $150 M, placing it in the lowest quartile for capital intensity for copper mineral processing projects.

The installed plant was a compromise between a high recovery but high capital intensity design, and a lower recovery but technically lower risk and low capital intensity design. The selective rougher flotation design was driven by the complex and variable mineralogy and high pyrite content, with over 90 per cent of pyrite required to be rejected in order to produce a final concentrate of over 23 per cent copper.

With increasing depth of the pit since the commencement of operations, the weathering profile of the feed has changed such that the ore became primary dominant in 2010, with chalcopyrite the main copper sulfide mineral. The complex folding and alteration of the ore zones has meant continued mining of supergene and oxidised areas within the pit, with the copper mineralogy remaining diverse and varying from native, oxide, secondary, and primary copper species within short time periods.

The development of the Phu Kham flow sheet was driven by the poor recovery in comparison to other low-grade copper-gold ores, and a need to counter decreasing ore grades from 2013. Major projects implemented up until 2011 included increasing rougher capacity by 25 per cent and increasing first cleaner capacity by 16 per cent, and the installation of a...
Jameson Cell in a cleaner scalper duty. In 2012, the operation is being upgraded to a nominal throughput of 16 Mt/a with installation of a second 13 MW ball mill, a further 33 per cent increase in rougher capacity, 40 per cent increase in second cleaner capacity, and 33 per cent increase in third cleaner capacity.

In 2009, a project to achieve step-change in copper and gold recovery from Phu Kham was initiated. A process development study was completed in 2011, which showed that it was technically feasible to generate a low-grade copper and gold mineral concentrate by bulk sulfide flotation of concentrator tailings suitable for leaching for recovery of copper and gold into high-grade products. During the study, opportunities for increasing copper and gold recovery in the existing concentrator using standard processing methods became apparent, and detailed mineralogical work and metallurgical test work was undertaken to determine the causes of copper and gold loss to tailings. The mineralogical work revealed that up to 60 per cent of copper sulfide mineral lost was in coarse non-sulfide gangue composites, and over 50 per cent of gold loss was in gold-pyrite composites. The work presented an opportunity to recover these composites by less selective rougher flotation, before upgrade and additional recovery for both copper and gold into a 23 per cent copper concentrate by regrinding of rougher concentrate to 20 μm and additional cleaning flotation capacity.

In 2013, the operation will increase total recovery of both copper and gold by six per cent into final concentrate. The recovery increase will be achieved through increasing mass recovery in ‘less selective’ roughing, with additional regrind capacity to reduce the rougher concentrate particle size to 20 μm, before upgrade to final concentrate in an expanded cleaning circuit. A second filter will be installed to dewater the additional concentrate produced.

**GEOLOGY AND MINERALOGY**

The Phu Kham copper-gold deposit is a complex copper-gold mineralised dyke system, which has undergone a number of faulting, folding and alteration events. Mineralisation is present in iron-rich skarns, silica-rich stockwork, and altered disseminated styles. Chalcopyrite and bornite are the dominant primary copper minerals in skarn, stockwork, and disseminated mineralisation. Gangue mineralogy is mainly quartz, mica and pyrite, with significant kaolinite clay and talc-related magnesium silicate content within the weathered zones.

A gold-enriched oxide zone on the Phu Kham orebody was the resource for the heap leach gold mine which was built and operated by Phu Bia Mining during the 2005 to 2010 period. Below the oxide zone, there is a zone of supergene weathering, with copper leached from the oxide zone re-precipitated in contact with pyrite grains as particles and coatings of chalcocite and covellite, with minor enargite and tennantite copper arsenic sulfides. Significant copper enrichment in the oxide and supergene zones is also present as oxide and native copper species.

Skarns are present as replacement of carbonate minerals, with disseminated grains of chalcopyrite and bornite in banded to massive pyrite skarns and veinlets containing pyrite, chalcopyrite and bornite in garnet, magnetite, and hematite-chlorite skarns. Pyrite skarns are common throughout the mineralised system.

Stockwork mineralisation is present as fine fractures in quartz veins. The fractures host pyrite, chalcopyrite and bornite sulfide minerals. Minor chalcopyrite mineralisation is also present in quartz-carbonate veins. Disseminated mineralisation consists of scattered grains of bornite and chalcopyrite in sericite altered host rock.

Gold occurs as small grains associated with pyrite and copper sulfides throughout the mineralised system.

**CIRCUIT DESCRIPTION**

The original 12 Mt/a concentrator design and commissioning in 2008 has been described in detail by Crnkovic et al (2009).

The crushing plant consists of a primary 55 in × 77 in gyratory crusher, with single truck dump point above a pocket designed to hold 200 t capacity equivalent to two 777D haul trucks. Crusher discharge drops to a crushed ore bin of 200 t capacity. The crushed ore bin is emptied by a variable speed apron feeder onto a crushed ore transfer conveyor belt (CV-001). The CV-001 conveyor transfers the ore to an 890 m long overland conveyor CV-002, which moves ore to the coarse ore stockpile with a live capacity of approximately 24 000 t. There is additional dead capacity for storage of up to 300 000 t of ore.

Ore is reclaimed from the crushed ore stockpile by two variable speed apron feeders onto a SAG mill feed conveyor. SAG mill grinding media is added to the ore feed conveyor via a spallage return hopper. Primary grinding is achieved in a dual pinion 13 MW variable speed slip energy recovery/ hyper-synchronous drive 34 ft × 20 ft SAG mill in closed circuit with scats return conveying including a high-lift conveyor to overcome topography constraints. SAG mill discharge is classified using an integral mill trommel, with minus 12 mm product reporting to a 1.85 MW cyclone feed pump. Cyclone feed is classified in a cluster of 18 650 mm diameter cyclones, with cyclone underflow reporting to a dual pinion 13 MW drive 40 ft × 24 ft ball mill. Quicklime is added to the ball mill for flotation pH control to depress pyrite. Ball mill product returns to the cyclone feed pump.

Cyclone overflow reports to a multiple stage feed sampler before a bank of 200 m³ tank cell roughers. The original plant had a single 200 m³ rougher feed conditioning tank.
before eight 200 m³ rougher cells, with the conditioning tank converted to a ninth flotation cell in 2009. In early 2011, a tenth 200 m³ rougher cell was installed and commissioned. Aero 9810 collector is used to recover copper sulfide minerals while maintaining selectivity against pyrite.

Rougher bank tailings passes through a static dual fin pipe sampler before reporting to final tailings mixing box where it is combined with cleaner scavenger tailings. The mixing box discharges to a metallurgical multiple stage sampler and a final tailings sump. The final tailings sump discharges slurry by gravity through two 750 mm diameter tailings lines, which transport tailings approximately 1.5 km to a cross-valley subaqueous tailings storage facility (TSF).

Rougher concentrate is classified in a cluster of six 400 mm diameter cyclones, with cyclone overflow reporting to a Jameson Cell feed hopper. Cyclone underflow reports to an open-circuit M10000 IsaMill™ regrind mill. IsaMill™ discharge reports to the Jameson Cell feed hopper. The 24 downcomer, 6500 mm diameter Jameson Cell was commissioned in March 2011 in a cleaner feed scalping duty, with concentrate passing through a pipe sampler for process control, before reporting to the final concentrate thickener feed sampler, and tailings reporting to the conventional cleaning circuit. A simplified flow diagram of the 12 Mt/a concentrator following installation of the Jameson Cell is shown in Figure 2.

The conventional cleaning circuit consists of three stages, with the first stage in open circuit. The original first stage of cleaning had a 70 m³ conditioning cell, followed by three 70 m³ first cleaners and three 70 m³ cleaner scavengers. In 2009, the conditioning tank was converted into an additional first cleaner cell. Cleaner scavenger tailings pass through a static dual fin sampler before reporting to the final tailings mixing box. First cleaner and cleaner scavenger concentrates report to the second cleaner, which consists of four 20 m³ cells. Second cleaner concentrate advances to the third cleaner of three 20 m³ cells, while second cleaner tailings returns to the first cleaners. Third cleaner concentrate passes through a pipe sampler, before reporting to the final concentrate thickener feed sampler. Third cleaner tailings return to the second cleaners.

Final concentrate (combined Jameson Cell and third cleaner concentrates) is sampled in a multiple stage metallurgical sampler, before gravitating to a 15 m diameter high-rate thickener. Thickener supernatant flows to a thickener overflow process water tank, while thickener surface froth is captured and discharged to a floor sump for return to thickener feed. Thickener underflow at a nominal density of 65 to 70 per cent solids is pumped to a mechanically agitated filter feed tank of approximately 24-hour surge capacity. The thickened concentrate slurry is dewatered using a 64-plate horizontal filter, with filter discharging into a covered storage shed. Concentrate is loaded into 20 t containers for transport by truck to the Sriracha port in Thailand.

Concentrator raw water is harvested from the Nam Mo River before being pumped to a crusher process water tank and mill header tank. The raw water is mainly used for cooling, pump glands, flotation froth wash showers, and fire water. Process water is recovered from the TSF supernatant, and transferred to a process water tank via two transfer stations.

A photograph of the 12 Mt/a concentrator in June 2008 is shown in Figure 3.

**PROCESS IMPROVEMENTS AND FLOW SHEET DEVELOPMENT**

**Flotation cell conversions**

In September 2009, the existing rougher conditioner and cleaner conditioner tanks were retrofitted with flotation mechanisms, thereby increasing the roughing capacity from eight to nine 200 m³ cells, and increasing the cleaner capacity from six to seven 70 m³ cells. The benefits arising from these changes amounted to increased copper recovery in the rougher flotation circuit by 3.5 per cent, and increased copper recovery in the cleaner circuit by 2.5 per cent. The increased residence time in each of the circuits, resulted in higher recovery of slow
floating secondary copper minerals, and composite particles particularly while transitional ore types were dominant at this time. The improved recovery performance was validated from using the database of daily rougher tail and cleaner scavenger tail reflation tests.

**FloatForce™ mechanisms**

An investigation into using Outotec FloatForce™ rotor-stator mechanisms in the rougher flotation circuit commenced in 2009, starting with the first rougher cell 1B. The design of the new mechanism was to deliver improved recovery from increased mixing efficiency, by not allowing any air in the central mixing area of the impeller thereby improving mixing efficiency without affecting slurry pumping. The installation of the FloatForce™ mechanism was simple, and was commissioned without any issues. The conclusions, drawn from extensive survey data around rougher cell 1B (before and after installation), demonstrated an improvement in rougher cell 1B copper recovery of 0.2 per cent. The basis of the survey data, the total predicted rougher copper recovery increase for the nine rougher cells with FloatForce™ mechanisms was 0.3 per cent. The installation period for the remaining eight mechanisms was completed in June 2011.

**Tenth rougher cell installation**

The ongoing rougher tail re-flotation tests continued to highlight that there was a further copper recovery benefit of approximately 0.6 per cent to be gained with the addition of another 200 m³ tank cell, which would increase rougher residence time by 2.4 minutes at maximum design throughput of 1750 t/h. In late January 2011, the tenth 200 m³ rougher tank cell was installed at the head of the rougher circuit. At the time that the tenth rougher cell was installed, the flotation feed rate was increased by about four per cent, which effectively reduced the overall rougher residence time increase from an expected 11 per cent to seven per cent. The overall surveyed copper recovery improvement was between 0.4 per cent and 0.6 per cent, depending on throughput rate, which met the project criteria.

**Cleaner circuit debottlenecking**

One of the major limitations in the original plant design had been a lack of cleaning capacity, particularly with respect to the second cleaner bank of four 20 m³ cells. The limitation was that of carrying capacity, rather than residence time. The cleaner circuit performance would generally deteriorate when the cleaner feed copper metal units exceeded 9.1 t/h copper, limiting copper metal production to a sustainable maximum of approximately 8.5 t/h.

To further investigate the cleaning circuit capacity, a cleaner circuit optimisation study was completed in February 2010. From plant data, a mineral based floatability component model was developed which allowed different cleaner circuit configurations to be simulated. The option which gave the optimum copper grade and recovery result was to install additional cleaning capacity ahead of the existing cleaner circuit, so effectively cleaner feed scalping. Different flotation cell technologies were considered for this application, with the Xstrata Jameson Cell meeting design criteria. The simulations indicated that a 0.6 per cent improvement in cleaner recovery could be achieved. The Jameson Cell was chosen because of low installed cost, simulated performance, low performance risk, moderate installation risk, and low production continuity risk during installation.

The circuit simulations, including the Jameson Cell as a cleaner feed scalper, indicated substantial recovery improvement over the existing circuit at circuit feed rates greater than 150 t/h. This was due to the elimination of the carrying capacity limitation as shown in Figure 4.

On this basis it was decided to proceed with this design, the Jameson Cell in a cleaner feed scalping simulation recovering approximately 60 per cent of the copper present in the cleaner circuit feed across a cleaner circuit feed rate range of 100 t/h to 300 t/h. The Jameson Cell concentrate grade from the simulation was 27 per cent copper, against a target of 25 per cent copper. The simulations showed that substantial unloading of the remainder of the cleaner circuit would occur. The result of this was that the third cleaner concentrate grade was low at less than 22 per cent copper; however, the net effect was to produce an overall circuit final concentrate grade of 24 per cent copper.

Although there was a small cleaner recovery improvement shown from the simulations performed using a cleaner scalper, the real benefit is in maintaining cleaner recovery when the cleaner feed rate is greater than 150 t/h.

The cleaner scalper cell was commissioned in March 2011. Commissioning was carried out over a period of one week, and no significant problems were encountered. The performance evaluation of the Jameson Cell was remarkably consistent with the expected performance from the equipment vendor, the simulation data, and from Phu Kham Metallurgical Laboratory flotation tests simulating performance of the Jameson Cell prior to commissioning. From surveys carried out in February 2012, with the Jameson Cell online and offline, the benefit of having the Jameson Cell in circuit was determined to be 0.8 per cent increase in copper recovery.

In terms of overall cleaner circuit debottlenecking, the objectives have been achieved. The cleaner circuit with cleaner feed scalping capacity is 10.1 t/h copper metal at 24 per cent concentrate grade, which is the current limit of the concentrate filtration circuit, for a total 16 per cent copper metal production increase.

**Phu Kham Upgrade Project**

The Phu Kham Upgrade Project commenced in March 2010 with a study to develop designs to ensure copper in concentrate production is maintained over 60 kt/a after 2013 when plant copper feed grade is expected to decrease. In order to maintain copper metal production, plant nominal design throughput will increase from 12 to 16 Mt/a (20001/h). Maximum instantaneous design throughput for the upgraded plant is 2250 t/h. The plant upgrade concept was not original, and had been studied in 2008 as part of a copper production expansion project. Key aspects of the upgrade designs for Phu Kham were the limitation of available space for additional
equipment, as the original 12 Mt/a plant design had not specifically made allowance for any expandability.

The initial phase of the upgrade included a plant debottlenecking study, which consisted of analysis of the actual plant performance and capacity data from 2008 to March 2010 against the original plant process design criteria. The purpose of the bottleneck study was to determine aspects of the original plant that either were, or would become bottlenecks with the 25 per cent increase in mill throughput. The key findings from the plant bottleneck study are shown in Figure 5, which shows that rougher copper recovery was 16 per cent below design, cleaner copper recovery was seven per cent below design, and mill throughput was three per cent below design at 12 Mt/a. The mill throughput variance was a function of rougher copper recovery and cleaning circuit capacity rather than limitations in the grinding circuit.

The crushing and concentrate dewatering plant capacities were also considered during the upgrade bottleneck study. The bottleneck study indicated that additional crushing capacity would be required with capability for handling wet and sticky ore, which is a common feature of the transition zones of the orebody. A mineral sizer in parallel to the existing crushing plant, with product reporting directly to the coarse ore stockpile was included in the upgrade designs. Although the concentrate thickener and filter performance had not indicated that future concentrate production rate would exceed capacity, limited data was available to confirm the capacity against upgraded plant design criteria. Test work was conducted to determine settling rates and filtration rates for concentrate during the upgrade project to obtain the required data.

The basis of design for the grinding circuit upgrade has been previously described by Hadaway and Bennett (2011). Two options for increasing grinding circuit throughput after the SAG mill to a nominal design of 16 Mt/a primary grind at 80 per cent passing 106 μm or 75 μm were reviewed. The first option was based on the original 2008 plant upgrade design incorporating an additional 6.5 MW single pinion ball mill, and the second for another 13 MW ball mill.

Data from JKTech grinding circuit modelling in 2009 was extrapolated using the Phu Kham mine schedule to
determine throughput estimates at 106 μm and 75 μm for the two mill options. Pebble crushing was not considered in the evaluation. The 6.5 MW mill had previously been shown to be able to increase throughput at the 106 μm primary grind to above the 16 Mt/a nominal design, however, was unable to meet the 16 Mt/a target at a significantly finer primary grind. The 13 MW mill is able to achieve above 18 Mt/a for the 106 μm primary grind, and will achieve above nominal design throughput at a 75 μm primary grind.

The effect of primary grind on flotation recovery was reviewed based on feasibility study work from bench scale batch tests in 2005. The study work indicated that the major primary ore sources, in particular stockwork primary, were relatively insensitive to primary grind size. Plant operations mineralogy data from 2008 to 2011 monthly composites indicates that minor sensitivity exists, with increases of over five per cent in copper sulfide liberation with a primary grind size decrease from 80 per cent passing 106 μm to 75 μm.

An economic analysis was conducted based on differences in capital and operating costs for the two options at 16 Mt/a throughput and 106 μm and 75 μm primary grind. The increase in operating cost for finer grinding versus revenue benefits in copper recovery showed that above $2.50/lb copper price the finer primary grind increased gross margin. Capital cost per installed megawatt was 26 per cent less for the 13 MW mill option, and the capital payback period for the 13 MW option was significantly shorter.

A risk assessment was conducted for the 6.5 MW option and the 13 MW option. The risk of the 13 MW option was considerably lower than for the 6.5 MW option, mainly due to the operating flexibility for periods of low-grade ore and ore types with higher sensitivity of recovery to primary grind. The 6.5 MW option was not able to take advantage of economies of scale gained by increased throughput, or the estimated one per cent increase in copper recovery at the finer grinds, and would not reach the nominal 16 Mt/a throughput at 106 μm primary grind after 2014. The throughput at 106 μm and 75 μm primary grinds for the two mill options is shown in Figure 6.

Based on the results of the risk assessment, the recommendation for installation of an additional 13 MW ball was accepted. Procurement of a second dual pinion 13 MW drive 40 ft × 24 ft ball mill commenced in November 2010.

The dominant cause of the 15 per cent rougher copper recovery shortfall shown in Figure 5 was a combination of lower than design rougher residence time due to five per cent lower rougher feed density, and a cleaner circuit capacity constraint which limited rougher mass recovery. The dominance of transition ores with significant slow-floating secondary copper mineral content milled during the March 2009 to February 2010 period and the under-representation of these ore types in the feasibility study test work provide explanation for some of the copper recovery shortfall in cleaning stages against design. The debottlenecking study was developed for the flotation circuit to determine increased capacity requirement at the design 16 Mt/a upgrade throughput.

Rougher flotation feed density design for the 12 Mt/a plant was 35 per cent solids. Actual operation rougher feed density averaged 30 per cent solids due to the higher slurry viscosity from kaolinite clay content not quantified during the feasibility study. An extra 200 m³ rougher cell was required to achieve the same residence time as at 35 per cent solids, which was achieved by conversion of the rougher conditioning tank to a cell in 2009. The reduced residence time from operating at the lower rougher density at design tonnage throughput resulted in a three per cent decrease in copper recovery, based upon plant residence time – recovery data from July 2009 to February 2010. The upgrade design therefore allowed for reduced rougher feed pulp density, and a residence time calculation confirmed that a 33 per cent increase in rougher capacity was required for the 25 per cent increase in mill throughput at 16 Mt/a, which would also provide an additional one per cent copper recovery. A total of five 200 m³ rougher cells in addition to the existing ten cells were included in the design, for a total of 15 cells.

The cleaning circuit was not expected to require significant expansion as a result of the 16 Mt/a upgrade, as the lower-grade mill feed would result in equivalent concentrate production to the 12 Mt/a design throughput rate. The upgrade design for the cleaner flotation circuit also included the Jameson Cell cleaner scalper although this had not been installed at this time. However, cleaner circuit mass balance simulation data including the Jameson Cell indicated that 40 per cent increase in the existing second cleaner residence time and lip length was required at 16 Mt/a. To gain this increase in second cleaner capacity, the existing three 20 m³...
third cleaner cells have been combined with the four 20 m³ second cleaner cells, and four new 20 m³ third cleaner cells added to for the upgrade. The simulation data for the cleaners also demonstrated that upgrade of the first cleaner capacity was not warranted, as the fine low-grade middlings recovered in the final cells were not able to be upgraded to near final concentrate specification.

Prior to commencement of the detailed design phase of the upgrade study, maximum sustainable production rate (MSPR) analyses were undertaken for the crushing and concentrate dewatering plants to determine whether capacity expansion was required for these areas of plant based on 16 Mt/a production schedules. The MSPR was defined as the best consecutive five days of performance, normalised using plant-specific industry standards for annual availability to allow for major scheduled maintenance. The MSPR for the crushing plant also included seasonal variation due to the tropical environment and the wet season impacts on crusher productivity.

The primary conclusions from the performance review of the Phu Kham crushing plant were that it had demonstrated the target upgrade production rate of 16 Mt/a over the June 2010 period, and approximately one third of total crushing plant downtime had been caused by events up and downstream of the crushing plant while the plant was available to crush. The low crusher utilisation of 64 per cent was equivalent to over 2.4 Mt/a of crushing capacity at the target throughput of 2400 t/h and target utilisation of 75 per cent of total time. With increases in haul fleet numbers for the upgrade, improvements in run-of-mine stockpile inventory, and a stand-by loader available when there were delays in truck presentation to the crusher, the MSPR demonstrated that increasing crushing capacity was not required.

The design specifications for the 64-plate and frame filter were for a filtration rate of 225 kg/m²/h, with an annual design production rate of 311 000 t of concentrate. Actual filter plant operating data was analysed to check filter performance against the design capacity. MSPR for the filter was determined to be 18 per cent above the life-of-mine maximum concentrate production schedule, leading to deferral of capital expenditure for the filtration plant. The main reasons for the higher than design performance were; optimisation of filter cycle settings following an improvement program including operations, maintenance, and vendor support input, and change to filter cloth media type.

Following the review of the upgrade design, engineering and procurement services commenced for the Phu Kham 16 Mt/a Upgrade project in January 2011, with commissioning scheduled for the third quarter of 2012. A simplified flow diagram for the upgraded plant is shown in Figure 7, with new equipment highlighted in mauve.

**Increased recovery project**

The Phu Kham feasibility studies between 2004 and 2006 identified two options for flotation processing of Phu Kham ore. The first option involved bulk flotation of the rougher feed targeting a 25 per cent mass recovery into rougher concentrate using non-selective amyl xanthate sulfdide mineral collector. The rougher concentrate was then reground to 80 per cent passing 38 μm and subjected to cleaner flotation at a pH of 12 for pyrite depression. This process produced high copper recovery results, however, there was difficulty achieving final concentrate grade of greater than 22 per cent copper across all ore types, particularly transition chalcocite-covellite secondary copper mineral dominant ores. A rougher feed photomicrograph (Shouldice and Mehrfert, 2009) is shown in Figure 8 with chalcocite-covellite intergrowth with pyrite and rimming of pyrite. There was also indication of copper activation of pyrite from soluble copper species in weathered and transition ores.
The second process option involved selective flotation in roughing at pH 11 - 12, using a copper sulfide selective collector. The rougher concentrates were again reground to 80 per cent passing 38 μm and lime to pH 12 and sodium cyanide was added to the cleaning stages to depress pyrite. The selective flotation option consistently achieved over 22 per cent copper final concentrate grade; however, ultimate copper recoveries to final concentrate were lower than the bulk flotation option.

The design of the original 12 Mt/a Phu Kham concentrator was a compromise between the two process options, with partially selective roughing being applied to minimise pyrite gangue recovery into cleaner flotation feed. Sodium cyanide addition to the cleaners was included in the design, however, has never been used with concentrate grade over 22 per cent copper consistently achieved since commissioning. This partially selective flotation process had significant capital cost advantages over bulk flotation at a time when the long-term copper price estimate was much lower than 2012 prices, due to the lower rougher concentrate regind and cleaning capacity required, and provided the best cost-benefit process alternative while reducing risk of being unable to achieve concentrate specifications using bulk rougher flotation.

Minimal work was performed during the feasibility studies to test the sensitivity of final copper grade and recovery on rougher concentrate regind product particle size. Grind size analysis was limited to two mineralogical examinations which concluded that reasonable copper and gold recoveries to rougher concentrate would result from a primary grind of 80 per cent passing 106 μm, and that a rougher concentrate regind to less than 45 μm was required to achieve an acceptable final concentrate grade. Mineral and liberation analysis showed that associations between copper sulfide minerals and pyrite did not indicate complex or fine intergrowths that would adversely impact on the metallurgy.

The Phu Kham increased recovery project (IRP) commenced in 2009, as part of a concept study to develop a process to increase copper and gold recovery from Phu Kham ore. Since commencement of operations in 2008, copper recovery had increased with increasing proportion of chalcopyrite-dominant primary ores replacing the chalcocite-covellite secondary copper mineral dominant high clay and talc transition ores. The increasing plant throughput and poor primary liberation with increasing pyrite content has caused copper and gold recovery to ‘flat-line’ as shown in Figure 9.

Bulk sulfide flotation bench tests in 2009 using isopropyl xanthate collector on Phu Kham rougher tailings indicated that up to ten per cent additional copper recovery and 70 per cent additional gold recovery could be achieved into a scavenger concentrate of approximately 0.8 per cent copper. The initial test program was designed to determine whether a low-grade copper-gold concentrate suitable for downstream hydrometallurgical processing to saleable products could be recovered from the concentrator tailings.

The preliminary test program showed that a bulk sulfide concentrate from plant tailings flotation could be upgraded to over ten per cent copper concentrate grade, depending upon copper sulfide mineral liberation, using a roughing, regrind, and two stage cleaning process similar to the Phu Kham concentrator process. Figure 10 shows the copper grade – recovery relationships with varying rougher concentrate regind power input of 10 kWh/t, 20 kWh/t and 40 kWh/t.

The test results presented in Figure 10 clearly demonstrated that finer regrinding of rougher concentrates from plant tailings flotation would improve both copper grade and recovery. Further flotation tests on plant final tailings samples using roughing at pH 9 with amyl xanthate, followed by regrinding of concentrate at 10 kWh/t power input, and two stages of cleaning consistently produced a low-grade flotation concentrate of approximately three per cent copper and 2 g/t gold. Average copper recovery was 69.6 per cent and average gold recovery was 54.1 per cent from 24 flotation tests as shown in Table 1.
A digital photomicrograph of flotation tailings is shown in Figure 12 (Shouldice and Ma, 2009). The wide size range of the chalcopyrite particles in non-sulfide gangue is evident.

Copper sulfide mineral grain size data for rougher tailings is presented in Table 2. The data shows that under selective rougher flotation conditions, recovery of coarse low-quality binary copper sulfide and gangue composite particles is poor due to the fine copper sulfide grain size.

Gold recovery by flotation at Phu Kham has been poor since commissioning, averaging approximately 40 per cent to final copper concentrate product. Prediction of gold recovery had also been demonstrated to be inaccurate during plant operation, due to a lack of understanding of the key mineralogical characteristics of gold occurrence and the variability of gold occurrence across different Phu Kham mineral assemblages. As part of the increased recovery project, the mineralogical reasons for gold loss into Phu Kham tailings were examined to ensure potential opportunities to increase gold recovery into final concentrate were included in IRP process design.

Diagnostic leach tests were conducted on Phu Kham plant tailings samples. Results of the diagnostic leaching are presented in Table 3.

The diagnostic leaching results in Table 3 demonstrated that over 60 per cent of the gold in tailings was available for cyanide leaching, either as free gold or partially liberated gold. This was supported by the Albion gold leach test work on acid Albion copper leach residues, which demonstrated that extraction of gold to solution did not significantly increase with increasing sulfide sulfur oxidation, as shown in Figure 13.

Laser ablation testing was conducted on the rougher tailings to determine the proportion of gold locked with pyrite and other sulfide mineral species. The laser ablation test results were combined with the results from the diagnostic leaching to provide a total gold association for the rougher tailings as presented in Table 4.

Diagnostic leaching provided a measure of the unlocked (cyanide soluble) and locked gold deportment, but the total unlocked gold could not be split between fully liberated gold particles, and partially liberated (exposed in composites) gold particles. Therefore the diagnostic leach measure of 57 per cent cyanide soluble gold in Table 4 could not provide definitive information for the cause of gold loss to tailings.
Final Phu Kham copper concentrate monthly plant composites were analysed by automated digital image scanning (ADIS) at G&T Metallurgical Services to determine the characteristics of gold and gold composite particles recovered in flotation (Shouldice and Johnston, 2012). Laboratory work on rougher tailings sample from the IRP laboratory test work was also conducted to produce a gold-rich concentrate by gravity concentration suitable for ADIS and photomicrograph analysis.

The ADIS work on the final concentrate showed an average gold particle size of 7.9 μm. The class and mass distribution summary and area as a percentage of the observed gold particles are presented in Table 5.

Binary particles of gold locked with pyrite in concentrate were of particular interest. Only eight per cent of the observed particles were gold-pyrite binary composite particles; however, 71 per cent of the total mass of observed gold was in these particles. The average surface area of the gold in the gold-pyrite binary particles was 87 per cent of the total particle surface area. An interpretation of this data indicated that for a gold-pyrite particle to float into concentrate, the gold:pyrite surface area ratio must be sufficiently large to overcome the depression of the attached pyrite particle under high pH flotation cleaning conditions. The liberated gold recovered was typically fine, with an average particle size of 7 μm.

Photomicrographs of the gold showing some typical particles in concentrate are presented in Figure 14. The ADIS work found that the average gold particle size was 19 μm in rougher tailing, approximately 13 times larger than the average gold particle in final concentrate. No binary particles were observed, only liberated gold and gold-chalcopyrite-pyrite-gangue multiphase particles as summarised by mass distribution in Table 6.

The liberated gold particles in tailings had a mean size of 35 μm, with the data indicating that liberated gold particles above 20 μm in size are unlikely to be recovered to final flotation concentrate, with a 13 μm particle the largest observed in the concentrate. Cyanide leaching would have extracted 69 per cent of the gold particles observed.
The significant characteristic of the multiphase particles observed in tailings was that greater than 95 per cent of the total mass of the particles were gangue mass. The gold in the multiphase particles had an average diameter of 9 μm, while the pyrite gangue had an average diameter of 121 μm.

Photomicrographs of the gold showing some typical particles in rougher tailings are presented in Figure 15.

Based on the concentrate and rougher tailings ADIS data, a summary of the estimated recoveries of the main Phu Kham ore gold association classes as presented in Table 7.

Two mineral process options for increasing recovery of the low quality copper sulfide–non-sulfide gangue binary particles were tested; mainstream inert grinding, and bulk sulfide flotation. The mainstream inert grinding process is classification of rougher tailings for recovery of the plus 53 μm fraction, regrinding to approximately 80 per cent passing 53 μm, and scavenger flotation. Bench scale test work on the mainstream inert grinding showed that an additional seven per cent recovery of copper and four per cent recovery of gold to final concentrate was achievable, however, the mass recovery of 45 per cent of the plant tailings into the plus 53 μm fraction required approximately 16 MW of additional installed power for regrinding, most of which would be wasted regrinding liberated gangue.

The bulk sulfide flotation process laboratory test work followed the original Phu Kham prefeasibility study process.
of maximising copper and gold recovery into a low-grade rougher concentrate using xanthate collector at natural pH. In developing the design for a bulk sulfide rougher, regrind and selective cleaning flotation process, three key parameters were required to be identified; rougher mass recovery to achieve maximum copper recovery across all ore types, rougher concentrate regrind product size, and cleaner capacity.

Rougher mass recovery design was determined by daily rougher flotation testing of plant rougher feed over a three month period in 2011, which allowed variability testing across all major ore types. The rougher tests were conducted using Aero 9810 copper selective collector but at up to 80 per cent liberation for both copper and gold grade and recovery into flotation concentrates.

The results for rougher mass recovery of 25 per cent. Rougher concentrate regrind size optimisation work commenced in 2011 to support the IRP process design. The benefits for copper recovery of finer grinding of concentrates from bulk flotation of Phu Kham tailings had previously been observed as shown in Figure 10. Samples of plant feed were tested at bench scale using roughing, rougher concentrate regrind at increasing power input, and three stages of cleaning to produce the copper and gold grade – recovery response curves in Figure 18. The increase in rougher concentrate regrind power input shows the benefit of the increased liberation for both copper and gold grade and recovery into flotation concentrates.

With the variable mineralogy of Phu Kham ore, the monthly composite mineralogy and laboratory scale test work was used to determine the rougher concentrate regrind size required to achieve maximum copper sulfide liberation in cleaner feed, with a target of 80 per cent copper sulfide liberation considered to be required to maximise recovery and maintain final concentrate specification. Scan data on two samples representing different Phu Kham ore types in Figure 19 showed that maximum copper sulfide liberation was typically achieved at 20 μm particle size, although 80 per cent liberation was not necessarily achieved for all ore types.

Power input to achieve 20 μm regrind product size was calculated from laboratory regrind signature plot data, and daily surveys of the existing M10000 IsaMill™ to be 18 - 25 kWh/t, with an additional 3 MW M10000 IsaMill™ included in the IRP design to provide a total of 6 MW power input at the maximum rougher concentrate mass recovery of 25 per cent.

**First cleaner design**

The increase in rougher concentrate mass recovery to 25 per cent of rougher feed, and the reduction in cleaner feed particle size to 20 μm required a corresponding increase in first cleaner flotation capacity. The 12 Mt/a plant had 25 minutes...
total first cleaner residence time for a 38 μm cleaner feed size, and test work was designed to determine whether this residence time needed to increase at the 20 μm cleaner feed size due to potentially reduced kinetics of the finer particles. Bench scale flotation cleaning tests and cleaner circuit model simulations indicated that flotation kinetics remained similar for copper sulfide minerals due to improved liberation, and that no increase in residence time was required for first cleaning until regrind product size was below 10 μm. Figure 20 shows the plant cleaner circuit chalcopyrite recovery by particle class and size. The recovery of liberated chalcopyrite remains high with decreasing particle size, which confirmed that the increased chalcopyrite liberation from gangue at a 20 μm regrind size will improve copper recovery.

The IRP design includes 100 per cent increase in first cleaner capacity to allow for the increased rougher mass recovery, and lower slurry density to provide improved dilution cleaning to minimise fine gangue particle entrainment.

To validate the extensive test work and mineralogy data results, four IRP process plant trials were conducted at Phu Kham between January 2012 and April 2012. The trial method used was to decrease SAG mill throughput by 50 per cent to avoid overloading the cleaning circuit, reduce rougher cell residence time to the equivalent post-16 Mt/a time of 30 minutes by reducing levels and air in three of the ten cells, and using increased collector and frother addition to increase rougher mass recovery to 25 per cent. The IsaMill™ power draw was increased to 2.8 MW to target a 20 μm regrind product size. Immediately prior to the trial periods, baseline plant surveys were undertaken to obtain comparison data for the same ore type. The key results of the four plant trials are summarised in Table 8.

The first plant trial period was over four hours to ensure critical test criteria could be achieved, in particular rougher mass recovery target of 25 per cent. Although this was a preliminary trial, survey results for the IRP process were
positive compared to baseline survey results, with eight per cent overall copper recovery and 19 per cent overall gold recovery achieved into a two per cent higher copper grade final concentrate. The second plant trial was conducted over a full 12-hour shift period, with improvement in both overall copper and gold recovery of 11 per cent at a 0.6 per cent copper grade improvement in final concentrate.

Two short four-hour variability trails were conducted in March and April 2012 on primary ore with a good flotation response and high pyrite ore with a poor flotation response. Copper recovery improvement was consistent at over five per cent for both ores, and gold recovery improvement was 13 per cent and 20 per cent into final concentrate. The short duration of the trials did not allow time for optimisation of cleaner circuit performance. The plant trial results provided significant confidence in the IRP concept and basis of design, which was finalised in May 2012 prior to commencing detailed design. The economic evaluation at six per cent increase in copper and gold recovery after the initial plant trials provided a compelling investment case, resulting in PanAust Ltd approval in February 2012 for the development of the IRP at Phu Kham.

The simplified process flow diagram for the IRP is shown in Figure 21, with upgrade of the existing rougher concentrate pumping and classification, a second 3 MW M10000 IsaMill™ regrind mill, seven additional 70 m³ first cleaner flotation cells, and upgraded pumping capacity in the cleaner flotation circuit. The increased copper recovery and resultant production of approximately 25 000 t/a additional mineral concentrate at a finer particle size was determined to exceed the existing 64-plate and frame filter MSPR, with an additional 40-plate and frame filter included in the design to meet IRP and future concentrate production output.

CONCLUSIONS

The installed 2008 Phu Kham 12 Mt/a concentrator was a compromise between a high-recovery but high-capital intensity design, and a lower-recovery but technically
low-risk and low-capital intensity design suitable for the prevailing low copper price market. The chosen selective rougher flotation design was driven by the complex and highly variable mineralogy particularly in the transition ore zones, and high pyrite content. Over 90 per cent of pyrite is required to be rejected in order to produce a final concentrate of over 23 per cent copper.

The optimisation of the Phu Kham flow sheet has been driven by low copper and gold recoveries and the requirement to maximise copper production by increasing plant throughput. Flotation circuit improvements and debottlenecking projects including additional roughing capacity and cleaner feed scalping have increased copper metal production by 16 per cent since 2009, while maintaining copper recovery at maximum design throughput of 14 Mt/a.

The plant optimisation and development program and major plant design at Phu Kham has been supported by extensive mineralogy, mineral association, and mineral liberation data from plant monthly composites and bench scale test products. The collection and analysis of this data has revealed reasons for copper and gold losses and mineral deportment, and significant opportunities for increasing recovery of copper and gold have been identified.

In 2012, the operation will be upgraded and debottlenecked to process a nominal throughput of 16 Mt/a with installation of a second 13 MW ball mill, a 33 per cent increase in rougher flotation capacity, a 40 per cent increase in second cleaner capacity, and 33 per cent increase in third cleaner capacity.

In 2013, the operation will increase total recovery of both copper and gold by six per cent into final concentrate by increasing mass recovery into rougher concentrate, and debottlenecking of rougher concentrate regrind, cleaning, and final concentrate dewatering plants.

**ACKNOWLEDGEMENTS**

The authors acknowledge the permission of PanAust Ltd and Phu Bia Mining Ltd to publish this paper. The authors would like to particularly thank company metallurgists for their

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**TABLE 8**

Phu Kham increased recovery project plant trial results.

<table>
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<tbody>
<tr>
<td></td>
<td>Baseline</td>
<td>Increased recovery project</td>
<td>Difference</td>
<td>Baseline</td>
<td>Increased recovery project</td>
<td>Difference</td>
</tr>
<tr>
<td>Rougher recovery (%)</td>
<td>83</td>
<td>92</td>
<td>+9</td>
<td>76</td>
<td>88</td>
<td>+12</td>
</tr>
<tr>
<td>Overall copper recovery (%)</td>
<td>71</td>
<td>79</td>
<td>+8</td>
<td>68</td>
<td>79</td>
<td>+11</td>
</tr>
<tr>
<td>Overall gold recovery (%)</td>
<td>48</td>
<td>66</td>
<td>+19</td>
<td>45</td>
<td>56</td>
<td>+11</td>
</tr>
<tr>
<td>Concentrate grade (%)</td>
<td>22.1</td>
<td>24.1</td>
<td>+2.0</td>
<td>24.2</td>
<td>24.8</td>
<td>+0.6</td>
</tr>
<tr>
<td></td>
<td>Baseline</td>
<td>Increased recovery project</td>
<td>Difference</td>
<td>Baseline</td>
<td>Increased recovery project</td>
<td>Difference</td>
</tr>
<tr>
<td>Rougher recovery (%)</td>
<td>87</td>
<td>93</td>
<td>+6</td>
<td>81</td>
<td>87</td>
<td>+6</td>
</tr>
<tr>
<td>Overall copper recovery (%)</td>
<td>81</td>
<td>87</td>
<td>+6</td>
<td>75</td>
<td>80</td>
<td>+5</td>
</tr>
<tr>
<td>Overall gold recovery (%)</td>
<td>40</td>
<td>53</td>
<td>+13</td>
<td>36</td>
<td>56</td>
<td>+20</td>
</tr>
<tr>
<td>Concentrate grade (%)</td>
<td>25.7</td>
<td>23.8</td>
<td>-2.0</td>
<td>25.5</td>
<td>21.7</td>
<td>-3.8</td>
</tr>
</tbody>
</table>

**FIG 20** - Phu Kham cleaner circuit chalcopyrite recovery by particle class and size.
work and technical assistance in the numerous projects to increase throughput and recovery at Phu Kham. The authors also gratefully acknowledge the support, technical reality checking, and encouragement of Mr Peter Munro and Dr Bill Johnson of Mineralurgy Pty Ltd, Dr Greg Harbort of AMEC Minproc for his flotation modelling work, and the work of John Glen and Tony Button at Burnie Research Laboratory, and Helen Johnston at G&T Metallurgical Services.

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From base metals and back – IsaMills and their advantages in African base metal operations

H. de Waal, K. Barns, and J. Monama

Xstrata

IsaMill™ technology was developed from Netzsch Feinmahltechnik GmbH stirred milling technology in the early 1990s to bring about a step change in grinding efficiency that was required to make Xstrata’s fine-grained lead/zinc orebodies economic to process. From small-scale machines suited to ultrafine grinding, the IsaMill™ has developed into technology that is able to treat much larger tonnages, in coarser applications, while still achieving high energy efficiency, suited for coarser more standard regrind and mainstream grinding applications.

The unique design of the IsaMill™, combining high power intensity and effective internal classification, achieves high energy efficiency and tight product distribution which can be effectively scaled from laboratory scale to full-sized models. The use of fine ceramic media also leads to significant benefits in downstream flotation and leaching operations.

These benefits are key drivers for the adoption of the technology into processing a diverse range of minerals worldwide, and offer major opportunities for power reduction and improved metallurgy for the African base metal operations.

**Keywords:** IsaMill, regrind, energy efficiency, inert grinding.

**Introduction**

The development of the IsaMill™, by MIM (now GlencoreXstrata) and Netzsch Feinmahltechnik GmbH, was initiated to enable the development of the fine-grained ore deposits at Mt Isa and McArthur River in Northern Australia. To liberate the valuable minerals and so produce a saleable concentrate this ultrafine-grained ore needed to be ground to a $P_{80}$ of 7 μm.

The conventional regrind technologies in the early 1990s, ball and tower mills, were unable to grind to these sizes economically. Additionally, the high steel media consumption and associated changes in pulp and surface chemistry were extremely detrimental to flotation metallurgy.

IsaMill™ technology was further transferred into mainstream inert grinding (MIG) and high throughput regrind applications with the development of the M10 000, 3 MW IsaMill™ in conjunction with Anglo Platinum at their Western Limb Tailings Retreatment facility near Rustenburg, South Africa. The successful commissioning of this mill in the Bushveld Complex resulted in rapid adoption of the technology by the platinum industry.

The development of the 3 MW IsaMill™, coupled with commercialization of the technology in the early 2000s, has accelerated the uptake of IsaMill™ technology. To date there are over 120 IsaMills™ installed around the world in a range of applications and industries, particularly in copper and lead zinc concentrate regrind duties.

**IsaMill principles of operation**

** Grinding mechanism**

The IsaMill™, as shown in Figure 1, is a horizontally stirred mill consisting of a series of 7-8 rotating grinding discs mounted on a cantilevered shaft THAT is driven through a motor and gearbox. The discs operate at tip speeds of 21-23 m/s, resulting in energy intensities of up to 300 kW/m$.^3$
Figure 1. IsaMill™ layout

Figure 2 shows the principle of operation. The mill is filled with suitable grinding media, and the area between each disc is essentially an individual grinding chamber. As a result, the mill effectively comprises eight grinding chambers in series. The media is set in motion by the action of the grinding discs, which accelerate the media radially towards the shell. Between the discs, where the media is not as subject to the high outwards acceleration of the disc face, the media is forced back in towards the shaft – creating a circulation of media between each set of discs. Minerals are ground as a result of the agitated media, the predominant mechanism being attrition grinding.

Figure 2. Operation of the IsaMill™
From base metals and back – IsaMills and their advantages in African base metal operations

The eight chambers in series mean that short-circuiting of the mill feed to the discharge is virtually impossible. There is a very high probability of media-particle collision as a result of the high energy intensity and the eight chambers in series.

Energy efficiency

The energy efficiency advantages of IsaMill™ technology over ball mills and other stirred mills are shown in Figures 3 and 4 respectively. Figure 3 compares the energy required to grind a gold-pyrite ore in a ball mill with 9 mm steel balls with an IsaMill™ with 2 mm media. The IsaMill™ is much more efficient below about 30 μm – to grind this ore to 15 μm would take 28 kWh/t in the IsaMill™, but 90 kWh/t in a ball mill. Traditionally, this has been attributed to the difference between attrition grinding and impact grinding. However, by far the most important factor is media size, as shown by Figure 4. Figure 4 shows the breakage rate in tower mills dropping dramatically with decreasing particle size - the breakage rate for 20 μm particles is ten times lower than the rate for 40 μm particles. Even though the tower mill mechanism is full attrition grinding, in practice it is constrained to using relatively coarse media, 9 mm balls in this case. In contrast, the IsaMill™ (Netzsch mill in Figure 4) can operate with much finer media and a much higher intensity of power input, meaning the peak breakage rate occurs at 20 μm, and does not drop as quickly as the tower mill below that size.

Figure 3. Grinding energy vs product size for gold-pyrite ore (Pease et al., 2004)
Particle size distribution

Due to the horizontal layout of the IsaMill™ and the multiple effective grinding chambers, the IsaMill™ is capable of producing a very steep product size distribution when compared to alternative technology such as ball and tower mills. As shown in Figure 5 the preferential size reduction of the coarser sized material over the finer material results in progressively steeper product size distributions at increasing specific energies.
The IsaMill™ preferentially targets the coarser size fractions in the slurry stream, producing a sharper particle size distribution curve. This is often a deciding factor in designing grinding and flotation circuits, as gangue entrainment in the flotation concentrate and the ultra-fine size fractions can influence final concentrate grades.

**Benefits of inert grinding**

After the initial comminution stages of crushing and/or SAG/AG milling, historical practice employed steel-charged ball or tower mills for the subsequent grinding stages. The impact of grinding using steel media often offset any benefits gained by improved liberation, particularly as the target size decreases below 25 μm. Grinding in a steel environment results in the precipitation of metal and iron hydroxides on the surface of ground particles. These conditions affect floatability and flotation selectivity, and lead to higher reagent consumptions to overcome the surface coatings and regain recovery (Trahar, 1984; Pease *et al*., 2006).

The benefits of inert grinding at several locations have been comprehensively reported (Pease *et al*., 2006, 2005, 2004; Young *et al*., 1997; Grano *et al*., 1994). While the negative impacts of steel grinding will be greatest at fine sizings due to the large surface areas and high media consumptions involved, inert grinding has also been shown to produce benefits at coarser sizings (Grano *et al*., 1994; Greet *et al*., 2004; Pease *et al*., 2006). For a long time, chrome media has been offered to, and investigated by, ball and tower mill operators as a means of improving pulp flotation chemistry by reducing the amount of iron released into the grinding pulp and contaminating freshly ground surfaces. Greet and Steiner (2004), analysed the surface of galena ground in three different environments for the presence of iron. It is clear from Table I that although grinding in a high-chrome environment reduced the surface iron contamination from 16.6% to 10.2%, grinding in a ceramic environment reduced the detectable surface iron to less than 0.1% - a significant improvement over both media types.

**Table I. Composition, determined by X-ray photoelectron spectroscopy (XPS), of the un-etched surfaces of Rapid Bay galena ground with different media (Greet, 2004)**

<table>
<thead>
<tr>
<th>Media type</th>
<th>Surface atomic adsorption (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>O</td>
</tr>
<tr>
<td>Forged steel</td>
<td>53.1</td>
</tr>
<tr>
<td>High chrome</td>
<td>50.0</td>
</tr>
<tr>
<td>Ceramic</td>
<td>33.6</td>
</tr>
</tbody>
</table>

**Applications in industry**

**Early IsaMill™ adoption in zinc – lead processing. – Mt Isa Mines**

The changes to the Mount Isa circuit as part of the ‘George Fisher Project’ re detailed by Young and Gao (2000) and Young, Pease, and Fisher (2000). In summary, the George Fisher project involved adding 6 x 1.12 MW M3000 IsaMills™ to the existing Mt Isa Pb-Zn concentrator to reground lead rougher concentrate to a P₈₀ of 12 μm, zinc rougher concentrate to 12 μm, and a zinc reground to P₈₀ of 7 μm as shown in Figure 6. As a result of the circuit changes, lead concentrate grade increased by 5%, and lead recovery 5% (equivalent to 10% increase in lead recovery at the same grade). Zinc recovery increased by 10%, in two steps, and zinc concentrate grade by 2% (equivalent to 16% increase in zinc recovery at the same grade).
Due entirely to extra liberation, the project predicted 5% higher zinc recovery and no increase in concentrate grade. This was achieved instantly, as shown in Figure 7. The surprise was the ‘second wave’ increase in recovery – a further 5% zinc recovery increase and 2% increase in zinc concentrate grade. This second wave occurred because although fines flotation improved after grinding finer, it took 6 months to understand how to optimize fines flotation and customize the circuit to suit. The biggest challenges to understanding the flotation of fines were:

- False expectations that a lot more reagents would be required after IsaMilling due to the significant fresh surface area created. Some reagent additions were expected to triple
- Not taking the depressant (lime to pH 11) off the zinc cleaners
- ‘Pulling’ flotation harder because of the misconception that flotation of fines would be slower.
In reality, even with the introduction of over 6 MW of extra grinding power, the operating cost per ton of feed did not increase. This was because of the following factors:

- Lower reagent additions – this was unexpected but logical. The inert grinding environment meant that even though there was significantly more mineral surface area (the surface area had been increased threefold), the cleaner surfaces achieved with inert grinding meant that less collector was required in solution to drive the diffusion of collector through any surface coatings to the surface of the mineral. The truly clean surfaces that were achieved by grinding in an inert environment meant that lower solution concentrations can achieve high surface coverage. Flotation after inert grinding is profoundly different.
- Elimination of circulation loads between roughing and cleaning by recovering minerals and removing them to final concentrate as soon as recoverable. A lot of power (and flotation capacity) was wasted in the conventional circuit pumping circulating loads of 100%–300%.
- New designs in pump boxes and pumps, reduction in circulating loads, and lower reagents eliminated spillage. When mineral surfaces are clean due to the inert grinding environment, bubble contact angles are predictable compared to those on the highly altered surfaces found in other systems. MIM was able to drop circulating loads from 200–300% to less than 50%. Liberated minerals with clean surfaces in narrow size distributions respond quickly and do not form large circulating loads. The low circulating loads mean lower density, which gives better dilution cleaning. Low circulation loads also mean fewer flotation cells are needed – eliminating a 100% circulating load doubles the residence time.
- The reduction in reagents and circulating loads reduces or eliminates spillage. The elimination of spillage stops the cycle where returning spillage disrupts a circuit, creating new circulating loads and even more spillage.

Figure 8 shows the recovery by size in the zinc cleaning circuit at Mt Isa after the adoption of IsaMill technology (with respect to rougher concentrate or IsaMill™ feed). Recovery is above 95% for all size fractions from 1 μm to 37 μm. Although recovery drops above 37 μm there are very few particles in this size range, and these are composites that the circuit directs to further regrinding. Interestingly, the highest recovery, over 98%, is in the 4–16 μm range. This size range is sometimes called 'slimes', and it is often said that ‘slimes don’t float’. Indeed, after steel grinding they often don’t.

**Figure 8. Zn recovery increase from IsaMill™ technology (Pease et al., 2006)**

**IsaMill transforming copper flow sheets – Phu Kham**

The Phu Kham copper-gold operation is located approximately 100 km north of the Laos capital Vientiane. It is owned by PanAust, an Australian-listed mining company.

The Phu Kham deposit hosts two distinct styles of mineralization: an oxide gold cap and beneath this transitional/primary copper-gold. The oxide gold cap is the principal deposit for the Phu Bia heap leach gold mine, the first phase of the development of the Phu Kham deposit, which entered into production in 2005. As discussed by
Bennett et al., (2012) the second phase of development, the 12 Mt/a concentrator designed to treat the high-pyrite copper-gold skarn ore, was commissioned in 2008.

The Phu Kham concentrator design was driven by the complex and variable mineralogy, where, due to the geological nature of the orebody the concentrator treats varying native, oxide, secondary, and primary copper species over a short time periods, and to the high pyrite content of the ore, with over 90% pyrite rejection required to achieve a 23% copper grade.

Expansions since the 2008 start-up have been driven by the relatively poor recovery when compared to other low-grade copper-gold deposits and have included an increase in rougher and cleaner capacity and the installation of a Jameson cleaner. Figure 9 shows the Phu Kham flow sheet following installation of the Jameson Cell cleaner in 2011. Of note is the original 3 MW M10,000 IsaMill™ installed to produce a cleaner feed size \( P_{80} \) of 35 \( \mu m \) and the 24 downcomer Jameson Cell installed in 2011 to improve flotation recovery.

Since installation of the Jameson Cell, the Phu Kham processing plant has undergone two further major upgrade projects. The Phu Kham Upgrade Project commenced in early 2010 and, as discussed by Bennett (2012), the focus of the project was to ensure the concentrator could maintain concentrate production at over 60 kt/a after 2013 as the copper grade starts to decrease. To maintain concentrate production, overall plant throughput had to be increased from 12 Mt/a to 16 Mt/a in the limited real-estate environment at site.

The Increased Recovery Project (IRP), for which initial studies started in 2009, was focused on increasing copper and gold recoveries to overcome the ‘flat line’ in recoveries (shown in Figure 10) that resulted from poor liberation coupled with increasing pyrite content. As part of the IRP project, two mineral processing options for increasing recovery of the low-quality copper sulphide – non-sulphide gangue binary particles were tested: mainstream inert grinding of the rougher tails and bulk sulphide flotation in the rougher circuit. The selected bulk sulphide flotation processing route targeted rougher mass recovery to maximize copper recovery across all ore types, rougher concentrate reground product size, and cleaner capacity.

Monthly composite mineralogy data and laboratory test work determined that a reground size of \( P_{80} \) 20 \( \mu m \) was required to achieve 80% copper sulphide liberation and so maximize copper sulphide liberation in the cleaner feed. To achieve this reground target at increased rougher concentrate production, a second 3 MW M10,000 IsaMill™ was
commissioned in the first half of 2013. The current Phu Kham flow sheet is shown in Figure 11. The initial operating results will be fully reported in later papers, but indications are that the IsaMills are achieving their design performance.

Figure 10. Phu Kham historical copper and gold recoveries (Bennett, 2012)

Figure 11. Phu Kham flow sheet following completion of the IRP (Bennett, 2012)
Conclusion

Necessity was the mother of invention for ultra-fine-grained orebodies, which were uneconomical to process with conventional ball and tower milling. High-intensity stirred milling was the breakthrough that transformed the economics of these deposits. It is the advantages of IsaMill™ technology – high power efficiency, simple installations, large single unit size, sharp size distributions in open circuit, and inert attrition of mineral surfaces that have resulted in the rapid transfer and uptake of the technology in base metal applications around the world.

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Enhancing Magnetite Returns – The Benefits of IsaMilling
M Larson¹, T Do² and J Siliezar³

ABSTRACT
With the expansion of magnetite projects in Australia has come the opportunity to take advantage of newer technologies that were not necessarily available when the magnetite industries of North America and Europe were maturing. One of these technologies is the IsaMill, originally developed for the economic treatment of ultra-fine grind (UFG) duties of Mount Isa Mines. As understanding of the IsaMill has increased, so too has the ability to process coarser feeds while maintaining the same energy efficiency advantage and tight product size distribution that originally made the IsaMill so attractive. In the case of magnetite the hard non-elastic ceramic grinding media is ideally suited to the grinding duty when compared to steel grinding media. The centripetal separator and naturally steep product size distribution curve provide a mill discharge that maximises the critical concentrate grade. This paper analyses the effect of the IsaMill grinding technology on the full-scale Ernest Henry Mine (EHM) magnetite operation along with pilot and lab scale demonstrations of optimisation. Issues encountered and the solutions to maximise the mill throughput and quality are examined in an effort to promote better understanding across the industry.

INTRODUCTION
In 2011, Ernest Henry Mine (EHM) started processing the magnetite from their copper/gold concentrator tailings. The flow sheet for this plant is shown in Figure 1.

The inclusion of the IsaMill was at the time unique for magnetite processing. The decision was made after comparative tests of an M4 IsaMill and a pilot 40 L tower mill. The results of this comparison are shown in Figure 2 (Burford and Niva 2008).

The crossover of energy efficiencies between the tower mill and IsaMill is common and will depend on feed size, ore hardness and media sizes employed. The same trend is seen in Figure 3 comparing a ball mill with an IsaMill for a different magnetite project (David, Larson and Le, 2011).

Since the time of the initial test work, and further developments have been made benefitting IsaMill operations in the manufacturing of reliable larger ceramic beads. While the EHM comparison test work was done with 3.5 mm ceramic beads 5 or 6+ mm beads are now commonly offered by a variety of manufacturers. This ensures a more complete top size reduction vital for magnetite operators to make final silica grade and also further reduces the already low amount of ultra-fines created which could complicate downstream processing with their propensity to lead to fine liberated silica entrainment in magnetic concentrates. The ceramic media has an advantage over steel media in that it is far less elastic and transfers energy more efficiently to the ore rather than just being deformed.

ISAMILL PERFORMANCE
During the first year the IsaMill was typically receiving a Fₚ₈₀ of 250 - 300 microns (compared to a design feed size of ~150 microns) and an absolute top size in excess of 1 mm. Cenotec ceramic media with a top size of 6.5 mm was used in the IsaMill with the goal of better breaking the coarse end of the feed.

Fortunately the deviation from design was realised prior to commissioning. This feed was similar in size to that tested at AMMTEC in May 2011. Some limited survey results are plotted on Figure 4 showing the AMMTEC signature plot with the good scale-up correlation typical of IsaMill installations. The mill net power draw (gross power ~ 80 kW) was also used as the AMMTEC results are expressed in net energy required.
The AMMTEC tests were performed with a top size 6 mm of Cenotec CZM media while the plant uses 6.5 mm Cenotec. Figure 5 shows the actual 3 MW M10 000 installed at EHM. The feed enters on the left of the picture at the non-drive end and discharges from the drive end where it is then sent to the regrind magnetic separators.

Also included in Figure 6 is an example of the IsaMill product size distribution. Considering the coarseness of the feed the steepness of the product size distribution is more than acceptable. Much of the coarse material in the product appears to be silica reporting to the discharge due to density differences with the finer higher specific gravity magnetite. The regrind magnetic separators concentrate during this survey contained zero material above 106 microns and minimal material above 75 microns.

**IMPROVING THE ISAMILL FEED**

As not all of the tails are currently processed, there is an opportunity to improve the grinding capacity of the magnetite circuit. This could be achieved by screening the rougher magnetic separator feed, concentrate or the cyclone underflow at 150, 212 or 300 microns and returning the oversize to the main copper concentrator. From an early survey shown in Table 1 it can be seen that this size fraction...
grades about 0.45 per cent copper and 40 per cent silica. This has the potential to improve the overall copper recovery of the EHM concentrator and reduce the amount of silica reporting to the magnetite cleaner circuit. In addition to any potential benefit in increased copper recovery, removing this material would reduce the wear on the IsaMill, increase the throughput of the IsaMill and improve downstream grade and filtering performance by reducing the amount of silica being fed to the IsaMill and potentially being turned into silica slimes.

It is an option at EHM to improve grade (assuming the liberated silica entrainment is controlled) by screening the rougher magnetic concentrate. This would result in a finer more homogenous feed to the IsaMill and also reduce the amount of coarse middling material discharging from the IsaMill. It would be expected that as a result the specific energy required would decrease to <20 kWh/t from the current 30 - 35 kWh/t. It would also increase the iron grade of the regrind feed by five to ten per cent Fe. The amount of actual magnetite processed through the IsaMill would increase from about 55 t/h to over 90 t/h under this scenario compared to the coarsest feed scenario. This has the potential to improve the concentrate grade without alternative treatment such as finer Derrick screens on the final concentrate. A setup for 300 micron, 212 micron or 150 micron screening of the rougher magnetic separator concentrate prior to the cyclones or IsaMill feed would also be more manageable than the finer Derrick screens. About half of the silica present and two-thirds of the non-magnetic material would be removed by a 300 micron screen. By not sending the material to IsaMill in the first place this non-magnetic material would not be able to be turned into fines more susceptible to entrainment further downstream.

The results of screening under four different sampled scenarios spanning three months is included in Table 2. By installing a 212 micron Derrick screen prior to the IsaMill on the rougher magnetic separator concentrate an improvement in actual magnetite throughput of about 30 - 40 per cent could conservatively be expected.

![FIG 6 - Ernest Henry Mine M10 000 IsaMill feed and discharge product size distributions.](image)

![TABLE 1](image)

<table>
<thead>
<tr>
<th>Size fraction (µm)</th>
<th>Cumulative % passing</th>
<th>% copper</th>
<th>% iron</th>
<th>% silica</th>
</tr>
</thead>
<tbody>
<tr>
<td>850</td>
<td>99.25</td>
<td>0.42</td>
<td>15.41</td>
<td>43.83</td>
</tr>
<tr>
<td>600</td>
<td>96.31</td>
<td>0.44</td>
<td>14.84</td>
<td>43.62</td>
</tr>
<tr>
<td>425</td>
<td>89.46</td>
<td>0.49</td>
<td>15.61</td>
<td>42.98</td>
</tr>
<tr>
<td>300</td>
<td>79.62</td>
<td>0.43</td>
<td>16.91</td>
<td>41.05</td>
</tr>
<tr>
<td>212</td>
<td>69.72</td>
<td>0.22</td>
<td>23.07</td>
<td>38.27</td>
</tr>
<tr>
<td>150</td>
<td>58.15</td>
<td>0.10</td>
<td>36.74</td>
<td>22.87</td>
</tr>
<tr>
<td>106</td>
<td>43.25</td>
<td>0.05</td>
<td>51.23</td>
<td>13.75</td>
</tr>
<tr>
<td>75</td>
<td>30.06</td>
<td>0.03</td>
<td>54.31</td>
<td>9.11</td>
</tr>
<tr>
<td>53</td>
<td>18.92</td>
<td>0.05</td>
<td>58.71</td>
<td>5.99</td>
</tr>
<tr>
<td>38</td>
<td>10.48</td>
<td>0.05</td>
<td>61.66</td>
<td>3.74</td>
</tr>
<tr>
<td>20</td>
<td>5.40</td>
<td>0.11</td>
<td>64.20</td>
<td>3.31</td>
</tr>
<tr>
<td>-20</td>
<td>0.15</td>
<td>49.66</td>
<td>8.60</td>
<td></td>
</tr>
</tbody>
</table>

![TABLE 2](image)

<table>
<thead>
<tr>
<th>Case 1</th>
<th>Fₘ (microns)</th>
<th>Energy (kWh/t)</th>
<th>Total throughput</th>
<th>Fe grade</th>
<th>Magnetite throughput</th>
<th>% improvement</th>
</tr>
</thead>
<tbody>
<tr>
<td>Normal</td>
<td>300</td>
<td>30</td>
<td>93.3</td>
<td>42.3</td>
<td>54.8</td>
<td></td>
</tr>
<tr>
<td>300 µm screen</td>
<td>180</td>
<td>23.8</td>
<td>117.6</td>
<td>49</td>
<td>80.1</td>
<td>46.0</td>
</tr>
<tr>
<td>212 µm screen</td>
<td>140</td>
<td>21.9</td>
<td>127.9</td>
<td>52.7</td>
<td>93.6</td>
<td>70.7</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Case 2</th>
<th>Normal</th>
<th>260</th>
<th>25.8</th>
<th>108.5</th>
<th>37.4</th>
<th>56.4</th>
</tr>
</thead>
<tbody>
<tr>
<td>300 µm screen</td>
<td>180</td>
<td>23.8</td>
<td>117.6</td>
<td>41</td>
<td>67.0</td>
<td>18.8</td>
</tr>
<tr>
<td>212 µm screen</td>
<td>140</td>
<td>21.9</td>
<td>127.9</td>
<td>43.6</td>
<td>77.4</td>
<td>37.3</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Case 3</th>
<th>Normal</th>
<th>155</th>
<th>22.7</th>
<th>123.3</th>
<th>46.4</th>
<th>79.5</th>
</tr>
</thead>
<tbody>
<tr>
<td>300 µm screen</td>
<td>120</td>
<td>20.5</td>
<td>136.6</td>
<td>49.2</td>
<td>93.3</td>
<td>17.4</td>
</tr>
<tr>
<td>212 µm screen</td>
<td>100</td>
<td>18.4</td>
<td>152.2</td>
<td>51.1</td>
<td>108.0</td>
<td>35.9</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Case 4</th>
<th>Normal</th>
<th>212</th>
<th>24.8</th>
<th>112.9</th>
<th>44.3</th>
<th>69.5</th>
</tr>
</thead>
<tbody>
<tr>
<td>300 µm screen</td>
<td>150</td>
<td>22.5</td>
<td>124.4</td>
<td>48.5</td>
<td>83.8</td>
<td>20.7</td>
</tr>
<tr>
<td>212 µm screen</td>
<td>125</td>
<td>20.9</td>
<td>134.0</td>
<td>52.5</td>
<td>97.7</td>
<td>40.6</td>
</tr>
</tbody>
</table>
In addition to the large amount of coarse material in the feed it was found by Davis tube analysis that 14 per cent of this feed was liberated non-magnetic material. Similar data from an earlier survey showed 20 per cent liberated non-magnetic material in the reground fresh feed.

Fine screening in general has been used for over 40 years in Minnesota (Devaney, 1985). From Healey, already in 1972 Erie Mining Company Hoyt Lakes plant had 200 mesh (74 micron) Rapifine screens supplied by Dorr-Oliver Company to screen the final concentrate. The actual paper cited came from 1967. The results of this screening are shown in Table 3.

### TABLE 3
Typical results from two-stage fine screening of finisher concentrate (Healy, 1967).

<table>
<thead>
<tr>
<th></th>
<th>% wt</th>
<th>% 325 mesh</th>
<th>% distribution of 325 mesh</th>
<th>% silica</th>
<th>% distribution of silica</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fine screen feed</td>
<td>100.0</td>
<td>88.5</td>
<td>100.0</td>
<td>8.5</td>
<td>100.00</td>
</tr>
<tr>
<td>Fine screen oversize</td>
<td>18.9</td>
<td>52.0</td>
<td>11.1</td>
<td>20.5</td>
<td>45.9</td>
</tr>
<tr>
<td>Fine screen undersize</td>
<td>81.1</td>
<td>97.0</td>
<td>88.9</td>
<td>5.7</td>
<td>54.1</td>
</tr>
</tbody>
</table>

EHM has recently commissioned pilot Derrick screen deck unit to prove the application for their plant. Table 4 shows the split of the iron around the screening pilot plant. The Derrick screen is widely accepted in the iron ore industry, though it is best known in application using finer screens in screening relatively coarse material (and thus lower grade) from final concentrate to improve iron grade.

The key points from Table 4 (Do, 2012):
- More than 90 per cent of the material in the stream was classified correctly.
- The per cent solids in the undersize is too low to constitute a feed to the IsaMill. This will require a magnetic densifying step prior to feeding the IsaMill.
- A detailed study will need to be completed to identify the preferred location of the screening plant in order to maintain the proper feed density to the IsaMill with minimal extra costs. Even so this application for the Derrick screen appears ideal. Rather than screening low-grade material after processing it is done prior to grinding saving the energy that would otherwise be spent grinding non-magnetics down to 45 microns.

If the whole of the EHM plant were to be thought of in terms of only processing magnetite with the copper concentrator rejecting the pesky gangue copper and gold, this proposed flow sheet fits in very well. The semi-autogenous grinding (SAG) mill and ball mill are ideally suited for the coarser grinding duties so the coarse fraction of the Derrick screen product returns to the ball mill where it is more efficiently ground. This leaves the IsaMill to grind the finer, iron-rich particles. The grind from 150 to 45 microns is the most energy intensive part of the process so it is best that the particles being ground to the final product size have more value.

By screening the feed, the grinding results can be reliably predicted using the squared function for fines production developed for the IsaMill JKSimMet model. As 45 microns is the size of interest in this case, energy can be predicted from one case to another as long as the per cent passing 45 microns in the feed is known and it is assumed that the feed size distribution curves are relatively normal. This is shown in Figure 7. This will be further proven with a 20 L pilot IsaMill at site. In addition this will be used to optimise the media size prior to installing the screens in the full-scale application.

One of the main advantages of the IsaMill is the steep product size distribution it produces. This is shown in Figure 8, where a 37 micron P43 results in a P90 of sub-75 microns and zero material over 105 microns (David, Larson and Le, 2011).

It could be expected that by reducing the feed top size a similar performance could be gained from the Ernest Henry M10 000. Figure 9 shows the product sizes from P80’s of 38 - 45 microns screening the feed at various sizes.

### MAGNETIC SEPARATOR OPERATION AND UPGRADEING

There is also room for improvement in the magnetic separator operation feeding the IsaMill. Some improvements have already been made, such as running the magnetic separators at a lower bath level. Magnetic separators are efficient on their own for recovery; the goal is to ensure that entrained silica is limited in the concentrate. By running at a high bath level and increasing the residence time it increased the amount of entrained non-magnetic material reporting to the concentrate. It is best to think of the magnetic separators as giant washing machines, where high flows of water are necessary to wash the silica from the magnetite. Figure 10 shows sanding in one of the regrind magnetic separators. By having the drums in contact with this bed of material it will result in it scooping up material with little selectivity. The picture on the right shows the same magnetic separator minutes later after opening the tailings valves.

In addition to running at low bath levels another change was to raise the drum levels. In the case of the regrind magnetic separators, virtually zero magnetite was in the tails stream but entrained non-magnetic was at over ten per cent of the concentrate. These results were with the separators running at a low bath level. With the drums raised, the tailings remained at below one per cent magnetite but the entrained non-magnetics dropped significantly to a point where the regrind magnetic separators were making a grade of about 68 per cent Fe. Most of the magnetite that will be lost under this operating mode will either be ultra-fine iron which is not efficient to filter or small pieces of magnetite in composites.
with much larger middling particles which will only act to lower the final concentrate grade. Any effort to recover either of these types of iron losses typically results in more of an increase in penalty in entrained silica than can be gained in improved iron recovery, and this is the reason most magnetite producers willingly sacrifice a small amount of recovery for a substantial improvement in grade. By raising the drum level it will also ensure that the drum is not rotating through a bed of settled solids, picking up silica and increasing the stress on the shaft. In addition results from a 19 July survey showed the incoming regrind feed was about 20 per cent liberated non-magnetic material. It should be noted that this was with the rougher magnetic separators bath levels still high.

Every effort should be made to reduce the entrainment of non-magnetic material in the regrind feed. By sending coarse silica particles to the IsaMill it is ensured that these will be ground down to create unnecessary silica slimes that are even more difficult to reject and negatively impact grade.

FIG 7 - Ernest Henry Mine production of 45 micron material.

FIG 8 - IsaMill product size distribution curve.

FIG 9 - Ernest Henry Mine IsaMill product size distributions at different feed size screens.

FIG 10 - Regrind magnetic separator sanded (left) and with tailings valves open (right).
and filtering. Figure 11 shows a sample of the +300 micron size fraction of the regrind circuit fresh feed, with the lighter colour coarse non-magnetic material obvious.

![Figure 11 - Regrind fresh feed oversize screened at 300 microns.](image)

During the second extended visit to site dedicated to operations improvement the rougher magnetic separator drums were raised from 50 mm to 75 mm of their mechanical limit. Short surveys of the tailings stream after showed that the tailings only contained about one per cent magnetic iron after the adjustment. This is consistent with good practices around the world and typical rule of thumb operational goals for copper magnetic separators. The next work to be undertaken on the rougher magnetic separator is to overhaul the distributor to even the slurry flow to the four banks of separators. Currently two of the banks receive most of the flow with the other two largely starved of feed. A general rule of thumb for sizing magnetic separators in North America is to have about 8-12 t/h of feed per foot of drum length (about 26-40 t/h per metre of drum length). Under ideal operating conditions and at a typical feed of 1400 t/h of fresh feed this would make the EHM separators undersized by about 50 per cent for the drums receiving the most feed.

Fixing the rougher magnetic separator distributor should allow EHM to further optimise the magnetic separators. There was a large amount of non-magnetic build-up observed under the drums that is likely contributing to liberated non-magnetic material in the concentrate. Adjustments to the flow pattern and water addition are planned for the future to deal with this.

Further it is important to point out the necessity of differentiating between magnetic and non-magnetic iron in the tailings; the grades and recoveries of these two iron species should be reported separately. The total Fe in the tail was assayed at about eight per cent; however only 0.5 per cent of the total mass was recovered in the Davis tube of the tailings stream, indicating that over 96 per cent of the iron in the tails is in the form of hematite or other low/non-magnetic form of iron. No effort should be made to recover this, as recovering non-magnetic iron will also result in the recovery of other non-magnetic penalty elements such as silica. Analysis of tailings streams should be done by Davis tube, Satmagan or even a hand-held magnet and never solely by straight assay for iron so that the recovery of the magnetic and non-magnetic iron can be fully understood and targeted. In the case of running a Davis tube on the magnetic separator tails, an assay should be done on the magnetic portion. Previous work done on testing different drum heights likely came to faulty conclusions due to assays not being done on the Davis tube fractions.

**CONCLUSIONS**

The M10 000 IsaMill scaled reliably from 4 L lab tests, even with a coarse feed and large media size. The product from the IsaMill was a wider size distribution than desired but it would appear that the top size was mostly liberated light silica and did not negatively impact downstream grade control. The feed to the IsaMill can still be optimised by screening the coarsest fraction and by improving the magnetic separator performance to better reject coarse liberated gangue material. These steps will allow the IsaMill to process more material of a higher grade and improve EHM’s overall magnetite production.

**ACKNOWLEDGEMENTS**

The authors wish to thank the staff of the EHM for their permission to publish this paper along with assistance in gathering the information presented. The authors also would like to thank the staff at ALS AMMTEC in Perth for the work done on numerous IsaMill signature plots done for EHM.

**REFERENCES**


Improving Energy Efficiency Across Mineral Processing and Smelting Operations – A New Approach

C L Evans¹, B L Coulter², E Wightman³ and A S Burrows⁴

ABSTRACT
With their commitment to sustainable development in an increasingly carbon-constrained world, many resources companies are focused on reducing the energy required to create their final products. In particular, the comminution and smelting of metal-bearing ores are both highly energy-intensive processes. If resource companies can optimise the energy efficiency across these two processing stages, they can directly reduce their overall energy consumption per unit mass of metal produced and thus reduce their greenhouse gas footprint.

Traditional grinding and flotation models seek to improve the efficiency of mineral processing, but they do not consider the energy used by downstream metal production, eg smelting and refining. Concentrators and smelters individually may be running efficiently, but are they optimising energy consumption of the overall system? A key step towards answering this question is to understand whether it takes less energy to remove an impurity in the concentrator or the smelter. Analysis of the complete concentrator – smelter process chain may show that there are times when the concentrator should use more energy to reduce the overall energy requirements across the mill and smelter.

The new methodology discussed here will integrate downstream processing energy with mineral processing energy, to find the overall most energy efficient circuit design and operating strategy. The proposed methodology includes:

- determining what grade and recovery positions can be achieved in the concentrator by using different combinations and amounts of grinding and regrinding;
- thermodynamic calculations of the energy used to process the different concentrate grades to metal, and
- integrating the concentrating and smelting models to find the lowest total energy for the system.

A case study which investigates the effect of increasing regrinding energy on the overall mineral processing – smelting energy consumption is presented. For the copper-nickel sulfide ore investigated the addition of regrinding results in 11 per cent less energy being used in the overall concentrator – smelter process chain per tonne of metal produced. Although regrinding is often seen as an energy-intensive process, in this case study the use of 1 kWh of energy for regrinding reduces the overall energy consumption by 12 kWh. The new mill to melt methodology will allow companies to model energy consumption from mill to smelter with a view to reducing the energy-intensity and greenhouse gas footprint of their processing and smelting operations.

INTRODUCTION

Global resource companies are currently operating under very challenging economic and regulatory conditions. In response to growing societal concern about the various impacts of the minerals industry, and the emergence of the concept of sustainable development as the key framework within which these impacts are analysed, most organisations in the sector are now reporting their performance in this area using a range of sustainability indicators. Among these, energy use and its impact on climate change are priorities, with many company sustainability reports including total energy use and associated greenhouse gas (GHG) emissions in both absolute and relative terms (ie normalised to a per unit product) among their key environmental sustainability indicators. Companies are setting targets to achieve improvements in these indicators, but at the same time there is a global trend to more complex and lower grade orebodies which are more energy-intensive to process. As a result, resource companies are having to become more innovative to improve the environmental sustainability and the efficiency of their operations. In particular, companies must address the specific energy consumption of their processes in order to reduce their emissions of greenhouse gases.

Mineral concentration and metal smelting are two energy-intensive stages in the production of metals. According to the Australian Government’s National GHG Inventory (2009), the production of mineral and metal products generated 26 million tonnes CO₂-e in 2005, equivalent to 4.5 per cent of Australia’s total GHG emissions. Smelting is energy- and GHG-intensive, with emissions from smelting up to three to ten times greater than those from mineral concentration processes (Ngora, Jahanshahi and Rankin, 2007). These two processes offer real opportunities to reduce the energy intensity of metal production.

In practice, mineral concentrators and smelters typically operate as independent units with their own product specifications and energy targets. We believe that the overall energy consumption across the metal production chain can be reduced by:

- scrutinising the energy consumers in the concentrator and smelter,
- defining the concentrate that achieves the lowest overall energy consumption, and
- modifying the product specification of the concentrator accordingly.

In particular, this paper looks at the effects of changing the amount of grinding energy on the required smelting energy and therefore the overall energy consumed to produce metal.

This paper outlines a proof of concept investigation to determine whether the process and energy models of concentrators and smelters can be integrated to produce an effective energy and GHG footprint model for metal production. The investigation includes a case study on a copper-nickel sulfide ore to demonstrate how the methodology would be applied. Proving that the mill to melt concept is viable is the first step in the journey towards developing a methodology for minimising energy consumption and associated GHG emissions along the entire concentrator and smelter metal production chain.

THE MILL TO MELT METHODOLOGY

The objective of the mill to melt methodology is to provide a framework for minimising energy use and GHG production per tonne of metal produced in mineral processing and smelting. Metallurgists recognise that there is scope to tune the operation of a concentrator to change the product quality and also scope to tune the smelter operation to smelt a different grade of

1. Senior Research Officer, The University of Queensland, Sustainable Minerals Institute, Julius Kruttschnitt Mineral Research Centre, Indooroopilly Qld 4068. Email: c.evans@uq.edu.au
2. Metallurgical Consultant, Xstrata Technology Ltd, 307 Queen Street, Brisbane Qld 4000. Email: bcoulter@xstratatech.com.au
3. Senior Research Fellow, The University of Queensland, Sustainable Minerals Institute, Julius Kruttschnitt Mineral Research Centre, Indooroopilly Qld 4068. Email: e.wightman@uq.edu.au
4. Senior Metallurgical Engineer, Xstrata Technology Ltd, 307 Queen Street, Brisbane Qld 4000. Email: aburrows@xstratatech.com.au
concentrate. However, there is no integrated tool currently available which allows companies to model the combined effect of these changes on overall energy consumption in the mill and smelter.

The mill to melt methodology aims to integrate mass and energy models of both mineral concentration and smelting stages to provide a tool for optimising overall energy consumption across these two energy-intensive processes. The methodology will identify the relationship between metal grade and recovery in the concentrator and the overall energy consumption across both the concentrator and smelter. This relationship can be expressed as a three-dimensional response surface and will allow the minimum energy operating point to be identified for a given operation. Note that the variation in ores, concentration and smelting processes between mining operations will mean that it is unlikely that a universal relationship which can model the response for all ores will be identified. In the mill to melt methodology the generic model will be fitted to the combination of ore, concentration and smelting processes at each site.

With the increasing focus on reducing GHG emissions from metal production a tool such as the mill to melt methodology is likely to find industrial application in both existing operations and new projects. It will be most beneficial in integrated operations where a single company owns both the concentrator and smelter.

Modelling mineral processing performance

Laboratory procedure

In the mill to melt methodology the process route used to concentrate an ore is implemented on laboratory scale equipment. This allows a variety of circuit configurations and operating strategies to be compared and is equally applicable to existing operations or greenfield sites. The laboratory procedure developed in the proof-of-concept work uses laboratory batch grinding mills and froth flotation cells in the same circuit arrangement as would be used in a full-scale industrial plant. Energy consumption during each of the grinding stages is monitored and recorded using an energy metre. A typical process flow sheet with two stages of grinding and separation is shown in Figure 1 and this is the process route used in the case study presented later in this paper.

![Fig 1 - A typical concentrator flow sheet with two stages of grinding and separation.](image)

The laboratory tests generate physical samples of ore particles from the feed, concentrate and tail streams which are characterised using the JKMRC/FEI MLA automated mineralogy system. Information about the minerals which occur in a range of particle size fractions is used in the modelling procedure, together with data which describe how the ore particles behave in the separation processes.

Mill to melt modelling procedure

Since the link between the mineral concentration and smelting stages requires information about both the mineral and elemental composition of the process streams the modelling procedure tracks the flow of minerals in the circuit. It is a simple step to convert the mineral data into elemental composition data.

The aim of the modelling procedure is to minimise the amount of laboratory work required to identify the operating point at which the minimum energy and greenhouse gas footprint is achieved. This will make the mill to melt methodology a practical and affordable tool.

One approach reported recently (Wightman et al., 2008; Wightman and Evans, 2009) shows that, for many ores, assumptions can be made about how the minerals liberate during comminution. These assumptions allow the comminution and liberation response of the ore to be calibrated from one set of physical tests. These calibrated values can then be used to predict the liberation characteristics at a wide range of grinding energy inputs without further laboratory testing. This 'measure few, model many' approach saves considerable time and money in applying the methodology.

The flotation modelling also requires laboratory tests to measure the response of the ore. This calibration of flotation response is essential because the variability between ores as naturally occurring substances means that each ore requires a set of flotation conditions tuned to its particular natural characteristics. The ongoing flotation modelling work seeks to minimise the amount of physical testing required by tracking the flotation response of the particles into the various concentrate and tailing streams.

Modelling smelter energy requirements

The smelter modelling stage uses computer-based, thermodynamic models of the smelting process to predict the input energy requirements. Depending on the ore being modelled there may be one or two final concentrates whose smelting energy requirements need to be taken into account separately. In the case study presented in this paper two concentrates are produced in the concentrator and thus two parallel smelting operations are included in the analysis, one for copper and one for nickel smelting.

Smelter models, based on thermodynamics, require information about both the chemical and mineral composition of the concentrate fed to the smelter. The mill to melt methodology generates information about the concentrate in terms of mineral mass flows which are readily converted to elemental flows.

A model of the smelting energy requirements can best be done on a case-by-case basis because there is variation in the equipment units employed at different smelters. To simplify the analysis, without loss of accuracy, it is sufficient that a case study should focus on modelling equipment units within a smelter that satisfy both of the following criteria:

- the greedy energy consumers, and
- the mineral-dependent equipment units.

Equipment units satisfying these two criteria are the most influential when making decisions about the concentrator-smelter interface and therefore the most important for the mill to melt optimisation.

An important consideration for smelter modelling is the inclusion of different energy sources. A smelter may have separate inputs of multiple fossil fuel types as well as electricity.
This case study includes different sources and expresses the model output as a common energy unit of kilowatt hours (kWh).

**A MILL TO MELT CASE STUDY**

The route chosen to prove that the concept of linking concentrator and smelter process and energy models is feasible involved developing a case study using real ore. The ore used in this proof-of-concept phase was a copper-nickel sulfide ore from an operating site in Canada. This ore is treated to produce both nickel and copper products. The smelter models used in this case study were based on the equipment units in the actual smelters which process these copper and nickel concentrates.

As with many base metal concentration process routes, the ore is subjected to two sequential stages of grinding and separation, as shown in Figure 1. When the smelters are included, there are four energy-intensive stages to be analysed in order to minimise overall energy consumption and the resulting GHG emissions of the operation.

In the example presented here, the effect on overall energy consumption of using varying amounts of energy in the second (or regrinding) stage is examined. The primary grinding energy is held constant and two levels of energy are applied in the regrinding stage. Laboratory tests and the corresponding mineralogical analyses of the products provide the information needed to model the grinding and flotation stages.

As Figure 2 shows, as the energy input to the regrinding of the rougher concentrate is increased the subsequent separation process is able to achieve a higher grade of flotation product at a given recovery of the target metal.

This shift in behaviour occurs because the comminution stage breaks the mineral grains in the rock apart, allowing the new separate grains of gangue mineral to be removed in the subsequent flotation stage. Removal of the gangue minerals in the concentrator means that energy is not spent removing it in the smelter. The question which the mill to melt methodology will allow us to answer is whether this will lead to a net reduction or increase in energy consumption across the concentrator – smelter production chain.

In Figure 3 and Figure 4 the effect on smelting energy requirements of increasing the amount of energy input to comminution, in this case to regrinding, is shown.

Note that the calculations for the smelter models include only the energy consumed by the concentrate-dependent, high energy consuming units (Somanathan and Tripathi 2008; Tripathi and Mackey 2009). For this case study, the relevant units were:

- nickel smelter – the electric smelting furnace; and
- copper smelter – the concentrate dryer, the production of industrial oxygen and compressed air for smelting/converting, the electric slag cleaning furnace.

These smelting units have been modelled assuming:

- the same concentrate moisture applies to the concentrate with/without regrind;
- the same smelter operating temperature for concentrate with/without regrind; and
- the concentrate is smelted at the typical smelting rate applicable at the case-study smelters, i.e., for the purposes of the energy model the smelter production rate was unconstrained by the concentrate production rate.

The energy data shown in Figures 3 and 4 indicate that the energy required to smelt the concentrates is reduced if regrinding is included in the mineral concentration stage. It is particularly interesting to note that the reduction in smelter energy consumption as a result of regrinding in the concentrator is many times larger than the energy consumed regrinding. The positive impact of regrinding on the total energy consumption is highlighted by the summary of the energy data presented in Figure 5 and Table 1.

**TABLE 1**

**Summary of energy consumption in the concentrator – smelter process chain with and without regrinding.**

<table>
<thead>
<tr>
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<th>Without regrinding</th>
<th>With regrinding</th>
<th>% energy saving from using regrinding</th>
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<tbody>
<tr>
<td>Total energy consumed per tonne of ore in feed (kWh/t)</td>
<td>55.7</td>
<td>50.9</td>
<td>8.6%</td>
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<tr>
<td>Total energy consumed per tonne of metal produced (kWh/t)</td>
<td>3910</td>
<td>3461</td>
<td>11%</td>
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As the data in Table 1 show, in this case study the introduction of a regrinding stage in the concentrator results in an estimated reduction of 11 per cent in the total energy consumption per unit mass of metal.

To assist in comparing various process operating points it may be useful to calculate the ratio of the net increase in regrinding energy to the resulting net reduction in smelter energy requirements. This ratio, termed the mill to melt energy dividend (Emm) is calculated as follows:

\[
E_{MM} = \frac{\text{Net smelter energy saving with regrinding}}{\text{Energy consumed in regrinding}}
\]  

For the current study, the Emm value calculated for the two process strategies shown in the case study is 12 which indicates the energy invested in regrinding is returned twelve-fold by the resulting reduction in energy consumption in the smelter. The Emm value may be a useful indicator of the relative benefits of regrinding in the concentrator.

The methodology has the potential to allow mining companies to optimise the use of energy across concentrator and smelter processes, reducing the carbon footprint of their operations. This proof of concept work indicates that the approach to integrating concentrator and smelter mass and energy models in the mill to melt methodology is worth investigating further.
Having shown that the concept of using concentrator and smelter process and energy models together to reduce energy use across the entire metal production chain is feasible, the next step is to develop the methodology further and to test its applicability to a range of ore types.

An important aspect of the ongoing flotation modelling work in mill to melt will be to optimise the flotation performance of ore particles after regrinding. The scope of the initial proof-of-concept work did not include flotation optimisation studies but there is potential for greater rejection of gangue after regrinding when optimisation is included in next phase of work. This has the potential to further reduce the total energy requirements.

The integration of the concentrator and smelter models is a key step which will allow the mill to melt methodology to be used in energy and GHG footprint minimisation studies.

With the addition of a small increment of complexity, the smelter energy modelling could be enhanced by the inclusion of mineral-dependent operating temperature calculations. In this scheme, the smelter operating temperature would be permitted to fall/rise to reflect the relative ease/difficulty of smelting particular minerals. This would more closely approach the real operation and would demonstrate the disproportionate energy savings that are achievable by selective rejection of problematic gangue species at the concentrator stage.

In some case studies, where the smelter and concentrator are geographically separated, the mill to melt approach could be expanded to include the logistics chain between the concentrator and smelter, because the impact of transport energy, per tonne of metal, is also dependent on concentrate grade.

Note that this paper has focused on one particular sustainability indicator, namely specific energy consumption. We acknowledge that changes in one area have the potential to influence other key sustainability indicators such as water use. The purpose of this case study was to demonstrate how changes in operating philosophy, implemented as a result of adopting a wider view of an integrated processing system, could result in net gains in terms of energy reduction. The further work described above could also eventually be expanded to include other key environmental sustainability indicators, depending on the context of the work.
ACKNOWLEDGEMENTS

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REFERENCES


Responding to the Challenge – Necessity Driving Circuit Change

A Taylor¹, V Lawson² and R Barrette³

ABSTRACT
Over the last year at Clarabelle Mill, three circuit reconfigurations were implemented. One circuit reconfiguration, ‘A Cleaning’, was initiated due to a furnace failure at the smelter complex, and two of the circuit reconfigurations, ‘increased copper (Cu) separation’ and the ‘Interim IsaMill™ circuit’, were seen as opportunities to increase cash flow through the mill. A Cleaning and the Interim IsaMill™ circuit took advantage of engineering design already completed for a major capital project. The engineering for each was modified, as needed, and construction of the required portions of the capital project was fast tracked. The increased Cu separation circuit changed existing test and plant equipment to increase the plant Cu concentrate production. Implementation of the projects required teams that included operations, maintenance, technical support, project management and capital project resources. This paper discusses the business reasons driving the changes and the forethought or synergies with other work, which made them a success, and what was learned through each project.

INTRODUCTION
Vale is the world’s second largest diversified mining company; having interests in iron mining and production, base metal mining and finishing, logistics, fertilisers as well as in the energy sector. Vale has production plants and assets in 38 countries worldwide, employing approximately 119 000 people. Clarabelle Mill is part of Vale’s nickel business within the Ontario Operations of the North Atlantic business group. Ontario Operations includes mines, a mill, smelter, matte processing plant, nickel refinery and a cobalt/precious metals refinery.

Clarabelle Mill is located in Sudbury, Ontario, Canada (approximately 350 km north of Toronto) on the southwest rim of the Sudbury basin, a meteoric impact crater that is 62 km long and 30 km wide. The result of the meteoric impact was concentrated nickel (Ni), copper (Cu) and platinum group metal (PGM) orebodies in the outer perimeter of the basin, described in more detail by Hanley (1957). Clarabelle Mill currently processes the 80 to 90 orebodies that are mined at Vale’s six operating mines in the Sudbury basin. Toll ores are also received from two KGHM (formerly QuadraFNX) mines at the mill.

The mill, commissioned in 1971, was originally built as one of four mills operated by then Inco Limited. Between the late 1970s and the early 1990s process changes were made to consolidate the Sudbury Operations milling processes into a single operating mill – Clarabelle Mill. In 1990, a semi-autogenous grinding (SAG) mill was installed, resulting in peak mill throughput of 11.9 million short tons per annum (Mst/a). In 1992, typical throughputs were ~10 Mst/a. Mill average feed grades through the 1990s and into the early 2000s were 1.4 per cent Cu and 1.2 per cent Ni. There are two flow sheets for Clarabelle Mill that are central to this paper. The current flotation flow sheet and the challenging ore recovery (CORe) flow sheet, which will be fully commissioned by spring 2013. The next two sections of the paper are descriptions of these flow sheets. Both flotation circuits are designed to create four product streams: Cu concentrate, Ni concentrate, Po tails and rock tails. Cu concentrate is sold to market. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrate is processed through the mill. Ni concentrates are sent to market.

Current circuit configuration
Clarabelle Mill receives run-of-mine ore by truck and railcar into the tipple bin, a subsurface storage bin. Clarabelle has a conventional crushing process as well as a SAG grinding circuit. The flow sheet for the crushing and grinding process is described by Bom et al (2009). Primary ball mill cyclone overflow reports to the flotation circuit (Figure 1).

The feed to the mill is a combination of three sulfide minerals: chalcopryite (Cp), pentlandite (Pn) and pyrrhotite (Po), with the balance being siliceous rock. Pyrrhotite is an iron sulfide floatable gangue, of which there are two crystal structures (monoclinic and hexagonal) present within the Sudbury ores, making it a more complex ore to deal with. Monoclinic Po is magnetic, and is essentially the only magnetic material in the ore, which historically drove the use of magnetic separators to separate this portion of the Po into a magnetic flotation feed.

¹ Senior Engineer Metallurgy, Vale Ltd, Copper Cliff, Ontario, Canada. Email: andrew.taylor@vale.com
² MAusIMM(CP), Manager – Mineral Separations Technology, Vale Ltd, Copper Cliff, Ontario, Canada. Email: virginia.lawson@vale.com
³ Section Leader – Mineral Separation Technology, Vale Ltd, Copper Cliff, Ontario, Canada. Email: renee.barrette@vale.com
Although Po is a gangue mineral, a portion is recovered to maintain the heat balance in the flash furnaces at the Copper Cliff Smelter. There are four distinct circuits in the flotation flow sheet as shown in Figure 1: the magnetic Po rejection circuit, the main flotation circuit, Cu separation circuit and the non-magnetic Po rejection circuit. The flow sheet is described in more detail by Taylor et al. (2012). Typically 20 per cent of the mill feed reports to the Magnetic Po rejection circuit, the remaining 80 per cent reports to the main flotation circuit feed.

Challenging ore recovery circuit configuration

During the 2007 and 2008 resources boom, plans were developed to increase the mill capacity at Clarabelle Mill in stages to a final capacity of 13.5 Mst/a. Vale’s life-of-mine plan (LOMP) for the Sudbury Operations identified ores for increased capacity. These ores were more complex, with an increased proportion of hexagonal pyrrhotite. Hexagonal pyrrhotite is non-magnetic and thus not recovered to the concentrate. A project was implemented to create a new flow sheet, allowing for the increased throughput and the processing of higher amounts of hexagonal pyrrhotite. The project, known as the Clarabelle Mill enhancement and recovery project (CMERP), was the base for the CORe flow sheet (Figure 2). The progression of the flotation circuits is described in detail by Lawson and Xu (2011).

The development of the CORe flow sheet is important to this paper because the engineering and construction schedule enabled the implementation of two of the projects discussed in this paper. Without the work performed in the early stages of development of the CORe project, the engineering and design required for A Cleaning and the Interim IsaMill™ circuit would have increased. Increased design and engineering would increase schedule and costs nullifying the benefits of the projects.

A CLEANING

In February 2011, the failure of a tapping block on the northeast wall of the #2 Flash Furnace (one of only two furnaces) at the Copper Cliff Smelter resulted in a period of 16 weeks of reduced nickel production. The expected resultant nickel lost was estimated to be five per cent of annual Vale production, (Barrette et al, 2012). In order to mitigate the production loss a series of actions were taken to maximise the nickel production from the mines and from Clarabelle Mill. With the total furnace throughput cut in half, a plan was developed to increase the nickel concentrate grade feeding the smelter. The fastest way to increase the nickel content in the nickel concentrate was to increase the mill feed nickel grade. To use ore blending to maintain high-concentrate grade, plant data were analysed to understand the mill feed grade versus concentrate grade and recovery trade-offs. The optimum feed grade was determined to be higher than 1.3 per cent Ni to maintain Ni concentrate grades of 16 per cent equivalent nickel reporting to the smelter. Reducing the pyrrhotite level in feed too low would simply replace pyrrhotite in the concentrate with rock that has zero nickel value. Consequently, the Po:Ni ratio needed to be controlled to ensure that dilution of the concentrate with pyrrhotite was minimised. Equivalent nickel grade is the Ni grade of the concentrate with the COP mathematically removed.

$$\text{EqNi} = \frac{100\%\text{Ni}}{(100 - (\%\text{Cu} \times 2.886))}$$

As an example, a concentrate grade that is ten per cent Ni and 11 per cent Cu has an EqNi of 14.6, so does a concentrate that has nine per cent Ni and 13 per cent Cu.

Based on these criteria, a new mine plan was created and the workforce was reassigned across Vale’s Sudbury mines to mine designated orebodies to create a blended ore, which met the division’s requirements for the smelter. The change in ore blend resulted in an immediate increase in the mill concentrate grade. With essentially no change in flotation circuit parameters the concentrate grade was increased by two equivalent nickel points from 14 to 16 due to the higher feed grades of good quality ore.

The second change to increase the Ni content in the Ni concentrate was to improve the rejection of entrained gangue from the Ni concentrate. This was done in two ways:

1. The first change was a modification of the frother control strategy to the main roughers, and the second was the A Cleaning flow sheet modification. The frother addition rate was initially tied to the plant feed tonnage and not based on a target rougher concentrate grade. Changing this control philosophy resulted in a reduction in the variability of the frother addition rate and thus concentrate grade.

2. The flow sheet modification to clean the rougher concentrate required capital funding and time to design, construct and commission the new circuit. The decision to implement cleaning of the primary rougher concentrate was made within two weeks of the furnace failure. The CORe project team had already designed an interim flow sheet that converted the required cells for A Cleaning from their current rougher/scavenger duty to a primary cleaner duty. At the time of this change CORe was in
The pipe route was a simple gravity and direct run with the magnetic re-cleaner capacity exceeded current plant on the following assumptions:

- The material would have another opportunity to redirect the rougher cleaner tail into the magnetic circuit Po re-cleaner cells. The rationale for this decision was based shown in Figure 3. To enable the construction to proceed rapidly, all the process decisions were made within a week of the decision to make the flow sheet change. Process options were considered using an empirical model in a Microsoft Excel simulation tool, based on existing plant data and CORE flow sheet design data. One major concern related to the potential of high Ni losses from the A Cleaner tail if the stream was incorrectly configured. This was mitigated by redirecting the rougher cleaner tail into the magnetic circuit Po re-cleaner cells. The rationale for this decision was based on the following assumptions:
  - The material would have another opportunity to float before seeing an exit stream.
  - The magnetic re-cleaner capacity exceeded current plant requirements.
  - The pipe route was a simple gravity and direct run with no need for pumping.

![FIG 3 - Clarabelle mill flow sheet post A Cleaning.](image)

**Process results**

The flow sheet change was commissioned on schedule, four weeks after construction mobilisation, and the improvement was immediate. Figure 4 shows that Stage 1 of the project (the process manipulations) did not shift the variability in the percentage of rock reporting to the Ni concentrate, but did shift the amount of entrained rock reporting on average from ~13 per cent to just over 11 per cent. In Stage 2, the addition of the rougher cleaning stage made further reductions from 11 per cent to ~9 per cent in the overall Rk recoveries to Ni concentrate, and reduced the variability of the Rk reporting to the smelter.

**Lessons learned**

There were three key learnings from this project:

1. The highest technical risk to the project was the pumping system. The time line did not allow for the procurement of pumps specified for the pumping system. Two decommissioned pumps from the plant had been identified as suitable for the service, based on pump curves and an estimated dynamic head of the piping. The risk was not communicated well to the plant personnel and there was a general perception that bad engineering design was the reason that the pumps failed to perform as planned. Better communication with plant personnel about the projects at the commissioning stage was deemed to be required.
2. A second learning was around typical standard engineering materials used at the mill. Clarabelle Mill uses rubber lined pipe for all slurry piping for wear protection. Due to the urgency of the project the field run piping was measured and fabricated in several days then sent out for rubber lining. The initial measurements did not take into account the rubber overlap of the pipe ends. This resulted in sections of pipe having to be fabricated on-site and field fit without rubber lining. The learning from this was that a single spool piece from any section of line should be left to fabricate as a final field fit in the piping run.
3. During the commissioning process, troubleshooting was ineffective because the process information was not being collected by the data historian used at Clarabelle Mill, called PI from OSIsoft. There were also some misunderstandings as to the specific goals of the commissioning and ramp-up and what data was necessary to show the success of the project. A formalised system has now been developed to prevent this type of occurrence at Clarabelle Mill.

**Successes**

Within days of the furnace failure a plan to mitigate losses from the Ontario division was established. The initial change to the operating control was completed within two days of the failure. Within one week the plan to move forward with the circuit reconfiguration was made. It took less than a month to get all approvals, tender the work and select a construction firm to perform the work. In just over six weeks the contractor received the approval, mobilised the workforce and completed construction of the new circuit. The time line presented in Figure 5 shows the nine-week span for the project.

The rejection of entrained gangue that was realised through both the process control changes and the addition of rougher cleaning, allowed the mill to supply the smelter with higher-grade Ni in concentrate with minor to no Ni losses. The changes allowed for the Ontario Operations to put in effect a mine plan to produce ore and produce finished Ni at an acceptable rate for the company’s customers.

Since the completion of the rebuild of #2 Flash Furnace, the continued operation of rougher cleaning has allowed for
less entrained rock in Ni concentrate reporting to the smelter, which increases the energy efficiency of the smelting process.

**INCREASED COPPER SEPARATION**

Clarabelle Mill’s execution plan for 2011 included a 50 per cent reduction in the variability in the Ni concentrate quality targets, Cu:Ni ratio, equivalent Ni grade and Po grade, as previously explained by Barrette et al (2012). At the same time the feed to the mill Cu:Ni ratio was increasing, which, in turn, was causing higher variability in the daily Cu:Ni ratio. The average feed Cu:Ni ratio increased from 1.1 to 1.5 in the last five years. As the copper removal capacity is bottlenecked by the floatation columns, an increase in feed Cu:Ni ratio at a fixed copper removal resulted in an increase in Cu:Ni ratio in the Ni concentrate. To remain on spec for Cu:Ni ratio in the Ni concentrate, the feed tonnage was reduced. A fundamental requirement for a quality organisation is the ability to manipulate feed metal units (using mill throughput) in order to maintain concentrate quality to the customer. This is shown in Figure 6, taken from Taylor et al (2012), where the mill tonnage capacity is plotted for varying Cu:Ni ratios in the feed at several Cu:Ni ratios in Ni concentrate with varying Cu concentration production capability. An example is that for a Cu:Ni ratio of 1.5 in the feed if the Cu concentrate tonnage constraint is moved from 650 to 800 st/h (590 to 725 mt/h) the potential mill throughput to remain on-spec increases from 21 000 st/d to 26 000 st/d for a Ni concentrate Cu:Ni ratio of 0.5. This represents a significant increase in value generation from the mill for the 2011/2012 mine plan.

![Mill throughput capability based on feed quality.](image)

**Advance planning and engineering**

The Cu Separation circuit at Clarabelle Mill was commissioned in 2006. The engineering design for the building, services and equipment allowed for a future expansion of the capacity, if needed. The original design capacity for commissioning in 2006 was 150 000 Mt/a of Cu concentrate or 485 Mt/d. Maximum design capacity with the expansion to four columns was estimated at 370 000 Mt/a. The circuit has successfully operated at up to 590 Mt/d capacity since commissioning. In 2009 and 2010, the Larox filter capacities were expanded by 40 per cent by installing four extra plates to each filter, as per the original design, which would allow for the full design throughput to be filtered. With the expansion of the filters completed the only limitation to throughput was flotation capacity. As the column surface area was the bottleneck the only way to increase production from the circuit was to install more flotation capacity.

In July 2008, an agreement between Xstrata Technology and Clarabelle Mill was reached for the rental of a Z1600 Jameson Cell test rig. The Z1600 is a single downcomer production sized Jameson Cell, which the Mineral Separation Technology group was going to use for research on several plant streams. Although the Z1600 is a semi-mobile unit, it could not be easily moved throughout the plant due to space limitations. Prior to the selection of the optimal Jameson Cell location, a test work was created; flotation capacity in Cu separation was one of the scenarios. The location and detailed engineering design for the installation took into account these factors so that continued test work could be performed with minimal future engineering requirements.

In December 2010, proposals for increasing Cu column capacity were investigated. The review included examining all current project studies completed, and identified several additional options. The options and details on the analysis are summarised by Taylor et al (2012). Based on the analysis, it became clear that the conversion of the Jameson Cell was low-risk, low-capital and quick to implement. The total project flow sheet included the addition of the Jameson Cell and the ability to bypass concentrate from the first Cu scavenger cell to Cu concentrate. The proposed modified flow sheet is shown in Figure 7.

Based on the project time line requirements, as well as the production forecast for Q3 2011 and into 2012, the project...
required rapid movement through the stages for a capital project. Figure 8 shows a high-level Gantt chart of the project time line. The project was 54 in duration from concept to commissioning. The construction phase, which was 18 weeks long, was driven by the delivery time of the control valves and instrumentation required to convert the Jameson Cell from a manually operated test cell to an automated cell.

**Process results**

The requirement for the removal of Cu from the circuit was changed late in the construction phase by a change in Vale’s mine production forecast for 2012. The increased Cu production was no longer part of the 2012 mining plan. This change impacted the commissioning of the increased Cu circuit time line due to a lack of Cu units entering the plant. The lack of Cu units was detrimental to commissioning as the lower Cu units would affect the quality of both the Ni and Cu concentrates.

The Jameson Cell was operated for metallurgical ramp-up at intervals when the Cu head-grade was high enough to run the new process. The current average production rate from the two 3.8 m SGS Minnovex columns is 590 Mt/d; the Z1600 shows a maximum capacity of approximately 112 - 115 Mt/d (based on mass pull surveys). Based on these numbers the columns have a mass pull rate of 1.19 Mt/h/m², while the Jameson Cell has been able to pull 2.31 Mt/h/m².

Metallurgically the Jameson Cell performed on par with the columns in its ability to make Cu concentrate that was within customer specification. Figure 9 shows that both the Jameson Cell, and the columns, are operating with the same upgrade capability for Cu and rejecting Ni in the same capacity. Figure 10 is a plot of the mass pull of concentrate in Mt/m²/h, calculated for both the Jameson Cell and the columns. The data from the columns is for 207 operating days from October 2011 through to the beginning of May 2012. The data for the Jameson Cell is based on 13 operating days where surveys were taken for analysis. Roughly 57 per cent of the time the columns operated at or above 1.1 Mt/m²/h, while the Jameson Cell has operated 83 per cent of the time at or above the 1.1 Mt/m²/h rate. The Jameson Cell reached a higher maximum mass pull rate than the columns: 2.44 t/m²/h versus 1.43 Mt/m²/h respectively.

**Lessons learned**

The two key learnings that came from rapid implementation of this project include:

1. Several flush points and clean-outs were missed in design of the pipelines. The lack of ability to flush and dump the line resulted in several plugged lines, at a low point and upstream of an isolation valve. More regular field walkthroughs with operations personnel during construction would most likely have identified the appropriate locations based on the field run.

2. The airflow control valve and flowmeter were not of appropriate size or configuration for the naturally aspirated Jameson Cell. The undersized instrumentation was not apparent until the installation of the control valve and flowmeter was complete. This would have been rectified prior to installation if Xstrata Technology had been contacted for input into the instrumentation installed on the Jameson Cell. More input from vendors and the

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**FIG 8 - Timeline for the increased Cu project.**
Vale Instrumentation group is required moving forward to ensure that all components installed in new systems are compatible and will function properly within the mill’s process network infrastructure.

Successes
The main purpose of the installation was to increase the plant’s ability to maintain Ni concentrate grade specifications at maximum tonnage milled when Cu feed grades increased. The circuit has shown its capability to perform that duty, even though its required runtime is low. It will be used as a swing system, as required, being able to be started and shut down with relative ease and with little upset to the overall plant operation. Having the ability to increase Cu separation capacity means that the mill operations can react to fluctuations in the Cu: Ni ratio of the feed in a timely manner and not reduce throughput of the mill to maintain the product quality.

Taking from the lessons learned in the A Cleaning project, during the field run sections of pipe, key spool pieces were not measured for fabrication until the majority of the piping was completed. Although the construction phase was still rapid the implementation of this fabrication phase caused less re-work and construction time on the piping than was required in the A Cleaning project.

ISAMILL™ INTERIM CIRCUIT
In 2007, based on the Ni market, the #8 ball mill was converted from scavenger regrind service to a primary ball mill for increased throughput capacity. In 2010, the #11 regrind ball mill for B Rougher and scavenger concentrates had a catastrophic structural failure, which rendered it unusable. Both of these regrind circuits were in the non-magnetic pyrrhotite rejection circuit. The recovery opportunity, by replacing both regrind mills with an Isamill™, was reviewed by the plant technical team. A plan to install an Isamill™ to replace the 2250 kW feeding the B Cleaners of 25 μm, it was estimated that the P_80 from the Isamill™ would need to be micron, with a Isamill™ of P_80 of ~40 μm required to attain this desired feed size. Although there was variation, for the 12 surveys completed, the P_80 for B Cleaner feed and Isamill™ discharge were 27 μm and 46 μm respectively (Figure 11).

Opportunity arises
As the Isamill™ is new technology to Clarabelle Mill and to Vale, it was decided to start the Isamill™ in an interim circuit. The ability to restate regrind capability to the B Cleaner feed was calculated to increase Ni recoveries by ~0.9 per cent by liberating Pn/Po binaries. A further benefit was that plant personnel could gain an understanding of the Isamill™ operation prior to CORe commissioning. This included the training of operators on optimum mill operation, training mechanics on maintenance and training the instrumentation and electrical departments on the new equipment.

Design modifications
The work required to allow the Isamill™ to operate in the interim B Rougher service, compared to its final Po regrind service in the CORe flow sheet, included:

- the modification of the last 50 ft of eight-inch discharge piping and two-inch lime addition piping to redirect the Isamill™ discharge and lime addition to a different pump box than will be used in the CORe flow sheet
- rebuilding the interim circuit pump box and associated pumps
- rebuilding of the cyclone pack associated with the mill earlier than originally planned

- installing pH probes into the B Cleaner cells to control the lime addition flow
- ensuring that all existing piping to be recommissioned was fit for use.

Process results
Low-plant flows resulted in higher levels of recycle to the mill feed and the increased discharge temperature was a concern. By reducing the setpoint on the mill power draw, the discharge temperature was maintained in an acceptable range.

The initial configuration of the cyclones was not appropriate for the low flow rates in the circuit. A new apex and vortex finder combination was calculated and then installed. A control philosophy to reduce pressure variation in the cyclopack was created and implemented.

Overall, the performance of the Isamill™ met design requirements. The target grind size for the circuit was a P_80 = ~40 μm required to attain this desired feed size. Although there was variation, for the 12 surveys completed, the P_80 for B Cleaner feed and Isamill™ discharge were 27 μm and 46 μm respectively (Figure 11).

Lessons learned
There are two key learnings from this project:

1. The implementation of the Interim Isamill™ circuit was a request of the CORe project team, by plant operations personnel. This required interaction of plant maintenance with an engineering procurement and construction management project. Planning work in the same area, using the same equipment was complex and at times became a scapegoat for both the project and plant for missed deadlines. Integrating the project team into the plant planning and scheduling system helped avoid the situation.

2. The Interim Isamill™ circuit was not part of the original tender package to the construction contractor. Because the Interim Isamill™ was essentially replacing #11 Mill, some of the existing piping that would eventually be modified needed to remain intact. Unfortunately, some of the piping changes that were made were required for the interim circuit. The changes were caught early enough to be corrected before commissioning. There is another interim circuit that will be implemented; closer discussion with the contractor will be required to ensure that this is not repeated.

Successes
The Interim Isamill™ circuit was designed and implemented with minimal issues as part of a major capital project. In
general the operating and control strategy worked as expected and where it did not changes were implemented in a timely manner to help facilitate the continued commissioning of the IsaMill™.

Table 1 shows the calculated operating work index (Wio), as calculated by Bond’s law of comminution. The data used for #11 Mill was from a detailed survey performed in 2004, where size by size analysis was completed. The table shows that although the current feed to the circuit has a finer distribution, the Wio during both commissioning runs was higher than when #11 Mill was operating. During commissioning, the IsaMill™ did not reach the initial calculated Wio of 8.5, which is attributed to the lack of feed to the IsaMill™ causing recirculation of product sized material to the feed.

<table>
<thead>
<tr>
<th></th>
<th>Average power draw (KW)</th>
<th>Average (t/h)</th>
<th>f80</th>
<th>P80</th>
<th>Wio</th>
</tr>
</thead>
<tbody>
<tr>
<td>#11 Mill (2004 data)</td>
<td>401</td>
<td>75</td>
<td>110</td>
<td>52</td>
<td>12.5</td>
</tr>
<tr>
<td>Commissioning run 1</td>
<td>659</td>
<td>62</td>
<td>92</td>
<td>20</td>
<td>9.6</td>
</tr>
<tr>
<td>Commissioning run 2</td>
<td>453</td>
<td>78</td>
<td>87</td>
<td>34</td>
<td>10.2</td>
</tr>
</tbody>
</table>

CONCLUSIONS
The A Cleaning project was driven by an emergency with necessary changes required to keep Ni losses to a minimum, while maximising the divisional throughput in a time when the smelter was the bottleneck due to the furnace failure. Process engineering was performed early on, but the smelter was the bottleneck due to the furnace failure. While maximising the divisional throughput in a time when necessary changes required to keep Ni losses to a minimum, the A Cleaning project was driven by an emergency with recycled material. Although the current feed to the circuit has a finer distribution, the Wio during both commissioning runs was higher than when #11 Mill was operating. During commissioning, the IsaMill™ did not reach the initial calculated Wio of 8.5, which is attributed to the lack of feed to the IsaMill™ causing recirculation of product sized material to the feed.

TABLE 1
Operating work index.

ACKNOWLEDGEMENTS
The authors would like to thank the Vale Canada Limited for permission to publish this paper. Moreover, the support and contribution of all personnel at Clarabelle Mill, the Mineral Separation Technology Group, the CORe Project and Vale Base Metals Research and Development should be recognised as instrumental in making these projects successful.

REFERENCES
Telfer Processing Plant Upgrade – The Implementation of Additional Cleaning Capacity and the Regrinding of Copper and Pyrite Concentrates

D R Seaman¹, F Burns², B Adamson³, B A Seaman⁴ and P Manton⁵

ABSTRACT

The Telfer concentrator, located in the Great Sandy Desert of Western Australia, consists of a dual train gold/copper operation processing ore from one underground and, currently, two open pit mines with differing mineralogy. The flotation circuit of each train was designed to operate in several modes depending on the feed mineralogy. The majority of ore mined at Telfer is processed in a sequential mode where copper minerals are first floated into a saleable copper concentrate followed by the flotation of an auriferous pyrite concentrate which is treated in an on-site hydrometallurgical plant (carbon-in-leach (CIL)). Gold is recovered as a gravity product within the primary grinding circuit, to the copper concentrate, and to a lesser extent, the CIL circuit.

Since Telfer was re-opened, with a new concentrator, in 2004, the processing plant has struggled with poor copper concentrate grades, partially due to the excessive entrainment of non-sulfide gangue minerals in the copper flotation circuit and, more recently, due to composite copper particles produced when processing ore from a supplementary satellite pit that has not previously been processed through the new Telfer concentrator. Gold recoveries in the CIL circuit have also been below industry standard.

This paper presents the implementation of recent changes made to the circuit to address these performance issues. The reconfiguration of the circuit has involved the installation of the following major equipment items: two ISAMills™ in ultra-fine grinding applications (one in the copper circuit and one in the pyrite circuit), two Jameson Cells to improve fine gangue rejection and a bank of 5 × Outotec TC30s to recovery copper and gold from the reground pyrite stream. The equipment was purchased direct from vendors and an engineering firm contracted to design and install the multi-vendor reconfiguration upgrade.

INTRODUCTION

Telfer is a gold/copper operation located in the Pilbara region of Western Australia. Open pit mining (Main Dome) recommenced in 2003, followed by an underground mine (Telfer Deeps) in mid-2006. Copper deportment in the ore varies significantly from mainly chalcocite in the open pit ore to predominately chalcopyrite in the underground ore. In both ore sources, gold is present as free gold and granularly locked in copper sulfides and pyrite.

The ore processing plant consists of two parallel trains, Train 1 and Train 2, which are currently treating a total of 21 Mt/a. This includes approximately 6 Mt from the underground mine. Train 1 receives a blend of the underground and open pit ores, while Train 2 treats open pit ore alone. Details of the mine geological and ore mineralogical information, the initial process plant design criteria and operating strategies as well as a summary of the commissioning phase, can be found in previous publications by Goulsbra et al (2003) and Benson et al (2007).

Ore is processed through both trains in a variety of configurations. The predominant configuration is sequential flotation, where copper bearing minerals are recovered to a saleable copper concentrate followed by re-activation and flotation of the pyrite, which is leached with cyanide to recover gold. Much of the free gold that is not recovered in the gravity stage during grinding is recovered to the copper concentrate.

More recently, Telfer has been processing ores from a supplementary satellite deposit, namely West Dome. Whereas, the Main Dome, Telfer’s primary ore source, is generally well liberated following grinding to a nominal target P₈₀ of 120 μm, the mineralogy of the West Dome ore differs significantly.
particular, the West Dome ore has a significantly higher sulfur content and the copper minerals tend to present as rimming around or veining through pyrite.

Zheng et al (2010) reported an initial change made to the Telfer processing plant to alleviate overloading of the Train 1 copper cleaner circuit and also presented the foundations of the reconfiguration that has taken place at the Telfer plant. This paper outlines three significant changes made to the processing plant over the last 12 months. These changes have been implemented in order to improve the metallurgical performance of the Telfer plant. These changes are summarised in Table 1.

**TABLE 1**

Telfer reconfiguration strategies to address different factors affecting plant performance.

<table>
<thead>
<tr>
<th>Factor/opportunity</th>
<th>Process plant reconfiguration</th>
</tr>
</thead>
<tbody>
<tr>
<td>High proportion of liberated, non-sulfide (non-value) gangue content in the copper concentrate stream preventing recovery of additional auriferous pyrite and gold containing composite particles</td>
<td>Cleaner scalper flotation – Two (E3432/8) Jameson cells in a cleaner-scalper configuration in front of the pre-existing two stage cleaning circuit</td>
</tr>
<tr>
<td>Poor copper concentrate grade when processing high proportions of supplementary, West Dome, ore – due to lower liberation of copper sulfides as compared to the primary ore source, Main Dome.</td>
<td>Copper regrind – A copper regrind mill and preclassification circuit for the regrinding of copper rougher concentrate (ISAMill M5000 – 1.1 MW)</td>
</tr>
<tr>
<td>Below industry standard recovery and high incremental operating cost of gold recovery from pyrite leach (carbon-in-leach) circuit.</td>
<td>Pyrite Regrind and Recleaning – A pyrite regrind mill and recleaner flotation cells to liberate and recover gold and copper from Pyrite concentrate prior to cyanidation. (ISAMill M5000 – 1.5 MW plus 5 × Outotec 30m3 tank cells with high shear stators).</td>
</tr>
</tbody>
</table>

Figure 1 and Figure 2 show the Telfer flow sheet pre- and post- the reconfigurations discussed in this article.

At the time of writing, the first modification (installation of Jameson cells as cleaner scalpers) had already been implemented and commissioned, with construction of the second two items well advanced and commissioning expected to take place within months of authoring this paper. Figure 3 shows a 3D model of the major equipment layout for the reconfiguration project. The equipment was kept to one area for ease of maintenance and also to minimise cost and disturbance to the running operation during construction and commissioning. The regrind mills share a common platform with a 10 t gantry crane overhead for ease of maintenance.

The major equipment was purchased directly from vendors by Newcrest Mining to reduce the time frame of the installation, and to allow a staged process of capital commitment during the project development phase. Process design and major equipment sizing was carried out by Newcrest Mining personnel (authors of this paper), and a third party engineering firm (GR Engineering Services Limited) was contracted under a lump sum EPC to complete the installation of the equipment. Xstrata Technology supplied a vendor package containing the pyrite regrind mill (the copper mill was purchased from a third party as a second-hand unused mill), the mill platform to support both mills, feed and discharge hoppers, media handling systems and all associated instrumentation and steel work. ISAMill™ technology was chosen for the regrind duty due to their proven energy efficiency and the inert grinding environment which prevents passivation of the sulphide surfaces (Pease et al, 2006). Several of the improvements outlined by Rule and de Waal (2011) were incorporated into the ISAMill configuration). Outotec supplied the flotation cells (5 × OT30s) used in the pyrite regrind circuit which have been fitted with Outotec’s float force mechanism (Coleman and Rinne, 2011) as well as high shear stators (Bilney, MacKinnon and Kok, 2006) to optimise the hydrodynamic conditions for fine particle flotation.

The on-site construction period will total approximately 12 months by completion of all three stages, with initial equipment orders placed approximately six months prior to
construction commencement. The final capital approval for the entire reconfiguration project was granted in July 2011, and commissioning of all equipment is due to be completed by mid-August 2012.

**Cleaner scalper flotation**

Based on a review of historical copper concentrate data, Seaman, Manton and Griffin (2011) showed that a large portion of liberated, non-sulfide non-floating gangue material was being recovered to the copper concentrate via entrainment.

Zheng, Crawford and Manton (2009) presented details on how a reconfiguration of Train 1 was completed in 2009 to assist the rejection of some of this gangue and to debottleneck the cleaning circuit. While this modification was successful, further improvements were identified and implemented as described by Seaman, Manton and Griffin (2011).

The new Jameson cells installed in a cleaner scalper duty were commissioned in November 2011. Each Jameson cell was installed to allow gravity discharge of tailings and concentrate to the existing plant, which meant the cells were installed on a steel structure 15 m off the ground, with the recirculation pumps located on ground level for ease of maintenance. The additional cost of elevating the cells was offset by lower operating and maintenance costs than if tailings and concentrate pumps and hoppers had been required. The cells have eight downcomers each, and are each driven by a 75 kW Warman 10/8 pump with a recirculating slurry flow rate of approximately 700 m$^3$/h. The fresh feed rate to each Jameson cell is in the order of 175 - 350 m$^3$/h (or approximately 20 to 60 t/h solids). The washwater system was designed to achieve a flow rate of up to 100 m$^3$/h per cell.

Jameson cells described by Evans, Atkinson and Jameson (1995) are highly efficient flotation machines that require a smaller footprint than conventional mechanical flotation cells and enable the efficient use of froth washing to improve gangue rejection. A schematic figure of the latest Jameson cell technology was presented by Young et al (2006).

Figure 4 shows a photograph of the installation. This stage of the Telfer Reconfiguration Project was initiated in
October 2010, major equipment order was placed in May 2011, and the cells were commissioned in November 2011.

**Performance of cleaner-scalper flotation circuit**

The benefit of the Jameson cell installation was derived from rejection of non-sulfide gangue allowing the recovery of slower floating valuable minerals (be they composites or liberated fines) and also the potential to replace liberated non-sulfide gangue with auriferous pyrite. Dilute batch flotation tests were conducted as part of the project development phase and the resulting selectivity curves used in a flotation model to predict the ultimate performance of the plant (Seaman, Manton and Griffin, 2011). It was assumed that a copper recovery of at least 50 per cent could be achieved across the Jameson cells.

Figure 5 shows the selectivity derived in the batch flotation tests compared with actual plant operating points generated from spot samples collected for four months following the Jameson cell commissioning.

It can be seen in Figure 5 that the actual operation of the Jameson cells matched the predicted test data reasonably well on Train 1, and with potentially poorer performance on Train 2 for the majority of the time. In addition, the stage recovery achieved by the cells, in most cases, well exceeds the expected 50 per cent metal recovery.

It is also clear from the operating data that at higher metal recoveries (over 80 per cent), the Jameson cells lose their selectivity. Thus, it is the ongoing focus of the Telfer operation to monitor and control this stage recovery below 80 per cent metal recovery.

Prior to installation of the Jameson cells, the copper recleaners were heavily loaded, and froth often built-up to a point where it would overflow the cells as shown in Figure 6. This phenomena has ceased since the Jameson cells have been commissioned as there is now much less floatable material reporting to the mechanical cells.

The Jameson cells themselves have proved to be fairly simple to operate, with little operator intervention required in terms changing operating conditions. Downcomer blockages are an ongoing problem with the cells, caused by scale in upstream pumps and tanks.

**Assay by size**

An assay-by-size survey and mass balance was conducted on the Train 2 cleaning circuit to investigate the performance of this circuit by size.

![FIG 5](Image5.png)

**FIG 5** - Mass versus copper selectivity for (A) Train 1 and (B) Train 2. Comparison of spot data (Nov 2011 - March 2012) with dilute batch flotation tests conducted as part of the project development phase.

![FIG 6](Image6.png)

**FIG 6** - Photograph of overloaded Train 2 reCleaners prior to Jameson cell installation.

![FIG 7](Image7.png)

**FIG 7** - Size by recovery across the Jameson cell and estimated non-sulfide gangue (NSG) content per size fraction of the Jameson cell concentrate and conventional Recleaner concentrate streams as a function of screen/cyclosize fraction (NSG, estimated from assay data).
Figure 7 shows the recovery by size fraction of different elements across the Jameson cell as well as the gangue content by size fraction of the Jameson cell (cleaner-scalper) and (conventional) recleaner concentrate.

It can be seen that the valuable (gold and copper) recoveries across the Jameson cell are particularly good across all size fractions, with some drop off on the coarse and fine ends. The NSG content by size demonstrates that the Jameson cells are rejecting NSG far better than the mechanical recleaner cells due to the addition of wash-water in these cells. Trials of wash-water addition to the mechanical recleaner cells have demonstrated that further NSG rejection is possible. The permanent extension of wash water addition to these cells is expected to be completed within months of authoring this paper.

**Cleaner block survey summary**

Metallurgical surveys were conducted in the two month period leading up to the Jameson cell commissioning. The data collected was mass balanced and cleaner performance data for the surveys conducted are shown below in Table 2.

It can be seen that, prior to Jameson cell installation, the cleaner block performance is variable and averages below 90 per cent for both trains – copper and gold. The lower iron and sulfur recoveries are a result of deliberate pyrite rejection in the cleaning circuit.

Table 3 shows a summary of cleaner block recoveries determined post installation of the Jameson cells. Note that in the case of Train 1, the Ro Con A stream is now included in the cleaner block, where previously it by-passed the cleaner block.

Comparing the pre- and post- Jameson cell cleaner block performance, it is clear that the overall cleaner block performance has significantly improved, with gold and copper cleaner recoveries in excess of 95 per cent.

This improved cleaner block performance was also observed in the cleaner scavenger tail grades which have significantly reduced on both trains as shown in Table 4.

Two months of operating data before and after installation was analysed to determine the circuit recovery improvement resulting from the installation of the cells. After taking into account known factors that affected recovery during this time.

### Table 2

<table>
<thead>
<tr>
<th>Cleaner recovery</th>
<th>Mass Cu</th>
<th>Au</th>
<th>Fe</th>
<th>S</th>
</tr>
</thead>
<tbody>
<tr>
<td>11.3</td>
<td>76.2</td>
<td>86.8</td>
<td>27.2</td>
<td>31.3</td>
</tr>
<tr>
<td>31.9</td>
<td>90.3</td>
<td>89.0</td>
<td>50.7</td>
<td>46.7</td>
</tr>
<tr>
<td>21.7</td>
<td>89.4</td>
<td>88.9</td>
<td>62.3</td>
<td>69.4</td>
</tr>
<tr>
<td>24.1</td>
<td>93.2</td>
<td>94.6</td>
<td>57.6</td>
<td>74.1</td>
</tr>
<tr>
<td><strong>Average T1</strong></td>
<td><strong>22.2</strong></td>
<td><strong>87.3</strong></td>
<td><strong>89.8</strong></td>
<td><strong>49.4</strong></td>
</tr>
<tr>
<td>26.7</td>
<td>50.0</td>
<td>80.1</td>
<td>51.5</td>
<td>61.6</td>
</tr>
<tr>
<td>29.5</td>
<td>88.1</td>
<td>94.2</td>
<td>74.8</td>
<td>85.2</td>
</tr>
<tr>
<td><strong>Average T2</strong></td>
<td><strong>28.1</strong></td>
<td><strong>69.0</strong></td>
<td><strong>87.2</strong></td>
<td><strong>63.1</strong></td>
</tr>
</tbody>
</table>

a. Cleaner circuit block represented by copper rougher concentrate as feed, copper recleaner concentrate and copper cleaner scavenger tail as products.

### Table 3

<table>
<thead>
<tr>
<th>Cleaner recovery</th>
<th>Mass Cu</th>
<th>Au</th>
<th>Fe</th>
<th>S</th>
</tr>
</thead>
<tbody>
<tr>
<td>89.5</td>
<td>99.5</td>
<td>99.6</td>
<td>98.2</td>
<td>95.4</td>
</tr>
<tr>
<td>43.6</td>
<td>92.4</td>
<td>93.4</td>
<td>45.7</td>
<td>47.3</td>
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<td>25.1</td>
<td>95.2</td>
<td>95.3</td>
<td>55.0</td>
<td>57.3</td>
</tr>
<tr>
<td>11.0</td>
<td>87.4</td>
<td></td>
<td>27.1</td>
<td>37.9</td>
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<td>26.7</td>
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<td>47.2</td>
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<td>32.2</td>
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<td>99.4</td>
<td></td>
<td>84.0</td>
<td>91.5</td>
</tr>
<tr>
<td>64.0</td>
<td>99.6</td>
<td></td>
<td>84.2</td>
<td>93.0</td>
</tr>
<tr>
<td><strong>Average T1</strong></td>
<td><strong>52.7</strong></td>
<td><strong>95.7</strong></td>
<td><strong>96.1</strong></td>
<td><strong>66.3</strong></td>
</tr>
<tr>
<td>68.6</td>
<td>94.3</td>
<td>97.2</td>
<td>82.6</td>
<td>90.0</td>
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<tr>
<td>25.5</td>
<td>97.4</td>
<td>98.0</td>
<td>67.9</td>
<td>83.8</td>
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<tr>
<td><strong>Average T2</strong></td>
<td><strong>47.0</strong></td>
<td><strong>95.8</strong></td>
<td><strong>97.6</strong></td>
<td><strong>75.3</strong></td>
</tr>
</tbody>
</table>

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(feed grade, ore mineralogy, plant throughput, concentrate grade, etc), it was found that the cleaner scalper installation had a payback of between two and seven months.

**Improving copper concentrate grade**

Within the last 12 months Telfer has started supplementing its primary Main Dome ore source with West Dome ore while development work is being carried out in the Main Dome pit. Historical test work and plant trials have demonstrated that copper concentrate produced from West Dome ore typically does not achieve a suitable grade for sale. This is owing to a number of factors which include but are not limited to:

- higher pyrite content in West Dome ore
- copper mineralogy of West Dome is primarily as secondary copper sulfides (chalocite, bornite etc) of smaller grain sizes than Main Dome copper sulfide minerals
- on average, the copper sulfide minerals are less liberated than in the Main Dome ore.

Limiting the quantity of West Dome ore in blend has overcome some of these limiting factors, however there have been, and will be in the future, times when the quantity of West Dome ore in feed exceeds the tolerable amount in blend.

Figure 8 presents a comparison between the Main Dome copper re-cleaner copper sulfide mineral liberation by free surface for Train 1 and Train 2 across two quarters (Q1 and Q2 – quarter 1 and 2 of FY2010 respectively) with West Dome laboratory re-cleaner concentrates. The comparison clearly demonstrates the poorer liberation of West Dome copper sulfides. This is supported by full-scale copper rougher concentrate liberation data included in Table 5. From the mineralogical studies conducted to date, it is known that the copper sulfide minerals most often occur in association with pyrite, either as veins within the pyrite minerals or as rims on the pyrite surface. Figure 9 shows two optical images, typical, of the copper mineral/pyrite association observed within the West Dome material.

![Figure 9](image)

**FIG 9** - Examples of copper sulfide inclusions in pyrite host particles.

As a result of this liberation issue, it is not surprising that earlier attempts to improve copper concentrate grade via depression, selective collection or chelation (of copper activating ions in solution) have been largely unsuccessful.

Following some promising laboratory tests incorporating a regrind stage between copper roughing and cleaning, a pilot ISAMill™ M20 (see Figure 10) was operated at Telfer during...

---

**TABLE 4**

<table>
<thead>
<tr>
<th>% Cu in Cl scav tail</th>
<th>Au (g/t) in Cl scav tail</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Train 1</td>
</tr>
<tr>
<td>Before</td>
<td>0.46</td>
</tr>
<tr>
<td>After</td>
<td>0.39</td>
</tr>
<tr>
<td>Significance (%)</td>
<td>84</td>
</tr>
</tbody>
</table>

**TABLE 5**

Liberation characteristics of West Dome copper rougher concentrate, collected during a plant trial.

<table>
<thead>
<tr>
<th>Mineral</th>
<th>% liberated (&gt;95%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pyrite</td>
<td>96.6</td>
</tr>
<tr>
<td>Chalcopyrite</td>
<td>69.8</td>
</tr>
<tr>
<td>Other copper sulfides</td>
<td>64.9</td>
</tr>
<tr>
<td>Other minerals</td>
<td>97.3</td>
</tr>
</tbody>
</table>
Figure 2. Copper rougher concentrate will be fed to a cyclone. P80s were approximately 40 and 20 μm respectively. In this case, the regrind feed and discharge were pumped to the existing cleaning circuit. A large discharge hopper with air-spargers will be installed to allow the oxidation state of the slurry to become suitably oxidised for flotation following the reducing conditions inside the ISAMill™.

**Improving gold extraction from pyrite concentrate**

Historically, carbon-in-leach (CIL) performance at Telfer has been below industry standard, currently averaging 75.2 per cent gold extraction from gold contained in the pyrite concentrate. Burns *et al* (2012) present a detailed diagnostic analysis of the factors affecting CIL performance at Telfer. They found, after conducting size-by-size analyses, diagnostic leach tests and detailed mineralogical studies, that poor liberation of gold grains in coarse pyrite particles was the primary factor limiting leach performance of Telfer’s pyrite concentrate. In addition, the study found that the high levels of cyanide soluble copper (mainly as composite particles with pyrite) entering the CIL circuit was responsible for high reagent consumptions in this circuit.

Burns *et al* (2012) showed that the optimal approach to improving CIL performance (in terms of both improving recovery and reducing operating cost) was to install a regrind mill on the current CIL feed stream, and then remove liberated gold and copper particles by flotation (pyrite releaning) prior to leaching the flotation tailings stream. In laboratory and pilot testing, the overall gold extraction increased to approximately 90 per cent overall gold recovery together with increased copper recovery and a reduction of approximately 25 per cent in cyanide consumption in the CIL circuit as a result of reducing the cyanide soluble copper concentration in the CIL feed.

The improvement is gained from liberating fine gold grains (~5 - 10 μm) locked in larger pyrite particles. Much of this gold (~50 - 75 per cent) will be recovered by flotation and combined with the copper concentrate for sale, prior to the leaching of the flotation tails. Figure 12 shows some optical micrographs of the typical gold inclusions in pyrite prior to regrinding at Telfer.

This circuit configuration performance and scale-up design criteria were developed using a mixture of pilot-scale and laboratory techniques. A 1.5 kW pilot ISAMill™ M20 (see Figure 10) was used to regrind the CIL feed material to different grind sizes in preparation for bench scale flotation tests and bottle roll leach diagnostics. In addition to preparing the feed for laboratory testing, operation of the pilot mill also allowed for the generation of power requirement curves which were subsequently used to scale-up the power required in the full scale installation.

Figure 13 shows the gold flotation recovery following regrinding to different sizes during the pilot studies (each data series, shows a different day on which the tests were carried out. At target regrind size of 25 μm, the flotation recovery on some of the tests achieved close to the current gold recovery of the CIL circuit (~75 per cent). Operating costs of the regrind circuit are expected to be approximately ten per cent per ounce of the CIL operating costs. If the high gold flotation recoveries can be achieved consistently at full scale, this may provide an option to de-commission the CIL circuit in the future should the leach costs become uneconomical. It is unclear why the recovery of gold reduced greatly when grinding to P80s finer than 20 - 25 μm. The authors speculate that this drop-off could be due to overgrinding of higher sulfide gangue (SG) gold/
gold composites relative to lower SG barren pyrite as a result of the ISAMill™ internal classification. This will be further investigated once the circuit is operational.

A 1.5 MW ISAMill™ (M5000) is being installed to regrind the pyrite concentrate to a target P80 of 25 μm at a maximum design throughput of 90 t/h. The mill will be fed underflow from an existing two-stage deslime hydrocyclone circuit. The mill will discharge into an oversized hopper (approximately five minutes of residence time), where plant air will be sparged into the slurry to assist with increasing the dissolved oxygen level of the slurry to facilitate pyrite depression and enhance gold flotation. The slurry will then be diluted upon transfer to a bank of 5 × 30 m³ Outotec flotation cells – pyrite recleaners.

The flotation cells will be fitted with high shear stators (Bilney et al., 2006) and float force mechanisms (Coleman and Rinne, 2011) in an attempt to provide optimal flotation conditions for the fine particles. The circuit will have flexibility to send the pyrite recleaner concentrate to the final copper concentrate tanks or to the copper cleaning circuit if further pyrite rejection is warranted. The pyrite recleaner tail will be sent to the existing leach circuit for further gold extraction, and can also be sent to final tailings to by-pass the CIL circuit.

CONCLUSIONS

The first phase (installation of copper cleaner scalper, Jameson cells) of the latest reconfiguration of the Telfer processing plant has been completed successfully and in accordance with expected improvements. The second and third stages of the reconfiguration are well underway at the time of writing this paper, with commissioning to be completed prior to presentation of this paper at the conference.

The copper regrind mill will improve copper concentrate grade when processing West Dome ores by liberating copper minerals from pyrite/copper (mostly chalcopyrite and chalcocite) binary particles. This modification to the circuit is necessary as Telfer commence processing of West Dome ores.

The pyrite regrind circuit will improve gold extraction from Pyrite while decreasing the operating cost of the existing CIL circuit.

The project construction will be completed within 12 months of initial mobilisation to site and within 14 months of the final capital approval being granted by Newcrest Mining for all stages of the project.

ACKNOWLEDGEMENTS

The authors acknowledge the permission of Newcrest Mining Limited to publish this work. Many people at Telfer’s Gold Mine were involved in the laboratory studies, piloting and concept design/development, as well as in the actual implementation of the works at Telfer. In particular, Craig Chase-Dunlop has played a significant role in co-ordinating the construction activities on site, and the project has been expertly managed by Barrie Greensill. The authors are indebted to all those at Telfer who have made this project a success.

REFERENCES


IsaMill-1:1 Direct Scaleup from Ultrafine to Coarse Grinding

Larson M., Anderson G., Barns K., Villadolid V.  
Xstrata Technology  
307 Queen Street, Level 4  
Brisbane, QLD 4000 Australia  
mlarson@xstratatech.com

Abstract

The IsaMill has been used commercially in concentrator plants for over 15 years. Improvements in ceramic grinding media, mill design and wear components have advanced the IsaMill to the point where it can readily accept F
80’s of +300 microns. One thing that has not changed since the early days of development is the robust 1:1 scaleup of the mill from the laboratory to the mine site. This paper examines Xstrata Technology’s efforts to both improve the grinding capability of the IsaMill and the work that has gone into ensuring the accuracy and precision of independent laboratories across the world that perform IsaMill signature plot scaleup work. Common issues encountered in design testwork are discussed in an effort to promote proper scaleup among all suppliers.

Introduction

Each year independent mineral processing laboratories around the world perform over 100 signature plots to provide an energy scaleup number for Xstrata Technology’s IsaMill. 8 laboratories around the world are now certified every two years to ensure the reliability of their work and the robustness of their technique. This is a time consuming process to gather and split the concentrate sample, run replicate tests and ship out concentrate and media to the individual laboratories around the world. Once the testwork has been done any non-conforming laboratories must be inspected to determine any deviation from normal. The total cost of the program exceeds A$50,000. However, the results provide a level of confidence in the testwork being done that ensures every test can result in a process guarantee without Xstrata Technology being involved in the actual testwork. Through this partnership with independent laboratories improvements have been made to the signature plot procedure to reduce error wherever possible and the range of media sizes tested has increased with improved knowledge of the inner workings of the M4 4 litre IsaMill.

Section 1. IsaMill scaleup testwork

Historically IsaMill testwork was done in-house by MIM and then Xstrata Technology. However, as testwork is not a core business and the interest in signature plot testwork increased it became desirable to have external independent laboratories take on this workload. The testwork itself is relatively simple and repeatable as long as the standard procedure is followed. This repeatability increases with practice.
Figure 1 and Table 1 represent the repeatability possible by someone with limited experience running the M4 (Larson 2012). The margin of error for these five tests is about 3.2% from the average energy to 10 microns. Each individual error from average is shown in the table.

![Figure 1](image_url)

**Figure 1. Copper concentrate replicate test**

<table>
<thead>
<tr>
<th>Test</th>
<th>Energy to 10 microns (kWh/ton)</th>
<th>Error from mean (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Test 7</td>
<td>84.16</td>
<td>+5.91%</td>
</tr>
<tr>
<td>Test 8</td>
<td>77.44</td>
<td>-2.54%</td>
</tr>
<tr>
<td>Test 9</td>
<td>76.94</td>
<td>-3.17%</td>
</tr>
<tr>
<td>Test 10</td>
<td>77.62</td>
<td>-2.31%</td>
</tr>
<tr>
<td>Test 11</td>
<td>81.12</td>
<td>+2.09%</td>
</tr>
</tbody>
</table>

Mean 79.46 kWh/ton

Standard Deviation 3.11

Table 1. Copper concentrate replicate test energy to 10 microns

Figure 2 and Table 2 represents the repeatability of the first standard sample, with a margin of error of 6.8% from the average of the repeatability tests. This was done for the 2009/2010 standard sample.

In this round of testwork 7 different labs were certified. The criterion was to fall within 5% of the average energy of 42.8 kWh/t to a P80 of 15 microns from the XT replicate tests. Of the 7 laboratories certified, five were completed with their first try, one on their second and one on their third. In the case of the latter, there were issues with their Malvern laser sizer sub-sampling that was fixed after a visit to audit their procedure. The energies reported by the seven labs averaged 41.5 kWh/t.
Table 2. 2009/2010 standard sample replicate test target energy and error

<table>
<thead>
<tr>
<th></th>
<th>Test 1</th>
<th>Test 2</th>
<th>Test 3</th>
<th>Test 4</th>
<th>Test 5</th>
<th>Average</th>
</tr>
</thead>
<tbody>
<tr>
<td>Energy to 15 microns (kWh/t)</td>
<td>38.14</td>
<td>40.58</td>
<td>42.38</td>
<td>45.81</td>
<td>47.01</td>
<td>42.78</td>
</tr>
<tr>
<td>Error from average energy</td>
<td>10.9%</td>
<td>5.2%</td>
<td>.9%</td>
<td>7.1%</td>
<td>9.9%</td>
<td>6.8%</td>
</tr>
</tbody>
</table>

Figure 3 and Table 3 are the replicate tests for the most recent standard sample done in 2011/2012. This has a margin of error of 2.1% to 12 microns and is performed by an XT engineer with the most recent practice of anyone performing signature plot tests. In this case the standard for the labs to maintain certification was adjusted to +/- 5% of the average energy rather than the +/- 2.1% shown possible by the replicate tests.

Figure 3. 2011/2012 standard sample replicate test (Villadolid 2011)
<table>
<thead>
<tr>
<th>Test 1</th>
<th>Test 2</th>
<th>Test 3</th>
<th>Test 4</th>
<th>Test 5</th>
<th>Average</th>
</tr>
</thead>
<tbody>
<tr>
<td>35.71</td>
<td>35.12</td>
<td>33.15</td>
<td>35.15</td>
<td>35.71</td>
<td>34.97</td>
</tr>
<tr>
<td>2.1%</td>
<td>.4%</td>
<td>5.2%</td>
<td>.5%</td>
<td>2.1%</td>
<td>2.1%</td>
</tr>
</tbody>
</table>

Table 3. 2011/2012 standard sample replicate test target energy and error

Section 2. Common sources of error

The IsaMill signature plot test will have an average margin of error of about +/- 5% under ideal conditions to common IsaMill product sizes. Standard techniques have been set in an effort to minimize this error.

The energy meter will have an error of about +/- 1% according to the common manufacturers of energy meters in use. The slurry flow measurement will have an error of about .5 seconds, or 1-1.5% depending on flow. The density measurement will have an error of about 1%. The actual timing of each pass will have a margin of error of about 1%. As each pass requires about 6-7 minutes to complete, an error of 2-3 seconds on calling the tank empty is a minimal error. The remaining measurement error is due to any sizing or sampling inconsistencies in the Malvern laser sizer.

These errors are almost inevitable, improper technique will only serve to amplify them. Some errors encountered while doing laboratory audits include the following (this is not to single out any particular lab for criticism but to promote proper techniques among all labs).

2.1 Flow measurement

Improper flow measurement can greatly increase the margin of error of a test. Flow can either be measured with a 1 litre Marcy scale cup or with a suitably large graduated cylinder. With the Marcy scale cup it is relatively straight forward as the time is measured for slurry- not froth- to discharge from the holes marking the 1 litre point. In this case time/volume is actually being measured.

With a graduated cylinder time/volume cannot be used as the froth will obscure where the slurry actually reaches the line for the volume of interest. In the case of using a graduated cylinder to measure slurry flow the slurry must be directed into the cylinder for a given amount of time and then the level analysed when the sample has settled. In this case the measurement is in volume/time.

2.2 Density measurement

The density sample is taken in the middle of each pass. This requires that the feed sample is properly mixed to avoid feed density segregation between the beginning and end of the pass. For this reason all tanks used for IsaMill signature plots must have internal baffles and a mixer capable of suspending .5mm material of an SG of 4-5. Failure to use baffles can cause segregation and also centrifuging of the sample as it nears the end of the pass, starving the feed pump of slurry and artificially extending the pass and increasing the measured energy consumption.

The slurry density sample is never to be calculated by using the Marcy scale, but rather by a wet/dry mass. The Marcy scale can overestimate the slurry density if flow is allowed to continue into the cup even after overflowing. It can also underestimate the slurry density if material is allowed to spill and less than 1 litre of slurry is measured.
2.3 Sample sizing

The sizing of the individual pass samples can be wrought with errors that are sometimes hard to detect. For this reason Xstrata Technology has general guidelines for Malvern use during scaleup tests. All subsamples should be mixed in a baffled beaker and samples taken with an adjustable micro-pipette. Use of magnetic stirrers should be avoided as the magnetic bar can be interfered with by the sample pipette. Baffles must be used to prevent segregation between the middle and outside of the sample container. The adjustable pipette is needed to ensure that all slurry drawn is actually put into the Malvern. In the case of magnetite testwork screen sizing is preferred to avoid the possibility of magnetic agglomeration affecting the laser sizer results.

In one case a labs standard sample P80 energy was coming out higher than acceptable but the line for the P98 signature plot energy matched what would be expected from the replicate test, as did the mill net power draw. After examination it was determined that faulty sub-sampling for the Malvern was the cause. In this case the Malvern was located across the room from the sample mixer and a bulb pipette was used to take the Malvern sample. As the obscuration level of the Malvern determines how much sample is discharged, the combination of excess sample and segregation in the bulb resulted in the top size being segregated to the bottom and ultimately biasing the Malvern sample. By using an adjustable micro-pipette only the amount of slurry actually needed is sampled and all of it discharged into the Malvern, preventing any segregation biases.

The Malvern itself is basically a black box poorly understood by much of industry. Without proper care readings can be sporadic and inaccurate without the user knowing. There are specific settings for minerals but the one setting that is most commonly wrong is the absorption value. This can commonly range from .01-1 depending on the Malvern model. Xstrata Technology generally recommends that this setting be placed at .1 so that particles of .3-1 microns are measured and accounted for. By setting the absorption improperly sub 1 micron particles will not be measured. For most applications this may not be a major problem but for some IsaMill applications that size range can make up 10-20% of an ultrafine size distribution.

2.4 Viscosity

Excess viscosity can also have detrimental effects to energy efficiency if not fully understood and appreciated. Xstrata Technology has started to provide all new operating sites and M4 laboratories with a Marsh funnel. The Marsh funnel is a simple yet effective tool for quickly determining the viscosity of slurry without the need to stop a test to perform more complex rheology measurements. It consists of just a funnel and a one quart (.946 Litres) container. One quart of water takes about 28-30 seconds through the funnel.

Typically in a M4 signature plot test the net power draw will drop from pass to pass. This is simply an effect of the material being ground finer and thus being easier to mix. It was found over the course of several M4 and M20 test programs that a set of conditions existed where this power draw would start to increase. Comparing to Marsh funnel readings taken during these same passes it was found that this increase in power draw correlated to Marsh funnel times of about +38 seconds for the one quart of slurry to pass through the funnel. This time limit will change with media size, as more of a void space between larger media won’t be as sensitive to viscosity changes but this aspect has not yet been fully investigated.
The Marsh funnel is not a rheometer, because it only provides one measurement under one flow condition. However the effective viscosity can be determined from following simple formula.

\[ \mu = \rho \left( t - 25 \right) \]

where \( \mu \) = effective viscosity in centipoise
\( \rho \) = density in g/cm³
\( t \) = quart funnel time in seconds

This is by no means meant to be a precision measurement. It does though act as an invaluable tool when each pass through the IsaMill only allows for 6-7 minutes for all measurements and it must be decided quickly if water is to be added to dilute the next pass.

Each pass through the IsaMill is separately accounted for with an individual energy so each pass at a different density does not affect the others.

2.5 Sample mass and segregation

The standard for most materials is to provide 15 kg of dry solids for each M4 signature plot. The correct amount of solids is critical to ensure that steady state is reached and a representative discharge is sampled without coarse solids being segregated and held in the mill. This requirement will be the same for all stirred mill tests where a continuous discharge sample is collected. An example of this is shown below in Figure 5 by Gao testing different feed masses in a 40 litre pilot Tower Mill. The smaller sample mass shows less energy required to reach equivalent particle sizes. Although this result may look good it is not realistic compared to the test done with more sample. For this reason Xstrata Technology recommends that to ensure proper scaleup, testwork is done with 3-4 x solid volume than mill void volume. For common sulphide ores this results in 15 kg of total mass. For something like magnetite with a higher solid SG more material is required. If a very coarse product and low energy is desired more sample still may be required to ensure steady state is reached in the first pass and that there is enough time per pass for the operator to reliably take all measurements with the high flowrate. If screen sizings are to be done where more mass is removed during each pass more starting mass will be required.
Figure 5. Effect of varying feed solids mass in a pilot tower mill (Gao)

Not all stirred mill test programs follow these guidelines however. Published conditions below compared to the IsaMill signature plot standard show recommended solid volumes of less than half of that used to size an IsaMill and barely enough solids to equal the total mill void space. (Nippon Eirich 2009 and Rahal, Erasmus and Major 2011)

<table>
<thead>
<tr>
<th>Mill Type</th>
<th>Mill Open Volume(L)</th>
<th>Sample Mass</th>
<th>Solid Volume(L)</th>
<th>Ratio</th>
</tr>
</thead>
<tbody>
<tr>
<td>M4 IsaMill(4L)</td>
<td>1.35</td>
<td>15kg</td>
<td>5.00</td>
<td>3.70</td>
</tr>
<tr>
<td>Nippon-EirichNE008(8L)</td>
<td>~2.35</td>
<td>10kg</td>
<td>3.33</td>
<td>1.42</td>
</tr>
<tr>
<td>Nippon-EirichKM-5(120L)</td>
<td>~35.20</td>
<td>150kg</td>
<td>50.00</td>
<td>1.42</td>
</tr>
<tr>
<td>Knelson-Deswik10(10L)</td>
<td>~4.72</td>
<td>20kg</td>
<td>6.67</td>
<td>1.41</td>
</tr>
</tbody>
</table>

Table 4. Regrind mill test volume ratios

Section 3. Larger grinding media

One of the most significant advances in IsaMill technology over the last 5 years has not come from Xstrata Technology itself but from media suppliers around the world. The original UFG IsaMills were run on either sand or slag, a major limitation to the coarseness of the feed that can be processed. Even the first widely used ceramic media in an IsaMill was limited to an effective size of 3.5mm. While progressing up the feed coarseness scale, there were still limitations as to what this was capable of grinding. In recent years 5 to 6.5mm high quality ceramic media has become available that has greatly increased the efficiency of grinding coarser material in the IsaMill. This media has also resulted in product size distribution curves that are much sharper and more efficient to process downstream. An example of this from Anderson, et al, is shown in Figure 6 for MRM ore. At an equivalent energy and feed size the 5-6mm media gives a much sharper size distribution curve with the top size completely ground.
In the case of the Ernest Henry Mine magnetite circuit the mill feed during startup was regularly between an F80 of 300-350 microns with a top size approaching 1mm. Due to the coarseness of this stream a 6.5mm ceramic media manufactured by Cenotec was chosen to ensure top size breakage. Signature plot work was done on the cyclone underflow of the rougher magnetic separator at AMMTEC prior to starting this circuit. The comparison between the AMMTEC signature plot (blue) and the EHM commissioning surveys (red) are shown below in Figure 7. In this case the test ore and commissioning ore were taken months apart and the media top size is 6mm for the AMMTEC test and 6.5mm for the EHM M10,000 but the scaleup to the mill target of 45 microns still falls within the 5% margin of error associated with the test. In this case the net energy for the M10,000 to grind to a P80 of 42 microns is 29.9 kWh/t versus 28.8 kWh/t.

It is currently unknown how large of a media can be tested in a standard M4 IsaMill and still scale to full size mills. At some point the wall effect will become apparent and the power draw will increase due to media shearing between the disc tip and mill shell. This has not happened at 6mm.
Despite being a 6.5mm top size and needing to grind coarse abrasive magnetite slurry the Cenotec media at EHM has remained very round and worn at a rate of 8-9 g/kWh. Figure 8 shows this media after being emptied into the media bin one month into operation. At this point the media would have turned over slightly more than 1 time.

Advances in the size of media available and the confidence that it will scale from a standard M4 test has led to other test programs to take advantage of IsaMill technology in new grinding duties. Work performed by AMMTEC from David, et al, shown below in Figure 9 demonstrates the energy efficiency of the IsaMill when taking a coarser magnetite feed and grinding to a target of 34 microns. This efficiency gain over the ball mill increases as the grind goes finer. This would further promote the possibility of running the IsaMill either
alone at a small scale or in series with a ball mill with 300-400 micron F80’s and being efficient to grind products below 60-70 microns.

![IsaMill/Levin test comparison](image)

Figure 9. IsaMill/Levin test comparison (David, Larson, Le 2011)

Similar to the MRM results with the sharper product size distribution the magnetite results (Figure 10) showed no material over 106 microns in a 37 micron P80 and very little over 75 microns. The P98/P80 ratio in the IsaMill is typically a positive factor even with smaller grinding media but the use of 5mm media on feeds typically served by 3.5mm media can improve this further. In the case of magnetite this results in less oversize middlings and an improved concentrate grade.

![IsaMill magnetite product size distribution curve](image)

Figure 10. IsaMill magnetite product size distribution curve (David, et al)

In this program the M4 was also run continuously in the pilot plant to confirm that the energy and product size distribution were not a result of coarse material segregation and to produce feed for the downstream processes. The M4 is ideally suited for this application as at a
common magnetite regrind energy of 10-15 kWh/t about 100 kg/hour can be processed through the mill.

The transition to 5mm media may also improve the energy efficiency of a process depending on the target P80. In all cases whether comparing different technologies or just media size for a particular technology there is a cross-over point where one becomes more efficient than the other. In the cases in Figure 11 the larger media is more efficient to a coarser product size. When an application requires a finer grind and more energy there is enough residence time for the 3.5mm media to break down the coarsest incoming feed at the same rate it comes in. Compared to 5mm media it will remove smaller pieces at a time so needs more time to grind a coarser feed. This is ideal for a finer grind and high energy application. At a coarser product and lower energy the high flowrate may overwhelm the 3.5mm media and fill up the mill with coarse material.

This is shown in Figure 11 for two different platinum applications. At product sizes above 30 microns the 5mm media will be more efficient than the 3.5mm media. In this case the smallest media size capable of breaking the incoming feed is not necessarily the most efficient option.

![Figure 11. Comparison of 5mm and 3.5mm media for platinum ore (Larson 2010)](image)

Use of +5mm media has the potential to improve energy efficiency and product size distribution for existing operations while increasing the range of feed sizes the IsaMill can treat in future applications.

**Conclusions**

By treating the independent mineral processing laboratories around the world as partners and involving them in advances in IsaMill technology Xstrata Technology has been able to maintain rigorous standards in the quality of scaleup work done by these labs. Advances in the ability to treat coarser feeds have not impaired the ability of large scales mills to still be accurately scaled from the standard 4 litre IsaMill. Large improvements in energy efficiency and product size distributions can be realized by utilizing the 4 litre mill in the lab or pilot plant setting.
Acknowledgements

The authors would like to thank Xstrata Technology for permission to publish this work and assistance in writing it. We would also like to thank Xstrata Copper’s Ernest Henry Mine for permission to publish the EHM magnetite IsaMill scaleup work and AMMTEC for performing this lab work.

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EDITORS:

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University of British Columbia

Kelly McLeod
Process Consultant

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A PILOT-SCALE EXAMINATION OF A HIGH PRESSURE GRINDING ROLL / STIRRED MILL COMMINUTION CIRCUIT

*J.A. Drozdiak¹ and B. Klein¹, S. Nadolski², A. Bamber³

¹NBK Institute of Mining
517 – 6350 Stores Road
Vancouver, Canada V6T 1Z4
(*Corresponding author: jeff@mining@yahoo.com)

²Koeppern Machinery Australia
73 Pavers Circle
Malaga, Australia

³BC Mining Research
122 – 1857 West 4th Avenue
Vancouver, Canada V6J 1M4
A PILOT-SCALE EXAMINATION OF A HIGH PRESSURE GRINDING ROLL / STIRRED MILL COMMINUTION CIRCUIT

ABSTRACT

In this paper we examine, through pilot-scale testing, the possibility of operating a circuit comprised of two stages of HPGR comminution, followed by grinding through a horizontal stirred mill. In order to assess whether the novel circuit design could achieve the reduced energy requirements indicated in the literature, two more-established circuits, a cone crusher / ball mill and an HPGR / ball mill, were examined using a combination of testing and flowsheet simulation. The results showed that, based solely on the specific energy requirements for comminution, the HPGR / stirred mill circuit achieved a reduction of 9.2% and 16.7% over the HPGR / ball mill and cone crusher / ball mill circuits, respectively.

KEYWORDS

High pressure grinding roll, stirred mill, energy, comminution, flowsheet development

INTRODUCTION

The mining industry will be faced with new challenges in the years ahead. The exponentially-increasing global population has resulted in an increased demand for raw resources. With the known rich, coarse-grained deposits being depleted, attention has turned to development of low-grade deposits requiring increased tonnages to achieve adequate metal production. This increased tonnage has resulted in an increased energy demand associated with metal extraction. Coupled with this, society is becoming increasingly conscious of their footprint on the environment, and serious attempts have begun to reduce carbon emissions and increase energy efficiency (Norgate & Haque, 2010). To adapt to this changing landscape, the mining industry must begin to accept and adapt new, more energy-efficient technologies and begin to focus on developing flowsheets capable of addressing the above issues.

Comminution, the process of crushing and grinding ore to liberate valuable minerals, is the most energy-intensive part of the processing flowsheet, and accounts for upwards of 75-80% of the overall energy consumption of the processing plant (Abouzeid & Fuerstenau, 2009). In addition, unit operations such as tumbling mills are as low as 1% efficient (Fuerstenau & Abouzeid, 2002). Currently, the main comminution circuits employed in the mining industry to process hard-rock, low-grade deposits include some form of tumbling mill. This equipment utilizes steel balls (ball mills), competent ore (Autogenous Grinding (AG) mills), or the combination of the two (Semi-Autogenous Grinding (SAG) mills) to fracture rock using the breakage mechanisms of impact and abrasion. The rotation of these large, cylindrical mills, coupled with the low probability of ball – particle collisions, results in a high demand for energy in the grinding process. Although their established circuit design and ability to process high tonnages is a huge benefit, the increased energy demand and inability to efficiently grind to liberation sizes below 45 µm (Shi, Morrison, Cervellin, Burns, & Musa, 2009) could slowly decrease their role in flowsheet designs of the future, especially as the increased demand for raw resources results in an increase in the development of finer-grained deposits.

In the past 20 years, new, more energy-efficient technologies have been developed and adapted for hard-rock mining comminution. The High Pressure Grinding Roll (HPGR), an innovative technology adapted from the cement and briquetting industries, has begun to be considered for more base metal
projects now that roll surfaces have been developed to treat hard, abrasive ores (Dunne, 2006). Operating with two counter-rotating rolls, HPGRs create a compressive bed of particles between the rolls, utilizing the process of inter-particle breakage. This form of breakage results in improved comminution performance with a decreased demand on energy (Klymowsky, Patzelt, Knecht, & Burchardt, 2006). Additionally, unlike tumbling mills, which require steel balls to act as an energy transfer medium, HPGRs transfer energy directly from the rolls to the bed of material, resulting in an increase in energy efficiency (Fuerstenau & Kapur, 1995). Another technology, known as a horizontal stirred mill or IsaMill™, was adapted from the pharmaceutical and related industries in the early 1990s to help effectively process fine-grained ore bodies (Johnson, Gao, Young, & Cronin, 1998). The IsaMill™ consists of a cylindrical tube with a centrally-rotating shaft, mounted with evenly-spaced grinding discs. Loaded with small ceramic grinding media (2 - 6 mm) and operated at high speeds, the equipment utilizes high-intensity attrition breakage to reduce particles in size. The rotation of a central shaft, as opposed to the entire grinding chamber (tumbling mills), results in decreased energy requirements, while the combination of small, hard (ceramic) grinding media and increased media velocity has been shown to improve the energy efficiency of grinding in fine particle sizes (Burford & Clark, 2007).

In this paper, we examine the possibility of incorporating the above-mentioned energy-efficient equipment into a single flowsheet, and eliminating the need for a tumbling mill. The biggest obstacle surrounding this research was that the proposed circuit would be operating both pieces of equipment outside of their normal operating range. As HPGRs began being adapted to the hard-rock mining sector, they found the most functionality in a tertiary crushing role, preparing feed for the ball mill (Morley, 2006). Therefore, the process envelope for an HPGR operating in hard-rock circuits typically has feed sizes of up to 70 mm, and products normally no finer than 4 mm (Gruendken, Matthies, & van der Meer, 2010). At the same time, horizontal stirred mill technologies such as the IsaMill™ have begun to be well-established in grinding applications as a regrind mill, providing a more energy-efficient alternative for processing rougher concentrates (Burford & Clark, 2007).

The combination of an HPGR and a stirred mill into a single flowsheet has been discussed several times in the literature. Valery and Jankovic (2002) proposed the first concept of a combination HPGR / stirred mill circuit in a study examining the need for a reduction in the energy requirements of comminution. Simulating results for a more energy efficient circuit, a high-intensity blasting, two stage HPGR / Vertimill® circuit was compared to a conventional blasting, SAG / ball mill circuit. The simulation results predicted an energy savings of 45%, but no actual testwork was conducted. Pease (2007) presented the concept of an HPGR / IsaMill™ circuit in his discussion of coarse stirred milling at McArthur River. No testing was carried out, but Pease predicted that this circuit could be an example of comminution flowsheet design of the future. Ayers, Knojjes, and Rule (2008) described the first operation of an HPGR / IsaMill™ circuit using pilot-scale equipment. The authors documented Anglo Platinum’s research into applying the IsaMill™ to coarser feed applications. A continuously operating circuit was established using an HPGR in closed circuit with a dry screen, followed by wet screening of the undersize, at a cut size of 850 µm. The screen product was fed to an M250 IsaMill™ operating with 3.5 mm MT1 ceramic grinding media. With an f80 of 300 µm and a product p80 of 45 µm, the IsaMill™ circuit achieved 1.3 tph, with a specific energy consumption of 75 kWh/t and a total circuit energy consumption of 80 kWh/t.

Using pilot-scale equipment, we developed an appropriate circuit layout and determined the potential specific energy for comminution required to process a copper-nickel sulphide ore from an f80 of 21 mm to a p80 of 75 µm. To determine whether the novel circuit arrangement could reduce the energy demand for comminution indicated in the literature, specific energy requirements for two established circuit layouts, a cone crusher / ball mill and an HPGR / ball mill, were explored using a combination of laboratory testing and circuit simulation. The ability to improve comminution efficiency, while providing the flexibility of grinding efficiently to fine product sizes, could help make the HPGR / stirred mill circuit an attractive alternative for future comminution flowsheets.
EXPERIMENTAL PROGRAM

Three circuits were examined for this energy comparison study, a cone crusher / ball mill circuit, an HPGR / ball mill circuit, and the novel HPGR / stirred mill circuit. The feed size to each circuit was fixed at an f80 of 21 mm, and a product p80 of 75 µm was chosen as a suitable feed size for flotation. The circuits were evaluated solely on the power consumed per tonne of material, in order to achieve an equivalently-sized product from an equivalently-sized feed. Energy requirements of material-handling equipment, such as conveyors, pumps and screens, were not taken into account.

Comminution Circuits

What follows is a discussion of the three comminution circuits examined in this paper. The approach in all cases was to determine an appropriate set of design criteria for each flowsheet and to calculate the specific work index for each stage of comminution, based on the work index determined and the transfer sizes selected.

Cone Crusher / Ball Mill Circuit

The first circuit we examined was a cone crusher / ball mill circuit, typically found in a three-stage crushing flowsheet. This circuit was the industry standard for hard-rock comminution prior to the establishment of SAG mill technology. The circuit is comprised of a cone crusher in closed circuit with a screen, followed by a ball mill in closed circuit with a cyclone. The flowsheet of this circuit is shown in Figure 1. Data for the circuit was generated from a combination of Bond grindability testing and flowsheet simulation using JK SimMet® software.

HPGR / Ball Mill Circuit

The second circuit we examined was an HPGR / ball mill circuit. This circuit mimics the standard HPGR comminution flowsheet currently being used in the hard-rock mining sector at operations such as Cerro Verde in Peru (Vanderbeek, Linde, Brack, & Marsden, 2006). The circuit is comprised of a high pressure grinding roll in closed circuit with a screen followed by a ball mill in closed circuit with a cyclone (refer to Figure 2). Data for this circuit was generated using a combination of HPGR pilot-scale testing, Bond grindability testing, and simulation using JK SimMet® software.
For HPGR pilot-scale evaluation, tests were carried out to assess the influence of different process parameters on comminution performance. These tests included the variation of specific pressing force, as well as closed-circuit testing with a 4 mm screen. Data from this study was entered into JK SimMet® to model fit an appropriate HPGR model. The T10H and HPGR power coefficient model parameters were fitted using the procedure outlined by Daniel and Morrell (2004). After calibration of the HPGR model, simulation was carried out for the HPGR / ball mill circuit to determine the appropriate transfer size between the HPGR and the ball mill.

**HPGR / Stirred Mill Circuit**

The final circuit we examined was the HPGR / stirred mill circuit. Daniel (2007b) determined from HPGR pilot-scale testing that two consecutive passes through the HPGR produced the highest size reduction ratios and further passes through the rolls resulted in diminishing size reduction and lower energy efficiency. Therefore, we determined that the novel HPGR / stirred mill circuit would incorporate two stages of high pressure grinding to prepare the feed for stirred milling. Drozdiak, Nadolski, Bamber, Klein, and Wilson (2010) demonstrated through pilot-scale testing that an appropriate transfer size between a two-stage HPGR circuit and a stirred mill circuit would be 710 µm. Using this data, two different HPGR / stirred mill circuits were examined. Circuit A comprised of the first-stage HPGR in open circuit feeding the second-stage HPGR in closed circuit with a 710 µm screen, and the undersize passing through a stirred mill in open circuit (refer to Figure 3), while Circuit B comprised of the first-stage HPGR in closed circuit with a 4 mm screen and the undersize feeding the same circuit layout as Circuit A (refer to Figure 4). Since no small-scale tests are available to determine the specific energy requirements for this equipment, pilot-scale testing was performed for the entire circuit.
Sample Description

The sample used for this study came from Teck Limited’s Mesaba copper-nickel deposit located in the Mesabi Range of the Duluth intrusive complex, situated in North-eastern Minnesota. This complex is comprised of mafic volcanics (tholeiitic basalt) with layered intrusions of primarily a gabbro-troctolite composite (Minnesota Geological Survey, 2010). Mineralogy of the Mesaba deposit comprises mainly of massive and disseminated sulphides with the main minerals of interest being chalcopyrite (copper), cubanite (copper), and pentlandite (nickel). The inferred resource stands at 700 Mt, with a grade of 0.46% Cu and 0.12% Ni (Infomine, 2001).

Approximately 5 tonnes of sample, at nominally 100% minus 100 mm, was shipped to UBC. We screened and crushed the material in a laboratory jaw crushe to 100% minus 32 mm, and homogenized and split the sample into sixteen 45-gallon drums using a rotary sample splitter. A representative sample was taken for size distribution, bulk density and moisture content determination. A moisture content of 1% and a bulk density of 2.16 t/m$^3$ with a Specific Gravity (SG) of 3.0 were established for the ore. The particle size distribution of the sample is shown in Figure 5.
Figure 5 – Feed particle size distribution of Mesaba ore

Equipment

The following section describes the main pieces of test equipment used and the methodology used for calculating specific energy consumption.

High Pressure Grinding Roll

HPGR testing was conducted using a pilot-scale unit manufactured by Koeppern Machinery Australia. The pilot unit is custom-made for obtaining design information for sizing and selection of industrial-scale units. Table 1 summarizes the technical data provided by Koeppern for the machine. Experimental data was recorded every 200 ms through the programmable logic controller (PLC) data logger and downloaded to a laptop. The computer system measures time, roller gap (left and right), pressing force (left and right), and power draw. A picture of the HPGR pilot unit is shown in Figure 6.

<table>
<thead>
<tr>
<th>Table 1 – HPGR machine specifics</th>
</tr>
</thead>
<tbody>
<tr>
<td>Roller Diameter</td>
</tr>
<tr>
<td>Roller Width</td>
</tr>
<tr>
<td>Press Drive</td>
</tr>
<tr>
<td>Feed System</td>
</tr>
<tr>
<td>Wear Surface</td>
</tr>
<tr>
<td>Installed Power</td>
</tr>
<tr>
<td>Maximum Pressing Force</td>
</tr>
<tr>
<td>Maximum Specific Pressing Force</td>
</tr>
<tr>
<td>Variable Speed Drive</td>
</tr>
</tbody>
</table>
A pilot test with the HPGR comprises the crushing of one 45-gallon drum of material (~375 kg). The material is loaded into a feed hopper with the use of an overhead crane and drum tipper. Once the machine conditions are stabilized, the slide gate of the feed hopper is opened and the test begins. The material flows with the aid of gravity through the HPGR rollers and drops on to the product conveyor located below the rolls. Once the test is complete, specific throughput and specific energy consumption are determined for the test using the power draw off the main motor and the throughput recorded during the stable operating period.

Since the HPGR does not grind uniformly across the roller width, a splitter gate is installed on the end of the product conveyor to separate the product into centre, edge and waste streams. The centre portion is finer than the edge portion and, during testing, a particle size distribution is performed on each to accurately predict size distributions for full-scale operations. For square rollers found in industrial units, where roll diameter is equal to roll width, the proportion of centre and edge product is observed to be approximately 85% centre and 15% edge. All of the HPGR product size distributions presented in this paper account for this through scaling of centre and edge size distributions at a ratio of 85:15. Material collected during unstable operation, initial response, and material run-out periods, was designated as waste material and only material which had been crushed during stable press operation was collected for analysis.

**Horizontal Stirred Mill**

Stirred mill testing was carried out using Netzsch’s M20 horizontal stirred mill. The mill has a capacity of 20 litres and is installed with an 18.6 kW motor. The mill was fitted with the current IsaMill™ internal grinding configuration and a Variable Frequency Drive (VFD). The installation of the VFD allowed for direct readings of mill power and mill speed. To monitor the mill, sensors were installed for feed pressure, and both feed and product temperature. A PLC interface and data logger was also installed to control the mill settings and record all important mill parameters during testing. The mill configuration,
including grinding disc design, was based on recommendations from Xstrata Technology and allows for the ability to scale-up results to what would be expected for industrial IsaMills™.

A Watson-Marlow and Bredel SPX 25 hose pump and corresponding VFD were used to feed the mill. The pump has a capacity of 25 L/min and was designed to handle viscous slurries. The installation of a VFD for the 1.5 kW pump motor allowed for accurate monitoring and control of mill flow rate. The corresponding mixing system was comprised of two 180 L-capacity mix tanks with corresponding 250 W variable speed agitators and was designed to mix slurries at upwards of 60% solids with a particle top size as coarse as 1.2 mm. The piping system for the circuit was set up so that each mix tank could easily be switched from product to feed with minimal delay. The final setup is shown in Figure 7.

![Figure 7 – M20 stirred mill installation at the NBK Institute of Mining](image)

For testing of the stirred mill energy requirements, a graph of specific energy consumption and p80 grind size was generated. This graph, known as a signature plot, is the common method used in the industry for accurate sizing of full-scale IsaMills™ and has a scale-up ratio of 1:1 (Gao, Weller, & Allum, 1999). The procedure entails running the material through the mill a select number of times and recording the energy requirements and product size after each pass. The passes are carried out consecutively in order to observe the energy consumption as the size of the product decreases. The results provide a series of points plotted on a log – log graph that shows the relationship between energy input and product size (p80).

Particle sizing for this work was done using a Malvern Mastersizer 2000. This laser sizing equipment utilizes the principle that grains of different sizes diffract light at different angles; a decrease in size produces an increase in diffracted angle. This equipment has become the standard for analyzing size ranges unrealistic for screening (Larson, Morrison, & Pietersen, 2008).
Vibrating Screen

All screening work carried out for HPGR closed-circuit testing was performed using a SWECO® Vibro-Energy® Separator. This vibrating screen, model ZS40, is equipped with a 373 W motor and a counterweight system to produce both vertical and horizontal vibrating motion. The screener is equipped with 1 m diameter wire mesh screens.

Bond Test Ball Mill

Energy requirements for ball mill grinding were determined using Bond ball mill work indices for cone crusher and HPGR product. Representative samples were screened at minus 3.35 mm and processed through a standard Bond ball mill measuring 305 mm in length and 305 mm in diameter, with a 285 ball charge weighing 20 125 g. Testing was carried out using the standard Bond Ball Mill Grindability Test procedure developed by Bond (1961). For the crushing work index, insufficiently sized material was available to perform impact testing. Therefore, a traditional approach was taken and the Bond work index was used. The resulting indices were then used with the Bond equation to calculate specific energy consumption for both crushing and grinding.

RESULTS

Cone Crusher / Ball Mill Circuit

The specific energy consumption of comminution for the circuit was determined with a flowsheet developed using JK SimMet® software. The circuit was designed for 250 tph capacity and equipment was sized based on a product p80 of 75 µm. Table 2 summarizes the equipment sized for the circuit.

<table>
<thead>
<tr>
<th>Table 2 – Equipment selection for the cone crusher / ball mill circuit</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Cone Crusher</strong></td>
</tr>
<tr>
<td>Closed Side Setting</td>
</tr>
<tr>
<td>Re-circulating Load</td>
</tr>
<tr>
<td><strong>Product Screen</strong></td>
</tr>
<tr>
<td>Aperture Size</td>
</tr>
<tr>
<td><strong>Ball Mill</strong></td>
</tr>
<tr>
<td>Diameter</td>
</tr>
<tr>
<td>Length</td>
</tr>
<tr>
<td>Critical Speed</td>
</tr>
<tr>
<td>Media Charge</td>
</tr>
<tr>
<td>Media Top Size</td>
</tr>
<tr>
<td>Re-circulating Load</td>
</tr>
<tr>
<td><strong>Hydrocyclones</strong></td>
</tr>
<tr>
<td>Quantity</td>
</tr>
<tr>
<td>Cyclone Diameter</td>
</tr>
<tr>
<td>Inlet Diameter</td>
</tr>
<tr>
<td>Vortex Finder Diameter</td>
</tr>
<tr>
<td>Apex (Spigot) Diameter</td>
</tr>
<tr>
<td>Length</td>
</tr>
<tr>
<td>Cone Angle</td>
</tr>
</tbody>
</table>
Simulation of the flowsheet predicted that the appropriate transfer size between the crushing circuit and the ball mill circuit would be 80% passing 2.12 mm. To calculate the overall specific energy consumption for the cone crusher and ball mill, work indices were determined for the material. In the case of the cone crusher, no material was available for the size requirements, 50-75 mm, necessary to perform impact testing. Therefore, a traditional approach was taken and the Bond ball mill work index was used. Locked-cycle testing was performed using two sieve sizes (106 µm and 150 µm), to allow for the comparison of different product sizes. The results for the work indices of the circuit are shown in Table 3.

### Table 3 – Bond work indices for the cone crusher / ball mill circuit

<table>
<thead>
<tr>
<th>Locked Cycle Screen Size (µm)</th>
<th>Feed p80 (µm)</th>
<th>Product p80 (µm)</th>
<th>Bond Work Index (kWh/t)</th>
</tr>
</thead>
<tbody>
<tr>
<td>150</td>
<td>2,133</td>
<td>119</td>
<td>15.9</td>
</tr>
<tr>
<td>106</td>
<td>2,241</td>
<td>80</td>
<td>16.5</td>
</tr>
</tbody>
</table>

Using the Bond work indices and the transfer size determined from flowsheet simulation, the theoretical energy requirements for the circuit are calculated using the Bond equation (Bond, 1961). Calculation of the cone crusher energy requirements will use the Bond work index at a sieve size of 150 µm. This coarser screen size provides a lower estimate for the energy requirements of a cone crusher and provides a best-case scenario for the crushing circuit. Calculation of the ball mill energy requirements will use the Bond work index at a sieve size of 106 µm. The final product for the circuit was set at a p80 of 75 µm and the Bond work index, at a sieve size of 106 µm, better reflects the energy requirements to grind to this finer particle size. A summary of the resulting energy requirements for the circuit are shown in Table 4.

### Table 4 – Summary of the cone crusher / ball mill circuit energy requirements

<table>
<thead>
<tr>
<th>Unit Operation</th>
<th>Feed p80 (mm)</th>
<th>Product p80 (mm)</th>
<th>Specific Energy Consumption (kWh/t)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cone Crusher</td>
<td>21</td>
<td>2.12</td>
<td>2.36</td>
</tr>
<tr>
<td>Ball Mill</td>
<td>2.12</td>
<td>0.075</td>
<td>15.47</td>
</tr>
<tr>
<td>TOTAL</td>
<td></td>
<td></td>
<td>17.83</td>
</tr>
</tbody>
</table>

### HPGR / Ball Mill Circuit

#### The HPGR Circuit

To evaluate the energy requirements for the HPGR / ball mill circuit, HPGR pilot-scale testing was performed to determine the ideal specific pressing force for the Mesaba material and assess how operating the HPGR in closed circuit with a 4 mm screen would affect comminution performance. Four initial tests were done to determine the effect of specific pressing force. Pressures of 2 N/mm², 3 N/mm², 4 N/mm², and 5 N/mm² were chosen and comparisons were made with respect to product size, net specific energy consumption, and specific throughput (m-dot). All tests were performed at a roller speed of 0.75 m/s and feed moisture content by weight of 2.5%.

The comparison of product particle size at different specific pressing forces is shown in Figure 8. As the pressing force increased, both the p50 and p80 decreased, although the effect on p80 was more pronounced than the effect on p50. This result is due to an increased force being exerted on the particles as they flow through the rolls. An increased force promotes increased breakage and the effect is more pronounced on larger sized particles, hence the steeper trend for the p80.
The comparison of specific throughput (m-dot) at different specific pressing forces is shown in Figure 9. As the pressing force increased, the specific throughput decreased. This trend is due to the gap between the rollers decreasing slightly with increasing pressing forces, resulting in the reduction of throughput in the machine.

The comparison of specific energy consumption at different specific pressing forces is summarized in Figure 10. As the pressing force increased, the energy consumption also increased. This is typical of the process because more energy is being transmitted into the material at higher pressures.
A specific pressing force of 4 N/mm² was selected for the remainder of pilot-scale testing. The results indicated that a pressing force of 4 N/mm² provided a fine balance between energy consumption and size reduction without a significant change in specific throughput.

To test the effect of closed-circuit operation, locked-cycle testing was conducted using a 4 mm screen. Material was processed through the HPGR at 4 N/mm² and the product screened at 4 mm using the SWECO® vibrating screen. Using the product size distributions from testing, the percentage of plus 4 mm was calculated (at 90% screening efficiency) and then combined with fresh feed and re-run through the HPGR. This process was repeated two more times to simulate closed-circuit operation. Results were generated to determine size reduction, specific throughput, and specific energy consumption for each cycle. The resulting product size for each cycle is shown in Figure 11. The chart shows that the introduction of a re-circulating load decreased the product size and began to stabilize by cycle four.
The effect on specific throughput for closed-circuit testing is shown in Figure 12. Closed-circuit operation had little effect on specific throughput. The variation between each cycle can probably be attributed to testing error.

![Figure 12 – Specific throughput for closed-circuit testing](image1)

The results for specific energy consumption are displayed in Figure 13. The closed-circuit testing had little to no effect on specific energy consumption.

![Figure 13 – Specific energy consumption for closed-circuit testing](image2)

Results from the last cycle of testing are summarized in Table 5. These results will be used for energy calculations, as well as the experimental data required for model fitting with JK SimMet®.
Table 5 – Results for the last cycle of closed-circuit testing

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>f₈₀</td>
<td>21.77</td>
</tr>
<tr>
<td>f₅₀</td>
<td>13.38 mm</td>
</tr>
<tr>
<td>P₈₀</td>
<td>6.61 mm</td>
</tr>
<tr>
<td>P₅₀</td>
<td>1.91 mm</td>
</tr>
<tr>
<td>Percentage Passing -4 mm</td>
<td>67.4%</td>
</tr>
<tr>
<td>Net Specific Energy Consumption</td>
<td>1.45 kWh/t</td>
</tr>
<tr>
<td>(-4 mm) Net Specific Energy Consumption</td>
<td>2.15 kWh/t</td>
</tr>
<tr>
<td>Specific Throughput</td>
<td>304 ts/hm³</td>
</tr>
</tbody>
</table>

Flowsheet Development and Circuit Energy Calculations

Using the results from the last cycle of HPGR closed-circuit testing, model fitting of an HPGR circuit was performed using JK SimMet®. The T10H and HPGR power coefficient model parameters were fitted using the model fit tool in JK SimMet®. This tool uses an iterative function to fit experimental data to simulated data by adjusting model parameters until a correlation can be achieved. The T10h and HPGR power coefficient parameters relate to the breakage mechanisms in the compression zone of the HPGR and the product size for closed-circuit testing was used as the experimental data. The procedure used for calibrating the HPGR model was outlined by Daniel and Morrell (2004). The resulting model fit was able to simulate a product size distribution similar to the one generated experimentally. Once an HPGR model was calibrated for use with Mesaba ore, a flowsheet was designed for 250 tph capacity with a product p₈₀ of 75 µm. The equipment sized for the flowsheet is summarized in Table 6.

Table 6 – Equipment selection for the HPGR / ball mill circuit

<table>
<thead>
<tr>
<th>High Pressure Grinding Roll</th>
</tr>
</thead>
<tbody>
<tr>
<td>Roller Diameter</td>
</tr>
<tr>
<td>Roller Width</td>
</tr>
<tr>
<td>Re-circulating Load</td>
</tr>
<tr>
<td>Product Screen</td>
</tr>
<tr>
<td>Aperture Size</td>
</tr>
<tr>
<td>Ball Mill</td>
</tr>
<tr>
<td>Diameter</td>
</tr>
<tr>
<td>Length</td>
</tr>
<tr>
<td>Critical Speed</td>
</tr>
<tr>
<td>Media Charge</td>
</tr>
<tr>
<td>Media Top Size</td>
</tr>
<tr>
<td>Re-circulating Load</td>
</tr>
</tbody>
</table>

Hydrocyclones

<table>
<thead>
<tr>
<th>Quantity</th>
<th>Cyclone Diameter</th>
<th>Inlet Diameter</th>
<th>Vortex Finder Diameter</th>
<th>Apex(Spigot) Diameter</th>
<th>Length</th>
<th>Cone Angle</th>
</tr>
</thead>
<tbody>
<tr>
<td>7</td>
<td>350 mm</td>
<td>175 mm</td>
<td>150 mm</td>
<td>113 mm</td>
<td>450 mm</td>
<td>20°</td>
</tr>
</tbody>
</table>
Simulation of the flowsheet predicted that the appropriate transfer size between the HPGR circuit and the ball mill circuit would be 80% passing 1.6 mm. Several publications have indicated that HPGR comminution and the presence of micro-cracks in the product, results in a decrease in the Bond work index when compared with conventionally crushed product (Daniel, 2007a; Muranda, 2009; Rule, Smit, Cope, & Humphries, 2008). To confirm this advantage, Bond ball mill work indices were determined for HPGR product at different specific pressing forces. Samples were taken from HPGR centre product and screened at 3.35 mm with no additional crushing. As with Bond work indices for cone crusher product, two separate screen sizes (106 µm and 150 µm) were tested to allow for the comparison of different product sizes. The results, including cone crusher product for comparison, are summarized in Table 7.

<table>
<thead>
<tr>
<th>Locked Cycle Screen Size (µm)</th>
<th>Feed Preparation Method</th>
<th>Feed f80 (µm)</th>
<th>Product p80 (µm)</th>
<th>Bond Work Index (kWh/t)</th>
</tr>
</thead>
<tbody>
<tr>
<td>150</td>
<td>Cone Crusher</td>
<td>2,133</td>
<td>119</td>
<td>15.9</td>
</tr>
<tr>
<td></td>
<td>HPGR - 3 N/mm²</td>
<td>1,854</td>
<td>125</td>
<td>14.5</td>
</tr>
<tr>
<td></td>
<td>HPGR - 4 N/mm²</td>
<td>1,849</td>
<td>124</td>
<td>14.5</td>
</tr>
<tr>
<td></td>
<td>HPGR - 5 N/mm²</td>
<td>1,497</td>
<td>118</td>
<td>13.3</td>
</tr>
<tr>
<td>106</td>
<td>Cone Crusher</td>
<td>2,241</td>
<td>80</td>
<td>16.5</td>
</tr>
<tr>
<td></td>
<td>HPGR - 3 N/mm²</td>
<td>1,765</td>
<td>81</td>
<td>15.8</td>
</tr>
<tr>
<td></td>
<td>HPGR - 4 N/mm²</td>
<td>1,764</td>
<td>81</td>
<td>15.7</td>
</tr>
<tr>
<td></td>
<td>HPGR - 5 N/mm²</td>
<td>1,682</td>
<td>79</td>
<td>15.7</td>
</tr>
</tbody>
</table>

A reduction in Bond work index was achieved between cone crusher and HPGR product. However, the reduction went from 8.8% to 4.8% with a decrease in screen size. Results also showed a reduction in Bond work index with increasing specific pressing force, although this effect may be attributed to the finer feed size. The reduction in screen size may have caused a decrease in the effectiveness of product micro-cracking and a relatively higher amount of energy was required to produce the finer product. Using the specific energy results obtained from Table 5, coupled with the Bond work index for 4 N/mm² at a screen size of 106 µm and the transfer size determined from JK SimMet simulation, the specific energy requirements for the HPGR / ball mill circuit can be summarized in Table 8.

<table>
<thead>
<tr>
<th>Unit Operation</th>
<th>Feed f80 (mm)</th>
<th>Product p80 (mm)</th>
<th>Specific Energy Consumption (kWh/t)</th>
</tr>
</thead>
<tbody>
<tr>
<td>HPGR</td>
<td>21</td>
<td>1.6</td>
<td>2.15</td>
</tr>
<tr>
<td>Ball Mill</td>
<td>1.6</td>
<td>0.075</td>
<td>14.2</td>
</tr>
<tr>
<td>TOTAL</td>
<td></td>
<td></td>
<td>16.35</td>
</tr>
</tbody>
</table>

**HPGR / Stirred Mill Circuit**

The HPGR / stirred mill circuit required considerably more pilot-scale testing than the previous two circuits, since very few operating examples could be found in the literature. The results of pilot-scale testing determined the appropriate layout for the two-stage HPGR circuit and provided the corresponding specific energy consumption for circuit energy summation.

The HPGR Circuit

Pilot-scale testing was conducted to produce suitable data for the HPGR section of the HPGR / stirred mill circuit. Since size reduction is limited with one stage of HPGR comminution (Daniel, 2007b), design of an HPGR / stirred mill circuit required at least two consecutive stages of HPGR comminution to
produce a particle size acceptable for stirred milling. With the transfer size between the second-stage HPGR and the stirred mill established by Drozdiak et al. (2010), work was done to determine the appropriate transfer size between each stage of HPGR crushing. Two options were examined to find the appropriate circuit layout. In Circuit A, the first-stage HPGR was placed in closed circuit with a 4 mm screen, while in Circuit B, the first stage remained open circuit and the second stage accepted product directly from stage one.

For Circuit A, closed-circuit testing product from the HPGR / ball mill circuit was processed again through the HPGR at the same roller speed (0.75 m/s) and specific pressing force (4 N/mm²). The use of the same specific pressing force for second-stage HPGR crushing stems from work performed by Rule et al. (2008), in which they found that no difference was observed when changing the specific pressing force in the second stage of two-stage HPGR crushing. For Circuit B, fresh feed was processed through two consecutive stages of HPGR comminution using the same roller speed and specific pressing force as Circuit A. The results for both options are summarized in Table 9. The size distributions for Circuits A and B are presented in Figure 14 and Figure 15, respectively.

Table 9 – Summary of results for the first-stage HPGR operating in open (Circuit A) and closed (Circuit B) circuit

<table>
<thead>
<tr>
<th>HPGR Stage 1</th>
<th>Circuit A</th>
<th>Circuit B</th>
</tr>
</thead>
<tbody>
<tr>
<td>f₈₀</td>
<td>21.77 mm</td>
<td>21.54 mm</td>
</tr>
<tr>
<td>f₅₀</td>
<td>13.38 mm</td>
<td>13.7 mm</td>
</tr>
<tr>
<td>HPGR p₈₀</td>
<td>6.61 mm</td>
<td>7.68 mm</td>
</tr>
<tr>
<td>HPGR p₅₀</td>
<td>1.91 mm</td>
<td>1.88 mm</td>
</tr>
<tr>
<td>Circuit p₈₀</td>
<td>1.86 mm</td>
<td>7.68 mm</td>
</tr>
<tr>
<td>Circuit p₅₀</td>
<td>489 µm</td>
<td>1.88 mm</td>
</tr>
<tr>
<td>Circuit p₈₀</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Circuit p₅₀</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Circuit Reduction Ratio</td>
<td>11.7</td>
<td>2.8</td>
</tr>
<tr>
<td>Net Specific Energy Consumption</td>
<td>1.45 kWh/t</td>
<td>1.54 kWh/t</td>
</tr>
<tr>
<td>Percentage Passing 4 mm</td>
<td>67.4%</td>
<td></td>
</tr>
<tr>
<td>(-4 mm) Net Specific Energy Consumption</td>
<td>2.15 kWh/t</td>
<td></td>
</tr>
<tr>
<td>Specific Throughput</td>
<td>304 ts/hm³</td>
<td>307 ts/hm³</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>HPGR Stage 2</th>
<th>Circuit A</th>
<th>Circuit B</th>
</tr>
</thead>
<tbody>
<tr>
<td>f₈₀</td>
<td>1.86 mm</td>
<td>7.68 mm</td>
</tr>
<tr>
<td>f₅₀</td>
<td>489 µm</td>
<td>1.88 mm</td>
</tr>
<tr>
<td>HPGR p₈₀</td>
<td>1.12 mm</td>
<td>2.79 mm</td>
</tr>
<tr>
<td>HPGR p₅₀</td>
<td>222 µm</td>
<td>462 µm</td>
</tr>
<tr>
<td>Circuit p₈₀</td>
<td>332 µm</td>
<td>339 µm</td>
</tr>
<tr>
<td>Circuit p₅₀</td>
<td>124 µm</td>
<td>142 µm</td>
</tr>
<tr>
<td>Circuit Reduction Ratio</td>
<td>5.6</td>
<td>22.6</td>
</tr>
<tr>
<td>Net Specific Energy Consumption</td>
<td>1.2 kWh/t</td>
<td>1.23 kWh/t</td>
</tr>
<tr>
<td>Percentage Passing 0.71 mm</td>
<td>71.3%</td>
<td>56.5%</td>
</tr>
<tr>
<td>(-0.71 mm) Net Specific Energy Consumption</td>
<td>1.68 kWh/t</td>
<td>2.18 kWh/t</td>
</tr>
<tr>
<td>Specific Throughput</td>
<td>235.81 ts/hm³</td>
<td>311 ts/hm³</td>
</tr>
</tbody>
</table>

TOTAL SPECIFIC ENERGY CONSUMPTION 3.83 kWh/t 3.72 kWh/t
Operating the first stage of HPGR crushing in open circuit required less energy compared with operating in closed circuit with a screen. If looked at strictly from an energy perspective, very little difference is gained choosing one circuit over the other, but if design and operating factors are considered, the choice of operating the first stage in open circuit becomes the better option. The ability to operate the circuit without a screen allows for the elimination of extra auxiliary equipment such as screens and conveyors, while the absence of an additional stage of wet screening would help to reduce the adverse effects that increased moisture content would have on HPGR performance (Fuerstenau & Abouzeid, 2007). Although the increased re-circulating load resulting in the second stage would require an increase in tonnage and machine size, this would be countered by the decreased machine size required for stage one. Overall, the reduced complexity offered by open circuit configuration led to us selecting this configuration for further testing.
Once the open circuit configuration was selected for stage one, additional pilot-scale testing was performed to evaluate how comminution performance would be affected by operating the second stage in closed circuit with a 710 µm screen. Testing was conducted in a similar manner to the locked cycle evaluation method used for the HPGR / ball mill circuit. Product from Circuit B was screened at 710 µm and a calculated split of oversize was mixed with fresh product from stage one and processed through the HPGR. This procedure was repeated two more times in order to simulate closed-circuit operation. The resulting product size for each cycle is shown in Figure 16. The product size increased slightly with the introduction of a re-circulating load. This is in contrast to the results for the HPGR / ball mill circuit, where the introduction of a re-circulating load caused a decrease in product size. This increase may have been the result of a finer re-circulating load reducing the breakage within the compressive bed.

Figure 16 – Product size for second stage closed circuit testing

The results for the effect of closed-circuit operation on specific throughput are displayed in Figure 17. The introduction of a re-circulating load had no effect on specific throughput.

Figure 17 – Specific throughput for second stage closed circuit testing
The results for the effect of closed-circuit operation on specific energy consumption are summarized in Figure 18. As with specific throughput, the introduction of a re-circulating load had no effect on specific energy consumption.

To achieve efficient screening at 710 µm for an industrial operation, the practice of wet screening is necessary. Fuerstenau and Abouzeid (2007) found that the introduction of moisture to an HPGR circuit leads to adverse effects on throughput and energy consumption. The effect of moisture on second-stage HPGR crushing was tested using product from the final closed-circuit cycle. We wet screened the sample over a 710 µm screen to determine the potential moisture content for oversize in a closed-circuit operation. The saturated oversize, with a measured moisture content of 10.5%, was then used to run an additional closed-circuit cycle. A summary of the results is presented in Table 10. To allow for a direct comparison of the effects of wet screening, the results from cycle four (dry) are presented as well. As expected, the results show an adverse effect on throughput and energy consumption, although the product size became considerably finer. The data generated for the wet screening cycle represents the worst-case scenario, and thus will be used for the energy calculations for the circuit.

Table 10 – Comparison of wet and dry screening for second stage closed circuit operation

<table>
<thead>
<tr>
<th></th>
<th>Dry Cycle</th>
<th>Wet Cycle</th>
</tr>
</thead>
<tbody>
<tr>
<td>Feed Moisture Content</td>
<td>2.4%</td>
<td>5.8%</td>
</tr>
<tr>
<td>f80</td>
<td>5.69 mm</td>
<td>6.41 mm</td>
</tr>
<tr>
<td>f50</td>
<td>1.79 mm</td>
<td>1.95 mm</td>
</tr>
<tr>
<td>p80</td>
<td>3.16 mm</td>
<td>2.88 mm</td>
</tr>
<tr>
<td>p50</td>
<td>718 µm</td>
<td>523 µm</td>
</tr>
<tr>
<td>Percentage Passing -710µm</td>
<td>49.8%</td>
<td>54.8%</td>
</tr>
<tr>
<td>Net Specific Energy Consumption</td>
<td>1.45 kWh/t</td>
<td>1.96 kWh/t</td>
</tr>
<tr>
<td>(-710 µm) Net Specific Energy Consumption</td>
<td>2.91 kWh/t</td>
<td>3.58 kWh/t</td>
</tr>
<tr>
<td>Specific Throughput</td>
<td>304 ts/hm³</td>
<td>232 ts/hm³</td>
</tr>
</tbody>
</table>
The Stirred Mill Circuit

To determine the specific energy requirements for the stirred mill circuit, two signature plots were performed using the minus 710 µm undersize from the second stage of closed-circuit testing. We chose operating conditions to target a specific energy input of 7-9 kWh/t per pass through the mill. This energy input would create an evenly-spaced set of data points on the signature plot and allow for accurate prediction of energy requirements for coarse grind sizes. Table 11 summarizes the operating conditions used for testing and Figure 19 shows the resulting signature plots. Testing showed that an average specific energy consumption of 9.73 kWh/t was required to grind to a p80 of 75 µm. This value was selected to be used for energy calculations of the HPGR / stirred mill circuit.

Table 11 – Summary of stirred mill operating conditions

<table>
<thead>
<tr>
<th>Characteristic</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>F80</td>
<td>340 µm</td>
</tr>
<tr>
<td>Feed Weight</td>
<td>100 kg</td>
</tr>
<tr>
<td>Percent Solids by Weight</td>
<td>57%</td>
</tr>
<tr>
<td>Percent Solids by Volume</td>
<td>31%</td>
</tr>
<tr>
<td>Flow Rate</td>
<td>20.4 L/min</td>
</tr>
<tr>
<td>Mill Speed</td>
<td>1,169 RPM</td>
</tr>
<tr>
<td>Grinding Media Top Size</td>
<td>6 mm</td>
</tr>
<tr>
<td>Grinding Media Type</td>
<td>Ceramic (Manufactured by Cenotec)</td>
</tr>
</tbody>
</table>

Figure 19 – Stirred mill signature plot results

The size measurements used to generate the signature plots in Figure 19 were performed using a Malvern Mastersizer 2000. For a comparison, pass one product was also sized using screens. Since Malvern sizing is based on volume, while screening is based on weight, results will not be identical. All other testwork performed for this flowsheet relied on size results from screening. Therefore, a comparison should be made. Malvern and screening comparisons for T1 and T2 are shown in Figure 20 and Figure 21, respectively. The screening results indicated a finer product than the Malvern results. These results show that the signature plots generated using Malvern sizing, can be considered a conservative estimate for energy consumption, since size results may have been finer using screens. Unfortunately, screening is
impractical below 38 µm (the product size after pass two), so Malvern sizing was used for all stirred mill products in order to remain consistent.

![Particle Size vs. Percentage Passing](image)

Figure 20 – Malvern and screen comparison for T1

![Particle Size vs. Percentage Passing](image)

Figure 21 – Malvern and screen comparison for T2

**Circuit Energy Summary**

The resulting specific energy requirements obtained from pilot-scale testing are summarized in Table 12. For a comparison, results from dry and wet screening for second-stage HPGR are included. Results show that the implementation of wet screening would result in a 4.7% increase in specific energy consumption for the circuit.
Table 12 – Summary of the HPGR / stirred mill energy requirements

<table>
<thead>
<tr>
<th>Unit Operation</th>
<th>Feed f80 (mm)</th>
<th>Product p80 (mm)</th>
<th>Specific Energy Consumption with Dry Screening (kWh/t)</th>
<th>Specific Energy Consumption with Wet Screening (kWh/t)</th>
</tr>
</thead>
<tbody>
<tr>
<td>First Stage HPGR</td>
<td>21</td>
<td>7.68</td>
<td>1.54</td>
<td>1.54</td>
</tr>
<tr>
<td>Second Stage HPGR</td>
<td>7.68</td>
<td>0.34</td>
<td>2.91</td>
<td>3.58</td>
</tr>
<tr>
<td>Stirred Mill</td>
<td>0.34</td>
<td>0.075</td>
<td>9.73</td>
<td>9.73</td>
</tr>
<tr>
<td>TOTAL</td>
<td></td>
<td></td>
<td>14.18</td>
<td>14.85</td>
</tr>
</tbody>
</table>

**DISCUSSION**

Using the energy requirements determined for each circuit, a bar graph is generated to summarize the specific energy consumption for each stage of comminution (refer to Figure 22). The graph shows that the HPGR / stirred mill circuit required the lowest specific energy consumption and achieved a reduction of 9.2% and 16.7% over the HPGR / ball mill and cone crusher / ball mill circuits, respectively.

The results presented in this paper were obtained from pilot-scale testing on a single test for each operating variable. Since pilot-scale testing required a significant quantity of material per test, 350 kg for the HPGR and 100 kg for the stirred mill, the reproducibility and standard deviation could not be determined for each changing variable. However, using five homogenized drums, we did perform some repeatability testing using the pilot-scale HPGR. Results showed that specific energy consumption had a standard deviation of 0.0167 and specific throughput, a standard deviation of 11.43. For stirred mill testing, since only two signature plots were generated at similar conditions, the standard deviation could not be calculated and instead the median of 0.23 was considered. The energy figures associated with Bond grindability testing were found to have a standard deviation of 0.0548, when comparing the three results of HPGR product at a screen size of 106 µm. With these results, testing errors were calculated for each circuit at a 95% confidence interval. Table 13 summarizes the statistics related to each circuit energy result. The error values show that the HPGR / stirred mill circuit contained the most potential for a variation in reported results. With the inclusion of testing error, the HPGR / stirred mill circuit still required the lowest specific energy consumption for comminution. The testing error presented here can be considered only an approximation because the results are based on only a few tests.

![Figure 22 – Summary of specific energy consumption for each circuit](image-url)
The energy values determined in this paper did not take into account any auxiliary equipment for the circuits. Each circuit would require additional energy requirements for feeders, conveyors, screens, pumps and cyclones. Additional energy requirements for the cone crusher / ball mill circuit would result from screens and conveyors for the crushing circuit and pumps and cyclones for the ball mill circuit. For the HPGR / ball mill circuit, increased energy requirements would result from screens and conveyors in the HPGR circuit and pumps and cyclones on the ball mill circuit. The energy requirements for the HPGR / stirred mill circuit would increase with a feed conveyor for first-stage HPGR crushing, screens and conveyors for second-stage HPGR crushing, and pumps for the stirred mill circuit. The extra energy required for the increased quantity of conveyors would be counteracted by the reduction in energy related to an open-circuit grinding configuration. The energy requirements for a de-agglomerator were not necessary for Mesaba ore, due to a low flake competency. However, this energy requirement would need to be considered for an ore that produced more competent flake. Overall, the increased energy requirements for all three circuits, when incorporating auxiliary equipment, should not affect the results significantly.

A preliminary design of an HPGR / stirred mill circuit is shown in Figure 23. The flowsheet design incorporates a closed-circuit crusher prior to HPGR processing. This step prevents oversized material from entering the HPGR circuit, ensuring the prevention of single particle breakage and damage of the metal studs. The design of the secondary crushing circuit is similar to the configuration installed at Cerro Verde (Vanderbeek et al., 2006). A cone crusher is placed in a reversed closed-circuit arrangement, which would reduce throughput and improve crushing efficiency for the cone crusher by screening out fine particles from the feed. A metal detector is placed on a conveyor prior to entering the HPGR circuit, to protect the HPGR roller lining from tramp metal. The two stages of HPGR comminution are designed with the first stage operating in open circuit to reduce materials handling requirement and decrease the amount of water (wet screening) entering the circuit. To achieve efficient screening for the second stage, inclusion of both de-agglomerator and wet screening steps were incorporated to handle competent flake. Alternatively, the possibility arises to implement an air classifier instead of wet screening for fine size classification, eliminating the addition of moisture inherent with a wet screening circuit. The use of an air classifier for fines production in an HPGR circuit has been shown to operate effectively in the cement industry (Aydogan, Ergun, & Benzer, 2006) and with careful design could be implemented in hard-rock circuits. To ensure optimal feed density for efficient stirred milling, undersize from the HPGR circuit would be fed to a mixing tank, where water would be added to control pulp density. Larson, Morrison, Shi, and Young (2008) stated that ideal operating conditions for stirred milling require a solids content of 40-50%, depending on viscosity of the slurry. This could be achieved with a simple process control loop installed between the water addition tank and the IsaMill™ feed tank. For operation of the IsaMill™ circuit, the simplicity available with the dynamic classifier and open-circuit configuration would eliminate any need for a recycle stream.

Table 13 – Statistics summary of circuit energy values

<table>
<thead>
<tr>
<th>Sample Set</th>
<th>Mean (kWh/t)</th>
<th>Standard Deviation</th>
<th>Standard Deviation of the Mean</th>
<th>95% Confidence Interval</th>
<th>Upper Limit</th>
<th>Lower Limit</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cone Crusher Specific Energy Value</td>
<td>3.23</td>
<td>0.0548</td>
<td>0.0316</td>
<td>0.0620</td>
<td>2.42</td>
<td>2.30</td>
</tr>
<tr>
<td>Ball Mill Specific Energy Value</td>
<td>3.15</td>
<td>0.0548</td>
<td>0.0316</td>
<td>0.0620</td>
<td>15.53</td>
<td>15.41</td>
</tr>
<tr>
<td>HPGR Energy Value</td>
<td>2.15</td>
<td>0.0167</td>
<td>0.00747</td>
<td>0.0146</td>
<td>2.16</td>
<td>2.14</td>
</tr>
<tr>
<td>Ball Mill Energy Value</td>
<td>3.14</td>
<td>0.0548</td>
<td>0.0316</td>
<td>0.0620</td>
<td>14.26</td>
<td>14.14</td>
</tr>
<tr>
<td>Stage 1 HPGR Energy Value</td>
<td>1.54</td>
<td>0.0167</td>
<td>0.00747</td>
<td>0.0146</td>
<td>1.55</td>
<td>1.53</td>
</tr>
<tr>
<td>Stage 2 HPGR Energy Value</td>
<td>3.58</td>
<td>0.0167</td>
<td>0.00747</td>
<td>0.0146</td>
<td>3.59</td>
<td>3.57</td>
</tr>
<tr>
<td>Stirred Mill Energy Value*</td>
<td>2.97</td>
<td>0.23</td>
<td>0.163</td>
<td>0.319</td>
<td>10.05</td>
<td>9.41</td>
</tr>
</tbody>
</table>

*Median used instead of standard deviation

Total Specific Energy Consumption

| Cone Crusher / Ball Mill Circuit | 17.83 +/- 0.09 |
| HPGR / Ball Mill Circuit        | 16.35 +/- 0.06 |
| HPGR / Stirred Mill Circuit     | 14.85 +/- 0.32 |

17.83 +/- 0.09
16.35 +/- 0.06
14.85 +/- 0.32
CONCLUSIONS

In this paper we demonstrated that an HPGR / stirred mill circuit is technically feasible and that, based solely on the specific energy requirements for comminution, the novel circuit could achieve a reduction of 9.2% and 16.7% over the HPGR / ball mill and cone crusher / ball mill circuits, respectively. Although this paper did not take into account the energy requirements for auxiliary equipment, the findings documented here provide an incentive to further explore the concept of an HPGR / stirred mill circuit. With the mining industry requiring a reduction in the energy demand associated with its processes, the design and implementation of novel flowsheets, such as the HPGR / stirred mill circuit presented here, should help keep the mining industry sustainable for future generations.

ACKNOWLEDGEMENTS

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EDITORS:

Ken Major
Mineral Processing Consultant
KWM Consulting Inc.
Maple Ridge, B.C.

Brian C. Flintoff
Senior Vice President, Technology
Metso Minerals Canada Inc.
Kelowna, B.C.

Bern Klein
Associate Professor and Department Head
Norman B. Keevil Institute of Mining Engineering
University of British Columbia

Kelly McLeod
Process Consultant

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STIRRED MILLING AT ANGLO AMERICAN PLATINUM

Chris Rule*

1Anglo American Platinum,
55 Marshall Street, PO Box 62179, Marshalltown, Johannesburg, South Africa,

(*Corresponding author: ChrisR@angloplat.com)
STIRRED MILLING AT ANGLO AMERICAN PLATINUM

ABSTRACT

This paper tells the story of the rapid commercial application of stirred milling technology at Anglo American Platinum’s many Concentrators. The first stirred mill – a 10 000 litre IsaMill™, the largest at the time, was commissioned in late 2003. The project development work was presented at SAG 2001. This operation was “proof of concept” and vindicated the decision to scale up the technology; allowing a rapid and extensive roll-out of mainstream regrind mills as well as several further flotation concentrate regrind mills, over the next six years.

Currently 23 stirred mills are in operation with 64 MW of installed drive capacity – this paper describes the projects’ roll-out and the significant results achieved to date. Many operational problems have been solved along the way, some of these will be described. The introduction of the technology has seen a step change in metallurgical results and progressively decreasing operating costs.

KEYWORDS

Process Mineralogy, PGM, Liberation, Plant Optimisation; UG2 reef, Merensky reef, Platreef; Stirred milling; Mainstream Inert grinding, MIG; Ultra Fine Grinding - UFG , Fine Grinding - FG

INTRODUCTION

The ore profile being mined at Anglo American Platinum has changed dramatically in the last two decades. Concurrently the production of PGMs has grown markedly. The ore sources have changed from being wholly based on the Merensky reef, in the 1980s, to the current split where UG2 makes up over 50% of ore treated in the fourteen managed Concentrator plants and is forecast at 16.7 million tonnes for 2011. Platreef at just under 11 million tonnes now ranks second with Merensky now only 4.5 million tonnes; total ore processed including other surface material is in excess of 41 million tonnes for 2011. See figure 1, which shows the profile change in the last 10 years.

![Managed Concentrators - Tons Milled per annum](Figure 1: Ore milled over the last 10 years at Anglo American Platinum Managed Concentrators)

The ore types have very different mineralogical characters which largely determine the metal extraction efficiencies. Unfortunately, the UG2 and Platreef are more mineralogically complex. They have smaller sized PGMs minerals and increasingly higher dissociation of PGM minerals from the larger and hence more easily flotable sulphide minerals; chalcopyrite, pentlandite and pyrrhotite. To compound the plant extraction challenge the head grades delivered have declined over the same time period. The
average feed grade has dropped from excess of 5.5 gpt combined PGMs, in 1999; to approximately 3.2 gpt in 2010.

Historically, AAP mining operations were mainly on the western Bushveld complex with the major operations at the Rustenburg, Amandelbult and Union section mining complexes. The Merensky reef mined there delivered grades often in excess of 6 gpt combined PGMs. See figure 2, for the location of the Anglo Platinum mining operations on the Bushveld complex and inset, a diagram of the Great Dyke in Zimbabwe.

Process mineralogy was recognized a fundamental capability that was required to facilitate a planned programme to address this declining trend in metals recovery. Investment in Anglo Research’s, (previously Anglo Platinum’s Research mineralogy section); has been considerable. (Schouwstra 2004)

In 2004, AAP’s Concentrator Technology function motivated to have a comprehensive programme for each plant’s monthly composites to be analysed by size fraction and further to submit selected samples to be analysed mineralogically. The volume of mineralogical work for twenty or so plants includes the non managed Concentrators; was enormous, so a priority process selected the reduced submissions for mineralogy. This priority rating was dictated by the plant’s relative ounce contribution and an ongoing assessment of the plant’s operational results at any time.

Anglo American Platinum has always had the capacity and desire to investigate potential beneficial new technologies. Investment in process mineralogy has been made bringing in the latest technology and hardware in the preceding years. The knowledge gained on the orebodies grew in line...
with this investment and the intensive exploration programmes on the tenements held by the company on the Bushveld and Great Dyke PGM provinces. Plant troubleshooting studies had already shown the loss profile of the UG2 and Platreef operations was principally due to incomplete liberation and to losses of sub 10 micron liberated PGM minerals. The following illustrations; taken from the monthly composite programme results, illustrates this clearly.

Table 1: Mineralogical association data for typical monthly composites for MF2 UG2 plant samples, size fraction analyses shows potential for further liberation

<table>
<thead>
<tr>
<th>Association</th>
<th>Feed</th>
<th>Concentrate</th>
<th>Tailings</th>
<th>Tailings -10</th>
<th>Tailings +10</th>
<th>Tailings +53</th>
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<tbody>
<tr>
<td>Liberated</td>
<td>49.2</td>
<td>53.1</td>
<td>51.3</td>
<td>62.5</td>
<td>15.5</td>
<td>2.4</td>
</tr>
<tr>
<td>Enclosed in BM/S</td>
<td>23.6</td>
<td>15.8</td>
<td>4.7</td>
<td>4.1</td>
<td>0.7</td>
<td>1.8</td>
</tr>
<tr>
<td>Attached to BM/S</td>
<td>7.9</td>
<td>12.7</td>
<td>0.3</td>
<td>0.4</td>
<td>0.6</td>
<td>-</td>
</tr>
<tr>
<td>PGM/BM/S/Silicate*</td>
<td>6.8</td>
<td>8.0</td>
<td>7.7</td>
<td>-</td>
<td>6.6</td>
<td>15.5</td>
</tr>
<tr>
<td>Enclosed in silicate</td>
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<td>3.4</td>
<td>30.0</td>
<td>2.7</td>
<td>44.3</td>
<td>57.0</td>
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<tr>
<td>Attached to silicate</td>
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<td>2.7</td>
<td>9.3</td>
<td>3.5</td>
<td>13.9</td>
<td>6.8</td>
</tr>
<tr>
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<td>7.6</td>
<td>4.2</td>
<td>6.0</td>
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<td>-</td>
<td>3.1</td>
<td>2.7</td>
<td>1.4</td>
<td>5.0</td>
</tr>
<tr>
<td>Middlings</td>
<td>7.2</td>
<td>14.8</td>
<td>7.0</td>
<td>8.3</td>
<td>14.8</td>
<td>-</td>
</tr>
<tr>
<td>Locked</td>
<td>43.6</td>
<td>32.1</td>
<td>61.7</td>
<td>6.4</td>
<td>66.7</td>
<td>67.6</td>
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*PGM/BM/S/silicate associated refers to PGM particles which are associated with BM/S which are then in turn associated with silicates.

Table 1 shows the potential for further grinding clearly, with the majority of losses of PGMs due to incomplete liberation, both locked and middlings particles in both the +10 micron and especially in the +53 micron fractions. The following illustrations shows the incomplete liberation in tabular form and in the false colour two dimensional images sourced form the MLA analyser.

Figure 3: Number of composite particles containing PGMs by size fraction in a typical UG2 tailings sample. Particle map (false colour), showing the association of PGMs, (red), with mainly gangue minerals in a final tailing sample analysed from the plant composite programme

A history for each plant is built up for the sampled streams; in table 2 the average PGM liberation in tailings is seen at low levels of 20-35% and generally these are in the sub 10 micron fraction which are more difficult to float.

Clearly then an economic comminution technology that could improve PGM particle liberation economically and without generating large amounts of “super-fines” would potentially be a key to reaching much higher PGM recoveries. Stirred milling potentially was that technology.
Table 2: Typical UG2 Concentrator monthly composite plant tailings data – assay and mineralogy

<table>
<thead>
<tr>
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<th></th>
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</thead>
<tbody>
<tr>
<td><strong>Grade (at% PGE)</strong></td>
<td>0.9</td>
<td>1.0</td>
<td>1.0</td>
<td>0.9</td>
<td>0.7</td>
</tr>
<tr>
<td><strong>Tailings d50 (µm)</strong></td>
<td>50</td>
<td>63</td>
<td>47</td>
<td>38</td>
<td>67</td>
</tr>
<tr>
<td><strong>All silica (mass %)</strong></td>
<td>2.4</td>
<td>2.5</td>
<td>3.3</td>
<td>3.3</td>
<td>2.1</td>
</tr>
<tr>
<td><strong>EMS liberation (%)</strong></td>
<td>56</td>
<td>60</td>
<td>59</td>
<td>72</td>
<td>69</td>
</tr>
<tr>
<td><strong>PGM liberation (%)</strong></td>
<td>24</td>
<td>27</td>
<td>33</td>
<td>33</td>
<td>24</td>
</tr>
<tr>
<td><strong>PGM lib (%)-10µm</strong></td>
<td>82</td>
<td>97</td>
<td>78</td>
<td>90</td>
<td>92</td>
</tr>
</tbody>
</table>

At the beginning of 2000, with the recent commercial availability of IsaMill™ stirred mill technology; a 4 litre bench scale mill was assessed at AAP’s research and development facilities along with other vertical stirred milling technology units. The three primary requirements for further regrinding in the PGM circuits were:

- Due to mainstream tonnages a large stirred mill was targeted; at that time the largest mills available were 1.1 MW 3000 litre for the IsaMill™ technology and for vertical stirred milling the 355 kW SMD. Multiple small units were not favoured,
- Inert or non steel media grinding was the preference due to potential surface chemistry and slurry chemistry issues with iron hydroxide precipitation and coatings on value minerals,
- The technology should be energy efficient and obviously must be economically viable with an expected recovery improvement potential of a further 2-5% PGM recovery.

Promising results lead to acquisition of larger units that could be used in pilot plant studies. Results at pilot scale led to the decision to use IsaMill™ technology in the new tailings re-treatment project, “WLTR” project, near Rustenburg. [5] The significant scale up decision, from a 3000 litre unit to a 10000 litre unit was made at that time by the author, thus facilitating the economic use of the technology for future mainstream applications and once a suitable ceramic media had been identified or developed. The initial mill was developed in a three party collaboration with Netzsch Feinmetal GmbH and Xstrata Technology; then known as MIM Technology. To mitigate the scale up risk a variable speed drive was chosen – the drive train was 2.6 MW due to the initial media being locally mined crushed and 2-5mm, screened silica sand.

**THE ISAMILL™ TECHNOLOGY ROLL-OUT PROGRAMME**

The circuit applications identified as being the optimal circuit interventions are illustrated in the following block flow sheets, figure 4 and figure 5. The mainstream application – given the name “MIG” or mainstream inert grinding was added on regrind ball mill product ahead of scavenger rougher flotation; resulting in better mainstream PGM extraction. The “UFG” or ultra fine grinding application was targeted at modifying the typical plant grade/recovery relationships in the cleaning circuits for the mainstream flotation products and results in better product grades and higher circuit PGM recovery by regrinding composites and applying intensive surface attritioning. The two acronyms were created to prevent confusion and recognized the essential features; MIG - inert grinding using non steel media in mainstream applications, typically targeting 80% -53 micron products and UFG, concentrate regrind targeting 80% less than 20 microns. It would be more correct to call this application “FG” or fine grinding to differentiate from the very fine regrinding targeting products that are sub 10 micron common in lead/zinc production and in leaching applications.

This typical PGM Concentrator, “MF2” flow sheet shown in figure 4, incorporating the stirred milling technology is applicable to the Platreef and Merensky plant circuits.
Mainstream Inert Grinding (MIG) is designed to improve recoveries by optimising liberation in an 'inert' media environment.

Ultra-Fine Grinding (UFG) reduces the mass pull through better liberation, surface cleaning and improving the selectivity between gangue and value minerals.

For UG2 reef, due to its essentially bimodal form – a mix of silicate minerals and chrome spinel minerals in the chromitite reef; a variant of this MF2 circuit had been developed. The new circuit applied MIG stirred milling technology in the split regrind circuit. The primary rougher flotation tailings are split with a hydrocyclone cluster prior to further processing. This produces two streams; an underflow rich in higher specific gravity and coarse chromite spinel and the overflow enriched in lower specific gravity silicates and fines. The majority of PGMs follow the silicates; their occurrence in the in-situ ore is predominantly in the silicate minerals in between the chromite spinel crystals. Either the MIG IsaMill™ follows a regrind ball mill or it treats the cyclone overflow stream directly; dependant on the available milling units at the plant.

Figure 4: Process flow diagram of the application of stirred milling as 1) mainstream or MIG and 2) concentrate regrind or UFG is a typical PGM industry stage grinding and flotation, or MF2, circuit.

Figure 5: Process flow diagram for a split regrind MF2 circuit for UG2 ore treatment – shows the MIG IsaMill™ on the cyclone overflow and the open circuit ball mill on the cyclone underflow.
The adoption of IsaMill™ stirred milling was a relatively fast adoption of a new technology. Key steps in this process can be listed:

- **2000/2001**: Scoping work carried out at AAP research facilities, “ARC”, at Germiston, east of Johannesburg and at the operational support centre, “DML” and pilot plant at Rustenburg. The pilot studies confirmed the potential for both mainstream stirred milling, “MIG” and concentrate regrind, “UFG” in mainstream flotation product cleaning.
- **2001**: Decision taken to use IsaMill™ stirred milling technology and to develop the first M10000 litre unit; development within a tripartite collaboration with Xstrata Technology, (then known as Mt Isa Mines Technology) and Netzsch Feinmetal GmbH. Key was the choice of scaled up unit; this was to allow future mainstream applications to be realized. Silica sand chosen as the grinding media in the absence of a proven ceramic media.
- **2001/2003**: Inclusion of UFG concentrate regrind M10000 IsaMill™ with variable speed 2.6 MW drive to mitigate design scale-up inaccuracy; in the tailings retreatment project, WLTR; at Rustenburg. Successfully commissioned in early 2004; proof of concept established. (Buys et al. 2004)

![Aerial photograph showing the WLTR site and a graph showing the change in PGM grade/recovery potential by applying UFG stirred milling of mainstream flotation concentrates](image)

**Figure 6: Aerial photograph showing the WLTR site and a graph showing the change in PGM grade/recovery potential by applying UFG stirred milling of mainstream flotation concentrates**

- **2004/2005**: Initiation of a comprehensive off-site and on-site pilot programme with on-site piloting of various stream feeds using a M20 IsaMill™ unit and associated “FCTR” mini flotation rig. Initiation of the routine monthly composite plant sample programme. Fractional analyses and mineralogical analyses on plant feeds, plant flotation product and plant tailings. Realization of the economic viability and then definition of the priority for establishment of a wide scale roll-out of IsaMill™ technology for both mainstream MIG and concentrate regrind UFG applications at major AAP Concentrators. Identification of an economically viable ceramic grinding media for MIG applications. (Rule et al., 2008)
2005/2006; the initiation of the ceramic media development programme. Silica sand, furnace slag and other natural medias could not be used in MIG applications due to the relative hardness of the ore and media. With the approval of the first MIG project at Mogalakwena South; the approving AAP body required that a multi-media ceramic supply base be put in place for the technology.

This was achieved by setting up a testing facility which included utilising both the 4 litre and 100 litre IsaMills at the Divisional Metallurgical Laboratory operational support facility in Rustenburg. A comprehensive test programme was established to allow assessment of the ceramic medias available in the world market. Simultaneously major suppliers were contacted and engaged with. Later in 2007, further QA/QC capability was established in order that ceramic media deliveries could be checked for adherence to specification. The programme to date has tested some 200 different medias from all the major ceramic media manufacturers and has led to the development of several improved formulations with various suppliers. A ceramic media manufacturing facility in South Africa was established by a third party in 2007. This proactive involvement has led to significant operating costs benefits and indeed the media consumable cost on a US$ per kWhr consumed by the IsaMills has dropped by more than 2/3rd from 2008 to 2011. (Bedesi et al, 2008)

2006; commissioning of the first mainstream or MIG IsaMill™ at Mogalakwena South Concentrator – C section project utilizing a 3MW M10000 unit; note the use of higher specific gravity media resulted in increase in drive train capacity. This mill installation proved the MIG concept was a workable technology and removed any risk constraint for the full roll-out of the technology throughout AAP operations.
2007/2008/2009: motivation, approval for execution and installation of a further 20 IsaMill™ units in MIG, (17 M1000 units) and UFG, (3 units, 2 M1000 and 1 M3000) duties at Bafokeng Rasimone, Waterval UG2, Waterval, Amandelbult Merensky, Amandelbult UG2 1 and UG2 2, Mogalakwena South A and B and Mogalakwena North Concentrators; (Rule et al. 2010)

Figure 9: IsaMill™ layout photographs, Waterval UG2, Amandelbult UG2#2 and Waterval Concentrators

Figure 10: MIG installations at Bafokeng Rasimone, top left, shows the typical “wrap around” equipment in the circuit and on top the right hand side; Mogalakwena South Concentrator; A and B mills installed on far left hand side. Below these are the model drawings for A and B section IsaMills

OPERATING EXPERIENCE DURING THE EARLY PERIOD

There were remarkably few problems with the first M10000 installed at the WLTR Project in 2003. The scale up resulted in operational performance that was almost exactly as predicted; hence the variable speed drive unit is/was seldom utilized. Test work for future MIG application on tailings material in 2008 to 2010 was conducted and some work was done at variable speed regimes with varying media loads etc.

Following the first MIG milling application start up of the Mogalakwena C section mill in late 2006; the success resulted in the rapid approval of the next four MIG mills, two each at Mogalakwena South; A and B and at Waterval UG2 with commissioning by the end of 2007.
A major difficulty identified initially was the management of the position of the media load. Variability in mill feed volume and slurry density resulted in incidents of media compression at low flows and media loss at high flows. The media compression incidents resulted in the early modification of the internal wear discs configuration, with the disc in position 1, i.e. nearest the feed inlet; being removed totally, effectively creating a larger first milling compartment. Modifications to the dimensions of the feed end discs have been made with reduced diameter discs being successfully trialled. Too high feed slurry flows result in significant media loss – seen visually downstream and also measured in higher than expected media losses.

![Figure 11: Photographs of grinding discs made of different rubber compounds and experimental ceramic tile shell liners after use](image)

A number of materials of construction test runs have been conducted; figure 10 shows the mill internals of an M10000 and some ceramic shell liners using tiles that were unsuccessfully trialled.

A summary of issues that have been addressed to date between 2008 and 2009 follows:

i) an inordinate number of drive train bearing over temperature trips were encountered on some of the mills – the bearing lubrication system was modified to a circulated/cooled oil rather than grease based system, before the modifications were completed the mills were run at lower power levels to reduce mill stoppage events

ii) The incidence of damage from tramp material became a very significant cost and operating downtime issue for the MIG installations particularly from the middle of 2008; this included premature severe damage to internals from repeated ingress of steel ball fragments, nuts and bolts, ceramic wear tiles, (from upstream launders/still boxes etc) and coarse oversize excursions from upstream ball mills ahead of the IsaMill™.

![Figure 12: Photographs showing metal steel ball scats and “rust” on surface of some of the 3.5mm top size media in the dumped mill media and resting on the grating below the mill – damage to the rubber and polyurethane wear parts in the mill was severe and rapid!](image)

iii) At times poor judgement on the wear life of grinding discs led to disc collapse and severe damage to mill internals from the broken pieces.
Whilst the problems were analysed the MIG mills especially were run at lower loads and in the extreme case where multiple internal damage incidents occurred the mills were taken off-line at Waterval UG2 for 6 months at the end of 2008.

A rapid identification of root cause and development of solutions phase followed and in 2009 when the next 16 mills were commissioned the design fixes were auctioned during the installation and before commissioning; fixes were actioned including to the first 6 mills which were taken offline for modifications:

a) Media position within the mill was controlled by modifying the existing slurry recycle stream allowing fixed volume flow set-point to be implemented in the control system; slurry density was controlled by optimizing the cyclone parameters upstream and optimization of the IsaMill™ feed circuit surge tank buffer control system, the 7 disc configuration was retained,

b) A linear screen was installed on the circuit feed system ahead of the surge tank feeding the mill density controlling cyclone clusters – a physical barrier to oversize material greater than 2 mm was thus imposed on the circuit. An interesting finding during the analyses exercise at one of the sites was the occurrence of copious fragments of steel in the mill contents – this was discerned form analyses of the dumped media where “rusting” as well as large fragments “spalled” from the steel ball charge of the primary mill was seen. How large fragments >20mm were able to pass through the upstream primary circuit closing vibrating screens with decks panels of 850 microns aperture, travel through the primary rougher flotation and then enter the MIG IsaMill™ circuit via the densifying cyclone cluster fed from the surge tank pumps begs the question around operating and maintenance vigilance and control!

c) Where ore quality processed varied another issue was apparent and is shown in the two photographs following, figure 14. The Mogalakwena South Concentrator occasionally treats extremely hard ore and during these periods the grind coarsens appreciably – exceeding the capacity of the FAG and ball mills ahead of the two A and B MIG IsaMills to provide a fine enough feed. Critical size build up results! The problem has been addressed by bringing back into the circuit the two redundant tertiary ball mills which now provide a further comminution step ahead of the MIGs.
d) Further more rigid operating recipe driven operation was rolled out across the sites – resulting in tighter control over all circuit parameters not just those impacting on “spikes” in volume and density flow through the circuit. A higher level of operating behaviour is now common place throughout the group Concentrators with benefits seen performance in all unit processes.

e) The application of a comprehensive data collection from both the DCS system, remotely, to the group control centre in Rustenburg and maintenance supervision programme on all mills was put in place; data on media consumption and stock levels is also collated and reported. This comprehensive data set and reports has prevented the incidence of wear failures to be almost eliminated by 2010. A monthly report is generated to all sites and support personnel and in parallel the routine wear data is collected at each maintenance stop for all mills and analysed and reported co-currently. Note all the wear supervision is completed by Xstrata Technology personnel and is used routinely and for testing for mill component and circuit optimization.

OPERATING RESULTS

The change in performance post MIG IsaMilling is well illustrated at the Mogalakwena South Concentrator, Waterval Concentrators and the Amandelbult Concentrators; the post commissioning period results at the Rustenburg and Amandelbult sites have been significantly better than predicted.

The initial evaluation for the Mogalakwena South Concentrator comparing the results post commissioning of AAP’s first MIG IsaMill™ on the “C” section ore processing line against the A and B section ore processing lines led to the rapid approval of further MIG mills at Waterval UG2 and for “A” and “B” sections. The statistical analyses conducted on the pre and post commissioning period concluded a PGM delta recovery increase of between 3.5 and 4%; due to MIG stirred milling in the mainstream circuit, refer figure 15.

Figure 15: Results from Mogalakwena South after commissioning of the first MIG mill on “C” section compared to the other two lines, “A” and “B” operating with MIG IsaMilling – statistical analyses showed a 3-4% delta recovery improvement for the technology.
Further assessment of the UG2 ore impact at the Waterval UG2 plant post commissioning in 2007 and through the early part of 2008 led to approval for the roll-out of the next phase of the technology’s application on UG2 and Merensky/UG2 mixed ore Concentrators – taking the IsaMill™ fleet deployed to 22 units and 64 MW of installed drive train installed power.

Rustenburg and Amandelbult mining operations are AAP’s biggest production units with roughly 2/3rd of the 2010 total managed Concentrators’ platinum output of 1.7 million troy ounces. Roughly 2/3rd of the mined and processed tonnage in 2010, 10 million tonnes; was sourced from the UG2 reef.

Figure 16 shows the impact on metals recovery since the commissioning period – an upward trend at both sites for platinum recovery. Tailings flotation scavenging achieves further metal recovery by allowing a lower grade product to be extracted from each site’s combined tailings; before disposal to surface tailings dams.

Analyses of the performance of the two UG2 plants at Amandelbult shows the reduction in tailings grades from the two plants over time; the dramatic step reduction since MIG IsaMill™ commissioning stands out. Tailings grades are now at the lowest levels since Amandelbult mining operations began in the 1970s.
The graph shown in Figure 16 shows the comparison of PGM recovery values achieved for UG2 processing at the two mining sites at Amandelbult and its neighbour Union, some 30 kilometres away. Historically PGM metal recoveries at the Mortimer UG2 plant at Union have always been similar to that achieved at Amandelbult and averages around 80-82% extraction of PGMs. Since MIG IsaMill™ commissioning a strong divergent and upward trend is seen for the Amandelbult operation.

CONCLUSIONS

The journey to fully install stirred milling technology in AAP’s Concentrator operations represents a very fast introduction of a new technology, by industry standards.
The economics of the recovery improvement potential at the plants is an overwhelming proposition due to the increasing value, with time, of the contained metal in the ROM ore. This justifies the level of investment in process mineralogy and a new comminution technology already made and continuing to be made at Anglo American Platinum. To illustrate this it is a sobering fact that an improvement of 1% in PGM recovery in the Concentrators at Anglo American Platinum is equivalent to >75 million US$ for each year at today’s metal prices and exchange rates.

Currently, further stirred mill installations are being motivated. Test work and circuit analyses has identified that further upside potential at existing operations remains to be achieved as the operations are optimised further.

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IsaMill™ TECHNOLOGY IN THE PRIMARY GRINDING CIRCUIT

*G.S. Anderson¹, D.T. Smith², S.J. Strohmayer²

¹Xstrata Technology
Level 4, 307 Queen Street
Brisbane, Australia, 4000
(*Corresponding author: GAnderson@xstratatech.com)

²Xstrata Zinc
McArthur River Mine
Northern Territory, Australia
IsaMill™ TECHNOLOGY IN THE PRIMARY GRINDING CIRCUIT

ABSTRACT

Originally developed as a step change in ultrafine grinding efficiency, the IsaMill™ has made inroads into conventional ball mill and tower mill mainstream grinding duties during the last 5 years - beginning with the Anglo Platinum mainstream grinding projects in South Africa. At the 2006 SAG Conference, plant testwork on the incorporation of IsaMill™ technology into the McArthur River (Northern Territory, Australia) Pb/Zn primary grinding circuit was reported. The IsaMill™ has now become an integral component of that circuit, treating up to 1mm top size feed and directly producing P80 45 µm rougher flotation feed. Two 1.1 MW IsaMills™ are now used in this duty contributing to a 27% increase in overall plant throughput and a reduction in rougher feed size distribution. In this paper, we report on the current state of development and performance of the IsaMill™ in the primary grinding duty at McArthur River and also consider future directions.

KEYWORDS

IsaMill™, McArthur River, efficient, energy, ceramic, inert, primary grinding, stirred milling.

INTRODUCTION

IsaMill™ technology was developed to address the ultra fine grinding (UFG) requirements of Mount Isa Mines (MIM, now Xstrata), particularly for the McArthur River lead/zinc deposit where a 7 µm grind was required, under efficient and inert conditions, in order to make a salable concentrate (Logan, Leung & Karelse, 1993). The 1.1 MW, 3000 litre IsaMill™ (M3000) became the enabling technology for McArthur River in 1995 and remains critical to its ongoing viability today. The history of McArthur River, the development of the IsaMill™ - between MIM and Netzsch Feinmaltechnik of Germany, as well as the operating characteristics of the IsaMill™ have been well documented elsewhere (Nihill, Stewart & Bowen, 1998; Enderle, Woodall, Duffy & Johnson, 1997; Pease, Anderson, Curry, Kazakoff, Musa, Shi & Rule, 2006).

Since its development and use for UFG duties, the IsaMill™ was commercialised and has now moved into coarser grinding, mainstream applications; allowing the key benefits of improved energy efficiency, small footprint and inert grinding environment to be applied to a range of different mineral types (Anderson & Burford, 2006). Key to this development was both the scaleup of the IsaMill™ from 3,000 litres (1.1 MW, M3000) to 10,000 litres (3 MW, M10,000) (Curry, Clark & Rule, 2005) and the development/use of ceramic media (Curry & Clermont, 2005).

Anglo Platinum lead the way in the rollout of IsaMills™ into mainstream duties based on the successful commissioning and operation of the first 10,000 litre IsaMill™ (in an ultra fine grinding application) in 2003 (Buys, Rule & Curry, 2005) and the significant evidence generated by Anglo Platinum’s own mineralogical studies, suggesting that finer grinding was required to improve PGM recoveries, particularly across the Platreef and UG2 orebody types (Rule, 2010). Between 2006 and 2009, 54 MW of MIG (Mainstream Inert Grinding) IsaMills™ were commissioned across the Anglo Group in South Africa. Despite some early commissioning and operational issues, the IsaMills™ have proved their value adding in excess of 3% to platinum recoveries (Rule & de Waal, 2011). Anglo Platinum currently has a total of 18 IsaMills™ in MIG duties and four IsaMills™ in UFG duties.

At McArthur River, life of mine optimisation studies determined that conversion from underground to open cut mining was required, with a subsequent increase in plant throughput requiring additional grinding capacity. Given their history of development and operational familiarity with the IsaMill™, McArthur River initiated their own testwork into coarse grinding applications around the same
time the Anglo projects were proceeding to full scale, with a view to investigating the potential of the IsaMill™ to contribute to the expansion of the primary grinding circuit at McArthur River.

MRM PRIMARY GRINDING CIRCUIT EVOLUTION

When commissioned in 1995 with nameplate capacity of 1.5 mtpa, the MRM primary grinding flowsheet consisted of a single stage 3.5 MW SAG mill. Mill discharge was screened over a 1.8 mm screen with the oversize recycled back to the SAG mill feed chute. Screen undersize was pumped to a set of cyclones that produced an overflow at around P<sub>80</sub> 45 µm for rougher flotation. Cyclone underflow returned to the SAG mill feed chute.

Over time, ore sourced from the initially targetted high grade number 2 orebody depleted. Additional ore was sourced from the number 3 and 4 orebodies with interbed inclusions. This material was harder and at lower head grade. As a result, numerous changes were implemented to the primary grinding circuit to increase throughput, including additional crushing stages ahead of the SAG mill, increased SAG ball loads, uprating the SAG Mill to 4 MW and the addition of a locally sourced second hand tower mill. This allowed plant throughput to be increased to 1.8 mtpa by 2006 but also resulted in the primary grind size increasing from P<sub>80</sub> of 45 to 70 µm. The 2006 flowsheet (after crushing) is shown in Figure 1.

![Figure 1 - McArthur River Primary Grinding Circuit 2006](image)

In 2005, a change from an underground operation to an open cut operation was under consideration, requiring the plant to process up to 2.5 mtpa - at further reduced head grade.

IsaMill™ LAB AND PILOT WORK

A testwork program commenced to determine the suitability of using the IsaMill™ to increase the plant throughput at McArthur River. Initial scoping work was carried out at laboratory scale in 2005, followed by two separate on site pilot plant testwork campaigns in 2006 using an 18 kW 20 litre (M20) IsaMill™ - this was reported at SAG2006 (Pease et al, 2006). The on-site pilot work, treating a bleed stream from the SAG mill classification circuit, predicted that a product size P<sub>80</sub> of 45 µm could be produced at approximately 10 kWhr/t using the IsaMill™.

FULL SCALE 1.1MW M3000 IsaMill™ TRIAL

Following the success of the pilot scale work, one of the six 1.1 MW M3000 UFG IsaMill™ units was reconfigured to operate in the MRM primary grinding circuit to confirm the results from the pilot work. The circuit was configured as per Figure 2 with the bleed stream originating from a modified SAG cyclone. A magnetic separator was included on the feed stream to protect the IsaMill™ from tramp SAG Mill steel - which had been identified as a potential issue during the pilot work. The IsaMill™ operated in open circuit...
with the product joining the SAG circuit cyclone overflow as feed to the rougher flotation circuit. The full scale trial commenced in May 2007 using 3.5 mm ceramic media with a feed size of \( F_{80} \) of 180-200 \( \mu \text{m} \). Figure 3 shows the 6 x M3000 IsaMills™ at McArthur River.

Figure 2 - Circuit Configuration for Full Scale IsaMill™ Trial

Figure 3 - 6 x M3000 IsaMills™ at McArthur River

This application stepped well outside the boundaries of previous IsaMill™ experience and a development phase was needed to optimise the grinding conditions to produce a satisfactory result. As for SAG or ball milling, in stirred milling it is also crucial to break the top size particles fast enough to avoid critical size buildup within the mill. The initial mill internal configuration and 3.5 mm media was unable to do this efficiently or quickly enough, resulting in the IsaMill™ becoming unstable after several hours of
operation due to a critical buildup of unbroken coarse material at the front end of the mill. This was confirmed during crash stop inspections of the mill which revealed a large volume of unbroken compacted and dewatered coarse material at the front of the mill and internal wear patterns indicating the incoming feed bypassing this built up stationary area. The buildup of stationary media and unbroken coarse material caused a 30% drop in power draw and resulted in the product from the mill becoming unacceptably coarse due to an effective reduction in the specific energy applied. This was initially overcome by changing to a coarser 5-6 mm top size media and the derating of the internal product separator rotor, which effectively decreased the degree of media compression at the front of the mill and allowed the media and fresh feed to agitate correctly, allowing the coarse feed particles to be ground.

Following several months of development and optimisation, a mill configuration and operating parameter regime was arrived at that resulted in the IsaMill™ being able to treat an F 80 of 200-250 µm to produce a P 80 of 35-45 µm at 10-15 kWhr/t. When the IsaMill™ was operating at 800-900 kW, treating approximately 60-70 tph of material, that would otherwise have been returned to the SAG Mill circuit, the overall plant throughput was able to be increased by 12%, from 260 tph to 290 tph, at an overall combined grind size P 80 of 55-65 µm.

Due to the success of the trial and the impact on the overall plant production, it was continued as a permanent addition to the primary grinding circuit. Eventually a second M3000 IsaMill™ was converted to operate in the primary grinding duty allowing a further increase in overall plant throughput.

GRINDING SURVEY DATA

Numerous surveys across the IsaMill™ circuit were taken under varying operating conditions including changes in internal disc configuration and wear compounds, media type and size, feed size distribution and ore types. Figure 4 illustrates some of the survey data collected over the last two years – all using 5-6 mm ceramic media, grouped by feed size. This is plotted against the graph originally published in SAG 2006 illustrating the testwork data and the predicted operating point for the IsaMill™.
It is evident that, while there is a fair degree of scatter, the points are generally in agreement with the size-energy relationship predicted from the pilot testwork. The different feed sizes show a fair degree of overlap – indicating that the feed size did not have a noticeable impact on the product size-energy relationship. Of note are the two samples where the feed size was F<sub>80>300</sub> µm which are within the grouping of data generated from the feed sizes that were F<sub>80<200</sub> µm. This suggests that, of the energy consumed in grinding to the product size, very little is required to reduce the top end of the distribution. This is however contingent on the fact that the media is large enough to break the particles at the top end of the distribution at an adequate rate.

**EFFECT OF MEDIA SIZE**

Ceramic media was trialled at two different sizes in the IsaMill™. Initially the same 3.5 mm ceramic used during the pilot plant work was used. This was changed out to a 5-6 mm top size ceramic during the early stages when there were problems with achieving the grind size and mill operational stability. Once the issues had been resolved, the 3.5 mm media was retrialled. It was found that both media sizes were able to grind the IsaMill™ feed without critical buildup of unbroken coarse material occurring, provided the density was controlled below 40% solids at the targetted specific energies. However, significant differences in the breakage rates and resultant product size distributions of the two media sizes were noted.

Figure 5 compares the 3.5 mm top size media and the 5-6 mm top size media where the same product P<sub>80</sub> size of 37 µm was produced. The feed size distribution was slightly coarser for the 5-6 mm media case. The 3.5 mm media consumed 16 kWhr/t to produce the P<sub>80</sub> of 37 µm and P<sub>98</sub> of 220 µm. The 5-6 mm media required only 11 kWhr/t to produce the same P<sub>80</sub> of 37 µm but a much finer P<sub>98</sub> of 110 µm. Therefore, the 5-6 mm media was able to produce the same P<sub>80</sub> as the 3.5 mm but consumed less energy and produced a tighter size distribution through more effective breakage of particles at the coarse end of the feed distribution.

![Figure 5 - Performance of 3.5 mm Media at 16 kWhr/t and 5-6 mm at 11 kWhr/t](image-url)

Figure 5 - Performance of 3.5 mm Media at 16 kWhr/t and 5-6 mm at 11 kWhr/t
The 5-6 mm media was able to successfully break the top size of the feed due to its ability to impart greater stress intensity as a result of its increased size (stress intensity is proportional to the cube of the media diameter). As a result, the 5-6 mm media was more efficient at achieving a given product target size. This reinforces one of the key points of stirred milling at coarser sizes – that the top size of the media must be chosen to adequately break the top size of the feed. If the top size is not broken quickly enough, it will remain in the mill leading to higher wear rates, coarser and wider product distributions, lower power efficiency and in worse case scenarios, sanding of the mill.

A second example, in Figure 6, shows the same curve at 16 kWhr/t for the 3.5 mm top size media with the energy input for the 5-6 mm top size media increased to 16 kWhr/t as well. Again, the feed size distribution for the 5-6 mm media was slightly coarser than that for the 3.5 mm media. The product size distribution from the 5-6 mm media ($P_{90}$ of 29 µm and a $P_{98}$ of 75 µm) was superior to that produced by the 3.5 mm media ($P_{90}$ of 37 µm and $P_{98}$ of 220 µm). The improvement was driven by the increased breakage rates at the top end of the distribution.

Based on this data and the fact that it was aimed to increase the target feed size distribution to a $P_{80}$ of +250 µm, 5-6 mm top size media became the standard for this project. This size media has proved adequate in handling the range of feed size distributions fed to the IsaMill™ up to an $F_{80}$ of 350 µm.

The best example of coarse particle breakage was achieved when the feed to the IsaMill™ was significantly coarsened for a two day trial period through adjustment of the sample preparation cyclone. This produced feed distributions with $P_{80}$ values >350 µm. Results of the two surveys taken during this period are included in the Figure 4 data. The distribution curves for one of the data points is shown in Figure 7 at a specific energy of 12.8 kWhr/t.

Figure 6 - Performance of 3.5 mm Media at 16 kWhr/t and 5-6 mm at 16 kWhr/t

Based on this data and the fact that it was aimed to increase the target feed size distribution to a $P_{80}$ of +250 µm, 5-6 mm top size media became the standard for this project. This size media has proved adequate in handling the range of feed size distributions fed to the IsaMill™ up to an $F_{80}$ of 350 µm.

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This can also be compared to the 11 kWh/t grind for 5-6 mm media (used in Figure 5), for a much finer feed size distribution. This reiterates the point made in the discussion of Figure 4 where the feed size distribution seems to have little impact on the energy required to reach a certain grind size. For a given media size, and provided that size is adequate, the overall energy consumption is largely determined by the target product size, rather than the feed size distribution.

Since the project was commenced a number of ceramic medias from different suppliers were trialled with the aim of optimising performance on a supplier cost/consumption basis. Media wear rate in
terms of g/kWh was monitored on a continual basis to allow the different media types to be compared on a consumption basis. Table 1 summarizes the data for the 5-6 mm media top size.

Table 1 - Media Performance Data

<table>
<thead>
<tr>
<th>Media</th>
<th>Cost $(AUD)/t</th>
<th>Consumption g/kWh</th>
<th>Max $/kWh $</th>
<th>Relative Max Cost</th>
</tr>
</thead>
<tbody>
<tr>
<td>Media A</td>
<td>$ 4.60</td>
<td>11-14</td>
<td>$ 0.064</td>
<td>1.00</td>
</tr>
<tr>
<td>Media B</td>
<td>$ 2.70</td>
<td>11-16</td>
<td>$ 0.043</td>
<td>0.67</td>
</tr>
<tr>
<td>Media C</td>
<td>$ 4.60</td>
<td>11-14</td>
<td>$ 0.064</td>
<td>1.00</td>
</tr>
<tr>
<td>Media D</td>
<td>$ 4.60</td>
<td>11-14</td>
<td>$ 0.064</td>
<td>1.00</td>
</tr>
<tr>
<td>Media E</td>
<td>$ 4.60</td>
<td>7-8</td>
<td>$ 0.037</td>
<td>0.57</td>
</tr>
<tr>
<td>Media F</td>
<td>$ 4.00</td>
<td>10-17</td>
<td>$ 0.068</td>
<td>1.06</td>
</tr>
</tbody>
</table>

The data gathered has not been able to discern any quantifiable information with regards to the effect of the different media types on grinding efficiency or mill internal wear. It is likely that these factors do exist but the changes have not been significant to be highlighted within the noise of the data obtained here and at the relatively coarse grind sizes. Xstrata Technology has previously conducted numerous controlled laboratory scale tests that highlight the effect of different media types on grinding efficiency. These effects are also more pronounced at finer (<20 µm P80 target) sizes than were targeted here.

COMPONENT WEAR RATES

In the ultra fine grinding duty at McArthur River, IsaMill™ inspections are typically at 2,000 hours. Typically a shell liner lasts 2,000 hours, feed flange lasts 10,000 hours, discharge flange lasts 15,000 hours and 3-4 of the 8 discs are replaced each inspection. While typical component lifetimes are longer - eg shell liner life is 6,000+ hours at Mt Isa – it is lower at MRM due to the high operating temperatures involved and the inability to rotate the shell liner due to the old style shells used. Newer IsaMills™ have a shell liner that can be rotated to maximise the rubber wear, therefore extending the useable life of the liner.

Initial wear rates of the IsaMill™ internal components in the primary grinding duty were significantly higher than UFG duties. The change in wear rates is a function of the feed size distribution and the presence of more gangue mineralogy - particularly silicates. In this case the feed size distribution increased from an F50 of approximately 45 µm to 200-350 µm and the silicate content of the ore is about 26-35% compared to 17-20% in the UFG duty. It should be noted that the use of coarser grinding media, whilst impacting the liners with a higher stress intensity due to its larger size, actually assists to reduce the amount of time the coarsest particles spend in the mill and therefore limit their influence on wear. This was demonstrated during the change from 3.5 mm to 5-6 mm media where it was assumed that the wear rates would increase due to the coarser media presence, however the overall wear rate remained the same due to the reduction in time the coarsest particles spent in the mill. Figure 9 shows some data from another IsaMill™ treating two separate feed types. The media size was the same in both cases but the change in feed size from 80% -75 µm to 40% -75 µm resulted in a nearly 5 fold increase in disc consumption. It is likely that larger media would have reduced the overall wear rate in the mill at the coarser feed size distribution.
In the initial primary grinding trials, the MRM M3000 IsaMill™ was serviced every 400 hours and usually required replacement of the shell liner as well as a complete set of discs. While it was still proving beneficial to the plant to operate the IsaMill™ in this duty there was a real need to reduce the maintenance cost and time.

Based on successful development work for the M10,000 IsaMills™ in South Africa (Rule & de Waal, 2011), a trial using reduced diameter discs (1,050 mm compared to standard 1,200 mm) was made on the M3000 IsaMill™ at McArthur River. The first 7 discs were replaced by the smaller discs, the final disc was left at 1,200 mm so as not to influence the product separator performance. The mill was able to run for 1,200 hours between inspections where a full disc and liner change was required. The drawback was a (not unexpected) reduction in power of about 35% as a result of the reduced diameter, limiting the power draw to about 650-700 kW maximum. The reduction in power draw limited the throughput by the same percentage in order to maintain the same grind performance. The disc consumption per MWh (Figure 10) reduced by more than half, however is still significantly higher than the data presented in Figure 9. Measures to address the power draw issue are discussed later in the paper.
Based on the positive results from the M3000 IsaMill™ testwork and subsequent conversion to full time production, MRM made the decision to purchase and install two 3.0 MW M10,000 IsaMill™ units with the intention that they be configured in the primary grinding duty for the plant expansion. During the project construction phase, the M3000 development work continued. For a variety of reasons including that a lot of development work on the coarse grind had been done on the M3000 IsaMill™ and at that stage there wasn’t scope to go through that learning experience again on the M10,000 IsaMill™, it was decided that the M10,000 IsaMills™ would be commissioned into a reconfigured UFG circuit (November 2008). This provided extra UFG capacity for the increased plant throughput and allowed free use of two of the M3000 IsaMills™ for the primary grinding duty.

Only one of the M10,000 IsaMills™ can be currently operated due to power supply limitations at site. The M10,000 currently treats all of the rougher concentrate (precyclone underflow) prior to grinding in the M3000 UFG IsaMills™ to produce the 7-8 µm cleaner feed.

**CURRENT PRIMARY GRINDING CIRCUIT PERFORMANCE**

Currently, two 1.1 MW M3000 IsaMills™ are used in the primary grinding duties. Each of these mills process 40-60 tph at 80-95 m³/hr. The mills are operating with the reduced diameter discs to extend the service life intervals to 1,200 hours. The operation of both coarse grinding M3000 IsaMills™ allows plant throughput to be increased from 260 tph to 330 tph – a 27% increase in primary tonnage at reduced rougher flotation feed size distribution.

Summarised data from a recent survey across the entire primary grinding circuit is included in Table 2 where the circuit was configured as per Figure 2. The data shows the plant operating at 330 tph with both primary grinding IsaMills™ on line operating at 15 kWh/t producing a product P80 sizing of sub 40 µm from a feed F50 of around 250 µm.

The primary grinding circuit is currently constrained by the dewatering capacity of the original double deck SAG discharge screen. The large recirculating load in the circuit is diverted away from the SAG mill and consequently the SAG discharge screen via the tower mill into the cyclone feed sump. The power to tonnage ratio of the tower mill is very low and the size reduction is minimal – however it plays a
significant role in this circuit by effectively bypassing slurry around the SAG discharge screen. As a result, when the tower mill is offline for shoe replacements, there is a plant throughput decrease of 60 – 70 tph required to ensure that the screen deck does not become dewatering constrained. New trial pipe work is currently being installed to allow a bypass of the tower mill (during maintenance) – this will allow an evaluation of the real impact of the tower mill on the overall throughput and grind size.

<table>
<thead>
<tr>
<th>Table 2 - Primary Grinding Circuit Survey Summary</th>
</tr>
</thead>
<tbody>
<tr>
<td>P80</td>
</tr>
<tr>
<td>Fresh Feed to SAG Mill</td>
</tr>
<tr>
<td>SAG Cyclone Feed</td>
</tr>
<tr>
<td>SAG Cyclone OF</td>
</tr>
<tr>
<td>SAG Cyclone UF</td>
</tr>
<tr>
<td>Tower Mill Feed</td>
</tr>
<tr>
<td>Tower Mill Discharge</td>
</tr>
<tr>
<td>1st IsaMill™ Feed</td>
</tr>
<tr>
<td>1st IsaMill™ Discharge</td>
</tr>
<tr>
<td>2nd IsaMill™ Feed</td>
</tr>
<tr>
<td>2nd IsaMill™ Discharge</td>
</tr>
<tr>
<td>Total Rougher Feed</td>
</tr>
<tr>
<td>Total</td>
</tr>
</tbody>
</table>

*kWhr/t is based on the fresh feed tonnage actually processed through the individual unit.

FUTURE DEVELOPMENT OF THE PRIMARY GRINDING IsaMill™ AT MCArTHUR RIVER

Experience under the extreme grinding conditions of the McArthur River circuit has improved the understanding of wear in the IsaMill™ and the interactions of feed size distribution, feed density, media size, disc design, disc tip speed etc. These learnings have been adapted to other duties to improve wear life in all IsaMill™ applications – UFG, regrind and primary mainstream grinding. Work is currently underway on a number of initiatives to further improve the wear performance at McArthur River.

Laboratory testwork of a redesigned IsaMill™ disc has shown a 25% increase in power draw. Several of these discs will soon be trialled at McArthur River with the aim of restoring the mill to its full 1 MW power draw, thereby allowing further throughput increases using the M3000 IsaMills™.

Investigations into retrofitting the latest design M5000 (5,000 litre) IsaMill™ shell to the existing M3000 layout are underway. This would permit the mill to operate with the standard 1,200mm diameter discs at lower wear rates and regain the full power draw potential. The M5000 was developed as a result of lessons learned from the McArthur River experience. It is aimed at several market areas but its key advantage is in its geometry which allows it the ability to treat coarse feed distributions and draw up to 1.5 MW without the high wear issues experienced in the M3000, which was specifically designed for UFG applications.

FUTURE DEVELOPMENT OF THE PRIMARY GRINDING CIRCUIT AT MCArTHUR RIVER

Xstrata Zinc has recently announced a proposed Phase 3 Development for McArthur River which includes increasing the processing capacity to 5 mtpa. The primary grinding circuit options for this increase are currently under investigation - with 3.0 MW M10,000 IsaMills™ included in each of the cases under
consideration. The M10,000 IsaMills™ will be used in either a secondary or tertiary duty to take F 80 150 µm feed to the rougher flotation feed size of P 80 45 µm. These mills will be configured in a similar manner to the two existing M10,000 mills, providing operational options for either primary or regrind duty. The project is scheduled to commission in 2014.

CONCLUSIONS

Successful laboratory, pilot scale and full scale testwork has seen the incorporation of IsaMill™ technology into the primary grinding circuit at McArthur River. The IsaMill™ has now been part of the primary grinding circuit for more than four years. The use of two IsaMills™ in the primary circuit has enabled a 27% increase in primary grinding throughput.

Significant improvements were made to the IsaMill™ wear performance albeit at some sacrifice of power draw and mill throughput. Initiatives are underway to recover the lost power draw without losing all of the gains made in wear performance. Learnings from this application have improved understanding of wear mechanisms and have led to improved designs for all IsaMill™ applications.

The success of the work is reflected in the fact that additional 3.0MW M10,000 IsaMills™ are under consideration for use in the primary grinding circuit of the upcoming expansion to 5mtpa processing capacity.

ACKNOWLEDGEMENTS

The authors wish to thank all those at McArthur River and Xstrata Technology involved in the original testwork and plant scale implementation associated with this project and for permission to publish the data.

NOMENCLATURE

kWht/t = specific energy

REFERENCES


IsaMill™ Technology in the Primary Grinding Circuit

G Anderson (Xstrata Technology)
D Smith (McArthur River Mining)
S Strohmayer (McArthur River Mining)

IsaMill™ Development - UFG

- Developed to address ultra fine grinding requirements for McArthur River and Mt Isa Pb/Zn deposits
- McArthur River required 7µm grind
- Conventional technology was unsuitable due to poor efficiency and surface contamination
- Looked outside of the mining industry
- Collaboration with Netzsch of Germany – history of manufacturing stirred mills for paint, inks, pharmaceuticals
- 1.1MW M3000 (litre) IsaMill™ developed – became the enabling technology for McArthur River Mine 1995
IsaMill™ Development - UFG

- Requirement for larger scale IsaMill™ identified by Anglo
  - Inert and efficient grinding to improve mineral liberation and recovery
- M10,000 IsaMill™ development with Anglo/Netzsch 2003
- 1st M10,000 in UFG at Western Limb Tailings Retreatment
- Development of 3.5mm ceramic media with Magotteaux
- 18 x 3MW M10,000 IsaMills™ installed within Anglo
- F80 of 75-100µm; P80 of 53µm
- Further 4 x 3MW on order for Anglo
- Worldwide over 100 IsaMills™
McArthur River – Plant Trial

- Plant trial commenced in May 2007
Effect of Media Size

IsaMill™ Survey Data

Effect of Media Size

IsaMill™ – 2020 Compendium of Technical Papers

Contents
IsaMill™ Wear Performance

- Initial 400 hour service life
- Disc and shell replacement
- Still beneficial to operate - trial became permanent production mill

- UFG mills 2000 hour service intervals
- Higher wear a function of
  - coarser feed/media size
  - mineralogy
  - mill geometry designed for UFG

F80 350µm @ 12.8kWhr/t

![Cumulative % Passing vs Size (µm)](image)
IsaMill™ Wear Performance

- Smaller Diameter Discs: 1200mm to 1050mm
- Service life increased 400 to 1200 hours

IsaMill™ – 2020 Compendium of Technical Papers

McArthur M10,000 IsaMills™

- Successful M3000 trial – permanent production mill
- 2 x M10,000 installed for plant expansion Nov 2008
- M3000 development work continued
- M10,000 commissioned in UFG circuit – allowed 2 x M3000 for primary grinding circuit

IsaMill™ – 2020 Compendium of Technical Papers

Contents
Primary Circuit Performance

- Plant throughput increased by 25% from 260 to 325 tph

<table>
<thead>
<tr>
<th>P90</th>
<th>tph</th>
<th>kW</th>
<th>kWh/t</th>
</tr>
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<tbody>
<tr>
<td>Fresh Feed to SAG Mill</td>
<td>6377</td>
<td>323</td>
<td>3800</td>
</tr>
<tr>
<td>SAG Cyclone Feed</td>
<td>386</td>
<td>1187</td>
<td>950</td>
</tr>
<tr>
<td>SAG Cyclone OF</td>
<td>87</td>
<td>245</td>
<td></td>
</tr>
<tr>
<td>SAG Cyclone UF</td>
<td>382</td>
<td>1303</td>
<td></td>
</tr>
<tr>
<td>Tower Mill Feed</td>
<td>582</td>
<td>1187</td>
<td>950</td>
</tr>
<tr>
<td>Tower Mill Discharge</td>
<td>577</td>
<td>1187</td>
<td></td>
</tr>
<tr>
<td>1st IsaMill™ Feed</td>
<td>253</td>
<td>39.6</td>
<td>600</td>
</tr>
<tr>
<td>1st IsaMill™ Discharge</td>
<td>43</td>
<td>39.6</td>
<td></td>
</tr>
<tr>
<td>2nd IsaMill™ Feed</td>
<td>247</td>
<td>38.7</td>
<td>600</td>
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<tr>
<td>2nd IsaMill™ Discharge</td>
<td>36</td>
<td>38.7</td>
<td></td>
</tr>
<tr>
<td>Total Rougher Feed</td>
<td>76</td>
<td>83.4</td>
<td></td>
</tr>
</tbody>
</table>

- With no IsaMill™ case, P80 85µm, OWi = 19.3kWhr/t
- With 2 x IsaMill™ case, P80 76µm, OWi = 18.4kWhr/t

Development

- IsaMill™ Development
  - Alternative disc designs and configurations to recover power draw
  - Shell design – M5000 IsaMill™ - can be retrofit to M3000 IsaMill™
    - Allow geometrical ratios established with SDD in M3000 to be scaled up

- Circuit Development
  - McArthur River Phase 3 Development 2014
  - Increase capacity to 5 mtpa
  - M10,000 included in study for upgraded primary circuit
    - Secondary or tertiary duty ahead of flotation
Conclusions

- Successful lab, pilot and full scale testwork
- **IsaMill™** now integral, permanent part of primary grinding circuit - operating for over 4 years

- Improvements made to wear life and performance with further development work ongoing

- Success of the program
  - **IsaMill™** permanent part of grinding circuit
  - 2 x M10,000 installation
  - Further **IsaMills™** under consideration for the Phase 3 expansion of the primary grinding circuit
Optimising Western Australia Magnetite Circuit Design

D David¹, M Larson² and M Li³

1. FAusIMM, Process Consultant, AMEC Minproc, Level 14, 140 St Georges Terrace, Perth WA 6000. Email: dean.david@amec.com
2. Senior Process Engineer, Xstrata Technology, 5th Floor, 509 Richards Street, Vancouver BC, V6B 2Z6, Canada. Email: mlarson@xstratatech.com
3. Senior Metallurgy Manager, Grange Resources, Level 11, 200 St Georges Terrace, Perth WA 6000. Email: Michelle.Li@grangeresources.com.au

ABSTRACT

The development of Western Australian magnetite deposits has lead to the design of some of the largest grinding mills and plants in the world. One of the projects demonstrates the efficiency gains possible by developing a simple yet thorough test program for circuit design. By drawing on the experience of current magnetite operations in Australia and the Mesabi and Marquette iron ranges in the United States, a basic flowsheet was developed. Through comprehensive testwork with AG, ball and stirred milling the flowsheet was optimized to take full advantage of each grinding mill’s strengths to reach the required final grind size. Laboratory work was verified in the pilot plant to optimize the energy efficiency of each grinding step while ensuring adequate liberation at each step for sufficient gangue rejection. By using three stages of grinding, the ball mill can best be employed in ensuring all top size gangue material is liberated and removed in the second magnetic separation step. The inclusion of the IsaMill, with its inherent steep product size distribution, as the tertiary grinding stage ensured that maximum grade was achieved and simplified the downstream process while giving further improvements in total grinding capital and operating costs. In this way the combination of the two technologies downstream from the AG mill is far more efficient than either would be on its own by reducing the total installed power by 1/3 and annual grinding media cost by 2/3.

INTRODUCTION

In the past decade the interest in Australia’s iron ore deposits has shifted to include the vast magnetite deposits scattered throughout Western Australia. These deposits have resulted in numerous magnetite concentrators currently in the design and construction phase. Previous magnetite concentrators in Australia have been limited to smaller operations such as Savage River and OneSteel’s Project Magnet.

With a feed tonnage of 3600 tph and a final grind P₈₀ of 34 µm the amount of grinding required for this project will be extensive. A lower than typical final silica grade of sub 3% SiO₂ is desired. Finally a small amount of pyrrhotite is present in the ore requiring a final flotation step to meet sulfur requirements. In this case the complete flowsheet is examined testing previously proven technologies from existing plants with the aim of optimizing each step of the process. Laboratory and pilot work is combined to ensure maximum economical efficiency while still maintaining a quality product.

TYPICAL MAGNETITE CONCENTRATOR DESIGN

Five magnetite concentrator flowsheets are shown below from Minnesota (Figs. 1, 4 and 5), Michigan (Fig. 3) and Australia (Fig. 2). Though they have a mix of rod, ball and autogenous milling (depending on what was commonplace or available when the plants were designed) two steps in the process should be pointed out.

The use of a hydroseparater is commonplace in plants with and without reverse silica flotation. This removal of slimes is absolutely necessary prior to a silica flotation step and also aids in removing fine silica that would be more likely entrained in a magnetic concentrate.
Whether called a hydroseparator, hydro-sizer, siphon-sizer or thickener sizer the basic principal is the same. A “thickener” is employed but with an upflow of water added. The coarser, heavier magnetite settles to the bottom and the upflow of water carries the fine silica to the overflow. A magnetic flocculator can also be used on the feed to increase the settling rate and minimise chemical flocculant consumption.

The fine screens are necessary to recirculate back to grinding the +105 µm material that would negatively impact the final concentrate grade, and, in the case of Savage River, result in unacceptable concentrate pipeline wear. In the case of the east Mesabi ores such as Minntac, a concentrate ground to a P80 of 325 mesh (44 µm) will typically have a grade of 8% SiO2 with half of that silica in the 10% of the mass making up the +140 mesh (105 µm) size fraction. By using fine screens that size fraction can be removed and the final concentrate grade reduced to 4-5% silica.

Of all these operations, only the Empire Mine flowsheet does not use fine screens. The Empire Mine grind results in a final P80 of 20 µm, making finisher screens both impractical and unnecessary. The Empire Mine is also the only one of these plants shown to use reverse flotation as a final upgrade step.

Reverse silica flotation is usually seen as a last resort when processing magnetite as there is a certainty that magnetite will be lost to tails. The finisher magnetic separators will have rejected all liberated silica (except for silica slimes that follow the water in the process) ahead of flotation. Consequently, virtually all coarse floated silica will have magnetite attached to it. Amine flotation of silica also presents issues with slimes and is particularly sensitive to water chemistry. While hydroseparators should probably be installed regardless to remove the slimes ahead of finisher magnetic separation, the added hydroseparator benefit of reducing soluble salts is only of benefit when a silica flotation step is performed. According to Iwasaki (1983), the fatty acid flotation step for iron ores is quite sensitive to magnesium and calcium ions in pulp solutions and for optimal performance the level of these ions should be kept below 100 ppm.

Figs. 1 and 2: Erie magnetite and Savage River flowsheets (Devaney, 1985)
The testwork at AMMTEC for this Western Australian magnetite deposit consisted of:

- Pilot autogenous primary milling,
- laboratory work (Levin test) and pilot secondary ball milling,
- laboratory and limited continuous secondary IsaMill testing,
- laboratory tertiary, limited continuous and pilot IsaMill testwork,
- Davis tube and pilot magnetic separation tests of the different intermediate and final products,
- hydroseparating tests of the final IsaMill magnetic concentrate,
- sulfide flotation tests of the final magnetic concentrate, and
- final concentrate filter testing by vendors.

**Autogenous primary milling**

All of the ore was ground in the AMMTEC pilot AG mill (1.74 m diameter inside liners and 0.46 m EGL) and then run through the magnetic separator. To protect the magnetic separator the AG mill was closed with a 1 mm screen. In the full scale plant it is intended that the AG mill will be closed at a coarser size, up to a maximum of 4 mm. The full scale plant is required to process 3600 tph of feed, ideally through two grinding lines, each with a large dual pinion or gearless drive primary.
autogenous mill. With the energy available from the pilot AG mill operating at about a 25% filling the P<sub>80</sub> of the magnetic separator feed (-1mm) was approximately 330 µm. The magnetic concentrate was significantly coarser than the non-magnetics with P<sub>80</sub>'s of 420 µm and 200 µm respectively. This first stage of magnetic separation removed 40% of the total mass as barren tails. The remaining 2200 tph is fed to the second stage of grinding. Additional analysis suggests that the P<sub>80</sub> of the magnetics will increase from 420 µm to 770 µm when a 3 mm screen is used at full scale compared to the 1 mm pilot plant closing screen.

**Ball Mill Results**

Prior to any pilot ball mill work, one Levin test was completed on the secondary grind feed shown in Fig. 6. The Levin test is a fine-feed substitute for a standard Bond Grinding Work Index test and it uses sequential open circuit milling passes applying a known energy at each pass. The results from this test, with the P<sub>80</sub> size in microns plotted against the specific energy (kWh/t) is shown in Fig. 6.

![Fig. 6: Levin test for the ball mill](image)

The Levin test for fine ball mill product predicts a net specific energy requirement of 52.5 kWh/t to grind from an F<sub>80</sub> of ~400 µm (AG mill/magnetic separator product) to a P<sub>80</sub> of 34 µm. An additional 2.3 kWh/t would be needed to reduce the plant feed P<sub>80</sub> of 770 µm to the test feed P<sub>80</sub> of 420 µm, giving a net energy requirement of 54.8 kWh/t. However, as the Levin Test does not incorporate a classification step to remove fines, energy is unnecessarily used grinding material that has already achieved final target grind size or less. Consequently, it is expected that the test will over-predict the specific power requirement for most duties, especially over wide size ranges. The actual results obtained with the AMMTEC 6’ pilot ball mill with 25 mm top size media was a P<sub>80</sub> of 37 µm using between 40 and 45 kWh/t of energy. At 2200 tph of combined feed for lines 1 and 2 and incorporating the 2.3 kWh/t to reduce from 770 µm to 420 µm, using single pass ball milling from 770 µm to 34 µm would require about 114 MW of ball mill energy, excluding that associated with cyclone feed pumps. This is equivalent to 6 of the largest ball mills in existence, 3 per line. In addition approximately $86 million would be spent on steel grinding media annually. The high media consumption is due to an unusually high average Bond abrasion index of 0.44, caused by the presence of gangue silicates and garnets. Typical magnetite ore will have an Ai of 0.25 or less. The requirement to use 25 mm steel balls to reach the required fine grind exacerbates the media wear rate. A $15 million capital cost would also be incurred in any ball milling circuit targeting a 34 µm product for finishing screens which are needed to minimise silica and to protect the concentrate pipeline. The ball mill cyclone combination cannot guarantee elimination of +100 µm particles from the pipeline feed.
IsaMill Secondary Grind

The same feed as was tested in the previous Levin test was run through an M4 IsaMill to generate a signature plot. A graded 5 mm ceramic media charge was used to ensure top size breakage. The signature plot test consists of two tanks, one feeding the mill and one acting as a discharge tank. In the middle of each pass a sample is taken for a sizing and density. The flowrate, time and net energy of the pass is recorded and the size is plotted against the specific energy (kWh/t). Valves are then changed to switch the feed and discharge tanks and the process is repeated multiple times, each pass incrementing the input energy. From this a signature plot is developed. This creates a straight line on a log-log size vs energy plot that scales directly to full scale IsaMills. The result of this signature plot test is show in Fig. 7.

![Fig. 7: IsaMill 400 µm signature plot](image)

The IsaMill was able to reduce the secondary grind feed of 400 µm to 34 µm using 31.7 kWh/t. With the 2.3 kWh/t required for reduction from 770 µm to 420 µm (in the previous ball mill test) the total predicted IsaMill power is 34 kWh/t (note that this calculation is theoretical as 770 µm is an impractical IsaMill magnetite feed F80 and the inefficiency of the IsaMills on this size of material would mean that 2.3 kWh/t is optimistic). This is an improvement in specific energy of 30% over the ball mill (for this theoretical coarser feed). This would result in an installed power of 78 MW if IsaMills were used compared to the 114 MW necessary for the ball mill. In addition, the annual grinding media cost would drop from $86M for steel balls to $57M for ceramic IsaMill media. Besides the abrasiveness of the ore, the galvanic action of the pyrrhotite present in the ore may be contributing to the high steel media wear (Iwasaki, 1999). This would not be a factor with ceramic media.

Even with this drastic improvement, as with the ball mill there are still issues with using the IsaMill in a single stage for this duty. Due to the coarse, hard garnet and gangue silicates present, and the friction created in mixing these particles, media wear was higher than expected. Despite showing better wear than the ball mill steel media, the IsaMill ceramic media wear was still about double the expected rate. The energy used in grinding this garnet down to 34 µm can be considered wasted. The garnet in question is a hard alumina silicate and actually has properties close to that of typical IsaMill grinding media. It would be much more efficient to remove these waste materials as close to liberation as possible. Fig. 8 shows a close up view of the garnet crystals fully liberated in the 125-150 µm size range.
Fig. 8: Liberated garnet after 150 µm grind

By plotting the Levin test and secondary grind IsaMill test together (Fig. 9) it can be seen that the gain in efficiency experienced with the IsaMill for this particular ore starts at a product P<sub>80</sub> of approximately 100 µm. At product sizes above that point the ball mill is more efficient. This is mainly a function of media size. A media larger than 5 mm used in the IsaMill would achieve more efficient size reduction to 100 µm but would not be as efficient to the finer 34 µm final target. With this magnitude of size reduction applied to a fairly hard ore, grinding the entire stream in one step will never achieve optimum efficiency across all size fractions with any grinding technology.

Fig. 9: IsaMill/Levin Test comparison

The IsaMill improvement in efficiency at the fine sizes can be explained by the grinding mechanism and media used. At fine product sizes below 70-100 µm, attrition becomes the main grinding mechanism. The 5 mm media provides not only more surface area compared to the 25 mm ball mill media, and a greater probability for collisions than the larger steel media, but it is also stiffer than the
steel media. The more elastic steel media does not transfer energy to the ore particles as efficiently as the ceramic media. The Vickers Hardness of the ceramic used is about 1000, whereas that for steel media will typically range from 500-600.

The efficiency changeover is also seen in Fig. 10 (Burford and Niva, 2008), a comparison of the IsaMill with the Tower Mill grinding an Ernest Henry magnetite feed taken from the copper concentrator tails. In this case below 70 µm the smaller harder IsaMill media is more efficient than the 12 mm steel media used in the Tower Mill. This test was conducted with 3.5 mm ceramic media. By using 5 mm media this intersection point would be shifted coarser.

![Signature Plot - P80: IsaMill™ vs Tower Mill](image)

Fig. 10: Ernest Henry Magnetite IsaMill/Tower Mill comparison

It is clear from Figs. 9 and 10 that ball or Tower milling with typical media is most efficient when applied to generating “coarse” products in the +80 µm range. Consequently ball milling was selected to bridge the efficiency gap that exists between AG milling with this ore to about 400 µm P80 and the application of IsaMilling to feed sizes of 100 µm and finer. The Levin test suggest that about 15 kWh/t is required to grind from the 400 µm feed P80 to a 100 µm product P80. Pilot testing achieved a ball mill grind (with 32 mm media) from 420 µm to 78 µm at a specific energy of 11.6 kWh/t, again significantly more efficient than the Levin Test prediction. When translated to full scale operation grinding 2200 tph from 770 µm to 100 µm, the ball mill installed power requirement is 34 MW, achievable in two large twin pinion ball mills (one per line).

By only requiring a product of 80 to 100 µm from the ball mill it gives the added benefit of being able to switch from the 25 or 32 mm steel balls used in the tests to 40 mm balls. The 40 mm balls will ensure adequate top size reduction when treating the 770 µm F80 feed and will significantly reduce the steel media wear.

**IsaMill tertiary grind**

The product from the intermediate pilot ball mill was subject to a magnetic separation step which removed a further 20% of the mass, the majority of this being quartz and some of it being garnet. This resulted in a less abrasive 76 µm feed to the IsaMill stage and this material was tested in the M4 IsaMill using the standard signature plot procedure and 5mm ceramic media. The results are shown in Fig. 11 indicating that from a F80 of 76.5µm, the IsaMill energy required for a P80 of 34 µm is 12.4 kWh/t.

The coarse IsaMill test (Fig. 7) indicated that 13.4 kWh/t would be necessary to grind from F80 of 76.5µm to P80 of 34µm. This difference of 7.5% could be the result of a softer feed due to the extra magnetic separation step. It is also very close to the 5% margin of error associated with these tests.

This signature plot compares well with other iron ore IsaMill testwork completed on similar iron ore feed with F80’s and energy adjusted to 76.5 µm as shown in Fig. 12 (Larson, 2011).
The Levin test indicates that the ball mill would require 30 kWh/t for the same size reduction. However, the difference between the two pilot scale ball mill trials (420 µm to 34 µm vs 420 µm to 100 µm) was only 22 kWh/t, a much better estimate of the ball mill power necessary to achieve the final grind. Again the Levin test has overpredicted the ball mill power requirement. The best available comparison between the two tested fine milling contenders is, therefore, 22 kWh/t for ball milling vs 12 kWh/t for IsaMilling, an advantage of 45% in specific energy alone to the IsaMill for this step. Additional installed power and capital cost savings are realized because cyclones are not required in the IsaMill circuit while they are necessary for ball milling.

Fig. 11: Fine magnetite feed IsaMill signature plot

Fig. 12: Common iron ore signature plots

This finer F_{80} 76.5 um feed was also run through the M4 IsaMill in a short 180 kg continuous test and then continuously as part of the pilot plant. An energy of about 12-13 kWh/t was targeted by adjusting the speed of the mill. The 34 µm P_{80} target size was maintained through these tests including the top size reduction.

There was also a benefit seen in ceramic media wear in grinding this finer, upgraded material. Media wear was reduced by 35% per kWh as compared to the secondary IsaMill feed (420 µm F_{80}).

Given the silica contamination and iron grade implications top size material can have, it is important to consider the complete product size distribution produced in regrinding. The size distribution from the signature plot pass that was closest to the 34 µm target is shown in Fig. 13.
The IsaMill product size distribution is 98.9% passing 75 µm and 100% passing 105 µm at a \( P_{80} \) of 37 µm. This removes the need for finisher screens which saves up to $15 million in capital costs together with operating costs associated with a potentially maintenance intensive piece of equipment. Typically 15 inch cyclones will produce a \( P_{98}/P_{80} \) ratio of about 3-4 and this would definitely necessitate finishing screens. In this case the IsaMill ratio is under 2.

Besides a natural improvement in concentrate grade without the need for fine screens, the steeper size distribution should have an additional benefit of producing less ultrafines. The intermediate magnetic separation step after the ball mill removes about 80% of the silica present in that stream. This results in less silica that has to be ground to 34 µm. This reduction in material that could be turned into silica slimes, along with the previously mentioned sharper size distribution, may result in a smaller hydroseparator design, reducing the footprint of one of the larger pieces of equipment in the flowsheet. Benefits should also be seen in filtering the concentrate. That testwork is still ongoing as of the writing of this paper.

**Final Results**

A combination of bench and pilot testing was used to determine the most efficient means of reducing primary crushed magnetite ore to a target grind \( P_{80} \) in the region of 34 µm. Efficiency was maximised by a combination of using appropriate machines for the various grinding duties and also by rejecting barren mass as early as possible in the flowsheet.

The AG milling stage was found to generate a relatively coarse product at pilot scale and this stream will only get coarser at full scale. The coarseness of the primary magnetic concentrate makes it unsuited for feeding to a fine milling device such as an IsaMill but it is ideal ball mill feed. However, the ball mill was found to be unsuitable for taking the ore from AG discharge to the final target grind in a single step as it becomes relatively inefficient for grinding finer than 80 µm. Testwork showed it was appropriate to use the ball mill to grind to about 100 µm \( P_{80} \), follow this with magnetic separation and complete the grinding to 34 µm in an IsaMill. Approximately 60 MW of power is saved (40% of total power and ~50% of the power for that grinding step) over the single stage ball mill circuit and $62M annually in grinding media costs as shown in Table 1. There will be the need for an additional magnetic separator step in the flowsheet but the finisher screens can be eliminated and the hydroseparator requirement will be smaller. The IsaMill has an internal centrifugal classifier so there is no need for classifying cyclones in tertiary grinding. Wherever possible, gravity will be used to avoid the installation of extra pumps. For example, the ball mill cyclone overflow will gravity flow to the intermediate magnetic separator distributor and the magnetic separators themselves will be positioned so that the magnetics gravitate directly to the IsaMill feed tanks.
Table 1. Installed power and annual media cost comparison

<table>
<thead>
<tr>
<th>Section Feed Rate (t/h)</th>
<th>Specific Energy (kWh/t)</th>
<th>Installed Power (MW)</th>
<th>Annual Media Cost Estimate</th>
</tr>
</thead>
<tbody>
<tr>
<td>Autogenous Mill 770 µm Product</td>
<td>3800</td>
<td>8.5</td>
<td>40</td>
</tr>
<tr>
<td>Single Stage Ball Mill 34 µm Product</td>
<td>2200</td>
<td>47</td>
<td>114</td>
</tr>
<tr>
<td>Single Stage IsaMill 34 µm Product</td>
<td>2200</td>
<td>34</td>
<td>78</td>
</tr>
<tr>
<td>Ball Mill 100 µm Product IsaMill 34 µm Product</td>
<td>2200</td>
<td>12</td>
<td>34</td>
</tr>
<tr>
<td></td>
<td>1720</td>
<td>13</td>
<td>24</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>58</td>
</tr>
</tbody>
</table>

While 58 MW of ball mill and IsaMill power will be installed, only about 50 MW of operating power will be necessary under average operating conditions. The installation of eight 3.0 MW M10 000 IsaMills (four per line) also leaves some room for expansion in the future. Although each mill is powered by a 3 MW motor only 2.7-2.8 MW will be necessary per mill on average. Also, in this case, when one IsaMill is shut down for maintenance, the other mills will be able to be ramped up to full power without affecting production.

The addition of the hydroseparator to the circuit is critical to achieve a silica content of less than 2%. The magnetic finishers are incapable of fully removing the slimes present and entrained at the final grind size. The hydroseparator has been shown to decrease the tested finisher magnetic separator concentrate silica content by about 1%.

After the hydroseparator and finisher magnetic separation there is a sulfide flotation step. While pyrite is removed to tails in the magnetic separation steps the pyrrhotite present in the ore is magnetic and would otherwise report to the final concentrate. The sulfur content of the final magnetic concentrate is 0.6% and this is unacceptable for pelletisation plants. Through the sulfide flotation step the sulfur content of the final product is reduced to less than 0.1% which is acceptable.

The final concentrator flowsheet developed for this project thus becomes:

Two 20 MW AG mills closed off with 3mm screens followed by magnetic separation. The concentrate of this feeds two 17 MW ball mills closed off by hydrocyclones, from which the overflow feeds a second set of magnetic separators. This magnetic concentrate feeds eight 3 MW M10 000 IsaMills. The IsaMill product is at the final grind size without further classification and feeds in succession to the hydroseparators, followed by the final magnetic separation step and finally the sulfide float. The underflow of the sulfide float step is pumped to a filter plant on the coast. Each separation step of magnetic separation, the hydroseparators and flotation will produce a stream of final tailings, with no recirculating loads planned.

CONCLUSIONS

This work can be considered a success in that significant improvements have been made over the preliminary design in both capital and operating costs. Further, the stringent requirements for silica and sulphur levels in the final concentrate have been achieved even with these cost savings. This is made all the more impressive by the short time frame in which the work and design were completed. The pilot plant campaign on the AG mill discharge lasted just over one week. This included running two different ball mills and adding the M4 IsaMill to run continuously in the pilot plant. The savings of +56 MW of grinding power and over $60M annually in grinding media speak for the effectiveness of having a simple, yet well thought out plan of testwork and adapting that plan as results became available.
ACKNOWLEDGEMENTS

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REFERENCES


IsaMill™ Design Improvements and Operational Performance at Anglo Platinum

C Rule¹ and H de Waal²

1. Head of Concentrators Technology, Anglo American Platinum, 55 Marshall Street, Marshall Town, Johannesburg, South Africa. Email: Chrisrule@Angloplat.com
2. Consulting Metallurgist, Xstrata Technology, Level 4, 307 Queens Street, Brisbane Qld 4000. Email: hdewaal@xstratatech.com

ABSTRACT

In 2003, Anglo Platinum, in a joint development with Xstrata Technology, installed the world’s first 10 000 litre IsaMill™ in a concentrate regrind duty at the Western Limb Tailings Re-treatment Plant. The success of that installation was the enabling event for Anglo Platinum to proceed with a substantial investment in horizontal stirred milling technology. Since 2006 an additional three IsaMills™ in concentrate regrind duties and a further 18 IsaMills™ in the more technically challenging coarse grinding mainstream applications, have been commissioned in group Concentrator operations - bringing the total number of IsaMills™ installed in Anglo Platinum plants to 22.

A collaborative approach between Anglo Platinum and Xstrata Technology towards improving milling efficiency and reducing operating costs, through internal mill component wear optimization and operating recipe development, has resulted in further improvements in the overall success of IsaMills™ in the flow sheets of many Anglo Platinum operations. The addition of IsaMills™ in the Anglo Platinum flow sheets has improved plant PGM recoveries by as much as 5%.

This paper explores the improvements made to the IsaMills™ flow sheet and mill internal design and shares some of the operating experience with IsaMills™ technology in Anglo Platinum

INTRODUCTION

IsaMill™ technology has been applied to a number of different ore types and process applications in an ongoing drive to liberate valuable minerals from complex ore bodies.

The need for ultrafine grinding arose at McArthur River mine in the Northern Territory of Australia where finely grained lead/zinc concentrates required a grind size P80 of 7 microns to produce a saleable product. The high energy intensity and inert grinding benefits of this unique stirred milling technology further contributed to producing a suitable final concentrate (Barns et al., 2006).

Anglo Platinum pursued stirred milling technology as an economically viable option to improve liberation following successful scoping work on bench scale initially using a 4 liter bench scale IsaMill™. Thereafter a series of off-site and onsite pilot scale tests were conducted which ultimately led to the installation of a large
scale 10 000 liter mill at their Western Limb tailings retreatment project. This mill has subsequently been deployed in a number of grinding duties with varying coarseness of feed, allowing investigation on the wear rates of mill internals (Rule C M, 2010).

Further requirements for liberation of finely grained platinum group metals (PGM’s) in the main stream process flow resulted in IsaMills™ allocated to a coarser grind duty as tertiary grinding applications, producing a pre-scavenging flotation product at an approximate P80 of 53 microns (µm).

The coarse grinding applications typically receive feed at F80 of up to 100 µm although coarser feed sizes of ten – 15 % >150 µm has been observed in certain cases (Rule et al.,2010)

The coarser feed sizes have been shown to impact negatively on wear performance of grinding discs and wear of shell liners to a lesser extent.

To date Anglo Platinum has installed 17 IsaMills™ in main stream inert grinding (MIG) applications and 4 units in an ultrafine concentrate regrind duty (UFG).

Since the installation of the first MIG mill a number of improvements have been made to subsequent flow sheet designs and mill internal components to optimize the component life and value gained from this technology.

This paper explores some of these changes and highlights the benefits realized and value gained from the collaborative efforts of Anglo Platinum and Xstrata Technology towards unlocking the true potential of IsaMill™ technology in the PGM industry.

**DESIGN IMPROVEMENTS**

The majority of IsaMills™ in Anglo Platinum have been installed in MIG grinding applications where the mills receive feed from a secondary or tertiary ball mill. IsaMill™ product will then report to a scavenging flotation stage to capitalize on the improved liberation state and flotation kinetics of the ground pulp.

Due to the inherent variability of the upstream operating plant; i.e. highly variable mainstream slurry flow rate and slurry particle size distribution, the need arose to modify the process flow sheet and mill internal component design parameters to better adapt to this variability. Media handling systems were also improved to facilitate faster reloading of media after maintenance inspections with a reduction in mechanical wear on this equipment and an improvement in spillage generation and media accounting.
Constant Flow Concept

The internal classification functionality of the IsaMill™ utilizes a product separator at the discharge end of the mill to centrifuge coarse material to the outer circumference of the mill and transport this material back towards the feed end of the mill.

![Fig. 1. IsaMill™ internal classification](image)

Fig. 1. depicts the internal configuration of the IsaMill™ detailing the classification function performed by the product separator (rotor).

Due to the pumping effect of the product separator towards the feed end of the mill a certain degree of compression is generated towards the feed end of the IsaMill™. Slurry feed into the mill counteracts this back pressure to provide a net positive flow through the mill. Conventional PI level control in the IsaMill™ feed tank cascades the change in feed tank level onto a flow rate set point for the feed pump. A reduction in volumetric flow rate into the IsaMill™ circuit due to upstream process changes will result in a reduction in flow rate into the IsaMill™ and subsequent increase in compression towards the feed end of the IsaMill™. This concentrates the grinding action towards the front end of the mill with the majority of the grinding chambers not contacting with a slurry / media mixture, and promoted uneven and accelerated wear rates of mill grinding discs particularly towards the feed end of the IsaMill™.

A constant flow concept was subsequently devised to fix the flow rate set point of the IsaMill™ feed pump and return a portion of the milled product to the feed tank, thus maintaining feed tank level set point while achieving a constant flow. The original valves in the discharge line was used during the starting and shut down procedures, these were replaced by ceramic lined variable split range control valves during the modification. The recycled portion should however be maintained to a minimum and advanced process controllers were utilized to maintain the recycle portion to a minimum as shown in Fig. 2.
Fig. 2. – IsaMill™ recycle flow control

Fig. 3. – Waterval UG 2 A-Section IsaMill™ Flow Control Impact

Waterval UG 2 Recycle Flow Control Impact - IsaMill™ Feed Flow Rate

- Pre re-cycle control
- Post re-cycle control

Fig. 3. – Waterval UG 2 A-Section IsaMill™ Flow Control Impact
The result of the new control system can be seen in Fig. 3. where the stability of feed flow increased drastically after the modifications to the recycle flow control were made to this mill. The operating recipe was also altered to achieve the best media distribution and grinding efficiency, thus the reduction in slurry feed flow rate set point to 180 m$^3$/h. The average flow rate achieved before re-cycle flow control was between 200 and 250 m$^3$/h, the variability in flow rate during normal plant operation can result in drastic changes in feed flow rate to as low as 50 m$^3$/h as displayed in Fig. 4. The modification to recycle flow control has proven to be significant during periods of low feed flow to the IsaMill™ circuit.

All MIG installations in Anglo Platinum have subsequently been fitted with this modification with similar improvements in flow rate stability realized.

**Reduced Diameter Discs (RDD’s)**

**Case Study One – Mogalakwena South Concentrator disc trial**

The major wear component inside the MIG IsaMills™ in Anglo Platinum is the grinding disc which enables the agitation of media for attrition grinding between agitated media and solid particles in the slurry stream.
The original grinding discs at 1720 mm in diameter showed some degree of wear particularly towards the feed end of the mill. The high tip speeds and distance between the disc outer circumference and mill shell resulted in accelerated wear of the discs and shell liner.

The MIG installations in Anglo Platinum have displayed the most aggressive wear rates of all main stream installations worldwide. Installations in a main stream copper processing plant in Australia have shown much lower wear rates at similar specific energy consumptions as the Anglo Platinum installations (refer to Fig. 8.) This could be attributed to the specific mineralogy of the reefs in the Bushveld complex mined for its PGM content. Many Anglo Platinum operations treat Upper Group 2 (UG2) reef known for its chromite spinel content, which comprise about 70 % of the Chromitite reef. The angular crystal structure and high specific gravity of chromite can contribute to higher wear rates on grinding discs if allowed to accumulate inside the IsaMill™. Chromite content in the mill can be controlled by ensuring that coarse chromite is treated separately from the IsaMill™ circuit or that the size fraction of chromite reporting to the IsaMill™ circuit has been sufficiently reduced to allow treatment through the IsaMill™. Typical F95 sizes suitable for treatment through the mill are approximately 110 - 115µm. This grind can normally be achieved with conventional ball mills as a pre-IsaMilling stage.

Merensky and Plat Reefs also form part of the PGM rich deposits in the Bushveld Complex. Less aggressive component wear rates have been reported for the mills deployed in a MIG duty on these reefs although the pyroxenite gangue rock associated with the Merensky reef and the weathering alteration effect on Plat Reef are known to be some of the hardest ore types to beneficiate.

The main wear dimension on the grinding discs was in thickness wear with much lower wear rates recorded in disc diameter. Disc replacement requirements are mainly as a result of width or thickness wear where the steel reinforcing inside the rubber moulded discs are exposed after the majority of rubber has worn away, thus necessitating the replacement of a disc.

The first IsaMills™ where reduced diameter discs were installed was at Mogalakwena South’s C Section IsaMill™. This mill was installed in a tertiary grinding application receiving feed from the secondary ball mill product stream. Slurry entered the IsaMill™ circuit at a P80 of 75 µm producing a product at P80 53 µm.
Fig. 5. show the reduction in disc thickness wear after changing to reduced diameter discs from normal discs. An immediate reduction in wear rates can be observed in grinding discs 1 and 2.
Of additional interest is the effect of slurry feed particle size distribution on disc thickness wear. The IsaMill™ was employed in secondary regrind duty for a period when the secondary ball mill was off line for major repairs. During this period the feed particle size distribution to the IsaMill™ circuit increased due to the absence of the secondary ball mill in the circuit, with a consequential increase in disc wear rates.

Fig. 6. displays the reduction in final grind during the period while the secondary ball mill was off line. No process data was available for IsaMill™ feed particle size distribution but the final tailings grind clearly shows a reduction in grind during the period in question. The operating recipe of the IsaMill™ remained the same with no changes to grinding media type or size and the same specific energy (kWh/ton) throughout.

Once the secondary ball mill was returned to duty the disc width wear rates reduced although not to previous rates observed when originally switching to reduced diameter discs. This could be attributed to a change in the pebble crushing circuit preceding the grinding circuit in C Section, where the cut point of final product from the crushing circuit was increased to capitalise on the expected additional grinding capacity with the re-introduction of the secondary ball mill into the circuit.
Case Study 2 – Western Limb Tailings Re-treatment Concentrator reduced diameter disc performance

The IsaMill™ at the Western Limb Tailings Re-treatment (WLTR) Concentrator was originally installed as a concentrate regrind mill. The WLTR plant processes PGM tailings material from historical dumps in the greater Anglo Platinum Rustenburg concentrator operations area. This mill was later converted from using sand as grinding media to ceramic beads to allow treatment of coarser slurries in main stream grinding applications.

Initial disc width wear while in concentrate regrind duty was between 0.19 – 0.02 µm/kWday. See Fig. 7.

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**Fig. 7. – WLTR Concentrator comparative disc wear rates**

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**WLTR Plant UFG v MIG Wear Rates**

<table>
<thead>
<tr>
<th>Disc no.</th>
<th>UFG</th>
<th>Primary Duty</th>
<th>Primary Duty RDD</th>
<th>Secondary re-grind RDD</th>
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</thead>
<tbody>
<tr>
<td>1</td>
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<td>2.44</td>
<td>1.96</td>
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<td>2</td>
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<tr>
<td>4</td>
<td>0.11</td>
<td>0.02</td>
<td>0.00</td>
<td>0.00</td>
</tr>
</tbody>
</table>

**Metallurgical Plant Design and Operating Strategies (MetPlant 2011)**

8 - 9 August 2011 Perth, WA
Disc wear rates increased rapidly to 2.44 µm/kWday when the mill was deployed as a primary grinding unit treating ore in a parallel side stream to the primary ball mill. Reduced diameter discs were then installed on discs one and two with normal diameter discs remaining in the rest of the mill. Disc width wear rates reduced on the first two positions with an expected increase in wear rate on disc three due to a shift in the media distribution profile towards the discharge end of the mill. In essence the discs after the first two discs started working a bit harder and this was illustrated in the increase in wear rates on disc three.

**Comparative wear data**

![Figure 8. Comparison between PGM and Copper reef disc wear rates](image)

Disc width wear rates in the majority of MIG installations in Anglo Platinum have shown significant differences when compared with “softer” ore types. A typical MIG installation on a copper ore as trended in Fig. 8. shows the difference in wear rates between this IsaMill™ and Mogalakwena C Section IsaMill™. Both these mills operate on very similar specific energy consumptions with similar operating recipes.
IsaCharger™ hydraulic media transfer

Fig. 9. – IsaCharger™ unit

The IsaMill™ require that the complete grinding media charge be removed from the mill prior to an internal inspection. A media hopper is situated underneath the mill shell to contain media scuttled from the mill. The media hopper is also used to store new media required to replenish the mill charge as the media is consumed as part of the grinding process. A screw feeder was originally used to transfer ceramic media from the media hopper into the IsaMill™ feed tank, from where it is pumped into the IsaMill™ as a slurry/media mixture.

The screw feeder consisted of numerous moving mechanical parts like gland seals, greased bearings etc. which required routine maintenance interventions to ensure high equipment availability was maintained. Xstrata Technology subsequently developed a hydraulic transfer device called “IsaCharger” to transfer media from the media hopper to the feed tank. The IsaCharger™ consists of no moving mechanical parts and utilizes a custom built high pressure venturi type device to transfer media by means of a high powered jet of water from underneath the media hopper into the mill feed tank. See Fig. 9.

Most IsaMill™ installations in Anglo platinum have been fitted with IsaChargers™ with reported improvements in equipment on-line and similar media recharge times as achieved by the original screw feeders.
To date no maintenance interventions have been required on any of these units installed on the Anglo Platinum IsaMills™. Some of the units have been in operation for up to 6 months.

**OPERATIONAL PERFORMANCE**

***Case Study – Amandelbult Concentrator UG 2 # 2 Plant***

![Image of Amandelbult UG 2 IsaMills™ showing MIG and UFG concentrate regrind mills.](image)

**Fig. 10. – Amandelbult UG 2 IsaMills™ showing MIG and UFG concentrate regrind mills.**

<table>
<thead>
<tr>
<th>Plant</th>
<th>Number and duty</th>
<th>Total Installed Power (kW)</th>
<th>Commissioning Date</th>
</tr>
</thead>
<tbody>
<tr>
<td>Merensky</td>
<td>2 Parallel Tertiary MIG M 10 000</td>
<td>6 000</td>
<td>April 2009</td>
</tr>
<tr>
<td>UG 2 no. 1</td>
<td>1 Tertiary MIG M10 000</td>
<td>3 000</td>
<td>April 2009</td>
</tr>
<tr>
<td>UG 2 no. 2</td>
<td>1 Tertiary MIG M 10 000</td>
<td>3 000</td>
<td>March 2009</td>
</tr>
<tr>
<td>UG 2 no. 2</td>
<td>1 UFG, M 3 000</td>
<td>1 500</td>
<td>March 2009</td>
</tr>
</tbody>
</table>

Table 1 – IsaMill™ installations at Amandelbult

Table 1 list the IsaMills™ installed at Amandelbult.

The Merensky IsaMill™ circuit comprises two 10 000 liter mills treating Merensky reef exclusively while the two UG 2 plants combined contribute approximately 57 % of the total production throughput at Amandelbult.

The importance of UG 2 reef as a contributor towards the total PGM ounces produced at Amandelbult is apparent and optimizing the PGM recovery for this reef type is vital.
The UG 2 # 2 plant IsaMill™ ties into the plant flow sheet as a tertiary regrind mill on the silicate stream. Chromite is removed from the process stream by means of cyclones after the primary roughing flotation stage. The chromite stream reports to a dedicated tumbling mill regrind circuit while the silicate stream is reground in a secondary ball mill prior to reporting to the IsaMill™ circuit. Fig. 11 illustrate the basic flow sheet of this circuit.

This case study on the Amandelbult UG 2 # 2 plant explores the performance of the IsaMill™ and quantifies the benefit realized after commissioning of the IsaMills™

**UG 2 Plant Mineralogy**

Table 2 details the mineral association in composite samples from the plant pre-IsaMill installation.

A significant portion of PGM’s are enclosed in gangue as a result of incomplete liberation, in particular the > 53 µm fractions in the final tailings sample comprise 57 % of mineral deportment in tailings in that size fraction.

The liberation issues on the final tailings streams could be addressed through main stream inert grinding with ultrafine concentrate regrinding of attached mineral particles in the sub 25 µm size ranges.
Table 2 – Mineral Association table for typical Amandelbult UG 2 process samples

<table>
<thead>
<tr>
<th>Association</th>
<th>Feed</th>
<th>Concentrate</th>
<th>Tailings &lt;10 µm</th>
<th>Tailings &gt;10 µm</th>
<th>Tailings &gt;53 µm</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liberated</td>
<td>49.2</td>
<td>53.1</td>
<td>31.3</td>
<td>82.3</td>
<td>18.5</td>
</tr>
<tr>
<td>Enclosed in BMS*</td>
<td>23.6</td>
<td>15.8</td>
<td>4.7</td>
<td>4.1</td>
<td>9.7</td>
</tr>
<tr>
<td>Attached to BMS</td>
<td>7.9</td>
<td>12.7</td>
<td>0.3</td>
<td>0.4</td>
<td>0.6</td>
</tr>
<tr>
<td>PGM/BMS/Silicate</td>
<td>5.6</td>
<td>6</td>
<td>7.7</td>
<td>-</td>
<td>5.6</td>
</tr>
<tr>
<td>Enclosed in Silicate</td>
<td>7.5</td>
<td>8.4</td>
<td>36.0</td>
<td>2.7</td>
<td>44.3</td>
</tr>
<tr>
<td>Attached to Silicate</td>
<td>0.6</td>
<td>2.7</td>
<td>9.3</td>
<td>3.5</td>
<td>13.9</td>
</tr>
<tr>
<td>Enclosed in Oxide</td>
<td>4.8</td>
<td>1.3</td>
<td>7.6</td>
<td>4.3</td>
<td>6.0</td>
</tr>
<tr>
<td>Attached to Oxide</td>
<td>0.8</td>
<td>-</td>
<td>3.1</td>
<td>2.7</td>
<td>1.4</td>
</tr>
<tr>
<td>TOTAL</td>
<td>100.0</td>
<td>100.0</td>
<td>100.0</td>
<td>100.0</td>
<td>100.0</td>
</tr>
<tr>
<td>Midlings</td>
<td>7.2</td>
<td>14.6</td>
<td>7.0</td>
<td>8.3</td>
<td>14.8</td>
</tr>
<tr>
<td>Locked</td>
<td>43.6</td>
<td>32.1</td>
<td>61.7</td>
<td>9.4</td>
<td>66.7</td>
</tr>
</tbody>
</table>

*BMS – Base metal sulphides

Grind Performance

The grind performance achieved by the MIG IsaMill™ on the UG 2 # 2 plant showed significant reduction in the coarser size fractions (>150 µm) with a marked improvement in IsaMill™ product towards the end of 2010. This improvement was mainly due to an increase in mill power draw from 1500 kW to 2000 kW as illustrated in Fig. 12.

Fig. 13 illustrate the increase in finer size fractions passing 53 µm primarily as a result of the increased power draw. This result is vitally important for the UG 2 operations as a significant proportion of PGM’s was locked in the > 53 µm fraction of the final tailings plant composite samples. Ref. Table 2.
Fig. 12. – Amandelbult UG 2 coarse fraction size reduction

Fig. 13. – Amandelbult UG 2 fine fraction size reduction
Recovery benefit

Fig. 14. – Amandelbult UG 2 fine fraction size reduction

Fig. 14. shows the historical tailings grades achieved in the two UG 2 plants as well as the contribution of these two process streams towards the final tailings grade of the concentrator complex. The IsaMills™ were commissioned in March and April 2009 and after a period of grinding and flotation circuit optimization, the tailings grades shows a marked improvement with tailings grades averaging between 0.5 and 0.6 g/t PGM and gold.

The reduction in final tailings grades on the other Anglo Platinum operations where IsaMills™ have been installed has shown similar trends as in the Amandelbult UG 2 # 2 plant.

CONCLUSIONS

IsaMills™ horizontal stirred mills have been successfully introduced in the majority of Anglo Platinum’s operations in South Africa. Further advances made in wear component design and circuit layout have further improved the results obtained from these mills from a component wear and grinding efficiency perspectives.

As an illustration to the importance of PGM recovery in the Concentrator Operations in Anglo Platinum, a one % increase in PGM recovery in the concentrators equates to approximately 75 million US$ at recent market prices and exchange rates. The following extract from the Anglo Platinum Annual report for 2009 puts further emphasises on the magnitude of the recovery improvements directly contributed to IsaMills™
“Early indications during the latter part of 2009 were promising; for example, platinum metal recoveries for the last quarter of 2009 at Rustenburg increased substantially post IsaMill™ commissioning by in excess of 3 percentage points.”

ACKNOWLEDGEMENTS

The authors would like to acknowledge Anglo American Platinum for permission to publish data from their concentrator operations in this paper.

REFERENCES


Iron Ore 2011

Unlocking the value in waste and reducing tailings:
Magnetite Production at Ernest Henry Mining

J Siliézar, D Stoll, J Twomey

Contact Author:
Full name: Jose Gerardo Siliezar
Position title: Magnetite Interface Metallurgist
Organisation Name: Ernest Henry Mining
Address: PO Box 527 Cloncurry Qld 4824
Phone: 07 4769 4500
Fax: 07 4769 4555
Email: jsiliezar@xstratacopper.com.au
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Unlocking the value in waste and reducing tailings:
Magnetite Production at Ernest Henry Mining

J. Siliezar (1), D Stoll (2), J. Twomey (3)

1.
Jose Siliezar (MAusIMM)
Position title: Magnetite Interface Metallurgist
Organisation Name: Ernest Henry Mining
Address: PO Box 527 Cloncurry Qld 4824
Email: jsiliezar@xstratacopper.com.au

2.
Dominic Stoll (MAusIMM)
Position title: Magnetite Project Metallurgist
Organisation Name: Ernest Henry Mining
Address: PO Box 527 Cloncurry Qld 4824
Email: dstoll@xstratacopper.com.au

3.
John Twomey (MAusIMM)
Position title: Manager Magnetite Project
Organisation Name: Ernest Henry Mining
Address: PO Box 527 Cloncurry Qld 4824
Email: jtwomey@xstratacopper.com.au
ABSTRACT

Ernest Henry Mining (EHM) is situated 38 km north-east of Cloncurry in the Mount Isa – Cloncurry mineral district of North-West Queensland. The EHM ore body is an iron oxide copper gold deposit with an average grade of 1 % copper, 0.5 g/tonne gold and 23% magnetite, with current reserves of approximately 88 million tonnes.

The copper concentrator is a single line plant with a nominal throughput rate of 1,300 tonnes per hour. An average of 350,000 tonnes of concentrate is produced each year containing around 100,000 tonnes of copper metal and 120,000 troy ounces of gold.

In December 2009 Xstrata Copper announced approval of a $589 million investment to extend the life of EHM to at least 2024, through the transformation of open pit mining operations to a major underground mine together with the construction of an associated magnetite extraction plant.

The EHM magnetite plant extracts the magnetite from the copper concentrator's tailings stream. It will produce approximately 1.2 million tonnes of magnetite concentrate per annum at full capacity for export to Asia, making EHM Queensland’s first iron ore concentrate exporter. Construction of the magnetite extraction plant commenced in July 2010 and was completed in January 2011.

This paper gives an overview of the development of the EHM magnetite extraction plant flow sheet from the early mineralogy and laboratory test work through to the plant commissioning and early operation of the base plant. It has a major focus on the commissioning learnings from the plant start-up and includes a review of the operational performance of the plant versus original design and expectations.

INTRODUCTION

In December 2009 the Xstrata board announced the approval of $589 million to transform EHM into a major underground mine with an associated magnetite extraction plant which will extend the mine life to 2024. This will create 400 permanent jobs from 2013 in addition to the jobs created during the construction phase. The transformation from an open pit to a large scale underground operation will enable the extraction of 6Mt pa of ore at full capacity to produce 1.2 Mtpa of tonnes of premium quality magnetite concentrate. The first exports of magnetite concentrate are planned for shipment from Townsville in the first half of this year.

The construction of the $79 million magnetite base plant was completed in January this year followed by a period of process commissioning. As part of the base plant Xstrata is developing an $8.6 million expansion of its 65 kt enclosed storage shed at the port of Townsville. While magnetite has traditionally been discarded as tailings at EHM, the magnetite extraction plant allows it to be captured as an important by-product of the copper-gold concentrating process, reducing the amount of tailings sent to the on-site storage facility. The production of magnetite concentrate enables EHM to maximise the value of its existing resources by creating an additional by-product, reduce the amount of tailings in the on-site storage facility and provide further employment opportunities in the region. The magnetite concentrate produced at EHM will be chiefly used to fuel Asia's steel industry, however, the commodity may also be utilised as a washing agent in coal operations.

The plant has 3 key circuits including an extraction plant which separates and upgrades the concentrator tailings to a marketable concentrate, a dewatering plant to remove the water from the magnetite concentrate for transport and a regrind circuit to maximise the recovery from concentrator tailings. The overview of the magnetite extraction process is shown in Figure 1.
MINERALISATION AND ORE RESERVES

The Ernest Henry ore body is an iron oxide copper gold deposit (IOCG) discovered in October 1991 by airborne geophysics. Similar examples are Olympic Dam and Prominent Hill in South Australia, and Osborne in Queensland. It is located in the Eastern Fold Belt of the Mount Isa Inlier under approximately 50m of sand and clay cover. The mineralisation at Ernest Henry has formed in a south-east plunging body of altered and variably brecciated felsic volcanic rock. The combined thickness of the mineralised sequence is approximately 250 m, width averages 300 m and the down dip length is approximately 1,000 m in a pipe like formation which dips 40 degrees to the SSE and is still open at depth (Ryan, 1998).

The primary ore mineralogy is dominated by chalcopyrite within a magnetite-carbonate gangue. The mean magnetite content of the primary ore is between 20 and 25 wt%. The copper grade increases with increasing magnetite grade (EHM Feasibility, 1995).

The Ernest Henry underground deposit is known to extend to at least 400 m below the open pit final stage 7 (575 m deep, 1.5 km x 1.3 km). In June 2010, lithology caving estimates were given at 76 million tonnes containing 1.3 %Cu, 0.7 g/t Au and 28% FeOx which is made up of measured and indicated resource, and 13 million tonnes containing 1.2 %Cu, 0.6 g/t Au and 26% FeOx of inferred resource. The width of the mineralised deposit is 200 m. The overview of the underground ore body is shown in Figure 2.

Figure 1: Overview of the EHM magnetite extraction process
The dominant iron oxide species in the ore is magnetite; varying significantly from 4.6% to 46.7%. Historical plant averages from the open pit are 17.6%. Monthly composite analyses of the final concentrator tail for the previous two years are in average 25% magnetite. Magnetite and hematite are well liberated (Middleditch, 2008) showing minimum association with chalcopyrite in the rougher tailings.

TESTWORK OVERVIEW

There have been numerous investigations into magnetite processing at EHM since operations started in 1997, with emphasis on how it can be removed, both for product development, and to assist in reducing the circulating load in the grinding circuit. Early studies using an onsite pilot plant in conjunction with large scale laboratory test work at CSIRO demonstrated that suitable product grades and production rates could be achieved. These test programs demonstrated that magnetite production via magnetic separation was viable, but limited by impurity grades and liberation (Zhu et al. 2007). Most of the programs contained either a comminution or impurity reduction stage. Where grinding was not employed, the yield achieved at acceptable product quality was very low. The results of the historical studies are shown in Table 1.

These early testwork campaigns showed that the magnetite in the EHM comminution circuit is preferentially milled due to the hydraulic rather than size-based nature of the magnetite in the primary classification circuit of the copper concentrator (Zhu et al 2004). The consequence is that the classifiers treat fine magnetite in exactly the same way as coarser gangue, concentrating it to the cyclone underflow resulting in preferential liberation of magnetite in the ball mill. Practically, due to the relatively liberated nature of the feldspar and magnetite, the hydro- cyclones operate with two independent efficiency curves (one for magnetite and one for feldspar), which overlap as a function of co-liberation of the two minerals (Zhu et al 2006).
Table 1
Historical testwork results

<table>
<thead>
<tr>
<th>Program</th>
<th>Tailings Sample</th>
<th>Product</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Grind P80</td>
<td>Yield (%)</td>
</tr>
<tr>
<td>EHM 2001 Laboratory</td>
<td>No</td>
<td>0.60</td>
</tr>
<tr>
<td>CSIRO August 2004 Pilot Run 1</td>
<td>No</td>
<td>47.2</td>
</tr>
<tr>
<td>CSIRO August 2004 Davis Tube</td>
<td>39μm</td>
<td>28.8</td>
</tr>
<tr>
<td>IMO Nov 2004 Pilot Phase 5</td>
<td>No</td>
<td>20.3</td>
</tr>
<tr>
<td>CSIRO Feb 2007 Laboratory</td>
<td>38μm</td>
<td>20.5</td>
</tr>
<tr>
<td>CSIRO June 2008 Laboratory</td>
<td>35μm</td>
<td>23.3</td>
</tr>
</tbody>
</table>

In 2008, after assessing the strengths and weaknesses of the previous programs, further metallurgical test work was conducted by Xstrata as part of a feasibility study to demonstrate if an iron ore concentrate meeting steel industry specifications could be produced at a reasonable yield by controlling particle size distribution. The work was completed at laboratory pilot scale, and used the laboratory flowsheet shown in Figure 3.

![Magnetite Production from EHM Tails](image)

**Figure 3 - Final design metallurgical testwork flowsheet**
The metallurgical testwork completed confirmed that magnetite concentrates containing 69% iron with minimal impurities can be produced with suitable mass recoveries by controlling particle size distribution of the magnetite product (Magee, 2009).

In addition extensive mineralogy was completed on test products, and demonstrated that the majority of the rougher magnetic separator losses were in the <20wt% Fe-Oxides liberation class. The majority of the losses from the cleaner magnetic separators were in the form of hematite, which is co-reported with magnetite as Fe-Oxides, but is non-magnetic. Approximately 15-20% of the Fe-Oxides in the EHM final tailings are hematite, rather than magnetite. The hematite is present in rims or inclusions around the partially liberated magnetite particles as shown in Figure 4.

![Figure 4 – Hematite rimming of magnetite particles](image)

**Figure 4 – Hematite rimming of magnetite particles**

H = Hematite  
M = Magnetite

In the 2009 testwork hematite represented the majority of the Fe-Oxide losses. Hematite recovery cannot be practically improved and hematite as it is not magnetic and will decrease with fine grinding due to derimming of the composite particles. Evidence of this exists throughout the circuit, but is clearest in the rougher non-magnetics stream, where 90% of the Fe Oxides that are lost are actually hematite rather than magnetite (Magee, 2009).
FLOW SHEET DEVELOPMENT AND EQUIPMENT

In order to benchmark the proposed flowsheet and to understand the process risks and plant operation site visits to La Candelaria and Los Pelambres in Chile were undertaken. These sites had similar magnetic separation and filtration technology to that being considered for the project. Ceramic disc filters were observed to be working effectively in concentrate filtration applications at these sites. The magnetic separation plant was observed to have poor process control but was achieving the desired product quality and demonstrated the robust nature of the magnetic separators to a range of operating conditions.

The magnetite plant is divided into 3 circuits. The extraction (magnetic separation) and dewatering (filters) form the base plant (which has been commissioned). The regrind circuit which is due for SMP construction in June is scheduled for process commissioning in August this year. A flow sheet of the base plant is shown in Figure 5.

Figure 5 – Magnetite extraction plant flowsheet for Ernest Henry Mine in Australia

Base Plant

The base plant is a simple beneficiation process that uses magnetic separators to extract the magnetite from the copper concentrator tailings a detail of the process parameters is shown in Table 2. It is divided into the following circuits:

- Rougher magnetic separator circuit that processes raw flotation tailings
- Primary cyclone cluster that separates out particles less than 75 µm
- Cleaner/finisher magnetic separator circuit that upgrades the rougher magnetic concentrate
- Dewatering circuit that removes the excess water from the recovered magnetite
Table 2
EHM base plant process parameters

<table>
<thead>
<tr>
<th>Circuit</th>
<th>Process Inputs</th>
<th>Process Outputs</th>
<th>Equipment Details</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rougher Circuit</td>
<td>Throughput: 721 t/h-1350 t/h</td>
<td>Mass yield: 30% - 33%</td>
<td>Rougher units constrains (m drum h):</td>
</tr>
<tr>
<td></td>
<td>Magnetite content: 14-24%</td>
<td></td>
<td>Magnetic loading: 26 t</td>
</tr>
<tr>
<td></td>
<td>Feed solids: 35-42%</td>
<td></td>
<td>Volumetric loading: 130 m³</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Classification Circuit</td>
<td>Throughput: 220 t/h-525 t/h</td>
<td>Product target: P&lt;sub&gt;99&lt;/sub&gt; 75 µm</td>
<td>Cyclone diameter: 400 mm</td>
</tr>
<tr>
<td></td>
<td>Feed solids: 33%</td>
<td>Mass yield to COF: 46% - 47%</td>
<td>No of cyclones: 8</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Product solids: 20%</td>
<td>Operating pressure: 80-120 KPA</td>
</tr>
<tr>
<td>Cleaner Circuit</td>
<td>Throughput: 95 t/h-230 t/h</td>
<td>Mass yield: 97% - 99%</td>
<td>Cleaner units constrains (m drum h):</td>
</tr>
<tr>
<td></td>
<td>Feed solids: 20%</td>
<td></td>
<td>Magnetic loading: 26 t</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>Volumetric loading: 105 m³</td>
</tr>
<tr>
<td>Dewatering Circuit</td>
<td>Throughput: 75 t/h-207 t/h</td>
<td>Moisture content: &lt; 8%</td>
<td>Filtration rate: 1.187 t/m²</td>
</tr>
<tr>
<td></td>
<td>Feed solids: 60%</td>
<td></td>
<td>Filtration area: 163 m²</td>
</tr>
</tbody>
</table>

The purpose of the rougher magnetic separation circuit is to separate the coarse magnetite material from the copper flotation tail. The circuit treats the material using four (4) twin magnetic separation drum units. The rougher magnetic separators are permanent low intensity magnets with a magnetic strength of 1000+ Gauss. Each unit is composed of two counter-current wet drum separators running in parallel with one another and fitted with respective launders for concentrate and tailings discharge. The concentrate produced from the separator consists mostly of fully liberated magnetite with some unliberated magnetite and middlings. The rougher magnetic separator recovers 78% of the magnetite from the rougher feed. Figure 6 shows the design and operating conditions of the rougher circuit.

![Figure 6 – EHM Magnetite rougher circuit](image)

The concentrate produced from the rougher magnetic separation still contains considerable amounts of impurities. The test work results from the rougher feed material suggests that particle sizes of less than - 53 µm consist mostly of liberated magnetite while the upper size is composed of unliberated magnetite, which requires further size reduction by milling. A size separation through a cyclone classifier is applied to reject most of the tailings and upgrade the rougher concentrate.

The primary cyclone circuit is designed to separate the fine liberated magnetite from the coarse material. This process stage incorporates a cluster of 8 Cavex 400 cyclones. The
feed density to the cluster is controlled to ensure the correct cut point is obtained while the number of cyclones operating can be altered to match the feed flow volume.

The cyclones produce an overflow product of 20% solids with a P80 of 53µm containing mostly liberated magnetite and a small amount of low-SG gangue material. The cyclone underflow which consists of mostly unliberated magnetite and low-SG gangue is sent to the existing flotation tailings sump. In the size classification circuit, 47% of the magnetite is recovered from the cyclone feed. A regrind circuit has been incorporated into the design to treat the cyclone underflow and recover the unliberated magnetite. Figure 7 shows the design and operating conditions of the classification circuit:

![Figure 7 – EHM Magnetite classification circuit](image)

The effective classification of the magnetite is paramount to the success of the separation process. This is complicated by the bimodal density properties of the high-SG magnetite and the low-SG gangue. This was evident throughout the metallurgical testwork campaigns and the JKTech was approached to model the suitability of 400CVX cyclones as a classification method.

A series of simulations were performed on the magnetite feed using the multicomponent Nageswararao cyclone model in JKSimFloat and independently verified using other tools, including JKSimMet. The plant feed was divided into three density categories and fewer than two feed conditions, underground and open cut mining, which produced different particle size distributions. The main objective was firstly to determine whether the proposed cyclone cluster could handle the feed flow rate and density in the magnetite recovery circuit. Secondly, the predicted flow splits, densities and size distributions of the overflow and underflow streams were also of importance. Finally, the recovery of magnetite to the overflow was a key variable. The outcome of the modelling was that the 400CVX cyclone was deemed to be suitable in the Magnetite recovery circuit at EHM.

The cyclone overflow is gravity fed to the cleaner/finisher circuit. The purpose of the cleaner/finisher circuit is to remove the final gangue that is entrained in the magnetite slurry. This stage incorporates three magnetic separator banks operating in parallel. Each bank comprises three magnetic separators operating in series. The cleaner/finisher magnetic separator works under the same principle as the rougher magnetic separator but at a different magnetic strength. The slurry is introduced into the cleaner separator first, which is a permanent low intensity magnet and operates with a magnetic strength of 750 Gauss. The cleaner separator produces a concentrate which flows into the feed launder of the following finisher drum separator which also operates at 750 Gauss. The last finisher drum separator is operated at 550+ Gauss and serves as a polishing separator for magnetite and non-magnetic gangue materials. At the end of this stage, 99.8% of the magnetite is recovered from the feed and a clean concentrate containing up to 98% iron oxides is produced, which is suitable for sale. Figure 8 shows the design and operating conditions of the cleaner circuit.
The purpose of the dewatering circuit is to condition the extracted magnetite to make it more suitable for storage and transportation. The circuit consists of two CC60 Larox ceramic disc filters. Due to the physical properties of the magnetite and the location of the dewatering plant, the clean magnetite is required to be maintained as a slurry prior to the dewatering process. This is achieved by agitation in holding tanks and continuous circulation of the slurry prior to entering the dewatering filters. Once dried, the product is transferred by gravity directly to the stockpile area located underneath the plant. Figure 9 shows the design and operating conditions of the dewatering circuit:

Regrind Circuit

In May 2005, Xstrata Technology completed ISAMill characterisation testwork on a sample of EHM tailings. The testwork sample had a P80 of 130 μm, and was ground progressively to 8.4 μm. The testwork showed that the ISAMill technology was able to produce material down to 13 μm from a feed size of 113 μm. Alternative technology tested failed to produce material less than 31 μm as shown in Figure 10 (Burford and Niva, 2008).

The reason for the difference between the mills was the smaller grinding media used in the ISAMill testwork. Media selection was based on what a full scale plant can realistically operate at. A full scale ISAMill can be supplied and operated with ceramic media from 1 to 3.5mm, however Tower Mills can only realistically operate with 12mm media and larger, which means that they cannot achieve the grind sizes that a full size ISAMill can.

ISAMills have been in fine grinding applications in base metal circuits since 1994. The ISAMill technology was selected due to its high energy efficiency and intense grinding action. Developments in this technology have allowed the mill to treat coarser feed sizes at high energy efficiency compared to traditional grinding technologies.
Further regrind studies before and after the magnetite project feasibility study included: Bond Index, Levin tests and ISAMill signature plots were carried out on EHM rougher magnetics. These additional studies identified that with rougher magnetic concentrate regrinding, Fe-Oxide recovery to final magnetite product could be increased from 48% to 80% (Magee, 2009). The regrind testwork showed that 1xM10000 (3MW) Isamill would be sufficient to process the cyclone underflow stream followed by rougher magnetic separation. The rougher magnetics will be pumped to the existing cleaner circuit in the base plant which has been designed to accommodate the extra flow.

COMMISIONING AND EARLY OPERATION

The EHM plant was designed to produce a premium product to supply steel industry requirements. The product specification will be published after the completion of the process commissioning. Process commissioning of the base plant commenced in January 2011 however it was delayed by the effects of the Brisbane Floods and Cyclone Yasi which affected a large area of South-East and North Queensland. Early results have achieved a product with the desired particle size distribution as shown in Figure 13. In addition the magnetite concentrate grade is approaching the desired target. The current objective is to produce a consistent product specification followed a yield optimisation stage.

In addition the magnetite project team is also working to complete the operational readiness plan to hand over the operation of the magnetite plant to the copper concentrator team. The operational readiness tasks include asset management plans such as maintenance schedules and critical spares, training material and the development of operating parameters, targets and KPI's.
The slurry commissioning plan includes a series of sampling campaigns to confirm the design parameters, optimise the operation of the plant, and develop guidelines for the daily operation. The sampling and testwork is focusing on collecting sufficient data points for statistical analysis.

- Confirm the specific gravities of all the process streams and compare with the design assumptions.
- Cyclone surveys to develop a size by size database to determine the optimum cut point for the cyclone overflow to achieve the desired magnetic separation.
- Determine the optimum cyclone operating parameters to achieve the target cut size.
- Determine cleaner efficiency and develop operational guidelines for water dilution to the cleaner circuit.
- Maximise ceramic filter throughput.
- Determine what impact if any water impurities could have on the final concentrate specification.

Future testwork is scheduled to understand and optimise the plant performance including full plant surveys, mineralogical analysis of all the process streams as well as IsaMill signature plots on the cyclone underflow.

CONCLUSIONS

The EHM magnetite extraction plant is part of the life of mine extension at Ernest Henry Mine and part of our strategic objective of maximising the value of our existing resources. It will provide an important source of growth and employment for the region. The EHM magnetite project has utilised a concise project management strategy and proven technology to maximise the value for shareholders in our pursuit for zero waste.
REFERENCES


Development of an Innovative Copper Flowsheet at Phu Kham

M F Young and I Crnkovic

Phu Kham Location in Laos
Phu Kham Plant Layout

Plant Layout showing

- Primary Grinding
- Rougher Flotation
- IsaMill Regrinding

Phu Kham Flowsheet
Phu Kham Flowsheet

- 12.8 Mt/a of copper-gold bearing ore from the open pit
- Plant Feed Grades are 0.75% Cu and 0.33g/t of Au and 3.8g/t of Ag
- Concentrate quality is +25% Cu, 7 g/t of Au and 60g/t of Ag

- Primary Grinding Circuit contains
  - 34 ft × 18 ft (13MW) SAG mill
  - 24 ft × 39 ft (13 MW) ball mill
- Regrinding Circuit contains
  - M10,000 (2.6MW) IsaMill

- Ore has fine locking of copper and gangue minerals
- Ore requires regrinding rougher concentrate to 38 microns to make good quality final copper concentrate

Cleaner Feed Performance

- IsaMill currently grinds from 90 microns to 38 microns
- Cleaner feed performance at 38 microns and 25 microns
- Laboratory Flotation Tests
• Cleaning Circuit was overloaded at times

• Fine grinding to improve concentrate quality, slowed flotation rate and increase frothing issues in the cleaner circuit, decreasing the performance of the cleaner circuit

• Jameson Cell with froth washing was installed at the head of the cleaner circuit to increase cleaning capacity

• This new circuit allowed final concentrate to be produced from the IsaMill regrind product in one flotation cell and remove more than half the load from the existing cleaning circuit
Layout of IsaMill and Jameson Cell

Jameson Cell Flotation at Phu Kham
Phu Kham Jameson Cell
Grade-Recovery Curve Performance

- IsaMill Cyclone Overflow P80=20 microns
- IsaMill Discharge P80=38 microns
- Feed (Rougher Conc) Grade = 3-5% Cu

![Graph showing Cu recovery and grade for different streams with IsaMill discharge]

Jameson Cell Size by Size Recovery on
IsaMill Cyclone Overflow Stream

![Graph showing size-by-size recovery for different cycles]
New Cleaning Circuit Results

- Jameson Cell was initially commissioned on the regrind cyclone overflow
- Cyclone Overflow was fine (P80=20 microns) and well liberated
- Jameson Cell produced 25.5% to 27.5% Cu grade at 75% to 90% Cu recovery on this stream
- Jameson Cell produced high recovery in all size fractions for this stream

- Jameson Cell then treated cyclone overflow plus IsaMill discharge
- Jameson Cell still produced high grade copper concentrate at 26% to 29% Cu grade at 55% to 60% copper recovery
- The recovery of the coarser size fraction was lower due to the locking and lower liberation in the size fractions

Redox Measurements in Plant

Low Eh reduces the copper mineral flotation performance
Impact of Inert Grinding - Chalcopyrite

The Effect of Eh – Mt Isa Cu Ore

The critical importance of the grinding environment on fine particle recovery in flotation - Stephen Grano - 2009

Cleaner Feed Performance

The Effect of Aeration in Laboratory Float
IsaMill Discharge Performance

The Effect of Aeration in Laboratory Float

Grade Recovery Curves (Total Cu) IMD

- IMD Without Aeration
- IMD With Aeration

HYPERSPARGE
Supersonic Gas Injection

Fine Bubble Generation
Able to generate much finer bubbles for fast gas transfer rates
Phu Kham – Effect of Aeration
Plant Surveys

Phu Kham Jameson Cell Performance
Before and After Aeration

HyperSparge installed in IsaMill Discharge line
Summary

- IsaMill regrinds rougher concentrate to 38 micron to improve the final copper concentrate quality

- Cleaner feed performance at 38 microns is 25-28% Cu in final concentrate
  and regrinding 25 microns can produce +30% Cu in final concentrate

- Jameson Cell installation was an innovative flowsheet change to removed load from cleaning circuit and increased plant capacity and decrease cleaner circuit frothing issues

- Aeration of Cleaner feed has improve performance and fines recovery and selectivity

- Jameson Cell has allowed the IsaMill to operate at Increase power to improve the circuit performance.
Regrind Mills: Challenges of Scaleup

Michael Larson1, Greg Anderson1, Dr. Rob Morrison2, Michael Young1
1- Xstrata Technology
2- Julius Kruttschnitt Mineral Research Centre

Abstract:

As ore deposits become finer grained the requirements for regrinding before cleaning or leaching have increased substantially. Despite this increasing need, there is no standard test to predict grinding energy requirement below 70 microns. The standard Bond Mill test applies for coarser ball milling, but is not appropriate for stirred milling with fine media grinding to 70 microns. With no industry-standard test, the energy requirement is often made on the basis of supplier estimates or benchmarking against similar applications. Yet suppliers use vastly different scale-up methods which result in widely different energy estimates. Estimates can differ by 100-300% even for similar mills. This must result in serious errors to install either too much or too little grinding power. This paper explores this by comparing the actual performance of full scale regrind mills against their original design estimate. It confirms that serious scale-up errors have been made, and have then been perpetuated by “benchmarking”.

The test conditions to achieve accurate power estimation are discussed. The essential test conditions are: continuous (not batch) tests, ensure steady state (no coarse fraction retained in test mill), correctly account for classification, measure energy directly (not inferred), and using the same media size as the full scale installation. Failure to correctly address even one of these factors can bias results by 40%; failure to address several multiplies the error.

Introduction:

The requirement for regrinding mills has increased as ore deposits have become more complex and fine grained. This increased requirement has not yet been matched with accurate industry design tools for fine grinding. There is no industry-standard test for fine stirred milling. The Bond test and other scale-up tests designed for coarser grinding do not apply below 70 microns for stirred milling. Using these coarse grinding techniques for fine grinding underestimates energy requirement because of the different ball sizes, different trajectories and different ball-particle and ball-shell interactions in small mills. Modified Bond tests such as the Levin test are more accurate but are often regarded to be too time consuming and requiring too much sample. As a result, regrind mill suppliers have developed their own scale-up methodologies. Yet in recent experience these different scale-up methods result in 300-400% variations in estimated energy for the same duty. In several recent cases, estimates for the same mill from different manufacturers have varied by over 100%. Clearly they can’t all be right. Plant designers need a standard technique to select the correct energy, and need rigorous data on actual operation performance (not design performance) for benchmarking. Without such a standard, the temptation is to choose the lowest energy estimate, perhaps supported by benchmarking against other designs rather than performance. This approach could perpetuate incorrect designs, and will lead to project failures.

This paper compares actual regrinding performance with design performance. The poor design record is explained by inadequate laboratory procedures. Different procedures and interpretations are reviewed, and the conditions required to achieve accurate scale-up for any regrinding technology are described.

Scale-up Procedures for IsaMill™

MIM developed the IsaMill™ in the early 1990’s in response to fine regrinding needs of its ore bodies. Because they were also the operators, the MIM engineers required accurate scale-up. Also, they were able to calibrate predictions against operating mills. As a result they developed procedures for accurate 1:1 scale-up from laboratory to plant, and achieved this with both the MIM George Fisher and the MRM 1.1 MW, 3,000 liter IsaMills (M3,000s) (Figure 1). In this scale-up procedure, an M4 (4 liter horizontal) IsaMill™ is operated in multiple pass mode where the product from one continuous pass becomes feed for the next pass to develop a “signature plot” of energy versus different product sizes. This
mill is filled 80% with media of the same size as final application (typically 1-3 mm), and agitated at 1500 RPM. The mill configuration and operation – discs, disc speed; media size, media type and media filling; and mill discharge classifier – are all very similar to the full scale operation. Therefore the feed particles “see” the same grinding and classification mechanisms as they do at full scale. Because the grinding media is small relative to the test unit, mill shell effects are minor, as demonstrated by the 1:1 scale-up performance.

The reasons for the accurate direct scale-up from the laboratory to full size IsaMills are:

- Using the same size and type of grinding media as full scale - no correction factors are needed.
- The mill configuration and mechanism is the same – horizontal mill, similar grinding discs with similar tip-speeds, and closing the mill with an internal classifier similar to full scale. Therefore the mechanisms, velocities and physics within the laboratory mill are very similar to the full scale mills.
- A large test mill relative to particle and media sizing, so shell effects are minimal.
- Testing with the same slurry feed size and density.
- Ensuring enough feed material is processed to reach steady state. Each point on the “signature plot” results from a continuous pass through the mill, during which the mill reaches steady state. The mill contents must be displaced several times in each pass, as retention of even a small amount of coarse particles will seriously underestimate full scale steady-state power. This is a crucial point that simply cannot be achieved in a batch test, because it does not continuously add new material and can’t ensure steady state. This requirement to achieve steady state is crucial for fine regrinding – fortunately it is also possible with relatively small samples (15 kg).
- Direct measurement of energy consumption from the agitator shaft, in the same way that energy is measured for full scale mills. This also done to accurately subtract the no-load power ensuring the lab mill will scale-up to any size mill.

Having developed this procedure for internal MIM operations use, it has now been successfully applied in other installations. Figure 2 demonstrates the scale-up performance for the 2.6MW (3500 hp) M10,000 IsaMill for Anglo Platinum.
Western Limb Tailings Retreatment Scale-up

<table>
<thead>
<tr>
<th>IsaMill Model</th>
<th>Installed Power (kW)</th>
<th>Chamber Volume (L)</th>
<th>Specific Energy (kWh/t)</th>
<th>Pulp % Solids</th>
<th>P80 (µm)</th>
<th>P10 (µm)</th>
</tr>
</thead>
<tbody>
<tr>
<td>M10,000</td>
<td>2,000</td>
<td>16,000</td>
<td>37</td>
<td>39</td>
<td>47.5</td>
<td>18.0</td>
</tr>
</tbody>
</table>

Figure 2. Western Limb M10,000 Scale-up (Curry D, Clark L, Rule C; 2005)

SMD Scale-up:

The scale-up performance of stirred mill detritors (SMD’s) at the Century mine has been reported. (Gao M, Reemeyer L, Obeng D, Holmes R; 2007). The duty at Century is fine lead-zinc regrind, very similar to the George Fisher and MRM duties reported above. However for Century, an error of 20% was reported from the 0.55 kW lab mill to the full scale 355 kW mill- that is the full scale mill was less efficient than predicted from testwork. This may be attributed to measuring laboratory power draw by grinding chamber reaction torque rather than by direct measurement (Nesset J, Radziszewski P, Hardie C, Leroux D; 2006), but the power measurement had been corrected to be a direct power measurement and there was still a 20% error from the 0.55 kW lab mill to the full scale 355 kW mill, so this error may be due to not reaching a steady state operating condition in the mill.

SMD Scale-up

The scale-up procedures developed for the IsaMill were then compared against the full scale Century SMD’s. The IsaMill M4 procedure accurately predicted the full scale Century performance on a P80 basis. This is intuitively correct – though the mills have different designs (horizontal versus vertical), essentially they stir the same media at similar speeds, so should be expected to have similar energy requirements for gross size reduction. The only difference between the tests was the finer P90 (sharper size distribution) from the IsaMill, attributable to the internal product separator and the bypass losses that occur in the full scale SMDs.

Figure 3. SMD Century scale-up (Gao et al, 2007)
Errors Caused by Inadequate Top Size Breakage

A continuous test that simulates full scale grinding mechanism should give accurate scale up. However rigour is required to ensure coarse particles do not accumulate in the mill, and are adequately broken during the test. The importance of avoiding build up is discussed later – using too small a sample to reach steady state will underestimate energy. But failure to break top size particles can overestimate grinding power. If grinding media is too small to break the largest particles they will accumulate in the mill and displace grinding media. This reduces grinding efficiency while increasing power draw (analogous to “critical size” build up in a SAG mill). This reduction in grinding efficiency is not always evident if the particles are retained in the mill.

This was almost certainly the cause of the analogous data reported by Farber et al. This work compared two different media types, A and B. The smaller media appears to be more efficient but the results are misleading because of coarse material holdup in the mill. When this coarse material is taken into account the energy requirement is much higher. The first graph in Figure 5 demonstrates the lower friction of Type B media, leading to lower power draw on water. As expected, for any media type, smaller particles need less energy to mix than coarser ones. (it is easier to push your hand through a bucket of 3mm balls than a bucket of 12 mm balls). The same effect should be seen when operating with slurry, yet the second graph in Figure 5 shows the opposite effect – both media types are reported to consume more energy using fine media in ore grinding (“UG2” platinum ore). This almost certainly indicates the tests were in error – the fine media (1.7mm Type B and 2 mm Type A) was too small to break the top size, causing reduced grinding efficiency and holdup of coarse material in the mill. Therefore the resulting signature plots (Figure 6) are incorrect.

Media Power Draw

Figure 5. Power draw in water (left) and UG2 slurry (right) for first pass of Signature Plot test. (Farber et al; 2010)
Indeed, there is a “fingerprint” of this error in Figure 6: the first point for 1.7mm Ceramic B does not fall on the same line as the other 3 points, rather it is finer than expected. This is a clear indication that coarse material was held in the mill in the first pass, that steady state had not been reached, and so the test is invalid. Note that this point’s deviation does not “look” very significant on the wide log-log scale used in Figure 6. However it is a large deviation that would be much more apparent if the graph was redrawn with a better scale (10 to 100 kwh/t rather than 1 to 1000 kwh/t). Beware of the dangers hidden by log-log scales in fine grinding!

Similar investigations which correctly ensured top size breakage and steady state were conducted by Larson. Different sizes and types of media were compared in water and grinding copper ore. This yielded the expected results – smaller media is more efficient, drawing less power in both water (Figure 7) and copper ore (Figure 8), so long as it breaks top size particles.
Figure 9 shows the difference in net power draw between the same two media charges but across different grind sizes. These were taken from individual signature plot tests on the same ore under the same operating conditions. Both media were capable of breaking the top size. The 5mm media resulted in a slightly smaller $P_{98}/P_{80}$ ratio. It was 2.6 compared to 2.7 for the 3mm.

![Media Size vs Power Draw in Copper ROM Ore](image)

Figure 9. Hira media size vs. power draw in a copper ROM ore. (He M; 2010)

Again, as shown in Table 1 and Figures 8 & 9 the smaller MT1 media will also draw less power under the same slurry conditions. Further, as the slurry is ground finer the power draw will decrease. In Figure 8 the first points at 1.2 kW are almost the same power draw. At these points the 2.5mm media produced a $P_{98}$ of 95µm and the 3.5mm media produced a $P_{98}$ of 63µm. The finer points had closer ratios as the 2.5mm media was more capable of grinding the top size in these passes.

<table>
<thead>
<tr>
<th>Media Size</th>
<th>1.5 mm</th>
<th>2.5 mm</th>
<th>3.5 mm</th>
</tr>
</thead>
<tbody>
<tr>
<td>Net Mill Power</td>
<td>.80 kW</td>
<td>.98 kW</td>
<td>1.07 kW</td>
</tr>
</tbody>
</table>

Table 1. Average net mill power draw for copper concentrate varying only media size (Larson M; 2010)

Morenci M10,000 Scale-up:

The design of the Morenci M10,000 is described below and in Table 3. Though the M4 IsaMill test accurately predicted energy requirement to achieve $P_{80}$, it demonstrates other important factors required for accurate prediction of the $P_{98}$, as explained below (Cole J, Wilmot J; 2009).

The design was for a $P_{80}$ of 7 micron and a $P_{98}$ of 15 micron. Actual feed sizes were $P_{80}$ of 11 micron and a $P_{98}$ of 34 micron with the Morenci concentrates. The Morenci concentrates contained variable amounts of pyrite which is more difficult to grind than a relatively pure chalcopyrite. Due to the mineralogy of Sierrita concentrates, containing approximately 85% chalcopyrite, a switch was made to process Sierrita concentrates. Sierrita concentrates were less variable in composition and softer than Morenci concentrates. Sierrita concentrates are also similar to the Bagdad concentrate which was used to size the IsaMill. The $P_{80}$ and $P_{98}$ from the Sierrita concentrates were finer than the Morenci concentrates, at 7.4 micron and 25 micron respectively, but were still not at the design criteria.

The design pyrite level in the concentrates was 22.5%, the actual varied from the low teens to as high as 55% on Morenci concentrates.
Morenci concentrate leach plant design criteria.

<table>
<thead>
<tr>
<th>Design Criteria</th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Feed Rate, mtph</td>
<td>26.4</td>
</tr>
<tr>
<td>Super Fine Grinding</td>
<td></td>
</tr>
<tr>
<td>Feed Rate, mtph</td>
<td>32.0</td>
</tr>
<tr>
<td>P80, µm</td>
<td>7.0</td>
</tr>
<tr>
<td>P98, µm</td>
<td>15.0</td>
</tr>
<tr>
<td>kWh/t</td>
<td>68.0</td>
</tr>
</tbody>
</table>

Table 2. Morenci M10,000 design criteria  (Cole J, Wilmot J; 2009)

Even with the difference in ore mineralogy, this inability to achieve the P98/P80 ratio as designed from the original testwork suggests that a small fraction of coarse, hard material was not fully ground and remained in the test mill. Though this is the only case so far when P80 scaled but the P98 did not, it suggests that if the objective of the testwork is highly accurate prediction of P98, then the standard 15 kg of sample for an M4 may be insufficient.

Centerra Gold’s Kumtor Mine has a similar duty with M10,000 IsaMill grinding gold/pyrite feed from 20µm to a P80 of 10µm prior to cyanide leach. Figure 10 shows this scale-up was successful, including the predicted P98 (in fact, the full sized mill produced a slightly better size distribution than predicted).

Figure 10. Kumtor M10,000 scale-up  (Kazakoff J, Mortimore A, Smith S, Curry D; 06)

Tower Mill Procedures:

The test conditions for the Nippon Eirich Tower Mill are shown in Figure 11. These procedures raise several questions when compared with the requirements described above. Firstly, recirculating mill discharge directly to the mill feed pump means that steady state can never be reached – mill feed is almost instantly contaminated with fines from the “first pass” (in contrast, the M4 IsaMill tests passes the sample continuously from the feed tank through the mill into a separate product tank, reproducing full scale grinding).

![Figure 11 - Laboratory Testwork Data (Corrected for Feed) vs. Commissioning Results (Tests 1 to 6)](image-url)
Secondly, these tests appear to use a sample size far too small relative to the size of the test mill. Mill void volume must be replaced at least 3-4 times to achieve steady state and avoid hold-up of coarse particles (which causes serious underestimation of energy). Table 3 indicates this Tower Mill testwork uses about half the volume required to reach steady state.

### Mill Type Test Feed vs. Void Space Ratio

<table>
<thead>
<tr>
<th>Mill Type</th>
<th>Mill Open Volume(L)</th>
<th>Solid Volume(L)</th>
<th>Ratio (recommended above 3)</th>
</tr>
</thead>
<tbody>
<tr>
<td>M4 IsaMill</td>
<td>1.35</td>
<td>5.0</td>
<td>3.70</td>
</tr>
<tr>
<td>NE008 Tower Mill</td>
<td>2.35</td>
<td>3.33</td>
<td>1.42</td>
</tr>
<tr>
<td>KM-5 Tower Mill</td>
<td>35.2</td>
<td>50.0</td>
<td>1.42</td>
</tr>
</tbody>
</table>

Table 3. Grinding test solids/open volume ratio comparison

The impact of too small a feed volume is demonstrated in Figure 12 from tests on a pilot Tower Mill. The signature plot shifts significantly from 50 liters of feed (0.7 times mill open volume) to 150 liters of feed (2 times open volume). Clearly the mill is not at steady state with the smaller volume. In this case the energy estimate to grind to a P<sub>80</sub> of 70µm increases by 30% at the higher feed volume. In fact, the feed volume needs to be even higher than this, so the true error is even higher.

Combining too small a sample size with the error of diluting new feed with mill discharge compounds the error further, and is likely to lead to serious underestimation of true power requirement in a plant situation.
Extensive testwork has been done to ensure this is not a problem with the standard 15 kg M4 IsaMill testwork. One example is shown below comparing the signature plots generated with masses of 15 and 35 kg of common iron ore concentrate. In this case both tests consisted of a magnetite concentrate with a $F_{80}$ of 65 microns being ground with 3.5mm media. The 3.5mm media was selected for its energy efficiency and ability to better break the top size than would happen with a smaller media. The resulting steep size distribution would show an improvement in final concentrate grade of 2-3% depending on the target grind. Both the $P_{80}$ and $P_{98}$ plots fall within the margin of error for the M4 test. The iron ore had a solid SG of 4.2. This is on the high side of what is normally tested. Anything lighter will take up more volume so the conclusions from this testwork will apply to those samples as well.

The two tests resulted in respective points at 16.5 and 16.7 µm $P_{80}$’s at just over 40 kWh/t. A comparison of those two size distribution curves is given below. The shape is nearly identical. This would also indicate that at both sample masses the mill is discharging the same steady state sample with no segregation effects.
Nesset Regrind Comparison:

The 2006 paper by Nesset et al compared the grinding efficiency of a ball mill, IsaMill, SMD and Vertimill. It concluded that the SMD procedure of measuring power by reaction plate torque would underestimate energy requirements by 32% and 39% for the two media types that were tested. Once the SMD power was measured from the motor there was little difference in energy requirement between the IsaMill and SMD on a P80 basis, though there were differences at the P98 level. This supports the findings reported above. However, Nesset’s paper also concluded that the laboratory Tower Mill was 43% more energy efficient than either the IsaMill or SMD, for the finest media tested in the Tower Mill. Since both the IsaMill and SMD were specifically developed and determined to be more efficient than Tower Mills for fine grinding, and this has been demonstrated consistently at plant scale, this conclusion needs further examination. It demonstrates another potential source of error in laboratory testing – using different size media than full scale.

Media size is crucial for fine grinding efficiency – the need to use finer media to increase energy efficiency drove the development of both the IsaMill and SMD. Nesset’s Tower Mill tests used 5mm media, a size which is never used in plant installations, and the coarser media that were tested in the Tower Mill were much less energy efficient. The finest media practically used in operating Tower Mills is 12 mm (otherwise media floats from the mill). Per cubic metre, 5mm media has 2.4 times the surface area of 12 mm media – a huge difference for attrition grinding in a stirred mill. Nesset et al report the most efficient media sizes for different devices in Table 4. Though 5 mm media is most efficient in the ball and Tower Mill, the key point is that it cannot be practically used at large scale, both in terms of media retention and energy intensity (size of installation). This is precisely why the high speed stirred mill technologies (IsaMill and SMD) were developed – to practically achieve the energy efficiency benefits of fine media.

Efficient Media Size

<table>
<thead>
<tr>
<th>Technology</th>
<th>Media Type</th>
<th>Media Size (mm)</th>
</tr>
</thead>
<tbody>
<tr>
<td>IsaMill</td>
<td>Ceramic beads</td>
<td>2.2</td>
</tr>
<tr>
<td>SMD</td>
<td>Ceramic beads</td>
<td>2.2</td>
</tr>
<tr>
<td>Vertimill</td>
<td>Steel shots</td>
<td>5</td>
</tr>
<tr>
<td>Laboratory Ball Mill</td>
<td>Steel shots</td>
<td>5</td>
</tr>
<tr>
<td>Brunswick Zn Regrind Mill</td>
<td>Steel slugs</td>
<td>16</td>
</tr>
</tbody>
</table>

Table 4. Most efficient media sizes (Nesset et al, 2006)

Even the inappropriate media size does not fully explain the unusual Tower Mill efficiency reported by Nesset. It is likely that the Tower Mill testwork also failed to reach steady state and retained coarse particles in the mill. Not enough experimental information is available to confirm this, but based on reported sample size and number of tests, it appears that the Tower Mill feed may only have been 1.5 times mill void volume, not enough to reach steady state. This suspicion is supported by a graph (Figure 15 below) comparing product size distribution for three different media sizes. This reports that the 5mm media
improved both $P_{80}$ and $P_{98}$ compared with coarser media. This is unexpected – usually coarse media breaks coarse particles more effectively (especially in stirred milling). This counter-intuitive result strongly suggests that coarse particles were retained in the Tower Mill in the test procedure, which would cause serious underestimation of energy requirements as described in earlier sections.

Vertimill Media Effect on Size Distribution

![Vertimill product sharpness and media size (Nesset et al; 2006)](image)

Figure 15. Vertimill product sharpness and media size (Nesset et al; 2006)

Comparison of Tower Mill Design and Operating Data

A short list of Tower Mill design and actual data is included in Table 5. This was compiled by publicly available data (Vendor installation list), AMIRA P336/P9 projects and Xstrata’s own operating data.

<table>
<thead>
<tr>
<th>Application</th>
<th>Energy requirements for a target product size</th>
<th>Design/ Actual</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Vendor Design</td>
<td>Actual*</td>
</tr>
<tr>
<td>Cu regrind</td>
<td>6 kWh/t for 45µm</td>
<td>13 kWh/t for 45µm</td>
</tr>
<tr>
<td>Pb secondary grind</td>
<td>4.4 kWh/t for 63µm</td>
<td>7.9 kWh/t for 63µm</td>
</tr>
<tr>
<td>Pb tertiary grind</td>
<td>7.1 kWh/t for 45µm</td>
<td>12 kWh/t for 45µm</td>
</tr>
<tr>
<td>Zn regrind</td>
<td>5.7 kWh/t for 30µm</td>
<td>9.4 kWh/t for 30µm</td>
</tr>
<tr>
<td>Zn regrind</td>
<td>19.4 kWh/t for 20µm</td>
<td>25.5 kWh/t for 20µm</td>
</tr>
<tr>
<td>Pb regrind</td>
<td>16.7 kWh/t for 20µm</td>
<td>31 kWh/t for 20µm</td>
</tr>
<tr>
<td>Ni regrind</td>
<td>11.7 kWh/t for 60µm</td>
<td>13.5 kWh/t for 60µm</td>
</tr>
<tr>
<td>Fe regrind</td>
<td>9.4 kWh/t for 30µm</td>
<td>13.8 kWh/t for 30µm</td>
</tr>
</tbody>
</table>

*Actual energy requirements were calculated using operating work index from plant surveys

Table 5. Tower Mill design and actual comparison

The operating work index was also calculated for all available full scale mill data and is shown in Figure 16. There is no design data included in this but it should serve as a guide as to what the mills are actually capable of. Recent regrinding designs falling in the lower area of the graph have significantly under quoted actual operating mills.
Modified Bond and Levin Tests:

The Levin test appears to be underused for sizing reground ball mills. Examples of the results are shown in Table 6. The requirement of 20-30 kg may be a prohibitive factor. The Levin test is a modified Bond Ball test, but makes use of finer screen sizes, and finer feed sizing. In the Levin test the lab ball mill is run at varying lengths of time (energy). At the end of each interval the entire mill contents are emptied and screened at that intervals size. Any undersize is replaced with new top size to maintain a constant volume. This is completed from 75, 53, 45 and 38 microns. With 4 tests each requiring about 2 liters of material this could result in a requirement for more material than is possible to produce in a small pilot plant run given that it will likely be a rougher concentrate. However, if the material mass is possible it would appear that the test will give more accurate results than many of the alternatives.

<table>
<thead>
<tr>
<th>Materials and Origin</th>
<th>kWh/t</th>
<th>Grindability test for fine material</th>
</tr>
</thead>
<tbody>
<tr>
<td>GOLD ore, East Driefontein</td>
<td>14.4</td>
<td>14.8</td>
</tr>
<tr>
<td>Gold ore, Libanon</td>
<td>10.8</td>
<td>11.3</td>
</tr>
<tr>
<td>Gold ore, Western Deep Levels, VCR Reef</td>
<td>14.7</td>
<td>13.8</td>
</tr>
<tr>
<td>Gold ore, Western Deep Levels, Carbon Leader</td>
<td>17.6</td>
<td>16.4</td>
</tr>
<tr>
<td>Fluorspar, Chemspar</td>
<td>4.3</td>
<td>3.3</td>
</tr>
<tr>
<td>Copper-lead-zinc ore, Black Mountain</td>
<td>12.8</td>
<td>11.1</td>
</tr>
<tr>
<td>Copper-zinc ore, Prieska</td>
<td>13.6</td>
<td>12.8</td>
</tr>
<tr>
<td>Sand tailing, Crown Mines</td>
<td>9.3</td>
<td>9.3</td>
</tr>
</tbody>
</table>

Table 6. Comparison of Levin test results with full scale ball mills (Levin J; 1989)
**Operating conditions:**

When downstream testwork has been done with product produced from one technology it should not be expected to transfer to a different grinding technology. Each will result in a different size distribution curve for a given P<sub>80</sub>. The ratio between the P<sub>98</sub> and the P<sub>80</sub> should be considered for each.

This is important for the recovery in leaching as shown below. The more top size material present the less liberation there will be. Material ground to the same target P<sub>80</sub> can have three times the top size present. This will also affect the final con grade during flotation as unliberated coarse particles will still float. Potentially these will contribute excess attached gangue minerals to the concentrate.

![Grinding Sharpness vs Leach Extraction](image)

Figure 17. Effect of size distribution sharpness on leaching recovery (Pease J, Young M, Curry D; 2007)

Care should also be taken to do the scale-up testwork at the correct density. The effects of viscosity become more pronounced as the fine sizes common to regrinding are met. At some point depending on the surface area created the slurry will begin to carry the media rather than mixing it to grind. Xstrata Technology currently recommends the use of a Marsh funnel during testwork to control the density to optimize the effects of viscosity on energy efficiency. The Marsh funnel is simply a cone through which the time for the flow of one quart of slurry is measured. The Marsh funnel is not a comprehensive measurement of rheology but does serve to give a quick easy point of measurement ideal for use in multiple lab tests or in the field for site surveys.

![Feed Density vs Energy](image)

Figure 18. Effect of feed density on fine grinding efficiency (Larson M, Morrison R, Young M; 2008)

**The Future:**

Xstrata Technology is also striving to improve the scale-up ability of the IsaMill through sponsored and internal research. One of the most recent developments was the development of a JKSimMet model of the IsaMill. The basis of this model is a relatively simple function for predicting the creation of fines. The normal signature plot has one limitation in that the line created does not usually pass through the feed at 0 energy. This limits the signature plot to predicting energy requirements coarser than the band of sizes produced on the plot. It is possible to extrapolate finer however it is always best to actually cover those sizes in the testwork to be sure of the results.
The Squared Function for Fines Production also creates a linear relationship but also consistently passes through the feed at 0 energy. The model itself was inspired by McIvor’s work (McIvor R and Finch J; 2007) demonstrating that new production finer than a certain target size is approximately linear for rod and ball mills. Plotting the percent passing a size vs. energy did not work for the IsaMill. It was found though that by squaring that % passing value a straight line through each point including the feed was developed as shown in Figure 19 for a variety of cooper ores. The only constraints are that the size shown has to be measureable in all of the samples. This will likely require a size between the F_{10} of the feed and the P_{90} of the product.

![Figure 19. The squared function for fines production applied to different cooper ores (Larson et al; 2008)](image)

This method also enhances the ability to properly size a mill if the design feed size changes after testwork. By keeping the slope of the line parallel and changing the feed size a new energy can be calculated. In Figure 19 a signature plot line is plotted as the squared function. A new coarser feed can be plotted by calculating the new squared value and starting the new energy vs. size line at that point to the left of the original feed line. By keeping the new squared function line parallel to the original one a new energy can be calculated to the desired product.

![Figure 20. Squared function for fines production energy prediction for coarser feed (Larson; 2010)](image)

The likely reason the squared function works is that the attrition grinding process is creating new surface area in a consistent manner. The IsaMill also results in a rounding of particles. As this should be the same for all attrition grinding processes there is a good probability that the squared function or something similar will apply to those as well.
Conclusions:

Currently a wide range of tests are used to predict regrinding requirements, and they result in enormous variation in predicted energy requirement. Clearly serious mistakes are being made, which will either lead to installing too much power, or too little power and resultant plant underperformance. The industry needs an accepted standard test to reliably predict regrinding energy. An essential part of this is an understanding of actual plant regrinding performance compared with design. The JKMRC Fine Grinding review aims to address these issues. Operators, designers and engineering companies should support a rigorous, independent review to establish industry performance and an industry standard.

Until this happens current scale-up methods have to be rigorously examined to ensure the regrind mills will perform as claimed. Designers should perform reality checks on all vendor estimates – not against other designs, but against actual performance of other installations.

Acknowledgements:

The authors wish to thank Xstrata Technology and the JKMRC for the research that went into this paper and for the helpful suggestions in writing it.

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PAPER 14

The Right Tools in the Right Place: How Xstrata Nickel Australasia Increased Ni throughput at its Cosmos Plant

Dan Curry¹, General Manager – Processing
Michael Cooper¹, Manager – Processing
Josh Rubenstein², Senior Process Engineer
Tom Shouldice³, President
Michael Young², Principal Metallurgist

¹ Xstrata Nickel Australasia
   Level 3, 24 Outram Street
   West Perth, WA 6005
   PH: +61 (8) 9213 1588
   Email: dcurry@xstratanickel.com.au
   Email: mcooper@xstratanickel.com.au

² Xstrata Technology
   Level 5, 509 Richards Street
   Vancouver, BC V6B 2Z6
   PH: (604) 699-6402
   Email: jrubenstein@xstratatech.com.au
   Email: myoung@xstratatech.com.au

³ G & T Metallurgical Services Ltd.
   2957 Bowers Place
   Kamloops, BC V1S 1W5
   PH: (250) 828-6157
   Email: tshouldice@gtmet.com

Key Words: IsaMill™, Jameson Cell, Wemco® SmartCell™, Nickel, Process Mineralogy

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ABSTRACT

Xstrata Nickel Australasia’s Cosmos plant in Western Australia is currently implementing a de-bottlenecking project to increase nickel output while maintaining metallurgical performance and a low unit cost for production. Mineralogical analyses indicated that the processing bottlenecks could be removed by simply reconfiguring portions of the circuit as well as adding capacity with technical advancements – notably an IsaMill™, Jameson Cell, Wemco® SmartCell™ and Larox Filter.

The Wemco® SmartCell™, which is particularly suited for coarse mineral flotation, and Jameson Cell, which is particularly well suited for fast floating and fine minerals, combine as rougher scalpers to complete the main stream portion of the flotation circuit. Addition of the two new cells meet the required flotation capacity to handle the forecast increase in nickel feed grade and tonnage.

The addition of an IsaMill™ for regrinding allows the SAG mill to produce a coarser primary product, thus enabling increased throughput. The advantages offered by the IsaMill™ include operation in open circuit (internal classification), an inert grinding environment (ceramic media) for better downstream flotation (“clean” new surfaces), compact footprint (easy to retrofit into existing plant) and energy efficiency. With the new rougher capacity addition, the remaining original plant cells were reconfigured to suit the new regrind mill and cleaning requirements. The existing 24 m² Larox pressure filter was replaced with a larger 32 m² unit to dewater the increased concentrate output.

The paper describes the upgrade at the Cosmos concentrator with particular emphasis on the mineralogical data and metallurgical benefits associated with the changes.

INTRODUCTION

The Cosmos concentrator is located 680 km northeast of Perth in Western Australia. The deposit’s development and operations were started in 1999 by Jubilee Mines NL, Cosmos’ previous owner, with the first production of nickel concentrate in April 2000. The Cosmos concentrator treats a komatite hosted high grade nickel sulphide ore, with sulphides being predominantly pentlandite and pyrrhotite with lesser amounts of pyrite and chalcopyrite.

Jubilee owned some of the most prospective and relatively under-explored nickel ground in the world thus representing an excellent expansion potential. This was well matched with Xstrata Nickel’s growth strategy and in October 2007, Xstrata bid for Jubilee Mines NL. In February 2008, Xstrata assumed management control of Jubilee and established Xstrata Nickel Australasia (XNA) as a new operating unit of Xstrata Nickel.

As part of XNA’s accelerated growth strategy, changes were required to accommodate increased nickel output from the Prospero mine at the Cosmos site (Figure 1). The increase in nickel production would be due to not only higher mined tonnes but also higher grades. The Cosmos concentrator was designed more than 10 years ago, and was built on the premise of
approximately 3 years of operation to treat the Cosmos ore deposit. In considering the age and planned lifespan of the original concentrator and its compact footprint, one might correctly assume there would be some challenges to complete timely upgrades for improved throughputs. Indeed, in November 2008, the Cosmos concentrator had a mass flow capacity of 25tph and nickel feed capacity of 1.3tph. Achieving the 2009 budgeted nickel output meant the concentrator needed to treat up to 2.8tph of nickel in ore feed (increase of 215%), and mass flow capacity through the SAG mill needed to increase to 45tph (an increase of 180%).

![Figure 1: Cosmos SAG Mill Throughput (2009 Actual YTD vs. Budget)](image1)

![Figure 2: Cosmos Concentrator Nickel Feed Tonnes per Month Budget 2009)](image2)

Mineralogical and metallurgical analyses were undertaken and the subsequent results from the various studies revealed that these ambitious targets could be achieved with the addition and rearrangement of several key pieces of equipment. Armed with this information and “the right tools” approach, a de-bottlenecking programme was initiated. While still in progress, the de-bottlenecking process has already demonstrated success and is on track to be completed within a twelve month schedule and, by industry standards, on a shoe-string budget.
Decisions and directions taken for the de-bottlenecking process have been, and continue to be, firmly based on a mineralogical approach. Size by size recovery, as well as liberation and modal analyses, have provided the scientific reasoning for the process changes being undertaken. Measurement of the size by size recovery of target minerals in key process streams was essential in determining the impact of the changes. Ensuring the “right tools” being in the “right place” included rearranging some equipment for a modified flow sheet and adding new technology, such as the IsaMill™, Jameson Cell and Wemco® SmartCell™, to help achieve the new metallurgical benchmarks.

A background description of the original flow sheet and corresponding process mineralogy is presented below, followed by a description of the de-bottlenecking project changes and impact on size by size mineral performance in the concentrator.

**ORIGINAL FLOW SHEET AND PERFORMANCE (JULY 2008 SURVEY)**

This section focuses on the plant design, flotation feed characteristics, flotation performance, final concentrate composition and tailing losses from before the de-bottlenecking process - the last two items being very important considerations. That is, what is present in the concentrate that should not be there and conversely, what in the tailings should not be present? Figure 3 is the original flow sheet of the Cosmos concentrator.

![Original Flowsheet](Image)

The comminution circuit comprised a single jaw crusher feeding directly into a single stage SAG mill in closed circuit with one pair of 400 mm Cavex cyclones. The classification design was inflexible and inefficient resulting in a significant amount of liberated pentlandite reporting to the cyclone underflow. Consequently the SkimAir® flotation cell performance was excellent, but
characteristic of poor grinding circuit efficiency rather than good plant design as these liberated particles should have reported to the cyclone overflow. The SkimAir® recovered the majority of the nickel (61%) at a concentrate grade of 20%. The remaining flotation circuit produced a concentrate grade of 16% nickel and a further 30% recovery. The overall circuit achieved 92% nickel recovery at a concentrate grade of 19% (Table 1).

The flotation circuit used two stages of roughing and three of scavenging in the main stream to treat the flotation feed (Table 2 and Table 3). The first rougher concentrate reported to final concentrate along with the SkimAir® product. Unusually, the cleaner concentrate and tail were in closed circuit with the rougher/scavenger circuit. This resulted in every flotation stream having a mixture of all flotation reagents; all competing simultaneously to activate, collect and depress. Apparently this made the circuit easier to operate but the large re-circulating loads, difficulties in controlling reagent dosages and continual dilution of the cleaner concentrate with rougher feed suggested simple and significant improvements could be made.

As further evidence of over-grinding, the cyclone overflow had pentlandite that was 86% liberated and non-sulphide gangue (NSG) 97% liberated. This degree of liberation was more suitable for a cleaner feed than a rougher circuit and suggested that the SAG mill throughput could be increased and a coarser cyclone overflow produced without compromising rougher recovery.

<table>
<thead>
<tr>
<th>Product</th>
<th>Mass (%)</th>
<th>Assays (%)</th>
<th>Distribution (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Recalculated Feed</td>
<td>100</td>
<td>4.61 0.17 12.2 8.69 0.08</td>
<td>100 100 100 100 100</td>
</tr>
<tr>
<td>Skim Con</td>
<td>13.8</td>
<td>20.6 0.68 34.5 30.6 0.24</td>
<td>61 54 39 49 43</td>
</tr>
<tr>
<td>Rougher 1 Con</td>
<td>8.5</td>
<td>16.4 0.77 27.7 26.1 0.13</td>
<td>30 37 19 25 15</td>
</tr>
<tr>
<td>Final Con</td>
<td>22.2</td>
<td>19.0 0.70 31.9 28.9 0.21</td>
<td>92 91 58 74 58</td>
</tr>
<tr>
<td>Final Tailings</td>
<td>77.8</td>
<td>0.5 0.02 6.5 2.9 0.04</td>
<td>8 9 42 26 42</td>
</tr>
</tbody>
</table>

As further evidence of over-grinding, the cyclone overflow had pentlandite that was 86% liberated and non-sulphide gangue (NSG) 97% liberated. This degree of liberation was more suitable for a cleaner feed than a rougher circuit and suggested that the SAG mill throughput could be increased and a coarser cyclone overflow produced without compromising rougher recovery.
Figure 4 shows the recovery by size in the SkimAir® circuit. Coarse particle recovery is excellent and gangue entrainment low (particularly considering the relatively high pulp density required in this circuit to manage the SAG water balance). Recovery of finer (<50µm) particles was poor however, these were recovered later in the main flotation circuit as shown in Figure 5. Approximately 30% of the liberated pyrrhotite in the cyclone overflow was recovered into the rougher 1 concentrate (maximum recovery occurred for particles sized between 10 and 50µm in diameter). NSG entrainment into the Rougher 1 concentrate was more extensive than with skim concentrate.

As shown in Figure 6, the two main diluents for the SkimAir® concentrate were pyrrhotite and NSG. Respectively, these represented 21% and 12% of the concentrate mass. The rougher 1 concentrate had approximately double the amount of NSG but about the same amount of pyrrhotite as the SkimAir® concentrate (Figure 7). Most of the major diluents were present as liberated grains. The theoretical grade-recovery curve shown in Figure 8 emphasizes that if more of these liberated diluents were rejected, a higher concentrate grade would be achievable.
Alternatively, a coarser grind (resulting in less liberation) would not necessarily affect the final concentrate grade. Using the cleaning circuit in a more conventional way and/or froth washing to reject liberated, entrained NSG could potentially assist in the production of higher concentrate grades.

![Distribution of Diluent Minerals in the SkimAir® Concentrate](image1)

**Figure 6:** Distribution of Diluent Minerals in the SkimAir® Concentrate

![Weight Distribution of in the Rougher Concentrate](image2)

**Figure 7:** Weight Distribution of in the Rougher Concentrate

![Mineralogical Limiting Grade Recovery Charts](image3)

**Figure 8:** Mineralogical Limiting Grade Recovery Charts
Eight per cent of the nickel in the ore was lost into the final tails. 60% of this was rejection of liberated pentlandite grains (Figure 9), a significant portion of which were less than 11µm. This is not unexpected given the absence of a true cleaning circuit and the conflicting pulp chemistry.

![Figure 9: Distribution of nickel losses by size and class into the tailings. Expressed as a Percent of Pentlandite in the Tailings.](image)

DE-BOTTLENECKED FLOW SHEET AND PERFORMANCE (APRIL 2009 SURVEY)

The SAG mill throughput has increased to accommodate the higher mined tonnage and as consequence produce a coarser flotation feed. Figure 10 shows the increase in SAG mill throughput. Significant changes around the SAG mill circuit included changing from a mix of 80/100/125mm balls to only 125mm balls, increasing the SAG shell lifter face angle from 15 to 20 degrees to take advantage of higher mill speed, and increasing the aperture size and open area of the discharge grates. Due to difficulties in balancing the cyclone operation over a wide range of flows with the 2 x 400mm Cavex cyclones, these were replaced with 4 x 250mm Cavex cyclones. Mineralogical analysis had shown that the flotation feed’s pentlandite and NSG liberation state could be lowered with a coarser grind size without harming rougher/scavenger flotation recovery. This would result in the recovery of a larger percentage of composite particles. If the final concentrate grade was to remain constant, then regrinding of these composites would be required prior to cleaner flotation. An IsaMill™ was added to the circuit for regrinding purposes.
A M500 IsaMill™ (as shown in Figure 11) was chosen for the regrind duty because of its small footprint; ideal for such a brownfield expansion. Its established ability to increase concentrate grades, use inert grinding media (clean particle surfaces), produce a narrow particle size distribution (in open circuit) and efficiently grind at low pulp density were all benefits useful to the Cosmos mineralogy and metallurgy. The fresh “clean” particle surfaces that inert media produces allow for optimised mineral separation, lower reagent consumption and higher flotation kinetics (Côté and Adante, 2009; Finch, Rao and Nesset, 2007; Huang, Grano and Skinner, 2006). The type of poorly liberated pentlandite composites that require regrinding are shown in Figure 12.
Lab scale tests on selected concentrates were reground and floated to test the concept. With a finer flotation feed, it was found that the nickel grade/recovery curve could be shifted upwards (Figure 13). The IsaMill™ also allows for much better MgO rejection and, when chemical conditions in the cleaners are set for it, optimal As rejection.

The typical size distribution for the M500 IsaMill™ product is given in Figure 14. The F80 and P80 are around 40 and 20µm, respectively. The specific energy needed for this size reduction is around 20 kWh/t. Of particular note is that the particle size distribution of the product is sharper and steeper than that of the feed. The IsaMill™ directs the grinding energy into grinding the coarse particles, not generating more fines. This is evidenced by the minimal change in the P10, P20 or P30 particle sizes. A 2mm ceramic media, Keramax® MT1™, supplied by Magotteaux, is being used for this reground duty.
Due to the higher head grade and plant throughput rates, additional flotation capacity was required. The first steps taken were to add a Z1600 Jameson Cell treating the SAG cyclone overflow as a rougher scalper and reconfigure the flotation circuit to be completely open (cleaner concentrates reconfigured to report to final concentrates and cleaner tails will report to final tails). After coarsening the SAG cyclone overflow product, the amount of liberated pentlandite in the cyclone underflow reduced and the SkimAir® recovery decreased (Figure 15). The SkimAir® was subsequently reconfigured to be fed from the cyclone overflow as a rougher.

![Figure 14: Typical Size Distribution for the IsaMill™ M500](image)

The second step was the addition of a 40m³ Wemco® SmartCell™ tank cell to increase the roughing capacity, initially added downstream of the Jameson Cell to provide sufficient overall

![Figure 15: SkimAir® Recovery versus SAG Cyclone Overflow Size](image)
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roughing capacity, thereby enabling the remaining original flotation cells to be reconfigured into the appropriate cleaning circuit. Upon commissioning of the Wemco® SmartCell™, it was found that frother carry-over from upstream additions caused excessive frothing, as well as concentrate pumping issues. Piping arrangements in place readily allowed the Wemco® SmartCell™ to be placed ahead of the Jameson Cell, which is the current configuration.

The Jameson Cell was selected for the Cosmos project because it is a small, high-throughput device. A picture of the Jameson Cell installed at Cosmos is given in Figure 16. Like the IsaMill™, the compact design makes it easy to retrofit into existing plants.

![Figure 16: Z1600 Jameson Cell at the Cosmos Concentrator](image)

The Jameson cell was commissioned in late March 2009 which coincided with a 2 unit increase in nickel concentrate grade while recovery was maintained at 90% (Table 4). As shown in Figure 17, the Jameson Cell concentrate grade is consistently above that of the final concentrate. Further, froth washing is used to reduce the entrainment of NSG as shown in Figure 18. It should be noted that the Jameson Cell was commissioned on cyclone overflow; several weeks before the SkimAir® was reconfigured.

<table>
<thead>
<tr>
<th>Month</th>
<th>Final Prod</th>
<th>Mass Assays - %</th>
<th>Distribution - %</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Ni</td>
<td>Cu</td>
<td>Fe</td>
</tr>
<tr>
<td>July 08</td>
<td>Con 22.2</td>
<td>19.0</td>
<td>0.71</td>
</tr>
<tr>
<td>Tails 77.8</td>
<td>0.5</td>
<td>0.02</td>
<td>6.5</td>
</tr>
<tr>
<td>Aug 08</td>
<td>Con 18.7</td>
<td>18.7</td>
<td>0.74</td>
</tr>
<tr>
<td>Tails 81.3</td>
<td>0.4</td>
<td>0.01</td>
<td>5.4</td>
</tr>
<tr>
<td>Dec 08</td>
<td>Con 19.9</td>
<td>19.6</td>
<td>0.65</td>
</tr>
<tr>
<td>Tails 80.1</td>
<td>0.5</td>
<td>0.02</td>
<td>6.4</td>
</tr>
<tr>
<td>Feb 09</td>
<td>Con 22.8</td>
<td>19.9</td>
<td>0.68</td>
</tr>
<tr>
<td>Tails 77.2</td>
<td>0.7</td>
<td>0.05</td>
<td>9.3</td>
</tr>
<tr>
<td>April 09</td>
<td>Con 22.5</td>
<td>21.8</td>
<td>0.75</td>
</tr>
<tr>
<td>Tails 77.5</td>
<td>0.7</td>
<td>0.03</td>
<td>15.5</td>
</tr>
</tbody>
</table>
The final flow sheet change was the reconfiguration of the original rougher/scavenger flotation cells into the cleaning circuit. In making the flotation circuit open, the sequential flotation stages are now more clearly defined. A diagram of the three main components of the circuit is shown in Figure 19. In this strategy the rougher and scavenger cells are designed to maximize nickel recovery. The cleaning section depresses and removes the NSG minerals via regrinding, washing/dilute cleaning and depressants. The re-cleaning stage can be employed when necessary for selective arsenic removal via pH control and cyanide addition.

The advantages of this strategy are three-fold. Sequential reagent addition ensures that reagents are only introduced at the intended stage and not upstream via a recycled stream. For example,
previous studies have shown that cyanide and copper sulphate addition can slow the kinetics of pentlandite. Secondly, the circuit is easier to operate with the various functions defined. Thirdly, the larger residence times allow higher throughput of ore and nickel units.

Size by size recovery analysis of the final concentrate after de-bottlenecking shows good recovery of liberated pentlandite by the flotation circuit (Figure 20), although there are still substantial losses of the sub 10 micron material. At the time of the April 2009 survey, the cleaner changes were not complete so residence times were achieved with high pulp densities. Dilution cleaning in the ultimate circuit should recover additional valuable fines and reduce entrainment of liberated NSG. Recovery of unliberated pentlandite had also improved (Figure 21). The new flotation circuit also showed an improvement in rejecting liberated pyrrhotite. As shown in Figure 22, liberated pyrrhotite made up less than 9% of final concentrate - a significant change from the 15% measured in July 2008 with the previous circuit.
In considering the theoretical grade-recovery curve given in Figure 23, it can be seen that despite some improvement from the original data, there remains opportunities to increase the concentrate grade without sacrificing nickel recovery.
As shown in Figure 24 and Table 5, nickel losses to the final tailings remained similar for the month of April as compared to the previous survey (Figure 9).
Table 5: Overall Distribution of Nickel Losses into the Final Tailings Stream

<table>
<thead>
<tr>
<th>Mineral Class</th>
<th>Distribution by Size Range and by Class</th>
<th>Losses by class</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>&gt;106µm</td>
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FINAL FLOWSHEET

The envisioned final flowsheet is shown in Figure 25. These changes will complete the plans for increased nickel output and lower production costs.
CONCLUDING REMARKS

Performance from the Cosmos concentrator has broadly met targets with respect to the changes made to date through the de-bottlenecking process. Overall concentrate grade and plant recovery have been maintained with respect to more difficult ore treated (nickel grades in feed have decreased, nickel to arsenic ratios in feed have increased and the ratio of massive to more disseminated ore has decreased - all negatively impacting nickel recovery) at increased plant throughput.

Mineralogically, Figures 27 to 29 demonstrate improved pentlandite recoveries across all size fractions, and similarly reduced pyrrhotite recoveries across all size fractions. NSG recovery has remained largely unchanged, but this will be the focus of the upcoming modifications to the cleaning portion of the new flow sheet.

![Figure 26: Pentlandite Recovery Comparison](image)

![Figure 27: Pyrrhotite Recovery Comparison](image)
Most of the de-bottlenecking plant design completed to date has focused on modifications to existing equipment and installations of several key additions such as the M500 IsaMill™, Z1600 Jameson Cell and Wemco® SmartCell™, with the primary objective of increased plant throughput capacity. As demonstrated in Figures 27 to 29, this has been achieved without sacrificing the plant performance.

Final modifications in the cleaning circuit will complete having the “right tools in the right place” to enable enhanced metallurgical performance. Analytical tools such as modal analyses and size by size recovery determinations were and will continue to be essential in identifying how best to use the processing tools installed for optimal mineral separations. The new circuit is key to enabling the Cosmos operation to be a low cost nickel producer into the future.

ACKNOWLEDGEMENTS

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Comparing Energy Efficiency in Grinding Mills

B D Burford¹ and E Niva²

1. Senior Process Engineer, Xstrata Technology, Level 4, 307 Queen Street, Brisbane Qld 4000. Email: BBurford@xstratatech.com.au
2. Undergraduate Student, Luleå University of Technology, Sweden, working for Xstrata Technology, Level 4, 307 Queen Street, Brisbane Qld 4000.

ABSTRACT

The IsaMill™ is challenging the way that plants are designed and operated. This paper challenges existing designs of concentrator flow sheets, particularly focusing on magnetite circuits.

From what has started out as a small scale ultra fine grinding mill in the pharmaceutical and pigment industries, it has been redesigned and improved upon for mineral processing, and has been the mainstay of fine grinding applications for over 10 years. These applications have required energy efficient grinding to succeed, and have been predominately in the base metals industry.

Further development of the IsaMill™ has now resulted in the machines being able to treat larger tonnages, with higher capacity motors. This development, along with the introduction of purpose designed ceramic media, has allowed the mill to treat coarser feed sizes. At the same time, the mill still offers highly efficient grinding, and has enabled it to be operated in coarser tertiary and secondary grinding applications.

The acceptance of the mill in coarser applications, predominately in base metals and PGM applications, has enabled the IsaMill™ to be a serious contender for beneficiation in other minerals. One such application is the potential for IsaMills™ to be part of magnetite flowsheets, which are being considered in Australia to meet the growing iron demand of China. The high energy efficiency of the IsaMill™ compared to conventional technologies, as well as the smaller infrastructure requirements, provides a great opportunity to reduce the power intensity of magnetite circuits.

This paper will examine the use of IsaMill™ technology in conventional grinding applications, including recent testwork in a primary grinding base metal circuit, as well as testwork on magnetite ore comparing a lab scale IsaMill™ with a lab scale Tower Mill.

The growing demand for minerals over the next decade, coupled with higher energy cost, will result in energy efficient technology, such as the IsaMill™, being included in standard circuit design.

INTRODUCTION and BACKGROUND

The development of IsaMill™ technology was driven by the metallurgical requirements of fine grained Lead/Zinc deposits at Mount Isa in Queensland and McArthur River in the Northern Territory, both of which were controlled by Mount Isa Mines Limited (now Xstrata).

The complex nature of both deposits required ultra fine grinding to sizes that were not possible to do economically with the technology that was available in the early 90’s.
McArthur River orebodies were mineralogically complex, and required regrinding down to 7 μm to achieve sufficient liberation to allow the production of a bulk concentrate (Enderle et al, 1997; Pease et al, 2006). In the case of the Mount Isa orebodies, there was a gradual decrease in plant metallurgical performance from the mid 1980’s as a result of decreasing liberation size and increased amounts of refractory pyrite in the ore that saw recovery decrease from 70% to 50% (Young et al, 1997; Pease et al, 2005; Pease et al, 2006). However using conventional ball and tower mill technology to achieve finer grinding for mineral liberation was uneconomic, as well as resulting in a high rate of steel media consumption which contaminated the mineral surfaces with iron deposits, resulting in poor flotation response post regrinding.

In both of these orebodies, a need had arisen for a technology that could grind to ultra fine sizes in metallurgical operations economically without serious contamination of mineral surfaces and pulp chemistry. Testwork was undertaken in the early 90’s at Mount Isa Mines into high speed horizontal stirred mill technology, which was used in pigment and other industries. It was shown at pilot scale that such mills could grind down to the ultra fine sizes required for mineral liberation.

Arising from these findings, a program of major mechanical modification of horizontal stirred mill technology was undertaken between Mount Isa Mines Limited and Netzsch-Feinmahltechnik GmbH (Enderle et al, 1997), the manufacturer of the stirred milling technology, to make the technology more applicable for the mining industry.

After many prototypes, the first full scale model was developed and installed at the Mount Isa Mines’ Lead Zinc Concentrator in 1994. The mill, the M3000 IsaMill™, was quickly installed in other circuits at this concentrator, and was installed in the McArthur River Concentrator in 1995 (Johnson et al, 1998). Later, in 1999, it was commercialised and sold outside of the Xstrata group.

Since commercialisation of the IsaMill™, there is now over 70 MW of installed IsaMill™ power operating around the world, treating materials such as copper/gold, lead/zinc and platinum. While the early installations were applied to ultra fine duties, the IsaMill™ today is being applied to coarser grinding applications, once the domain of tower and ball mills. The application to coarser duties, means all the advantages of the IsaMill™ that was developed for ultra fine grinding can now be applied to the coarser applications.

**IsaMill™ OPERATION**

The IsaMill™ is a horizontally stirred mill consisting of a series of discs rotating around a shaft driven through a motor and gearbox, at speeds ranging from 21-23m/s, with energy intensities up to 300kW/m². The general layout of the IsaMill™ is displayed in Figure 1, while the grinding mechanism is displayed in Figure 2.
In operation, the mill is filled with grinding media between each disc, each one of these segments acting as an individual grinding chamber. When 8 discs are used in the mill, it effectively acts as 8 grinding chambers in series, which also minimises any short circuiting of the mill feed to the discharge. The action of the grinding discs when rotating, radially accelerates the media towards the shell. Between the discs, where the media is not subject to the high outwards acceleration of the disc face, the media is forced back in towards the shaft – creating a circulation of media between each set of discs. Minerals are ground by an attrition action, as a result of the agitated media, with the resulting high energy efficiency being achieved due to the high probability of media-particle collision.

**Energy Intensity**

The high tip speed of the IsaMill™ results in a high energy intensive environment. Energy intensity of the IsaMill™ is significantly higher than any other commercially available grinding equipment, as illustrated in Table 1. Combining the energy intensity and the high grinding efficiency leads to a compact mill, able to be fitted into existing plants where floor space is limited.
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<thead>
<tr>
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<th>Mill Volume (m³)</th>
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<tr>
<td>IsaMill™ - M10,000</td>
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</tbody>
</table>

Table 1: Comparative Energy Intensity of Grinding Technologies

Media

The key to the energy efficiency of the IsaMill™ is the ability to use fine media. While tower mills are typically limited to 10-12mm fresh media sizing, the IsaMill™ can use media as small as 1mm. This results in significantly more surface area per unit volume of media in the IsaMill™ compared to a tower mill.

The IsaMill™ is versatile and able to use a range of media types, including low cost, locally available media such as sand or smelter slag, to provide good grinding performance at acceptable energy efficiency. However, the need for improved energy efficiency at many installations has resulted in the use of high quality, high density ceramics, designed specifically for stirred milling applications, such as Magotteaux Keramax ® MT1™.

Media Retention

Grinding media is retained in the mill without the need for screens, which is why IsaMills™ can use fine media. At the end of the mill is a patented product separator consisting of a rotor and displacement body. The distance between the last disc and the rotor disc creates a centrifuge, so that coarse particles and media move to the outside of the mill, which are pumped back towards the feed end of the mill from the action of the rotor. Meanwhile the fine particles flow through the rotor and discharge from the mill, which means no screens or cyclones are required, and allows the mill to be operated in open circuit without cyclones, reducing capital and simplifying circuit configuration.

Product Size Distribution

In open circuit operation, the IsaMill™ is able to produce a sharp product size distribution, which reduces overgrinding and the creation of ultra fines. Typically the ratio of the P98 to the P80 is around 2.5. This is a direct result of the individual grinding chambers acting in series, preventing short circuiting, as well as the classification action of the product separator. The ability to operate the mill in open circuit greatly simplifies the operating and maintenance strategies of the circuit. Also the sharp product size is beneficial in pipeline design and filtration, due to the reduction of ultra fines and oversize particles.

Inert Grinding

The operation of the IsaMill™ using sand or slag, or more often ceramic grinding media, has a big advantage over steel media in conventional grinding systems, as it greatly reduces the generation of ferrie ions. These are detrimental to flotation and leaching circuits, as the ions form a coating on the mineral...
surface, which hinders the action of the flotation or leaching reagents, resulting in more reagents being used, and may also result in poor metallurgical performance.

Maintenance

The IsaMill™ has been designed to keep maintenance simple. The shell of the mill is simply rolled away from the mill on a set of rails, enabling the disc and internal wear surfaces to be examined and changed if required (Figure 3). The shell liner of the mill is easily replaced as the shell is designed in two pieces.

Wear within the mill is determined by the specific size reduction of the mill, as well as wear characteristics of the minerals, and it is common for IsaMills™ to be operating with availabilities 96% and higher.

![Figure 3 – IsaMill™ Maintenance](image)

**COARSE GRINDING TRANSFORMATION**

The IsaMill™ was developed to enable the fine grained orebodies of McArthur River and Mt Isa to be developed (Enderle et al. 1997; Pease et al, 2006). Grinding down to a P80 of 7μm at high energy efficiency was a big step forward in mineral comminution, however only a small number of mine sites needed grinding down to this size.

However, the development of ceramic media and M10,000 IsaMills™ in recent years has enabled the IsaMill™ to treat coarser feed materials in tertiary and even secondary grinding duties, which has resulted in greater application of IsaMills™ in most concentrators, (Burford, 2007).

While the use of low cost natural media and slag was used in initial IsaMill™ applications, the quality variability and the certainty of supply had a big impact on IsaMill™ operation. In particular the variability of the media shape, SG and size constrained the energy efficiency of the mill when operated with sand or slag, (Curry et al 2005b)
The development of ceramic media designed for use in IsaMills™ by Magotteaux International, was a major step forward for application of IsaMills™ in coarse grinding. This was due to the media having good structural integrity, tough, high SG as well as being designed up to 3.5mm in diameter. (Anderson et al 2006)

In terms of the energy that can be provided by the media particle, the development of larger diameter media made from ceramic increases the energy available for grinding due to the increased diameter of the media, and the increased SG of the ceramic. In terms of the Keramax® MT1™, the SG of the ceramic is 3.7, over 40% higher than that of sand, (SG = 2.6). This is described by the Stress Intensity Relationship in Table 2 (Pease 2007).

\[ E \propto d^3 \cdot v^2 \cdot SG \]

\( E \) = Energy per Media Particle  
\( d \) = media diameter  
\( v \) = media velocity  
\( SG \) = media density

**Table 2 – Stress Intensity Relationship**

The other development in the transformation of the IsaMill™ from fine to coarse grinding applications, was the development of the larger M10,000 IsaMill™ (Curry et al 2005a). As previously described, the WLTRP project by Anglo Platinum in South Africa required large scale grinding mills to treat 53 tph, up to a maximum of 65tph, from a P80 of 75μm down to a P90 of 25μm. This duty required 35 kWh/t, and would have involved multiple numbers of the M3000 IsaMills™. However a joint development between Anglo Platinum, and Xstrata Technology, enabled the much larger M10,000 to be developed. Not only was this mill nearly 3 times bigger in volume than the M3000, it was powered by a 2.6MW, and provided considerable more energy available for grinding, (Figure 4). Later versions were supplied with 3MW motors (Anderson 2006), such versions offering 300kW/m², Larger flow rates could now be treated by IsaMill™ technology.

With the developments of ceramic media and M10,000 IsaMills™, energy efficiency and other benefits that were common in fine grinding circuits, were now transferred to coarse grinding applications.
**IsaMills™ AND THE POTENTIAL OF MAGNETITE GRINDING**

The increasing worldwide demand for iron ore has triggered the development of Australia’s magnetite resources. While once regarded as uneconomic to process, they are now being seriously considered as a potential iron source at the current iron ore prices, driven by high demand from the China.

Magnetite has been regarded as uneconomic to process in Australia due to the infrastructure requirements to produce an iron concentrate, as well as the high energy requirements to grind the ore. Haematite deposits in the Pilbara region of Western Australia have always been the preferred source of iron ore in Australia, due to its relatively low cost mining and processing methods, and its high quality and abundance. However, the surging iron ore price and the high demand for iron, has resulted in many magnetite projects, that were once regarded as marginal, being regarded as commercial propositions.

To date, there are approximately 5 projects planning to treat magnetite ore, scheduled to be started up over the next 5 years, with many others being considered. By 2014 this would mean approximately 40MT of ore will need to be processed yearly (Gardner-Bond, 2008; Australian Resources, 2008).

Magnetite is also a key source of iron in many countries where haematite resources are not present, and regions such as Northern Africa, Central Europe, China, and North America have operations mining and treating magnetite. In these operations, traditional comminution technology such as ball mills and tower mills are common practice, and the large tonnages treated in these plants results in large amounts of power being used for grinding the ore.

However, with the development of IsaMill™ technology for coarse grinding applications in base metals, there is also significant potential for the IsaMill™ to be used in magnetite operations. More energy efficient grinding in these plants with IsaMill™ technology, could result in less installed power being required, compared to traditional plant design utilizing ball mills and tower mills, with less infrastructure required such as cyclones and pumps.
MAGNETITE TESTWORK

The objective of the test was to do a side by side comparison for a magnetite sample being ground by an IsaMill™ and a tower mill to determine the signature plot of each mill. Both mills were to be run in open circuit. Davis tube testing was also undertaken on the samples after grinding.

The testwork was undertaken at the CSIRO facilities at Pullenvale, Queensland. At this site there is a M4 IsaMill™ lab scale unit, as well as a small scale tower mill.

Magnetite ore for the testwork was sourced from Ernest Henry Mine tailings (EHM). This mine is a copper mine in North West Queensland, with the host rock being a breccia, which is comprised of strongly altered and replaced felsic volcanic fragments in a matrix assemblage of predominantly magnetite, chalcopyrite and carbonate minerals. Post flotation, the majority of the chalcopyrite and some other sulfides have been separated from the gangue stream, leaving it rich in magnetite.

The M4 IsaMill™ is a 4 litre mill, containing 7 discs for this testwork, operating with 3.5mm ceramic Keramax® - MT™ media. The tower mill is 40L capacity and operates using 12mm steel media. These are displayed in Figure 5.

![M4 IsaMill and small scale Tower Mill](image)

**Figure 5: M4 IsaMill and small scale Tower Mill**

For the tests, the as received sample under went sample preparation and preliminary grinding to reduce the feed size from 163μm to 113μm to eliminate oversize particles blocking the test rigs.

In each test, 20 to 21kg of sample was made into slurry of 50% solids and pumped through each mill for a number of runs. The power used for each run is recorded, and a small sample of the discharge is taken for laser sizing. This procedure is carried out through each mill for a minimum of 12 times, or until there is no significant reduction in the sizing, i.e. the mill cannot reduce the sample any further. The data is then used to draw a signature plot (a log-log graph of P80 size versus the specific energy to obtain that size), as displayed in Figure 6.

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Grinding Results

The signature plot for the two tests, for each grinding technology, shows a good level of reproducibility.

The IsaMill™ was able to reduce the top size of the feed, at a F80 of 113µm, down to a P80 of 13µm. The tower mill also treated the same feed size, however couldn’t produce grind sizes down any further than 31µm. For the testwork, a charge similar to what a full scale grinding unit would use was used. In the case of the IsaMill™, 3.5mm ceramic media (Magotteaux Keramax® MT1™) was used, while in the tower mill, 12mm steel media was used. As expected, the smaller media in the IsaMill™ enables finer grind sizes to be achieved, while the 12mm balls in the tower mill limits how fine the tower mill can grind.

The flatter curve for the IsaMill™ signature plot indicates less energy required to achieve grind size, than the steeper curve that was obtained with the tower mill. This difference has a big impact on the energy needed to grind down to the required product sizes. For instance, to reduce a theoretical sample with a P80 from 100 µm to 30µm, using the tower mill will take approximately 39 kWh/t. However to do the same size reduction with an IsaMill™ will take only 13 kWh/t, only a third of the tower mill power. The key to the signature plots is that for the coarser sizes, there is only a small power requirement to grind the coarse sizes. However as the size required becomes smaller, there is an exponential increase in the power required. Therefore while the tower mill may be more power efficient at sizes greater than 65µm, a reduction in particle size less than 65µm for this sample will result in the IsaMill™ being more efficient.
One scenario that was not tested was using the tower mill in a closed circuit, as is often the case in practice. However setting up such an experiment is difficult due to recirculating loads and ensuring the cyclones cut efficiently. This is one of the practical drawbacks of closed circuits, in that cyclones never operate efficiently, and are often poorly maintained, and small diameter cyclones required for fine cuts, are prone to blockages. Also the associated power of pumping at a reasonable pressure for the cyclone to operate needs to be taken into account in the energy use in these circuits.

The signature plot was restrictive in the scale that could be achieved with the IsaMill™, as the feed for each test was maintained at a P80 of 113µm. In practice, coarser grinding can take place at increased sizes between P80’s of 250 to 300µm, and larger media is being developed to handle even coarser sizes.

**Davis Tube Results**

The samples from the grinding testwork underwent Davis Tube testing, which involved the separation of the magnetics from non-magnetics using a small scale magnetic separator. The iron grade versus iron recovery obtained from this testwork is shown in Figure 7. The grade recovery relationship indicates the maximum grade for the ore type was 71% iron, at a maximum iron recovery of 90%. The assay from the magnetic separation from the Davis Tube testwork gives an indication of how the liberation of the minerals occurs as the particle size reduces.

![Fe Grade vs. Recovery](image)

**Figure 7 – Iron Grade vs. Recovery**

The EHM tailing that was received and used in the test work contains 43% iron, with a P80 of 163µm. When this material underwent magnetic separation without regrinding a concentrate containing 61% iron was achieved at an iron recovery of 93%. At this grade, the silica, sulfur, alumina and phosphorus levels are 9.3%, 0.34%, 2.01 and 0.02% respectively.

The data was also plotted to produce a grade versus size curve as shown in Figure 8.
Figure 8 - Grade vs. Size Curve

Figure 8 shows that a 70% iron grade is achievable with a grind size P80 of 50 μm. Further grinding will improve the grade marginally to a 71% iron grade. As in any grinding circuit, the benefits of increased concentrate grade needs to be weighed up against the extra grinding power that is required. In cases where the grind size is quite fine, the increase in grade requires an exponential increase in grinding energy and could well require another grinding unit to achieve.

Figures 9, 10, 11 and 12 also show the particle size grade relationships of the impurity elements, silica, sulfur, alumina and phosphorus. As expected, grinding finer and separating the magnetics from the non-magnetics, will result in less of the non-ferrous impurities reporting to the magnetics as they are liberated by finer grinding. At a grind size P80 of 50 μm, silica, sulfur, alumina and phosphorus levels have dropped to 1.6%, 0.045%0.035 and 0.003% respectively.
Figure 10 – Size vs. Sulfur Grade

Figure 11 – Size vs. Alumina Grade

Figure 12 – Size vs. Phosphorus Grade
Table 2 – Assay per Size Fraction – Magnetic Concentrate

In relation to other magnetite ores, there have been several M4 IsaMill™ tests conducted on other deposits, although not as many as undertaken on base metal deposits. The magnetite samples come from deposits in the Yilgarn Craton in Western Australia, and Central Europe.

The M4 IsaMill™ on the magnetite material to date, indicate that it is in the middle of the range in terms of the power required to reduce the size of the sample. Magnetite, copper and PGM (Platinum Group Minerals) signature plots using an M4 IsaMill™ have been plotted on Figure 13, as well as in Table 3, for the power required to grind samples from an F80 of 45 µm, to a product size P80 of 25µm. This range was chosen as the majority of magnetite samples that have been tested have been in this range. However, as observed in Table 2, the IsaMill™ could treat much coarser magnetite feed sizes than these.

Figure 13 – Signature Plots of Different Materials

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<table>
<thead>
<tr>
<th>Material</th>
<th>Circuit</th>
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<td>Isa Cu Con</td>
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</table>

Table 3 – Power Requirement for Grinding from 45μm to 25μm

The information from Table 3 and Figure 13 indicates that the 4 magnetite samples required 8 to 16 kWh/t to achieve the grind required. The hardest of the magnetite material was the EHM tail sample. In comparison to the EHM copper concentrate sample, EHM Cu Con, which is floated off before the tailing, and contains mainly chalcopyrite and other sulfides with low levels of magnetite, required less than half the energy compared to energy required to do a similar size reduction for the EHM Tails sample.

The PGM Material came from several sites across the Bushveld in South Africa. The power required for the grinding duty ranged from 26 to 33 kWh/T, and was significantly higher than the energy to grind the magnetite sample, which ranged in energy from 8 to 16 kWh/t. Both PGM Site 2 and PGM Site 3 have M10,000 IsaMills operating, treating the PGM minerals which are associated with the hard chromite host rock. The other site in the list where a M10,000 IsaMill™ operates is the Kumtor mine in Kyrgyzstan, which treats a pyrite concentrate at 21.5 kWh/t.

NEW INSTALLATIONS USING IsaMill™ in COARSE DUTIES

Phu Kham Project

The Phu Kham deposit is located approximately 100km north of the Laos capital Vientiane. It is owned by Pan Australian, an Australian listed mining company. The Phu Kham deposit hosts two distinct styles of mineralization: an oxide gold cap and beneath this transitional/primary copper-gold. The Phu Kham oxide gold cap is the principal deposit for the Phu Bia heap leach gold mine, the first phase of the development of the Phu Kham deposit, which entered into production in 2005. Feed to the concentrator will consist of 12MT on average, with planned annual output from this mine being over 200,000 tonnes of concentrate (grading 25% copper), containing 50,000 tonnes copper, 40,000 ounces gold and 400,000 ounces silver, (on average). The concentrate will be exported for further treatment and refining by custom smelters in the Asia Pacific region.

Process technology employed for Phu Kham Copper-Gold is conventional comminution at the head of the circuit, followed by flotation to produce a copper-precious metal concentrate, (Pan Australian, 2006)

Rougher concentrate will be treated through a M10,000 IsaMill™, powered by a 2.6MW motor, treating approximately 168 tph and reducing the feed size from a F80 of 106μm to a P80 of 38μm, before further flotation. The grinding media for the operation will be MT1. The Phu Kham Copper-Gold operation is planned for start-up in mid 2008.
Prominent Hill Project

Oxiana Limited owns 100% of the Prominent Hill copper-gold project located 650 kilometers north west of Adelaide, and 130 kilometers north west of BHP Billiton’s Olympic Dam in South Australia.

The ore body consisting of a copper gold breccia, will be mined via an open pit. The ore will be treated through a conventional grinding and flotation processing plant, with a designed capacity of 8MTPa. The initial planned concentrate production will be on average 187,000 tpa, peaking at 230,000 tpa in 2009, with average concentrate grades of 45% copper, 19g/t gold, 57g/t silver. The high grade concentrate will be sold to smelters in Australia and Asia. (Oxiana, 2007)

One M10,000 IsaMill™, powered by a 3.0MW motor, has been selected to treat the rougher concentrate. It will treat approximately 138 tph, reducing the feed size from a F80 of 125μm to a P80 of 24μm for further flotation. The planned commissioning of the mill will be mid to late 2008. The grinding media for the operation will be MTI.

Anglo Platinum Installations

In 2007, Anglo Platinum had ordered five, M10,000 IsaMills™, following the successful installation and operation of the first M10,000 IsaMill™ at their Western Limb Tailings Retreatment Project in 2003. The typical duty of these installations is from 75-100 μm feed size down to 53 μm product size.

Installations have been successfully commissioned at Potgietersrust Platinum mine (C-Section), Potgietersrust A and B Sections (2 mills), and two more at the Rustenburg Waterval UG2 operation.

The Potgietersrust Platinum mine (C-Section) mill is designed to operate with a 3MW motor and use MT1 media, treating scats from A and B section primary milling circuits, with the ore having a Bond Work Index (BWi) over 30 kWh/t. Figure 14 illustrates the simplified C section flowsheet using an IsaMill™.

![Figure 14: Simplified PPL C Section Flowsheet with a M10,000 IsaMill™](image-url)

McArthur River Mines (MRM)

McArthur River Mine (MRM) is a zinc/lead mine in the Northern Territory, Australia, and is operated by Xstrata Zinc. It was commissioned in 1995, and was where the IsaMill™ was developed to regrind streams down to a P80 of 7 μm, which was the enabling technology that allowed the mine to be developed. Initially there were 4 x M3000, 1.1MW IsaMills™ in the regrind duty. This has since been expanded to 6 IsaMills™ with a combined installed power of 6.7MW. The current plant flowsheet is shown in Figure 15.

![Figure 15: McArthur River Flowsheet](image)

MRM have a need to increase milling capacity from 230tph to 305tph to account for decreased head grades as the operation shifts from underground to open cut. Flotation feed size is also to be reduced from the current P80 of 75μm, back to the original size of a P80 of 45 μm. At the same time there was a desire to reduce downtime and reduce operating cost by eliminating the Tower Mill from the circuit, hence MRM have been keen to explore the effectiveness of a M10,000 IsaMill™ in the primary grinding circuit.

Testwork has been undertaken using a M4 and M20 IsaMills™ treating SAG mill cyclone underflow (Anderson 2006, Burford 2006), with further testwork using a M20 IsaMill™ in late 2006 designed to overcome the presence of scats which caused problems in the earlier testwork. Figure 12 displays the
flowsheet that was used for this testwork that incorporated a magnetic separator to remove steel seats in the cyclone underflow. Feed to the mill was also screened at 1mm.

![Diagram showing flowsheet with a magnetic separator and an IsaMill™ testwork circuit.]

**Figure 16: Site Testwork at MRM Using M20 IsaMill™ with a Magnetic Separator on Feed**

The M20 IsaMill™ was able to treat material from a feed sizing of 300μm, down to a product sizing of 20 to 25 μm, (finer than the 40 μm target), in a single pass. The data was able to permit a size energy relationship to be established, as shown in Figure 13, compared with the current Tower Mill operation in that circuit.

![Graph showing size versus net energy comparison for IsaMill™ and Tower Mill.]

**Figure 17: Size versus Net Energy Comparison for IsaMill™ and Tower Mill**

Using the energy data from the M20 IsaMill™ testwork, and the current energy use for the tower mill in the primary circuit, it has been conservatively estimated that the IsaMill™ could produce a P80 size of 45 μm to 50 μm, while the tower mill could produce a P80 of 100 μm using a similar amount of energy.
However the flowsheet was not the most efficient use of both of the technologies, especially the IsaMill™, as IsaMill™ operation was hampered by the need to operate the mill to control the coarse particles, rather than achieve target grind size. Simulations followed the testwork with different circuit configurations, which lead to a much better circuit design based on the main advantage of the IsaMill™, open circuit operation.

MRM is planning to use IsaMills™ to treat the flotation feed instead of cyclone underflow, with the eventual circuit designed to eliminate cyclones and the tower mill. The eventual circuit configuration is displayed in Figure 18.

![Diagram](image)

**Figure 18 – IsaMill™ in MRM Grinding Circuit**

This benefit of this circuit is that it allows the IsaMills™ to concentrate on the particles that it is ideally suited to, with an estimated F80 of 300μm, to produce the F80 of 45μm.

To date, testwork is being undertaken with one M3000, 1.1MW IsaMill™ continuously operating in this circuit with different grinding media, wear materials and feed sizing being trialed by the mill. Later in 2008, two M10,000 IsaMills™ powered with 3MW motors will be installed in the primary grinding circuit, with the rated capacity of the primary grinding circuit to be increased from 1.9 Mtpa to 2.5 Mtpa.

It is expected that the introduction of open circuit IsaMilling™ on the flotation feed using inert media will improve metallurgy as has been observed at other lead/zinc circuits (Pease et al 2005, Young et al 1997), such as improving the selectivity of fines, improving flotation rates, and reducing circulating loads and flotation reagents.
CONCLUSION

IsaMill™ technology is becoming a realistic alternative in coarse grinding applications since the development of ceramic media as well as the large scale M10,000 IsaMill™. These developments have enabled coarse grinding to be undertaken at a number of mineral processing sites, where energy efficient grinding is required.

The application of IsaMill™ technology to new applications, such as grinding in magnetite concentrators, offers magnetite operators an exciting alternative to conventional ball mill and tower mill technologies. Magnetite ore was found to be amenable to grinding with IsaMills™, in much the same manner as base metal ore is, with several of the magnetite samples being softer than base metal ore types that are treated using IsaMill™ technology.

With the pressure being applied to all industries today for improved sustainability and the potential cost implications of carbon emission in the future, the need for increased energy efficiency in grinding is as important as ever.

ACKNOWLEDGEMENT

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AMMENDMENT

Figure 6 - Size versus Specific Energy – M4 IsaMill™ and Tower Mill

Equations updated with more data post publishing of original paper.
GRINDING COMPARISON TEST OF IsaMills™ WITH TOWER MILLS USING MAGNETITE

INTRODUCTION
IsaMills™ have been in fine grinding applications in base metal circuits since 1994. The IsaMill™ has been enabling technology, that has allowed these applications to be developed, due to its high energy efficiency and intense grinding action. Over the years the IsaMill™ has evolved from its initial design as an “ultra fine grinding mill” to its use in more traditional regrind duties following the increase in capacity of the units and the development of a ceramic media by Magotteaux – Keramax® MT1™. These developments have allowed the mill to treat coarser feed sizes at high energy efficiency compared to traditional grinding technologies. There is a range of models now available from the 500kW M1000, to the largest stirred mill currently available, a 3MW M10,000.

Today the mills undertake coarse grinding duties in PGM, copper/pyrite, and zinc/lead applications. While most of the work to date has been with base metals, there is no reason why the benefits of base metal operation cannot be transferred to other minerals.

RESULTS OF TESTWORK ON MAGNETITE RECENTLY CONDUCTED (SEE OVERLEAF)

TESTWORK
The current testwork compares the operation of a lab scale M4 IsaMill™, with a lab scale Tower Mill using Ernest Henry Mine tailings. The tailings are the result of the flotation process, where chalcopyrite has been recovered from the breccia host rock, which is comprised of strongly altered and replaced felsic volcanic fragments in a matrix assemblage of predominantly magnetite, chalcopyrite and carbonate minerals.

The M4 IsaMill™ is a 4 litre mill, containing 6 to 9 discs, operating with 3.5mm ceramic Keramax® - MT1™ media. The Tower Mill is 40L capacity and operates using 12mm steel media.
In each test, 20 to 21kg of sample is made into a slurry of 50% solids and pumped through each mill for a number of runs. The power used for each run is recorded, and a small sample of the discharge is taken to permit laser sizing. This procedure is carried out through each mill for a minimum of 12 times, or until there is no significant reduction in the sizing, i.e. the mill cannot reduce the sample any further. A log-log, P80 size versus the specific energy to obtain that size graph, is then drawn (signature plot).

The test was repeated for each mill twice.

DISCUSSION
The results of the two test for each grinding technology shows a good level of reproducibility.

Key findings were:

- The IsaMill™ was able to produce material down to 13µm from a feed size of 113µm. However the Tower Mill failed to produce material less than 31µm. The reason for the difference between the mills was the grinding media used in the IsaMill™ testwork was a lot smaller. However media selection was based upon what realistically a full scale plant can operate with. A full scale IsaMill™ can be supplied and operated with ceramic media from 1 to 3.5mm, however Tower Mills can only realistically operate with media 12mm and upwards, which means that they cannot achieve the grind sizes that a full size IsaMill™ can achieve.

- The IsaMill™ used less power to produce a product 65µm or smaller compared to the Tower Mill. This means for this magnetite feed, an operator targeting a product of 40µm with an IsaMill™ will need approximately 14 to 18Kwh/T, while a Tower Mill will need a lot more energy between 26 to 27Kwh/T. The implications for this is smaller installed power for an IsaMill™ circuit and the resultant lower installation cost due to it operating in open circuit without the need for cyclones, as well as having a smaller footprint. It may also result in less grinding units compared with a Tower Mill circuit when large power requirements are needed.

Testwork is ongoing to determine the energy efficiency of the IsaMill™ on other magnetite deposits.

For further information please contact:
Lindsay Clark, Mineral Processing Business Manager – Xstrata Technology
Telephone +61 7 3833 8500; Email L.Clark@xstratatech.com.au; Web www.isamill.com
Case Study

Coarse IsaMilling at McArthur River

Joe Pease

General Manager Xstrata Technology

September 2007
At this conference in 2006, I concluded with this slide, a picture of a possible energy efficient circuit of the future. A circuit without any tumbling mills at all. I said that perhaps in a decade we would see the first plant built without any tumbling mills. Or with only a small conventional grinding stage between HPGR and IsaMills.

Well, we aren’t there yet, but we are a lot closer than we thought we were even a year ago.

This is a story about how you can find answers in unexpected places.
And how solving one problem may end up solving other, even bigger, problems.
The mother of invention

Figure 2: Different Grain Size of Broken Hill and McArthur River Ores
(both photos at the same magnification)

The story starts at McArthur River. This huge lead zinc deposit was discovered in 1955, but remained undeveloped for 40 years. There was simply no technology that could economically treat the ultra fine grained minerals.

As always, necessity was the mother of invention. Keeping the orebody, and survival in our other fine grained orebodies, simply required that we find a more efficient way to grind fine, or else go out of business. We had to make a step change in fine grinding. Conventional grinding, in ball or tower mills, was uneconomic for 3 reasons:

- it used too much energy
- the media cost was too high
- the large amount of steel media consumed harmed subsequent flotation.
Mt Isa’s head of research, Bill Johnson, knew that the answer to the problem didn’t lie in the minerals industry – we had looked there for decades. So he asked the question, who else has to grind fine?

In fact, there was good established technology for fine grinding of high value manufactured products – like printer inks, pharmaceuticals, paint pigments, chocolate.

The pioneers in the field, and still the leaders, were Netzsch of Germany. We chose to work with them.

As an aside, my wife says that if Bill had had only listened to her, he would have looked to chocolate to find answers long before he actually did.
Crossover to Minerals

Redesign and scale up to:

- Much bigger scale
- Continuous
- Low cost Media

So the concept was in manufacturing, but it had to be modified to minerals.

The manufacturing applications were for very high value products – like printer ink and chocolate. They are much higher value than zinc, which at the time was trading at about 40 c/lb. So these applications were small mills, often batch, and used very high cost media – ensuring no contamination of the product was much more important than the cost of grinding media.

To be economic to treat large tonnages of low value streams, we had to make the mills much bigger, operate continuously and robustly, and be able to use low cost media.
The end result was the IsaMill. This was the enabling technology for McArthur River, and then for the George Fisher and Black Star orebodies at Mt Isa.

This shows the 6 mills at McArthur River. They are 1 MW drive – 6 times bigger than the previous biggest Netzsch mills.

They operate continuously, and for the first 7 years their grinding media was ore gravel screened from the SAG mill discharge. That is, this was fully autogenous grinding to 7 microns!

The previously untreatable orebody became economic, and achieves over 80% recovery into a 55% Zn+ Pb bulk concentrate.
3 MW, M10,000 IsaMill

So by looking in an unexpected area, a problem was solved, and a new technology was developed.

For many in the audience there is nothing new in this story, it has been told before. But now there is a new, unexpected twist to the story, and again McArthur River is at the forefront.

They are currently installing IsaMills to grind SAG mill discharge.
In early 2008, two 3 MW M10,000 IsaMills will be installed to treat SAG mill product before flotation.

So we are a lot closer to a circuit without tumbling mills than most of us expected.
Firstly, what is the IsaMill?
The IsaMill was developed at Mt Isa in the early 1990’s as an economical grinding solution for fine grained ore bodies.

Pictured here is a 3.0 MW M10,000 (litre) IsaMill. This is the largest IsaMill currently available with up to a 3.0MW motor. Motor, gearbox, bearings, mill

The motor turns a horizontal shaft within the mill, the shell remains stationary.
The IsaMill

- **High intensity**
  - Small footprint – 3MW in 10 m³ grinding volume
- **High Power efficiency**
  - Small media
- **Inert grinding**
  - Clean surfaces, no steel effects
- **Internal Classifier**
  - Low cost media
  - Sharp size without cyclones
- **Horizontal**

These are the characteristics of the IsaMill that make it quite different from conventional grinding.

The high intensity means a small mill footprint and installation size. The IsaMill has a power intensity of 300 kW per cubic metre, versus about 20 for a ball mill or Tower Mill - that is about 10 times higher. This means a significantly different installation, even for things like media – the entire first charge of media for a 3 MW mill is only 7 cubic metres.

The high power efficiency simply comes from the small media, as discussed in my previous presentation. In fact, the high power intensity and high power efficiency are linked in practice. A slow stirred ball or Tower Mill using 2 mm media would also be efficient for fine grinding. But the low power intensity would make the installation uneconomic – a huge installation, with prohibitively high consumption of prohibitively high cost media. The high power intensity in the IsaMill comes from the high stirring speed – about 20 m/second. This means that the fine media can do a lot of work in a small volume.

The internal classifier really is the great innovation of the IsaMill. How do you keep fine media, eg 1-2 mm, inside a mill, while allowing product to exit, and without using screens. We knew any solution that used screens just wasn’t going to work at a large scale – the screens would block and peg, and would be an operating and maintenance nightmare. Further, having screens would dictate that you could only use high quality, high cost media, because low quality media would break and block the screen. The answer was the product separator, basically a centrifuge to keep media in the mill without screens. It was the breakthrough that allowed us to use virtually free media – ore gravel at McArthur River, discard smelter slag at Mt Isa.

A happy consequence of this is that the mill doesn’t need to be closed circuited with cyclones. This is hard to get used to – we have always known that grinding mills need cyclones and a circulating load. But the IsaMill doesn’t.

The final point is obvious – this is a horizontal mill. But this simple point has big implications.
Our partners, Netzsch, make both vertical and horizontal stirred mills for manufacturing. They told us that both configurations have advantages. But there was one overwhelming factor in favour of the horizontal configuration – scale up to larger sizes. This is because of start-up torque. Netzsch advised that they simply wouldn’t build a stirred mill over 400 kW, because the mechanical design would be dominated by the start-up torque on the bottom stirrer after a shutdown – the bottom stirrer has to be able to remobilise the settled load. For big mills this would dominate the design of stirrer, shaft, crane etc. In contrast, in a horizontal mill, all 8 stirrers are available. This helps explain why the IsaMill has scaled up 10 fold from 300 kw to 3 MW in a decade, already almost 3 times bigger than the biggest installation Tower Mills have achieved in over 50 years.

The picture shows the 8 grinding discs (black) and the product separator (yellow) at the mill discharge.
In operation the mill is 70-80% filled with media, which is stirred at high speed – up to the stirrer tip speed of about 20 m/s. New feed has to pass through 8 different grinding chambers before it gets to see the product separator, or centrifuge, at the end of the mill. At this point media is returned to the grinding discs, and fine solids and water is discharged. The mill operates full and pressurised, with average residence time in the mill is 30-60 seconds. Once again, this is fundamentally different from conventional grinding. It also explains why the mill doesn’t need cyclones to make a sharp product size – there is little time for overgrinding, but particles have to pass through 8 grinding chambers in series before leaving the mill.

The best way to describe this is to have a look at a small mill in operation.
This video shows why the IsaMill gets a sharp size distribution. The action of the product separator compresses the media between the 8 grinding discs. Feed has to pass through effectively 8 grinding chambers in series before it reaches the exit. When dye is introduced to the mill, its slow movement down the mill demonstrates the almost plug-flow path for new feed.
Developing the mill from grinding printer ink and chocolate to grinding zinc was the big step.

Compared with that, the move from fine grinding minerals to coarser grinds was relatively simple.

We were always fascinated by the potential to take the energy efficiency, small footprint, and inert grinding to mainstream applications. But there were two things we needed to do. We needed a bigger mill. We now have this with the 3 MW M10,000 mill.

Secondly, we needed better media. The ability to use low quality, free media like slag, ore gravel, or local sand, is remarkable, but it restricts the use of the mill to fine grinding. These grinding medias have relatively low density, and small natural grain size. This restricts the size of the ore particle they can break, even at the high stirring speed in an IsaMill. To be practical for mainstream coarse grinding, the media needed to be consistent quality, and high enough density and size to break the biggest particles in mill feed. And still be lower operating cost than using steel balls in conventional mills.

This was achieved by the development of MT1 Keramax ceramic by Magotteaux.
Energy per Media Particle

\[ E \propto d^3 \cdot v^2 \cdot SG \]

- \( d \) = media diameter
- \( v \) = media velocity
- \( SG \) = media density

I promise that this will be the only formula in this presentation. But it is important to the McArthur River coarse grinding story.

Media particles need enough energy to break the largest particles in the mill feed, as quickly as they are entering the mill. If they don’t, the coarse particles will build up in the mill and reduce the grinding efficiency – we are all familiar with critical size fractions from Autogenous and SAG milling.

In a ball mill, some steel balls are picked up and dropped the diameter of the mill – plenty of energy. In fact, probably too much energy, so some is wasted, but the coarsest particles will be broken.

Similarly in the IsaMill, we need enough energy in enough collisions to break the feed. We can either increase the SG of the media, increase its diameter, or increase its speed. The much high speeds in the IsaMill, ~20 m/sec, explains why small media can grind coarse particles. We don’t want to increase speed any more, so we increased the media diameter and density. The MT1 ceramic has density of 3.7, compared with 2.4 for sand. Maggoteaux currently produce it up to 3.5 mm diameter, enough for grinding duties up to 300 micron P80 feed.
3 MW IsaMill, ceramic media

Mill product direct to circuit

Feed in

Screw feeder for media addition
McArthur River Expansion Project

- From underground to open cut mining
- Increase plant tonnage 33%
- Reduce Flotation feed sizing: 75 to 45 microns
  - Back to original design
- Continue 7 micron regrind of cleaner feed

McArthur River is now going through its next phase of development. It is moving from underground mining to Open Cut mining. Feed tonnage will be increased by 33%, from 230 t/h to 305 t/h (2.4 Mt/y). The project will also reduce flotation feed sizing from 75 microns back to the original design of 45 microns (increases in feed tonnage over the years caused the coarsening flotation feed).

Of course, to make concentrate grade and recovery, all minerals have to be ground to 7 microns before cleaning.
The original circuit was a SAG mill in closed circuit with cyclone and double deck screen to produce a P80 45 microns product for flotation. Media for the IsaMills was from the second screen deck (0.8-1.8mm). The IsaMills grind rougher con from 50 to 7 microns. Later, the IsaMill media was changed to sand. This decoupled SAG mill operation from IsaMill operation, allowing both to improve efficiency and throughput.

Feed tonnage has increased since start up. A 935 kW Tower Mill was installed to grind part of the cyclone underflow to partly compensate the tonnage increase. This is not the ideal duty for a Tower Mill, but it was chosen simply because a second hand unit was available – these were the days of 37 cent zinc! Even so, flotation feed has coarsened to about 70 microns at the higher tonnage. Increasing the efficiency of the IsaMills helped compensate for the coarser feed, by maintaining a 7 micron grind of cleaner feed.

The ore from the open cut is lower grade and a bit more complex than the original orebodies treated. So McArthur needed to both increase tonnage, and return to design flotation feed size, while maintaining regrind size at 7 microns. IsaMills offered an efficient way to do this, while also bringing some of the advantages of inert grinding to rougher flotation.

McArthur River undertook a test program to evaluate the IsaMill for the coarse grinding duty.
The first trial was to test the mill on a similar stream to the Tower Mill, ie treat a portion of SAG mill underflow.

Initial testwork was done initially in the 4 litre M4 laboratory mill. This indicated that 3.5 mm ceramic media would grind the stream at higher efficiency and with a sharper size distribution than the Tower Mill. As a result, the product could be sent straight to flotation rather than back to the SAG mill cyclone.

We then did in-plant pilot work using the M20 (20 litre) mill, treating SAG cyclone underflow. The target was to produce approximately 45 micron product from the IsaMill.
Pilot IsaMill Coarse Grinding at MRM

- 3.5 mm media could grind the feed
- More efficient than Tower Mill
- Sharper size distribution than Tower Mill

But:
- Hampered by steel scats and +1 mm chips
- Had to reduce throughput, increase power to prevent “critical mass”
  - Meaning grind was finer than target (25 micron)

The site pilot testwork confirmed the laboratory work that the IsaMill with 3.5 mm ceramic media could grind the cyclone underflow with the +1mm screened out. Efficiency and size distribution was better than the Tower Mill. But the operation was hampered by build up of steel scats and coarse (up to 1mm) particles in this stream.

The steel scats could be removed with a magnetic separator. However the breakage rate of coarse particles was still too low. To prevent build up of coarse particles, the mill power had to be kept high, but throughput reduced. This meant an increase in kwh/t, and the feed was ground finer than necessary, to 25 microns rather than 45 micron target. Theoretically the grind was efficient, but we didn’t need to grind that fine.
As expected, the IsaMill was more efficient than the Tower Mill – for the same energy input (7kwh/t) it would produce a 55-60 micron product versus a 100 micron product from the Tower Mill. However this comparison is not particularly valid, as neither mill is suited to this grinding application. The IsaMill had to be operated to control coarse particles rather than achieve target grind size. So it became clear that this was not the right location for the IsaMills.

So our first attempt told us that the IsaMill would certainly grind, but it needed to be in a better position in the circuit. We were trying to do too many stages of grinding in a single mill.
Dealing with +1mm fraction

- **Increase breakage energy**
  - \( E \approx d^3 \cdot v^2 \cdot SG \)
  - 7mm media has 8 times the breakage energy of 3.5mm

- **Find a better location**
  - Screened SAG discharge, not cyclone underflow

- **2006-7: SAG simulation, pilot testing, full scale testing in existing 1 MW IsaMill**

- **Project approval early 2007**

The breakage rate of coarse particles could be improved by returning to our formula – by increasing speed, media SG, or media size. In fact, the full scale mills operate with higher tip speed than the M20 unit, so should have performed better. But the biggest impact would be from bigger media – doubling the top size of media will mean an 8 times increase in maximum breakage energy available.

Currently the coarsest size of MT1 commercially available is 3.5 mm. This will increase in future. But the main point is that it became obvious we were trying to grind the wrong stream. A much better solution was to take the IsaMill out of the SAG mill circulating load. Let the SAG mill grind SAG mill feed, the IsaMill can grind the final product of that circuit.

During 2006 and 2007, McArthur River surveyed and simulated a different SAG mill configuration. This included closing the SAG mill with a 1 mm screen. Screen undersize is nominally -1mm, with a P80 of 300 micron feed. This stream would be ground in open circuit IsaMills, then sent to flotation. One of the existing 1MW has been used for full scale evaluation of IsaMilling and different medias, including ceramic and coarse sand. As a result, McArthur River are proceeding with a full scale project to install IsaMills to grind flotation feed.
This is the new circuit being installed at MacArthur River. Two 3 MW M10,000 IsaMills will be installed in early 2008. Screen undersize will be sent direct to open circuit IsaMills, which will produce a P80 45 micron flotation feed. The Tower Mill and cyclones will be removed from the circuit.
MRM Expansion Project

- Add 2 IsaMills to treat 305 t/h at 40% solids \textit{(open circuit)}
- Remove Tower Mill and cyclones
- Increase feed 1.9 Mt/y to 2.5 Mt/y (33%)
- Reduce flotation feed P80 75 to 45 microns
- Potential downstream benefits
  - Inert grinding before roughing
  - Reduced regrinding energy (finer float feed)

The mills will be fed at 40% solids.

The finer flotation feed sizing should improve roughing performance, and will slightly reduce regrinding energy need. Primary roughing may also benefit from the inert grinding – experience at other installations is improved flotation rates and selectivity for fines, and reduced reagent needs. These gains can only be evaluated at full scale, where the full impact on water chemistry and circulating loads is evident. So any gains from inert grinding are an “upside”, not part of the justification.
McArthur River have taken the coarse grinding applications of IsaMilling to the next level. Anglo Platinum have already committed to a major program of IsaMilling. This is the first installation at their PPL concentrator, which grinds scavenger feed in open circuit from around 100 microns to around 50 microns. This mill was commissioned in December 2006.
Coarse Grinding at Anglo Platinum

- Mainstream Coarse IsaMilling to Increase liberation and recovery
- Concentrate Regrind to increase grade
- Currently installing 64 MW of IsaMills (mainly coarse)

The success of the Anglo Platinum installations has led them to undertake a major installation program for IsaMills. Over the next 18 months they will be installing 64 MW of IsaMilling capacity. Much of it will be grinding mainstream feeds like the PPL installation – to increase liberation and recovery in roughing and/or scavenging. Other mills will be installed to increase concentrate grade, to increase smelter throughput, and reduce smelting energy and cost.

By the end of 2008, there will be 100 MW of operating IsaMill capacity. The technology has quickly moved from fine grinding to coarse – over 70% of the installations by the end of 2008 will be in conventional regrinding or mainstream grinding. (product sizes 25 to 60 microns).
Mineral Processing without Tumbling Mills?

So I return to my opening slide. Most people think of the IsaMill as a fine grinding mill. No more. A little over a decade ago it moved from grinding ink to grinding zinc. It has quickly made the adjustment to coarser grinding.

With the work at McArthur River, we are much closer to the hypothetical circuit without tumbling mills than any of us imagined even a year ago.

We aren’t there yet, but we are a lot closer than we were even in 2006. The work at McArthur River for the first time will take a SAG mill discharge directly to IsaMills rather than to conventional ball mills. So further advances in increasing the feed size of IsaMills (eg with coarser or higher density media), and/or reducing the product size of HPGRs could enable this circuit. Reducing the product size of HPGRs could be achieved in several ways, including by using two or more HPGR stages in series, with each stage in either closed circuit, or in a simpler open circuit configuration.

Perhaps the HPGR discharge sizing and IsaMill feed sizing will not converge quite close enough for them to meet as we have shown in the above simplified diagram. Perhaps the P80’s will be mismatched, or maybe just the coarse fraction from HPGR will be too coarse for the IsaMill. In this case, the simplified circuit above could be modified to include a conventional grinding and/or classification stage between HPGR and the IsaMill. In this way, the benefits of energy efficiency and low steel consumption of HPGR and IsaMills would be maintained, with the addition of a relatively small conventional grinding and classification step (eg ball mill or Tower Mill with or without a classification stage).
Increasing the Energy Efficiency of Processing

Joe Pease

General Manager Xstrata Technology

August 2007
Can we double the energy efficiency of grinding?

- Diminishing returns on research?
- Radical new technologies?
- 10 years or longer?

Mineral comminution consumes about 3% of the world’s energy.

So improving the efficiency of grinding has long been an important objective for operators and researchers. Much progress has been made. Entire research institutions like the JKMRC have dedicated decades of study with almost the sole objective of improving comminution efficiency.

Does all this work mean that the big gains have been made, and that we are on diminishing returns from our research?

Has conventional technology reached its limit? Is equipment just getting bigger, but not more efficient?

Do we need radical new technologies like microwave grinding to make a step change in grinding efficiency?

A new collaborative research program, AMSRI has a $20 M budget and the ambitious target to double grinding efficiency in 10 years. Is this a realistic target, given the enormous amount of work already done?

No, we don’t think 10 years is a reasonable target. We think it can be done today!
The tools are already available

- **Laboratory grinding and flotation characterisation**
- **Quantitative mineralogy**
- **Basic thermodynamics**

In fact, processing efficiency can be significantly increased by using tools already developed by our previous research.

We don’t need radical new technologies, we need to combine the tools we already have in the right way.

These tools are:
- Laboratory grinding and laboratory flotation tests
- Quantitative mineralogy
- Basic thermodynamics of smelting and refining

To grind the right mineral in the right place to the right size.

In many cases this approach can double the energy efficiency of **processing**. Note that I refer to processing efficiency, not grinding efficiency. Grinding is not an end in itself. It is just one step in the processing chain. It cannot be considered in isolation of the downstream process. This is why I have included smelting thermodynamics in the list of tools. It is no point reducing grinding energy if this increases smelting energy.
We already have the tools to double the efficiency of **processing**:

- *grind the right streams,*
- *in the right place,*
- *to the right size*

In fact, processing efficiency can be significantly increased by using tools already developed by our previous research.

We don’t need radical new technologies like microwave grinding, we need to combine the tools we already have in the right way.

These tools are:
- Laboratory grinding and laboratory flotation tests
- Quantitative mineralogy
- Basic thermodynamics of smelting and refining

To grind the right mineral in the right place to the right size.

In many cases this approach can double the energy efficiency of **processing**.

Note that I refer to processing efficiency, not grinding efficiency. Grinding is not an end in itself. It is just one step in the processing chain. It cannot be considered in isolation of the downstream process. This is why I have included smelting thermodynamics in the list of tools. Sometimes we should increase grinding energy, if we can reduce smelting energy by more.
So of course, the real question is about the energy efficiency of the whole process, not just the grinding step.

It is no point reducing grinding energy if it increases downstream energy by more.

This is the same logic we all understand about mine-to-mill. One of the advances in the last decade has been to recognise the importance of the mine-to-mill interaction. Most of us will be familiar with the classic example of the mining engineer who reduced his blasting cost by 20 c/tonne, only to reduce SAG mill throughput by 10% and increase milling cost by $1/tonne. Sometimes the right answer is to spend a bit more and blast finer in the mine, and save much more in the mill.

Well of course, the same applies between the mill and the smelter. Yet how many of us check the thermodynamics of smelting, and see what we can do in the mill to reduce smelting energy? Maybe that’s not a fair question, because we don’t have simple tools to help us do it. Have you ever seen a grinding model that includes some simple smelting thermodynamics to help you minimise processing energy? Which research group is working on it? We have research proposals to develop more and more sophisticated grinding simulations, to seek the last little bit of grinding efficiency. But where do they consider smelting energy?

I think this is a huge oversight, and a huge opportunity — we are just like the mining engineer who tried to reduce his blasting cost with no consideration of what it would do to grinding efficiency.
This is a flowsheet of a common circuit. How can we use existing tools to reduce the energy need.

Often the primary grind can be coarsened. We don’t need complete liberation for roughing, we just need to liberate enough gangue away from the valuable mineral to get high rougher recovery. We can then achieve the liberation we want for cleaning by regrinding a smaller tonnage of rougher concentrate (or a part of it).

This is the biggest single impact on grinding energy – why grind big tonnages of silica finer than you need to? Grinding too fine up front also causes us to put in more roughing flotation capacity than we really need. Coarsening primary grind also has other advantages for tailings storage and reactivity.
Where is the lowest energy place to remove this impurity?

- Mine
- Preconcentration
- Roughing
- Cleaning
- Smelting/refining

Grind it out.

Or add Silica and limestone, and melt it out at 1350°C

Improving energy efficiency means asking a simple question: where is the cheapest, lowest energy place to remove this impurity?

- in the mine (by dilution control) – requiring no energy
- in a preconcentration stage (removing some gangue while it is still coarse)
- in roughing – ie by grinding the whole mill feed
- in cleaning – ie by grinding a much smaller tonnage of rougher concentrate
- in the smelter – by melting it out?

Of course, cleaner feed tonnage is much lower than rougher feed tonnage. So ideally, so it is obviously better to grind the lower tonnage of rougher concentrate rather than rougher feed. So long as you grind rougher feed just enough to get recovery into the rougher concentrate.

The photo of the composite shown is taken from a paper on Nickel metallurgy. The gray is pentlandite, the white is MgO. Excellent work was done in the flotation circuit to allow recovery of this low grade composite. Even so, this particle causes a problem for the smelter – MgO causes highly viscous slag. The smelter has to add Silica and lime fluxes, then heat the whole lot up to 1350°C to remove the MgO. This consumes a lot of energy – over 10 times the energy it would take to grind the MgO out. The high temperatures also make smelting difficult and reduce refractory campaign life. So to improve energy efficiency, the concentrator should regrind this particle – so long as the mill can still recover the pentlandite.
How can we use existing tools better?

- Simplify use of quantitative mineralogy
- Make better use of regrinding
- Look at the mill-smelter interface

I think there are three areas where we can improve our use of the existing tools to improve energy efficiency, which I will discuss in the next few slides.
Quantitative Mineralogy ...

.... See the forest for the trees

**What do we want to achieve in each part of the circuit, eg**

- Primary grind and rough for recovery
  - Look for gangue liberation, not mineral liberation
- Regrind and clean for grade
  - Look for mineral liberation and entrainment

**Distinguish symptoms from causes**

- Eg fines in tails may mean ...
  - ... you are grinding too fine
  - OR you are grinding too coarse!

Because quantitative mineralogy is so powerful, sometimes we overcomplicate it. It can generate so much data that we get lost in detail, and can’t see the forest for the trees. We need to keep it simple, and apply the 80/20 rule. Avoid the temptation to find out everything about every mineral in every liberation class in every stream. Ask a few simple questions about what you are trying to achieve in each stage of the process. And look for root causes rather than symptoms. The mineralogy can only be interpreted with a good understanding of how the circuit is operated.

Two simple questions are:

1. In roughing, how coarse can I grind and still be able to get good recovery to rougher concentrate. You don’t need to worry about grade yet, that is, you don’t need to worry about the mineral liberation, but rather the gangue liberation – is enough gangue liberated from the mineral to allow it to float into rougher concentrate.

2. In cleaning, the question changes to how fine you need to grind to make a high grade concentrate. Now mineral liberation is important. So is entrainment of liberated gangue. If there are coarse composites in concentrate that could be easily liberated, these are “low hanging fruit”. Even if you are already at concentrate grade, by removing the easy impurities you make room to “pull” the rest of the circuit harder.

This point is an example of distinguishing root causes from symptoms. You will always find fine liberated values in cleaner tailings and rougher tailings. Perhaps this means you are grinding too fine – “overgrinding” your valuable minerals. But, paradoxically, more often it means you aren’t grinding fine enough! The next slide shows why.
Problem or Symptom?

This is a photomicrograph of zinc concentrate. Look at the composite particle of sphalerite and pyrite at the bottom. It has a fair amount of exposed sphalerite surface, so it floats well, and is too valuable to throw out. Yet it is also too low grade to accept into concentrate - it probably assays about 20% Zn, it needs to be about 50% to be payable.

Because it floats relatively easily, the operator only has two choices:

• If he accepts it in concentrate, it lowers con grade. He has strict limits on the concentrate grade and Fe content, so he now has to pull the roughers and cleaners more gently to keep the concentrate grade. He can’t chase coarse composites, or slow floating fines in tailings, because he doesn’t have the processing power to recover them at a high grade.

• Alternatively, he tries to reject this composite, in this case by starving collector and running high pH. But these are the very conditions that will cause low recovery of the fines. Look at the fine liberated sphalerite particles near the composite – they have less sphalerite surface than the composite, so depressing the composite will probably depress them as well.

So no matter what the operator does, if he has to make a concentrate grade, then dealing with this composite will cause him to lose fines (and coarse composites). Instead, regrounds the composite removes the pyrite which increases concentrate grade without depressant. The higher concentrate grade leaves him room to pull his roughers and cleaners harder. Regrinding will improve his fines recovery.

This shows the importance of using quantitative mineralogy like forensics, not like a computer sledgehammer. Don’t get hung up on there being fines in tails. Look at the composites in concentrates, if you fix this root cause many of the symptoms disappear.

Before you can ask the right questions of your mineralogy, you have to understand how the circuit is operated.
Grind the right minerals in the right place

- **Primary grind as coarse as you can**
- **Regrind as fine as you need**
- **Regrind cleaner feed, not cleaner tail**
  - No step change from grinding a minor stream

In summary, use quantitative mineralogy to ask these simple questions. In roughing, how coarse can I grind and still get recovery to rougher concentrate.

In cleaning, how many composites can I remove from the concentrate by regrinding finer.

We think the result is that many circuits could grind the main stream coarser, but need to regrind the cleaner feed finer than they do. They can make a higher grade at the same recovery, with less energy.

So many of our circuits regrind cleaner tailings, not cleaner feed. But if regrinding is a powerful tool, why save it until after you have made most of your concentrate? If composites like the one in the previous photo go straight to concentrate, the damage is already done. Cleaner tailing has a small portion of the metal, so you can’t get a step change by regrinding it.
Energy Efficient circuit

Compare this flowsheet to the one on page 6. Often the primary grind can be coarsened. We don’t need complete liberation for roughing, we just need enough liberation of valuable from gangue to allow high recovery in roughing. We can then achieve the liberation we want for cleaning by regrinding a smaller tonnage of rougher concentrate (or a part of it).

This is the biggest single impact on grinding energy – why grind big tonnages of silica fine when you don’t need to. Grinding too fine up front also causes us to put in more roughing flotation capacity than we really need.

The regrind mill shown here is an IsaMill. These mills grind in open circuit, producing a sharp sizer distribution than conventional grinding and hydrocyclones.

Coarsening primary grind also has other advantages for tailings storage and reactivity.
Do we do enough Regrinding?

- **Fines are hard to float?**
  - Not with clean surfaces and good size distributions

- **Regrinding hurts chemistry?**
  - Not with inert grinding

- **Regrinding is inefficient?**
  - Not any more

We think many concentrators don’t regrind fine enough. When they do regrind, they regrind small difficult streams like cleaner tails or scavenger concentrate, not the main cleaner feed.

This probably reflects some perceptions about regrinding and subsequent flotation. Indeed, these concerns are valid, and most plant operators have experienced them. Conventional regrinding to fine sizes is inefficient, it uses lots of media, and the steel media hurts flotation. Therefore we avoid regrinding cleaner feed so we don’t risk harming our main stream; we save regrinding to cleaner tailing, a relatively small and difficult stream.

But this has all changed with the advent of high intensity inert grinding. The IsaMill is much more efficient than ball or Tower Mills for fine grinding, it uses inert media that produces clean surfaces for flotation, and produces sharp size distributions.

Regrinding used to be one step forward for liberation, one step back from chemistry. Not any more.
Fines float very well when:

<table>
<thead>
<tr>
<th>They have clean surfaces</th>
</tr>
</thead>
<tbody>
<tr>
<td>• Inert regrinding</td>
</tr>
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<table>
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<tr>
<th>You add reagents to suit</th>
</tr>
</thead>
<tbody>
<tr>
<td>• Don’t try to depress composites at the same time</td>
</tr>
<tr>
<td>• Avoid circulating loads</td>
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</table>

96% of recovered particles at MRM are less than 2.5 microns

Traditionally we thought that fines don’t float well, that we need special reagents and flotation cells, and still get poor performance.

We now know this isn’t true. Numerous operations get excellent fines flotation, with conventional flotation cells and standard reagents. We just have to give them clean surfaces, and make sure they aren’t competing with composites – that is, we need to be able to add enough collector to float the fines, without having to worry about floating unwanted composites at the same time (as we discussed before).

When we do this, we find that fines down to 1 micron float just as well as intermediate size particles, and almost as quickly. In the Mt Isa lead and zinc circuit, cleaning recovery after IsaMilling of cleaner feed is above 95% for all size fractions above 1 micron, until dropping above 37 microns due to composites.

For another example, consider Macarthur River, an operation whose entire concentrate is finer than 7 microns, floated in conventional cells. Let’s look at what this means in terms of individual particles. I know we only get paid for weight, not number of particles. But flotation is about individual particles connecting with bubbles.

So 96 percent of the successful particle-bubble interactions at McArthur River occur for particles less than 2.5 microns.

So fines do float, and very well indeed.
A lot has been written about grinding efficiency, and the relative efficiencies of ball mills and Tower Mills and high speed stirred mills. In fact, it is fairly simply. Overwhelmingly the biggest effect is media size. If you run a ball mill slowly with 12 mm media, it will be as efficient as a Tower Mill. Its just that we don’t normally run ball mills with 12 mm media. Similarly, a Tower Mill with 5 mm or smaller media will be more efficient again. But practically, no one ever runs a Tower Mill with 5 mm media – the media cost and wear would be prohibitive. As an aside, if you get testwork done, make sure it is done with the same media size you will use in the plant. We have heard of occasions when a Tower Mill energy need has been predicted from a test with 6 mm media, whereas the plant will use 12 mm or coarser, which will require significantly more energy.

The other important parameter in this slide is power intensity – it is no point being power efficient (like a Tower Mill with 5 mm media) if the power intensity is so low that you need a huge installation to get the necessary installed power.

So let me summarise – a Tower Mill with small media will be just as efficient as a high speed stirred mill with small media (the optimum media size for each device will be different). But the low power intensity of Tower Mills means you need a much bigger installation. These points are made by Nesset et al in their paper “Assessing the Performance and Efficiency of Fine Grinding Technology”, CIM 38th Annual Canadian Mineral Processing Operators Conference, Ottawa 2006.
This is a picture of a 3 MW IsaMill installation, showing the features that make it such a different technology.

Firstly, note the scale from the people at the discharge – this is a 3 MW grinding mill. Consider the size of a 3 MW ball mill or three, 1 MW Tower Mills.

Note the simple installation. The mill is pressured, with feed in, outlet pipe straight to process (no closed circuit cyclones) and simple media system under the floor. Media is added to the mill feed pump by the screw feeder. Note the low footprint, low head height, and simple crane needs.
To give an idea of difference in scale – the IsaMill is not an incremental change, it is a step change – an order of magnitude smaller than conventional technology. Until now, the Tower Mill was considered the most efficient, modern technology for regrinding. After 55 years, the biggest model is still only 1.1 MW (shown).

The IsaMill next to it has been scaled to keep the people about the same size. This is a 3 MW IsaMill – three times the power of the much bigger Tower Mill. This is what we mean by power intensity. Think of the difference in installation cost if you need 3 MW of Tower Mills (with associated media handling, cranes, and closed circuit cyclones).

This is why we say that the IsaMill changes the way we can think about circuit design. It is much easier to imagine putting 3 MW of open circuit IsaMill in the middle of a flotation circuit.
A new tool for plant design

Regrinding that
• is energy efficient
• improves mineral surfaces and chemistry
• has a small footprint and head height
• doesn’t need cyclones
• can be installed throughout the circuit

The tools to grind the right minerals in the right place

The IsaMill is a step change in regrinding technology. It was developed to change the economics and efficiency of ultrafine grinding. But it has now brought the same advantages to conventional grinding and regrinding, where it will have its biggest effect on the industry.

This fundamentally changes the way we can think about the design of circuits. A lot of regrinding power can be installed in a small space and in a simple installation that can be located where it is needed throughout the plant.
Using existing tools better ...

- Simplify use of quantitative mineralogy
- Make better use of regrinding
- Look at the mill-smelter interface
  - A bigger problem than Mine-to-Mill?

The third reason we don’t take achieve the energy efficiency we could is because of the mill to smelter interface.

Some time ago we recognised the mine to mill interface as a problem. We knew it was a problem because we could hear the miners and metallurgists squabbling on sites. Since then a lot of work has been done to better co-ordinate mining and milling to improve overall efficiency – eg is it better to spend a bit more on mine blasting (drilling and chemical energy) to save more energy in the SAG mill.

But how often do we hear the same questions being asked about concentrating and smelting? Or do we just accept that the target grade is 25% Cu because that suits the smelter contract. I think the mill-to-smelter interface is a more serious problem than mine to mill, because there is virtually no dialogue between us. You can’t even hear us fighting, because we don’t talk. The smelter may be owned by another company, and based overseas. My commercial people talk to your commercial people, and they argue about supply/demand balance. Can anyone imagine that they are negotiating about the most energy efficient way to remove impurities.
How do we choose concentrate grade?

By considering the optimum energy/cost trade off between grinding and smelting?

- Mineralogy plus smelting thermodynamics to minimise energy?

OR:

By getting maximum recovery at “target” grade?

- With target grade set by commercial negotiation?

It is logical for the concentrator to maximise return based on the smelter contract. They will work out what grade concentrate gives them the best return given contract terms and transport costs. Then they will maximise recovery and minimise costs at this grade. They will set KPI’s, and be rewarded to minimise cost and maximise recovery – which will “lock in” that target grade. What incentive do they have to question the grade? Why would they volunteer to increase costs, maybe even drop some recovery, in order to make a higher concentrate grade, if this just helps another company’s smelter. Of course, they won’t.

The concentrator can only respond to the contract. And can we expect the contract negotiators to be optimising the trade off between mineralogy, smelting thermodynamics and energy efficiency? Of course we can’t.

This is why I think mill to smelter is a bigger problem than mine to mill. There is simply no dialogue. The metallurgy and calculations are fairly easy, but we don’t do it. We don’t even recognise it as a problem. Of all the research projects you have heard of to improve grinding efficiency, how many include the smelter?
Let me give an example. I return to the Nickel composite we looked at earlier. This photo was taken from a paper by G. Senior and S. Thomas, “Development of a New Flowsheet for the Flotation of Low Grade Nickel Ore”, International Journal of Mineral Processing 78(2005), Elsevier.

It is a photo of a low grade composite floated into scavenger concentrate at Mt Keith circuit. Excellent work was done in the concentrator to be able to recover this low grade composite, without requiring a finer primary grind.

So the nickel recovery has been increased, and the process is more efficient. It could be even more efficient if we could then eliminate some of the additional MgO from the final concentrate, while still recovering the pentlandite. This is a liberation issue.

We were originally asked by what was then WMC to look at the potential for IsaSmelt for the Kalgoolie Nickel smelter, to handle the high MgO content. But the more we looked at smelting, the more we suspected that it would be easier to deal with the MgO in the concentrator.

So we worked with the Leinster operation to see whether we could improve the Nickel-MgO separation with inert regrinding in an IsaMill.
Grind or Smelt?
- Nickel – MgO separation at Leinster

The answer is yes. This graph is from D. Seaman et al, “Process Design of a Regrind Facility at the Leinster Nickel Operation to Improve Concentrator Recovery”, AUSIMM, 9th Mill Operators Conference, Freemantle 2007.

The graph shows the results of pilot work conducted at Leinster. The coarse cleaner scavenger concentrate is shown with and without IsaMilling. The regrinding is effective in significantly increasing the Nickel concentrate grade at the same recovery. Most of the increase comes from removing MgO, as shown by the next graph.
Big impact MgO - and smelting energy

In this case, MgO can be removed from the concentrate without any loss of recovery with relatively little regrinding energy. This requires much less energy than smelting it out from concentrate.
The next example is from Anglo Platinum. This mill is their first installation, at the Western Limb Tailings Retreatment Plant. In many ways this was typical of operations that needed a step change in technology – old mill tailings that had been in dams for up to 100 years, fine grained and with altered surfaces. The ability of the IsaMill to efficiently liberate, and provide clean new surfaces to flotation, was enabling technology for this operation.

This was the development site for the big M10,000 IsaMill – this unit is 2.6 MW and operates with sand media, the same mill model is now rated at 3 MW with ceramic media.

This is a fine grinding mill, operating on a difficult stream. But Anglo Platinum realised that the advantages of IsaMilling didn’t stop with these difficult fine grained applications. They saw that the ability to improve liberation, and at the same time improve flotation selectivity, had much wider application to their mainstream applications.
The Big Picture at Anglo

- **Installing 64 MW of IsaMill**
- **Coarse grinding scavenger feed**
  - F80 100 microns, P80 50 microns
  - Liberate gangue to increase PGM recovery
- **Regrinding rougher concentrate**
  - P80 25-35 micron
  - Increase concentrate grade

- **Higher recovery and higher grade**
  - Significantly increased smelter capacity
  - Significant reduction in processing energy

The end result is that Anglo Platinum have embarked on a major program to fit IsaMills to their concentrators, in a combination of mainstream grinding and regrinding duties. This program will significantly improve both recovery and concentrate grade. The increase in concentrate grade means a significant increase in capacity through existing smelting, and significant reduction in overall processing energy.

The mainstream mills will grind from typically F80 100 microns to P80 45 to 60 microns, and will treat rougher tail/scavenger feed. Each mill will process around 300 t/h of solids.

The regrind mills will treat rougher concentrate, and will regrind to P80 15-20 microns before cleaning.

It is clear that the IsaMill is now a mainstream grinding and regrinding mill, not a niche fine grinding mill.
We can build plants which:
- Use half the power
- use a fraction of the footprint and height
- Deliver better results – grade and recovery

We already have the tools:
- Laboratory grinding and flotation
- Quantitative mineralogy
- Smelting thermodynamics
- Efficient regrinding technology

To summarise, we think we already have all the tools we need to double the energy efficiency of processing. We just need to use them together in the right way.
We have the tools to get it right

This is an illustration of what we can currently do. We can use laboratory tests to develop a range of grade/recovery curves at different grinding and regrinding sizes.

We can use quantitative mineralogy to help us design and interpret these tests, and interpolate between them.

We can predict the grinding energy needed to create the different grade/recovery curves.

We can use simple smelting thermodynamics to calculate the different smelting energy needed for the different concentrates.

This gives us enough information to find the most efficient operating point; the lowest energy trade off between grinding and smelting for that ore.
Challenge to Researchers:

Develop a standard “energy index” to rank ores
- Develop grade/recovery curves at different grinds
- Calculate grinding energy and smelting energy for different options
- Use for new ores, and benchmark existing plants

Reduce the energy of processing, not just energy of grinding

So in closing, I issue a challenge to researchers. If we have the tools, why don’t we use them in the right way to optimise energy efficiency.

I think the previous concepts can be captured in a relatively simple “energy index” to describe an ore. For a long time the industry has used the Bond Work Index to summarise grinding energy needs. Even though there are much more sophisticated grinding models, we still find a single index to be useful.

So why don’t we develop a modern energy index, like the Bond Index, but to include grinding and regrinding energy, the grade recovery curve, and basis smelting thermodynamics. A simple measure that for each ore predicts how much energy will be needed to produce metal (or final product). And a simple technique to indicate the lowest energy route for that ore, the best trade off between grinding and smelting/refining.

The same index and technique could be used to benchmark our existing operations – how do we perform against the estimate of the most efficient way to treat the ore.

Our focus is to improve the energy efficiency of processing, not just the energy efficiency of grinding.
IsaMill™ Technology Used in Efficient Grinding Circuits

B.D. Burford¹ and L.W. Clark²

High intensity stirred milling is now an industry accepted method to efficiently grind fine and coarse particles. In particular, the IsaMill™, which was invented for, and transformed the fine grinding industry, is now being included in many new comminution circuits in coarser applications. While comminution has always been regarded as important from a processing perspective, the pressure being applied by environmental concerns on all large scale power users, now make highly energy efficient processes more important than ever.

The advantages that were developed in fine grinding in the early IsaMill™ installations have been carried over into coarse grinding applications. These advantages include a simple grinding circuit that operates in open circuit with a small footprint, the ability to offer sharp product size classification, as well as the use of inert media in a high energy intensive environment.

This paper will examine the use of IsaMill™ technology in fine grinding (P80 below 15 micron), and examine the use of the technology in conventional grinding applications (P80 20 - 150 µm). Recent installations will be examined, including fine and coarse grinding applications, as well as the recent test work that was undertaken using an IsaMill™ in a primary grinding circuit, and the resulting circuit proposal for this site.

While comminution has been relatively unchanged for the last century, the need to install energy efficient technology will promote further growth in IsaMill™ installations, and result in one of the biggest challenges to traditional comminution design.

¹. Senior Process Engineer, Xstrata Technology, L4, 307 Queen Street, Brisbane 4000, Qld, Australia
². Business Manager – Mineral Processing, Xstrata Technology, L4, 307 Queen Street, Brisbane 4000, Qld, Australia
INTRODUCTION and BACKGROUND

The development of IsaMill™ technology was driven by the metallurgical requirements of Lead/Zinc deposits at Mount Isa in Queensland and McArthur River in the Northern Territory, both of which were controlled by Mount Isa Mines Limited (now Xstrata).

The McArthur River deposit was discovered in 1955 but, despite the efforts of numerous mining companies, an economic method for treatment of the fine grained deposit to produce saleable Pb/Zn concentrates was not achieved in 25 years of investigations (Enderle et al, 1997; Pease et al, 2006). In 1989 it was determined that regrinding down to 80% passing 7 µm was necessary to achieve sufficient non sulphide gangue liberation to allow the production of a bulk concentrate. Figure 1 is a comparison of the relative grain sizes of McArthur River and Broken Hill ore and illustrates the complexity of the mineralogy at McArthur River.

**Figure 1: Comparison of McArthur River and Broken Hill Ore Grain Size (Grey square is 40 µm)**

In the case of Mount Isa, there was a gradual decrease in plant metallurgical performance from the mid 1980’s as a result of decreasing liberation size and increased amounts of refractory pyrite in the ore. Concentrate grade targets were reduced to maintain zinc recovery, however plant performance continued to deteriorate to such an extent that by the early 1990’s the zinc recovery had decreased from 70% to 50% (Young et al, 1997; Pease et al, 2005; Pease et al, 2006).

Significant work was conducted at Mt Isa on projects investigating finer regrinding using conventional ball and tower mill technology however the power consumptions necessary to achieve the required fine liberation sizes made them uneconomic. Further, the high rate of steel media consumption contaminated the mineral surfaces with iron, resulting in poor flotation response post regrinding.

A real need had arisen for a technology that could grind to ultrafine sizes in metallurgical operations economically and without serious contamination of mineral surfaces and pulp chemistry. However in 1990, there was no generally accepted technology for regrinding economically to such sizes in base metals. So testwork was undertaken at the time into high speed horizontal stirred mill technology, which was used in pigment and other industries. It was shown that such mills could grind down to the ultrafine sizes required for mineral liberation.

Arising from these findings, a program of major mechanical modification of horizontal stirred mill technology was undertaken between Mount Isa Mines Limited and NETZSCH-Feinmahltechnik GmbH (Enderle et al, 1997), the manufacturer of stirred milling technology used for other industries.
After many prototypes of increasing capacities, the first full scale model was developed and installed at the Mount Isa Mines’ Lead Zinc Concentrator in 1994. The mill, the M3000 IsaMILL™, was quickly installed in other circuits at this concentrator, and was installed in the McArthur River Concentrator in 1995 (Johnson et al, 1998). Later, in 1999, it was commercialised and sold outside of the Xstrata group.

Since commercialisation of the IsaMill™, there are now over 50MW of installed IsaMills™ operating around the world, treating materials including copper/gold, lead/zinc and platinum. While the early installations treated only ultrafine sizes, the current mill installations are treating courser sized materials, once the domain of tower and ball mills. The need for energy efficient grinding circuits will only result in more IsaMill™ circuits being applied in the future.

**IsaMill™ OPERATION**

**Grinding Mechanism**

The IsaMill™ is a horizontally stirred mill consisting of a series of 8 discs rotating around a shaft driven through a motor and gearbox. The discs operate at tip speeds of 21-23m/s resulting in high energy intensities of up to 300kW/m3. Figure 2 illustrates the layout of the IsaMill™.

![IsaMill™ Layout](image)

The mill is filled with a suitable grinding media and the area between each disc is essentially an individual grinding chamber. As a result the mill is effectively 8 grinding chambers in series. The media is set in motion by the action of the grinding discs which radially accelerate the media towards the shell. Between the discs, where the media is not as subject to the high outwards acceleration of the disc face, the media is forced back in towards the shaft – creating a circulation of media between each set of discs. Minerals are ground as a result of the agitated media, the predominant mechanism being attrition grinding. The mechanism is best illustrated in Figure 3.
As a result of having 8 chambers in series, short circuiting of mill feed to the discharge is virtually impossible. There is a very high probability of media-particle collision as a result of the high energy intensity and the 8 chambers in series.

**Media**

The key to the efficiency of the IsaMill™ is the ability to use fine media. While tower mills are typically limited to 10-12mm fresh media sizing, the IsaMill™ can use media as small as 1mm. This results in significantly more surface area per unit volume of media in the IsaMill™ than in a Tower Mill – a 2mm charge has 90 times more particles per unit volume compared to 12mm media. As a result, there is a significantly higher chance of media-particle collision, particularly at fine sizes.

The IsaMill™ is able to use a range of media types. Typically, low cost, locally available media such as sand or smelter slag have been used, which provide good grinding performance at acceptable energy efficiency. However, the need for improved energy efficiency at many installations has resulted in the use of high quality, high density ceramics, designed specifically for stirred milling applications, such as Magotteaux’s Keramax MT1.

**Media Retention**

Grinding media is retained in the mill without the need for screens, which is why IsaMills™ can use fine media. At the end of the mill is a patented product separator consisting of a rotor and displacement body (refer to Figure 3). The close distance between the last disc and the rotor disc centrifuges any coarse particles towards the outside of the mill. Ground product flows into the rotor area where it is essentially pumped back towards the feed end of the mill. This pumping action retains the media in the mill. The balance of the product (equivalent to the feed flowrate) exits the mill through the displacement body. This unique mechanism means that screens or cyclones are not required to retain media in the IsaMill™ which can therefore be operated in open circuit without cyclones, which simplifies the circuit, and reduces capital.
Energy Intensity

The high tip speed of the IsaMill™ results in a high energy intensive environment. Energy intensity of the IsaMill™ is significantly higher than any other commercially available grinding equipment as illustrated in Table 1. Combining the energy intensity and the high grinding efficiency leads to a compact mill, able to be fitted into existing plants where floor space is limited. On the other hand, Tower Mills require a settlement zone at the top to separate the media from the slurry – this limits the agitation speed to a tip speed of 3m/s (compared to IsaMill™ at 21-23m/s) and therefore limits the energy intensity, while ball mills can only have a relatively low amount of ball loading before media empties from the mill.

**Table 1: Comparative Energy Intensity of Grinding Technologies**

<table>
<thead>
<tr>
<th></th>
<th>Installed Power (kW)</th>
<th>Mill Volume (m³)</th>
<th>Power Intensity (kW/m³)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Autogenous Mill</td>
<td>6400</td>
<td>353</td>
<td>18</td>
</tr>
<tr>
<td>Ball Mill</td>
<td>2600</td>
<td>126</td>
<td>21</td>
</tr>
<tr>
<td>Regrind Mill</td>
<td>740</td>
<td>39</td>
<td>19</td>
</tr>
<tr>
<td>Tower Mill</td>
<td>1000</td>
<td>12</td>
<td>42</td>
</tr>
<tr>
<td>IsaMill™</td>
<td>3000</td>
<td>10</td>
<td>300</td>
</tr>
</tbody>
</table>

Product Size Distribution

In open circuit operation, the IsaMill™ is able to produce a sharp product size distribution. Typically the ratio of the P98 to the P80 is around 2.5. This is a direct result of the effect of 8 chambers in series preventing short circuiting and the classification action of the product separator. The ability to operate the mill in open circuit greatly simplifies the operating and maintenance strategies of the circuit. Figure 4 illustrates a typical IsaMill™ product distribution in open circuit at varying power inputs – note the steepness of the curve and the lack of ultra fines that would be expected from a Tower Mill distribution.

**Figure 4: Typical IsaMill™ Product Distribution in Open Circuit (South American pyritic gold concentrate)**
Inert Grinding

After the initial comminution stages of crushing and/or SAG/AG milling, the following grinding stages is usually carried out using steel charged ball or tower mills. The impact of grinding using steel media can offset any benefits gained by improved liberation, particularly as the target size decreases below 25 µm. Grinding in a steel environment results in the precipitation of metal and iron hydroxides on to the surface of ground particles. These conditions affect flotability, flotation selectivity and lead to higher reagent consumptions to overcome the surface coatings and regain recovery (Trahar, 1984; Pease et al, 2006). The benefits of inert grinding at several locations have been well reported (Pease et al, 2006, 2005, 2004; Young et al, 1997; Grano et al, 1994).

While the negative impacts of steel grinding will be greatest at fine sizings due to the large surface areas and high media consumptions involved, inert grinding has also been shown to produce benefits at coarser sizings (Grano et al, 1994; Greet et al, 2004; Pease et al, 2006). For a long time, chrome media has been offered to, and investigated by, ball and tower mill operators as a means of improving pulp flotation chemistry by reducing the amount of iron released into the grinding pulp and contaminating freshly ground surfaces. Greet and Steiner, 2004, analysed the surface of galena ground in three different environments for the presence of iron. It is clear from Table 2 that while grinding in a high chrome environment reduced the surface iron composition from 16.6% to 10.2%, grinding in a ceramic environment reduced the detectable surface iron to less than 0.1% - a significant improvement over both media types.

<table>
<thead>
<tr>
<th>Media Type</th>
<th>Surface Atomic Composition (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>O</td>
</tr>
<tr>
<td>Mild Steel</td>
<td>53.1</td>
</tr>
<tr>
<td>High Chrome</td>
<td>50.0</td>
</tr>
<tr>
<td>Ceramic</td>
<td>33.6</td>
</tr>
</tbody>
</table>

IsaMill™ RECENT DEVELOPMENTS

Until recently, inert grinding to coarser sizes (+25µm) was generally impractical in traditional milling equipment due to the high capital investment required to fill the mill. However two recent developments in IsaMill™ technology have allowed coarse grinding in an inert environment to become a reality, namely MT1 ceramic media and the scale up of the IsaMill™ to the M10,000 model.

MT1 Ceramic Media

The development of high intensity stirred mills, such as the IsaMill™, increased energy efficiency by the use of higher agitation speeds and finer media, compared to those associated with tower mills or ball mills. (Gao et al, 1999, 2002). The IsaMill™ has been able to use low cost, cheap media, such as sand, discarded slag, river pebbles, scats etc to obtain very fine grind sizes. However while the mills have been run with low cost media, the low quality media limits the energy efficiency of the mill. Quality issues such as particle shape, grain size, specific gravity etc all constrain the mills energy efficiency as well as the size of feed that can be milled. It is this latter limitation that has
constrained the mill to grinding in fine size ranges. Previously available ceramics were of variable quality, high cost, high wearing and low SG which generally made their use uneconomic.

A recent initiative between Magotteaux International and Xstrata Technology brought about the development of a ceramic specifically designed for use in the IsaMill™ to address the limitations of current media. The product, Keramax MT1, was designed to address all the key characteristics for the ideal media, listed by Lichter and Davey, 2002, as:

- Hardness
- High sphericity
- High roundness
- Mechanical Integrity
- SG
- Definite initial charge PSD and top up size
- Chemical composition

Keramax MT1 is a high density, high quality alumina ceramic media. It has been shown to have a consistent hardness when sectioned with no air bubbles or structural defects (Curry et al, 2005). It has been specifically designed for grinding applications, with relatively high SG and low wear characteristics. Keramax MT1 properties are listed in Table 3.

Table 3: Keramax MT1 Properties

<table>
<thead>
<tr>
<th>Keramax MT1</th>
<th>Properties</th>
</tr>
</thead>
<tbody>
<tr>
<td>Composition</td>
<td>79% Al2O3</td>
</tr>
<tr>
<td></td>
<td>6.5% SiO2</td>
</tr>
<tr>
<td></td>
<td>14.0% ZrO2</td>
</tr>
<tr>
<td>Hardness</td>
<td>1300-1400 HV</td>
</tr>
<tr>
<td>Fracture Toughness</td>
<td>5-6</td>
</tr>
<tr>
<td>Specific Gravity</td>
<td>3.7</td>
</tr>
<tr>
<td>Bulk Density</td>
<td>2.3-2.4</td>
</tr>
</tbody>
</table>

The surface of the media is smooth and ‘pearl-like’ to touch. These surface properties mean that the energy loss in grinding due to friction is minimized, improving the efficiency of grinding. At the same time there is negligible contamination of ground material with deleterious ions.

The true benefits of this material can be seen when it is compared to other media in a lab environment. This was undertaken by Xstrata Technology and Magotteaux separately in 2004 on gold bearing concentrate from the Eastern Gold fields region of WA, using a M4 IsaMill™ (Curry et al, 2005). The results are summarized in Table 4.
Table 4: Media Type vs Energy, Consumption and Net Power for 250t/hr Treatment Rate

<table>
<thead>
<tr>
<th>Media Type</th>
<th>Consumption Rate (g/kWhr)</th>
<th>Specific Energy (kWhr/t)</th>
<th>Consumption Rate (Kg/t)</th>
<th>Net Power (MW) for 250 t/hr example</th>
</tr>
</thead>
<tbody>
<tr>
<td>MT1 (-4 +3mm)</td>
<td>15</td>
<td>7.6</td>
<td>0.11</td>
<td>1.9</td>
</tr>
<tr>
<td>Alumina 1 (-4 +3mm)</td>
<td>128</td>
<td>13.1</td>
<td>1.68</td>
<td>3.3</td>
</tr>
<tr>
<td>Alumina 2 (-4 +3mm)</td>
<td>295</td>
<td>12.4</td>
<td>3.66</td>
<td>3.1</td>
</tr>
<tr>
<td>Australian River Pebble (-4 +3mm)</td>
<td>200</td>
<td>27.9</td>
<td>5.58</td>
<td>7.0</td>
</tr>
<tr>
<td>Australian Silica Sand (-6 +3mm)</td>
<td>781</td>
<td>11.2</td>
<td>8.77</td>
<td>2.8</td>
</tr>
<tr>
<td>Ni Slag (-4 +1mm)</td>
<td>1305</td>
<td>17.8</td>
<td>23.23</td>
<td>4.4</td>
</tr>
</tbody>
</table>

Keramax MT1 was found to be more efficient than the other ceramics or sands tested. This has big implications to grinding circuit design, as it requires less installed power for an application on MT1 compared a to much larger installed power that would be required if sand was used. This is highlighted in the final column of Table 4, where for a hypothetical throughput of 250tph, 1.9MW is required to do the grinding duty using MT1, compared to 7.0MW on River Pebble (-4mm to +3mm).

Keramax MT1 has all the attributes of inert media that has been in use in IsaMills™ over the last decade, but has the toughness and high SG to make it a more efficient and harder wearing media than slags and sand, while at the same time being able to handle coarser feed size distributions.

**M10,000 IsaMill™**

In late 2002, Xstrata Technology, in collaboration with Netzsch and Anglo Platinum, developed the M10,000 IsaMill™; the worlds largest stirred mill. The mill was developed for Anglo’s Western Limb Tailings Retreatment Project near Rustenburg in South Africa. The project involved the retreatment of dormant platinum containing tailings, which represented a major economic resource. Fine grinding testwork had established that the PGMs could be recovered from the tailings. Pilot scale work resulted in the circuit needing to produce a primary grind, no less than 80% passing 75 µm, and a rougher concentrate regrind of no less than 85% passing 25 µm, which would enable good recovery of PGM’s. The tonnage required for the regrind circuit was 53tph (max 65tph), to produce a P90 of 25 µm, requiring 35kWh/t. This duty could not be performed in a single M3000 IsaMill™. To maximize the economics of the project, a larger mill was required leading to the development of the 2.6MW, M10,000 IsaMill™, (Curry et al 2005).

Results from the Western Limb Tailings Retreatment Project have confirmed the accuracy of the scale up from the lab scale M4 IsaMill™ and pilot plant IsaMill™ testwork to the full scale unit, Table 5.
Table 5 Laboratory versus Full Scale Grinding Efficiency

<table>
<thead>
<tr>
<th>IsaMill™ Model</th>
<th>Installed Power (kW)</th>
<th>Chamber Volume</th>
<th>Specific Energy (kWh/t)</th>
<th>Pulp Solids</th>
<th>P98 (µm)</th>
<th>P98 (µm)</th>
</tr>
</thead>
<tbody>
<tr>
<td>M4</td>
<td>4</td>
<td>3.5</td>
<td>37</td>
<td>39</td>
<td>47.5</td>
<td>16.0</td>
</tr>
<tr>
<td>M10,000</td>
<td>2,600</td>
<td>10,000</td>
<td>37</td>
<td>42</td>
<td>42.5</td>
<td>16.5</td>
</tr>
</tbody>
</table>

Since Anglo’s Western Limb Tailings Retreatment Project, there have been 9 other M10,000 IsaMills™ either installed or designed at the time of writing. They are either installed with 2.6MW or 3.0MW motors, and all use ceramic media. The rapid growth of IsaMill™ technology is shown in Figure 5.

Figure 5 – Time versus Cumulative Power for IsaMill™ Installations

Today, large scale coarse grinding using inert media is both practical and economic, due to the development of large scale IsaMill™ technology and MT1 ceramic media. The use of the IsaMill™ with inert media in coarse grinding applications not only has significantly improved energy efficiency, but also has improved downstream separation processes due to the production of minerals with clean, iron free surfaces.

The following Case Studies describe IsaMill™ installations being applied to coarse grinding duties.
CASE STUDY 1 - KUMTOR M10,000 INSTALLATION

The M10,000 IsaMill™ installation at the Kumtor gold mine in the Kyrgyz Republic, is one of the largest gold mines operating in Central Asia, producing over 500,000 ounces of gold per annum. It is based 4000m above sea level in the Tien Shan Mountains, and is owned by the Kumtor Operating Company, a fully-owned subsidiary of the Canadian company Centerra Gold Inc.

The Kumtor installation treats a hard refractory orebody of finely disseminated gold linked with significant quantities of pyrite. The flowsheet is conventional crushing, SAG and ball mill circuits, before the ball mill cyclone overflow is floated. The tailings are thickened and then undergo CIL before passing to the tails disposal circuit. The concentrate is also thickened, and is then reground in a conventional ball mill, prior to CIL, before stripping, electro winning and refining to doré. The regrind ball mill, operates in closed circuit with 25 mm diameter balls and very high recirculation loads (∼600 %), to produce a product with a P80 of ~20 µm. Liberation studies undertaken by Kumtor established the gold recovery benefits of grinding finer, with an optimal P80 sizing of 10 µm selected for the design.

In late 2005, a 2.6MW M10,000 IsaMill™ was commissioned at Kumtor, treating the regrind ball mill discharge to a product sizing of 80% passing ~10 µm. Figure 6 illustrates the before and after flowsheets. This installation represented the first commercial application of the Magotteaux MT1 media. The decision to use the MT1 media was based on several key factors:

- Low wear rates resulted in less media required to be transported to the remote site.
- High SG ceramic media translates into more efficient power usage compared to sand, enabling 1 x M10,000 IsaMill™ operating in open circuit to do the work equivalent to 2 x M3000 IsaMills™ operating on sand, reducing capital cost and simplifying the circuit.
- Cyanide consumption would be reduced

Soon after commissioning the IsaMill™ was treating on average 72 tph, compared to the design tonnage of 65 tph. The current power draw was approximately 1950 KW, which equates to 23 KWhr/t specific grinding energy. Shortly after commissioning, metallurgical results had seen a drop in the gold tailings grade by 30%, (which represents an extra 20,000 ounces of gold per annum being recovered).

Figure 6: Pre and Post IsaMill Circuits at Kumtor
Replacement of Kumtor Ball Mill with IsaMill™ During Ball Mill Maintenance

While the Kumtor IsaMill™ is designed as an ultra fine grinding application, it was noted by Kumtor management that the IsaMill™ would be required from time to time to operate without the ball mill, when the ball mill required maintenance, Figure 7. This would allow the Kumtor operation to continue without any lost production from the maintenance. In this operation the mill would be expected to produce a coarser product, although it was hoped to match the ball mill discharge.

Figure 7: Circuit Configuration during Ball Mill Maintenance

Testwork was conducted on samples of rougher concentrate, which predicted that a coarse charge of ceramic media could produce a similar output to the ball mill, i.e. a P80 of 20 to 25um.

In March 2006, shortly after the IsaMill™ was installed for fine grind duty, the ball mill was relined for a 4 week campaign, and the IsaMill™ feed was coarsened from 20 um to 150um. However during the relining of the ball mill, a much finer ball charge was actually used in the IsaMill™ than used during the testwork, while the IsaMill™ was also operated with a non optimized feed density and rotor selection. The 4 week timeframe did not permit these key variables to be optimized, which, along with the fine media, impacted on the power draw, which was only 1885 KW (average) throughout the trial.

The key findings of the trial were

- IsaMill™ has reduced the feed from F80 of 130µm to 150µm to P80 of 60µm to 65µm
- Size reduction has been undertaken by finer media than used in the testwork
- Media consumption has been good at 17.5g/KWhr
- IsaMill™ operated continuously during the 4 week ball mill shutdown enabling full scale production to be achieved, with no breakdowns
- No significant wear issues were noticed during the trial, with the same mill liner and disc used throughout and after the trial
- Coarser media, optimal density and the use of a high flow rotor could have improved the power draw and further reduced the discharge sizings
CASE STUDY 2 - NEW INSTALLATIONS USING IsaMill™ in COARSE DUTIES

Phu Kham Project

The Phu Kham deposit is located approximately 100km north of the Laos capital Vientiane. It is owned by Pan Australian, an Australian listed mining company. The Phu Kham deposit hosts two distinct styles of mineralisation: an oxide gold cap and beneath this transitional/primary copper-gold. The Phu Kham oxide gold cap is the principal deposit for the Phu Bia heap leach gold mine, the first phase of the development of the Phu Kham deposit, which entered into production in 2005. The Phu Kham Copper-Gold operation is planned for start-up in mid 2008. Feed to the concentrator will consist of 12MT on average, with planned annual output from this mine being over 200,000 dry metric tonnes (dmt) of concentrate (grading 25% copper), containing 50,000 tonnes copper, 40,000 ounces gold and 400,000 ounces silver, (on average). The concentrate will be exported for further treatment and refining by custom smelters in the Asia Pacific region.

Process technology employed for Phu Kham Copper-Gold is conventional comminution at the head of the circuit, followed by flotation to produce a copper-precious metal concentrate, (Pan Australian, 2006)

Rougher concentrate will be treated through a M10,000 IsaMill™, powered by a 2.6MW motor, treating approximately 168 tph and reducing the feed size from a F80 of 106um to a P80um of 38um, before further flotation. The grinding media for the operation will be MT1.

Prominent Hill Project

Oxiana Limited owns 100% of the Prominent Hill copper-gold project located 650 kilometers north west of Adelaide, and 130 kilometers north west of BHP Billiton’s Olympic Dam in South Australia, Australia.

The ore body consist of copper gold breccia, and will be mined via an open pit. The ore will be treated through a conventional grinding and flotation processing plant, with a designed capacity of 8MTPa. The initial planned concentrate production will be on average 187,000 dry t/a peaking at 230,000t in 2009, with average concentrate grades of 45% copper, 19g/t gold, 57g/t silver. The high grade concentrate will be sold to smelters in Australia and in Asia. (Oxiana, 2007)

One M10,000 IsaMill™, powered by a 3.0MW motor, has been selected to treat the rougher concentrate. It will treat approximately 138 tph, reducing the feed size from a F80 of 125um to P80um of 24um for further flotation. The planned commissioning of the mill will be mid to late 2008. The grinding media for the operation will be MT1.

Anglo Platinum Installations

Anglo Platinum installed the first M10,000 in 2003, in their Western Limb Tailings Retreatment Project, after working with Xstrata Technology to scale up the existing M3000 IsaMill™. The resulting mill, the M10000, was a powered by a 2.6MW motor, which had a variable speed drive (this was a precautionary measure, as it was the first M10,000 to be built).
The mill treats oxidized PGM’s from a tailings dam in a precleaner tail stream, and reduces feed from 75µm to 25µm.

Four years later, in 2007, Anglo Platinum has ordered another five, M10,000 IsaMills™. This time the mills come with fixed speed drives, and are powered by 3MW motors. The high energy efficiency demanded by the sites have called for ceramic media, and the duty of the mills will be similar, i.e. typical duty from 75-100 um feed size down to 53 um product size.

To date, the first IsaMill™ has been successfully commissioned at Potgietersrust Platinum mine (C-Section), and there are mills being manufactured/installed at Potgietersrust A and B Sections (2 mills), followed by two more at the Rustenburg Watervaal UG2 operation in late 2007.

The Potgietersrust Platinum mine (C-Section) mill is designed to operate with a 3MW motor and use MT1 media, treating scats from A and B section primary milling circuits, with the ore having a Bond Work Index, BWi, over 30 kWh/t. Figure 8 illustrates the simplified C section flowsheet with an IsaMill™.

![Figure 8: Simplified PPL C Section Flowsheet with a M10,000 IsaMill™](image)

The rougher tail is milled in closed circuit with a ball mill. Cyclone overflow, at a nominal P80 of 75 µm, proceeds to the IsaMill™ predensifying cyclone which increases the percent solids for efficient IsaMill™ operation, while sending the fines to the scavenger flotation circuit. The IsaMill™ then treats the predensifying cyclone underflow, before the product enters the scavenger flotation circuit.

The IsaMill™ circuit is based on testwork which found very favourable flotation results after IsaMilling™ the cyclone overflow from an F80 of 75 µm down to a product sizing P80 of 53 µm. The circuit is designed to treat 162t/hr (nominally) through the IsaMill™, which is then rejoined with the predensifying cyclone overflow, before it passes to the scavenger flotation circuit. Design energy consumption grinding from F80 75µm to P80 53µm is 9 kWhr/t. The IsaMill will operate with top size 3.5mm MT1 ceramic media.

Before the implementation of the IsaMill™, it was calculated that a 8MW conventional ball mill operated in closed circuit, would be needed to do this duty. However with the M10,000 IsaMill™, less than half the power is required due to open circuit operation and the use of ceramic.
CASE STUDY 3 – IsaMills™ in McARTHUR RIVER SAG CIRCUIT

McArthur River Mine (MRM) is part of Xstrata Zinc, an operating subsidiary of Xstrata PLC. It is a zinc/lead mine, operating in Northern Territory, Australia, and was commissioned in 1995. The development of the IsaMill™ for regrinding down to a P80 of 7 µm was the enabling technology that allowed the mine to be developed. Initially there were 4 x M3000, 1.1MW IsaMills™ in the regrind duty. This has since been expanded to 6 with a combined installed power of 6.7MW. The current plant flowsheet is shown in Figure 9.

Figure 9: McArthur River Flowsheet

The grinding circuit consists of a primary SAG mill, closed by a double deck screen and cyclones. All screen oversize is returned to the SAG Mill. Cyclone overflow (currently at P80 of 70µm) feeds flotation, while cyclone underflow is split between a Tower Mill or returned to the SAG Mill. The Tower Mill product is returned to the SAG sump where it is pumped with the SAG screen underflow to the primary cyclones.

MRM have a need to increase milling capacity to account for decreased head grades as the operation shifts from underground to open cut. At the same time there was a desire to reduce downtime and reduce operating cost by eliminating the Tower Mill from the circuit, hence MRM have been keen to explore the effectiveness of a M10,000 IsaMill™ in the primary grinding circuit. In the primary grinding circuit, the IsaMill™ will be treating material of the order of 300 to 350µm, the coarsest any IsaMill™ has been designed for. The IsaMill™ in the primary grinding circuit will be operating in open circuit, producing a P80 of 40µm product that could be pumped directly to flotation, with the SAG cyclone overflow. The Tower Mill will be decommissioned.

Testwork has been undertaken using a M4 and M20 IsaMills™ and reported by Anderson and Burford (2006). The findings from this work indicated that while the IsaMill™ could treat the screened cyclone underflow feed at 623µm, the presence of SAG mill scats also in the underflow
caused blockages of the small scale mill. The feed to the IsaMill™ can be summarised by the Figure 10.

**Figure 10: McArthur River SAG Cyclone Underflow and IsaMill M20 Feed Distribution**

![Figure 10](image)

The average results from the two site test conducted with the M20 IsaMill™ indicted that the mill could reduce an average feed size of F80 of 350μm, to a product with a P80 of 20 to 30 μm, using MT1 as media (3.5mm in diameter). The testwork also indicated that an energy input of 10-15kWhr/t was required to reduce the feed, screened at a top size of 623μm, to a product P80 of 40μm. The data is shown in Figure 11, marked as “3.5mm Test 1” and “3.5mm Test 2”.

Results were also presented on Figure 11 showing two trials that were also conducted using the 4 litre laboratory IsaMill™, (“M4 test 1” was conducted with unscreened SAG cyclone underflow; “M4 test 2” was conducted with SAG cyclone underflow top screened at 1.7mm).

Data from the M20 site tests shows higher efficiency for a given target size compared to the M4 IsaMill™. Some of the increased efficiency may have been due to the decreased feed size distribution of the site test, but it is unknown what is the impact of this difference, if any. Also the variability and accuracy of the data is affected by the fact that the mill was filling up with steel during the trials, which indicated further testwork was required to confirm the initial results...

**Figure 11: Product Size - Energy Relationship for 3.5mm MT1**

![Figure 11](image)
Continuing Testwork

Further testwork was conducted later in 2006 to support the initial testwork, as well as overcome the presence of scats in the feed stream that lead to blockages in the M20 IsaMill™. Figure 12 displays the flowsheet that was used in this testwork that incorporated a magnetic separator to remove steel scats in the cyclone underflow. Also the feed was screened this time at 1mm (previous screening was at 623um).

Figure 12: Site Testwork at MRM Using M20 IsaMill™ with a Magnetic Separator on Feed

The M20 IsaMill™ was able to treat material, that was slightly finer than the previous work, from a feed sizing of 300um, down to a product sizing of 20 to 25 um, (finer than the 40um target). The data was able to permit a size energy relationship to be established, as shown in Figure 13, compared with the current Tower Mill operation in that circuit.

Figure 13: Size versus Net Energy Comparison for IsaMill™ and Tower Mill

Using the energy data from the M20 IsaMill™ testwork, and the current energy use for the Tower Mill in the primary circuit, it has been conservatively estimated that the IsaMill™ could do the same duty as the Tower Mill for a third of the energy, a saving of 5KWhr/T less energy, i.e. Tower Mill
uses 7 to 8 KWHr/T for a P80 of 100um, while the IsaMill™ achieves same sizing for 2 to 3 KWHr/T.

**McArthur River Mine Future Circuit**

McArthur River Mine has included two M10,000 IsaMills™ in the upgrade of the concentrator to cope with the increased tonnage as the mine goes from underground to open cut. As at March 2007, the concentrator has been planned to increase throughput from 1.8MTpa to 2.5MTPa, with a resulting increase in concentrate production from 320,000T of zinc lead concentrate to 430,000T of zinc lead concentrate. It is planned that the circuit will be ready for commissioning in mid 2008, (Xstrata Zinc, 2007). The IsaMills™ will take 1 year to be manufactured before they are delivered to site. They will take between 1 to 2 weeks to be installed.

McArthur River Mine managements’ selection of IsaMills™ technology for their upgrade was due to the higher efficiency of the mills compared to existing technology. This overcome the need to increase the number of power generators at the remote mine site, while the small footprint of the mills enabled them to fit into the existing concentrator.

Final design parameters are still to be finalised at the time of writing, but the new circuit will have the cyclone underflow from the SAG mill reporting to magnetic separators and oversize screens to ensure all iron scats are taken from the stream before treatment by the two M10,000 IsaMills™.

Further testwork will also be undertaken during 2007 to trial one of the M3000 IsaMills™ in the fine grinding circuit to undertake the coarser duty. The first trial is planned for May, 2007.
CONCLUSION

IsaMill™ technology is becoming the preferred technology in efficient coarse grinding circuits. The development of reliable ceramic media, such as MT1, as well as the development of high capacity M10,000 IsaMills™, have lead to the IsaMill™ being a realistic alternative in coarse grinding applications.

IsaMills™ in coarse grinding applications have all the advantages of fine grinding applications, such as simple circuit design, small footprint and iron free contamination when used with ceramic media, which has profound beneficial impact on metallurgy, recirculating loads and reagent use. However the biggest advantage of IsaMills™ is its high energy efficiency, as a result of its high speed stirring action in a packed bed.

With the pressure being applied to all industries today for improved sustainability and the need for increased energy efficiency, the adoption of IsaMill technology in coarse applications is good news for operators, as the development of the IsaMills™ allows for significant energy and capital savings for the application.

IsaMill™ technology is a true alternative to ball mill and tower mill circuits.
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Transforming Flowsheet Design with Inert Grinding - the IsaMill

J.D. Pease¹, D.C. Curry¹, K. E. Barns¹, M.F. Young², C. Rule³

¹ Xstrata Technology
Level 2, 87 Wickham Tce
Brisbane, QLD, 4000
Australia
Ph: +61 7 3833 8555

² Xstrata Zinc
Mount Isa, Qld,
Australia

³ Anglo American Platinum Corporation
55 Marshall St, Johannesburg
South Africa

Key words: grinding, flotation, inert media
ABSTRACT
The IsaMill was developed for fine-grained ores that required at least double the grinding efficiency of ball or tower milling to be economic. This was achieved, but in practice, the benefits of using inert media (e.g., sand, slag, ore, ceramic) have proven to be at least as important as the higher grinding efficiency. Flotation selectivity and rate was improved for all particles, particularly fines. This allowed a dramatic simplification of the Mount Isa lead zinc flotation circuit – adding 6 MW of ultrafine grinding power reduced reagent addition and flotation volume and increased plant energy efficiency. This was quite unexpected.

This paper examines four orebodies that were enabled by inert grinding – McArthur River, George Fisher and Black Star Open Cut Mines (complex fine-grained lead zinc orebodies in Australia), and the Western Limb Tailings Re-treatment plant (a PGM operation in South Africa). In a unique case history, the operating performance of the Mt Isa lead zinc concentrator is explained by size-by-size mineralogical data collected over 25 years, systematically explaining the impact of declining ore quality, and the effects of additional conventional grinding, inert grinding, and circuit and reagent redesign.

The mineralogical case histories are so compelling that it is argued that the advantages of inert grinding should not be confined to difficult, fine-grained orebodies. The availability of large-scale efficient inert grinding mills could have profound impact on circuit design for many orebodies.

INTRODUCTION
The IsaMill was developed by MIM (now Xstrata) to enable the development of the McArthur River Orebody in the Northern Territory of Australia. This ultra-fine grained ore needed a grind size of 80% passing 7 microns to produce a saleable concentrate. Such a fine grind was not economic with conventional ball or tower milling. Below about 25 microns the low energy efficiency and high media consumption of these mills is prohibitive. Critically, the high steel consumption and associated change in pulp and surface chemistry also seriously affects flotation metallurgy.

Figure 1 demonstrates the need for stirred milling to reduce power consumption for fine grinding of a pyrite concentrate. A grind P80 of 12 microns requires over 120 kwh/t in a ball mill using 9mm balls, but only 40 kwh/t in an IsaMill with 2mm media. The curve demonstrates the difference between “conventional” grinding and fine grinding. The Bond Work Index is developed for the flat “A” section of the curve, but does not apply to the exponential increase in grinding energy after the “knee”, section “B”. Here, very small changes in target P80 mean huge increases in power requirement, which cannot be achieved with conventional mills.

The key to efficient fine grinding is fine grinding media. A charge of 2mm media has 91 times more particles than the same volume of 12mm media; the same charge of 1 mm media has 730 times more particles. More particles means more breakage events, but to build a practical mill to use fine media requires three things:
- The small media particles need sufficient momentum to quickly break small ore particles. Slow stirring speeds may still grind efficiently, but the slow breakage rates will make the mill size prohibitive.
- An economic source of fine media.
• A practical means for the mill to retain media and pass target size product.

The IsaMill achieves these needs by combining high power intensity (280 kw/m³, compared with about 20kw/m³ for a ball mill), high stirring speed (22 m/s tip speed), and a centrifugal discharge to retain the media. The screenless centrifugal discharge allows low cost media to be used – eg granulated slag, silica sand, gravel fractions of the ore.

Figure 1: Grinding Energy vs Product size for a pyrite concentrate

The grinding mechanism and high power efficiency of the IsaMill is discussed elsewhere (Pease 2004). This paper will focus on another aspect of the technology that has been equally important in enabling orebodies – the impact of high intensity, inert grinding on flotation performance. Compared with steel grinding, inert grinding profoundly changes fines flotation, demolishing many common theories about fines behaviour. For example, at McArthur River, 96% of the individual particles recovered are less than 2.5 micron, and they are recovered at high grade and recovery using conventional cells. At Mt Isa, recovery from cleaner feed is above 95% for all size fractions from 1 micron to 37 micron, also using conventional (second hand) flotation cells (Pease 2004). Contrary to the common belief that “slimes don’t float”, the best performance, above 98% cleaner recovery, is in the 4 to 16 micron fraction.

A key question is – are the metallurgy gains of inert grinding confined to fines flotation? Until recently this was a hypothetical question – there was no practical technology for inert grinding at coarser sizes. This paper contends that recent developments have changed that – the availability of large scale (2.6MW) IsaMills and low cost ceramic grinding media have extended the gains in power efficiency and metallurgy to coarser grinds. In 1995 stirred milling “crossed over” from industrial manufacturing to the ultrafine mineral grinding. It is now ready for a more significant crossover to coarse mineral applications.
IMPACT OF STEEL GRINDING ON FLOTATION
Historically fine grinding applications use conventional ball or Tower Milling. As well as the high cost and low power efficiency, the chemical impacts of steel grinding on flotation offset the benefits of better liberation. Operations often respond to the poor flotation chemistry by using higher reagent additions or intensive conditioning to clean mineral surfaces, and pre-aeration steps to increase pulp potential and mineral hydrophobicity. These responses are expensive and only partly effective, and do not address the root cause.

In the early 1960s investigations by Ray and Formanek looked at beneficial effects on flotation of grinding lead-zinc ore in porcelain mills compared with grinding in iron mills or the addition of iron powder to porcelain mills (Kocabag, 1985). This work was confirmed by Fahlstrom (1960) and Thornton (1973) on chalcopyrite, galena and sphalerite ores. Later work by Greet (2004), Cullinan (1999 and 1999b), Pietrobon (2004), Grano (1994) and Johnson (2002) and Fleahy (1994) supported the benefits of inert grinding for a wide range of ores including nickel.

If a mineral or metal is immersed in water it assumes an electrical potential with respect to the water. The principal reactions are oxidation of the mineral to form metal ions and elemental sulphur, oxidation of xanthates to dixanthogens, and reduction of oxygen to hydroxyl ions (Figure 2).

Oxidative reaction: \[ \text{MS} \rightarrow \text{M}^{2+} + \text{S}^0 + 2\text{e}^- \] oxidation of mineral sulphide
\[ 2\text{X}^- \rightarrow \text{X}_2 + 2\text{e}^- \] oxidation of xanthate

Reductive reaction: \[ \frac{1}{2}\text{O}_2 + \text{H}_2\text{O} + 2\text{e}^- \rightarrow 2\text{OH}^- \] reduction of O$_2$

**Figure 2: The anodic and cathodic domains in a mineral system**

Sulphide minerals are semi-conductors. When they are brought into contact with steel media an electrochemical cell is formed (Figure 3). The steel media has the highest rest potential and is the anode; the sulphide mineral is the cathode.

Oxidative Reaction: \[ \text{Fe} \rightarrow \text{Fe}^{2+} + 2\text{e}^- \] oxidation of steel media: anode
Reductive Reaction: \[ 2\text{e}^- + \text{H}_2\text{O} + \frac{1}{2}\text{O}_2 \rightarrow 2\text{OH}^- \] reduction of oxygen: cathode

Further reduction of Fe ions from solution results in iron hydroxide deposits on the mineral surface, as shown in Figure 3.
As a result, grinding in a steel media environment has several detrimental effects:

- **The Eh effect**: the reducing environment lowers dissolved oxygen and Eh of slurry. Since collector adsorption is Eh dependent and may require oxidation of xanthates to dixanthogen, this will reduce floatability. Providing pre-aeration steps before flotation reduces this impact, but is unlikely to completely reverse it. This effect of grinding media on Eh and floatability has been well documented by Trahar (1984).

- Oxidation of steel grinding media causes *iron hydroxide coatings on mineral surfaces*. This reduces flotation selectivity in all sizes, but particularly for fine particles. This is even evident at the coarse sizes of autogenous grinding (figures 4-6). The impact is more significant in secondary and regrinds mills, as more fines are created and steel consumption is much higher.

- Oxygen reduction on mineral surfaces promotes *precipitation of hydrophyllic, insoluble metal hydroxides on the surface of sulphide minerals*. The impact on flotation is more pronounced for fine particles (Figure 6). Surface coatings can be offset by higher reagent addition, but at the cost of lower selectivity for clean minerals, as well as the higher reagent cost.
Some of the flotation impacts of steel media can be overcome by increasing pH and higher reagent addition. But flotation selectivity will still be low – some coarser minerals will have clean surfaces and will suffer from higher reagents, but the coated fines surfaces will demand them. Further, some flotation circuits require low pH (ie: pyrite/arsenopyrite/gold flotation circuits).

A much better solution is to address the root cause of the problem – keep all mineral surfaces clean by using inert media. While the benefits of high chrome media over forged steel media have been well documented, ceramic media has a much greater impact. Table 1 (Greet, 2004b) shows that changing from forged steel to high chrome media reduces the surface atomic composition of iron from 16.6% to 10.2%. Grinding with ceramic media reduces this surface iron measure to below 0.1%. Considerable work has been done to demonstrate the flotation advantages of high chrome media. By contrast, little commercial based work has been done on fully inert media since it has not been a practical option until recently.

With steel media, the liberation benefits of grinding are offset by the chemistry impacts – a case of “two steps forward and one step back”. For grinding below 25 microns the chemical impact may be dominant – grinding is “one step forward, but two steps back”. The availability of large scale inert grinding allows operators to improve both liberation and flotation performance, getting full value from the installed grinding power.
Table 1: Composition determined via XPS, of the unetched surfaces of Rapid Bay Galena ground with different media (Greet 2004b, Cullinan 1999)

<table>
<thead>
<tr>
<th>Media Type</th>
<th>Surface Atomic Composition (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>O</td>
</tr>
<tr>
<td>Forged Steel</td>
<td>53.1</td>
</tr>
<tr>
<td>High Chrome</td>
<td>50.0</td>
</tr>
<tr>
<td>Ceramic</td>
<td>33.6</td>
</tr>
</tbody>
</table>

CASE STUDY 1: MCArTHUR RIVER MINING (MRM)

The McArthur River orebody was discovered in 1955, with a resource of 227 Mt at 9.2% Zn and 4.1% Pb. However it remained undeveloped for forty years since existing technology could not economically treat the extremely fine grained minerals (Figure 7). The IsaMill was developed specifically to treat this orebody. It allowed economic regrinding to 80% passing 7 microns, fine enough to reduce silica in bulk concentrate to marketable levels.

Figure 7: Different Grain Size of Broken Hill and McArthur River Ores (Grey Square is 40μm)

The plant started mid 1995 with 4 IsaMills regrinding rougher concentrate. Media for the mills was screened ore gravel from the SAG mill discharge – a fully autogenous ultra-fine grind! Two more mills were installed to increase production and recovery (in 1998 and in 2001). In 2004 the media was changed from ore gravel to screened sand – the higher efficiency of the sand increased mill capacity, and reduced wear on mill components at the higher throughput.

Table 2 shows production performance at MRM – very high concentrate grades are achieved at over 80% recovery for the fine grind. The concentrate sizing is P80 7 microns, and P50 2.5 microns. Looking at this from the perspective of a flotation bubble, 50% by weight means that 96% of the particles in MRM concentrate are finer than 2.5 microns. So 96% of the successful particle-bubble collisions at MRM happen for particles finer than 2.5 microns. This is achieved in conventional flotation cells – the selection criteria was to design for adequate lip length, then buy the cheapest cells available.
Fines float extremely well if they have fresh clean surfaces.

Table 2: Performance of McArthur River since commissioning

<table>
<thead>
<tr>
<th>Year</th>
<th>Tonnes</th>
<th>Head Grade</th>
<th>Tonnes</th>
<th>Zn Recovery</th>
<th>Con Grade</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td>%Zn</td>
<td>%Pb</td>
</tr>
<tr>
<td>1995/96</td>
<td>707,994</td>
<td>12.9%</td>
<td>759,519</td>
<td>66.4%</td>
<td>39.3%</td>
</tr>
<tr>
<td>1996/97</td>
<td>1,035,222</td>
<td>14.4%</td>
<td>1,026,150</td>
<td>73.5%</td>
<td>43.5%</td>
</tr>
<tr>
<td>1997/98</td>
<td>1,127,000</td>
<td>16.1%</td>
<td>1,139,000</td>
<td>74.3%</td>
<td>43.3%</td>
</tr>
<tr>
<td>1998/99</td>
<td>1,222,238</td>
<td>16.4%</td>
<td>1,220,957</td>
<td>79.5%</td>
<td>45.0%</td>
</tr>
<tr>
<td>1999/00</td>
<td>1,254,227</td>
<td>16.3%</td>
<td>1,262,639</td>
<td>80.9%</td>
<td>46.9%</td>
</tr>
<tr>
<td>2000/01</td>
<td>1,226,499</td>
<td>15.4%</td>
<td>1,270,319</td>
<td>82.4%</td>
<td>46.8%</td>
</tr>
<tr>
<td>2001/02</td>
<td>1,398,109</td>
<td>14.9%</td>
<td>1,404,539</td>
<td>82.7%</td>
<td>46.8%</td>
</tr>
<tr>
<td>2002/03</td>
<td>1,505,306</td>
<td>12.7%</td>
<td>1,511,856</td>
<td>82.4%</td>
<td>46.6%</td>
</tr>
<tr>
<td>2004</td>
<td>1,523,243</td>
<td>12.7%</td>
<td>1,579,762</td>
<td>80.0%</td>
<td>47.1%</td>
</tr>
</tbody>
</table>

CASE STUDY 2: GEORGE FISHER OREBODY AT MOUNT ISA

The changes to the Mount Isa circuit as part of the “George Fisher Project” are detailed elsewhere (Young & Gao, 2000, Young, Pease & Fisher, 2000). In summary, the project involved adding a further 6 IsaMills, to regrind lead rougher concentrate to P<sub>80</sub> of 12 µm, most zinc rougher concentrate to 12 µm, and a zinc regrind to P<sub>80</sub> of 7 µm (Figure 8). Lead performance increased by 5% concentrate grade and 5% recovery (equivalent to 10% increase in lead recovery at the same grade). Zinc recovery increased by 10%, in two steps, and zinc concentrate grade by 2% (equivalent to 16% increase in zinc recovery at the same grade). The story of zinc metallurgy is told in Figures 8, 9 and 10.

The project predicted 5% higher zinc recovery (and no extra concentrate grade) due to extra liberation. Figure 9 shows this was achieved instantly. The surprise was the “second wave” of a further 5% zinc recovery increase and the 2% increase in zinc concentrate grade. This was because fines flotation improved after grinding finer.

It took about 6 months to discover how much better the fines could perform because we were so used to flotation after conventional grinding rather than after IsaMill. Our three biggest mistakes were:

- Expecting to need a lot more reagents after IsaMill due to the huge new surface area created. Some reagent additions were forecast to triple.
- Not taking the depressant (lime to pH 11) off the zinc cleaners.
- “Pulling” flotation harder because we thought flotation rates of the fines would be slower.
To our surprise, even though we introduced 6MW of extra grinding power, operating cost per tonne of feed did not increase. This was because of:

- Lower reagent additions
- Elimination of circulation loads between roughing and cleaning – a lot of power (and flotation capacity) is wasted in conventional circuits by pumping circulating loads of 100%-300%.
- Virtual elimination of spillage – due to new designs for pump boxes and pumps, the lower reagents, and especially the reduction in circulating loads.

The reduction in reagents was most unexpected, but perfectly logical in hindsight. During mill commissioning we increased reagents because we were increasing mineral surface area threefold. We were wrong. While more surface means more collector on the surface, this doesn’t necessarily mean more collector in the pulp. If surface coatings mask mineral surfaces, then high collector in solution is needed to drive the diffusion through the coating to the surface. But if surfaces are truly clean – something that is never seen in a conventional steel grinding circuit but can be achieved by grinding in an inert environment – lower solution concentrations can achieve high surface coverage. Flotation after inert grinding is profoundly different – we had to forget everything we had learnt in our circuit and start with a “clean slate”. Later we realised that the answers we needed were in the classic flotation texts. Bubble contact angles in clean mineral systems may not seem relevant to the plant operator who has badly altered mineral surfaces. But they are quite relevant in a system that quickly creates fresh clean surface.

The new circuit cut circulating loads from 200-300% to less than 50%. Liberated minerals with clean surfaces in a narrow size distributions respond predictably. They don’t form large circulating loads of “undecided” particles. This creates a virtuous circle – lower circulating loads in cleaning means lower density, which gives better dilution cleaning. It also means less flotation cells are needed – eliminating a 100% circulating load doubles residence time. Mt Isa had to shut down some zinc cleaning capacity after installing the IsaMills.

The reduction in reagents and reduction in circulating loads almost eliminated spillage. This is another virtuous circle – returning spillage disrupts a circuit, creating new circulating loads and spillage.

Figure 10 shows the recovery by size in the zinc cleaning circuit at Mt Isa after IsaMilling (the recovery with respect to rougher concentrate, which is the feed to IsaMilling). Recovery is above 95% for all size fractions from 1micron to 37 micron. Recovery drops above 37 microns but there are very few particles here – these are composite that the circuit directs to regrinding to fine liberated, high recovery fractions. The highest recovery, over 98%, is in the 4-16 micron range. This size range is sometimes called “slimes”, and it is often said that “slimes don’t float”. Indeed, after steel grinding they often don’t.
Mt Isa Pb / Zn Concentrator Flow Sheet

Figure 8: Mt Isa Pb/Zn Concentrator Flow Sheet

Figure 9: Zinc Recovery Increase from IsaMilling
Figure 10: Mt Isa Zinc Recovery from Rougher Concentrate by Size

The Big Picture – Mineralogy, Liberation, Chemistry and Recovery Explained

Figure 11 is a graphic summary of the impacts of ore type, liberation, steel grinding, inert grinding and chemistry changes unfolding over twenty years at the Mt Isa operation. Each month Mt Isa collects inventory samples of all feed and exit streams. Each exit stream is sized and cyclosized, and size fractions are submitted for quantitative mineralogy. A huge amount of data is generated about different mineral classes in different size fractions. However one summary variable plotted in Figure 11 tells a compelling story. It plots sphalerite liberation in “recalculated feed” and sphalerite recovery. “Recalculated feed” is created by mathematically combining all the plant exit streams according to their relative tonnage. Therefore it captures the impact of all the grinding and regrinding stages of the circuit. The Mt Isa journey unfolds in four stages, as described below.
Stage 1: The sickening decline

During the 1980’s ore sources became increasingly fine grained. Liberation declined, and recovery inexorably declined with it. Like most operators under profit pressure we increased feed tonnage. This increased revenue but further reduced liberation. We tried dozens of reagents, dozens of circuit changes, and dozens of metallurgists. Finally we accepted we were just “shuffling the deck chairs on the Titanic” – reagents don’t grind. A fundamental rule seemed to guide plant performance in spite of our metallurgical endeavours – recovery equals liberation plus 10%. A small amount of composite particles can be accepted into concentrate, until the quality constraint on impurities is reached.

Liberation dropped by 15%
Recovery dropped by 20%
**Stage 2: Use a bigger hammer**
We had to increase liberation to increase recovery. The only available technology was conventional ball milling, and the then emerging Tower Milling. We installed 6MW of ball and Tower Mills, effectively doubling the grinding capacity. While we knew this would increase liberation, we also knew there was a downside – creating finer particles using so much steel media would create surface chemistry problems. As expected, this delivered two steps forward from liberation, and one step backward from surface chemistry. Liberation increased by 18%, while zinc recovery increased by 7%. The old rule “recovery is liberation plus 10%” had broken down, but this was the inevitable price of grinding fine with steel media.

<table>
<thead>
<tr>
<th>Liberation increased by 18%</th>
</tr>
</thead>
<tbody>
<tr>
<td>Recovery increased by 7%</td>
</tr>
<tr>
<td>Concentrate grade unchanged</td>
</tr>
<tr>
<td>(not shown)</td>
</tr>
</tbody>
</table>

**Stage 3: A pleasant surprise**
The IsaMill technology for McArthur River was developed in the Mt Isa concentrator. The first installations were to regrind lead cleaner feed. The intent was to liberate some of the galena/sphalerite binaries to reject some zinc from lead concentrate. This was achieved, though the total effect on sphalerite liberation was small. However zinc recovery to zinc concentrate increased by 4%. This was because the clean surfaces after IsaMilling allowed us to improve flotation chemistry – we were able to reject fine liberated sphalerite particles that had previously misreported to lead concentrate. We had achieved what no amount of chemistry or circuit changes had been able to achieve in the past – increasing recovery increase without significantly changing liberation.

<table>
<thead>
<tr>
<th>Liberation change negligible</th>
</tr>
</thead>
<tbody>
<tr>
<td>Recovery increased by 4%</td>
</tr>
</tbody>
</table>
Stage 4: Cooking with Gas

Breaking the quandary between increasing liberation and harming surface chemistry was a revelation. With steel milling we had reached to stage of one step forward from liberation, but two steps back from chemistry. Suddenly a new option was available—improve both liberation and chemistry in the same step. We put 4 MW IsaMills into the zinc circuit. Liberation increased 4%, but recovery increased by 10%. Concentrate grade also increased by 2%—we could have taken this as an extra 6-8% recovery, but the higher grade was more profitable. Making the higher grade had not been possible in the past.

<table>
<thead>
<tr>
<th>Liberation increased by 4%</th>
</tr>
</thead>
<tbody>
<tr>
<td>Recovery increased by 10%</td>
</tr>
<tr>
<td>Concentrate grade increased 2% (not shown)</td>
</tr>
</tbody>
</table>

Figure 10 shows that fine sphalerite recovery is over 95% after IsaMilling (ie: recovery of sphalerite from rougher concentrate). Figure 11 however shows total circuit sphalerite is still only 80% with respect to plant feed. Of the 18% sphalerite losses, 7% reports to lead concentrate, and 11% reports to tailing. In the tailings about half is in coarse composites and half in fine liberated particles. If a particle is too coarse, or too badly surface altered by steel grinding, then it doesn’t float in the roughers, and doesn’t get a chance to see the IsaMills.

Undoubtedly rougher performance would be better after inert grinding, but the technology was not available when we installed the extra secondary grinding. Mount Isa is now considering the feasibility of a large open cut operation which would necessitate building a new concentrator. Our design for that concentrator uses large IsaMills (3.3MW or bigger) in place of ball mills to grind rougher feed, and further mills to regrind rougher concentrate. This reduces footprint, reduces capital cost, increases energy efficiency, and will achieve higher recovery with lower reagent additions than the current circuit.

CASE STUDY 3: MT ISA BLACK STAR OPEN CUT

Surface resources at Mt Isa had long been a target for open cut mining. However the poor metallurgical response was always a barrier to production. Much of the ore is “transitional” between surface oxides and deeper primary sulphides. The transition ore is lower grade than primary ore, has fine grained mineralogy, and leaching has activated pyrite and sphalerite, leading to non-selective flotation. Constant attempts over the last 80 years failed to make the ore economic, with flotation unable to make smelter quality concentrates at any recovery.

The development of the IsaMills and the flowsheet to treat George Fisher ore changed this. The fine grinding achieves mineral liberation and cleans the mineral surfaces by attrition, and the combination of high intensity inert grinding and the correct water chemistry in flotation...
stops re-activation of unwanted minerals. The impact is shown by the grade recovery curve in Figure 12 - target concentrate grades can now be made at acceptable recoveries.

As a result, IsaMill inert grinding technology enabled production from the Black Star Open Cut resource. Production commenced in the first half of 2005, targeting 1.5M t/y to supplement underground production, produced from a mineral resource of 25Mt at 5.1%Zn and 2.7%Pb. This represents only a small portion of the potential open cut resources at Mt Isa. The success to date has led reassessment of the economics of the entire open cut resource and future production scale at Mt Isa.

![Figure 12: Pb Grade/Recovery Curve – ISA Lead-Zinc Transition Ore](image)

**CASE STUDY 4 : MERENSKY PLATINUM TAILINGS RETREATMENT PLANT**

In 2001 Anglo Platinum assessed the retreatment of dormant tailings dams in the Rustenberg area in South Africa. These tailings contained economic amounts of Platinum Group Minerals (PGMs) if new technology could address two issues:

- The fine grained mineralisation of the PGMs in tailings (why it wasn’t recovered first time)
- Surface oxidation and oxidation products which harmed flotation – some of the tailings were placed over 100 years ago.

Anglo Platinum and Xstrata Technology worked together to find an economic treatment route. To achieve economies of scale for the project the IsaMill was successfully scaled up from 1,000 kW to 2,600kW. This proved to be the enabling technology for this project due to:

- The ability to grind fine at low cost – the mill operates in open circuit, and uses cheap local sand as the grinding media.
- The clean mineral surfaces resulting from the inert grinding environment. This was crucial to achieve target grades and recoveries after regrinding.

Figure 13 shows the improvement made by an IsaMill regrinding rougher concentrate before cleaning (Buys et al, 2004). The mill increases flotation kinetics in cleaning, just as it does at Mt Isa and McArthur River. This is in contrast with the common observation that regrinding in a steel mill slows kinetics of all minerals.
Anglo Platinum commissioned the Western Limb Tailings Retreatment Plant in 2004. At the end of 2004 they concluded that (Buys, 2004):

- IsaMill technology was enabling for the WLTR project since it allowed acceptable concentrate grades to be made from oxidised slow floating tailings.
- Flotation kinetics improved after fine grinding due to both extra liberation and the removal of iron oxide surface coatings. Inert fine grinding of rougher concentrate was necessary.
- The scale up to the M10,000 IsaMill (from 1 MW to 2.6 MW) was successful.

**Figure 13:** Improvement in Platinum Grade/Recovery After IsaMilling for Western Limb Tailings Retreatment

**Figure 14:** Western Limb Tailings Retreatment Flowsheet

**INERT GRINDING BEFORE LEACHING**
Just as the ability to grind to 10 microns in an inert grinding environment has been the enabling technology for the flotation of fine grained ore bodies so to this ability has been an enabling step for several hydrometallurgical technologies. The high surface area of fine particles means high leaching rates at relatively low temperature and pressure, reducing capital and operating costs. High intensity fine grinding also reduces the activation energy required to leach minerals by creating a highly stressed surface, reducing the crystalline nature to amorphous phases. This effect of mechanical (or mechanochemical) activation of minerals is well reported (Balaz, 2000; Juhasz and Opoczky, 1990; Grelach et al, 1989), and it means that minerals leach under much less aggressive conditions. The higher power intensity of grinding would enhance this effect. Several emerging leaching processes have been based on fine grinding of feed – the Activox process, the UBC/Anglo process, the Phelps Dodge Process, and Xstrata’s Albion Process.

In flotation the chemical benefits of inert grinding can be dramatic. In leaching there is a lesser, but still important impact. Conventional grinding before leaching will directly input steel to the leach feed. This may require additional preoxidation of leach feed, and also result in higher reagent consumption. An IsaMill will soon be trialed to replace a ball mill regrinding concentrate before gold cyanidation. The ball mill is currently consuming 10t/day of steel media, which enters leach feed. This will provide a future case study of the impact of inert grinding on leach performance and reagent consumption.

THE CROSSOVER TO COARSER GRINDING
The principles learnt for fines flotation also apply for coarser particles (above 30 microns). The impact of steel on flotation is not as dramatic as it is for fines, but it is still there. This is why so much work has been done on the use of high chrome media. The principles of floating with clean surfaces, in narrow size distributions, with fast kinetics, low reagents and low circulating loads are vital to achieve good fines recovery, and will also improve flotation of coarser particles.

Until recently the prospect of inert grinding at coarse sizes was generally impractical. This has changed with IsaMills now operating at 2.6 MW, and the availability of low cost, high efficiency ceramic (MT1) developed specifically for stirred milling (Curry and Clermont, 2005). Pilot work on coarser feeds (eg 200 microns) consistently shows that IsaMills with ceramic media have significantly higher power efficiency and lower capital cost installations (eg open circuit) than conventional ball milling. The crossover to coarse grinding may be a “transformational” technology change, delivering lower capital, higher efficiency grinding, as well as better metallurgy.

The pilot predictions need to be demonstrated at full scale. The first industrial application of a 2.6 MW IsaMill with MT1 ceramic will be commissioned in late 2005, at Centerra Gold’s Kumtor operation in the Kyrgyz Republic. This operation which produces 500,000 oz/y gold will use the IsaMill to grind rougher concentrate before leaching. This application will allow full scale comparison of grinding rates and size distributions achieved by SAG mills, ball mills and IsaMills.

Another full scale coarse grind application is currently under design, and is expected to be operating in late 2006. This will involve a 3.3 MW IsaMill using MT1 grinding rougher
tailings in open circuit, from 150 micron to 55 micron. Based on current plant data, an 8 MW conventional ball mill in closed circuit would be required to match this grinding performance. Ball milling cannot be justified for this application due both to the high capital cost and the relatively poor metallurgy compared with inert grinding.

CONCLUSION
Necessity was the mother of invention for ultra-fine grained orebodies. They were simply intractable with conventional ball and Tower milling. High intensity stirred milling was the breakthrough that transformed the economics – low capital and installation cost, low media cost, high power efficiency, simple installations, sharp size distributions in open-circuit, and inert attrition of mineral surfaces.

It is most unlikely that only fine grained ores will benefit from these features. The rapid increase in scale to 3.3 MW enables large scale coarser grinding applications for IsaMills. Several projects are under design to transfer this technology to the mainstream.

The performance of these projects will answer the question: is this the next big thing in grinding?

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THE APPLICATION OF LARGE SCALE STIRRED MILLING TO THE RETREATMENT OF MERENSKY PLATINUM TAILINGS

Stephan Buys\textsuperscript{1}, Plant Superintendent
Chris Rule\textsuperscript{2}, General Manager - Concentrator Technology
Dan Curry\textsuperscript{3}, Technical Superintendent – Mineral Processing

\textsuperscript{1}Anglo Platinum Management Services (Pty) LTD
Process Technology Division
Central Deep
Rustenburg
Republic of South Africa
+27 14 598 2312
E-mail: stephanb@angloplat.com

\textsuperscript{2}Anglo Platinum Management Services (Pty) LTD
Process Technology Division
55 Marshal Street
Johannesburg
Gauteng
Republic of South Africa
+27 11 373 6998
E-mail: chrisrule@angloplat.com

\textsuperscript{3}Xstrata Technology Pty Ltd
Level 2, 87 Wickham Terrace
Brisbane
Queensland 4000
Australia
Tel: +61 7 3833 8500
E-mail: dcurry@xstrata.com.au

Key Words:
Platinum, PGM’s, tailings, fine grinding, IsaMill, liberation, surface chemistry, fine flotation
ABSTRACT

The use of fine grinding (FG) in mineral processing is well established. The application of FG in the South African platinum industry is more recent and its use in the recovery of PGM’s from dormant tailings dams is unique. Anglo Platinum uses IsaMill FG technology at their Western Limb Tailings Retreatment project (WLTR) near Rustenburg in the North West Province of South Africa.

In December 2003, Anglo Platinum commissioned the first purpose built tailings re-treatment facility for PGM recovery in South Africa. The concentrator re-treats Merensky ore tailings from concentrators that operated early last century. The WLTR facility takes advantage of modern technology, such as FG, to economically recover PGM’s from material historically considered as waste. The WLTR concentrator has a capacity of 4.8 Mtpa, and has been designed to easily expand to 10.8 Mtpa. The flow sheet includes recovery of tailings by high pressure water monitoring, ball milling, rougher flotation, rougher concentrate reground and cleaner/recleaner flotation.

The rougher concentrate is reground in an IsaMill, which is a stirred mill that operates with inert silica sand grinding media. For this project, Anglo Platinum required a stirred mill of a unit size not previously available. Xstrata Technology and Anglo Platinum collaboratively developed the largest wet, fine grinding mill available; the M10,000, 2.6 MW IsaMill.

The use of IsaMill technology was enabling for the WLTR project, as it allowed smeltable concentrate grades to be produced from the oxidised, slow floating tailings. The economics and practicality of fine grinding has fundamentally changed with the M10,000 IsaMill. Flotation is transformed by high intensity attritioning in the inert media environment, and the narrow size distribution of product in open circuit configuration. The use of conventional steel media grinding to improve fine particle flotation is limited due to the oxidation of liberated mineral surfaces inherent in these environments.

INTRODUCTION

During 2000 and 2001, Anglo Platinum conducted metallurgical and geological investigations into the retreatment of dormant tailings dams in the Rustenburg area. At the current market value of platinum and palladium, the concentration of PGM minerals in the dams represented a possible economic resource. Metallurgical test work identified a significant proportion of these minerals that could be recovered via fine grinding and flotation.

A pilot plant program was developed to confirm the effectiveness of fine grinding with respect to flotation recovery. The results from the program assisted with the proposed concentrator process design and samples were generated for vendor test work of critical equipment.

A collaborative design project between Anglo Platinum and Xstrata (then MIM – Mt Isa Mines) was initiated to run in parallel with the pilot plant tests and plant design. This project required a
detailed design for a fine grinding IsaMill of 2,600 kW (3500 hp) capacity, to maximise economies of scale with this unit process.

**METALLURGICAL TEST PROGRAM**

**Bulk Sample Collection**

The Klipfontein Central tailings dam was selected to provide the bulk sample for pilot testing. A total of 56 holes were drilled, producing approximately 424 tonnes of sample. The resource for the Rustenburg section was calculated to be 186 Mt at 1.08 g/t PGM + Au. Over 4 Moz of platinum were contained within this resource. The Klipfontein Central bulk sample was representative of the total Rustenburg resource.

**Test Program Detail**

Using previous laboratory scale results in conjunction with certain process design criteria, the pilot scale program was developed to confirm the integrity of the concepts identified previously. Four campaigns were designed to provide the necessary information. The grinding circuits are summarised in Table 2.

<table>
<thead>
<tr>
<th>Table 1 : Grinding Targets and Mill Type</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Primary Grinding</strong></td>
</tr>
<tr>
<td><strong>Mill Type</strong></td>
</tr>
<tr>
<td>1</td>
</tr>
<tr>
<td>2</td>
</tr>
<tr>
<td>3</td>
</tr>
<tr>
<td>4</td>
</tr>
</tbody>
</table>

A nominal grind of 80 % passing 75 μm was achieved by using a primary and a secondary ball mill (secondary mill in closed circuit with cyclone) for Runs 1 and 3. An IsaMill was used in open circuit to grind the cyclone product to 90 % passing 75 μm for Runs 2 and 4. All tests used rougher flotation with a residence time of 45 minutes. A cleaning /recleaning stage was used to increase the rougher concentrate grade to 50 g/t (final concentrate target). An IsaMill was used in open circuit to regrind the rougher concentrate to 90 % passing 25 μm prior to cleaning for Runs 3 and 4. The rougher and cleaner bank flotation tails reported to final tails. The recleaner tail reported to the cleaner feed. Reagent selection, addition rates and locations were based on results from laboratory scale test work.

**Head Assay Analysis**

The bulk sample was mixed prior to being fed to the pilot plant. The statistics in Table 2 indicate that this mixing process was effective, as very little variation in grade is evident. Samples were collected over three one-hour periods for each Run.
Table 2: Head Assay Analysis

<table>
<thead>
<tr>
<th></th>
<th>3E (g/t)</th>
<th>Cu (%)</th>
<th>Ni (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Mean</td>
<td>1.17</td>
<td>0.027</td>
<td>0.076</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>0.04</td>
<td>0.003</td>
<td>0.010</td>
</tr>
<tr>
<td>% RSD</td>
<td>3.45</td>
<td>9.49</td>
<td>13.86</td>
</tr>
</tbody>
</table>

Pilot Plant Results

A summary of the flotation results, with respect to PGM, for each run is presented in figure 1. The overall recovery achieved for the four runs were similar with runs 1 and 3 achieving slightly better results. Final concentrate grade improves significantly by introducing a finer product to the cleaner flotation circuit.

![Figure 1: Final PGM Recovery and Concentrate Grade](image)

The final concentrate grade is dependent upon grind size. Table 3 demonstrates the particle sizes of the rougher and cleaner concentrate streams.

Table 3: Rougher and Cleaner Stream Sizings

<table>
<thead>
<tr>
<th>Run #</th>
<th>Rougher Feed % Passing 75 µm</th>
<th>Cleaner Feed % Passing 25 µm</th>
<th>Rougher Conc Regrind</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>77.0</td>
<td>50.3</td>
<td>No</td>
</tr>
<tr>
<td>2</td>
<td>93.4</td>
<td>85.4</td>
<td>No</td>
</tr>
<tr>
<td>3</td>
<td>76.6</td>
<td>83.4</td>
<td>Yes</td>
</tr>
<tr>
<td>4</td>
<td>94.6</td>
<td>88.1</td>
<td>Yes</td>
</tr>
</tbody>
</table>

The cleaner feed size distributions were similar for Runs 2, 3 and 4. Despite Run 2 not having a rougher concentrate regrind stage, the finer primary grind (routher feed) resulted in a similar cleaner feed size as Runs 3 and 4 where regrinding was used.
Rougher Flotation

Figure 1 confirmed the laboratory scale observations that the recovery of PGM’s from the tailings dams was highly dependent upon fineness of grind. At equivalent concentrate grade, PGM recovery was approximately 10 % higher with a primary grind of 90 % passing 75 µm, than with 80 % passing 75 µm.

![Figure 2: Rougher Flotation Grade /Recovery Curve](image)

Whilst the flotation kinetics of PGM’s are typically slow, it was significant that the rate of flotation increased with the finer grind. Mineralogical examination showed that the improved flotation performance at the finer grind was due to additional liberation of PGM’s as well as the removal of iron oxide layers from base metal sulphide surfaces by grinding in the inert media environment of the IsaMill.

![Figure 3 : Rougher Flotation Recovery /Time Curve](image)
Cleaner Flotation

It was anticipated that the use of a conventional cleaner/recleaner circuit would enhance the grade of the rougher concentrate with minimal loss in recovery. However, it was not possible to shift the grade/recovery curve for Run 1 (80% passing 75 µm primary grind) as shown in Figure 4. As shown in Table 3, the cleaner feed size distribution was the coarsest of all Runs (50.3 % passing 25 µm). Limited success came from additional laboratory investigation to identify methods of upgrading the rougher concentrate using alternative reagents, pH modification and novel chemistry techniques. However, mineralogical examination showed that poor liberation and iron oxide coating of base metal sulphide surfaces were responsible for the poor upgrade potential of the rougher concentrate. Regrinding of the concentrate was required to improve PGM liberation and freshen-up the heavily oxidised mineral surfaces.

Figure 4: Run 1 – Grade /Recovery for 80%-75µm Primary Grind

Figure 5: Run 2 – Grade /Recovery for 90%-75µm Primary Grind

Run 2 adopted a finer primary grind of 90% passing 75 µm. The benefit of this to rougher recovery has been discussed, but the additional liberation and removal of base metal sulphide
oxide surface coatings permitted significant upgrading in the cleaners. Figure 5 shows the shift of the grade/recovery curve.

Runs 3 and 4 used an IsaMill rougher concentrate regrind stage to test what improvements in cleaning could be gleaned from additional liberation and particle surface cleaning. With the coarse rougher concentrate size distribution produced in Run 1, it was expected that regrinding prior to cleaning would have the greatest benefit for this case. Figure 6 confirms this, with the grade/recovery curve shifted to the right of the rougher plot.

![Figure 6: Run 3 – Grade/Recovery After Rougher Concentrate Regrind at 80%-75 Primary Grind](image)

The cleaner feed size after regrinding in Run 3 was 83.4 % passing 25 µm. This compares well to the cleaner feed size in Run 2 (finer primary grind) of 85.4 % passing 25 µm. As Run 2 and Run 3 both involved a stage of inert media grinding in an IsaMill (clean mineral surfaces), and possess similar size distributions it was expected that the overall flotation performance of these two cases would be similar. Figure 1 shows Run 3 exhibiting a marginally better final concentrate grade and recovery, despite having a lower feed grade and slightly coarser cleaner feed size distribution than Run 4. It is hypothesised that the better performance of Run 3 was due to IsaMilling applied immediately before cleaning, rather than before roughing (to produce a finer primary grind) in Run 2. It is likely that the clean, reactive mineral surfaces produced by the inert IsaMill environment oxidise during the 45 minute rougher flotation period. Application of IsaMilling after roughing as performed in Run 3, produces fresh mineral surfaces which are likely to be responsible for better cleaner flotation performance.

Run 4 used IsaMilling to produce a primary grind of 90 % passing 75 µm, and for regrinding the rougher concentrate. The cleaner feed size of Run 2 (finer primary grind only) was 85.4 % passing 25 µm, compared to 88.1 % passing 25 µm in Run 4. Only a small increase in fineness was produced by regrinding the rougher concentrate, but the flotation performance improved, showing that the cleaning potential is sensitive to rougher concentrate particle size distribution and mineral surface condition.
Pilot Plant Test Conclusions

The following conclusions were made from the pilot plant tests:

- The PGM grade and recovery targets can be met with the use of conventional flotation and IsaMill inert grinding.
- A primary grind of no less than 80% passing 75 µm, and rougher concentrate regrind of no less than 85% passing 25 µm is required to meet the grade/recovery targets.
- A finer rougher concentrate regrind is likely to improve cleaner flotation – flotation is sensitive to grind size.

Most importantly, the combination of improved liberation and inert media grinding is required to maximise flotation potential. The impact of changes in mineral surface chemistry (such as oxidation from iron in grinding media) to flotation recovery is generally much larger than the effect due to liberation alone (refer to AMIRA P336, Report P336/26). There are examples where the actual change in recovery due to increased liberation from grinding is much lower than predicted due to the negative impact of mineral surface oxidation. It is even possible that the overall flotation recovery decreases after regrinding due to surface chemistry changes, despite increasing mineral liberation (Frew, Davey and Glen, 1994).

PLANT DESIGN

The Western Limb Tailings Retreatment Project (WLTRP) simplified flow sheet is shown in Figure 8. A staged approach was taken with the design and installation of the circuit, to limit the capital exposure and project risk to the novel treatment of dormant PGM tailings. The project was divided into two Phases, with the detailed design and installation of Phase 2 dependent upon the operating knowledge of Phase 1.
The Phase 1 WLTR concentrator was commissioned in the fourth quarter of 2003. The design is based on 400,000 tonnes per month (4.8 Mtpa), in a single grinding /flotation line. Much of the Phase 2 civil work was put in place during construction of Phase 1, making the addition of Phase 2 a relatively simple task. Phase 2 would basically add a duplicate grinding /flotation line to take capacity to 900,000 tonnes per month, or 10.8 Mtpa.

![Figure 8: Simplified WLTRP Flow Sheet](image)

The deslime cyclone under flow reports to the single 7.3m x 10.06m, 10.5 MW (14,000 hp) ball mill in closed circuit with cyclones. The cyclone over flow feeds rougher flotation, which comprises 9 x 130 m³ tank cells. Rougher tail reports to final tail. The first rougher concentrate is of smeltable grade, and is sent to final concentrate. The remaining rougher concentrate is thickened and then floated in a pre-cleaner prior to regrinding.

A single, open circuit, 2.6 MW (3,500 hp) M10,000 IsaMill grinds the rougher concentrate to a P₉₀ of 25µm using local silica sand as grinding media. The IsaMill operates in open circuit and produce a narrow product particle size distribution, because of the internal classification system this technology uses. The ground product is then sent to a two stage cleaning circuit where final concentrate is produced, and cleaner tailings are discarded as final tailings. The cleaner circuit comprises of 2 x 10 m³ pre-cleaners, 6 x 20 m³ cleaners and 4 x 10 m³ re-cleaners.

**IsaMill Design and Scale-Up**

The WLTRP design required grinding of a nominal 53 th⁻¹ and maximum 65 th⁻¹. Test work demonstrated that a reduction of F₈₀ = 75 µm to P₉₀ = 25 µm required 35 kWh⁻¹ using -5 +3 mm silica sand grinding media from a local quarry. As grinding media, the local sand was extremely low cost and demonstrated a consumption rate of 50 g/kWh (1.75 kg/t). The purpose of the scale-up project was to use only one IsaMill. Considering the maximum duty, a 2.6 MW IsaMill with
10,000 litre grinding chamber was designed. This would be the largest fine grinding mill available to the minerals processing industry.

The design represented a significant challenge, as the scale-up of absorbed grinding power was more than 2.5 times the existing design. Historically, the scale-up process of the IsaMill to the 1.1 MW (M3000) model was based on conservation of power intensity. At the 2.6 MW scale, the media agitator (grinding disc) tip velocity would be excessive using this method. A ‘constant tip velocity’ method was developed which proved more complex than power intensity models, as power does not scale-up linearly with volume.

A variable frequency drive was designed for the first M10,000 installation to reduce process risk and permit testing with fine (regrind duty) and coarse (secondary grinding duty) media types, important to Phase 2 of the WLTR Project.

Component and materials selection was critical, and importantly the disc surface abrasion rate was kept within 3 % of the 1.1 MW mill’s rate. As the M10,000 disc design was larger in diameter and thickness, the increased volume of rubber would mean a longer disc life than in the 1.1 MW mill, as abrasion rates are a function of area.

The larger diameter of the M10,000 required modifications to the design of the Product Separator; a centrifuging /pumping device that classifies the mill product (and retains grinding media). As centripetal acceleration decreases with increasing diameter (at constant tip speed), and pumping efficiency decreases with lower radial velocity, both the centrifuging and pumping actions of the Separator had to be re-designed. The new design focused on improving efficiency of the rotor suction region, rotor pump finger shape and product classification efficiency.

The disc tip velocity and power calculations have proven to be accurate. Figure 9 shows the correlation between calculated power draw, and observed power draw at different disc tip speeds during commissioning. Higher tip speed (and power draw) was not required, as the grinding efficiency was high and product size specification was met. Operation of the new Product Separator was successful above 76 % motor speed. Classification efficiency was poor at this speed, however the minimum operating design speed for the M10,000 was 20.5 ms⁻¹ or 81 % output. Figure 11 shows how the product particle size distribution of the IsaMill is narrower than that of the feed. This is due to the classification effect of the Product Separator. The IsaMill was operating at 80 % output, or 20 ms⁻¹ tip speed at the time of that survey.
<table>
<thead>
<tr>
<th>Tip Speed (m/s)</th>
<th>Calculated Power (kW)</th>
<th>Observed Power (kW)</th>
<th>Motor Output (%)</th>
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</thead>
<tbody>
<tr>
<td>19</td>
<td>1500</td>
<td>1400</td>
<td>76</td>
</tr>
<tr>
<td>20</td>
<td>1750</td>
<td>1700</td>
<td>80</td>
</tr>
<tr>
<td>21</td>
<td>2026</td>
<td>2000</td>
<td>84</td>
</tr>
<tr>
<td>22</td>
<td>2329</td>
<td>N/A</td>
<td>88</td>
</tr>
</tbody>
</table>

Power Correlation

\[ R^2 = 0.9992 \]

Figure 9: Power Scale-Up For M10,000 IsaMill

Figure 10: WLTRP Site And IsaMill
FINE GRINDING AND FLOTATION CIRCUIT OPERATION

Commissioning

The IsaMill circuit was commissioned late December 2003, and began operation in early February 2004 once steady rougher concentrate feed was available. Figure 11 shows the first size survey data across the mill during commissioning. At this time, the feed sizing was finer than design (low primary mill feed tonnage) and of lower pulp density (1.13 kg/L, 17 % solids). As the IsaMill can operate successfully with low density feed, the product specification could still be met ($P_{90} = 25 \, \mu m$). Notably, the product particle size distribution is significantly narrower than the feed, which is a function of the internal classification of this type of mill.

![Figure 11: IsaMill Size Distributions – Commissioning](image)

The cleaner flotation banks had already been commissioned, so a qualitative comparison of froth characteristics before and after IsaMilling could be made. Figure 12 shows the observed difference between the recleaner froths. Without regrinding, the froth was pale, watery and barren; stability was poor and mass pull was very low. After regrinding, the froth was dark grey and heavy in mineralisation. Mass pull was higher and froth stability better.
Current Operation

The main focus after the commissioning of the newly built WLTRP was the optimisation of the main stream circuit as this is the key to ensure optimum PGM recovery from the recovered tailings material. Optimisation of the IsaMill and cleaner flotation circuit will be in full swing when the first phase is completed.

Operational data generated from the IsaMill and cleaner circuit has revealed that the unit is producing a P₉₀ of 40µm compared to the design P₉₀ of 25µm. The optimisation of this circuit is currently in progress. The first obvious change in the circuit was the increase in mass pull from the rougher circuit from a design of 8% to a current mass pull of between 10 to 12%. This is a 50% increase in feed tonnage to the unit. Consequently, the energy input is 14.5 kWh/t compared to design of 35 kWh/t which explains the coarser grind. Increasing the power draw and specific energy of grinding will form part of the cleaner circuit optimisation program.

Figure 1 indicates the average of the feed and product size distributions taken over a week period. The IsaMill produced a reduction ration of 1.6 from a F₈₀ of 40µm to a P₈₀ of 25µm. The current feed to the mill is much finer than originally anticipated, so the use of a finer grinding media (to increase efficiency) will also be investigated.
Fine grinding is a new and relatively unexplored field in the platinum industry and optimisation of PGM flotation after fine grinding will be the key to success. Anglo Research Laboratories has initiated several projects to investigate the optimisation of the flotation parameters to ensure maximum benefit from the IsaMill installation.

CONCLUSIONS

- The use of IsaMill technology was enabling for the WLTR project, as it allowed smeltable concentrate grades to be produced from the oxidised, slow floating tailings.
- The collaborative IsaMill design project between Anglo Platinum and Xstrata was successful and is a sound model for technology development.
- Flotation kinetics increased after IsaMilling. The improved flotation performance was due to additional liberation plus the removal of iron oxide surface coatings by grinding in an inert media environment.
- Rougher concentrate IsaMilling was required for the WLTRP, to liberate PGM’s and clean-up heavily oxidised base metal sulphide surfaces.
- A staged approach was taken with the WLTRP design to limit capital exposure and minimise risk associated with the novel treatment of dormant PGM tailings.
- The design PGM grade and recovery targets were met using conventional flotation and IsaMill inert grinding technology.
- The IsaMill M10,000 scale-up (from 1.1MW to 2.6MW) was successful.
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Fine Grinding as Enabling Technology – The IsaMill

Pease, J.D. (1), Young, M.F. (2), Curry, D.C. (1)

(1) Xstrata Technology
(2) Xstrata - Mount Isa Mines

Introduction

The new generation of stirred mills like the IsaMill has fundamentally changed the economics of fine grinding. This has made them enabling technology for several existing and planned operations, and has opened new fields of processing in hydrometallurgy. These opportunities are made possible by the unique combination of features of stirred mills:

- Very high intensity attrition grinding mechanism, suited to fines grinding
- Small media size, essential to increase grinding efficiency for fines
- The use of inert grinding media. This can deliver dramatic improvements to flotation kinetics and recovery, and improved leaching leaching rates and chemistry.

Stirred milling was developed for fine grained ores that required an economic grind to sub 10 micron sizes. The first examples were lead zinc deposits – McArthur River, George Fisher and Mt Isa Blackstar orebodies enabled by the IsaMill, and Century which uses the stirred mill detritor (SMD).

The original application for ultra-fine grained orebodies is a relatively small niche, but it is now clear that there will also be applications in coarser grinding applications, particularly when power efficiency, space, and flotation surface chemistry are important. Two features specific to the IsaMill that make it attractive for coarser grinds are:

- the internal product classifier, which allows low cost open-circuit installations with a sharp product size
- the large unit size (currently up to 2.6 MW) suitable for large scale applications.

Installations at Lonmin Platinum and Anglo Platinum are examples where the IsaMill was chosen for coarser grind applications because of the flotation benefits of inert grinding. The case study of the Anglo Platinum tailings re-treatment plant shows that the 2.6MW mill was the enabling technology for the operation.

It is expected that the lower cost of fine grinding will also enable the economics of many leaching technologies operations, eg Activox, Albion Process. For example, in the Albion process atmospheric leaching of otherwise refractory minerals is feasible at fine sizes.
Stirred Milling Technology

Three features of stirred mills that transform the economics of fine grinding are:
- the high intensity attrition grinding environment
- the ability to use fine grained media (eg 1 mm) to suit to the fine grained feed
- the ability to use cheap natural products (local sand, slag, ore) as grinding media

These features distinguish stirred mills as fundamentally different from both ball mills and Tower Mills, as demonstrated by Tables 1 and 2.

### Table 1: Power Intensity of Different Grinding Devices

<table>
<thead>
<tr>
<th>Mill</th>
<th>Mill Diameter (m)</th>
<th>Mill Length (m)</th>
<th>Installed Power (kW)</th>
<th>Mill Volume m³</th>
<th>Power Intensity (kW/m³)</th>
</tr>
</thead>
<tbody>
<tr>
<td><em>Autogenous Mill</em></td>
<td>10</td>
<td>4.5</td>
<td>6400</td>
<td>353</td>
<td>18</td>
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<tr>
<td><em>Ball Mill</em></td>
<td>5</td>
<td>6.4</td>
<td>2600</td>
<td>126</td>
<td>21</td>
</tr>
<tr>
<td><em>Regrind Ball Mill</em></td>
<td>3.2</td>
<td>4.8</td>
<td>740</td>
<td>39</td>
<td>19</td>
</tr>
<tr>
<td><em>Tower Mill</em></td>
<td>2.5</td>
<td>2.5</td>
<td>520</td>
<td>12</td>
<td>42</td>
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<tr>
<td><em>IsaMill</em></td>
<td>1.3</td>
<td>3</td>
<td>1120</td>
<td>3</td>
<td>280</td>
</tr>
</tbody>
</table>

Table 1: Power Intensity of Different Grinding Devices

<table>
<thead>
<tr>
<th>Power Intensity (kW/m)</th>
<th>Media Size (mm)</th>
<th>No. Balls / m³</th>
<th>Surface Area (m)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ball Mill</td>
<td>20</td>
<td>20</td>
<td>120</td>
</tr>
<tr>
<td>Tower Mill</td>
<td>40</td>
<td>12</td>
<td>200</td>
</tr>
<tr>
<td>IsaMill</td>
<td>280</td>
<td>1</td>
<td>3600</td>
</tr>
</tbody>
</table>

Ball Mill is a 5.6m D x 6.4m L @ 2.6MW
Tower Mill is a 2.5m D x 2.5m L @ 520KW

Table 2: Mill Comparison of Media Size, Power Intensity, number of grinding media

The ability to use smaller media is probably the dominant impact on grinding efficiency. It dramatically increases the grinding surface area and the number of grinding “events”, essential to efficiently grind fine particles. Figure 1 shows the grinding power required to grind a sample of KCGM pyrite concentrate to different target P80 grind sizes, using a ball mill (with 9mm steel media) or an IsaMill with sand media. There is little difference at coarser sizes, but below 30 microns the advantage of stirred milling becomes dramatic. Ball milling simply cannot produce a 10 micron product at any practical power consumption. In this case the IsaMill has extended the economic range of grinding from about 20-30 microns to 10 microns – enabling technology if a 10 micron grind is needed, as it was for the KCGM cyanide leach.
Chemistry Impacts

The use of inert grinding media gives a crucial advantage to stirred milling in fine flotation and leaching applications. Even if it were economic to grind to 10 microns in a steel mill with very small balls, the amount of iron in solution would almost certainly ruin downstream flotation or leaching processes. The chemical impacts of steel grinding have been well reported (Trahar, 1984; Frew et al 1994; Greet & Steinier 2004), and compete with the benefits obtained from better liberation. Many plant metallurgists still believe that “slimes don’t float”, in spite of the fact that between them, Mt Isa, McArthur River and Century produce over 1.5 Million tonnes a year of concentrate below 10 microns, at high recovery. At Mt Isa recovery in the zinc cleaners is above 95% in all size fractions from 1 micron to 38 micron. At McArthur River, 96% of individual particles recovered are under 2.5 microns (Pease et al 2004).

While high-Chrome media can reduce the chemistry impact, the cost is higher and the impact for fines is only marginal compared with inert media.

Fine Grinding Before Leaching

Unlike flotation, leaching applications do not suffer as much from the same surface chemistry impacts from steel media. The use of steel media, however, can still be detrimental to a leaching process. When fine grinding pyritic concentrates of precious metals, it is common to follow the fine grinding stage with a pre-aeration stage to remove active pyrite and pyrrhotite before cyanidation. Worn steel media in the ground pyrite can significantly increase the pre-aeration time needed. In a recent application of the Isa mill, the existing regrind mill before gold leaching was consuming 10 t/day of steel balls. This reduced pulp Eh and extended residence time in subsequent pre-aeration and increased cyanide consumption in leaching.

Three mechanisms are important when fine grinding before leaching:

- **the liberation impact** – in simple cases the grinding is simply to expose fine grained minerals to leachant (eg exposing fine gold to cyanide). In this case dissolution of the host mineral is not needed.
- **the sizing impact** – “refractory” minerals often do react, but are passivated by reaction products forming a 2-3 micron “rim” on the particle. For a 30 micron particle this rim prevents the molecular transfer necessary to keep the reaction proceeding deeper into...
the particle. But for a 9 micron particle this rim is sufficient for the mineral to disintegrate (Figure 2).

- **The mechanical activation impact** - the high energy intensity of fine creates a highly stressed surface, reducing the crystalline nature to amorphous phases (Figure 3). The surface defects act as electron transfer sites, accelerating the rate of surface oxidation reactions, and lowering the activation energy required to oxidise the mineral. This effect of mechanical (or mechanochemical) activation of minerals is well reported (Balaz, 2000; Juhasz and Opoczky, 1990; Grelach et al 1989). It means that subsequent leaching of the minerals can take place under much less aggressive conditions, with a reduction in the capital cost of the leach plant.

![Unleached 9 micron particle](image1)

![Leached particle](image2)

**Figure 2**: The sizing impact of fine grinding. For a bigger particle, a 2.5 micron passive layer will prevent further leaching, but for a 9 micron particle it is sufficient for the mineral to be consumed.

![Chalcopyrite Concentrate prior to Grinding](image3)

![Chalcopyrite Concentrate Following Ultrafine Grinding](image4)

**Figure 3**: Impact of intense grinding on surface appearance of Chalcopyrite, the stressed and fractured surfaces on the right leach faster and with lower activation energy, even at the same size particles (from Balaz, 2000).
In practical situations all three effects of liberation, size reduction and surface activation occur together. Each increases leaching rate but it is difficult to distinguish the relative contributions. However the combined impact can be dramatic – eg in Xstrata’s Albion Process, the Isamill grinds and activates minerals to a point where bacteria or pressure are no longer required, and leaching can be carried out in simple open tanks. The extremely high power intensity in the Isamill compared with other grinding methods suggest it would enhance mechanical activation.

**Peculiarities of Fine Grinding – tips for new players**

Some aspects of fine grinding are not immediately intuitive to operators of conventional grinding. While there is nothing fundamentally different about small particles, some effects that are minor at coarser sizes become dominant at fine sizes. Some important tips for those designing fines circuits are:

- **beware the “knee” of the signature plot**: Figure 1 shows that stirred milling extends the practical range of grinding, but at some point the signature plot still goes “vertical” (the “knee” can be pushed finer by using smaller media). Sometimes clients tell us their target grind size is “about 8 or 10 microns” – but the power to get to 8 microns may be double the power to get to 10 microns.

- **The importance of consistent sizing technique**: this follows from the first point. A one micron difference between two sizing machines can change an estimated power draw by 50% ! To compare different grinding devices or media near the “knee” of the signature plot, it is essential that you use the same sizing machine (ideally operated by the same person), otherwise noise in the sizings will overwhelm the results.

- **The importance of classification**: every grinding operator knows that sharp classification is important for grinding efficiency. But this is difficult to achieve for ultrafine grinding. A sharp cut at 10 microns needs 2 inch cyclones – but no-one who has ever operated a cluster of 2 inch cyclones in a concentrator will want to do it again. As a result, operators usually choose bigger cyclones, but the operability comes at the expense of grinding efficiency. The solution offered by the Isamill is to classify within the mill by the centrifugal product separator, which produces a sharper cut than fine cyclones. It also eliminates the extra capital and operating cost of closed circuit cycloning.

- **Density and Viscosity impacts**: stirred mills operate at lower pulp densities than conventional mills. The efficiency of the Isamill is much less affected by density than conventional mills. While efficiency does generally still increase with feed density, the maximum density will be limited by viscosity, and viscosity effects are much more apparent for fine products. Though it is ore dependant, as a general guide sub 10 micron applications will be limited to about 45-50% feed solids.

**Power and Energy Efficiency**

Grinding energy is one of the major costs of mineral processing. Choosing the right grinding machine and the best media are certainly important. Some other important factors that are sometimes overlooked:

- Energy efficiency should be defined in terms of **power per unit product** recovered, not per tonne of ore. Well targeted grinding will improve recovery.

- The **energy usage of all production steps** should be considered. This includes energy in the blasting, mining, milling and smelting. Eg, higher concentrate grade will reduce smelting fuel and fluxing needs. It also includes the **energy content of grinding media** (eg steel balls versus local slag or sand), and flotation reagents (typically lower consumption after inert grinding).
• The circuit design should aim to apply the right grinding power in the right place, on the smallest possible stream, and avoiding circulating loads (Pease et al 2004; Young et al 1997). The circuit needs to be designed to the size-by-size ore mineralogy.

• Efficient classification. Grinding before flotation or leaching should narrow the size distribution by reducing the top size particles; it should not waste energy grinding already fine particles. A sharp grinding curve, characterised by a low ratio of P98 to P80 is vital. This is demonstrated by Figure 4.

As an example, in the George Fisher circuit 6 MW of additional grinding power was installed, but total energy efficiency was increased. This was because the inert milling increased flotation rates, increased recovery, dropped circulating loads (less pump and flotation energy and less spillage rehandling), and increased concentrate grade (less fuel in smelters). The circuit design of applying efficient regrinding to small streams meant that up-front grind size targets could be relaxed. The energy content of IsaMill media is free (granulated smelter slag). As a result, even with the extra grinding power total unit cost per tonne of ore did not increase, yet grades and recoveries increased significantly (Case Study 2).

Figures 4 and 5 demonstrate why the energy efficiency of the IsaMill is so high. Unlike conventional grinding, the size distribution of the IsaMill product sharpens with additional grinding. This unique behaviour is because:

• There is no short circuiting – particles have to pass through 8 consecutive grinding chambers, then pass the centripetal field of the product separator before leaving the mill.

• The low volume/high intensity means a short average residence time in the mill (typically 90 seconds). So a particle can travel through the 8 grinding chambers and “see” the product separator within 90 seconds. Fines will exit the mill, coarse particles will be retained.

Figure 4: Increasing grinding in an open circuit IsaMill sharpens the product size.
Recent Developments in Stirred Milling

Stirred milling to enable fine grinding and flotation operations is well established - currently 21 MW of IsaMills are installed worldwide, and have produced over 10 Mt of concentrates at Mt Isa and McArthur River alone. Because these applications are in the "niche" of fine grained lead zinc ores, it is easy to overlook the potential for conventional grinding. Recent developments in IsaMills bring some crucial advantages to more conventional grind size:

- improvements in design and materials for wear parts – eg "slip-in" shell liners. IsaMills routinely achieve availabilities over 97% at McArthur River and Lonmin Platinum (lower at Mt Isa since mills are frequently taken off line to conserve ore).
- Taking advantage of the internal classifier in the circuit design. In early installations (eg MRM) we took a "belts and braces" approach to classification, backing up the product separator with cyclones. In fact, cyclones reduce circuit performance, resulting in a flatter size distribution than produced by an open circuit IsaMill. The ideal IsaMill installation is shown in Figure 6, precyclone mill feed if necessary, but run the mill in open circuit.
- Developments in grinding media. The product separator allows cheap local grinding media to be used (there is no screen to block if some media degrades). For example, Mt Isa operates on waste granulated smelter slag, MRM ran for 6 years on autogenous ore chips, and Lonmin and Anglo use local sand at $US 0.08/tonne milled. Concurrently, there are new developments in manufactured media – higher cost, but low wear rates and much higher grinding efficiency in some applications. The availability of a standard product, with a choice of media size to suit the application, is important for the stirred mills to be accepted as a mainstream option.
- The successful scale up to the 2.6MW mill.

In combination, these developments mean that IsaMills may be a low cost, high efficiency alternative in some mainstream grinding applications. The low cost comes from simple installation – low footprint, low crane heights and loads, no need for closed circuit cyclone installations – an IsaMill installation is fundamentally different from a conventional grinding mill installation.
Figure 6: Recommended circuit configurations for IsaMilling, taking advantage of sharp internal classification.

Figure 7: The IsaMill range (left); ‘slip in’ rubber shell liner
Case Study 1: McArthur River Mining (MRM)

The McArthur River lead zinc deposit was the driving force behind the development of IsaMills. The orebody was discovered in 1955. It had a resource of 227 Mt at 9.2% Zn and 4.1% Pb, however no existing technology could economically treat the extremely fine grained minerals (Figure 8). The development of the IsaMill was truly enabling for this orebody. It allowed economic regrinding to 80% passing 7 microns, fine enough to reduce silica in bulk concentrate to marketable levels. Note that even at this size there is not adequate galena-sphalerite liberation to allow separate lead and zinc concentrates.

Figure 8: Different Grain Size of Broken Hill and McArthur River Ores (Grey Square is 40um)

The plant started mid 1995 with 4 IsaMills regrinding rougher concentrate. Media for the mills was provided by screening a fraction of ore gravel from the SAG mill discharge – a fully autogenous ultra-fine grind! Two more mills were installed to increase production and recovery (in 1998 and in 2001). In 2004 the media was changed from ore gravel to screened sand – the higher efficiency of the sand increased mill capacity, and reduced wear on mill components at the higher throughputs.

Table 3 shows production performance at MRM – very high concentrate grades and recoveries are achieved in spite of the ultrafine minerals. This disproves the view that “fines don’t float”. Consider the following perspective: a P80 of 7 micron means a P50 of 2.5 microns at MRM. While 50% of concentrate weight is finer than 2.5 microns, this means that 96% of individual particles recovered are less than 2.5 microns. Since flotation depends on individual particles attaching to bubbles, this means that 96% of the successful particle-bubble collisions at MRM happen for particles finer than 2.5 microns, into a high grade concentrate at high recovery. Fines float very well indeed after IsaMilling.
<table>
<thead>
<tr>
<th></th>
<th>MINING</th>
<th></th>
<th>METALLURGY</th>
<th></th>
<th></th>
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</thead>
<tbody>
<tr>
<td></td>
<td>Tonnes</td>
<td>Head Grade</td>
<td>Tonnes</td>
<td>Zn Recovery</td>
<td>Con Grade</td>
<td></td>
</tr>
<tr>
<td>1995/96</td>
<td>707,994</td>
<td>12.9%</td>
<td>759,519</td>
<td>66.4%</td>
<td>39.3%</td>
<td>11.2%</td>
</tr>
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<td>1996/97</td>
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<td>14.4%</td>
<td>1,026,150</td>
<td>73.5%</td>
<td>43.5%</td>
<td>11.0%</td>
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<td>1997/98</td>
<td>1,127,000</td>
<td>16.1%</td>
<td>1,139,000</td>
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<td>43.3%</td>
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<td>1998/99</td>
<td>1,222,238</td>
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<td>1,404,539</td>
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<td>46.8%</td>
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<tr>
<td>2002/03</td>
<td>1,505,306</td>
<td>12.7%</td>
<td>1,511,856</td>
<td>82.4%</td>
<td>46.6%</td>
<td>10.5%</td>
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<tr>
<td>2004</td>
<td>1,523,243</td>
<td>13.2%</td>
<td>1,579,762</td>
<td>80.1%</td>
<td>47.1%</td>
<td>10.4%</td>
</tr>
</tbody>
</table>

Table 3: Performance of McArthur River since commissioning
**Case Study 2: George Fisher Orebody**

The IsaMill technology for MRM was developed in the lead zinc concentrator at Mt Isa. It was clear that the technology would have benefits for the Mt Isa lead zinc orebodies, and it was to prove enabling for the George Fisher orebodies north of Mt Isa. While not as fine grained as MRM, components of George Fisher require a 7 micron grind to achieve acceptable concentrate grades and recoveries.

A circuit was designed to treat the mix of ores from George Fisher, Hilton, and Mt Isa lead zinc orebodies (Young et al, 2000). The circuit included eight 1 MW IsaMills, grinding lead rougher concentrate and intermediate zinc streams as shown in Figure 9. The principles of the circuit design were:

- **Only grind the minerals you have to** – this needs a thorough understanding of size by size mineralogical performance throughout the circuit. Recover what minerals you can at coarser sizes, then apply successively finer grinding and flotation stages to recover the finer grained minerals.
- **Float in narrow size distributions and tailor the flotation conditions to suit** – this is achieved in the staged grind and float circuit in Figure 9, with separate zinc recovery stages for 37 micron, 15 micron and 7 micron particles. A vital principle is to avoid circulating loads – if particles don’t float in the 37 micron circuit, don’t send them back to roughing, send them to regrinding and a custom designed circuit. The circuit may look more complex on paper, but in reality is much simpler to operate.

In fact the benefits of the inert grinding and the staged flotation design were so profound that it took us 6 months to appreciate them. Figure 10 shows the immediate 5% zinc recovery gain after installing. This was the gain due to extra liberation of sphalerite, and was all we expected. The “second wave” of even higher benefits happened when we realised that the clean surfaces from inert milling, and the staged flotation circuit, fundamentally changed mineral behaviour. In spite of the finer grind we didn’t need more reagent, we needed less. We didn’t need more flotation capacity, we needed less – fine minerals floated quite fast in conventional cells when they had clean surfaces and the right reagent conditions. The net impact of the circuit changes was:

- Lead recovery increased by 5% and lead concentrate grade increased by 5%
- Zinc recovery increased by 10% and zinc concentrate grade increased by 2% (in economic terms equivalent to 18% recovery increase at the same grade).
- Unit cost per tonne of ore was unchanged in spite of 6MW of extra grinding power.

Figure 11 demonstrates the combined effect of the staged grinding and cleaning approach – high zinc recovery (+95%) in all size fractions from 1 micron to 38 micron.
Baseline

Reduced grinding & flotation capacity, due to equipment relocation during construction.

1st Wave +5% Zinc Recovery

2nd Wave +5% Zinc Recovery

IsaMills Commissioned

+ 2% Conc Grade (not shown)

Figure 9: Mt Isa Pb/Zn Concentrator Flow Sheet

Figure 10: Zinc Recovery Increase from IsaMilling
Figure 11: Mt Isa Zinc Recovery from Rougher Concentrate by Size
Case Study 3: Mt Isa Black Star Open Cut

Surface resources at Mt Isa had long been a target for open cut mining. However the poor metallurgical response was always a barrier to production. Much of the ore is “transitional” between surface oxides and deeper primary sulphides. The transition ore is lower grade than primary ore, has fine grained mineralogy, and leaching has activated pyrite and sphalerite, leading to non-selective flotation. Constant attempts over the last 80 years failed to make the ore economic, with flotation unable to make smelter quality concentrates at any recovery.

The development of the IsaMills and the flowsheet to treat George Fisher ore changed this. The fine grinding achieves mineral liberation and cleans the mineral surfaces by attrition, and the combination of high intensity inert grinding and the correct water chemistry in flotation stops re-activation of unwanted minerals. The impact is shown by the grade recovery curve in Figure 12 - target concentrate grades can now be made at acceptable recoveries.

As a result, the IsaMills were the enabling technology that led to the approval of the Black Star Open Cut project at Mt Isa. Stripping commenced in 2004 and ore mining will commence in the first half of 2005, targeting 1.5M t/y to supplement underground production, produced from a mineral resource of 25Mt at 5.1%Zn and 2.7%Pb. This project represents only a small portion of the potential open cut resources at Mt Isa, the economics of which will also be reassessed once this project has been successful.

Figure 12: Pb Grade/Recovery Curve – ISA Lead-Zinc Transition Ore
Case Study 4: Merensky Platinum Tailings Retreatment Plant

During 2001 and 2002, Anglo Platinum assessed the retreatment of dormant tailings dams in the Rustenberg area in South Africa. These tailings represented a possible economic resource with the new grinding technology. Two processing issues were:

- The fine grained mineralisation of platinum (why it wasn’t recovered first time)
- Surface oxidation and oxidation products which harmed flotation – some of the tailings were placed over 100 years ago.

A collaborative project was undertaken by Anglo Platinum and Xstrata Technology to find an economic treatment. To achieve economies of scale for the project the IsaMill had to be scaled up from 1,000 kW to 2,600 kW.

The program was successful, and the large IsaMill proved to be the enabling technology for this project due to:

- The ability to grind fine at low cost – the mill operates in open circuit, and uses cheap local sand as the grinding media.
- The clean mineral surfaces resulting from the inert grinding environment. This was crucial to achieve target grades and recoveries after regrinding.

Figure 13 shows the improvement that IsaMill regrinding makes on cleaning rougher concentrate (Buys et al, 2004). The mill grinds rougher concentrate, thereby reducing power input compared with targeting a fine primary grind. It was found that this also improved cleaner flotation compared with fine primary grinding before roughing. It is likely that the clean surfaces produced in the IsaMill would re-oxidise during the 45 minute roughing period. In contrast, fine grinding immediately before cleaning increased flotation kinetics in cleaning (in contrast to the common observation that regrinding in a steel mill slows kinetics of all minerals).

Anglo Platinum commissioned the Western Limb Tailings Retreatment Plant in 2004. At the end of 2004 they concluded that (Buys et al, 2004):

- IsaMill technology was enabling for the WLTR project since it allowed acceptable concentrate grades to be made from oxidised slow floating tailings.
- Flotation kinetics improved after fine grinding due to both extra liberation and the removal of iron oxide surface coatings. Inert fine grinding of rougher concentrate was necessary.
- The scale up to the M10,000 IsaMill (from 1 MW to 2.6 MW) was successful.
Figure 13: Improvement in Platinum Grade/Recovery After IsaMilling for Western Limb Tailings Retreatment

Figure 14: Western Limb Tailings Retreatment Flowsheet
Case Study 5: Hydrometallurgical Processes and The Albion Process

The ability to efficiently grind minerals to 10 microns is an enabling step for several hydrometallurgical technologies. Fine grinding improves both kinetics and thermodynamics of leaching. The high surface area of fine particles gives high leaching rates at relatively low temperature and pressure, reducing capital and operating costs. Fine grinding also reduces the activation energy required to leach minerals. Several patented processes rely on fine leach feeds, e.g., the Activox process, the UBC/Anglo process (Driesinger and Marsh, 2002; Hourn and Halbe, 1999), the Phelps Dodge Process (Marsden and Brewer, 2003), and Xstrata’s Albion Process (Hourn and Halbe, 1999).

In these processes, metals are leached from a sulphide concentrate by oxidation. Oxidation is typically achieved using ferric iron or oxygen. Fine grinding facilitates the action of both ferric iron and oxygen, making the mineral easier to leach. Fine grinding also ensures that the mineral disintegrates before the leaching surface is passivated by the deposition of leach products.

Fine grinding can also help leach precious metal from sulphide concentrates where oxidation is not required. Preferential breakage of minerals along grain boundary fractures, where occluded gold and silver often accumulate, can significantly improve precious metals recovery.

The Albion Process is a graphic example of the powerful combination of fine grinding plus leaching. Minerals traditionally regarded as “refractory” leach easily at atmospheric pressure and temperatures below 100 degrees when ground finely. Xstrata has demonstrated this process for leaching of sphalerite, chalcopyrite, pyrite, arsenopyrite, stibnite, pentlandite, cobaltite and enargite.

Xstrata Technology’s 1 t/d Albion Process pilot plant recently operated for 20 months treating McArthur River zinc concentrate, achieving 98% Zn recovery in leaching at 18 - 24 hrs residence time. Leaching was carried out at atmospheric pressure and 80 - 90 degrees. The pilot produced over 30 tonnes of full scale SHG zinc cathodes. Xstrata Zinc is currently undertaking a feasibility study for a full scale plant using this technology.

Advantages of the Albion Process are:

- The ability to treat a lower grade copper or zinc concentrate than conventional roasting or smelting. This may allow a lower ore cut-off grade. For MRM, the ability to leach a low-grade concentrate would increase economically recoverable zinc from 3.6 million tonnes of recoverable zinc to 11.5 million tonnes.
- Compared with conventional zinc refining, the Albion Process avoids either the high pressure autoclave leach, or the roasting step before leaching. The capital cost savings are estimated at $1 000 per annual tonne of recovered zinc, with the roaster, acid plant, acid storage facilities and concentrate filtration and storage sheds eliminated.
- Either a Goethite or Jarosite stage can be used for iron control. Goethite is favoured since it precipitates as coarse particles, which are easy to filter and environmentally stable. Any arsenic present in concentrate is fixed in residue as ferric arsenate, a stable phase for tailings impoundment. This could prove enabling for high Arsenic copper ores that cannot be treating by primary smelting methods.
- The simplicity of operating a grinding mill and rubber lined atmospheric tanks means that the Albion process can suit small operations, e.g., refractory gold operations in remote sites. Such operations would be unlikely to bear the cost and complexity of high pressure or bacterial leaching processes. The rapid leach kinetics typical of Albion Process
residues in cyanide leaching can mean that the Albion leach circuit can be retrofit using existing tanks. In such cases, the only substantial new equipment may be the Isamills.

The mechanisms of leaching fine particles (particle size and surface stress) highlight the importance of the grinding and classification circuit. A high intensity environment like the IsaMill would be expected to create highest surface activation. It is hard to quantify this effect separately, because the IsaMill also produces a much sharper size distribution than other mills, also contributing to higher leaching rate. This impact has not been quantified, because it is hard to distinguish it from the sizing effect. Figure *** above demonstrates the crucial importance of P98 in leaching. A 30 micron particle simply may not leach after the 3 micron passive layer has formed – so a P98 of 30 microns may give 2% lower recovery than a P98 of 19 microns, even for the same P80. Similarly, if the objective of grinding is simply to expose finely disseminated gold to a cyanide leach, then gold disseminated in 30 micron particles will be lost.

Figure 15 illustrates the different size distributions of two large scale industrial applications – IsaMills at McArthur River, and detritors at Century (Reemeyer, 2004). Both operations produce a similar P80, but the IsaMill produce a much finer P98. In flotation this difference is not dramatic, but it is for leaching. There is a vital conclusion for project economics – achieving a sharper size distribution (lower P98/P80 ratio) may mean that the same recovery can be achieved at a coarser grind size (measured by P80). In ultrafine grinding, this may mean a 30% lower power requirement (see Figure 1), or alternatively higher recovery at the same power consumption. This highlights the crucial importance of sharp classification in circuits before leaching. The sharp size distribution from open-circuit IsaMills is because each particle has to first pass through 8 consecutive grinding segments, then has to escape the centrifugal forces of the product separator before leaving the mill.

Figure 15: Comparison of Sizings from MRM (IsaMill) and Century (SMD)

This effect has been demonstrated frequently in laboratory scale leaching. Table 4 and Figures 16 and 17 show the effect of different grinding technologies on sizing and copper recovery for chalcopyrite leached in the Albion Process. For the same P80, copper recovery varied by 3 % for different grinding machines, due to a combination of different P98 and different surface condition. A measure of the sharpness of the sizing curve is the ratio P98/P80. A value closer to unity means a sharper size distribution, and a higher leach recovery.
Table 4 and Figure 16:
Particle Size Distributions for Chalcopyrite Concentrate with Varying Grinding Mills (laboratory grinding with different mills, with the same feed and same sizing device)

<table>
<thead>
<tr>
<th>% Passing - microns</th>
<th>IsaMill</th>
<th>ECC Mill</th>
<th>GK Vibratory Mill</th>
<th>Metprotech Mill</th>
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</thead>
<tbody>
<tr>
<td>98</td>
<td>23.1</td>
<td>34.4</td>
<td>42.8</td>
<td>51.90</td>
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<td>95</td>
<td>17.44</td>
<td>26.33</td>
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<td>90</td>
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<td>1.42</td>
<td>1.32</td>
<td>1.51</td>
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</table>
Figure 17: P98/80 Ratio vs Leached Copper Extraction (Copper Bulk Concentrate)

Of course, other grinding mechanisms can also produce a sharper cut by placing mills in series to minimise short circuiting, and/or by closed circuit cycloning of the mill discharge. But closed circuit cycloning is expensive and difficult, requiring a cluster of small cyclones to cut sharply at fine sizes. Further, a sharp cut in closed-circuit cyclones usually requires water dilution of cyclone feed – which may then necessitate a cost intensive dewatering stage before leaching. In this case the IsaMill can produce a sharp size distribution with two less unit processes, at lower capital cost, lower operating cost, and better energy efficiency.
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Improving The Efficiency Of Fine Grinding – Developments In Ceramic Media Technology

Curry D C¹; Clermont B²
¹Technical Superintendent - Mineral Processing; Xstrata Technology
²R&D Project Manager – Comminution; Magotteaux International

Abstract. The use of ceramic beads as grinding media within high intensity stirred mills (such as the IsaMill) is desirable to maximise the energy efficiency of these processing units. Using a ceramic media with properties tailored to high intensity stirred milling further increases energy efficiency and extends the practical operating range of these mills to coarser feed and product sizes. Historically, the economics of using ceramic media types in stirred mills in the minerals industry has not been attractive. Magotteaux International and Xstrata Technology cooperated in the testing and product development of a new ceramic media, known as Keramax MT1, designed specifically for the minerals industry. Comparative media consumption tests are described and show how the low wear rate and high efficiency of this new ceramic offers an economic alternative to existing media types. The paper proposes the sliding friction coefficient as a new tool in characterising stirred milling grinding media. The first application of this ceramic media will be regrinding a gold bearing, pyrite rich sulphide flotation concentrate in the industry’s largest stirred mill; the M10,000 IsaMill.

1 INTRODUCTION

The high energy efficiency of stirred mills compared to ball mills is well understood within the industry. The use of tower mills as an energy efficient alternative to secondary and regrind ball milling became a standard flow sheet inclusion in the latter part of the previous century and is still common today. The modern high intensity stirred mills (such as the IsaMill) further extend the energy benefits of this technology by using higher agitation speeds and smaller media particles [1].

Media selection has a major influence on mill parameters such as energy efficiency, internal wear and operating costs. An inert grinding environment is beneficial to avoid mineral surface degradation and obtain downstream processing and cost advantages [2]. Ceramic media has a profound implication to these parameters and the availability of an economic ceramic media could give significant benefits to the users of IsaMills.

1.1 Grinding Media Types

To date, all IsaMill installations have taken advantage of the technology’s ability to use a low cost, but relatively low quality grinding media such as silica sand, river pebble, smelter slag or fine primary mill scats (autogenous milling). Whilst the IsaMill produces high energy efficiencies compared to conventional milling when using these media types, these ‘naturally occurring’ materials handicap the technology in several ways. The energy efficiency is low compared to what is possible with higher quality media, such as ceramic based compounds. The angular shape and small grain size of the natural media types limit the size of media, and therefore size of feed that can be milled. For example, sand media typically breaks down to its natural grain size when exposed to the high intensity milling environment. Generally sands have grain sizes finer than 5 mm. This limits the feed size that the mill can treat. From a mill wear perspective, it is preferable to use a higher SG media to increase media forces rather than larger, low SG media. The ideal media type for high intensity stirred mills has consistent, reproducible characteristics as shown below, and is further detailed by Lichter and Davey [3]:

- Definite initial charge PSD and top up size
- Chemical composition
- Hardness (related to chemical composition and grain size)
- High sphericity
- High roundness
- Competency (mechanical integrity)
- SG (as designed for machine operation /ore breakage requirements)

A media type that can be manufactured to the ideal qualities is therefore desirable. This reduces power costs and extends the benefits of the technology to treatment of coarser feed sizes. Despite this, the use of manufactured media such as ceramic beads has generally been uneconomic, as the combination of low consumption rate, high energy efficiency and low unit cost have not converged. This has limited the application of such energy efficient grinding technology such as the IsaMill, and restricted its application to regrind and ultra fine milling only.

1.2 Keramax MT1 Development

Magotteaux International has developed a ceramic grinding media specifically applicable to high intensity stirred milling in the minerals
industry. In cooperation with Xstrata Technology, the performance and cost effectiveness of the Keramax MT1 grinding media has been tested and verified using laboratory, pilot and full scale IsaMills. The first industrial application combining Keramax MT1 and IsaMill technology will be commissioned late 2005.

2 CERAMIC GRINDING MEDIA OVERVIEW

Two manufacturing processes may be distinguished in the production of ceramic media commonly used in fine grinding for non-contaminating applications:

- Sintered ceramic beads obtained by a cold forming of ceramic powder and by firing in high temperature kilns.
- “Fused” ceramic formed by electric fusion of oxides. The majority of these ceramic beads are named “zirconium silicate”.

Ceramic beads are usually classified according to their chemical composition and physical properties such as bulk density, hardness and fracture toughness. The bulk density has a large influence on the mill power draw. Hardness and fracture toughness could give an indication of the bead’s wear resistance.

The zirconium oxide beads are the highest quality grinding media with the highest initial cost. Keramax MT1 beads have been developed as an economic ceramic media for the mineral processing industry.

2.1 Keramax MT1 Properties

Keramax MT1 is a high density alumina grinding media with consistent microstructure to provide high resistance to wear and high energy efficiency. The media features a smooth bead surface which is ‘pearl’ like to touch. The surface properties suggest that the energy loss in grinding due to friction could be far lower than with other media and that the life time of the internal mill wear components will be increased.

Some low unit cost alumina media types have previously been tested in the IsaMill; both at laboratory and full scale. The consumption rate of these media types was very high (higher than silica sand) due to inconsistencies throughout the bead cross section. Some beads were very soft in the centre and others had air inclusions. The microstructure of MT1 is consistent throughout the cross section.

Table 1: Summary of ceramic bead compounds

<table>
<thead>
<tr>
<th>Ceramic Beads</th>
<th>Chemical Composition</th>
</tr>
</thead>
<tbody>
<tr>
<td>Alumina</td>
<td>Al₂O₃ ≥ 85% - SiO₂</td>
</tr>
<tr>
<td>Yttrium stabilized zirconium oxide</td>
<td>ZrO₂ (95%) - Y₂O₃ (5%)</td>
</tr>
<tr>
<td>Cerium stabilized zirconium oxide</td>
<td>ZrO₂ (80%) - CeO₂ (20%)</td>
</tr>
<tr>
<td>Magnesium stabilized zirconium oxide</td>
<td>ZrO₂ (97%) - MgO (3%)</td>
</tr>
<tr>
<td>Zirconium silicate</td>
<td>ZrO₂ (69%) - SiO₂ (31%)</td>
</tr>
<tr>
<td>Aluminum silicate</td>
<td>Al₂O₃ (34%) - SiO₂ (62%)</td>
</tr>
</tbody>
</table>

Table 2: Summary of ceramic bead physical properties

<table>
<thead>
<tr>
<th>Ceramic Beads</th>
<th>Bulk Density</th>
<th>Hardness (HV)</th>
<th>Fracture Toughness (K1C)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Alumina</td>
<td>2.0 – 2.1</td>
<td>1500-1700</td>
<td>3 - 5</td>
</tr>
<tr>
<td>Yttrium stabilized zirconium oxide</td>
<td>3.6 – 3.8</td>
<td>1300-1400</td>
<td>13</td>
</tr>
<tr>
<td>Cerium stabilized zirconium oxide</td>
<td>3.9 – 4.0</td>
<td>1100-1200</td>
<td>13</td>
</tr>
<tr>
<td>Magnesium stabilized zirconium oxide</td>
<td>3.2 – 3.4</td>
<td>900-1100</td>
<td>6</td>
</tr>
<tr>
<td>Zirconium silicate</td>
<td>2.2 – 2.4</td>
<td>600-800</td>
<td>3</td>
</tr>
<tr>
<td>Aluminum silicate</td>
<td>1.7 – 1.8</td>
<td>800-900</td>
<td>3</td>
</tr>
</tbody>
</table>

Table 3: Keramax MT1 compound

<table>
<thead>
<tr>
<th>Ceramic Bead</th>
<th>Chemical Composition</th>
</tr>
</thead>
<tbody>
<tr>
<td>Keramax MT1</td>
<td>Al₂O₃ (79%) – SiO₂ (6,5%) – ZrO₂ (14%)</td>
</tr>
</tbody>
</table>

Table 4: Keramax MT1 physical properties

<table>
<thead>
<tr>
<th>Ceramic Bead</th>
<th>Bulk Density</th>
<th>Hardness (HV)</th>
<th>Fracture Toughness (K1C)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Keramax MT1</td>
<td>2.3 – 2.4</td>
<td>1300-1400</td>
<td>5 - 6</td>
</tr>
</tbody>
</table>
Table 5: Keramax MT1 cross sectional hardness

<table>
<thead>
<tr>
<th>Location</th>
<th>1</th>
<th>2</th>
<th>3</th>
<th>4</th>
<th>5</th>
<th>6</th>
<th>7</th>
<th>8</th>
<th>9</th>
<th>10</th>
<th>Ave</th>
<th>SD</th>
</tr>
</thead>
<tbody>
<tr>
<td>HV</td>
<td>1308</td>
<td>1301</td>
<td>1308</td>
<td>1351</td>
<td>1322</td>
<td>1294</td>
<td>1366</td>
<td>1322</td>
<td>1396</td>
<td>1294</td>
<td>1326</td>
<td>34</td>
</tr>
</tbody>
</table>

Figure 1: Keramax MT1 cross section – hardness measurement points 1-10, left to right

3 SURFACE PROPERTIES OF GRINDING MEDIA

Grinding media selection is usually focused around the parameters listed in 1.1. In testing of MT1 during product development, it was found that grinding efficiency was greater than other media types of similar size, shape factors and SG. In fact, grinding efficiency was better than could be predicted with stress intensity calculations. Of significance was the power trend that occurred during test work under the conditions of media-water and media-slurry. In all other media types ever tested by Xstrata Technology (including ceramics), an IsaMill will draw more power when operating with a media-water system than with a media-slurry system (ie motor power would always decrease when slurry was introduced to a mill operating with grinding media and water – in a start up situation for example). With MT1, the reverse occurs. It was hypothesized that slurry ‘lubricates’ a typical media charge as the frictional loss between media particles in motion is lower in the presence of fine feed slurry than media on media interaction in the presence of only water. Further to this, and considering the unique reverse trend with MT1, it was proposed that the media on media interaction of MT1 in a water environment resulted in minimal frictional loss, compared to what now would be considered an abrasive slurry environment with slurry being fed to the mill.

The appearance of the MT1 surface and observing how easily the media flows certainly supported the theory of lower frictional loss. Xstrata’s Discrete Element Method (DEM) model was used to verify the sensitivity to energy distribution and power draw of surface friction coefficients.

3.1 Media Surface Sliding Friction Coefficient

Surface rolling friction ($\mu_r$) does not have a measurable effect on energy distribution, shaft torque or media velocity and force, so was fixed at $\mu_r = 0.005$ which is consistent with other DEM simulation work [4].

Two sliding friction coefficients ($\mu_s$) were used in this simulation to illustrate the difference between media of mid and low $\mu_s$. These were $\mu_s = 0.01$ and 0.30 which compare well to measured $\mu_s$ of < 0.10 for MT1 and > 0.30 for silica sand and river pebble. All other media parameters were held constant.

Figure 2 shows the steady state flow patterns of media at a cross section near a disc surface, located at the shaft mid-point along the axial direction of the IsaMill. The colour represents the velocity of particles. Figure 3 shows the steady-state force distribution of the same media with colour representing total forces acting on media particles. The simulations use a mono-sized media charge of 4 mm diameter, in a 4 litre chamber and ignore the effect of slurry, so physical interactions are solely a function of media specification.
The velocity in both cases is highest near the disc hole meaning that the disc surface properties and holes are responsible for ‘lift’ of the media. Media with low sliding friction move to the periphery and have higher velocity. Media with high sliding friction stay closer to the shaft and possess lower velocity. The transfer of energy between the disc and media for particles with low sliding friction is more efficient.

Particles with low sliding friction have much larger forces than those with high sliding friction. Grinding should consume less energy when media types having low sliding friction coefficients are used.

Figure 4 shows the resultant torque acting on the IsaMill shaft for the two different media types. The IsaMill took longer to reach steady state with the low $\mu_s$ media but the stable torque was 14,000 torque units compared to 21,000 torque units with the higher $\mu_s$ media. In this case, the IsaMill operating with the higher $\mu_s$ would consume 50% more power than the lower $\mu_s$ case. Again, this demonstrates that an IsaMill operating with a low $\mu_s$ media charge would have better energy efficiency.
4 TEST WORK

4.1 Laboratory

The aim of laboratory tests was to obtain comparative results of different grinding media in terms of relative consumption rate and grinding efficiency. Combining the energy and consumption data produces a kg/t consumption figure for each media type. This would present the first evaluation of the economic potential of MT1 in IsaMilling applications.

Tests were performed in Netzsch LME4 (IsaMill M4) machines at both Magotteaux’s Belgian facility and Xstrata’s Brisbane laboratory. This section describes test work performed on a gold bearing pyrite concentrate from the Eastern Gold Fields region of Western Australia.

The simplified test procedure is described below:

- Media is pre-conditioned in water before testing on slurry to simulate a conditioned charge. Media is dried, weighed and re-loaded in the mill. (Note the MT1 media charge did not lose any mass after 60 minutes of grinding in water).
- Pyrite concentrate slurry with a F80 = 170 µm is pumped through the LME4 in a single pass of ± 10 minutes. Energy, flow rate and pulp density are measured during this time.
- A sample is taken for particle size analysis using a Malvern Microsizer.
- This process is repeated a minimum of three times to produce size / energy data pairs.
- The test rig is then placed in closed circuit with the slurry stock, and operated for a further 60 minutes to maximise the accuracy of energy measurement and media consumption.
- The loss in mass is measured at the end of the test. Dividing this mass by total net energy, gives a g/kWh media consumption rate.
- A size value (eg P80) is plotted against its respective energy value and a measurement of grinding efficiency is obtained (a Signature Plot).

The consumption rates of different media types on the pyritic concentrate, under identical grinding conditions are shown in Table 6. Keramax MT1 exhibited the lowest consumption rate, giving a relative wear ratio of 1. The relative wear ratios of the other media types are shown in the right hand column.

The size distribution of the media types (with exception of the nickel slag) were not representative of fully conditioned charges; most media types only had particles in the -4 +3 mm range as this was how they could be ordered from the suppliers. The absence of -3 mm media would lead to inefficient grinding to ultra fine sizes. Because of this, a coarser target product size was selected where all media types could produce a product of acceptable efficiency. A P80 = 47 µm was selected, as it met the above criteria for the narrowly sized media charges, and also presented as the only data point produced for Alumina 2 (due to a pump failure during testing, more data points were not produced for this media type). Therefore, the energy efficiency of all media types are compared at a P80 = 47 µm and shown in Table 7. All tests were performed in open circuit.

The next best media to MT1 required 50% additional energy to produce the same product size. Significantly, this media was an Australian silica sand which required less energy than the two alumina ceramic media types. The silica sand was however, coarser than either of the alumina media which would result in higher wear rates of internal mill components, so maintenance costs would need to be evaluated.

Using the available media consumption and energy data, the media consumption per tonne of concentrate treated can be calculated and is shown in Table 8. It is clear that MT1 has economic potential in IsaMill applications, as the consumption per tonne milled in this test is at least an order of magnitude less than the next best media. The higher energy efficiency also has capital cost benefits, as a lower installed power can be selected to optimise project economics. For example, Table 9 shows the net power requirement for this concentrate type to treat an arbitrary 250 t/h. Once a specific project's power cost is factored in, the relative performance of MT1 further improves. Based on this outcome, further work at pilot scale was planned.
Table 6: Grinding media consumption rates - per unit energy

<table>
<thead>
<tr>
<th>Media Type</th>
<th>Consumption Rate (g/kWh)</th>
<th>Relative Consumption</th>
</tr>
</thead>
<tbody>
<tr>
<td>MT1 (-4 +3 mm)</td>
<td>15</td>
<td>1.0</td>
</tr>
<tr>
<td>Alumina 1 (-4 +3 mm)</td>
<td>128</td>
<td>8.5</td>
</tr>
<tr>
<td>Alumina 2 (-4 +3 mm)</td>
<td>295</td>
<td>19.7</td>
</tr>
<tr>
<td>Australian River Pebble (-4 +3 mm)</td>
<td>200</td>
<td>13.3</td>
</tr>
<tr>
<td>Australian Silica Sand (-6 +3 mm)</td>
<td>781</td>
<td>52.1</td>
</tr>
<tr>
<td>Ni Slag (-4 +1 mm)</td>
<td>1305</td>
<td>87.0</td>
</tr>
</tbody>
</table>

Table 7: Grinding media energy efficiency from F<sub>80</sub> = 170 µm to P<sub>80</sub> = 47 µm

<table>
<thead>
<tr>
<th>Media Type</th>
<th>Specific Energy (kWh/t)</th>
<th>Relative Energy</th>
</tr>
</thead>
<tbody>
<tr>
<td>MT1 (-4 +3 mm)</td>
<td>7.6</td>
<td>1.0</td>
</tr>
<tr>
<td>Alumina 1 (-4 +3 mm)</td>
<td>13.1</td>
<td>1.7</td>
</tr>
<tr>
<td>Alumina 2 (-4 +3 mm)</td>
<td>12.4</td>
<td>1.6</td>
</tr>
<tr>
<td>Australian River Pebble (-4 +3 mm)</td>
<td>27.9</td>
<td>3.7</td>
</tr>
<tr>
<td>Australian Silica Sand (-6 +3 mm)</td>
<td>11.2</td>
<td>1.5</td>
</tr>
<tr>
<td>Ni Slag (-4 +1 mm)</td>
<td>17.8</td>
<td>2.3</td>
</tr>
</tbody>
</table>

Table 8: Grinding media consumption rates - per tonne treated

<table>
<thead>
<tr>
<th>Media Type</th>
<th>Consumption Rate (kg/t)</th>
<th>Relative Consumption</th>
</tr>
</thead>
<tbody>
<tr>
<td>MT1 (-4 +3 mm)</td>
<td>0.11</td>
<td>1</td>
</tr>
<tr>
<td>Alumina 1 (-4 +3 mm)</td>
<td>1.68</td>
<td>15</td>
</tr>
<tr>
<td>Alumina 2 (-4 +3 mm)</td>
<td>3.66</td>
<td>32</td>
</tr>
<tr>
<td>Australian River Pebble (-4 +3 mm)</td>
<td>5.58</td>
<td>49</td>
</tr>
<tr>
<td>Australian Silica Sand (-6 +3 mm)</td>
<td>8.77</td>
<td>77</td>
</tr>
<tr>
<td>Ni Slag (-4 +1 mm)</td>
<td>23.23</td>
<td>203</td>
</tr>
</tbody>
</table>

Table 9: IsaMill net power requirement – illustration at 250 t/h treatment rate

<table>
<thead>
<tr>
<th>Media Type</th>
<th>Net Power (MW)</th>
</tr>
</thead>
<tbody>
<tr>
<td>MT1 (-4 +3 mm)</td>
<td>1.9</td>
</tr>
<tr>
<td>Alumina 1 (-4 +3 mm)</td>
<td>3.3</td>
</tr>
<tr>
<td>Alumina 2 (-4 +3 mm)</td>
<td>3.1</td>
</tr>
<tr>
<td>Australian River Pebble (-4 +3 mm)</td>
<td>7.0</td>
</tr>
<tr>
<td>Australian Silica Sand (-6 +3 mm)</td>
<td>2.8</td>
</tr>
<tr>
<td>Ni Slag (-4 +1 mm)</td>
<td>4.4</td>
</tr>
</tbody>
</table>
4.2 Pilot

Pilot testing was conducted at a South African platinum concentrator treating MF2 tail (i.e., secondary ball mill/flotation circuit tailing) on UG2 ore in the Western Limb of the Bushveld Igneous Complex near Rustenburg. This site was selected because a full scale IsaMill is already operating in cleaner circuit regrind on silica sand which would permit full scale verification of the pilot results. Also, the M100 pilot IsaMill (55 kW) was located near this site which is an ideal mill for this work. Four media types were selected, however only two were able to grind the extremely hard chromite in the plant tail. The two media types that failed to grind the feed were a silica sand and alumina ceramic (both having SG = 2.6 t/m³). Neither of these media types are available in coarse enough sizes to break the chromite in the feed. ‘Ceramic A’ was of similar SG to MT1.

The simplified test procedure is described below:

- Equivalent volume of media is loaded into the IsaMill.
- The power draw is recorded, and maintained by pumping fresh media in with the feed slurry.
- Media consumption is calculated by recording the mass added during the test and by measuring the mass variation of the load before and after each test.
- Composite samples are taken every day for particle size analysis.
- All the operating conditions such as slurry flow rate, pulp density and mill speed are identical for each test.
- Each media type was tested for 5 days to generate sufficient data.

4.3 Full Scale

At the time of writing this paper, two full scale IsaMill tests with MT1 at different sites were underway. The sites agreed to perform full scale testing based on the encouraging results from the laboratory and pilot test work. It is planned to publish the test results in the future, however the following conclusions can be made:

- The IsaMill was running in a continuous mode in a single pass operation.

Figure 5 plots the operating points of the IsaMill with the two media types during the test (specific energy against % passing 25 µm). Table 10 summarises the consumption and specific energy data per % -25 µm produced.

Ceramic A required 46% more energy to produce the same product size, whilst consuming over 400% more media in the process.

During the MT1 test, some media was crushed in the gland seal area of the mill which had worn during the Ceramic A test. The seal arrangement was subsequently modified to a similar design as the full scale mills. However, it was suspected that the observed 20 g/kWh consumption rate was high. This could be confirmed during full scale tests.
Table 10: Specific energy and media consumption rate

<table>
<thead>
<tr>
<th>Media Type</th>
<th>Specific Energy (Per % -25 µm)</th>
<th>Consumption Rate (g/kWh)</th>
</tr>
</thead>
<tbody>
<tr>
<td>MT1 (-3.5 +1.8 mm)</td>
<td>74</td>
<td>20</td>
</tr>
<tr>
<td>Ceramic A (-3.4 +1.7 mm)</td>
<td>108</td>
<td>81</td>
</tr>
</tbody>
</table>

Table 11: Kumtor – relative consumption and motor power draw @ P₈₀ = 12.5 µm

<table>
<thead>
<tr>
<th>Media Type</th>
<th>Relative Consumption</th>
<th>Relative Motor Power</th>
</tr>
</thead>
<tbody>
<tr>
<td>MT1 (-2.2 +1.8 mm)</td>
<td>1.0</td>
<td>1.0</td>
</tr>
<tr>
<td>Ceramic B (-2.4 +1.4 mm)</td>
<td>4.6</td>
<td>1.3</td>
</tr>
<tr>
<td>Australian River Sand (-2.4 +1.2 mm)</td>
<td>9.2</td>
<td>2.2</td>
</tr>
</tbody>
</table>

5 FIRST INDUSTRIAL APPLICATION

The first application of MT1 and large scale IsaMilling is at Centerra Gold’s Kumtor operation in the Kyrgyz Republic. Kumtor is the largest gold operation in Central Asia producing over 500,000 oz pa. MT1 media and the large M10,000 IsaMill enabled the highest energy efficiency possible on a small foot print. Due to the variation in (pyritic) rougher concentrate mass pull, a high intensity stirred mill with good turn down was required; the IsaMill is designed to operate between 1.3 and 2.6 MW power draw by varying the amount of media inside the mill.

The IsaMill will process an existing regrind mill product to ultra fine sizes of P₈₀ < 12.5 µm. Several media types were evaluated, including ceramics and silica sand. The lower energy efficiency of silica sand would have required the additional capital cost of a second mill. The media evaluation is summarised in Table 11.

6 CONCLUSION

Historically ceramic type grinding media has not been economically viable for mineral processing applications with high speed stirred mills such as the IsaMill, due to high operating costs. The availability of an economic ceramic media will provide considerable benefits to the users of IsaMills, and allow the energy benefits of the technology to be applied at coarser grind sizes.

Keramax MT1 ceramic grinding media was tested on various mineral concentrates and offers very high energy efficiency and extremely low consumption rates. The surface properties of MT1 contribute to the high energy efficiency of IsaMilling with this media type. DEM simulations explain the more efficient distribution of energy when using a grinding media with low sliding friction coefficient.

MT1 offers the design metallurgist a way of lowering capital and operating costs for large scale stirred milling projects. The footprint of IsaMill installations can be decreased because of the lower installed power requirement and smaller media handling system. The Kumtor gold project will be the first to demonstrate the effectiveness of combining a high efficiency grinding media with large scale stirred milling.

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Collaborative Technology Development –
Design And Operation Of The World’s Largest Stirred Mill

Curry D C¹, Clark L W² and Rule C³

¹Technical Superintendent - Mineral Processing; Xstrata Technology
²Business Manager - Mineral Processing; Xstrata Technology
³Head of Concentrator Technology; Anglo Platinum Limited

Abstract. Since late 2003, Anglo Platinum has been operating the world’s largest stirred mill at their Western Limb Tailings Retreatment Project near Rustenburg in South Africa. The mill is a 2.6 MW, M10,000 IsaMill which was a development of the 1.1 MW M3,000 machine. Only 16 months was required from project kick-off to commissioning – a considerable undertaking in view of the large scale-up step. A novel collaborative development approach was taken between Anglo Platinum and Xstrata as a vehicle for robust technology advancement.

The IsaMill has operated successfully since commissioning in terms of product size, scale-up and mechanical availability. The M10,000 used new designs to reduce internal component wear rates compared to previous IsaMill designs and the relative operating costs of the large IsaMill models (M3,000 and M10,000) are discussed. Laboratory, pilot and full scale operating data is used to validate the success of this installation.

1 INTRODUCTION
Anglo Platinum commissioned the first large scale tailings retreatment facility in the South African platinum industry in December 2003. The facility (known as the Western Limb Tailings Retreatment Project – WLTRP) uses modern concentrator technology to recover PGM’s from dormant tailings dams.

Pilot test work demonstrated the need for inert media regrinding of rougher concentrates before cleaning, to produce a smeltable concentrate grade. IsaMill technology was selected to regrind rougher concentrate to 90 per cent passing 25 µm. Due to the scale of the WLTRP, the regrind power requirement was significant at 2.2 MW absorbed. For this duty, Anglo Platinum and Xstrata developed a larger IsaMill; the M10,000.

2 COLLABORATIVE DEVELOPMENT OF TECHNOLOGY

2.1 Introduction
IsaMill Technology was first developed at Mt Isa in the mid 1990’s when a major research program was undertaken by MIM as it was then known (now Xstrata) to enable fine grained ore bodies to be economically developed. The first installation of the technology was at the Hilton Concentrator in 1992. This machine was an M500 with a 205 kW motor. The machine that was ultimately developed as an economic full scale machine was an M3000 IsaMill, with a 1.12 MW motor.

Initially, two M3000 IsaMills were installed in the Mount Isa lead/zinc concentrator in 1994. This was then followed with four M3000 IsaMills in the McArthur River concentrator in 1995 [4]. These were required to achieve ultrafine grinding for liberation of fine grained minerals and then achieve good separation of the minerals by flotation, after ultrafine grinding.

The subsequent development of the 2.6 MW M10,000 in late 2002 meant that in just 10 years since the first unit was installed in an operating plant at Hilton Concentrator, the units have increased their capacity 13 fold from 205 kW to 2.6 MW and their volume has increased 20 times.

Further to this, and the focus on this section of the paper, the design, fabrication, installation and commissioning of the project for the capacity increase from 1.12 MW (M3000) to 2.6 MW (M10,000) took only 16 months; an incredibly short time frame.

The project was initiated during a range of meetings between Xstrata Technology (or MIM Process Technologies as it was then) and Anglo Platinum. The scale up was always contemplated by both Netzsch and Xstrata as a future plan, even during the early development of IsaMill at Mt Isa, however it was brought forward as a focused development by a need of Anglo Platinum. The project had an aim to develop a larger IsaMill that has a higher capacity than two M3000 IsaMills for a similar capital and operating cost.

2.2 Driving Forces behind the Development
The reasons behind each company being involved in this collaborative development were all different. Anglo Platinum were trying to reduce capital cost for equipment and foresaw potential applications of large scale stirred milling requirements for whole of ore applications in their concentrators in the future. Anglo Platinum also had a current project with a specific application
that deemed it necessary for an IsaMill with over 2 MW of grinding power.

Xstrata Technology had considered this development to be a natural expansion of the marketed product range. There were also a number of internal projects that were under consideration at the time which envisaged use of larger scale IsaMills. To have the technology developed in this way under a collaborative development with an external client was clearly seen as bringing benefit to Xstrata. Finally, for Xstrata the application of IsaMills was limited due to the size of the largest existing available machine (1.12 MW). For large concentrators with whole of ore, or high tonnage regrind applications, having a significant number of smaller M3000 machines was not considered desirable from a capital, operating and maintenance perspective.

For Netzsch, the long established IsaMill manufacturer, the appeal was more from a challenge to their technical expertise and improved demand for their products.

2.3 Collaborative Development Process

The Process used for the design of the M10000 development was very unique. It was a major departure from all previous technology developments which were performed in house within Xstrata, and given the developments underway within MIM at this time it was quite incredible that it actually happened.

The process used was to first crystallise the need of the customer. Considerable time and effort by both Anglo Platinum and Xstrata Technology was involved examining the application, the requirements for regrinding and the potential of IsaMill to handle the duty. The result from this was the "Defined Application". The second step was for Xstrata Technology and Netzsch to examine the potential for the M3000 to be scaled up to meet the needs of the "Defined Application". This was conducted by initially Xstrata and Netzsch to individually review the requirements and then come together for a meeting to discuss and agree on the scale up mechanism, and the recommended machine size.

The final part of the collaboration involved Anglo Platinum undertaking an Engineering Review of IsaMill Technology and this involved a visit and audit of Netzsch’s manufacturing and engineering facilities in Germany.

These three process steps resulted a signing of The Collaborative Development Agreement on October 5th 2002 between Anglo Platinum and Xstrata.

3 DESIGN

3.1 Project Description And Pilot Testing

Anglo Platinum investigated the retreatment of dormant tailings dams in the Rustenburg area in 2000. Given market value of platinum and palladium and US dollar exchange rates at the time, the concentration of PGM minerals in the dams represented a major economic resource. Metallurgical test work identified that a significant proportion of these minerals could be recovered via fine grinding and flotation.

The pilot scale program was developed to confirm concepts previously identified during laboratory scale testing [2]. The laboratory tests suggested that the recovery of PGM’s from the tailings dams was sensitive to the grind size presented to flotation. Pilot tests confirmed this, showing PGM recovery in rougher flotation was 10 % higher with a primary grind of 90 % passing 75 µm, than with 80 % passing 75 µm. Significantly, flotation kinetics increased with the finer grind and the additional liberation of PGM’s and associated base metal sulphides.

![Figure 1: Flotation kinetics as a function of grind size](image-url)
Mineralogical examination showed poor upgrade potential of the rougher concentrate, due to insufficient liberation and oxide coatings on base metal sulphide surfaces. A conventional cleaner/re-cleaner circuit was not able to shift the grade/recovery curve, so regrinding of the rougher concentrate was employed to improve PGM liberation and freshen-up the heavily weathered mineral surfaces.

The main findings from the pilot plant results were [2]:

- The PGM grade and recovery targets could be met with the use of conventional flotation and IsaMill inert grinding.
- A primary grind of no less than 80% passing 75 µm, and rougher concentrate regrind of no less than 85% passing 25 µm is required to meet the grade/recovery targets.

IsaMill technology was enabling for the WLTR project, as smeltable concentrate grades could be produced from the oxidised, slow floating tailings.

### 3.2 Application

The WLTR concentrator has a capacity of 4.8 Mtpa, and has been designed to easily expand to 10.8 Mtpa. The flow sheet includes recovery of tailings by high pressure water monitoring, ball milling, rougher flotation, rougher concentrate regrinding and cleaner/re-cleaner flotation [1]. The process design called for an IsaMill nominal feed rate of 53 th⁻¹ and maximum 65 th⁻¹. The production of a P₉₀ = 25 µm required 35 kWht⁻¹ using -5 +3 mm local silica sand. The IsaMill product is fed to a two stage cleaning circuit to produce final concentrate [3].

### 3.3 IsaMill Design

Phase 2 of the WLTR Project (pending) requires the IsaMill to treat a far coarser feed size. Because of this, and to reduce process risk with the first installation, a variable frequency drive was installed.

The previous method of scale up (to the 1.1 MW M3,000 size) was based on conservation of power intensity. With the larger M10,000, the media agitator (grinding disc) tip velocity would be excessive using this method. A 'constant tip velocity' method was developed which proved more complex than power intensity models, as power does not scale-up linearly with volume.

The larger diameter of the M10,000 required modifications to the design of the Product Separator; a centrifuging/pumping device that
classifies the mill product (and retains grinding media). As centripetal acceleration decreases with increasing diameter (at constant tip speed), and pumping efficiency decreases with lower radial velocity, both the centrifuging and pumping actions of the Separator had to be re-designed. The new design focused on improving efficiency of the rotor suction region, rotor pump finger shape and product classification efficiency [1]. Component and materials selection was vital to the operating cost competitiveness of the new mill. A key design aspect was to control disc wear, and the design proposed that the disc surface abrasion rate was within 3% of the M3,000 mill’s rate. As the M10,000 disc design was larger in diameter and thickness, the increased volume of rubber would mean a longer disc life than in the M3,000, as abrasion rates are a function of area and the power density per unit area would be lower.

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![Figure 4: 22 x 11 m footprint of WLTR M10,000 IsaMill grinding circuit](image)

4 OPERATION

4.1 Scale Up

The theoretical power draw can be calculated for a given volume, media filling and disc tip speed. With a fixed speed mill, the accuracy of the power calculation can only be verified at a single point. The power based scale up calculations could be verified over a wider range than usually possible, due to the use of a variable speed drive with the WLTRP M10,000. At the time of testing the power scale up, the IsaMill power draw had been reduced by operating at a lower media volumetric filling of 56% (calculated). The IsaMill feed tonnage was well below design, so the mill was turned down to avoid over grinding. This explains why the motor power draw in Figure 5 is lower than normal. The actual motor power draw at different mill speeds is within a 2% error band of the scale up calculation.

Test work on similar Anglo Platinum Merensky tailings material was conducted at laboratory scale in 2001. A range of tests were performed, at varying feed sizes and energy inputs. The grinding data for a laboratory test is compared to survey data of the M10,000 at similar energy input (Table 1). Allowing for minor differences in the ore samples, the milled products should exhibit similar particle size distributions as the mineralogy is broadly consistent and the same grinding media (-5 +3 mm local crushed silica sand) and specific energy were used.

The M10,000 exhibits marginally better top size control, and a narrower overall product PSD. The CSI (Coarse end Size Index - or $P_{98} : P_{80}$ ratio) is 2.6 for the M10,000 whereas the CSI for the laboratory mill is 3.0.

The M10,000 scaled up accurately, both in terms of power and product PSD. Power efficiency appears equal to the laboratory mill, with control of the product PSD marginally better.

![Figure 5: Theoretical vs actual motor power at varying disc tip speed](image)
Table 1: Laboratory vs Full Scale Grinding Efficiency

<table>
<thead>
<tr>
<th>IsaMill Model</th>
<th>Installed Power (kW)</th>
<th>Chamber Volume (L)</th>
<th>Specific Energy (kWh/t)</th>
<th>Pulp % Solids</th>
<th>P90 (µm)</th>
<th>P80 (µm)</th>
</tr>
</thead>
<tbody>
<tr>
<td>M4</td>
<td>4</td>
<td>3.5</td>
<td>37</td>
<td>39</td>
<td>47.5</td>
<td>16.0</td>
</tr>
<tr>
<td>M10,000</td>
<td>2,600</td>
<td>10,000</td>
<td>37</td>
<td>42</td>
<td>42.5</td>
<td>16.5</td>
</tr>
</tbody>
</table>

4.2 Turn Down
Section 4.1 describes the M10,000 operating at low power (< 50% of installed motor capacity) by reducing the media load in the grinding chamber. This method of power control is common to all IsaMills. Decreasing power by reducing media load, or ‘turn down’ is necessary in single mill grinding lines to avoid over grinding (and over heating) during periods of low feed tonnage. Turn down would be important to the WLTRP, as the rougher flotation mass pull would fluctuate with head grade. Figure 7 shows the M10,000 operating at similar energy inputs, but greatly different power draws. Despite these extreme operating situations, the IsaMill produces the same P80 in both cases.

4.3 Flotation
A series of IsaMill on/off tests were performed between August and October 2004, to confirm the effect of fine grinding on flotation [5]. Figure 8 shows the impact of fine grinding on final concentrate grade, by plotting the IsaMill on/off data pairs over time. Only data pairs that represent a full 24 hour period (of
Table 2: Mean final concentrate and cleaner flotation tail grades

<table>
<thead>
<tr>
<th>IsaMill Status</th>
<th>Final Conc (g/t)</th>
<th>% Change</th>
<th>Cleaner Tail (g/t)</th>
<th>% Change</th>
</tr>
</thead>
<tbody>
<tr>
<td>On</td>
<td>57.3</td>
<td>+ 42 %</td>
<td>3.05</td>
<td>- 28 %</td>
</tr>
<tr>
<td>Off</td>
<td>40.4</td>
<td></td>
<td>4.26</td>
<td></td>
</tr>
</tbody>
</table>

The higher flotation kinetics and improved PGM liberation that come with regrinding the rougher concentrate, resulted in higher concentrate grades in the reclaimer circuit. As the reclaimer concentrates report to final concentrate, the ability to maximise the grade in these cells is vital to meeting overall concentrate specification. The difference in reclaimer concentrate grades with the IsaMill on or off is detailed in Figure 9 which shows a down-the-bank survey. Cleaner flotation grade and recovery is improved with IsaMilling.
4.4 Maintenance, Availability and Operating Cost

Being a completely new machine, the focus placed on commissioning activities was great. Representatives from Xstrata (Technology (Brisbane), Zinc (Mount Isa) and South Africa business units), Netzsch (Germany) and Siemens (Germany and Australia) were involved in the commissioning process. Despite having built contingencies upon contingencies in the planning stage, the end result was an uneventful and simple commissioning. Hot commissioning commenced on 15th December 2003, and was complete by the 18th December. The mill was then 'stood down' until the tailings dam reclaim mining and main plant ramp up were complete.

Following several brief shut downs in the first month of operation to inspect the mill internals, the M10,000 was put into a normal PM cycle of an 8 hour shut down every 8 weeks (or 1300 hours mill run time). These shut downs are used to rotate and/or shuffle the grinding discs to maximise their life. The disc life is 6000 hours, or 38 weeks (at current circuit availability). The first-fill shell liner is still in use, and was 30 % worn at the July 2005 shut down.

Down time for the IsaMill grinding circuit in the months June and July 2005 is shown below, and separated into stoppages related only to the IsaMill, and stoppages related to ancillary equipment in the circuit. Examples of the most common stoppages of ancillary equipment include pump trips and bypass events due to upstream disturbances that require the IsaMill feed thickener to be placed in recycle. The overall grinding circuit availability is greater than 95 %, with the IsaMill machine availability above 99 %. The month of July included a planned 8 hour IsaMill maintenance shut down.

As the M10,000 design was able to take advantage of hindsight and adjust several parameters from the M3,000 design, it was expected that the larger IsaMill would have a lower operating cost. Another platinum producer operates an M3,000 IsaMill only kilometres from the WLTR site, and uses the same grinding media, with similar feed and product sizes. This allows for the most accurate comparison in spares cost between the two machines. Table 4 shows that the smaller IsaMill spares cost is three times the cost per tonne milled and almost double the cost per kilowatt hour of the larger machine.

5 CONCLUSION

The use of a novel technology development method has been beneficial for Anglo Platinum and Xstrata. The design process was rapid, exhaustive in detail, transparent and above all successful. The M10,000 IsaMill has scaled up accurately and is operating below forecast cost and above budget availability. Most importantly, the down stream cleaner flotation results are matching those seen in pilot testing, whilst plant tests have confirmed the need for regrinding with the IsaMill prior to cleaning to achieve final concentrate grade.

<table>
<thead>
<tr>
<th>Table 3: Down time analysis for June and July 2005</th>
</tr>
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<tbody>
<tr>
<td></td>
</tr>
<tr>
<td>A Hours in month</td>
</tr>
<tr>
<td>B Main plant running</td>
</tr>
<tr>
<td>C IsaMill planned maintenance</td>
</tr>
<tr>
<td>D IsaMill break down</td>
</tr>
<tr>
<td>E Total IsaMill down time = (C + D)</td>
</tr>
<tr>
<td>F Ancillary break down</td>
</tr>
<tr>
<td>G IsaMill Availability = (B – E) / B x 100</td>
</tr>
<tr>
<td>H Overall Circuit Availability = (B - E - F) / B x 100</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Table 4: Normalised operating cost of M10,000 and M3,000 IsaMills</th>
</tr>
</thead>
<tbody>
<tr>
<td>IsaMill Model</td>
</tr>
<tr>
<td>----------------</td>
</tr>
<tr>
<td></td>
</tr>
<tr>
<td>M3,000 (with 1,100 kW motor)</td>
</tr>
<tr>
<td>M10,000 (with 2,600 kW motor)</td>
</tr>
</tbody>
</table>
REFERENCES


ABSTRACT

"Conventional wisdom" has deemed that fine particles have low flotation recoveries. Plant size-recovery graphs often have the classic "hill" shape - high recovery in the mid sizes and low recovery at the fine and coarse ends. Yet if mineral liberation is poor, low fines recovery may be because you don't grind fine enough!

This apparent paradox is explained by the old concept of "sand/slimes" circuits, which recognises the different flotation needs of fine and coarse particles. This concept is overlooked in the push for simpler circuits and larger equipment. Most plants now treat all particles together in a wide size-distribution. Reagent conditions are set for the dominant coarser particles, so fines are starved of collector. Worse still if there are significant mid-sized composites - often these have to be rejected in cleaning to achieve target concentrate grade. But the conditions which reject mid-size composites - collector starvation and high depressants – also reject fine liberated particles. In fact, fines flotation can be excellent when flotation chemistry is tailored to fines.

After first recovering fast floating liberated particles, correct grinding to liberate remaining composites is essential to increase fines recovery. Firstly, liberating composites allows lower depressant and higher collector additions, since composites will not dilute the concentrate. Secondly, finer grinding narrows the size distribution to flotation, allowing reagent conditions to be set to suit all particles. Additionally, fine grinding in an inert attritioning environment like an IsaMill removes surface deposits that may have made some fines slow-floating.

An excellent case study is the installation of IsaMills in the Mount Isa lead zinc concentrator to grind lead and zinc rougher concentrate to 12 microns and zinc cleaner tailings to 7 microns. Most plant losses had previously been in the sub 10-micron fraction, yet ultra-fine grinding increased plant recovery by 5% lead and 10% zinc. Circulating loads dropped, reagent additions dropped in spite of the much higher particle surface area, and the plant became much more operable and responsive.
INTRODUCTION: THE CONVENTIONAL VIEW

The conventional view of the flotation size-recovery curve is shown in Figure 1. There is a good reason for this view — if you sample almost any flotation plant you will produce a similar curve. The numbers speak for themselves — fine particles float poorly in most plants. Operators carefully avoid “overgrinding” and “sliming” of feed.

Yet this obvious conclusion is challenged by practice in some other plants. For example, Xstrata’s McArthur River Mine (MRM) produces 380,000 t/y of concentrate at a P80 size of 7 microns (µm), and a recovery of 82%. And Mount Isa Mines produces 260,000 t/y of lead concentrate and 350,000 t/y zinc concentrates at P80 about 15 µm and over 80% recovery. The biggest problem for these plants is coarse particles — anything over 20 µm is considered as “gravel” which will reduce recovery.

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Figure 1: Typical Recovery vs Size Curve

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Improving Fines Recovery By Grinding Finer: MetPlant 2004 AusIMM
RESOLVING THE PARADOX

Why is it that most plants fear fines production, yet others rely on it? The answer lies in three areas:

- Mineralogy and liberation
- Operating constraints and strategy
- Flotation rates and surface area

Mineralogy: make no mistake, you don’t want to grind fine if you don’t have to. It is power intensive and expensive. But if you have fine-grained mineralogy, you have no choice. Figures 2 shows photomicrographs of Broken Hill ore and McArthur River (MRM) ore at the same magnification. Broken Hill will never have a liberation problem – its metallurgists can focus on simple grinding and flotation circuits and avoiding “overgrinding”. But for MRM and Mount Isa, no amount of fiddling with circuit design or reagent testing will help (trust us, we wasted years of our lives trying!). For these ores, there is one question that must be answered before anything else: “Will this circuit change/new reagent increase mineral liberation?” If not, do yourself a favour and keep your money in your pocket.

Figure 2: Different Grain Size of Broken Hill and McArthur River Ores
(both photos at the same magnification)
SOME DEFINITIONS AND PERSPECTIVES

We often note that communication between different operating plants is confusing because we all use different definitions of “coarse”, “intermediate” and “fine” particles. An operator of a “coarse” grained orebody may call minus 37 \( \mu \text{m} \) “slimes”. To an operator at Mount Isa or McArthur River this is gravel. To them, anything above 20 \( \mu \text{m} \) is coarse, between 10 and 20 \( \mu \text{m} \) is intermediate, and less than 10 \( \mu \text{m} \) is fine or ultrafine.

We avoid using generic terms like “fines” in this paper. But if you hear someone saying that “fine” particles don’t float well, ask them two questions:

- What do they mean by “fine”?
- Are they aware that in the last decade, Mount Isa Mines and MRM have produced over 10 million tonnes of concentrates at an average sizing less than 10 \( \mu \text{m} \), at over 80% recovery, in conventional flotation cells, with a simple xanthate reagent system?

Particle surface area per tonne increases rapidly as size gets finer. One tonne of 7 \( \mu \text{m} \) particles has 5 times the surface area of a tonne of 37 \( \mu \text{m} \) particles; for 2.5 \( \mu \text{m} \) particles the surface area triples again. This explains why grinding energy increases exponentially as grind size decreases (Figure 3), and why finer media with higher surface area is needed for grinding below about 30 \( \mu \text{m} \). It also explains the higher collector need of fines.

In this paper reference to a 7 \( \mu \text{m} \) particle means a single particle of that diameter. By convention a grind size of 7 \( \mu \text{m} \) refers to the 80% passing size (P80). For example, the regrind size at MRM is 7 \( \mu \text{m} \), so only 20% of final product is 7 microns or coarser. Fifty percent weight of MRM concentrate is below 2.5 \( \mu \text{m} \). Since flotation works on individual particles interacting with bubbles, consider this from a different perspective - 50% by weight less than 2.5 \( \mu \text{m} \) means that 96% of individual particles recovered at MRM are less than 2.5 \( \mu \text{m} \) ! In spite of popular perception, fine particles float very well indeed.
ULTRAFINE GRINDING CIRCUIT DESIGN

If you have to grind below 25 μm, then you need to choose the right equipment. Three issues are particularly important:

- **power efficiency**
- **classification within the grinding circuit**
- **the impact of grinding on flotation performance**

**Figure 3**: Grinding Energy versus Product Size for a Gold Ore

- **Power efficiency** is demonstrated by Figure 3, comparing the power required to grind a gold ore in a ball mill with 9 mm balls with an IsaMill with 2 mm media. The IsaMill is much more efficient below about 30 μm – to grind this ore to 15 μm would take 28 kWh/t in the IsaMill, but 90 kWh/t in a ball mill. Traditionally this has been attributed to the difference between attrition grinding and impact grinding. However by far the most important factor is media size, as shown by Figure 4, which shows the breakage rate in Tower Mills drops dramatically - the breakage rate for 20 μm particles is ten times lower than the rate for 40 μm particles. Even though the Tower Mill is full attrition grinding, practically it is constrained to using relatively coarse media, 9mm balls in this case. In contrast, the IsaMill (Netzsch mill in Figure 4) can operate with much finer media and much higher intensity of power input (Table 1), meaning the peak breakage rate occurs at 20 μm, and doesn’t drop as quickly below that.
Figure 4: Breakage Rates in Different Grinding Devices

![Graph showing breakage rates in different grinding devices.](image)

Table 1: Comparison of Various Grinding Technologies

<table>
<thead>
<tr>
<th>FEATURE</th>
<th>ISAMILL</th>
<th>TOWER MILLS</th>
<th>VERTICAL PIN MILLS</th>
</tr>
</thead>
<tbody>
<tr>
<td>Grinding Intensity (kW/L)</td>
<td>0.54</td>
<td>0.005</td>
<td>0.15 - 0.18</td>
</tr>
<tr>
<td>Residence Time to 15 µm (min)</td>
<td>0.6</td>
<td>154</td>
<td>7 - 9</td>
</tr>
<tr>
<td>Power Usage to 15 µm (kWh/t)</td>
<td>17.4</td>
<td>59.6</td>
<td>37.5 - 39.0</td>
</tr>
<tr>
<td>Media Material</td>
<td>Various</td>
<td>Steel</td>
<td>Steel</td>
</tr>
<tr>
<td>Media Size (mm)</td>
<td>0.8 - 1.6</td>
<td>9 - 12</td>
<td>6 - 8</td>
</tr>
</tbody>
</table>

Figure 4 and Table 1: Extracts from AMIRA P336, Gao M and Weller K, Review of Alternative Technologies for Fine Grinding, November 1993.

Improving Fines Recovery By Grinding Finer: MetPlant 2004 AusIMM
• **Good classification** is vital for power efficiency in ultrafine grinding, just as it is in conventional grinding. However it is not generally practical to use cyclones to close-circuit a grinding mill with a target below about 15 µm. To get good cyclone efficiency at these sizes requires small cyclones, eg two inch (50 mm) diameter or smaller. This is virtually inoperable on a large scale, so the circuit is either compromised (and less power efficient) by using bigger cyclones, or an alternative solution is needed. The IsaMill achieves this by the internal classifier mechanism, using the high centripetal forces generated inside the mill to classify the discharge, ensuring a very sharp product size without external cyclones. The very short residence time in the IsaMill also minimises "overgrinding", further contributing to the sharp product size distribution. As an added advantage this mechanism also retains fine media very effectively, meaning that low cost media can be used, eg local sand, or granulated smelter slag.

A cautionary word to those designing circuits – the benefits of good classification on power efficiency and media retention does not show up in laboratory tests. These tests are done in batch mills, and many technologies will show the same power efficiency in a closed device. The crucial questions are, what is the power efficiency, media retention, and product size distribution in a full-scale continuous installation.

• Managing the **impact of grinding on flotation performance** is the third crucial factor in plant design. Even if you can accept the low power efficiency of a mill with steel balls, you may not be able to deal with its impact of surface chemistry. Consuming so much power in a steel environment means high retention time and lots of steel contamination. The resultant low pulp potential changes flotation behaviour, requiring additional reagents and reducing selectivity. One early response to this problem was to use High Intensity Conditioning (HIC), eg at Hellyer, to reverse the negative impact of Tower Milling on surface chemistry. Processes like IsaMilling are far more efficient by providing this high intensity as part of the grinding action, and grinding in an inert environment. Later we will show how IsaMills significantly improved the flotation behaviour of ultrafine particles at Mount Isa.
FLOTATION CIRCUIT DESIGN AND OPERATING STRATEGY

The view that “fine” particles don’t float is caused by circuit design and the constraints of operating strategy. Simply, flotation works best when applied to narrow size distributions. A 5 µm particle has 10 times the surface area of a 50 µm particle, and fundamentally different hydrodynamics. Yet often our circuit designs assume they will behave the same, and treat them together in flotation. Texts as old as Taggart described the benefits of “sand/slimes” splits into separate circuits. This simple concept has been largely ignored in the push for circuit simplification and larger flotation cells.

We are not advocating complicated flotation circuits. However if you have fine-grained minerals then you must design your circuit to suit the needs of fine particles, not coarse particles. The Mount Isa circuit developed into an excellent balance of the needs of different minerals, relying on several stages of grinding and flotation. The design principals are:

- **Don’t grind anything more than you need to.** Fine grinding is expensive – technically the best solution for Mount Isa would be to grind everything to 12 µm, but this would not be economic. Therefore stage grind and float to suit the mineral behaviour – at Mount Isa this means a 37 µm grind before roughing. Some mineral is liberated at this size and can go to concentrate. Other minerals in rougher concentrate need to be ground to 12 µm. Some of these are rejected in cleaning and need to be reground to 7 µm.

- **Float minerals in narrow size distributions** – this happens automatically with the staged grinding approach described above, and is assisted by the inherent sharp size distribution produced by the IsaMills.

- **Minimise circulating loads, and open-circuit as much as possible** – this is another automatic outcome of staged grinding. It is pointless to recirculate a composite particle unless you are going to grind it to liberation. If you do reground it, you should now float it separately with similar sized particles.

These principles can be seen in the simplified Mount Isa flotation circuit in Figure 5. Though the circuit may appear complicated, it is better than the alternatives of either:

- Grinding everything to 12 µm and floating together (too expensive)
- Recirculating reground products and trying to float them with coarser minerals (causing poor performance of the reground minerals, high circulating loads and low recoveries).
Contrary to appearance, these developments at Mount Isa greatly simplified circuit operations. Lead recovery increased by 5% and lead concentrate grade by 5%, zinc recovery by 10% and concentrate grade by 2%. More surprisingly, reagent needs dropped, circulating loads dropped, and the circuit became far more stable. Flotation suddenly became as easy and predictable as the textbooks say it is!

Figure 5: Mt Isa Pb / Zn Concentrator Flow Sheet

Overall Recovery: Pb = 79 %  
Zn = 82 %
COMPETING FLOTATION RATES OF DIFFERENT PARTICLE SIZES

The profound impact of a narrow size distribution to flotation feed is explained by mineralogy and operating constraints. In a system with just pure liberated sphalerite and quartz, flotation could achieve good recovery in all size ranges, even though the “fines” have slower flotation rates. But in real circuits there are two crucial constraints:

- Other contaminant minerals such as pyrite and pyrrhotite also exhibit some floatability. If so, a “coarse” pyrite particle may have the same flotation rate as a “fine” sphalerite particle.

- Composite particles. To explain the problem with composites, imagine a simple $37 \mu m$ sphalerite-quartz binary. This particle has to be rejected since typically zinc concentrates must be less than 3% silica. The low collector and high depressant needed to reject the $37 \mu m$ composite will also depress the slower floating $10 \mu m$ liberated sphalerite particle.

Some further problems arise when floating coarse and fine particles together:

- It is no point depressing the $37 \mu m$ composite particle unless you can liberate it. While plants often “send it to regrind”, this is often a conventional ball mill or Tower Mill that has very low breakage rates on sub $30 \mu m$ particles. This causes high circulating loads of composite particles.

- The high circulating loads then take up volume and reduce residence time in roughing and cleaning. Since fine particles are slower floating, this drop in residence time further hurts their recovery.

- If high pH is used for depression, and lime is used to get high pH, then a surface chemistry problem is introduced. Circuit water can become super-saturated in Calcium ions. This leads to reaction with sulphate ions, which causes gypsum to precipitate on the nearest surface – usually a mineral particle. SEM work at Mount Isa before IsaMills showed that up to 80% of sphalerite surface was masked by gypsum. This has a more serious effect on sub $20 \mu m$ particles.
RECONCILING THE REALITY WITH THE PERCEPTION

The common perception is that “fines” don’t float, the reality is that they will in the right conditions. Figure 6 explains this conceptually – in most plants, sub 20 µm particles do perform poorly because they are mixed with coarser particles with much different needs. If these particles were floated in a narrow size distribution, flotation conditions can be tailored to them and they perform well. This explains the performance of the Mount Isa staged grind and float circuit.

Figure 6: Conceptual description of staged grind and float circuit
CASE STUDY – MOUNT ISA LEAD ZINC CONCENTRATOR

The changes to the Mount Isa circuit as part of the “George Fisher Project” are detailed elsewhere (Young & Gao, 2000, Young, Pease & Fisher, 2000). In summary, the project involved adding a further 6 IsaMills, to regrind lead rougher concentrate to P80 of 12 µm, most zinc rougher concentrate to 12 µm, and a zinc regrind to P80 of 7 µm (see Figure 5). Lead performance also increased by 5% concentrate grade and 5% recovery (equivalent to 10% increase in lead recovery at the same grade). Zinc recovery increased by 10%, in two steps, and zinc concentrate grade by 2% (equivalent to 16% increase in zinc recovery at the same grade). The story of zinc metallurgy can be told in Figures 7, 8 and 9.

The project predicted 5% higher zinc recovery (and no extra concentrate grade) due to extra liberation. Figure 7 shows this was achieved instantly. What surprised us was the “second wave” of a further 5% zinc recovery increase and the 2% increase in zinc concentrate grade. This was because fines performance improved after we ground finer.

It took us about 6 months to discover how much better the fines could perform because we were so used to flotation performance after conventional grinding rather than after IsaMilling. The three biggest mistakes we made were:

- We thought we would need a lot more reagents after IsaMilling. Since we created a huge amount of new surface area we expected some reagent additions to triple.
- We didn’t take the depressant (lime to pH 11) off the zinc cleaners.
- We thought flotation rates would be slower, so we “pulled” flotation banks harder to compensate.

Figure 7: Zinc Recovery Increase from IsaMilling

Reduced grinding & flotation capacity, due to equipment relocation during construction.

IsaMills Commissioned

1st Wave +5% Zinc Recovery

2nd Wave +5% Zinc Recovery

+ 2% Conc Grade (not shown)

Baseline

Improving Fines Recovery By Grinding Finer : MetPlant 2004 AusIMM
This resulted in four fundamental changes:

- Circulating loads had dropped dramatically, because we had finally addressed the liberation problem. This created significantly more flotation capacity than we had expected, and reduced reagent need.

- We no longer needed depressant on the cleaners since there weren't many composites left. The liberated minerals “behaved” properly. This was a crucial step, since removing the lime dropped the calcium in circuit water, dropping gypsum formation and further increasing flotation rate of sub 15 µm particles.

- The narrow size distribution in each stage of flotation made flotation easy. Together with the low circulating loads and open-circuit design, the flotation circuit became steadier and simple to control.

- The inert grinding environment and attritioning in the IsaMill improved mineral behaviour and changed reagent needs. Our initial expectations were way off – we added far too much reagent, and in the wrong places.

Over the next 3 months we learnt to challenge every known “truth” about the behaviour of our circuit, and accept that grinding could improve mineral flotation behaviour rather than reduce it. Once we understood that we had to approach the circuit as a “clean sheet”, we rapidly improved recovery by the 5% shown in the “second wave”. In spite of adding 6 MW of grinding power to the circuit, we were amazed that unit costs ultimately did not increase:

- Overall reagent consumption remained the same in spite of the extra surface area – since we no longer had competing depressants and collectors battling with composites.

- Total power consumption did not increase as much as we thought, since circulating loads of over 100% simply disappeared. As a result we were able to shut down some flotation capacity.

- With lower reagents and no circulating loads, spillage was almost eliminated.

- The very steady nature of the circuit meant that simple stabilising control worked, keeping equipment at high efficiency.

- The IsaMill grinding media is free, granulated lead smelter slag that would otherwise be discarded.
The improvements in flotation are further described by Figure 8, which records size-by-size performance for sphalerite. Not only did the IsaMills put more material in the liberated sub 15 µm size range, they also improved recovery of all sub 25 µm particles. For example, recovery of 8 µm liberated particles increased from about 78% to about 88% after Isa Milling. Though recovery in the coarser fractions has dropped, this was what was needed to improve total recovery since:

- The coarser fractions had contained the composites, previously rejecting these composites also rejected fines.
- After IsaMilling, very fast floating coarser particles (ie liberated ones) were still accepted, but the slower floating ones were sent for regrinding and liberation.
- There was much less material in these coarse size ranges after IsaMilling – coarse composites were reground into finer size ranges with higher recovery (and higher grade since they were more liberated).

Figure 8: Increased Fines Recovery After Fine Grinding in the IsaMills
THE BIG PICTURE

The history of changes at Mount Isa is described by Figure 9, which considers sphalerite liberation and zinc recovery. As ore became finer grained in the 1980’s, liberation dropped and recovery dropped accordingly. In 1992 we installed conventional ball and Tower Milling to improve liberation. Sphalerite liberation improved by 25% (from 55% to 80%), but recovery only improved by 10%, due to the extra difficulty of floating after grinding in ball and Tower Mills. In 1995 we installed the prototype IsaMills in the lead circuit. Not only did this improve lead performance, but zinc recovery also increased even though total sphalerite liberation was little changed. This was because the improvements from IsaMilling allowed sphalerite to be redirected from the lead concentrate to the zinc circuit. In 1999 the installation of additional IsaMills increased sphalerite liberation by a further 5%, but zinc recovery increased by 10% and zinc concentrate grade by 2% (net impact equivalent to 16% recovery increase at the same concentrate grade). This demonstrates the fundamental changes to fines flotation made possible by ultrafine grinding in IsaMills.
Ball & Tower Milling Added
Liberation increased by 20%
Recovery increased by 5%
Recovery increased less than liberation due to negative impact of steel media on flotation of fines

First IsaMills Added in Lead Circuit
Sphalerite liberation constant
Zn recovery increased by 4% due to better rejection from Pb conc after IsaMilling

IsaMills Installed In Zinc Circuit
Liberation increased by 7%
Recovery increased by 10%
Conc grade increased by 2%
Recovery benefit higher than liberation increase because of improved flotation after IsaMilling
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ABSTRACT

Fine grinding has the potential to revolutionise the mineral processing industry due to the benefits attained through being able to economically grind finer to enhance particle liberation for improved flotation and to introduce a high degree of strain into the mineral lattice to improve leaching rates in hydrometallurgical processes.

The development of fine grinding technology has enabled mining operations to be developed at a number of mine sites previously considered uneconomic. It has also enhanced the performance of existing mining operations at a number of mine sites and provided the key such that leaching can be carried out under relatively mild conditions in low cost open tanks. This paper is a review of a number of case studies that demonstrate the benefit that fine grinding has provided to a wide range of applications.

The use of inert media has provided further improvements in circuit performance and recovery.

The Mt Isa operation, MRM operations and the Albion Process are reviewed in this paper.
INTRODUCTION

Fine grinding technology has come a long way in the last decade. The ability to grind ore down to sizes below 10µm has lead to the establishment of new mines such as McArthur River, Century and George Fisher mines, as well as improving recoveries of lead and zinc at the established Mount Isa deposits.

Fine grinding has also opened up the door to new leaching processes. Leaching has always been regarded as a simple process, but as orebodies have become more complex, and oxide deposits giving way to more refractory ore, more advanced leaching is required. One of the technologies that fits this duty is the ALBION Process, however unlike competing technologies that incorporate complex processing steps, the development of fine grinding has made this process a very simple and effective process.

Both processing routes was made possible by the innovation of the IsaMill and other fine grinding processes, and this paper investigates the impact of this technology on the processing of complex ores through mineral processing and leaching.
BACKGROUND – MOUNT ISA

Mount Isa is a long established mining town situated in North West Queensland. Lead carbonate ore was first found by John Campbell Miles in 1923, which later revealed one of the biggest lead/zinc sulphide deposits of its time. Mount Isa Mines Limited was formed soon after the initial discovery, and lead mining, concentrating and smelting operations began.

Over time, the easy to treat orebodies gave way to harder to treat orebodies, which had lower grades and more complex mineralogy. The two main lead/zinc orebodies found at Mount Isa were named Racecourse and Black Star, described as: (Pease, Young et al, 1997):

- “Black Star” orebodies are massive and are mined by open stoping at lower mining cost, and therefore have a lower cut-off grade. Generally this ore is finer grained and has considerably more fine-grained carbonaceous pyrite, whilst core replacement of pyrite by galena (atolling) is more common and at a more advanced stage.

- “Racecourse” orebodies are narrow and mined by bench stoping with a higher cut-off grade. Generally these orebodies are higher in grade for lead and silver, coarser grained and lower in pyrite. The pyrite is more the euhedral type than the fine-grained carbonaceous type.

The other major lead/zinc orebodies in the region, also owned and mined by Mount Isa Mines Limited, was the Hilton and Hilton North orebodies, 20km north of Mt Isa. The mine was later renamed as the George Fisher mine. The orebodies were similar to the orebodies at Mount Isa, with the upper part of the orebodies having characteristics of the “Black Star” ore, and the lower orebodies similar to “Racecourse” ore (Pease, Young 1997).

The presence of pyrite was described as two types. The euhedral pyrite was described as coarse grained, while the fine grained carbonaceous pyrite is present in the 5 to 30µm range containing natural carbon and is naturally floating, (Munro 1993).

CONCENTRATOR PRACTICE AT MOUNT ISA

Ore at Mount Isa was mined and treated through the Lead/Zinc Concentrator. There have been various concentrators built on the site. The current lead/zinc concentrator, No. 2, operating today, was installed in 1966, and produces lead and zinc concentrates through sequential lead, then zinc flotation. It also had a LGM circuit which was a zinc rich bulk concentrate where “low grade middlings” were processed into a saleable product.

As the operations at Mt Isa progressed, the Racecourse orebodies provided the bulk of the ore during the early days of mine development, being higher grade. However over time more of the Black Star material was processed. The introduction to mining the Hilton orebodies began in 1987, and a lot of this material was trucked into Mt Isa to be treated through the Lead/Zinc Concentrator. (a lead/zinc concentrator was built at Hilton, and was commissioned in late 1989 for processing some of the ore).

Along with the changing orebodies, and the resultant mineralogical changes, the mining rate was also increased. This is highlighted in figure 1 (Pease, Young et al 1997), where head grades of the concentrator feed dropped while the throughput rate increased.
As more Black Star and eventually Hilton ore was being treated, the ore became finer grained and more complex. This was observed for all sizes of concentrator feed, having less liberated mineral species. Figure 2 shows the amount of sphalerite liberation per size class deteriorating over time, from 84/85 to 91/92.

Also as more of the Black Star was mined, the proportion of carbonaceous pyrite compared to the euhedral pyrite increased. The carbonaceous pyrite is hydrophobic under most conditions and cannot be depressed easily, as well as consuming large amounts of reagents. The net effect of this change in terms of mineralogy is a greater concentration of iron sulphides in the concentrates (Pease, Young et al 1997). Any attempt made to depress the presence of iron sulphides in the concentrate such as the use of lime or dextrin, resulted in composites being depressed as well in the concentrate, which formed high recirculating loads in the circuit. This in turn limited cleaning and flotation capacity.
In summary the Lead/Zinc operation at Mount Isa was not recovering enough metal. The main concerns were:

- More fine grained ore
- Less liberation in the concentrator
- Greater presence of carbonaceous pyrite
- Increasing throughput

The net result was the detrimental impact of metal recovery. Figure 4 shows the gradual deterioration in zinc recovery following the decrease in zinc liberation.

Figure 4: Zinc Liberation and Recovery
At the same time as this trend was occurring, MIM Holdings, the parent of Mount Isa Mines Limited, was embarking on processing the ore from the McArthur River deposit in the Northern Territory. This material was extremely fine grained and required grind sizes down to 7µm to liberate the particles. Obviously, new processing techniques were required to turn around the trend.

ADDRESSING THE LIBERATION AND SEPARATION PROBLEMS PRE 1994

At the start of the nineties there was very little equipment available for fine grinding in mineral processing, (fine being defined as below 25µm at least). There was conventional ball milling, and tower milling, and the emergence of metprotech mills, but these were inefficient, as well as impacting on downstream flotation.

Nevertheless, Mount Isa Mines tried to address the impacts of deteriorating ore by implementing the following as described Pease, Young et al 1997:

- Tower mill regrinding installed in the LGM circuit (1991), dropping the p80 to 12µm
- The “Fine Grinding Project,” which doubled grinding and flotation capacity and instituted a “cold” lead reverse cleaning circuit (1992). This project addressed both key issues, ie liberation in the zinc circuit and separation (of carbonaceous pyrite) in the lead circuit. The flotation feed dropped from a P80 of 80µm to a p80 of 37µm due to an increase in secondary grinding from 6.3 MW to 11.5 MW
- Improvement in liberation allowed circuit simplification, the increased use of conventional tools (eg high pH zinc circuit cleaning) and relocation of regrinding duties from the LGM circuit to the zinc retreat circuit.
- Generally, the application of process control became more effective with the improvements because of more achievable targets.

Details of these circuit changes were described in detail by Pease, Young et al 1997. A brief summary of the observations at this stage was that by grinding finer up front in the LGM circuit some of the liberation problems were being addressed. The liberated carbonaceous pyrite contamination of the concentrates were being improved by the reverse flotation in the lead circuit, the use of traditional depressants to knock out the pyrite, as well as the increased capacity to allow for the adequate separation separation.

Of the changes in the circuit it was reported that major gains had been achieved by the extra regrinding with some of the composite zinc that had reported to the LGM having been liberated enough to report to the zinc concentrate. This achieved an extra 5% of zinc recovery to the zinc concentrate. However, zinc liberation, had been increased to 20%, and higher recoveries were expected at this level of liberation than what had actually been achieved.

THE IsaMill DEVELOPMENT

In 1984 to 1986, fine grinding work had been conducted in the pilot plant at Mt Isa to investigate fine grinding circuits. Tower mill test work had also been investigated in 1991, with poor results due to the inability of the equipment to grind fine enough. The other drawbacks of these circuits was the high iron content in the mill charges and the resulting impact on the surface chemistry due its effect on the redox potential.

The introduction of a small stirred mill to the Mt Isa site could only be described as a major turning point in fine and ultrafine grinding in mineral processing. The mill, a ½ litre bench scale mill, resembled a milk shake maker, and initially used fine copper smelter slag to grind the ore and concentrate. It was provided by Netzsch, an experienced fine grinding equipment supplier, in the paint and food processing industry! Testwork on McArthur River Ore started in 1991, and by January 1992, a small pilot scale mill, LME100, had been designed and installed at the pilot plant at Mt Isa.

In a relatively short time, the influence of the bench scale testwork lead to the development of the first pilot scale mill. Mill development is shown in the following table.
The table highlights the major mill models that evolved, but does not attempt to show the detail of testwork that was involved with both the mill and flotation circuit downstream of the mill. To go into the development is a case study within itself. Needless to say there were hundreds of trials involved to develop the commercial mill, the M3000, including various component trials for disc materials, spacing of components, separator, liners, chamber lengths, plant and lab testwork, mineralogical investigations and examinations, and most importantly the development of the screenless discharge – ie. product separator.

It is important to note that in just 11 years since the first unit was installed in an operating plant at Hilton Concentrator, the units have increased their capacity 13 fold from 205KW to 2.6MW. Mill volume has increased 20 times in this time frame. This is a very rapid increase in capacity. In comparison, autogenous milling technology took 19 years to increase power draw just 6 times from 1940 to 1959, and this was for a technology that had been around since 1907 in the goldfields of South Africa, (Weiss 1985), (Metso 2004).

The current model IsaMill, M10000, is powered by a 2.6MW motor, and has enabled large scale mineral processing applications to consider fine grinding as an economic option.

Once installed at Mt Isa and McArthur River, and the resultant success at the operations (as described later), MIM Holdings, allowed the IsaMill to be sold on a commercial basis. This still continues under Xstrata Technology, who actively promote the mill in mineral processing and leaching operations.
Today there are 27 IsaMills installed throughout the world. 14 are installed in Xstrata mines, McArthur River and Mt Isa, the rest are at non Xstrata mines. The mills have been installed by Xstrata Technology and Netzsch.

The technology, while developed to overcome serious liberation problems at the McArthur River and Mt Isa deposits, has become an industry standard in large scale fine grinding applications. It was probably due to its conception at a mine site, and the need to be reliable and robust at these operations that have made it accepted technology in fine grinding.

Of the 27 mills operating at mine sites, approximately 650T/Hr of material is now ground with the IsaMill, or 5.2M tonnes per annum. The product sizes ranges from a P80 of 7µm to 25µm, for materials ranging from lead and zinc sulphides, platinum concentrates, industrial minerals, iron oxide and refractory gold concentrate. Such a new technology has been embraced by those operations that rely on the finer grinding to achieve metal recovery.

A summary of the Australian lead/zinc operations using IsaMills are as follows:

<table>
<thead>
<tr>
<th>Operation</th>
<th>Material</th>
<th>Annual Treatment (T/Yr)</th>
<th>P80 (µm)</th>
<th>Recovery (% Metal)</th>
<th>Total Treatment Since IsaMill Installation (MT)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Mt Isa Zn Cons</td>
<td>350,000</td>
<td>15</td>
<td>&gt;80%</td>
<td></td>
<td>1.7 MT A</td>
</tr>
<tr>
<td>Mt Isa Pb Cons</td>
<td>260,000</td>
<td>15</td>
<td>&gt;80%</td>
<td></td>
<td>3.0 MT B</td>
</tr>
<tr>
<td>McArthur River Zn/Pb Cons</td>
<td>380,000</td>
<td>7</td>
<td>82%</td>
<td></td>
<td>2.9 MT C</td>
</tr>
</tbody>
</table>

A: 1994 to June 04  
B: 1999 to June 04  
C: 1995 to June 04 – this figure equates to 4.9MT of feed to IsaMills

Table 2: IsaMill Pb/Zn Operations in Australia

In short almost 1MT of lead/zinc concentrate is produced by IsaMills every year in Australia alone! The average of this material is well under 15µm, which before the development of the IsaMill would never have been economical to treat!
IMPACT OF IsaMills AT MOUNT ISA

As discussed earlier, the addition of more conventional secondary grinding and Tower Mill regrinding resulted in sphalerite liberation improving by 20% (from 55% to 75%), but recovery only improved by 5%. The lack of expected recovery was due to the extra difficulty of floating after grinding in ball and Tower Mills, i.e., the influence of iron media reducing the slurry.

In 1995 the first full scale prototype IsaMill (to become the M3000) was installed in the lead circuit. Not only did this improve lead performance, but zinc recovery in the zinc circuit also increased even though total sphalerite liberation had changed little. This was because the improvements due to grinding by iron free media from IsaMilling allowed sphalerite to be redirected from the lead concentrate to the zinc circuit. The lead circuit was able to operate more efficiently and separate the lead from the zinc, leaving the zinc to report to the following zinc circuit.

In 1999 the installation of additional IsaMills for the treatment of the George Fisher orebody, increased sphalerite liberation by a further 5%, but zinc recovery increased by 10% and zinc concentrate grade by 2%! This equated to a 16% recovery increase at the same concentrate grade. This demonstrates the fundamental changes to fine flotation made possible by ultrafine grinding in IsaMills with inert media compared to conventional means.

This is well described in the following graph, figure 7, showing the dramatic turnaround in liberated sphalerite that was recovered to zinc concentrate. During this period ore quality still deteriorated, as more Black Star and George Fisher was being mined, containing the fine grained ore, as well as increasing levels of carbonaceous pyrite.
Figure 7: Zinc Liberation and Recovery

**Relationship Between Sphalerite Liberation And Recovery Over 20 Years**

**BENEFITS OF IsaMills**

From the Mt Isa experience, size is important. It was important to get to 10µm to liberate the complex ore in the zinc circuit. The lead circuit was courser at 25µm. However even with the use of conventional mills and Tower Mills that could get to 25µm, the expected recovery was not achieved. It was only with IsaMills that the size and flotation performance improved. Why?

Three major benefits are gained from IsaMills over competing technologies reported by Pease et al 2004. They were:

1. **Ball & Tower Milling Added**
   - Liberation increased by 20%
   - Recovery increased by 5%
   - Recovery increased less than liberation due to negative impact of steel media on flotation of fines

2. **First IsaMills Added in Lead Circuit**
   - Sphalerite liberation constant
   - Zn recovery increased by 4% due to better rejection from Pb conc after IsaMilling

3. **IsaMills Installed in Zinc Circuit**
   - Liberation increased by 7%
   - Recovery increased by 10%
   - Conc grade increased by 2%
   - Recovery benefit higher than liberation increase because of improved flotation after IsaMilling
• **Impact of grinding on flotation performance.**

Milling using steel balls as media will effect the Eh as the balls reduce the pulp, creating negative redox potential. This reduces metal recovery, and will only improve as the flotation is oxidised. This is well documented as shown by work in figure 8 (Trahar 1984). The graph showing the impact of ceramic media is especially relevant to IsaMills as the sand media, like ceramics, are inert and has negligible effect on the pulp chemistry. Also conventional and Tower Mills generally have much higher residence times than IsaMills, resulting in lots of steel contamination. In these circuits additional reagents will be required, reducing selectivity or use innovative processes such as High Intensity Conditioning (HIC), eg at Hellyer, to reverse the negative impact of Tower Milling on surface chemistry. Processes like IsaMilling are far more efficient by providing this high intensity as part of the grinding action, and grinding in an inert environment.

![Figure 8: Redox Potential vs Recovery](image)

• **Power efficiency**

Even if size reduction could have been achieved by conventional milling, tower mills, or other stirred mill technology, only IsaMills provide efficient use of energy. Figure 9 compares the power required to grind a gold ore in a ball mill with 9 mm balls with an IsaMill with 2 mm media. The IsaMill is much more efficient below about 30 µm; to grind this ore to 15 µm would take 28 kWh/t in the IsaMill, but 90 kWh/t in a ball mill. Traditionally this has been attributed to the difference between attrition grinding and impact grinding. However by far the most important factor is media size, as shown by Figure 10, is the breakage rate. In Tower Mills this drops dramatically; the breakage rate for a 20 µm particles is ten times lower than the rate for 40 µm particles. Even though the Tower Mill is full attrition grinding, practically it is constrained to using relatively coarse media, 9mm balls in this case. In contrast, the IsaMill (Netzsch mill in Figure 11) can operate with much finer media and much higher intensity of power input (Table 3), meaning the peak breakage rate occurs at 20 µm, and doesn’t drop as quickly below that.
Figure 9: Efficiency of IsaMills Compared to Ball Mills

Figure 10: Particle Size vs Breakage Rate for Fine Grinding Mills
The influence of the sand media greatly improves the chances of contact of the grinding medium and particle. It comes down to simple geometry; the smaller the size the bigger the surface area, and the more particles per volume. This creates a greater chance of contact, as highlighted in Table 4.

<table>
<thead>
<tr>
<th>Feature</th>
<th>ISAMILL</th>
<th>TOWER MILLS</th>
<th>VERTICAL PIN MILLS</th>
</tr>
</thead>
<tbody>
<tr>
<td>Grinding Intensity (kW/L)</td>
<td>0.54</td>
<td>0.005</td>
<td>0.15 - 0.18</td>
</tr>
<tr>
<td>Residence Time to 15 µm (min)</td>
<td>0.6</td>
<td>154</td>
<td>7 - 9</td>
</tr>
<tr>
<td>Power Usage to 15 µm (kWh/t)</td>
<td>17.4</td>
<td>59.6</td>
<td>37.5 - 39.0</td>
</tr>
<tr>
<td>Media Material</td>
<td>Various</td>
<td>Steel</td>
<td>Steel</td>
</tr>
<tr>
<td>Media Size (mm)</td>
<td>0.8 - 1.6</td>
<td>9 - 12</td>
<td>6 - 8</td>
</tr>
</tbody>
</table>

Table 3: IsaMill, Tower Mill and Vertical Pin Mill Comparison

The influence of the sand media greatly improves the chances of contact of the grinding medium and particle. It comes down to simple geometry; the smaller the size the bigger the surface area, and the more particles per volume. This creates a greater chance of contact, as highlighted in Table 4.

<table>
<thead>
<tr>
<th>Power Intensity (kW/m³)</th>
<th>Media Size (m)</th>
<th>No. Balls / m³</th>
<th>Surface Area (m²/m³)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ball Mill</td>
<td>20</td>
<td>0.02</td>
<td>95541</td>
</tr>
<tr>
<td>Tower Mill</td>
<td>40</td>
<td>0.012</td>
<td>442321</td>
</tr>
<tr>
<td>IsaMill</td>
<td>280</td>
<td>0.001</td>
<td>1146496815</td>
</tr>
</tbody>
</table>

Table 4: Mill Comparison of Media Size, Power Intensity

- **Good classification**

Like all grinding operations, good classification is vital for power efficiency in ultrafine grinding. But the problem posed by particles less than 15 µm is how is it possible to classify this low using conventional technology? In short it is difficult. It is not generally practical to use cyclones to close-circuit a grinding mill with a target below about 15 µm. To get good cyclone efficiency at these sizes requires small cyclones, eg two inch (50 mm) diameter or smaller. This is virtually inoperable on a large scale, so the circuit is either compromised (and less power efficient) by using bigger cyclones, or an alternative solution is needed. The IsaMill at the Mt Isa operations achieved the cut size of 10µm in the zinc circuit by using an internal classifier mechanism, figure 11. This device uses the high centripetal forces generated inside the mill to classify the discharge, ensuring a very sharp product size without external cyclones. At the same time it prevents the fine media from passing out of the mill meaning that low cost media can be used, eg local sand, or granulated smelter slag. Also the very short residence time in the IsaMill also minimises “overgrinding”, further contributing to the sharp product size distribution. This novel classification device replaced the conventional screen that was on the prototype mills in the early 90’s, as it was found impractical to screen out fine particles.
IMPACT OF IsaMills at McARTHUR RIVER

As discussed earlier, the trialing of IsaMills at Mt Isa was to evaluate the use of this new technology for McArthur River. It was well documented the orebody was terribly fine, and was one of the reasons why the orebody had not been developed. It had been discovered in an exploration campaign in 1955 (Legge 1990), but the fine grain structure meant that conventional grinding couldn't liberate the grains fine enough. Figure 12 below, shows a sample of the ore compared to the courser grain structure of a Broken Hill sample (Pease, 2004)

Figure 12 : Different Grain Size of Broken Hill and McArthur River Ores
(Grey Square is 40µm)

However, the success of the IsaMill at the Mt Isa Lead/Zinc concentrator, meant that this deposit could be processed. The IsaMills were described as "enabling technology", and had the ability to grind down to 7µm to produce concentrate economically.

The plant started mid 1995 with 4 IsaMills, followed by another mill in 1998, and another in 2001. Circuit changes over the years have now resulted in a bulk concentrate being produced, where all rougher concentrate is presented to the IsaMills. A cycloning stage separates approximately ½ of the rougher concentrate for regrinding by the IsaMills. It is interesting to note that while the P80 target is 7µm, approximately 50% of the final concentrate is less than 2.5µm. If this was viewed from a number of particles basis, it would mean 96% of the particles are less than 2.5µm report to the final concentrate, (Pease 2004). This shows it is possible to grind fine and float fine particles.
INFLUENCE OF IsaMilling ON HYDRMETALLURGICAL APPLICATIONS

IsaMills are finding a potential use in fine grind leaching applications. It is currently a very important part in the Albion flowsheet, and has been looked at for other hydrometallurgical leaching processes for nickel and copper sulphides.

In the hydrometallurgical industry the Albion Process is one of the simpler processes. It is a process designed for the oxidative leaching of refractory base and precious metal bearing sulphide ores. The leaching occurs in conventional, non-pressurised reactors, which significantly reduces the capital cost compared to pressure and bacterial leaching processes.

What allows the leaching to be undertaken in relatively mild conditions is the addition of an IsaMill in the grinding circuit. The IsaMill produces mineral particles with a high degree of residual strain in the crystal lattice, and a very high surface area. This results in very high defect density within the individual mineral grains, resulting in the mineral being extremely active toward oxidation, (Hourn 1999). The conditions required to oxidise the mineral particle are less extreme than other leaching processes, with oxidation carried out at atmospheric pressure in agitated tank reactors. This is due to the mineral particle being highly fractured from the IsaMilling stage, which enables it to fall apart as it is leached. When the mineral is copper sulphide, the disintegration of the particles prevents the formation of the passive layer that prevents further leaching of the particles. The leach residence time is typically less than 24 hours, and the leach does not use any reagents other than acid, limestone and oxygen.

While other fine grinding technologies can be applied to the ALBION flowsheet, the IsaMill has the advantage of producing a very tight particle size distribution, as displayed in figure 13. The IsaMills are constructed with 8 grinding disc in a small chamber. Every disc acts as a separate grinding operation, therefore 8 disc implies 8 grinding operations. When this is coupled with the product separator, all chances of short circuiting are eliminated. However, other fine grinding practices have a tendency to produce a broad particle distribution as they short circuit some of the feed material.

The tight size distribution is an important factor in leaching operations. Where a flotation process can have slightly oversized particles, which can either report to concentrate or regrinding stages without any major effect on circuit recovery or grade, a leach circuit can be greatly impacted by the presence of large particles. Leaching processes need to have particles that are small enough to allow the leachant to fully leach the particle, otherwise the oversize will represent a recovery loss as the mineral hasn’t had the opportunity to be leached. That is why P98 is important in leach circuits.

Work conducted by Hydrometallurgy Research Laboratory (now a part of Xstrata Technology) conducted several test using different fine grinding bench scale mills, leaching copper sulphide concentrate using the ALBION process. While all fine grinding methods produced a similar P80, the IsaMill produced the finest P98. When the products were then leached, the IsaMill produced an extra 3% of leachable copper compared to other processes. This is displayed in figures 14, where the ratio of P98/P80 is plotted for each grinding process. The ratio of P98/P80 is an indication of how tight the feed sizing is, ie the closer to 1 the sharper the cut.
Figure 13 – Particle Size vs % Passing per Fine Grinding Method (Copper Bulk Concentrate)

<table>
<thead>
<tr>
<th>% Passing - microns</th>
<th>IsaMill</th>
<th>ECC Mill</th>
<th>GK Vibratory Mill</th>
<th>Metprotech Mill</th>
</tr>
</thead>
<tbody>
<tr>
<td>98</td>
<td>23.1</td>
<td>34.4</td>
<td>42.8</td>
<td>51.90</td>
</tr>
<tr>
<td>95</td>
<td>17.44</td>
<td>26.33</td>
<td>30.61</td>
<td>33.40</td>
</tr>
<tr>
<td>90</td>
<td>12.31</td>
<td>18.6</td>
<td>19.44</td>
<td>23.41</td>
</tr>
<tr>
<td>80</td>
<td>9.11</td>
<td>9.56</td>
<td>10.2</td>
<td>8.95</td>
</tr>
<tr>
<td>50</td>
<td>4.85</td>
<td>4.71</td>
<td>5.12</td>
<td>4.10</td>
</tr>
<tr>
<td>40</td>
<td>3.76</td>
<td>3.55</td>
<td>4.03</td>
<td>3.31</td>
</tr>
<tr>
<td>30</td>
<td>2.66</td>
<td>2.21</td>
<td>2.41</td>
<td>2.31</td>
</tr>
<tr>
<td>20</td>
<td>1.94</td>
<td>1.86</td>
<td>2.02</td>
<td>1.96</td>
</tr>
<tr>
<td>10</td>
<td>1.42</td>
<td>1.32</td>
<td>1.51</td>
<td>1.61</td>
</tr>
</tbody>
</table>

Table 5 – Fine Grinding Method vs Product Sizing (Copper Bulk Concentrate)
Finally, after the oxidative leach stage of the ALBION Process, the iron is precipitated out as goethite. This stage precipitates the iron and acid liberated during the oxidative leach, as well as neutralising any acid remaining in the slurry ahead of thickening and filtration. Goethite precipitates grow as coarse particles, considerably improving the settling properties of the finely ground leach residue.

The particle size distribution of the goethite residue is significantly coarser than the leach feed. A goethite precipitation stage also has the added benefit of precipitating any arsenic in the feed concentrate as a ferric arsenate, improving the stability of the arsenic phase in a tailings impoundment.
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Ultra Fine Grinding - A Practical Alternative to Oxidative Treatment of
Refractory Gold Ores

S. Ellis
Kalgoorlie Consolidated Gold Mines
Kalgoorlie, WA.

Abstract

Since early 2001, Kalgoorlie Consolidated Gold Mines (KCGM) has successfully operated an ultra fine grinding (UFG) circuit to supplement its roaster capacity for the treatment of the refractory gold flotation concentrate. A second UFG mill was installed in 2002 taking the total UFG capacity to over 20tph while grinding to 11-12 microns and achieving over 90% gold recovery. A development program in 2002/3 involving plant trials, pilot plants and laboratory testwork resulted in process improvements and a better understanding of the milling and gold leach processes that assisted in narrowing the gap between UFG and roasting. This paper details the operation of the ultra fine grinding process at KCGM as a non oxidative treatment for the extraction of gold from a refractory ores.
**Introduction**

The refractory nature of gold ores is often associated with gold finely disseminated in sulphide minerals, such as pyrite, at conventional grind sizes. Conventional milling can liberate the pyrite from the gangue allowing a low mass pyritic concentrate to be produced by a process such as flotation. However, direct leaching of the concentrate results in poor gold extractions as the cyanide lixiviant is unable to contact the gold locked or included within the pyrite (Figure 1).

![Figure 1 Pyrite Locked Gold Within an Ore](image)

The traditional approach for such refractory material has been to liberate the locked gold by chemically destroying the pyrite through oxidation. Roasting, pressure oxidation, and bacterial oxidation all employ various degrees of temperature, pressure and catalysis to react the pyrite with oxygen to produce an iron oxide and sulphur by-products. This method efficiently liberates finely disseminated gold or gold in solid solution.

Whilst such oxidative reactions are metallurgically sound and are capable of achieving high metal recoveries, the environmental aspects of treating the reaction products can alter the economics of the process.
For example, capture and disposal of sulphur dioxide from the roasting of sulphides or the neutralisation of the acidic liquors from pressure oxidation, may add significant additional costs to the process. In certain cases, these additional costs may make an alternative process route more economically attractive.

**Fine Grinding**

An alternative, applicable to the liberation of disseminated gold from the host mineral, is to continue the grinding process to further reduce the particle size of the host mineral thereby exposing a part of the gold surface for contact with cyanide solution. A benefit of this technique is that the host mineral is not destroyed in an oxidative chemical reaction with the resultant problems of treatment of the reaction products. Such fine grinding, however, has proven to be increasingly energy intensive with each size reduction step. In pit blasting, primary crushing, secondary crushing, SAG and ball milling are all able to exploit natural fracture planes in the ore allowing breakage along these features. As progressive size reduction occurs, the naturally occurring minerals are liberated and a point reached where the crystal structure of the mineral has to be broken for further size reduction to be achieved. This may present a significant barrier to further breakage with higher power intensities required to achieve a breakage event.

In the past, inefficiencies associated with conventional milling have made fine grinding unattractive to the mineral processing industry. The desired grinds for the harder minerals could only be achieved by prolonged milling with resultant low throughput and high power consumption. The use of smaller media in closed circuit can assist tumbling mills to achieve fine grinds but remain fundamentally limited in the manner in which they impart kinetic energy to the media as well as having large "dead zones" where little media movement occurs (Kalra, 1999). As finer media is used, the kinetic energy imparted to the media lessens thus significantly reducing the available energy transfer in a media/particle contact event.
Ultra Fine Grinding

UFG mills overcome these limitations by the use of rotating stirrers inside a stationary mill shell. Ultra fine grinding mills have been in use for many years in a large number of every day applications such as pharmaceuticals, dyes, clays, paint and pigments before being used in the mineral processing industry. They usually fine grind in a range of 1 µm to 10µm, and impart a significantly increased surface area as well as other potentially desirable properties such as colour, ease of absorption into the blood stream and increased chemical reactivity.

Figure 2 compares the power consumption of a laboratory ball mill to a UFG mill in grinding KCGM concentrate.

![Figure 2  Comparison of Grind Product Sizes](image)

The use of UFG grinding in the minerals processing industry is a relatively new development being based on the smaller low mass, batch UFG mills being used by other industries for high value products.
The chief requirement of the minerals processing industry was a mill that could process quantities in the order of several tonnes per hour in continuous operation whilst maintaining cost efficiency in power and media usage.

In achieving finer grinds, UFG mills use a finer media size (2-3 mm) than conventional milling (12-100mm) with a much higher installed power per mill unit volume (Table 1).

![Table 1. Typical Mill Grinds and Power Intensities](image)

<table>
<thead>
<tr>
<th>Type of Mill</th>
<th>Typical Lower Grind Size P80 µm</th>
<th>Power Intensity kW/m³</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ball Mill</td>
<td>75</td>
<td>20</td>
</tr>
<tr>
<td>Tower Mill</td>
<td>20</td>
<td>40</td>
</tr>
<tr>
<td>UFG Mill</td>
<td>5</td>
<td>280</td>
</tr>
</tbody>
</table>

Two basic types of UFG mills are available, the vertical stirred mill and the horizontal stirred mill. Both use rotating stirrers within a stationary mill shell to impart kinetic energy to a fine media charge (usually sand). The breakage mechanism is the same for the two mills, the differences being related to stirrer speed, method of media retention, and size of currently available mills.
Breakage and Particle Reactivity

As well as the resultant increase in the degree of liberation of the mill products, UFG also increases the surface area of the products enhancing the rate of downstream chemical reactions. The application of intensive non breakage stress events is believed to distort the mineral crystal lattice creating new defect sites which have high localised electron densities. These high electron densities facilitate the transfer of electrons to an oxidant thereby significantly increasing the rate of chemical reactions (Hourne and Halbe, 1999). This can result in a lowering of the activation energy for chemical reactions of the mill products and allow the reaction to proceed at lower temperatures and pressures than for the unmilled material.

This increase reactivity is demonstrated by the very high oxygen demand observed for pyrite after it is subjected to UFG.
The Activox and Albion processes make use of this phenomenon in their leach steps to achieve sulphide dissolution at reduced temperature and pressures to conventional pressure oxidation. However, for gold recovery, this increased rate of chemical reaction of the pyrite presents unwanted side reactions that can result in increased cyanide and lime consumptions.

**Sizing Measurements of the Product**

Over time, laser sizing has gained popularity and has now become the de facto standard for fine particle size measurement. Limitations with the earlier laser machines have been overcome with the development of sufficient on board computing power to use the Mei Equation relating light scattering characteristics to particle size. The laser measures an average particle volume and converts this data to an average particle diameter. The laser method is fundamentally different from a screen sizing which measures the diameter of a particle that can pass through a screen aperture.

Testwork conducted at KCGM (Turton-White, 2003) showed that the source of the greatest error in carrying out a laser sizing measurement was the sub-sampling of the bulk sample to the very small sample presented to the machine. It was shown that samples of greater than 50µm required an additional sub-sampling step to ensure a reproducible measurement.

Caution should be exercised with the practice of screen sizing out the coarse particles and laser sizing the screen undersize and combining both sets of results. Figure 4 highlights the different results than can be obtained when size fractions from a dry screening are analysed by the laser.
Method of Breakage

Stress Intensity has been examined as a critical determinant of the kinetic energy contained by the media in motion (Becker, 1997) and can be described in terms of the media diameter \( d \), the stirrer tip speed \( v \), and the media density \( \rho \).

\[
\text{Stress Intensity (Nm)} = d^3 \times v^2 \times \rho \quad \text{Equation 1}
\]

As shown in Equation 1, the Stress Intensity is related to the cube of the diameter of the media. For the same media type and mill speed, the Stress Intensity is increased by a factor of eight if the media size is increased from 3mm to 6mm. This highlights the importance of the inter-relationship between the top size of the mill feed, the selection of media size, and the wear on mill internals.
Kwade (1999) concluded that impact breakage rather than attrition was the main breakage mechanism in stirred mills for particles larger than 1µm. Recently Yue and Klein (2003) have examined breakage in successive grinding cycles and have found that a grinding limit exists where increasing grind time no longer results in particle breakage. They have postulated that this grinding limit is a function of the applied stress intensity and have noted that as this limit is reached, attrition rather than fracture becomes the chief breakage mechanism.

Sample Testwork

Not all refractory gold ores give a large recovery improvement after fine milling. Gold locked in arsenopyrite for example does not achieve the same gold recovery as gold disseminated in pyrite due to the smaller gold particle size of the locked gold (Figure 5).

Figure 5. Gold Recovery of Arsenopyrite and Pyrite Ores
Preliminary metallurgical testwork can be readily undertaken to determine the leach response to UFG. This involves the milling of a samples to a variety of nominated grind sizes and the leaching of the milled product. This not only determines the potential recovery but also gives a first pass economic assessment of the main economic drivers - power consumption, reagent costs, and gold recovery. Such preliminary data allows a comparison to be made with other processing routes.

Along with the determination of the grind/recovery curve, a mineralogical assessment of the gold deportment and particle size may provide additional information. Mineralogical scans such as QEMScan and MLA are able to give good data as to the mineral liberation and particle size of gold down to about 5µm. Below this size, an alternative method such as secondary ion mass spectrometry (SIMS) should be used.

**Ultra Fine Grinding at KCGM**

KCGM examined many concentrate treatment options that could provide an alternative to the roasting process in use. Chief among these were pressure oxidation, bacterial oxidation, and ultra fine grinding. An economic study carried out in 1997, determined that UFG had the superior Net Present Value return of the alternative options examined but was still a higher cost option than roasting as conducted at KCGM's Gidji site.

To progress UFG to the detailed engineering stage, further testwork was undertaken to define the flowsheet. A full pilot plant trial of the milling and leach process was carried out by Amdel (Adelaide) in 1999 which confirmed the recovery and power consumption indicated by laboratory scale testwork. The grinding of KCGM concentrate was amongst the highest energy consumers when compared to other UFG applications (Table 2); demonstrating that power consumption is very case specific.
Table 2. IsaMill Comparative Milling Data for Various Mineral Concentrates

<table>
<thead>
<tr>
<th>Type of Ore</th>
<th>Lead Rougher Concentrate (PbS)</th>
<th>Zinc Rougher Concentrate (ZnS)</th>
<th>Lead/Zinc Rougher Concentrate (CuFeS)</th>
<th>Copper Concentrate (FeS2)</th>
<th>KCGM Gold Concentrate (FeS2)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Feed F80 micron</td>
<td>25</td>
<td>30</td>
<td>65</td>
<td>120</td>
<td>120</td>
</tr>
<tr>
<td>Product P80 micron</td>
<td>10</td>
<td>10</td>
<td>10</td>
<td>10</td>
<td>10</td>
</tr>
<tr>
<td>Net Energy kWh/t</td>
<td>10</td>
<td>20</td>
<td>30</td>
<td>100</td>
<td>80</td>
</tr>
<tr>
<td>Throughput t/h</td>
<td>96</td>
<td>48</td>
<td>32</td>
<td>9.6</td>
<td>12.0</td>
</tr>
</tbody>
</table>

The work demonstrated that the recovery of gold from KCGM concentrate could be increased from 75% to 92% by grinding to 10µm.

**Mill Selection**

In the first instance equivalent samples of final concentrate were sent to all major mill suppliers for assessment. Results highlighted the lack of uniformity in the test methods and in the methods of sizing. No useful comparison could be made between results.

Site visits were useful in ascertaining the maintenance requirements of the mills, and general impressions of operators but with the variety ore types, hardness, and feed size to the mill any difference in metallurgical performance was difficult to gauge. From the preliminary assessment, the choice of mill was narrowed to the Detritor and the IsaMill. To discriminate between these mills, a pilot scale comparison was made by bring both the 18.5 kW pilot Detritor and the 55kW pilot IsaMill to site. The two mills performed similarly with no measurable difference in product quality or in power consumption under equivalent conditions. Given that no significant difference in metallurgical performance was apparent, KCGM opted for the IsaMill based on its mill size, low maintenance requirements, and its proven performance.
KCGM Flowsheet

For KCGM the main process risk lay with the unproven cyanide leaching of 10µm pyritic ore with acceptable reagent consumptions. In order to minimise this risk and to reduce capital costs associated with the process, KCGM determined to use single stage milling acknowledging that some inefficiency in power consumption and general mill and cyclone operation would result from the large size reduction step. To minimise the impact of the large feed size, a pre-treatment cyclone step was used to lower the feed size to the UFG mill by rejecting the coarser material to the cyclone underflow. The underflow was directed to the roasters for processing where the size was not of major significance. Whilst this approach assisted in reducing the feed size to the mill, it was subject to variations driven by the limited amount of suitably sized material in the concentrate. A further disadvantage lay in the pre-treatment cyclones directing an increased quantity of fine gangue material to the mill. The final flowsheet for the KCGM process is shown in Figure 6.

Figure 6. KCGM Flowsheet

The key design criteria for the process is shown in Table 3.
Table 3. UFG Key Design Criteria

<table>
<thead>
<tr>
<th>Item</th>
<th>Design</th>
</tr>
</thead>
<tbody>
<tr>
<td>Concentrate Processing Rate</td>
<td>10 t/h</td>
</tr>
<tr>
<td>Deslimed Concentrate Solids Size Distribution (F80)</td>
<td>120 microns</td>
</tr>
<tr>
<td>UFG Feed Prep Product Solids Size Distribution (P80)</td>
<td>50 microns</td>
</tr>
<tr>
<td>Initial Concentrate Gold Grade</td>
<td>40 to 50 g/t</td>
</tr>
<tr>
<td>UFG Feed Prep Product Solids Concentration</td>
<td>45 % to 55% w/w</td>
</tr>
<tr>
<td>UFG Product Solids Size Distribution (P80)</td>
<td>10 microns</td>
</tr>
<tr>
<td>Gold Recovery</td>
<td>92 %</td>
</tr>
</tbody>
</table>

Media Selection

The media size selected for the duty was based on the top size of the feed likely to be introduced to the mill. Pilot plant experience had shown that running too little media or too small a media size could result in a build up of unbroken feed top size in the mill. This build up in feed material within the mill led to a locking and centrifuging of the charge with a marked drop off in power draw. To ensure that the feed top size was adequately broken under all conditions, 6mm top size sand media was selected for the duty. Silica sand was selected due to its relatively low cost with initial supplies sourced from Northern New South Wales. Despite Australia wide searches for alternative supplies, only a few locations have been identified as having sand media of sufficient competency to achieve acceptable media consumption rates. Steel, smelter slag and ceramic medias were tested but rejected on the ground of cost, high consumption rates, and/or the introduction of deleterious materials (iron) to the leach process.
The use of a large media size (6mm) whilst ensuring top size particle breakage, resulted in severe wear on the mill internal components. Typically, wear associated with the leading mill disc, necessitated a mill stoppage every ten days compared to several months at other IsaMill installations running with 3mm media.

Media size is a critical determinant of the kinetic energy imparted to the media and is related to the cube of the diameter of the media. For the same media type and mill speed, the energy is increased by a factor of eight if the media size is increased from 3mm to 6mm. This highlights the importance of the inter-relationship between the top size of the mill feed, the selection of media size, and the wear on the mill internals.

**Cyclones**

Testwork had shown that closed circuit grinding with classification cyclones could significantly improve the mill throughput rate over open circuit. This was particularly apparent with the broad feed size distribution being presented to the mill as new feed. A key aspect of the KCGM flowsheet was the ability to provide a suitably sized product to the leach process for gold extraction whilst maintaining a suitable mill feed density (55%) with good fines rejection to overflow.

Closed circuit was also beneficial to the downstream gold leaching process in that the pyrite mineral carrying the gold have high SG's which preferentially reports to the cyclone underflow over an equivalent sized particle of a lower SG. This natural preference for high SG particles to report to the underflow, results in a further passage through the mill until such time as the reduced size counters the differential SG and allows passage out of the cyclone overflow and then to leach at a finer grind size.

KCGM selected 68mm cyclones for the classification duty on mill discharge. These cyclones have been essential in achieving a 10µm P80 product to leach but have presented a number of operational problems in spigots blockages and high wear rates.
With 10mm spigots, blockages were a constant problem with trash, scale, and sand media all contributing to blockages. Such blockages allowed a direct bypassing of mill discharge to leach with the resultant drop off in recovery. Significantly fewer spigot blockages have occurred since a vibrating screen was installed between the mill discharge and the cyclone feed hopper to remove this material. The main source of blockage nowadays is bridging of the spigot if the cyclone underflow becomes too high resulting in roping.

**Thickening**

The low pulp density of the cyclone overflow (8%) necessitated a thickening step to achieve the targeted 50% leach density. Testwork showed that the 10µm product would flocculate and settle rapidly at around 0.4 t/m²/hr using the standard non-ionic polyacrylamide flocculant in use at a dosage rate of 140g/t. Plant experience confirmed these figures with thickener underflow densities of up to 55% being readily achieved.

**Leaching**

The leaching of sulphidic ores is often problematic with a number of side reactions likely to occur that can lead to high reagent consumption and poor gold recovery (Deschenes, 2001). The KCGM concentrate to be leached not only contained a high percentage of pyrite milled to a very fine particle size but also cyanide soluble copper (0.15%) as chalcopyrite. The presence of telluride gold (calaverite) also added to the leach complexity requiring specialised leach conditions.

In the laboratory testwork and early plant practice, a very high oxygen demand and slow leach kinetics were noted. Typical gold recovery of around 90% was achieved at a cyanide consumption rate of 15 kg/t. A leach development program was conducted in 2002 where intensive laboratory and pilot leach tests were carried out to improve gold recovery and lower reagent consumptions. The testwork resulted in a revised leaching regime where pre-oxidation was abandoned, lead nitrate was introduced at the start of the
leach, and low dissolved oxygen levels (3ppm) maintained in the leach. It was also found that a high lime environment (30 kg/t) and use of carbon in leach (CIL) assisted the leaching of the gold tellurides.

By the use of these new leach conditions, the leach residence time was reduced from 72 hours to 24 hours, the cyanide consumption reduced from 15 kg/t to 8 kg/t, and average gold recovery increased by 2%. These improvements are shown in Figures 7, 8, and 9.

**Figure 7. Improved Gold Recovery**

![Gold recovery before and after revised leach regime](image-url)
Figure 8. Reduced Cyanide Consumption

Figure 9. Improved Leach Kinetics

<table>
<thead>
<tr>
<th>Leaching Time (hours)</th>
<th>% Gold Recovery</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
<td>Old leach conditions</td>
</tr>
<tr>
<td>10</td>
<td>New leach conditions</td>
</tr>
<tr>
<td>20</td>
<td></td>
</tr>
<tr>
<td>30</td>
<td></td>
</tr>
<tr>
<td>40</td>
<td></td>
</tr>
<tr>
<td>50</td>
<td></td>
</tr>
<tr>
<td>60</td>
<td></td>
</tr>
<tr>
<td>70</td>
<td></td>
</tr>
<tr>
<td>80</td>
<td></td>
</tr>
</tbody>
</table>
Despite the presence of an amount of sub 10µm material in the leach, no issue with the loss of carbon activity due to fine particles blocking pores in activated carbon has been apparent.

**Costs**

The capital cost of each of the KCGM UFG milling installation was $4.5 million. With ancillary equipment for the leach process the total project costs (including EPCM, owners management costs, and contingency) were around $6 million for each circuit.

Operational costs with a breakdown to the respective activity are shown in Figures 10 and 11.
The Future

With UFG being in operation for over two years at KCGM and having successfully proven the concept of economic gold extraction from refractory sulphide ore, albeit with lower gold recovery than roasting, the challenge is to further reduce process operating costs. The developmental work carried out has highlighted a number of areas where further improvements may be possible. Further work is in progress to quantify the benefits of introducing a primary milling stage. It is believed that a significant reduction in mill and cyclone wear components would result if a finer mill feed was produced as this would allow the use of a smaller media size thereby bringing wear rates in line with other similar mill users. Some improvement in power efficiency may also result from such an approach as well as some further improvement in gold recovery resulting from improved cyclone classification of a finer mill product.

The cost efficiency of ceramic media is likely to improve considerably in the future due to recent improvements in the quality and a lowering of the unit price with an increasing
market size. As the number of UFG mills increases, the availability of sand of sufficient size and quality may be limited.

The application of ultra fine grinding to refractory gold ores has proven to offer a viable alternative to conventional oxidative processes in particular cases. Its application to a hydrometallurgical extraction route for other metals beyond gold is open and likely to lead to alternative process routes for these metals also.

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ISAMILL FINE GRINDING TECHNOLOGY AND ITS INDUSTRIAL APPLICATIONS AT MOUNT ISA MINES

Mingwei Gao¹, Senior Metallurgist
Michael Young², Metallurgical Superintendent
Peter Allum³, Concentrators Manager

¹ MIM Process Technologies
Level 2, 87 Wickham Terrace,
Brisbane, Qld 4000,
Australia
PH: +61 7 3833 8500, Ext: 8505
E-mail: mingwei.gao@mim.com.au

² Mount Isa Mines Limited,
Mount Isa, Qld 4825,
Australia
PH: +61 7 4744 2011, Ext: 2926
E-mail: mfyoun@isa.mim.com.au

³ Mount Isa Mines Limited,
Mount Isa, Qld 4825,
Australia
PH: +61 7 4744 2011, Ext: 2640
E-mail: pjallu@isa.mim.com.au

Key Words: IsaMill, ultra-fine and fine grinding, grinding media, product size distribution, regrind circuit, Jameson Cell, tower mill, maintenance, wear components, gland seal, power intensity
Tough economical conditions and high grade but fine-grained ore processing have forced the mining industry to look for more efficient processes. IsaMill Technology for ultra-fine and fine grinding is one such process employed at several mining operations in Australia including Mount Isa Mines Limited in Australia. Mount Isa Mines is a business unit of MIM Holdings Limited. MIM is an Australian-based international mining and mineral processing company with around 8000 employees worldwide. Major products include copper, gold, zinc-lead-silver, coal and technology sales.

The IsaMill is a horizontal stirred mill with sizes up to 3 m³ chamber net volume and 1120 kW installed motor. It was jointly invented by Mount Isa Mines of Australia and Netzsch Feinmahltechnik GmbH of Germany for ultra-fine and fine grinding duties in the minerals industry. IsaMill operations started in 1994 at the Mount Isa Mines Lead/Zinc Concentrator and then at MIM’s McArthur River Mine in 1995. These IsaMills have been used to liberate sphalerite and galena at Mount Isa and silica at McArthur River by grinding to less than 7 micron.

This paper discusses the IsaMill Technology and reports on the metallurgical performance of the IsaMills at Mount Isa Mines, including work done over the last 9 years to improve mill design, process efficiency and maintenance, which has resulted in the current state-of-the-art IsaMill Technology.

The current lead/zinc deposits at Mount Isa Mines (George Fisher, Hilton and Isa) have an average grade of 10.5% of zinc, 7.2% of lead and 165g/t of silver. Fine grinding, in particular in the zinc retreatment circuit requiring a milled product P₈₀ of 7 micron, is essential to treat the ores from the Mount Isa Mines deposits. Every one micron size reduction in the zinc retreatment circuit below a P₈₀ of 10 micron improves the overall plant zinc recovery by 1%, resulting in an increase in revenue of about 1 million dollars per annum.

Eight IsaMills operate in the Lead/Zinc Concentrator at Mount Isa Mines, with two installed in 1994 and six commenced operation at the end of 1999. (Enderle et al, 1997; Johnson et al, 1998; Young et al, 2000). These are required for ultra-fine grinding to achieve liberation of fine-grained minerals, which improves separation of the minerals in the flotation circuits.

The process parameters and IsaMill designs have been continuously improved to increase the IsaMill grinding performance and the component life between maintenance periods. This paper reports the latest progresses in developing the IsaMill process and discusses the new features of the mills that have enhanced their performance.

IsaMill is a major development in grinding technology jointly invented by Mount Isa Mines of Australia and Netzsch Feinmahltechnik GmbH of Germany for ultra-fine and fine grinding duties in the minerals industry. The IsaMill was introduced by MIM over a period of 9 years, as there
was no existing ultra-fine grinding technology that could be used or modified for use in the minerals industry. The IsaMill technology demonstrated its ability for very power efficient grinding to less than 10 micron. The major hurdles that required extensive investment in time and money were the high wear rates, characteristic of this type of technology, and the separation of media from the product. Other technologies use exotic materials in construction and screens for these purposes but they are not consistent with a low cost, large scale and reliable grinding technology. The result of MIM’s development is a combination of smart component design and careful selection of rubber and polyurethane compounds to give long wear lives with relatively cheap materials.

IsaMills are currently operating in three mining operations in Australia. Namely, Mount Isa Mines, McArthur River Mine and Kalgoorlie Consolidated Gold Mines. A total of 15 IsaMills are in use at these operations. Figure 1 shows the eight IsaMills at the Mount Isa Mines Lead/Zinc Concentrator.

The IsaMill is a horizontal high speed stirred mill that operates with very high power intensities (up to 350 kW/m³). In comparison, the power intensity of a ball mill is about 20 kW/m³. The high power intensity enables the IsaMill to process fine particles at a high throughput that is essential for the economics of the minerals industry. The largest IsaMill currently available has a horizontally mounted grinding chamber shell of about 4 m³ in total volume (or approximately 3 m³ working volume). Inside the shell are rotating grinding discs mounted on a shaft which is coupled to a motor and gearbox. To allow quick and simple removal of the grinding chamber to expose the mill internals for maintenance purposes, the shaft is counter-levelled at the feed inlet end. The grinding discs agitate the media and ore particles in a slurry that is continuously fed into the feed port. A patented product separator keeps the media inside the mill allowing only the product to exit and simple control strategies based on power draw enable the IsaMill to produce a constant target product size. The invention of the product separator eliminates screens from ultra-fine grinding which delivers a process with the robustness required by the mining industry. The
product separator uses the high centrifugal forces generated to retain the media inside the grinding chamber.

The IsaMill uses fine media, stirring it with very high disc rotation speed, which increases the probability and energy of media/particle collisions. The mill has been designed to break particles using the attrition/abrasion mode of breakage where particle surfaces are chipped repeatedly producing very fine sized progenies at relatively low power consumption.

The compact IsaMill operates under pressure with "stage-by-stage" grinding occurring between the high speed rotating discs, which ensures grinding events are evenly distributed throughout the grinding chamber. This homogeneous grinding mechanism results in significant improvements to the particle size distributions from the feed to the product. The IsaMill Technology grinds the particles requiring size reduction, without over-grinding material at or below the required P 80. The uniform grinding mechanism is also the reason why IsaMill scale-up is 100% direct from laboratory to pilot to full scale (Weller et al 1999).

Currently, all operating IsaMills are installed with 1120 kW (1500 hp) motors and a grinding chamber with a net volume of 3 m³. A range of different size IsaMills below the 1120 kW motor model are also available. The schematic of the mill is shown in Figure 2.

![IsaMill Schematic indicating Major Components](image)

**Figure 2: IsaMill Schematic indicating Major Components**

**Mount Isa Lead/Zinc Concentrator Flowsheet**

The process flowsheet of the Mount Isa Mines Lead/Zinc Concentrator is shown in Figure 3. The circuit has been through various stages of modifications to address the declining ore qualities in the last ten years (Young et al 1997). It contains three IsaMills in the lead circuit and five IsaMills in the zinc regrind and retreatment circuits.
Figure 3: Mount Isa Mines Lead/Zinc Concentrator Flowsheet

The designed specifications of the IsaMills in the lead and zinc circuits are listed in Table 1. The product P80 sizing of 12 micron for the lead and zinc regrind IsaMills was required to liberate the galena and sphalerite from the gangue materials in the lead rougher concentrate, zinc column cleaner tailings and zinc rougher/scavenger concentrate. The zinc cleaner tailings were further reduced to a product P80 of 7.5 micron in the zinc retreatment circuit.

Table 1: Design Specifications of the IsaMills

<table>
<thead>
<tr>
<th></th>
<th>Lead Regrind</th>
<th>Zinc Regrind</th>
<th>Zinc Retreat</th>
</tr>
</thead>
<tbody>
<tr>
<td>Feed F80, micron</td>
<td>25</td>
<td>25</td>
<td>20</td>
</tr>
<tr>
<td>Product P80, micron</td>
<td>12</td>
<td>12</td>
<td>7.5</td>
</tr>
<tr>
<td>Net Energy, kWh/t</td>
<td>10</td>
<td>25</td>
<td>45</td>
</tr>
</tbody>
</table>

The ores from lead/zinc deposits in Mount Isa Mines contain minerals with a range of textural associations (Bojcevski et al, 1998). These textural associations allow a portion of the minerals to be liberated at coarse grind sizings (P80 = 37 micron), a portion to be liberated at fine grind sizings (P80 = 12 micron), while the remainder require ultra-fine grind sizings (P80 = 7.5 micron) for liberation.

This range of textural associations allows the use of a flowsheet using staged grinding and staged flotation to remove liberated valuable minerals with a coarse grind, while composite particles are reground for further liberation. This reduces the overall grinding energy consumption of the
flowsheet and decreases the flotation capacity required, as coarse particles have faster flotation kinetics than fine particles.

**THE ISA MILL PERFORMANCE**

**Lead Regrind Circuit**

The original lead regrind circuit was part of the bulk lead/zinc retreatment circuit that consisted of a tower mill operating in closed circuit configuration with 6 inch Kreb cyclones as shown in Figure 4. The cyclone overflow was processed by Agitair flotation cells. This circuit had problems of over-grinding the galena-bearing minerals as the cyclones returned the smaller but heavier galena particles and coarser but lighter gangue particles to the underflow for further grinding. The effect of this phenomenon is illustrated in Figure 5.

![Figure 4: Closed-Circuit Tower Mill for Treating Bulk Lead/Zinc Concentrate](image)

![Figure 5: Component Size Distributions of Mount Isa Mines’ Tower Mill Feed and Discharge](image)
With the introduction of the IsaMills in 1994, better liberation was achieved. This eliminated the bulk lead/zinc concentrate regrind circuit (Young et al, 1997). In 1999, three IsaMills configured in open circuit were installed in the lead regrind circuit, followed by a Jameson Flotation Cell. This new circuit is shown in Figure 6.

As part of the new circuit the recycle stream in the milling stage was eliminated to ease the problem of over-grinding the galena minerals. The new circuit also incorporated a Jameson Flotation Cell that generates finer bubbles to process the IsaMill product that has a P80 sizing finer than 12 micron. The immediate benefit from this simple circuit modification was to increase the overall lead recovery by 5%.

Another feature of the new circuit is the use of inert lead granulation slag as grinding media in the IsaMill. This improves the pulp chemistry and hence flotation selectivity, compared with using 12 mm steel balls in the tower mill.

**Zinc Regrind Circuit**

Figure 7 shows the flowsheet of the zinc regrind circuit. It includes a pre-cycloning stage, a closed-circuit tower mill and two open-circuit IsaMills that operate in parallel. The feed to the pre-cycloning stage includes the zinc column cleaner tailings and the zinc rougher and scavenger concentrate. As well as removing particles already at or below the required P80, the pre-cycloning stage also increases the feed density to the regrind circuit.
Figure 7: Zinc Regrind Circuit

Figure 7 shows that the IsaMills in the zinc regrind circuit reduced the feed F\textsubscript{80} of 20 micron to a product P\textsubscript{80} of 11.3 micron with a net energy consumption of 22 kWh/t. In comparison, the size reduction across the tower mill was only from a feed F\textsubscript{80} of 56.7 micron to a product P\textsubscript{80} of 49.4 micron.

**Zinc Retreatment Circuit**

Figure 8 shows the zinc retreatment circuit. There are three IsaMills in the circuit. The feed to the pre-cyclones in the retreatment circuit includes the zinc cleaner tailings and the zinc retreatment cleaner tailings. These streams are pre-cycloned to remove material already at or below the required P\textsubscript{80} and to increase the feed density to the IsaMills. The IsaMill product combines with the cyclone overflow and feeds to the zinc retreatment rougher flotation cells.

Figure 8: Zinc Retreatment Circuit
The feed to the zinc retreatment IsaMills is extremely hard to grind as it contains a large amount of composite particles with silica content as high as 30%. This is because the liberated valuable minerals have been removed in previous stages of flotation. Figure 8 shows the performance of one of the zinc retreatment IsaMills. It reduced a feed F<sub>80</sub> of 22.8 micron to a product P<sub>80</sub> of 7.8 micron with a net energy consumption of 48 kWh/t.

ISA MILL VERSUS TOWER MILL

The existing tower mill was in operation since 1991, firstly in the lead/zinc bulk concentrate regrind circuit, then in the original zinc regrind circuit. It was re-located to the current zinc regrind circuit in 1999 as the first stage of zinc regrinding. The purpose the tower mill was to produce a finer IsaMill feed, so the available IsaMills could be operated in open circuit with a fine lead slag media of 1 to 2 mm. This fine media sizing achieves the best power efficiency for products below 10 micron.

Using the survey data collected from the zinc regrind circuit, a comparison of the IsaMill versus the tower mill was possible. Figure 9 shows the size distributions of the feed and products for the tower mill. The difference in the 80% passing sizings of the feed (F<sub>80</sub>) and the products (P<sub>80</sub>) was in the order of 5 micron. Many surveys of the same tower mill in previous applications have also indicated that when being fed with a feed F<sub>80</sub> finer than 50 micron, the tower mill was capable of a size reduction of only 5 to 10 micron.

![Tower Mill Discharge Size Distributions](image)

Figure 9: Feed and Product Size Distributions for Zinc Regrind Tower Mill
In comparison, as shown in Figure 10, the size reduction across the zinc regrind IsaMill was from a feed $F_{80}$ of 55 micron to a product $P_{80}$ as fine as 12 micron, depending on the feed flow rate.

The excellent size reduction performance with the IsaMills is a result of using small media in the order of 1 to 2 mm. The advantage of operating with smaller media size is that it provides an increased number of media particles in the mill for grinding. For example, 1172 spheres of 1 mm media occupy the volume of only one 12.5 mm ball. Therefore, an IsaMill using 1 mm media will have 1172 times more media particles per mill volume available for grinding compared to a tower mill using 12.5 mm balls.

The IsaMill is also more efficient in using small media by stirring them with very high disc circumferential speed (21m/s), increasing the probability and energy of media-particle collisions. The IsaMill consists of a horizontal stationery mill chamber operating under a pressure of 100 to 200 kPa. This grinding mechanism ensures the grinding events are evenly distributed throughout the mill chamber. In comparison the tower mill is pressurised only by gravity, which results in the grinding events occurring mainly near the base of the tower mill’s vertical grinding chamber.

Another major difference in the performance of the IsaMill and the tower mill is the profile of the product size distributions as indicated in Figures 9 and 10. The tower mill product has a long tail at the finer end of the particle size distribution due to over-grinding of the fines. The IsaMill on the other hand grinds mainly the coarser part of the feed size distribution and produces a minimum amount of new fines. This ability of selectively grinding coarser particles in the feed is ideal to liberate the valuable minerals from the inter-locked particles and maximise their recovery in the downstream beneficiation processes. The tighter size distributions produced by the IsaMill allow the flotation circuit to operate more efficiently, since the flotation feed is more uniform. Optimising the flotation circuit is also made easier with a feed having a consistent particle size.
distribution. The IsaMill Technology efficiently grinds the particles requiring size reduction, without over-grinding material at or below the required P80.

**EFFECT OF SLURRY FEED SIZING**

Over the years of operating the IsaMills at Mount Isa Mines, it has been found that the slurry feed size distribution is important for the IsaMill performance when using local smelter slag as the grinding media. A finer feed size would normally help to reduce the energy consumption for a finer product when operated in open circuit. When the feed size becomes coarser the media sizing should be increased to provide the momentum and media-particle size ratio needed for breaking larger particles.

To demonstrate the effect of feed size, the IsaMill performance with feed F80s of 35 and 20 micron are compared in Figure 11. Reducing the IsaMill feed F80 from 35 to 20 micron enabled the IsaMill to produce a finer product P80 of 11.3 micron compared to 14.4 micron for the coarser feed. Also the net energy consumption was reduced by about 30% from 33 to 22 kWh/t.

The less efficient grinding when processing the coarser feed was due to the size of the fine slag media. Figure 12 shows the size distribution of the slag media collected during the survey. It contained about 50% fines below 0.5 mm, which is at the lower limit of media size for effective grinding in the IsaMill. A finer media normally helps to produce a finer product but has limited ability to cope with a coarser feed.

For an IsaMill operating in open circuit configuration with a fine grinding media, the feed size needs to be controlled operationally to achieve target grind sizes at maximum power efficiency.

![Figure 11: Effect of Feed Size on Zinc Regrind IsaMill](image-url)
**Effect of Media Sizing**

The tower mill circuit prior to the zinc regrind IsaMills is required to maintain the IsaMill performance when the feed size becomes coarser. Another option is to coarsen the slag media size used in the IsaMill to process the coarser feed. If an ultra-fine product is required with both coarse feed and coarse grinding media then the IsaMill should be operated in closed circuit to maintain power efficiency.

Trials with different lead slag media sizes were conducted in one of the zinc regrind IsaMills, with unscreened slag containing about 50% fines below 0.5 mm and screened slag containing about 8% fines below 0.5 mm. Figure 13 shows the size distributions of the screened and unscreened slag.

![Figure 13: Effect of Media Sizing](image_url)
Figure 14 shows the plant survey data using the same feed material, but with screened and unscreened slag as the media. The feed F_{80} was 30 micron in both cases. The figure shows that screened slag media achieved a P_{80} of 15 micron using less than 30 kWh/t, compared to unscreened slag which required more than 40 kWh/t for the same product size.

Selection of the grinding media size is undoubtedly one of the key criteria for achieving the maximum IsaMill power efficiency.

![Figure 14: IsaMill Performance with Screened and Unscreened Lead Slag Media](image)

**NEW DESIGN FEATURES OF THE ISA MILL**

The initial prototype IsaMill installed at Mount Isa Mines in 1993 had a grinding chamber volume of 1500 litres and operated with 5 grinding discs. The purpose of this mill installation was to aid in the design of the full-scale production IsaMills. The first production mills installed at MIM’s Mount Isa and McArthur River mines operated initially with 6 discs. Subsequently the number of grinding discs was expanded to seven and then to the current standard of eight discs. The net mill volume of the production mills with eight discs was 3000 litres.

Continuous development and improvement of the IsaMills from 1994 to the present has lead to a design providing superior grinding performance, increased component life, and decreased downtime during routine maintenance periods.

With the developmental focus of the IsaMill being minimum operating costs and maximum availability, the equipment has been designed for fast and straightforward maintenance. Locating all major components at ground level with the shell components on rails results in very easy access to the wear items. The major mechanical items such as motor and bearings are standard components with typical maintenance programs and life expectancy.

**Internal Wear Components**

The materials of construction and design, and method of fabrication of all IsaMill internal wear components have improved since the first installation. Over 20 different types of wear materials
have been tested. These tests determined the characteristics that components need for long wear life.

The internal steel surfaces on the original production IsaMill grinding chambers and end flanges were cold rubber lined. These wear components were required to be sent to the nearest rubber lining workshop after a certain amount of operating time for relining. This meant that a spare set of these steel components had to be held in stock.

To avoid this requirement components of the feed and discharge end flanges were redesigned, so that only a replaceable rubber wear component had to be held in stock. These rubber components simply slide on to the steel surfaces and then are bolted in place. This allowed improvements in the fabrication technique of these rubber components, which lead to an improvement in wear life.

The final challenge in terms of wear components was the IsaMill grinding chamber. The latest IsaMills were designed with a split shell (Figure 15), in which a replaceable slip-in liner could be installed. The IsaMill shell can be unbolted, opened into two halves and the worn shell slip-in liner removed and a new one installed. This removes the need to send the shell away to be cold rubber lined and the need to have spare shells to install when one is worn. All wear components are fabricated off-site then stored on-site until required. All components are steel backed with rubber or similar coating. These are relatively low cost, light and easy to handle and store, but still provide long wear lives.

![Figure 15: IsaMill with a Split Shell Design](image)

**Shaft Sealing System**

In addition to the internal wear components of the IsaMill, the shaft sealing system also requires routine maintenance. This system prevents slurry and grinding media discharging from the point where the rotating shaft enters the grinding chamber. The gland sealing system can be adjusted
externally while operating, similar to a slurry pump, and can be maintained when the discs are removed from the shaft during scheduled maintenance.

The commercial full-scale IsaMills use gland packing and gland water sealing for sealing the shaft. The design of this gland sealing system was a great improvement over several of versions trialed during the scale-up phase of the IsaMill development program. The gland life of these trial systems was shorter than desired, especially since the life of the internal wear components (grinding discs and shell liners) had been extended.

Improvements were made to install an independent gland water supply system that increased the gland life. The new design for the gland system also includes an increased bearing spacing to decrease the deflection of the shaft at the gland, and introduces water flushing on the slurry side of the gland to reduce the amount of slurry and media in this area. The new gland system has been working well since installed about 2 years ago.

**Mill Access**

Figure 16 shows the support system for the IsaMill. The mill grinding chamber shell and discharge end flange are both fixed to trolleys which are mounted on four wheels that sit on rails. The IsaMill has hydraulic rams on the rails under the wheels. These rams are used to move the grinding chamber and discharge end flange along the support rails so that the internal components of the mill are exposed for maintenance. The original installations had fixed hydraulic power packs on each IsaMill, which tended to get covered in dust and other debris, thus requiring more than expected maintenance. The new IsaMill design has a portable hydraulic power pack that plugs into the rams on the mill when maintenance is required and is then stored in a clean area after maintenance has been completed.

This quick and simple system to access the mills is achievable since the internal shaft is counter-levelled at the feed inlet end. In other words there are only bearings at the feed inlet end of the mill and at the discharge end the shaft is free and does not protrude through the end flange.
MAINTENANCE PROGRAM

Maintenance of the IsaMill is as simple as routine maintenance for a slurry pump. As previously mentioned the internal rotating shaft is counter-levered at the feed inlet end. This means that the discharge end flange and grinding chamber can be removed to expose the mill internals without having to remove bearings, shaft etc. The discharge end flange is firstly unbolted from the grinding chamber and the trolley fixed to the flange is then connected to the hydraulic ram and driven out along the rails. Next the chamber shell is unbolted from the feed inlet flange and is also moved along the rails using the hydraulic ram to fully expose the mill internals, including shaft, grinding discs and product separator.

To remove the grinding discs the locking nut at the end of the shaft is undone so that discs can be slid off the shaft. New discs are then slid on the shaft. The bolt-on wear components on the feed and discharge end flanges and the shell liner can then be replaced, if required. After replacing any worn components the chamber shell is moved back into position using the hydraulic ram and rebolted to the feed inlet flange. The discharge end flange is then wheeled up and bolted against the chamber shell. The IsaMill is now ready for to be put back into service.

The maintenance schedule is built around the grinding discs. The average life of a set of discs at Mount Isa is about 4000 hours. The other components last several cycles of disc replacement. The lives of all the components are continually being improved to decrease downtime and maintenance costs.

The maintenance program at Mount Isa Mines is currently based upon a 1,500 hour operating cycle. At the end of each cycle the IsaMill is shutdown and opened for inspection and replacement of wear parts if required. The shutdown has a total duration of 8 hours.

The inspection has the following procedure:

- Shutdown of mill
- Flushing of mill followed by dumping of mill contents
- Removal of end flange bolts
- Removal of Shell
- Inspection of discs, shell liner, product separator and gland seal for wear
- Replacement of any worn discs
- Replacement of gland seal if required
- Re-assembly of shaft, discs and spacers
- Replacement of shell and end-flange
- Recommencement of operation

On each third operating cycle a 16-hour shutdown is undertaken for detailed inspection of bearings and tolerances etc.

At Mount Isa Mines the shell liner, product separator and end-plate linings have typical wear lives of 18 months. The shaft sleeves (disc spacers) are low wear items and their life significantly exceeds that of the other mill internals.
The gland seal packing is comparable to that used in slurry pumps and can be expected to have a similar life. Clean and consistent water supply will ensure maximum life. Recent design changes as mentioned above has increased gland life by at least a factor of two. Gland packing life is estimated to be at least 6 months.

At Mount Isa Mines the overall IsaMill availability on an annual basis is 99%.

**BENEFITS TO PLANT METALLURGY**

The installation of eight IsaMills since 1994 in the Mount Isa Mines Lead/Zinc Concentrator, along with other circuit modifications, has increased the liberation of galena and sphalerite and increased the metal recoveries at the target grades.

Figure 17 shows the concentrator zinc recovery performance for the last twenty years with recent operation reaching the highest zinc recoveries. Significant liberation of valuable minerals as a result of using the IsaMill technology is one of the most important technical breakthroughs that has contributed to the improved plant metal recoveries.

**SUMMARY**

IsaMill technology was successfully developed and operated at Mount Isa Mines over the last nine years. Currently 15 full-scale IsaMills operate in three mining locations in Australia, economically producing products as fine as 7 micron.
Compared to other fine grinding technologies, the IsaMill has high power input per mill volume (up to 350 kW/m³) and operates under pressure (100 to 200 kPa). These two features enable the small grinding media of a few millimetres to work effectively to process a wide range of feed sizes and to produce the required fine products.

Media selection is important for the IsaMill operation. Mill scale-up data can be obtained from laboratory data but design of IsaMill circuits needs to be conducted carefully. Feed size and media size effects as well as variations in the circuit need to be appreciated.

The new IsaMill design has increased the grinding performance and improved the maintenance cycle giving greater component life and ease of maintenance.

ACKNOWLEDGMENT

The authors wish to acknowledge the management of Mount Isa Mines Limited and MIM Process Technologies for permission to publish this paper. For this work, the important input of many employees at Mount Isa Mines Limited and MIM Process Technologies is acknowledged, as well as the contribution of their colleagues at Netzsch Feinmahltechnik GmbH.

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IsaMill Ultrafine Grinding for a Sulphide Leach Process

by
Harbort, G(1), Hourn, M(2), Murphy, A(1)

(1) MIM Process Technologies
(2) Hydrometallurgical Research Laboratories

ABSTRACT

Fine grinding mills have improved in design and efficiency in recent years, allowing major opportunities for treatment of materials where liberation to grind sizes below fifteen microns are required. The successful development of the IsaMill, a horizontal stirred mill, has produced equipment capable of grinding the larger tonnages which exist in mineral processing operations, to product sizes below ten microns. Initially developed for use with base metal flotation circuits, significant test work conducted in 1998 shows that major economic gains can be achieved by producing finely ground material for leaching.

IsaMill development and operation is reviewed. Results from test work to produce a feed stock for both straight cyanidation and sulphide leaching are discussed, with emphasis on energy consumption, product size and leachability. A number of options for ultrafine grinding and leaching are also discussed.

INTRODUCTION

In operations where flotation products are produced as a saleable concentrate, decreasing liberation size may result in decreased metallurgical performance due to gangue impurities and lower recoveries. Ultra fine grinding is gaining significant acceptance as a cost effective means to provide an optimum grade/recovery response in flotation.

In some gold leaching operations, ores contain gold in close association with sulphide minerals such as pyrite and arsenopyrite. Ultrafine grinding provides an effective method of liberating physically locked gold, or for producing a feedstock that is amenable to oxidative processes.

ISAMILL DEVELOPMENT

Decreasing liberation size and increased amounts of refractory pyrite in Mt Isa’s lead/zinc ore resulted in a gradual decrease in concentrator metallurgical performance over time. Work had been conducted at Mt Isa between 1975 and 1985, involving regrinding to ultra fine sizes to increase liberation, using conventional grinding technologies. It was found that the conventional technologies had a very high power consumption to achieve the required sizing and the flotation performance was worse than expected due to contamination from the iron media used.

In 1990, there was no accepted technology for regrinding economically to ultra fine sizes in metalliferous operations. Test work in 1990 and 1991 indicated that high speed horizontal mills could efficiently grind to a product of 80%...
passing 7 microns at laboratory scale and subsequently provide a major increase in metallurgical performance. To make an ultrafine grinding mill capable of treating the tonnages required at Mt Isa, a program of development was undertaken between Mount Isa Mines Limited in association with NETZSCH-Feinmahltechnik GmbH (NFT).

Scale-up was tested using trial installations at the Hilton and Mt Isa lead/zinc concentrators. By the end of 1994, the first full scale IsaMill (1.1MW) was installed in the lead/zinc concentrator. This has allowed a suitable, low cost grinding media to be proven in operation and provided a system for separating grinding media from product, without the disadvantages of screens or sedimentation zones. Furthermore, it has allowed development of cost effective wear materials. A diagram of the 1.1MW IsaMill is shown in Figure 1.

In 1998 the rights for commercialisation of the IsaMill where transferred from Mount Isa Mines Limited to MIM Process Technologies and under an exclusive agreement with NFT, on the 14th of December, 1998 the IsaMill technology was launched to the metalliferous industry as a cost effective means of grinding down to and below 10 microns. There are currently two 1.1MW IsaMills operating in Mount Isa, treating lead concentrate and five 1.1MW IsaMills operating at the McArthur River site, treating bulk lead/zinc concentrate. A further six 1.1MW IsaMills will be installed in Mt Isa to treat zinc concentrates from the George Fischer Mine, commencing June, 1999.

APPLICATIONS

Ultrafine Grinding for Flotation

Several 1.1MW IsaMills have been installed in both the lead and zinc cleaner flotation circuits at Mount Isa.

Operations on lead rougher concentrate have shown that a throughput of 73tph per mill could be achieved at 45% solids. The net energy consumption for reducing a feed of 20 microns to a product 80% passing twelve microns was 6kWh/t. Figure 2 shows the improvement in silica rejection with and without the IsaMills in operation.

IsaMills were also installed to treat zinc retreatment circuit cyclone underflow. From a feed of 80% passing 46 microns a product of ten microns could be produced with a specific energy of 65kWh/t. Figure 3 shows the improvement in the zinc grade/recovery curve after ultrafine grinding with the IsaMill.

Of specific interest in the Mount Isa operations is the wide choice of grinding media available for use in the IsaMills. Mt Isa has variously used heavy media plant reject and lead or copper smelter slag for grinding media. These materials had previously been considered waste and provided an extremely inexpensive grinding media. In addition they are inert, having no effect on flotation chemistry, unlike steel media.

Ultrafine Grinding for Direct Cyanidation

Many refractory gold bearing ores contain gold in close association with sulphide minerals, such as pyrite and arsenopyrite. Refractory gold may be present in several forms, ranging from fine free gold housed on boundaries between mineral grains, to gold that is in solid solution with the sulphide matrix. Techniques used to recover gold from refractory ores range from ultrafine grinding for improved liberation of physically locked gold, to oxidative processes where the
sulphide matrix is wholly or partially destroyed through chemical oxidation.

The effectiveness of ultrafine grinding as a method of liberating physically locked gold is illustrated in Table 1. The data in Table 1 is also displayed graphically in Figure 4.

A sample of pyrite ore, grading 4.59 g/t gold and 4.85 % sulphur was finely ground in a laboratory scale IsaMill to 80 % passing 20 and 10 microns respectively. The finely ground product was then leached in a conventional agitated cyanide leach test for a period of 48 hours at pH 10, and a free cyanide level of 500 ppm.

<table>
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<th>80 % passing size - microns</th>
<th>% gold recovery</th>
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</table>

Table 1. The Effect of Ultrafine Grinding on Gold Recovery from Pyritic Ore.

Pressure leaching involves the oxidation of sulphide minerals in acidic solutions at elevated temperatures and pressures, where the aggressive leaching conditions are used to improve the kinetics of the oxidation process.

Ultrafine grinding resulted in a significant improvement in the gold recovery through cyanidation. The most common oxidative processes used to recover gold from refractory sulphides are pressure leaching, bacterial leaching and roasting.

Roasting involves the reaction of sulphide minerals with hot gasses containing oxygen, at temperatures in the range 500 - 800 °C. Sulphur dioxide gas, a byproduct of the roasting process, is usually captured for production of sulphuric acid.

In recent years, there has been a considerable amount of development carried out on processes which couple ultrafine grinding and oxidative leaching in an effort to reduce the capital costs associated with Pressure Leaching, Bacterial Leaching and Roasting. One such processes is MIM Holdings’ ALBION Process.

CASE STUDY

Comparative fine grinding and oxidative leaching testwork was carried out on a sample of pyrite concentrate using MIM Holding’s proprietary ALBION process and two common laboratory scale ultrafine grinding mills. The two mills tested were the IsaMill and a Vertically Stirred Mill (VSM). A head analysis of the pyrite concentrate is listed in Table 2.

<table>
<thead>
<tr>
<th>Element</th>
<th>Assay</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fe</td>
<td>28.5</td>
</tr>
<tr>
<td>S</td>
<td>23.6</td>
</tr>
<tr>
<td>Au</td>
<td>50</td>
</tr>
<tr>
<td>Ag</td>
<td>18</td>
</tr>
</tbody>
</table>

Table 2. Head Analysis of Pyrite Concentrate Sample used in the Fine Grind/Oxidative Leach Testwork.
Samples of ground concentrate were produced from both mills at a range of particle size distributions. The ground concentrate samples were then leached under the conditions typically specified for the ALBION process, which involves oxidation of sulphide minerals at atmospheric pressure in conventional agitated tanks. Cyanide leaching of an oxidised concentrate produced from each type of mill was also carried out.

**Experimental Setup**

The laboratory scale IsaMill consisted of a horizontal milling chamber, drive motor and 6 disc radial impeller. The mill was fitted with a 4.0 kW variable speed drive and Yokogawa power meter to determine the specific energy requirements of each grind. A diagram of the laboratory ISAMILL set up is shown in Figure 5.

![Diagram of the IsaMill Experimental Setup](image)

**Figure 5. Diagram of the IsaMill Experimental Setup**

The pyrite concentrate sample was pumped by variable speed peristaltic pump from a feed tank to the IsaMill feed port. Slurry exited the mill through a separator located before the mill mechanical seal, which allowed slurry to pass but retained media within the mill.

Ground slurry was discharged from the mill through a slurry discharge port and was collected in a sealed product tank. Samples of ground slurry were collected for the leaching tests, and each of the slurry samples was filtered and the filter cake stored frozen prior to leaching.

The Vertically Stirred Mill consisted of a 10 litre chamber and pin style impeller. The mill shaft held seven radial arms, each 160 mm long and 10 mm in diameter. The mill motor was a 2.7 kW variable-speed drive, fitted with a 3 phase AC inverter for power consumption measurements. The mill media was 2.3 – 6.6 mm steel shot. A diagram of the laboratory Vertical Mill set up is shown in Figure 6.

![Diagram of the Vertically Stirred Mill Experimental Setup](image)

**Figure 6. Diagram of the Vertically Stirred Mill Experimental Setup**

Feed slurry was pumped through the mill feed port located at the base of the mill, and was discharged from the mill overflow port. Ground slurry was collected in a sealed product tank. Each of the slurry samples was filtered and the filter cake stored frozen prior to leaching.

**Results.**

The results of the ultrafine grinding and oxidative leaching tests are listed in Table 3. Figure 4 shows the effect of particle size on the extent of sulphur oxidation achieved for the two different types of fine grinding mill.

The IsaMill was the most efficient of the two mills tested, grinding to 80 % passing 8.61 microns at a specific energy input of 64 kWh/t. In comparison, the Vertically Stirred Mill required 105 kWh/t to achieve a product at 80 % passing 8.2 microns. The relationship between specific energy and particle size for both of the mills is shown in Figure 7.

The specific energy curves for the two mills were similar for product sizes down to 80 % passing 25 microns, at which point the slope of the curve for the Vertically Stirred Mill increased significantly relative to the IsaMill. The curves in Figure 4 show that the IsaMill was much more efficient at grinding in the particle range below 12 microns.
Table 3. Results of ultrafine grinding and oxidative leaching of a pyrite concentrate sample.

<table>
<thead>
<tr>
<th>Mill Type</th>
<th>Specific energy Input – kWh/t</th>
<th>80 % passing size</th>
<th>% sulphur oxid.</th>
</tr>
</thead>
<tbody>
<tr>
<td>Head</td>
<td>0</td>
<td>108.39</td>
<td>24</td>
</tr>
<tr>
<td>IsaMill</td>
<td>11</td>
<td>51.9</td>
<td>61</td>
</tr>
<tr>
<td>IsaMill</td>
<td>19</td>
<td>36.4</td>
<td>76</td>
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<tr>
<td>IsaMill</td>
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<td>17.94</td>
<td>91</td>
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<td>IsaMill</td>
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<tr>
<td>IsaMill</td>
<td>64</td>
<td>8.61</td>
<td>96</td>
</tr>
<tr>
<td>VSM</td>
<td>12</td>
<td>41.5</td>
<td>74</td>
</tr>
<tr>
<td>VSM</td>
<td>24</td>
<td>26.7</td>
<td>88</td>
</tr>
<tr>
<td>VSM</td>
<td>75</td>
<td>10.68</td>
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</tr>
<tr>
<td>VSM</td>
<td>105</td>
<td>8.2</td>
<td>96</td>
</tr>
<tr>
<td>VSM</td>
<td>148</td>
<td>7.54</td>
<td>94</td>
</tr>
</tbody>
</table>

Figure 7. Comparative Milling Curves for the IsaMill and Vertically Stirred Mill.

In terms of breakage rate, the mill residence time to achieve a specified grind showed a similar pattern to that of specific energy. The relationship between mill residence time and particle size for both of the mills is shown in Figure 8. The residence time for the two mills were similar for product sizes down to 80% passing 25 microns and below this point the IsaMill required significantly less residence time to achieve a specified grind. At the target grind of 80% passing ten microns the IsaMill required 12.49 minutes grinding time, compared to 18.46 minutes residence time for the Vertically Stirred Mill.

The superior efficiency of the IsaMill in grinding to below 12 microns was believed to be due to the smaller media used in the mill. It is well known that the efficiency of any fine grinding process improves with finer media, and the mechanics of the IsaMill allow for agitation of finer media at higher tip speeds than is possible for a vertically stirred mill.

Figure 8. Comparison of residence time for the IsaMill and Vertically Stirred Mill.

The degree of oxidation achieved from leaching the concentrate samples did not show a large variation with type of mill, as indicated by the closeness of the oxidation versus particle size curves in Figure 5, below. Both mills achieved a maximum of 96% oxidation at approximately 80% passing 15 microns.

Figure 9. The Effect of Particle Size on Sulphide Oxidation for the two Mill Types

CONCLUSIONS

The use of ultra fine grinding in combination with cyanidation or chemical leach processes is a viable alternative to traditional expensive treatment of refractory gold concentrates.

A laboratory scale case study comparing an IsaMill (a horizontal stirred mill) and a vertical stirred mill has shown that a more energy efficient outcome is likely with the former.

In addition, a superior residence time performance of the IsaMill suggests a more compact, lower cost mill, when installed.
Developments in Milling Practice at the Lead/Zinc Concentrator of Mount Isa Mines Limited from 1990.

M.F. Young¹, J.D. Pease², N.W. Johnson³ and P.D. Munro⁴.

Submitted for:
AusIMM Sixth Mill Operators Conference, 6-8 October 1997
Madang, Papua New Guinea.

ABSTRACT

The Lead/Zinc Concentrator of Mount Isa Mines Limited processes complex fine grained ore from the Isa and Hilton silver-lead-zinc orebodies, producing lead concentrate, zinc concentrate and (until 1996) a low grade middlings (LGM or bulk) concentrate.

Metallurgical performance declined dramatically during the 1980’s because of declining ore quality, as the ore became both finer grained, and contained increasing amounts of refractory pyrite. Developments in milling practice to restore performance focused on two areas: liberation and separation. Increased mineral liberation was achieved by more than doubling grinding and regrinding capacity to increase sphalerite liberation. This successfully recovered an extra 20 per cent zinc metal to zinc concentrate, which previously reported to final tailing, lead concentrate or LGM concentrate. There was also a small increase in galena liberation, increasing lead recovery to lead concentrate and reducing contamination of the zinc concentrate by lead. The increased sphalerite and galena liberation also significantly reduced the production of LGM concentrate.

The second area of development was to improve the separation of galena and sphalerite from gangue minerals. This was achieved by circuit rationalisation, better understanding of water chemistry leading to an improved reagent scheme, and improved process control. These changes both improved performance and simplified the circuit, giving better and steadier concentrate grades and recoveries.

The combination of increased liberation, improved separation, and circuit simplification has dramatically increased the metallurgical performance of the Lead/Zinc Concentrator when treating very complex fine grained ore.

2. Lead/Zinc Concentrator Manager, Mount Isa Mines Ltd, Mount Isa Qld 4825
3. Minerals Processing Research Manager, Mount Isa Mines Ltd, Mount Isa Qld 4825
4. Principal Engineer, Metallurgy, MIM Holdings, 410 Ann Street, Brisbane Qld 4000
INTRODUCTION

Processing of Mt Isa silver-lead-zinc ore commenced in 1931 at the No.1 Concentrator, initially treating a mixture of oxidised and sulphide ore, but by 1935 treating only sulphide ore (Kruttschnitt et al, 1939). Over the years the flowsheet was developed to improve metallurgical performance of the fine grained and difficult to treat Mount Isa ore (Kruttschnitt et al, 1939, Cunningham, 1953, Challen et al, 1968, Davey and Slaughter, 1970). Over the period 1952 to 1960, ore reserves were increased substantially, and the decision was made to increase the treatment rate, firstly by modernising and expanding the existing No.1 Concentrator, and secondly by constructing a new (No. 2) Lead/Zinc Concentrator (Challen et al, 1968).

The No 2 Concentrator was commissioned in June 1966 and total silver-lead-zinc ore treatment was transferred from the No. 1 Concentrator in May 1967. Various improvements in the 1970’s (Bartrum et al, 1977) were followed by the installation of larger flotation cells in a single circuit conversion (Johnson et al, 1982). A major increase in throughput occurred in 1982 with the commissioning of a Heavy Medium Preconcentration plant to reject 30 per cent of waste ore before flotation (Fiedler et al, 1984). From 1987, ore from the nearby Hilton mine was introduced to supplement the Mt Isa ore. By 1992, treatment rate had reached 5 Mt/y, of which 30 per cent was from the Hilton Mine and 70 per cent from Isa Mine.

During the 1980’s, the increase in throughput and decline in head grade were exacerbated by a significant increase in ore complexity. This resulted in a severe liberation problem, with a finer mineral liberation grain-size than the plant grinding capacity could achieve, and a worsening separation problem caused by increasing amounts of naturally floating carbonaceous pyrite. Metallurgical performance declined dramatically as the plant did not have technologies to deal with either problem. This paper chronicles the change in ore characteristics, along with the highly successful technological changes to return good metallurgical performance with the more complex ore.

MINERALOGY

The mineralogy of the silver-lead-zinc orebodies can be characterised as fine intergrowths of galena and sphalerite with both sulphide and non-sulphide gangue. The main silver material is freibergite which is intimately associated with the galena. The non-sulphide gangue is quartz, dolomite and carbonaceous
shale. Minor amounts of chalcopyrite are present. The sulphide gangue is both pyrite and pyrrhotite, with pyrite predominant. Pyrite is present as two distinct types; firstly, as normal relatively coarse grained euhedral pyrite, and secondly, as fine grained (5 to 30μm) spheroidal pyrite. This second type is refractory and contains elemental carbon, sometimes forming atoll rims on galena. The carbonaceous pyrite is hydrophobic under almost any conditions and is the dominant sulphide diluent in both lead and zinc concentrates (Davey et al, 1970, Munro, 1993).

Feed to the Lead/Zinc Concentrator consists of both Isa and Hilton ores. Isa ores can be classified into two broad categories:

- The upper Isa orebodies, No’s 1, 2 and 5, referred to as “Black Star” orebodies. They are more massive and are mined by open stoping at lower mining cost, and therefore have a lower cut-off grade. Generally this ore is finer grained and has considerably more fine-grained carbonaceous pyrite, whilst core replacement of pyrite by galena (atolling) is more common and at a more advanced stage (Davey et al, 1970).

- The lower Isa orebodies, No’s 7 to 14 and Rio Grande are referred to as “Racecourse” orebodies. These are more narrow and mined by bench stoping with a higher cut-off grade. Generally these orebodies are higher in grade for lead and silver, coarser grained and lower in pyrite. The pyrite is more the euhedral type than the fine-grained carbonaceous type. Though this ore has a higher mining cost per tonne, it is the more profitable ore since it has the highest grade and the best metallurgy. The gradual displacement of this ore by Black Star and Hilton ores is the reason for the continual decline in ore quality experienced in the Concentrator.

The “Racecourse” and “Black Star” categories are used to describe the mineral types in the ore and its metallurgical performance, as well as its geological location.

Hilton ore has been treated through the Isa Lead/Zinc Concentrator since 1987 and is divided into two similar categories. The upper orebodies, No’s 1, 2 and 3 (“Black Star” type ore), are more massive in size and thus allow for open stoping, contain more fine grained, naturally floating carbonaceous pyrite than the lower orebodies and contain more pyrrhotite than Isa orebodies. The lower orebodies, No’s 4 to 7 (“Racecourse” type ore), are narrower and deeper orebodies, mined by bench stoping, with more euhedral pyrite and non-sulphide gangue dilution than the upper orebodies. The silver minerals in the Hilton orebodies are less associated with galena than in the Isa orebodies, hence silver recovery to lead concentrate is lower than from the Isa orebodies. Hilton orebodies also contain a wider range of silver minerals.

The two microphotographs show the difference in complexity between coarse grained, high grade ore (Figure 2) and the fine grained ore with refractory fine pyrite dilution (Figure 3) (Riley and McKay, 1976). Both of these samples were taken from the No 5 orebody at Mt Isa. Over the years, as the tonnage has increased and the head grade declined, more of the ore feeding the Concentrator has been of the Figure 3 type and less has been of Figure 2 type.

![Figure 2.](image1.jpg) ![Figure 3.](image2.jpg)

Microphotographs of Mount Isa No 5 orebody showing the different grain sizes and complexities that occur in the orebodies at Mount Isa (Riley and McKay, 1976).
The mineralogy at Mount Isa presents two distinct problems that affect metallurgical performance – achieving adequate mineral liberation during grinding, followed by separation in flotation. Clearly the ore in Figure 3 needs much finer grinding to achieve equivalent mineral liberation. While the main separation problem is the increase in refractory pyrite, finer grinding to solve the liberation problem increases the difficulty of separation.

Ore Type Performance

The impact of the more difficult separation because of ore type on flotation performance is demonstrated by Figures 4 and 5: at the same grind size and ideal laboratory conditions, lead performance can vary from 60 per cent Pb concentrate grade at 90 per cent recovery (characteristic of the best “Racecourse” ore) to 15 per cent Pb concentrate grade at 50 per cent recovery (characteristic of the worst “Black Star” ore) (Figure 4). Similarly, at the same (fine) grind size, zinc recovery at target concentrate grade can vary by 20 per cent (Figure 5).
The laboratory flotation tests shown in Figures 4 and 5 were conducted to evaluate ores using a laboratory flowsheet similar to current plant operation. Crushed ore was initially ground to P80 = 37 μm, followed by a lead rougher, rougher concentrate regrinding to P80 = 15 μm and final lead concentrate production by three stages of dilution cleaning. Lead rougher tailing and first cleaner tailing were combined to feed the zinc rougher, with zinc rougher concentrate regrinding to P80 = 15 μm and final zinc concentrate production by three stages of dilution cleaning. The flotation tests were conducted with a very fine regrind size (15 μm) to maximise liberation, since less regrinding gives less liberation, and hence both lower zinc recoveries and more zinc contamination of the lead concentrate.

Lead circuit laboratory flotation performance (Figure 4) varies widely for different ore types depending on the refractory nature of the ore and the type and content of the iron sulphides. Different ore types yield significantly different performance. Since different ore types are being mined at any one time from many sources, so the mixture of ore types feeding the concentrator is continuously changing. This causes the performance of the plant to be continuously changing in the absence of intervention by the control room operator.

This leads to two effects in the short term operation of the concentrator:
- there is a change in performance and the operator is unable to tell if it is an ore change or another input change (eg: reagents, mechanical failure, uncontrolled input); and
- the operator makes a controlled change and the performance changes in an expected or unexpected manner. Was the effect from the operator’s change or an ore change?

These issues also need to be addressed when trying to improve plant performance.

![Pb Grade Recovery Curves](image)

*Figure 4 - Lead circuit laboratory flotation performance of different ore types.*
Zn Grade Recovery Curves wrt Plant Feed

Figure 5 - Zinc circuit laboratory flotation performance of different ore types.

Zinc metallurgical performance (Figure 5) varies over a smaller band, due to low zinc losses in the lead circuit and the ability to be more selective against pyrite.

The methods used in the laboratory have shown good agreement with plant performance and provide confidence in using laboratory testwork to predict plant performance.

A SCIENTIFIC APPROACH TO THE PROBLEM

The rapidly deteriorating metallurgical performance in the 1980’s was attributed to continual changes in ore mineralogy. Until the nature of these changes were fully understood, the response consisted of an endless circle of circuit changes, reagent changes, operator changes, metallurgist changes and so on. Fortunately, this was a brief, (though tense) period. It was clear that the solutions could only come from a rigorous scientific approach based on the mineralogy.

Fortunately, good scientific tools were in place to understand the nature of the changes. The two fundamental tools were size-by-size mineralogical analysis and liberation analysis. These two tools were applied to routine plant inventory samples, detailed plant surveys and laboratory and pilot plant testwork. The data combined to provide a unique mineralogical profile of plant performance, which captured both the decade-long decline in performance, and the results of the step changes in improvement.

Routine Analysis of Plant Inventory Samples

Inventory samples of plant products are taken and assayed every shift for metallurgical accounting purposes. Great care has been taken with the design and operation of inventory samplers to ensure there is no size or assay basis. Shift samples are combined into weighted daily composites, which are further combined into weighted monthly composites. In addition to chemical assays, the monthly composite samples are subjected to the following analyses:

- Screen sizing to 37 um, followed by fine sizing (infrasizing until 1992/93, then cyclosizing after 1992/93. The cyclonizer part of the sizing is extended to C7 by a precyclone, followed by collection of the normal C1 to C5 cyclosizer fractions and then to C6 by a centrifuging of the minus C5 fraction. This procedure allows extension of the size analysis to finer sizes, as well as collection of the finest sample. The C6 fraction is typically 4 to 7 um for sphalerite (Johnson, 1992).
- Chemical assay of all size fractions, providing a fully sized mass balance for the plant each month.
Liberation analysis of size fractions. Until 1992/93 this was done by manual point counting and afterwards by QEM*SEM (Quantitative Evaluation of Materials by Scanning Electron Microscope). This provided a full size-by-size monthly mineralogical mass balance of plant operations.

Plant Surveys

In addition to monthly balances, more detailed information was obtained from occasional full or part plant surveys. The surveys are carefully designed to provide a complete snapshot of the operation, with a full mass balance including cyclone splits and down-the-bank flotation performance. All samples are assayed, and selected samples sized and analysed mineralogically.

Laboratory and pilot plant testwork

Laboratory and pilot plant testwork was used to test and identify solutions to problems, the size and quantity of potential performance improvements and the flowsheets required to achieve performance gains. As before, all samples were assayed, and selected samples sized and analysed mineralogically.

THE LIBERATION PROBLEM

The first step change in sphalerite liberation occurred in July 1980 when target zinc concentrate grade was dropped from 52 per cent Zn to 50.5 per cent Zn to maintain recovery above 70 per cent (Johnson et al, 1982). This change caused the adoption of the rigorous mineralogical approach to quantify future ore changes. Figure 6 shows a graphic summary of the changes to sphalerite liberation after 1980.

In Figure 6, ‘sphalerite liberation’ represents the percentage of sphalerite in plant feed which has been liberated before exiting the plant in either concentrate or tailings. This is achieved by the tonnage-weighed mathematical combination of all plant products to form a recalculated plant feed, which represents the total effect of all grinding and regrinding in the plant. A sphalerite grain is considered liberated if it is more than 90 per cent sphalerite in two-dimensional analysis (Gottlieb et al, 1994).

From 1984 to 1991, sphalerite liberation dropped from almost 70 per cent to just over 50 per cent. This was attributed to finer grained ore, although the recalculated feed sizing coarsened from P80 = 50 um to P80 = 80 um because of increases in throughput with no extra grinding power. A decrease in sphalerite liberation causes a drop in zinc recovery, since the maintenance of zinc concentrate grade at 50.5 per cent Zn allows no additional lower grade composites in the concentrate.
There were two possible responses to the liberation problem: either grind finer to increase liberation, or place low-grade middlings into a new, lower grade concentrate. Because of the high capital cost of additional grinding, production of a Low Grade Middlings (LGM), or bulk concentrate became necessary in 1985/86, to maintain total zinc recovery. This concentrate typically assayed 13 per cent Pb and 34 per cent Zn, compared with zinc concentrate which had to contain 50 per cent Zn and less than 3 per cent Pb. Figure 6 shows the effect of the LGM Concentrate on total zinc recovery. The difference between the zinc recovery to zinc concentrate and overall combined zinc recovery is the zinc recovery to the LGM concentrate.

As liberation continued to drop, recovery to zinc concentrate fell with concomitant increases in LGM concentrate production. By 1988, total zinc recovery had to decline further as the target 34 per cent zinc in LGM concentrate was unattainable and the LGM concentrate market had become saturated.

In hindsight, production of LGM concentrate distracted management from the true severity of the problem, since zinc recovery was still quoted as over 70 per cent until 1989. This was really a misrepresentation, since only 55 per cent was recovered to zinc concentrate, with 15 per cent to LGM concentrate. It should also be noted that at the beginning of LGM production, revenue was high because of good contract terms for LGM concentrate. As production rose, contract terms declined until zinc in LGM concentrate was worth less than half that of zinc in zinc concentrate.

Size by size analysis

During the 1980’s the sphalerite liberation declined in three major steps (as shown in Figure 6):

- Mid 1985 from 70 per cent to 60 per cent,
- Mid 1987 from 60 per cent to 55 per cent, and
- Early 1991 from 55 per cent to 50 per cent.

Figures 7 and 8 show the recalculated plant feed sphalerite liberation (by size fraction) and the zinc recovery to zinc concentrate (by size fraction) for selected months over the period 1984/1985 to 1991/1992. Size fractions displayed in Figures 7 and 8 are infrasizer fractions (F7, F6 and F5) and sieve size fraction +400# (+37um). The unsized sample is shown as ALL.

It can be seen from the sphalerite liberation by size (Figure 7) that sphalerite liberation in all size fractions decreased almost uniformly. Consequently, zinc recovery by size also decreased uniformly
(Figure 8). Figures 7 and 8 combine to show that all size fractions were becoming more complex and difficult to liberate, not just the coarse size fractions.

**Sphalerite Liberation in the Recalculated Plant Feed by Size Fraction**

Declining over time in three major steps.

**Zinc Recovery to Zinc Concentrate by Size Fraction**

Declining over time.

**THE SEPARATION PROBLEM.**

While declining liberation posed the most serious problem for zinc metallurgy, decreasing separation efficiency posed the greatest problem for lead metallurgy, and was a secondary issue for zinc metallurgy. The separation problem was caused by increasing amounts of fine grained, carbonaceous
pyrite. As shown by Figure 9, lead head grade declined during the 1980’s with a concomitant increase in iron sulphides. This was exacerbated by an increasing proportion of the pyrite occurring as the naturally floating carbonaceous type, rather than the “well behaved” euhedral pyrite.

Carbonaceous pyrite is hydrophobic under almost any flotation conditions and consumes large quantities of reagent, making flotation difficult to control. Figure 10 shows the changes to lead metallurgy from 1973 to 1990, with lead concentrate grade and recovery both decreasing and the high viability of lead concentrate grade. Although the natural floatability of the carbonaceous pyrite results in a greater impact in the lead circuit, Figure 11 shows that there was also some impact on the zinc circuit. For the same concentrate grade 1980-1992, iron sulphide content rose from 4 per cent to 6.5 per cent. This also contributed to falling zinc recovery, since the higher iron sulphides content in zinc concentrate left less room for composite particles containing sphalerite.
Figure 10.

Zinc Grade and Iron Sulphide Grade in Zinc Concentrate

Figure 11.

The falling lead concentrate grade also posed a serious problem for the lead smelter. Smelter throughput was limited by sinter plant sulphur removal capacity. Increased pyrite in lead concentrate caused lead grade to decrease and sulphur to increase (Figure 12), reducing smelter throughput.

Figure 12.

Numerous circuit and reagent alternatives were trialed to minimise the impact of the carbonaceous pyrite. After several decades of work, the most effective response remains a dextrin depressant at natural pH (7.5 to 8). Other changes instituted in the 1980’s to improve lead circuit separation included:
Installation of preflotation before lead roughing which operated between December 1986 and February 1987. Preflotation concentrate was cleaned before discarding at an assay slightly higher than head grade for lead, zinc and silver. Preflotation operation was stopped partially because of recovery loss and partially due to the resultant unstable operation of subsequent lead roughing and cleaning. Without preflotation, the majority of lead circuit xanthate addition is added to mill feed, but with preflotation the addition is made to rougher feed. Addition of the xanthate to rougher feed appeared to cause instability in the lead circuit, particularly in summer with high pulp temperatures (+45 deg C).

Use of a “high-low split” lead cleaning circuit during 1991 and 1992 (Figure 13). This circuit collected a high grade concentrate from the first part of the lead cleaners, whilst a low-grade concentrate from the second part of the cleaners was sent to further cleaning in a Jameson cell, the latter producing a high-grade concentrate and a low-grade tail. High-grade concentrates were sent to the lead smelter whilst the low-grade Jameson cell tailing was stored until smelter capacity was available.

Installation of a Heavy Medium Plant (HMP) slimes roughing and cleaning circuit in 1988. These “slimes” are generated in the mining process, represent about 15 per cent of the lead in the flotation feed, and have different reagent requirements to normal feed (Grano et al, 1988). A flowsheet including roughing at pH 9 (with lime), zinc sulphate, and cleaning in a Jameson cell was developed, producing a lead concentrate from HMP slimes that averaged over 60 per cent Pb, compared with an estimated 45 per cent Pb when included in conventional feed. The separate HMP slimes roughing and cleaning contributed a two per cent increase in overall lead concentrate grade.

The LGM concentrate circuit from 1986 assisted by directing difficult lead middling streams into a low grade concentrate. This raised lead concentrate grade, and helped match concentrator and smelter capacity by directing some metal away from the lead concentrate.

In the case of zinc separation, little improvement was possible because of the poor liberation. Use of traditional iron sulphide depressants (eg. lime and dextrin) became severely restricted because of uncontrollable circulating loads of composites and fine free sphalerite (Johnson et al, 1982). In turn, the circulating loads consumed limited cleaning and retreatment flotation capacity. Attempts to improve zinc circuit selectivity included:

- Operating the LGM circuit to provide an exit for the most difficult composite streams, eg zinc scavenger concentrate and zinc cleaner tailing;
- installation of new column cleaning capacity in the zinc retreatment and LGM circuits (Espinosa-Gomez et al, 1989 and Espinosa-Gomez and Johnson, 1991);
- substitution of lime with dextrin in zinc cleaning, with later restriction of dextrin additions to minimise circulating loads;
- trials of a number of supplementary collectors which promised, but did not provide, increased selectivity; and
- use of a hot reverse cleaning circuit developed in the pilot plant. However, by the time this approach was developed it was clear that circuit simplification was the priority rather than the addition of new equipment.

The changes made during the 1980’s were individually effective in achieving performance improvements, however the rate of improvement did not match the ore type rate of decline. Further, the succession of small changes had dramatically increased the complexity of the combined lead, zinc and LGM circuits as the changes had treated symptoms rather than the underlying mineralogical cause. By 1992, the Concentrator had 14 exit streams: 8 concentrates and 6 final tailings (Figure 13) and suffered operational difficulties with respect to stable operation.
1990’s metallurgical performance improved dramatically as several projects addressed the underlying mineralogical issues:

The projects were:

- The “Fine Grinding Project,” which doubled grinding and flotation capacity and instituted a “cold” lead reverse cleaning circuit (1992). This project addressed both key issues, i.e., liberation in the zinc circuit and separation (of carbonaceous pyrite) in the lead circuit.
- Improvement in liberation allowed circuit simplification, the increased use of conventional tools (e.g., high pH zinc circuit cleaning) and relocation of regrinding duties from the LGM circuit to the zinc retreat circuit.
- New ultrafine regrinding technology at Mt Isa was introduced for both zinc and lead regrinding (1994).
- Generally, the application of process control became more effective with the improvements because of more achievable targets.

The fine grinding project

While it was well-known prior to 1992 that finer grinding was required, the high capital cost prevented the acquisition of the requisite additional grinding equipment. Conversion of the Copper Concentrator to SAG milling in 1991 solved this problem by releasing two 5m by 6.1m, 2.6 MW ball mills. A project was approved to install these mills for secondary grinding in the Lead/Zinc Concentrator, along with a 520 kW Tower Mill for additional regrinding in the LGM circuit. The finer feed sizing and more dilute pulps necessitated extra flotation capacity, which resulted in the installation of two banks of nine Dorr Oliver DO600 cells for lead secondary roughing and scavenging, and three banks of 12 Dorr Oliver DO600 cells for zinc roughing and scavenging. Existing flotation cells were used to provide additional roughing, retreatment, or cleaning capacity. In summary, the new equipment provided the following changes:
• An increase in primary and secondary grinding power from 6.3 MW to 11.5 MW. Recalculated feed sizing dropped from P80 = 80um P80 = 37um microns;
• increase in regrinding power from 0.75 MW to 1.27 MW. Regrinding size in the LGM circuit dropped to P80 = 12um; and
• a doubling of flotation capacity.

The Tower mill was commissioned in December 1991, and the two secondary mills in October and November 1992. Figure 14 shows the effect on sphalerite liberation and zinc recovery. The key features are:
• Sphalerite liberation increased by 25 per cent by 1993;
• an increase in total zinc recovery of eight per cent. More importantly, the amount of zinc reporting to the low value LGM concentrate was reduced, yielding an increase in zinc recovery to zinc concentrate of over 15 per cent. These recovery gains not only achieved feasibility estimates almost immediately after commissioning, but also quickly exceeded the same estimates.

**SPHALERITE LIBERATION IN RECALCULATED FEED VS ZINC RECOVERY**

Smoothed Data: 3 period rolling average

![Graph showing sphalerite liberation and zinc recovery](image)

NOTE: The change to QEM*SEM in 1992/93 increased the sphalerite liberation by +4%.
An interesting feature of Figure 14 is that sphalerite liberation exceeded recovery quite significantly. Traditionally, ‘Johnson’s rule of thumb’ stated that combined zinc recovery equaled sphalerite liberation plus 10 per cent, reflecting the amount of diluents that could be tolerated in a zinc concentrate of 50.5 per cent Zn and an LGM concentrate of 34 per cent Zn. All the liberation gains of the Fine Grinding Project were not converted to recovery, since more minerals were now in the difficult to separate size ranges (eg 20 per cent of sphalerite is now less than 4 um). This does not imply that liberation is no longer an issue. Indeed, the pursuit of increasing levels of liberation since the Fine Grinding Project was installed has been a major theme of development. However, it creates an environment where pulp chemistry and flotation separation are now more productive areas of research.

Figure 15. - Flowsheet after the installation of increased grinding and flotation capacity (nine products).

The Fine Grinding Project and Separation Improvements

The ‘cold’ lead reverse cleaning circuit was installed at the same time as the Fine Grinding Project to remove carbonaceous pyrite from lead concentrate. Conventional lead cleaner concentrate is raised to pH 12 with lime to depress galena but not carbonaceous pyrite. A pyrite concentrate is floated, cleaned, and discarded (Figure 15). The pyrite concentrate assays around 30 per cent Fe, 32 per cent S, and 19 per cent Pb. Typically, the reverse cleaning trades off one per cent Pb and 1.2 per cent Ag recovery for each one per cent increase in lead concentrate grade (and accompanying 0.5 per cent lower Fe and 0.3 per cent lower S). The maximum upgrading capacity of the circuit (because of physical constraints) is 4 per cent Pb. Operation of this circuit is intermittent, depending on current ore type, metal prices, and smelter performance. The circuit’s major advantage is the provision of independent control of lead grade/recovery decisions. In the conventional cleaners there is very little ability to trade lead recovery for grade, since the lead cleaner tailing assay has to be kept low to keep galena out of the zinc circuit. A process control system varies the reverse cleaning circuit air addition to control a setpoint lead concentrate grade. This gives the lead smelter a much steadier grade concentrate, while minimising the recovery loss of galena. The circuit is shut down when better ores are encountered. The improved lead concentrate quality resulting from this circuit (by decreasing the iron sulphides) is shown in Figure 16.
In the two years after the implementation of the Fine Grinding Project, further performance gains were made as the circuit was adjusted and simplified. Effectively, operating personnel had to “unlearn” many of the circuit rules essential when poor liberation and insufficient flotation capacity were the root of many problems. The most significant of these were:

- Reintroduction of high pH zinc cleaning using lime. This had been abandoned prior to the Fine Grinding Project because of unmanageable circulating loads of composites.
- Relocation of some LGM circuit regrinding capacity into the zinc circuit (Figure 17). Together with the reintroduction of lime, this helped increase recovery to zinc concentrate by a further 5 per cent.

**MOUNT ISA MINES LIMITED - LEAD/ZINC CONCENTRATOR FLOTATION FLOWSHEET**

![Flowsheet](image)

*Figure 17. - Flowsheet after the relocation of regrinding capacity. (eight products).*
• Use of some fresh water, instead of process (recycle) water, as dilution water in both zinc and lead cleaning. This reduced the impact of salt deposition (especially gypsum) on fine minerals surfaces. It also reduced the lime requirement in zinc cleaning as the slurry was previously supersaturated in calcium, and helped reduce frothing problems in both lead and zinc cleaning.

• Reintroduction of basic process control loops. Enormous efforts in advanced process control had previously yielded little gain, as the process was inherently unstable. Supervisory loops have been gradually introduced as the circuit has been simplified and stabilised. Tonnage/size/load grinding loops are used by operators over 85 per cent of the time on all grinding lines, and 18 flotation loops are used 70 per cent of the time. Tonnage based feed forward reagent ratio controllers are used for cyanide, copper sulphate and xanthate additions in the lead and zinc circuits. Adaptive controllers are used in the cleaners, adjusting both air and xanthate additions.

New Regrinding Technology

• Mount Isa Mines Limited developed revolutionary new ultrafine grinding technology for the McArthur River deposit, with prototypes developed in the Lead/Zinc Concentrator. The circuit has had two 1.1 MW mills regrinding lead concentrate, since 1995 (Figure 18) (Enderle et al, 1997). These mills have further increased liberation and recovery and simplified the circuit. The lead regrinding mills increased zinc recovery by 5 per cent by liberating sphalerite from composites that previously reported to lead concentrate. These mills regrind lead rougher concentrate to P80 = 15 um.

• The regrinding mills also eliminated the bleed stream of difficult lead middling particles to the LGM circuit, leaving the zinc treatment as the only remaining feed to the LGM circuit.

MOUNT ISA MINES LIMITED - LEAD/ZINC CONCENTRATOR FLOTATION FLOWSHEET

![Flowsheet after the installation of regrinding of lead rougher concentrate.](image)

Size by size analysis

The increase in sphalerite liberation in the recalculated plant feed is shown on a sized basis in Figure 19 for selected months from 1991-1995. Figure 19 shows that liberation increased across all size fractions for each project which increased grinding power or grinding efficiency. The liberation increased in four main stages (Figure 14):
• Installation of Tower Mill in LGM circuit,
• installation of increased primary and secondary grinding power during the Fine Grinding Project,
• relocation of grinding power from LGM circuit to zinc circuit, and
• installation of lead rougher concentrate regrinding.

The zinc recovery by size data increased in a similar manner to the liberation by size data (Figure 20). The figure shows that all size fractions were more liberated with finer grinding and not just the coarse size fractions. The liberation was therefore improved by two methods; firstly by increasing the liberation of each size fraction and secondly, and more importantly, by moving particles from the coarse, less liberated size fraction to the finer, more liberated size fractions.

![Diagram of Sphalerite Liberation in the Recalculated Plant Feed by Size Fraction](image)

Figure 19.

![Diagram of Zinc Recovery to Zinc Concentrate by Size Fraction](image)

Figure 20.
CONCLUSION

Adoption of a rigorous, size-by-size mineralogical approach to plant operations was crucial to identify and solve the dramatic decline in ore quality and metallurgical performance.

The result was a 20 per cent increase in zinc recovery to zinc concentrate, 5 per cent increase in lead recovery to lead concentrate, improved quality for both lead and zinc concentrates, and 70 per cent reduction in the production of the low value LGM concentrate.

Important also is the simplification of the circuit. From 14 exit streams in 1992, the circuit had eight exit streams by 1995. This produced a dramatic improvement in circuit stability and increased ease of circuit operation. Three main indicators of circuit stability are:

- the willingness of operators to use simple process control loops to assist their decisions;
- the speed of achieving stability after plant start ups, ie. metallurgical results on start-up shifts are now indistinguishable from normal operating shifts; and
- plant spillage and hygiene. High side rubber boots are no longer issued, nor needed!

This case study is an excellent example of the benefits of applying a scientific approach to routine operations over a long period of time.

POSTSCRIPT - RECENT CHANGES

The metallurgical improvements described in this paper were driven by technology changes targeted at the fundamental nature of our fine grained, complex ore. The changes were highly successful and economically essential to business as ore quality declined. However, the improvements came at a price - high capital and operating cost. In 1996, the next improvement came from a comprehensive examination of the mine/mill/smelter business. This led to elimination of the LGM (bulk) concentrate and increased lead and zinc concentrate grades and recoveries, as well as providing considerable circuit simplification. These changes were achieved without capital and without extra operating cost. They will be the subject of future publications.

ACKNOWLEDGMENT

The consent of Mount Isa Mines Limited to the publication of the paper is gratefully acknowledged. The authors wish to thank the professional engineers, foreman, operators, maintainers and postgraduate and vacation students who were concerned with the research, development, commissioning and operation of the plant and with data collection and analysis for their contributions during the 1980’s and 1990’s.
REFERENCES


At the Coal21 Annual Conference in 2007, the author proposed an alternative pathway to low CO₂ emissions electricity with coal as the key fuel. The proposed pathway avoids or minimises many of the issues facing current clean coal technologies focused on CO₂ capture and geological storage (CCS). These include supercritical pulverised coal with post-combustion capture (PCC), gasification with capture, and oxy-pulverised coal with flue gas liquefaction. The current focus on CCS is because improvements in efficiency for conventional coal technology will not give anything like the step reduction in greenhouse gas emissions required. This means that only limited reductions in CO₂ will occur (via efficiency improvements) until around 2025 when CCS is expected to be rolled out on a meaningful scale.

Presently, if a new coal plant were built in Queensland or NSW, there would be a strong preference for it to be a 700MW, dry-cooled, advanced supercritical unit, with provision for future staged CO₂ capture using PCC. With capture at the current stage of development, this technology would give a delivered cost of electricity (including capture, transmission and distribution penalties), of around $125/MWh, and with a fuel cycle thermal efficiency for delivered electricity of around 30%.

This article considers an alternative technology pathway to that of current clean coal technologies.

**EFFICIENCY CHALLENGE FOR CLEAN COAL**

The following graph (Figure 1) shows that the delivered thermal efficiency of coal-based generation (i.e. with transmission) has improved at a relatively constant rate over the last 115 years. Although there have been a number of disruptive technologies (the steam turbine, suspension or pulverised fuel (pf) firing, large unified designs, and supercritical steam conditions), when graphed over a longer time interval up to the present, there has been a steady increase in efficiency.

Based on best current estimates from the technology developers, efficiencies for dry-cooled plants under Australian conditions will increase by around 4% points by 2030. However, if CO₂ capture were adopted, there will be an efficiency penalty of around 8% points – which will effectively wipe out 50 years of efficiency improvements.

**ENERGY PENALTY INCREASES STORAGE**

The important point is not the efficiency number itself, but the proportional increase in the amount of CO₂ that will need to be stored as efficiency is reduced: if the overall efficiency of the system is reduced, then more CO₂ is produced and more must be captured, which increases the energy penalty for capture and further lowers the overall efficiency. This results in a disproportionate increase in the amount of CO₂ to be captured and stored to meet a given emissions intensity.

Figure 2 shows the amount of CO₂ to be stored to meet an emissions intensity of 150 kg/MWh versus delivered thermal efficiency of the coal fired generation plant. With best current black coal plants, for each MWh delivered, around 1000 kg of CO₂ will need to be stored. This increases to around 1550–1600 kg/MWh for current brown coal plants.
EMISSION REDUCTION ACTIVITIES

Continued from page 5

Making a current versus future technology comparison, if the delivered efficiency were increased to 45%, then the amount of CO₂ to be stored would decrease to 600 kg/MWh (a 40% reduction for black coal, and a 62% reduction for brown coal). If a delivered efficiency of 65% could be achieved, this would reduce the amount of storage required for black coal by 62%, and by around 76% for brown coal.

These higher delivered efficiencies would markedly reduce the aquifer volume required for CCS. As an example, to store the CO₂ from NSW’s current coal plants would require injecting liquid CO₂ into around 8000 m³ of reservoir pore volume every hour. Using nominal Department of Primary Industries values for aquifer porosity and the effectiveness factor, this means that around 1 000 000 m³ of reservoir aquifer will be filled every hour. The higher efficiency value would reduce this to around 380 000 m³ per hour for black coal.

As the cost of CO₂ mitigation per unit of power will also decrease in the same proportion, anything which can increase the system efficiency, and reduce the amount of CO₂ required to be stored, will be valued much more highly in the future.

THE ALTERNATIVE PATHWAY

The alternative pathway being proposed by CSIRO has different priorities to those of the current clean coal approach:

• Achieving the highest possible efficiency for coal-based generation at smaller scale,

• Using the advantages of higher efficiency, smaller coal plants to underpin a high penetration of renewables, thereby giving more cost-effective renewables and a proportional reduction in CO₂ intensity of the system, and

• Using niche (rather than major) CO₂ capture and storage as the last step in meeting emissions targets.

So while current clean coal technologies are based on large-scale centralised plants with major CO₂ capture and storage – the alternative pathway is about smaller, more efficient plants that underpin renewables, and using niche capture as a last step.

MAXIMISING THERMAL EFFICIENCY

Figure 3 compares the thermal efficiency (from delivered fuel to delivered electricity – without capture) for a range of technologies. Size-for-size, the most efficient means of converting fuel energy to electricity is by using large diesel engines or direct carbon fuel cells (not yet commercially available, though the high efficiency has been demonstrated in the laboratory).

Coal could achieve a much higher efficiency if these technologies could somehow be used, enabling a delivered efficiency of around 50% from a direct injection coal engine (DICE) in the short-medium term, and at least 65% in the longer-term future from the direct carbon fuel cell (DCFC). Both could be achieved at a smaller scale than for conventional coal technologies – an attribute that could give a number of other important advantages associated with more decentralised electricity generation, including a reduction in transmission losses (electricity transmission losses are almost always significantly higher than those from the transport of fuel).

Although DICE and the DCFC require ultra low ash coals (say below 2% ash), technologies already exist to produce suitable fuels from coals (including from Victorian brown coals).

DIRECT INJECTION COAL ENGINE

Although never commercialised, the coal engine is not new, being the subject of a number of development programs over the last century (with key programs every 20 years approximately). The early work was led by Diesel and then co-worker Padowski, with engines running for many years on everything from lignite dust to coke. However, the most comprehensive test program was undertaken by the US Department of Energy (DOE) over the period 1978 to 1992, mostly for transportation applications. This included adapting and testing engines ranging from a 2MW single cylinder test engine (90 rpm), a 2MW locomotive (1000 rpm), and an 800 kW haul truck engine (1900 rpm). The most successful demonstrations used ultra low ash coal water fuels which were injected using modified conventional direct injection (i.e. solid injection) systems. Although the technical issues were overcome for the direct injection coal engine, with combustion efficiencies of over 99% being achieved at up to 1900 rpm, and with thermal efficiency equal to diesel fuel, the program was eventually terminated due to persistently low oil prices and before a commercial engine was developed.

WHAT’S CHANGED?

In revisiting this technology, it is important to consider a number of different drivers: the impending cost of CO₂ abatement, cooling water availability, the need to support a step increase in renewables, energy security and changes to the structure of the electricity supply industry.
DICE would have a number of advantages, including being implemented with smaller capital steps, and would be suitable for baseload, peaking and providing grid security or ancillary services. DICE would also be capable of cofiring of biomass fuels (char, crude bio-oils and algae soups), which could enable biomass to be utilised at double the efficiency of current biomass plants. Also, as waste heat is at sufficient temperature to provide the low temperature heat energy needed for CO₂ capture, this could be added with a smaller cost and energy penalty than for current clean coal technologies (instead of converting this energy to electricity with very small and inefficient turbo machines).

In addition, there have been a number of key technology changes, including developments in coal cleaning, micronising, engine technology/materials/sizes and costs (about the same as for pf), which are all likely to further increase the viability of the coal engine over that in previous development programs, and reduce the time required for adoption of the technology. Currently, under the European HERCULES Project, new engine technologies are being developed which will increase the efficiency of large engines to over 60% (LHV, mech).

**DIRECT CARBON FUEL CELL**

The ultimate in thermal efficiency from coal could be achieved by using ultra low ash coals in direct carbon fuel cells. The DCFC works similarly to the hydrogen fuel cell, but is superior in terms of thermal efficiency, with over 75% being achieved in US laboratories for a range of carbon types, including a sample of ultra low ash coal produced by UCC Energy. There are a number of possible configurations being developed. An example schematic arrangement is shown in Figure 4, which comprises a solid oxide fuel cell bank, and a flue gas split (capture is not required) for CO₂ sequestration.

The DCFC is being commercialised for small applications (e.g. the US Army are funding the development of a unit to convert ration pack wrappers to electricity for field use), and current projections from US developers are that it will be available for large-scale commercial electricity production from around 2030. Development is also expected to piggy-back on technologies being developed for the hydrogen fuel cell.

From a system perspective, the DCFC would be highly advantageous, as the technology would be suitable for baseload and peaking duties, is highly suitable for biomass char; and as it produces essentially pure CO₂, it would have a smaller efficiency and cost penalty for CO₂ capture and storage. In the alternative pathway, it is envisaged that DICE power plants would ultimately be converted to DCFC, say post 2030.

In Australia, the DCFC is being researched for coal applications by the CSIRO, and the Universities of Newcastle and Queensland.

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![Diagram of main coal cleaning processes](image)

**ULTRA LOW ASH COAL**

The alternative pathway depends entirely on cost-effective production of ultra low ash coals. Although the cleaner the better, detailed cost specifications for both large diesel engines and the DCFC remain unclear. The DOE work in the 1980s and early 1990s concluded that up to 2% ash is suitable for DICE, and this could be produced from chemical and physical coal cleaning processes.

The DOE studies showed that the most suitable method for fuel delivery was direct injection of micronised coal water fuel or CWF (so drying is not required after cleaning). Depending on engine size and operating speed, the micronised coal should have a top size of 20–30 µm, and a coal concentration of around 45%. This gives a CWF viscosity of around 300–500 mPa.s (at 100/s) which is sufficient to enable effective atomisation.

Although there are a number of methods of utilising CWF in the DCFC (including integrated flash drying and partial gasification combinations), it is expected that a similar quality coal would be required.

There are three main process types for cleaning coal and producing an engine grade CWF from bituminous coals, as shown in Figure 5. These involve either removing mineral matter from the coal, or the coal from the mineral matter. None of these processes are currently commercial, but have been operated at up to pilot scale.

**CHEMICALLY CLEANING LOW ASH FEED COAL**

The first process involves chemically cleaning a relatively low ash feed coal, producing a coal water slurry, and then micronising to produce an engine grade CWF. Examples are the AMAX and UCC processes which use a combination of caustic and acid treatments to dissolve away the ash and InterTech which does a similar thing using hydrofluoric acid.

**FINE COAL PROCESSING**

The second approach uses fine coal processing to remove as much of the ash as possible by conventional flotation, then uses chemical cleaning to remove the remaining ash, followed by slurrying and micronising to produce a CWF for an engine. This was developed by AMAX in the 1980s. However, there are limitations on how much grinding can be tolerated without compromising the materials handling in the chemical cleaning process, the ability of the pre-cleaning step to reduce ash levels significantly (below that of normal coal preparation technologies) will be highly coal dependent.

**ULTRA FINE COAL PROCESSING**

The third approach involves slurrying and micronising the coal first (not last), and then using ultra fine flotation or selective agglomeration to remove the ash. Although several previous studies have shown that this process is technically feasible, the results obtained in the past have been highly variable, especially with respect to coal recovery rates. However, recent test work with a range of Australian coals (including tailings) has shown that if coal is ground finely enough, a consistent and low ash product is possible with very high coal recovery. As the micronised raw coal slurry has the appearance of crude oil prior to refining, this process has been termed micronised coal refining, and the product micronised refined coal (MRC).

As there is a trade-off between product ash and (nominal) processing costs (see Figure 6), the most suitable cleaning process will depend on the target application. In particular, there is a significant increase in production cost with ash contents below around 1.5–2%, because chemical cleaning is then required for most coals. For this reason, chemically cleaned coals at 0.2–1% ash are likely to be cost competitive with natural gas.
Coal at around 2% ash would be cost competitive with thermal coals used in conventional coal power plants.

ULTRA LOW ASH COAL PROJECTS
Over the last few years there has been a steady increase in efforts to produce ultra low ash coals. By far the most advanced is the process being developed by UCC Energy, which is aiming to demonstrate the production of ultra clean coal (UCC) with ash contents down to 0.2%. This very low level is achieved by chemical cleaning. Although originally developed for gas turbines, UCC Energy is now aiming to optimise the process to produce a higher ash specification fuel for a wide range of diesel engines in applications that could replace natural gas/LNG turbines. It is projected that the revised process will give a step reduction in processing costs and CO₂ emissions from coal processing. A small-scale demonstration is presently being undertaken as part of the Asia Pacific Clean Fossil Fuel program (a component of the Asia-Pacific Partnership on Clean Development and Climate).

MICRONISED Refined COAL
To compete with black coal pf plants, CSIRO, TUNRA Clean Coal, the University of Newcastle and Xstrata Technology are collaborating to investigate the use of new developments in milling and ultra fine coal cleaning to produce low-cost micronised refined coal (MRC) suitable for very large diesel engines. The process involves micronising run of mine coal to a d90 of <20 µm using an Isamill, and then physical cleaning and partial dewatering using advanced ultra fine coal flotation.

Cost estimates show that MRC could be produced at substantially lower costs than for processes using chemicals, which should enable MRC-DICE technology with the largest engines (mostly likely to be able to cope with higher residual mineral matter and other ash forming material in the fuel) to compete favourably with pf plants for CO₂ costs above $30/t.

UNDERPINNING RENEWABLES
The second aspect of the alternative pathway is to use smaller high efficiency coal power plants to provide the grid security necessary to enable a higher penetration of renewables. Coal could have a more important role in underpinning the uptake and development of renewable energy, particularly those which are intermittent – wind and solar of course, but also seasonal biofuels. By doing this, coal when treating higher ash coals. Since the extensive coal engine R&D for the DOE in the 1980s, new large mills have been developed for the minerals industry, routinely milling up to 10 000 tonnes of hard rock per day to sizes below 10 µm (e.g. see photograph of a typical Isamill in Figure 7). Preliminary studies by CSIRO have shown that coal can be micronised by an Isamill suitable for producing engine grade fuel with a projected energy consumption of around 60 kWh/t.

Another interesting development is that TUNRA Clean Coal have dewatered MRC to 12% total moisture using selective agglomeration (at very low levels of organic additions). MRC is produced as a dry prill, which would give the option of transporting and storing MRC in solid form. While most effort is on bituminous coals, Exergen are developing a diesel engine fuel from Victorian brown coal and lignites. As the ash content of these coals is already low, fuel production is focused on producing a suitable CWF by upgrading these coals by hydrothermal treatment.

MICRONISING
Efficient micronising of coal is essential for either coal engine or fuel cell applications. Conventional pf mills are totally unsuitable for producing the ultra fine particle size which is thought to be required for diesel engines (d50 of <10 µm and a top size of 30 µm). Conventional mills are even less suitable when treating higher ash coals. Since the extensive coal engine R&D for the DOE in the 1980s, new large mills have been developed for the minerals industry, routinely milling up to 10 000 tonnes of hard rock per day to sizes below 10 µm (e.g. see photograph of a typical Isamill in Figure 7). Preliminary studies by CSIRO have shown that coal can be micronised by an Isamill suitable for producing engine grade fuel with a projected energy consumption of around 60 kWh/t.

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effectively reduces its own CO₂ intensity, and supports the transition to more sustainable generation.

Done effectively, this avoids the costs and inefficiency of energy storage or of providing the ancillary services required for a secure grid.

There are at least three ways that coal plants can do this – Providing longer-term backup, shorter-term spinning reserve, and efficient utilisation of biofuels.

While the backup and spinning reserve roles are well known, to do this efficiently, without increased CO₂ emissions (especially with a higher penetration of renewables), requires technology which can give higher part-load efficiency and short start times. As large diesel engines have excellent part-load efficiency and also relatively short start times, DICE-based power plants should be better than pf and gas plants for this role.

The ability of DICE to efficiently utilise other biofuels is also likely to greatly assist the development of renewables. These could range from micronised chars, through a wide range of crude bio-oils, to soups of high lipid content algae. Relative to gas turbines, diesel engines are extremely tolerant to alkalis in the fuel, and compared to pf plants have only a marginal penalty for fuel containing up to 60% water. While these attributes are also shared with gasification-based technologies, DICE could do this with a much higher efficiency (8–15 percentage points depending on unit capacity) and remain effective at smaller scale.

**CAPTURE AND STORAGE**

Even with high efficiency and a high penetration of renewables, some CCS may eventually be required. As yet there appears to have been no R&D specifically undertaken for capture from a coal engine or a direct carbon fuel cell. However, as the coal engine exhaust will have properties between that of pf boilers and NG gas turbines, it should not pose any special difficulty for post-combustion capture. As engines reject their heat at a temperature which is usable for CO₂ stripping, this will greatly reduce the energy penalty for capture – especially if an integrated capture process is used to replace the small turbo machinery normally used to recoup extra exhaust energy.

Capture from a DCFC would be even easier, as the flue gas produced would be similar to that from oxy-pf. Capture would not be required, only dehumidification and liquefaction, giving an energy penalty only one-third that for oxy-pf.

It is likely that either technology has the potential to halve the energy penalty for capture, thereby further reducing the amount of CO₂ that has to be stored.

**IMPLICATIONS FOR CO₂ INTENSITY**

The diagram in Figure 8 shows the likely reduction in CO₂ intensity of delivered electricity for the current approach based on clean coal technologies with CCS, and that proposed for the alternative pathway. Although both pathways remain unproven, both could achieve a CO₂ intensity of around 100 kg/MWh into the future.

With clean coal technologies only a small reduction in intensity is likely until the rollout of major CCS – which by current projections is unlikely before 2025–2030. As the alternative pathway can be implemented at smaller unit capacity (60–100 MW), it is proposed that this should result in smaller capital and risk hurdles which could facilitate earlier implementation. This would start with using ultra low ash coals into large diesel engines from 2015 giving a 20% reduction in CO₂ for new plants. These plants could be used to give both direct and indirect support to a high penetration of renewables (say 20%), providing a cumulative reduction in CO₂ intensity of around 40%. DICE plants would be replaced by DCFC/niche CCS as required post 2030 to meet future emission targets, and with a 400% reduction in the amount of CCS required. It is proposed that additional cumulative CO₂ savings could accrue from earlier implementation.

WHERE TO FROM HERE

It is acknowledged that the proposed alternative pathway needs considerable development and demonstration to match the level of technical development of the clean coal technologies. Despite this, the proposed pathway has strong technical merit because of the ability to carry out a near-commercial scale demonstration at a relatively small scale. This greatly reduces development and commercialisation risk hurdles. In addition, while the overall approach is novel, most of the component technologies are based on adapting commercial and mature processes (and many are Australian). An additional feature is the high degree of flexibility and adaptability of all of the technologies.

A number of proposals are under development to establish small-scale demonstration projects in Australia. These mostly involve a two-stage program starting with pilot testing of fuel production at around 1 t/day for use in a 500 kW pilot engine (or several cylinders of a larger engine), together with engineering for a larger demonstration plant. This would be followed by a 10 000 hour demonstration plant with a 10 MW power plant. Even at this small scale, the demonstration plant would exceed the sent out thermal efficiency of Australia’s best pf plants. In addition, this would be sufficient to scale up (through both multiples and scale) to a 100 MW power plant within 5 years.

As the most costly component of the demonstration is expected to be the power plant (around $25M for 10MW), there is considerable scope for cost-sharing this facility between a number of ultra low ash coal projects. Hopefully this could be achieved with strong interest from the large engine manufacturers.

In the meantime, CSIRO, partners and several industry groups continue to undertake research on how best to prepare coal water fuels from different coals/different cleaning technologies, and to better understand coal-engine interactions. Several Chinese groups are also working in this area.