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Evolution of SAG Mill Process Control at the Xstrata Nickel Raglan Operation

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ABSTRACT

The Xstrata Nickel Raglan Concentrator, located at the northern limit of Quebec's Nunavik region has been in operation since 1997. Originally designed to process 800,000 tonnes per year of high grade nickel ore, plant capacity is currently approaching 1,200,000 tonnes per year due to an Optimization Project (FAG to SAG conversion) in 2005 and many recent process improvements.

In 2006, the Raglan Metallurgical Group with assistance from the Xstrata Process Support group decided to make SAG control one of its priorities in order to further increase throughput and improve stability. The original primary mill control logic delivered by the Optimisation Project in 2005 consisted of a simple PID loop controlling the mill charge (bearing pressure) through feed rate changes. Due to the multivariable, non-linear and disturbance dominated nature of the SAG circuit it was quickly found that this traditional PID control strategy was not able to adequately control the mill; resulting in poor stability, frequent operator interventions and less than optimum performance.

This paper describes the successful integration of advanced field systems such as mill feed image analysis (Wipfrag) and crusher gap controller (ASRi), into a multi-variable fuzzy logic SAG mill controller. The process of how a strategy for control was developed and implemented directly in the existing control system (exploiting standard system capabilities) is discussed. It is also shown how this approach has provided mill automation which is simple, robust (despite erratic feed characteristics), and delivers both an increase in throughput and a reduction of variability.

INTRODUCTION

Originally commissioned as a Fully Autogenous Grinding (FAG) circuit in 1997, preparations to convert the Katinniq Concentrator Mill to SAG had in fact commenced as early as 1999, but were halted owing to alternative process improvements which achieved *up to* 1Mtpa as early as 2000. Up until this time, plant *process controls* had remained more or less static; modifications having been limited to those required by new equipment. Primary mill process controls and related measurements were at this stage limited to:

- Mill feed tonnage (automatic *proportional-integral* (PI) control with operator set-point)
- Mill water addition (valve – operator determined % open)
- Mill rotation speed (operator set-point)
- Mill power and bearing pressure indication
- *Flotation Feed* assays : Cu, Ni, Fe (via Courier 30 x-ray system)

It was at this stage that the Process Control Group of the former Falconbridge Technology Centre was invited to identify potential areas for process control improvements.

INITIAL MILL PROCESS CONTROL DEVELOPMENT

The process control system initially installed comprised a network of Modicon Quantum PLCs connected to a number of WonderWare Intouch SCADA HMI (Human-Machine Interface) stations. While the network of operator stations was upgraded in 1998 to address an overloaded communication network, the PLC based controls had remained largely as commissioned.

All the above controls had been programmed in *ladder logic* using the Modicon MODSOFT programming utility. While this method of programming is well suited to automation of motor start/stop and other discrete electrical equipment, it is not well suited to continuous control processes. Although *basic* regulatory controls can be relatively simple programmed, they are inflexible and very quickly become convoluted and unreadable once more complex controls are attempted.

It was largely the inflexibility and lack of user-friendliness of the control system that hampered development of improved process controls up until around 2003. Changes were onerous to perform and application of more advanced regulatory controls was difficult to execute. This was a frustrating situation for mill operations, who recognised the potential that improved process control could deliver.

This situation resulted in acceptance of an earlier recommendation to convert the PLC programming software to Schneider's 'Concept' (Thwaites, 1999; Bartsch, 2001). 'Concept' is an IEC-61131 compliant programming package which would deliver *DCS like* functionality to the control system; on the same PLC hardware platform! This conversion was a turning point for the advancement of effective process control at Raglan. Similar conversions have been reported elsewhere (Atasoy & Price, 2006) and (Karageorgos, Genovese & Baas, 2006), which have led to improved implementation of effective process controls.

FAG TO SAG CONVERSION PROJECT

It was the combination of increased throughput targets associated with anticipated harder ore with finer grind requirements, that finally pushed the decision to convert the primary mill to semi autogenous grinding (SAG) in October 2005. The circuit delivered by the conversion project is shown in Figure 1, and included the SAG mill itself, a new oversize recycle Crusher, a double deck discharge screen as well and improved and additional downstream equipment.

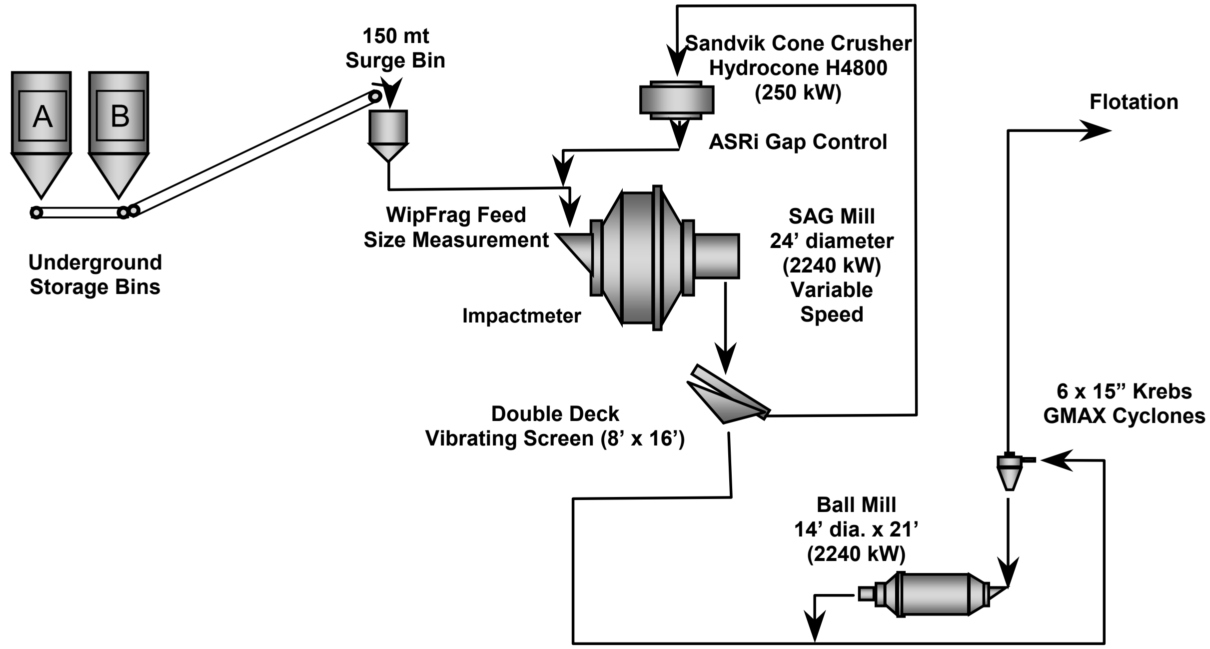


Figure 1: Raglan Concentrator circuit after FAG to SAG Conversion/Optimisation Project

In addition to the new process equipment, the SAG mill was commissioned with all controls implemented on the new PLC software platform. This removed the control strategy limitations previously described. A*dditional* mill control loops and measurements now included:

- Mill *feed tonnage* (automatic PI control – operator or remote set-point from mill *Bearing pressure* controller.)
- Mill water addition (automatic PI control – operator or remote set-point from mill *Density* controller.)

However, during the production ramp-up it was quickly evident that despite much effort, the cascade (PI) bearing pressure control was poor and resulted in large swings in both feed and mill charge (Figure 2). While some success has been reported using PI controllers to maintain charge (Viklund, et al., 2006 ; Edwards, Vien & Perry, 2002), other operations have experienced similar difficulties with this traditional control technique (Veloo, et al., 2006; Van Drunick & Penny, 2006).

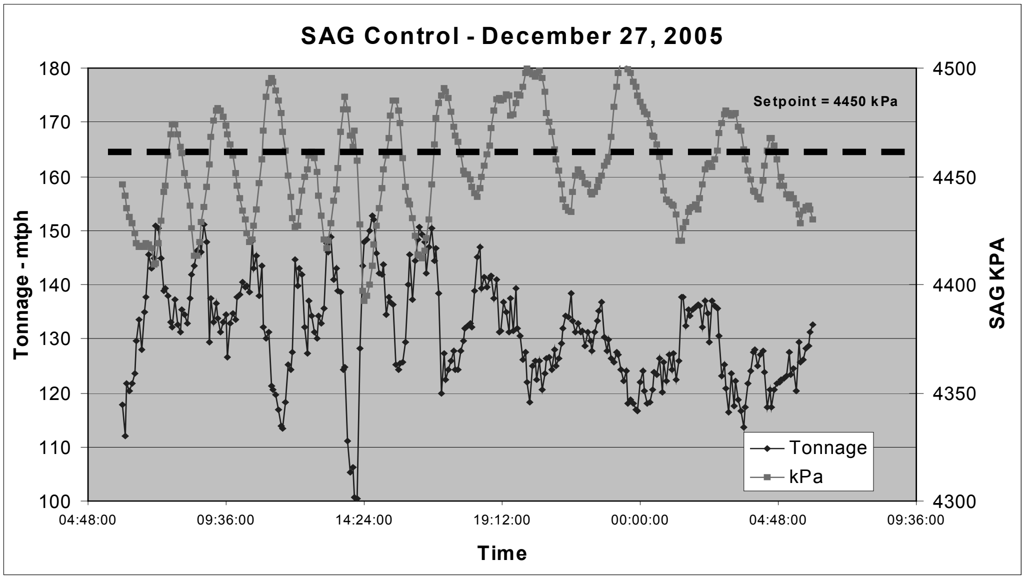


Figure 2: Bearing Pressure control using PI cascade strategy (Hardie, 2006)

The main reasons why conventional PI control was determined to be inadequate for control of mill bearing pressure was:

- Mill feed was highly variable in terms of feed size, grade and hardness.
- The relationship between the feed and pressure was both variable and non-linear in nature.
- There are multiple (and often unexplained) interactions that occur between the various variables.

The automatic pressure control loop was eventually abandoned, and control reverted to the operator making frequent changes to the feed set-point in an attempt to control charge.

Difficulties with maintaining a stable mill charge were also exacerbated by an incident which revealed a thermal limitation in the 3000hp LCI variable speed drive system at low mill speeds. This event resulted in variable speed operation being suspended, and the mill being operated at a fixed speed (76% of critical) until a thorough analysis of the event could be completed.

DEVELOPMENT OF A SAG MILL CONTROLLER

Owing to the highly variable feed conditions, the mill required the constant attention of plant operators in order to prevent overloading or losing the SAG charge resulting in consequent damage to liners. Focus was therefore placed specifically on improving SAG control.

The optimisation project had delivered a mill that should have been set up for success. The control system had full DCS functionality and the following additional instrumentation had been added:

- ASRi (Sandvik) recycle crusher gap control.
- Wipfrag™ feed size distribution measurements – upgraded software and ruggedised installation. This system had already been in operation, but had not been used for control purposes.
- FFE Shell Impactmeter measurements.
- Mill motor current draw and winding temperatures.

Unfortunately little thought had been put into integrating all these new 'toys' into a multivariable controller which was able to consider all the new information simultaneously. It was time to move beyond single-input, single-output (SISO) control!

Which way forward?

Advanced control for the Raglan mill had been on the minds of the concentrator team long before the SAG conversion project. Since the involvement of the Xstrata Process Control Group, both model-based and expert-system approaches had been investigated. It was the recognition that a far stronger *regulatory control base* should first be established, that had postponed further development in this area. Now that this had been delivered, was it not the time to follow through on the above preliminary investigations, and implement an advanced control system?

Fortunately, process control development at Raglan had been guided by a steering committee comprised of both *internal* (Raglan & Xstrata Process Support) and *external* resources (AIA Automation, Soutex & TopControl). The founding vision of this steering committee was: "Raglan does not aim to be the best, but above all to be the most robust" (Jacob, 2005). This vision, resulted in the following guiding philosophies:

- Standardisation ("non negotiable")
- Judicious and low risk introduction of new technology
- Awareness of change management issues

Considering the multiple changes that the concentrator had already been through coming out of the SAG conversion project, there was uncertainty whether the operation was ready to sustain yet another new system in the short term. Many changes had already been implemented over a short period of time. Could not more be done with what had been delivered?

Identifying the opportunities

As mentioned previously, the control system in place now had full 'DCS like' functionality which was only beginning to be exploited. The IEC-61131 standard was developed to allow for PLC programming which was powerful and yet simple to develop and read. There appeared to be potential to leverage this characteristic.

An example of the above is the ability to program *standard* functionality into a single object, (e.g. advanced PID controller) and expose only the inputs and outputs. One can then use this standard object multiple times, *without exposing the complicated internal logic within that* *object*. What is actually a relatively complicated function, can now be viewed from a greatly *simplified perspective.* This makes it possible to program several layers of complexity, without making the PLC program overly difficult to read or develop. It enables the PLC programmer to take a high-level view of the programmed controls, and *drill down* only if required. This *object-orientated* model of software programming has been main-stream in desktop software development for some time, but has only recently been adopted in some industrial process automation platforms.

Technology selection

While model-based approaches have been reported for mill control (Sandoz et al., 1999), it was felt that owing to the high variable and disturbance dominated nature of the SAG operation, a rule based approach would deliver more robustness. This is certainly the technology underlying the majority of mill 'expert systems' in use today (Lanthier, 2007)

Introduction to Fuzzy Logic

Traditional logic results in an evaluation which is always TRUE or FALSE. A property is compared against a limit, and a firm evaluation is reached. For example, one may say that if bathwater is over 50 deg C it is HOT, otherwise it is NOT HOT.

Fuzzy Logic relates to a theory of "ambiguous" or "non binary" interpretation. This is when a statement cannot be definitively described as being either TRUE or FALSE. Most physical measurements do not have a sharp dividing line between their quantitative descriptions. As an example, bathwater is not adequately described as either HOT or COLD, but by the *intermediate stages* in between. The main idea of Fuzzy Logic is to apply a degree of *membership* to the intermediate stages between HOT and COLD. Through the use of fuzzy logic, these intermediate stages can be mathematically described.

Use of further *linguistic terms* such as PLEASANT can give a more comprehensive picture of the quantitative description. For example, we may describe our bathwater as 0% cold, 60% PLEASANT & 40% HOT.

This *fuzzy* quantification is also applicable to consequent control actions. If our control objective is 100% PLEASANT bathwater, we may decide to open the cold water tap "A LITTLE" or "A LOT" depending on how the bathwater responds to the opening of the cold-water tap!

It is not difficult to see how the above principles may be applied to mill control, through evaluation of measured variable (inputs) such as *power consumption* and manipulated variables (outputs) such as *feed-rate set-point*.

Testing the Idea

The Concept programming tool has a built in 'Fuzzy' library, which provided all the components required to build a fuzzy controller within the existing plant controllers. In order to test the principle, a fuzzy controller was configured (on an offline PLC) with very basic (SAG control) rules relating to control of feed tonnage. The controller was then 'fed' the same process data that was available to the operator and the output was compared to actual operator changes to feed tonnage.

Despite the test controller being a very crude *first pass* effort. It was immediately evident that the controller was not only able to mimic operator actions, but was often able to make the required feed tonnage moves *earlier* than the operator did, thereby potentially having a stabilising effect on the process through faster response to disturbances.

It was the above successful demonstration of the potential of a simple *PLC resident* 'fuzzy' controller; that prompted the acceptance of this approach in providing a solution for the SAG control issue, without resorting to the application of any new systems.

Control Strategy

The control objective for SAG milling has historically been centred on producing a stable target grind size (Edwards, Vien & Perry, 2002 ; Viklund et al., 2006). However, high metal prices may shift the control objective to be more about increasing mill throughput . Increasing throughput will generally result in a coarser or less stable grind, resulting in lower or less stable recoveries. In the Raglan case the economic loss caused by the reduced recovery, is small compared to the value of the additional metal produced by the higher tonnages.

Since the existing SAG operation was seeing wide swings in mill charge which was upsetting the secondary grinding and other downstream circuits, the primary control objectives were defined as follows:

- Stabilise mill charge (protect liners, increase mill availability)
- Maximise tonnage throughput whenever possible (increase metal production)

Manipulated variables would be

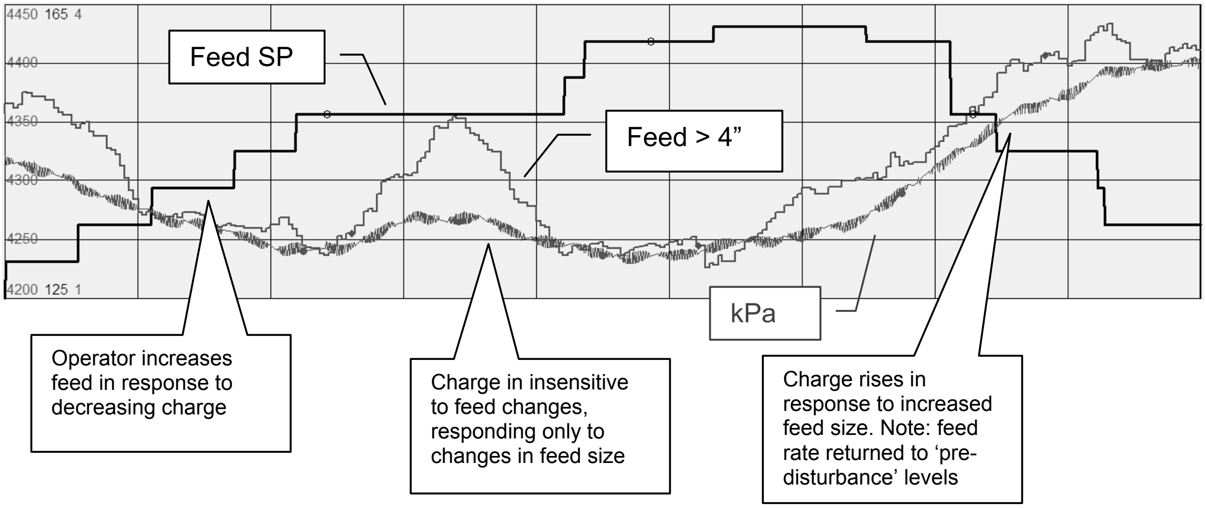
- SAG feed rate (Feed controller set-point)
- SAG water addition (Density controller set-point)
- Crusher gap (ASRi controller set-point)

The Realities

Despite much effort by the industry, accurate measurement of mill charge still represents a challenge. Owing to limitations in the technology, it had been decided not to install a 'charge analyser' on the SAG mill during the 2005 Optimisation project. Retrofit of mill load-cells was also deemed as not feasible. Studies (Evans, 2001) have shown that bearing pressure can be used as a good *indication* of charge level, if one is aware of the sources of error (e.g. density, particle size, ball charge) and the prerequisites (e.g. stable oil temperature etc). It was furthermore felt that once control of bearing pressure had been achieved, a better indication of mill charge could then be sought as a future control parameter.

The lack of control of mill speed was a significant control constraint considering the required control objectives. Mill speed is generally recognised as the most effective manipulated variable in the control of a stable mill charge (Sims, Lacouture, McKay, 2006; Viklund, et al., 2006) and so the lack of any ability to manipulate the speed would be a significant constraint on the effectiveness of the controller.

The relationship between feed size distribution and mill charge is significant. Figure 3 shows a situation where the operator is attempting to maintain mill charge by increasing the feed-rate as the bearing pressure is dropping. Because the feed size is decreasing, this is ineffective in maintaining the charge. The charge is only retained once feed size has returned to normal levels! Owing to this strong relationship and the fixed speed limitation, dealing with these disturbances would thus be on a 'best effort' basis! The value of feed size information as a *feed forward* variable is however important; enabling *pre-emptive* response to a change in feed.



Operator increases feed in response to decreasing charge

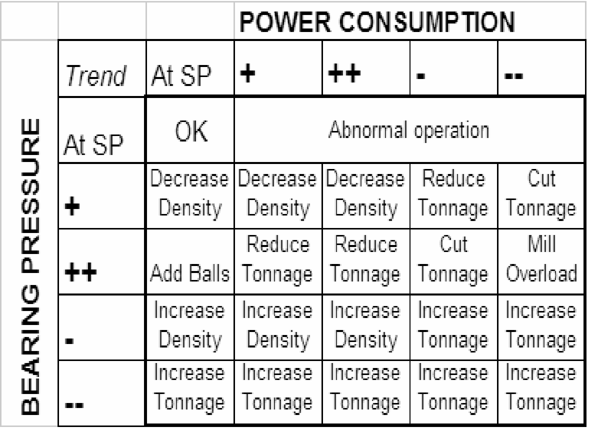
Charge in insensitive to feed changes, responding only to changes in feed size

Charge rises in response to increased feed size. Note: feed rate returned to 'pre-disturbance' levels

Figure 3: Effect of feed size on SAG charge (manual feed control)

Classic control matrix

The basis for the control strategy was derived from the classic control matrix seen in Figure 4 which describes various dynamic relationships between mill power consumption and bearing pressure; and prescribes required action to both feed rate and water addition.



			POWER CONSUMPTION			
	Trend	At SP	+	++	-	--
BEARING PRESSURE	At SP	OK	Abnormal operation			
	+	Decrease Density	Decrease Density	Decrease Density	Reduce Tonnage	Cut Tonnage
	++	Add Balls	Reduce Tonnage	Reduce Tonnage	Cut Tonnage	Mill Overload
	-	Increase Density	Increase Density	Increase Density	Increase Tonnage	Increase Tonnage
	--	Increase Tonnage	Increase Tonnage	Increase Tonnage	Increase Tonnage	Increase Tonnage

Figure 4: SAG Control Matrix

An important aspect in execution of this control matrix was consideration of the *rate-of-change (ROC)* of both the controlled variables. Previously the operator would determine this information by visual inspection of graphic trends. This was both difficult owing to the 'noisy' nature of the measurements and most often resulted in trend behaviour being identified some time after the change had started.

It was therefore required that the ROC of measured variables would have to be *calculated* in order for the controller to function effectively. Owing to the noisy nature of the power and pressure signals, there are significant challenges in producing a 'slope' variable for these parameters. The solution implemented, involved used of multiple stages of 1st order and moving average filters, but it is felt that improvement can still be made in this area.

IMPLEMENTATION

Fuzzy Controller Development

The elements required to build a fuzzy control are indicated in Figure 5. All the tools for implementing these elements were available in the PLC via the *Concept* fuzzy programming library.

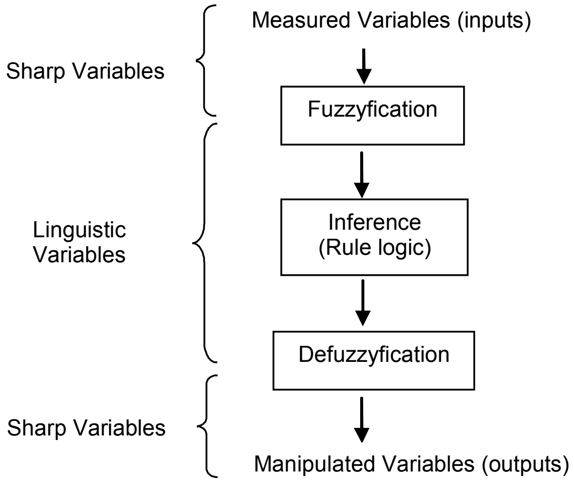


Figure 5: Basic Fuzzy Controller Structure

Linguistic (fuzzy) *variables* were created for power, bearing pressure and for feed size distribution (Wipfrag data). These are shown in Table 1 together with example membership degree values.

Table 1: Fuzzy Input Variables (incl. sample membership degrees in italics)

	Input Variable (sharp)		
Linguistic Variable (fuzzy)	Power	Bearing Pressure	Feed Size > 4”
VERY HIGH	0%	0%	0%
HIGH	***30%***	0%	0%
OK	***70%***	***50%***	***40%***
LOW	0%	***50%***	***60%***
VERY LOW	0%	0%	0%
RISING FAST	0%	0%	0%
RISING	0%	0%	0%
STEADY	***55%***	***65%***	0%
FALLING	***45%***	***35%***	***80%***
FALLING FAST	0%	0%	***20%***

Translation of the control matrix (Figure 4) into a series of rules was the thereafter required in order to process the linguistic variables determined from Table 1. Samples of the resultant rules appear below.

(Rule 3)
IF Bearing pressure (OK or LOW) and (FALLING or STEADY) and
Power (LOW) and (FALLING) then
INCREASE FEED and *DECREASE WATER*

(Rule 5)
IF Power (VERY LOW) and (FALLING FAST) then
INCREASE FEED QUICKLY

Additional rules to detect overload & allow feed-forward of feed size data resulted in a total of 34 rules being developed. Processing of these rules (inference) results in a set of fuzzy output variables (Table 2) being calculated.

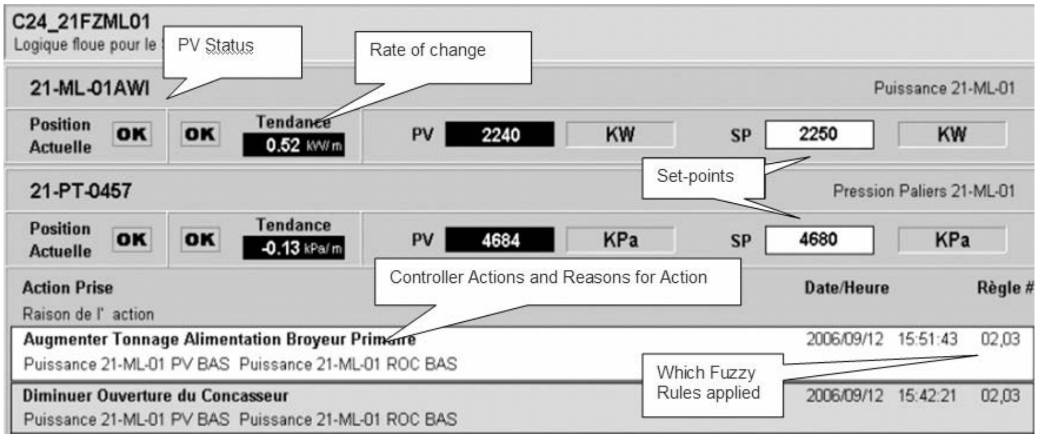
Table 2: Fuzzy Output Variables (incl. sample membership degrees in italics)

	Output Variable (sharp) – Controller SP		
Linguistic Variable (fuzzy)	Feed Rate	Water	Crusher Gap
INCREASE +	0%	0%	0%
INCREASE	***30%***	0%	0%
NO ACTION	***70%***	***35%***	***80%***
DECREASE	0%	***65%***	***20%***
DECREASE +	0%	0%	0%

The process of defuzzyfication determines the most appropriate manipulation of the controller setpoint through evaluation of the membership degree of each fuzzy output variable. As an example, for the membership degrees indicated in Table 2; the controller would decrease the water set-point.

Operator Interface

A *fuzzy controller faceplate object* was developed on the existing WonderWare Intouch HMI (Figure 6). This faceplate allows entry of controller set-points as well as visualisation of current controller status. Each time the controller makes a move, reasons behind the move are indicated to the operator. The two strongest rules active at the time are also indicated for troubleshooting purposes.



Action Prise / Raison de l' action	Date/Heure	Règle #
Augmenter Tonnage Alimentation Broyeur Primaire Puissance 21-ML-01 PV BAS Puissance 21-ML-01 ROC BAS	2006/09/12 15:51:43	02,03
Diminuer Ouverture du Concasseur Puissance 21-ML-01 PV BAS Puissance 21-ML-01 ROC BAS	2006/09/12 15:42:21	02,03

Figure 6: Operator interface (part-of)

COMMISSIONING AND OPERATION

The commissioning strategy was very conservative in order to maintain the confidence of operations. The fly in/out nature of the operation meant a longer commissioning period than would otherwise have been the case. Plant metallurgists and technicians were consulted in order to refine the initial rules. Information sessions were held with plant operators to describe how the controller would work as well as to include as many of the team as possible prior to commissioning.

Advisory Mode operation

Because the controller had been set up to make *discrete changes* to existing controller set-points, the controller could be set to run in an *advisory* mode without actually making changes to the process. This way the operator had the chance to evaluate the controller move, before manually changing the set-point. This was not only useful in terms of fine-tuning the controller, but also useful in gaining the trust of the mill operators.

Initial Controller 'tuning'

The following tuneable parameters were set-up over the course of a few days operation in advisory mode (controller not visible to the operator)

- Membership function breakpoints (e.g High & Very High threshold)
- Rule weighting (importance)
- Magnitude and base interval of control actions

On-line testing

Following an initial tuning period, the controller faceplate was made visible to the operator. A plant metallurgist spent the few hours with each operator the first time he/she was exposed to the new controller. During this initial period, the operator manually executed the *advisory* moves of the controller. While this was an important confidence building step, this soon proved tedious and the controller was switched to supervisory mode at the operator's request. Monitored testing in this mode continued for a total of 14 day shifts, during which time the controller proved capable of both complying with the control strategy and dealing with significant process upsets without operator intervention. At this point it was deemed commissioned and was switched to full time operation. Operators were encouraged to comment on moves made by the controller in the case of any disagreement.

Fine-tuning

The operators kept a 'black-book' in which they could note any controller behaviour which appeared irregular. These comments were used to modify rules or rule weightings. During the first month of operation, the following control deficiencies were identified by operations:

- Changes to feed rate and water addition (mill density) were effected in parallel in order to control charge as quickly as possible. This may result in the mill running at a higher than ideal density, resulting in a lower 'steady-state' feed rate representing a potential loss of throughput. This issue corrected by weighting of the rules relating to increased water addition.
- With experience, rules were modified and automatic constraints were added. An example of this was the inclusion of a *maximum metal tph constraint*, above which the tonnage would not be increased. This prevented overloading downstream circuits during periods of less competent ore.
- There was no unanimous operating strategy for the recycle crusher gap. Normally operated at minimum gap it would only be opened when maximum tonnage was insufficient to maintain mill charge (mill speed could not be lowered). However, continued operation at non-minimum recycle crusher gap setting resulted in other undesirable conditions such as:

    - Change to the kW/kPa relationship
    - Less efficient power utilisation in the mill (Comeau, 2007)

  Incomplete understanding of all of the above phenomena has hampered development of an absolute strategy for recycle crusher gap control, and some improvements could be made in this regard. Current strategy represents a compromise between various opinions.

PERFORMANCE

Utilisation

Even with a statistical approach; evaluation of performance is difficult owing to many factors including:

- Variability of mill feed and consequent variability of *baseline performance capability.*
- Natural tendency of humans to perform better when they being observed/tested (Hawthorne effect)
- Inability to isolate the effect of one change, from other simultaneous changes which may be occurring in the process.

However ***operator utilisation*** is a metric which is very telling to the success of any process control project. Figure 6 illustrates this success, showing a rapid rise in utilisation. Key to the robustness of the controller was the ability for the operator to revert to manual control of each of the manipulated variables ***individually***. This meant that if the operator found fault with (for example) the water addition controller, he/she could take *only that controller* off supervisory control, leaving the other controllers ON supervisory mode. This allowed deficiencies to be identified and corrected. As an example: note the increase in utilisation of the density (water) controller from 57% in March to 96% in April. This increase in utilisation followed an increase to the importance of the rules to decrease density. This resulted in a decreased charge (through increased flush), which left room to increase feed-rate).

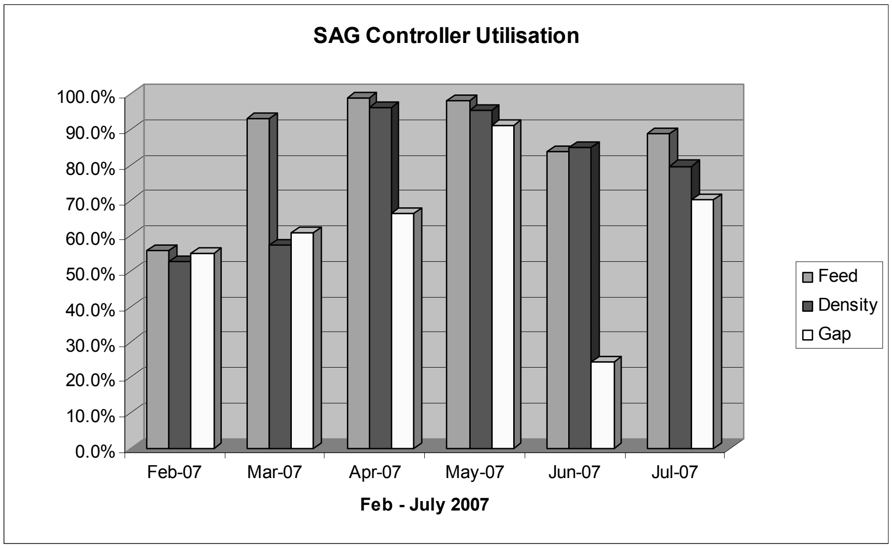


Figure 7: SAG Controller Utilisation (commissioned 17 Feb 2007)

Throughput Increase

Although no formal ON/OFF testing was performed, a throughput increase of between 3% - 6% was indicated during the first 4 months of operation. This represents a possible gain of up to 8tph. This increase in throughput was expected to be associated with a coarser grind and consequent loss in recovery. To determine the economic effect of this reduction in recovery, regressions between Feed-tonnage and grind (P80) were performed. The data is shown in Figure 8.

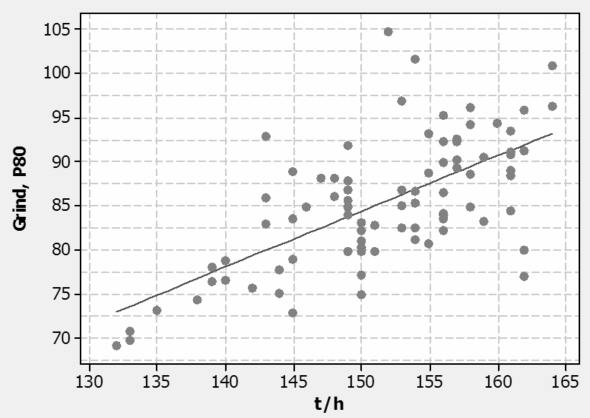


Figure 8: Effect of increased tonnage on grind size and recovery

The fitted curves confirm the expectation that an increase in tonnage does result in a larger grind and consequent reduction in recovery. This recovery loss was quantified to be only 0.09% per additional tph. Even when this loss of recovery was factored in, each additional 1 tph was calculated to worth nearly $4 million per year!

Variability Reduction

A decrease in variability was also observed as shown in Figure 9. The reduction of the variability in the SAG feed rate set-point was particularly significant because this infers a reduction in feed variability to downstream circuits. Therefore the maximum potential of the metallurgy could be exploited owing to less variation in the feed rate and/or feed size.

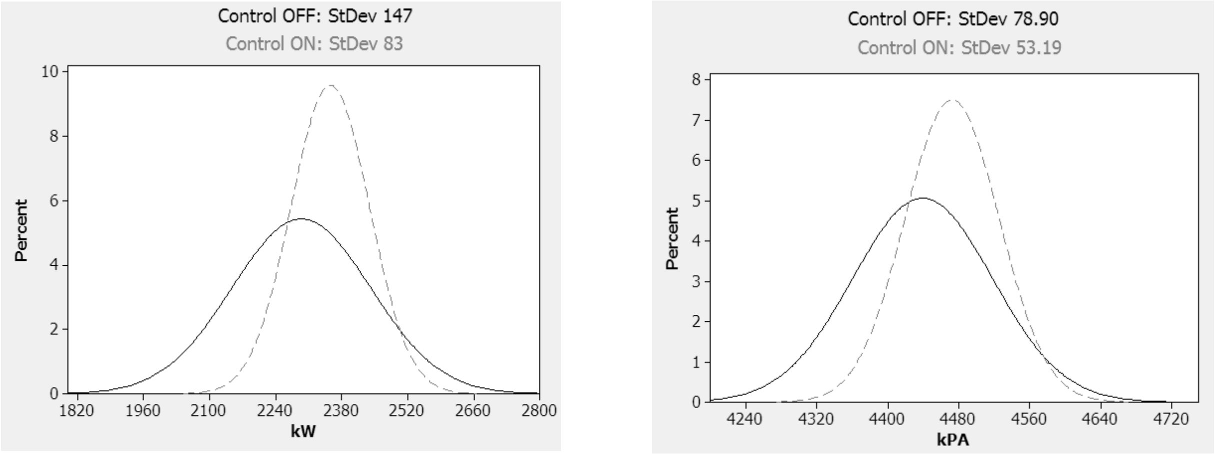


Figure 9: Histograms indicating reduction in variability of SAG mill parameters

Early identification of trend (rate-of-change) behaviour.

In order to respond as early as possible to changing conditions in the SAG, rate-of-change is an important characteristic of SAG mill controlled variables (Austin & Flintoff, 1987). Operator identification of trend behaviour is usually through visual inspection of trend displays. Because of the noisy nature of these measurements, it can take time to interpret the resultant ‘smudge’ as either ascending or descending. Through the use of appropriate filtering techniques (Edwards, Vien & Perry, 2002) it has been shown to be possible to *automatically identify* and respond to these changing conditions preemptively. This is particularly true for the “% Feed size > 4 inches” feed-forward variable (coming from the Wipfrag system), which is only used for its rate-of-change value. (i.e. the *absolute value* is not used in the control strategy).

Advantage of constant vigilance

The grinding operator has other duties which take him/her out of the control booth from time to time. This may result in less than optimum throughput through running the SAG conservatively to avoid problems while it is not being ‘attended’. This was particularly evident during shift changes.

During the first few weeks of controller operation it was usual for operators to continue this practice by placing the controller OFF supervisory control during shift change. This practice soon faded away however as operators gained confidence in the controller to deal with most disturbances.

Consistent response to process conditions

Based on their own experiences, plant operators often have differing ideas of how best to control a process. This may result in each shift running the plant with a somewhat different operating strategy. This situation is obviously sub-optimum, and is difficult for the plant metallurgist to deal with from a *change management* perspective. The controller obviously applies the rules and consequent control actions in a consistent manner.

It was nonetheless found to be important to keep an open mind and not be dogmatic about how the mill should be operated. Many tests and subsequent modifications were performed in order to find the most robust compromise solution that would be accepted by all. While there remain some ‘discussions’, the SAG controller has been useful in unifying the operating strategy of the SAG circuit.

Embedded Controller versus Expert System

The ‘under-the-hood’ control principles and technology of the PLC based SAG controller are similar to those applied in full-blown ‘expert systems’. While the estimated cost of using the existing control system to implement an ‘expert’ controller is significantly less than that of a vendor supplied expert system, this was never a motivation for the project. (As already discussed, the potential rate-of-return offered by good SAG control is so high that it makes these projects relatively insensitive to high initial capital costs.) No! The motivation behind this approach was to leverage existing resources and not add additional complexity and additional external support requirements. This was deemed particularly important given the limitations of a fly-in fly out operation.

It is recognized that the embedded approach does provide less in the way of *Tools* and *Diagnostic Utilities*, which had to be created as deemed required. There is no denying that vendor supplied packages may offer a more ‘flashy’ and ‘polished’ result. However they do not necessarily deliver more in terms of performance.

The truth is that the *fuzzy rules* approach is so robust, that it doesn’t have to be *perfect* to be successful! If one uses the analogy of the controller being a committee of mill operators, metallurgists and engineers, continually voting on what control actions are to be taken, one can see that even spurious recommendations will normally be filtered out by the others around the table. This has been seen in this implementation. Even if ‘questionable’ moves are made, they are quickly automatically reversed. The decision must then be made whether these occasional ‘rule/logic glitches’ are worth troubleshooting, or whether process control resources are best applied to other plant areas.

Robustness of the embedded approach is unsurpassed. All the control logic runs on the same control platform which controls the rest of the plant and is not vulnerable to failure of communications to 3rd party system. These 3rd party systems are often Microsoft Windows based, and are therefore subject to associated vulnerabilities and instabilities (!).

FUTURE WORK

Incorporation of Shell Impact Information.

Owing to the relatively low ball charge (5-6%) of the Raglan SAG mill, the impactmeter has not yet delivered information which can be reliably used for *automatic* process control. There is however the potential for this instrument to deliver information about SAG charge which is not evident by bearing pressure (e.g. indication when the charge appears to 'stick' to the shell). This is important to prevent liner damage caused by low charge levels.

Restoration of Variable Speed

The lack of ability to vary the speed of the mill has presented significant challenges to having effective control of charge. Controlling the charge with feed and water only has been likened to controlling the speed of a car by opening/closing windows and doors while leaving the accelerator untouched!

While some progress has been made towards re-instating control of mill speed (additional motor and transformer cooling) the complete evaluation of the drive-train capability is not complete.

Determination of optimum charge (kPa) set-point

Analysis of performance data indicates that variability decreases and throughput increases at lower kPa set-points. This increase in throughput as lower rock charge levels has been observed at other operations (Dunne, et al., 2001). Work to determine the *optimum* kPa set-point remains to be done as well as testing to determined optimum ball to ore ratio.

Improved Charge measurement

It is recognised that bearing pressure is not a replacement for a true charge measurement. Since the ability to control around desired kPa set-point has now been obtained, it may be useful to identify a more acute indication of actual charge. Appropriate technologies are being investigated.

CONCLUSION

Raglan SAG mill process control development has resulted in a controller which automates mill operation, increases throughput and reduces variability. Utilisation of the controller is consistently greater than 96% and has required little maintenance in the way of rule modification.

An upgrade of the plant control software to IEC-61131 delivered DCS like functionality which allowed the fuzzy logic rule based controller to be programmed directly on the plant control system using standard libraries. Since the controller is embedded into the plant control system it is extremely robust, being completely independent of 3rd party software and systems. This reduces overall control system complexity for Raglan maintenance personnel.

Effective change management proved to be essential for a successful implementation.

Progress is currently underway to extend this control approach to other areas of the concentrator operation.

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