ABSTRACT

The main objective of this investigation was to demonstrate the suitability of a state-of-the-art fine coal cleaning and dewatering circuit through on-site testing. The demonstrated circuit included a newly developed centrifugal flotation system, known as the Imhoflot™ G-Cell, and the Steel Belt Filter (SBF) dewatering system. The Creek Paum Mine of the Knight Hawk Coal Company, located at Ava Illinois, served as the host-site for this demonstration project.

Highlights of the experimental results obtained from this study include high rejection of sulfur and ash as well as satisfactory clean coal yield obtained from the G-Cell. A slip-stream from the plant cyclone overflow stream was utilized as the G-Cell feed slurry. It had an ash content varying from 40% to 60%, feed solid content between 3% to 6% by weight, and sulfur content of nearly 1.5%. The lowest ash content produced from G-Cell treatment was in the range of 7% to 8% at yield values in the range of 35% to 40%. The maximum sulfur rejection achieved from the G-Cell operation was nearly 70%. Based on the results obtained from a long-term test program, the 95% confidence interval for the true mean ash content of the plant cyclone overflow (G-Cell feed) stream was predicted to be between 47.6% and 55.3%. The corresponding confidence intervals for the mean mass yield and the mean product ash content obtained from the G-Cell and SBF circuit were 33.0% to 37.1% and 10.2% to 11.6%, respectively. The confidence interval for the circuit product moisture (surface moisture) content established over the same time period was 38.3% to 40.7%. The relatively high moisture content of the product was attributed to the extremely fine size consist of the flotation concentrate having a $d_{80}$ (80% passing size) of 60 micron.

An economic analysis conducted indicates the requirement of a $1.37 million total investment for this type of fine coal circuit in an average size coal preparation plant treating nearly 700 ton per hour (tph) of raw coal. The G-Cell and SBF fine coal circuit, having a capacity of 35 tph of raw coal in the size range of minus 150 micron, may generate additional annual income of $3.18 million. This translates to a payback period of less than 6 months for the state-of-the-art fine coal cleaning and dewatering circuit demonstrated in this project. The cost of fine coal recovery was estimated to be $5.00/ton of clean coal and $3.00/ton of raw coal, which suggests a great potential for plant profitability.
EXECUTIVE SUMMARY

Commercially available technologies for dewatering fine coal (finer than 150 micron) are known to not be very cost effective. In addition, froth flotation based technologies commercially available for cleaning fine coal are not particularly suitable for removing sulfur from fine coal due to the weak hydrophobicity associated with coal pyrite particles. Therefore, many coal preparation plants in the US are quite indifferent towards fine coal cleaning. In fact, nearly 50% of the coal preparation plants in Illinois discard their fine coal, i.e. coal finer than 150 micron, without any attempt to recover the clean coal from this size fraction of the run-of-mine coal. The Creek Paum Mine of the Knight Hawk Coal Company mines high-sulfur coal in Illinois and is one such operation that continues the aforementioned wasteful practice due to the lack of effective and relatively inexpensive fine coal cleaning and dewatering methods. The coal company participated in this project by providing its Creek Paum Mine to serve as the host-site to help evaluate the proposed fine coal recovery system.

The new circuit is a combination of a novel centrifugal flotation system, known as the Imhoflot\textsuperscript{TM} G-Cell flotation technology, and the innovative Steel Belt Filter (SBF) dewatering system. Originally developed in Europe, the Imhoflot\textsuperscript{TM} G-Cell technology has been tested for the first time in the US as a part of this investigation. Similar to the Jameson Cell technology, the G-Cell is a self-aspirating flotation system, in which the bubble-particle attachment takes place within seconds inside a downcomer operating with a very high air fraction. However, unlike the Jameson Cell and Muto-Tech Column technologies, the coal slurry containing particle-bubble aggregates is introduced into the separation vessel of the G-Cell through tangential entries, thus creating a centrifugal field due to the rotary motion of the coal slurry. This phenomenon provides two special features to the G-Cell technology. First, fine coal pyrite particles, which may be weakly attached to the air bubbles, are effectively rejected to the vessel-wall and ultimately to the tailings stream under the action of centrifugal force. Second, the resulting centrifugal force also helps in enhancing the separation kinetics between particle-laden air bubbles, which rise due to buoyancy effect and the hydrophilic gangues (ash bearing mineral particles), which sink and report to the tailing stream. Experimental observations with the G-Cell suggest that a residence time of 30 to 60 seconds in the separation vessel may be sufficient to achieve a complete separation. Therefore, the G-Cell separation vessel of a specific size can treat a significantly larger volumetric flow of coal slurry in comparison to column flotation technologies, for which nearly 3 minutes of residence time may be commonly recommended for effective separation. This phenomenon may make the volumetric throughput capacity of the G-Cell technology 3 to 6 times greater in comparison to other advanced flotation technologies.

The SBF dewatering system (a South-African technology), the other component of the fine coal recovery circuit evaluated, was extensively studied as an individual unit operation for the first time in the US as a part of a previous ICCI-funded research project (Mohanty, 2003). The unique features of this innovative technology include the combined use of pressure and vacuum, which enables it to achieve the maximum reduction of moisture content at a reasonably low flocculant dosage.
During the initial project months, a 0.8 meter diameter G-Cell unit was procured from the equipment vendor and assembled at the Illinois Coal Development Park. The G-Cell unit was transported and installed at the plant site on a suitable level ground along with its ancillary equipment as well as the pump and sump assembly. Several series of exploratory tests were conducted to not only obtain a better understanding of the G-Cell process but also establish the operating parameter ranges for the subsequent optimization test program. A slip-stream obtained from the overflow stream of the raw coal cyclones operating in the plant was used as the feed slurry for the G-Cell unit. After completion of the optimization test program for the G-Cell, an SBF unit consisting of a 0.6 meter wide belt, was integrated into the test-circuit to dewater the G-Cell product (flotation concentrate). A long term test program was conducted over a period of three days to study fluctuation in the feed characteristics and the resulting changes in clean coal quantity and quality recovered from the new fine coal circuit.

Highlights of the experimental results obtained from this project include high rejection of sulfur, relatively low ash content of product and satisfactory clean coal yield obtained from the G-Cell. It was inferred that a two stage (routher-scavenger) G-Cell operation will be required to obtain high clean coal yield. The feed slurry treated by the G-Cell and SBF circuit had an ash content in the range of 40% to 60%, feed solid content between 3% to 6% by weight, and sulfur content of nearly 1.5%. The lowest ash content produced from G-Cell treatment was in the range of 7% to 8% at yield values in the range of 35% to 40%. The maximum sulfur rejection achieved from the G-Cell operation was nearly 70%. Based on the results obtained from the long-term test program, the 95% confidence interval for the mean ash content of the plant cyclone overflow (G-Cell feed) stream was found to be between 47.6% and 55.3%. The corresponding confidence intervals for the true mean mass yield and the mean product ash content obtained from the G-Cell and SBF circuit were 33.0% to 37.1% and 10.2% to 11.6%, respectively. The confidence interval for the circuit product moisture (surface moisture) content established over the same time period was 38.3% to 40.7%. The relatively high moisture content of the product was attributed to the extremely fine size consist of the flotation concentrate that had a d80 of 60 micron.

An economic analysis conducted in consultation with manufacturers of the G-Cell and SBF indicates that a total investment of $1.37 million would be required for the proposed fine coal circuit for an average size coal preparation plant treating nearly 700 ton per hour (tph) of raw coal. The G-Cell and SBF circuit, having a cleaning capacity of 35 tph and 2500 gallon per minute (gpm) of raw coal slurry in the particle size range of minus 150 micron, may generate additional annual income of $3.18 million. This translates to a payback period of less than 6 months for the G-Cell and SBF circuit demonstrated in this project. By amortizing the total capital and installation cost over a period of 15 years using a discount rate of 12%, the cost of recovering (cleaning and dewatering) fine coal using this circuit was calculated to be $5.00 per ton of dry clean coal and $3.00 per ton of dry raw coal. This per ton cost may be significantly lowered by dewatering flotation concentrate and spiral product together in the SBF unit instead of flotation concentrate alone.
OBJECTIVES

The overall goal of this project was to demonstrate a high-efficiency fine coal cleaning and dewatering circuit in the field for its near-term commercialization in coal preparation plants in Illinois. In order to accomplish this overall goal, the specific research objectives were:

- To optimize the performance of an Imhoflot\textsuperscript{TM} G-Cell to effectively clean high sulfur Illinois coal fines using a reasonably large pilot-scale unit in a plant environment.
- To optimize the performance of a fine coal cleaning and dewatering circuit including a G-Cell and a Steel Belt Filter and conduct a sensitivity analysis of the circuit performance with respect to a fluctuating plant environment.
- To conduct an economic analysis to justify the commercial viability of the new fine coal cleaning and dewatering circuit.

INTRODUCTION AND BACKGROUND

**Fine Coal Cleaning:** The common fine coal cleaning technologies already in commercial use in the US include:

- Conventional sub-aeration flotation cell: Most prominent are Denver and Wemco Cells.
- Flotation Columns: The flotation columns commercialized in the US coal industry include:
  - MicroCell\textsuperscript{TM} Column
  - Jamseon Cell
  - CPT (Canadian Process Technology) Column
  - Turbo Column

The majority of coal preparation plants in Illinois that have a fine coal cleaning circuit use conventional froth flotation cells to clean the minus 150 micron (100 mesh) size coals. Since conventional froth flotation cells fail to effectively treat the ultrafine high ash materials due to the absence of a wash water system, many preparation plants deslime their minus 150 micron coal slurry prior to flotation cleaning. However, the desliming achieved by small diameter cyclones allows a significant amount of ultrafine misplacement to the flotation feed, and thus, the product quality is negatively impacted. To address this issue, a high-efficiency froth washing system is being developed by the Illinois State Geological Survey with research funds received from DCEO/ICCI (Khan, 1998, 1999). This froth-washer can be retrofitted to conventional flotation cells as well as flotation columns to produce a high-quality clean coal product.

Flotation columns were commercialized in the US coal industry in the later half of the 1990s. Although, the flotation column product has typically very low ash content, column
throughput is relatively low. Evidently, the use of flotation columns has been very limited in the State of Illinois. Several past studies (Honaker et al., 1998; Luttrell et al., 1996; Honaker, 1995) have proved the inferior removal of coal pyrites in flotation processes in comparison to enhanced gravity concentrators, such as the Falcon concentrator. However, these enhanced gravity separators provide excellent ash and pyrite cleaning performance only in the particle size range of 600 x 45 micron. Their efficiency significantly deteriorates below a particle size of 45 micron.

**Fine Coal Dewatering:** The common methods of dewatering fine clean coals and fine tailings materials in the coal mining industry include:

- **Mechanical Dewatering:**
  - Screen bowl centrifuge
  - Solid bowl centrifuge
- **Filtration:**
  - Vacuum disk filter
  - Vacuum drum filter
  - Horizontal vacuum belt filter
  - Plate and frame filter press
  - Belt filter press

The mechanical dewatering process, especially screen-bowl centrifugation, is widely used for dewatering the relatively coarser portion of the fine coal product. The optimum particle size range for screen bowl centrifuge dewatering is 1 mm x 45 micron. The perforated portion of the screen bowl, which helps reduce the product moisture content, also causes the loss of a significant portion of the ultrafine clean coal in the size range below 45 micron. The centrate stream is commonly recirculated back to the feed system on a continuous basis to prevent the loss of significant amount of ultrafine clean coal particles present in the centrate. However, this phenomenon effectively reduces the fresh feed handling capacity of the screen bowl centrifuge. On the other hand, the solid bowl centrifuge tends to achieve a near complete recovery of solids to the product stream but at the cost of higher moisture content despite the use of nearly 3000 g-force. Thus, the solid bowls are occasionally used for tailings dewatering but rarely for clean coal dewatering.

Vacuum disk filtration is a commonly used process for dewatering clean coal fines in the particle size range of minus 150 micron in preparation plants in the US, whereas the horizontal vacuum belt filter is used more in Australia. Although moisture content of 20% to 25% is commonly achievable with these filtration processes, the presence of higher proportions of minus 45 micron size material in the feed slurry typically results in surface moisture content as high as 40%. The necessity of using special polypropylene filter cloth having a very short useful life also adds to the high operating cost. The pressure filters such as plate and frame presses are extremely useful for dewatering feed slurry containing higher proportion of minus 45 micron material. Product moisture content of nearly 25% is commonly achieved for clean coal fines containing more than 80% minus 45 micron solid particles. However, the plate and frame presses are well
known for their high capital and operating costs. For example, a plate and frame press having a product throughput capacity of 30 tph may cost more than $1 million. Also, operating cost is high because quick-wearing filter cloths need to be replaced at frequent intervals.

The drawbacks of existing fine coal cleaning and dewatering technologies suggest the need for the present investigation evaluating a new fine coal recovery circuit consisting of the Imhoflot G-Cell and the PSS Steel Belt Filter (SBF). A brief description of these two technologies is provided next.

**Imhoflot™ G-Cell:** The G-Cell technology is a new and improved version of the Imhoflot pneumatic flotation technology, which was developed at the Technical University of Clausthal, Germany more than 20 years ago (Bahr et al., 1982). The original Imhoflot technology is a self-aspirating device, in which bubble-particle contact takes place inside a vertical downcomer in a highly turbulent water-air phase having a very high air fraction. The downcomer discharges into a vertical separating vessel, where a crowder mechanism is used to separate froth from the tailings slurry. Thus, the bubble-particle contact and froth separation takes place in two physically separate sections of the device. Although, bubble-particle attachment takes place in a few seconds inside the downcomer due to a highly congenial flotation environment, nearly two to three minutes of residence time is allowed in the vertical cell for the separation of froth. This latter sub-process was considered unnecessary and that led to the development of the G-Cell, in which the bubble-particle aggregates from the downcomer are injected tangentially at multiple peripheral-locations into a static bowl as shown in the schematic diagram in Figure 1. The inlet nozzle sizes and the volumetric flow rate determine the rotational speed of the flotation slurry containing the bubble-particle aggregates. The rotational speed generates a centrifugal field that causes the interface between the pulp and the froth to incline towards the center of rotation (Battersby et al., 2003). This inclined interface facilitates a smooth flow of the froth to a central launder. The height of the launder is an operating variable to control the froth height in the bowl. The remaining slurry in the bowl is discharged through tangential tailings outlet of adjustable size. A pilot scale investigation of the G-Cell technology was recently completed in Germany (Battersby et al., 2003) and South Africa (Lubbe, 2003). The test model is shown in Figure 2.

**Steel Belt Filter (SBF):** The Steel Belt Filter (SBF) technology used in the test circuit was invented in South Africa by Particle Separation Systems (PSS). A pilot scale SBF Model SBF 600, shown in Figure 3 was tested at the Illinois Coal Development Park as a part of a previous DCEO/ICCI project (Mohanty, 2003). As shown in the schematic diagram of Figure 4, the SBF unit consists of two special woven steel mesh belts traveling at the same speed with the top belt in the anti-clockwise and bottom belt in the clock-wise direction. The feed slurry is evenly introduced on the bottom belt by a feed distributor so that the entire width of the belt is covered by the slurry. Before getting pressed between the two belts, the slurry material is pre-dewatered under the action of vacuum by the vacuum boxes placed underneath the bottom belt. The thickened slurry is then subjected to mechanical pressure between the two belts through a series of offset top
Figure 1. A schematic diagram of an Imhoflot™ G-Cell (Battersby et al., 2003)

Figure 2. Imhoflot G-Cell Model IMF G22 (Battersby et al., 2003)
And bottom rollers. In the total system, water is withdrawn from the solid material under the action of simultaneous mechanical pressure and suction force provided by the vacuum boxes placed in the inner side of bottom belt. Thus, the SBF technology combines the dewatering action of two conventional dewatering technologies; i.e., horizontal vacuum belt filter and conventional belt filter press into one novel system. The unique features of the SBF technology were presented in a previous final technical report (Mohanty, 2003) submitted to the ICCI.

Figure 3. PSS™-SBF Model 600 being used in the Illinois Coal Development Park

Figure 4. A schematic diagram of the SBF technology (Buisman, 2002)
As a part of this project, the G-Cell technology was optimized using a statistically designed experimental program for its possible application to clean Illinois coal fines. After identifying the optimum operating condition for both G-Cell and SBF, a long-term test program was conducted over a period of three days to do a sensitivity analysis of the circuit performance with respect to the fluctuating feed characteristics in the plant environment at the Creek Paum mine site. An economic analysis was conducted to determine the pay-back period for the proposed fine coal circuit (consisting of full-scale units) proposed in this project and also to determine cost per ton of fine coal recovery.

**EXPERIMENTAL PROCEDURES**

The experimental program for this project was conducted in two phases. The first phase concentrated on optimizing the performance of the G-Cell since this was the first experimental program using the G-Cell to process US coal. In the second phase, a complete fine coal circuit consisting of a 0.8 meter diameter G-Cell as the cleaning unit and a 0.6 meter wide SBF unit as the clean coal dewatering unit was tested. Photographs of the on-site test setup are provided in the Appendix.

As shown in the experimental layouts (Figures 5 a & b), a slip-stream was obtained from the overflow of the classifying cyclones operating inside the plant at the Creek Paum Mine. Being a high ash stream of nominally minus 150 micron particle size, currently, this stream is rejected directly to the plant thickener without any attempt to recover the clean coal present. The slip-stream from the cyclone overflow discharged directly to the feed sump having a capacity of nearly 800 liters. A majority of the G-Cell tests were conducted without any tailings recirculation. A batch of 800 liters of flotation feed was collected in the feed sump and conditioned with the desired dosage of fuel oil (#2) for nearly five minutes before beginning the flotation tests. A variety of frother types were tested during the experiment; subject to its solubility in water, a frother was either added directly to the feed line or the feed sump at the desired concentrations. The tests conducted with a partial tailings re-circulation were carried out with a continuous supply of plant feed slurry and partly recirculated tailings to the feed sump. The amount of fresh feed material had to be adjusted by controlling a valve to maintain a constant level in the feed sump. For example, to conduct a test with a 25% tailings recirculation and G-Cell feed flow rate of 100 L/Min, 75 L/Min of fresh feed and 25 L/Min of tailings material were continuously supplied to the feed sump.

Initially, several series of exploratory tests were conducted to select a proper size of feed inlet orifice, select a suitable type of frother, gather an estimate of the plant variability and establish the range of values for the key process parameters to be varied in the optimization test program. The process parameters and their respective range of values selected based on these exploratory tests are listed in Table 1. As shown, four process parameters, including feed flow rate, wash water rate, frother concentration and collector dosages, were varied in the range of 80 to 100 L/min, 5 to 8 L/Min, 10 to 20 ppm and 0.4
Figure 5(a). A schematic diagram of the experimental layout used during the G-Cell optimization test program

Figure 5(b). A schematic diagram of the experimental layout used during the G-Cell and SBF circuit test program
Table 1. Key operating parameters of the G-Cell and their ranges maintained during the optimization test program

<table>
<thead>
<tr>
<th>Operating Parameter</th>
<th>Low</th>
<th>Medium</th>
<th>High</th>
</tr>
</thead>
<tbody>
<tr>
<td>Feed flow rate</td>
<td>80 L/Min</td>
<td>90 L/Min</td>
<td>100 L/Min</td>
</tr>
<tr>
<td>Wash water rate</td>
<td>5.0 L/Min</td>
<td>6.5 L/Min</td>
<td>8.0 L/Min</td>
</tr>
<tr>
<td>Frother Concentration</td>
<td>10 ppm</td>
<td>15 ppm</td>
<td>20 ppm</td>
</tr>
<tr>
<td>Collector Dosage</td>
<td>0.4 kg/tonne</td>
<td>0.6 kg/tonne</td>
<td>0.8 kg/tonne</td>
</tr>
</tbody>
</table>

Feed pressure varied from 1.7 Bar (25 psi) to 2.4 Bar (35 psi) by changing the feed flow rate in the aforementioned range. The height of the tailings discharge port (which controls the froth height in the G-Cell) was maintained constant at nearly 12 cm below the lip of the internal product launder since it was not practicable to vary it over a reasonably wide range. The other process parameters that were maintained constant included the Pine oil type of frother and the orifice size of 10 mm for the four feed inlets. In total, 29 tests were conducted during the optimization test program in accordance with a test matrix generated utilizing Central Composite Design. A set of three samples (feed, product and tailings) was collected for each G-Cell test for subsequent analysis of different types of assays, including ash and sulfur. A composite feed sample was collected during the entire optimization test program lasting nearly sixteen hours for conducting release analysis.

Upon completion of the G-Cell optimization test program, the SBF unit was integrated to the layout, as shown in Figure 5 (b), to dewater the G-Cell concentrate on a continuous basis. SBF operating parameters were previously optimized (Mohanty, 2003). Several more rounds of SBF tests were conducted using different flocculant types (both polymer and cellulose based) and dosages to lower the product moisture content. After identifying a suitable flocculant, a long-term test program was conducted by keeping the controllable operating parameters constant for both G-Cell and SBF. This test program continued over eighteen hours on three different days and eleven relevant sets of samples were collected at an average interval of nearly two hours.

RESULTS AND DISCUSSIONS

Task 1: Procurement of Major Equipment

The major equipment purchased for this project included a 0.8 meter diameter G-Cell and an on-line slurry density gauge. Delay in initial project account set up, the delivery of the G-Cell unit and the initial manufacturing defects with the G-Cell unit caused an overall delay in project completion. During the initial project months, the G-Cell unit was assembled on a 22 ft long trailer along with its pump and sump assembly to be transported to the Creek Paum mine site. An on-line slurry density gauge (ALARA) was also purchased to control the solid content of the plant cyclone overflow or, in other words, the G-Cell feed slurry, which varied over a range of 3% to 6%. The ALARA slurry density gauge was calibrated on a 3 inch line. However, fluctuation in the solid
content reading was more than two to three percentage points, which could not be controlled even with help from the technical support group of the equipment vendor (Berthold Inc.). Therefore, it was decided not to use the slurry density gauge during the G-Cell test program.

Task 2: On-site Circuit Installation

A 2 inch diameter slip-stream was obtained from the overflow pipe of the raw coal classifying cyclones operating on the third floor of the plant. A platform for the G-Cell and SBF test program was prepared by pouring a 6 inch thick, 10 ft x 30 ft concrete pad in a relatively level open area near the plant. A special switch box was provided by the plant for electrical connections required to operate the test rig. A mobile diesel air compressor was used for compressed air required to operate the SBF and the diaphragm pump used in the test rig.

Task 3: G-Cell Optimization

Metallurgical Performance:

The maximum cleaning potential of a given coal is measured by the characteristic recovery-grade curve generated from release or tree analysis. As exhibited by Figures 6 (a) through (c), the proximity of the G-Cell data to the recovery-grade curve generated from release analysis is indicative of the superior ash and sulfur cleaning achievable with G-Cell technology. The G-Cell data, shown in Figure 6, were obtained from the experiments conducted as a part of the optimization test program utilizing a Central Composite Design (CCD) test matrix. A few test data had to be eliminated from the analysis due to the extreme variability in plant conditions. The feed quality variation that was considered normal for the plant condition and thus acceptable for analysis ranged from 40% to 50% ash. The sulfur content of the G-Cell feed was typically around 1.4% to 1.5%. The product ash, combustible recovery and sulfur rejection achieved by the G-Cell varied from 6.25% to 17.4%, 23.3% to 69.5% and 41.7% to 89%, respectively. The corresponding ash rejection values ranged from 88.4% to 97.6%. The best separation efficiency of nearly 53% was obtained at a test condition that achieved a mass yield of 45.0%, a combustible recovery of 69.5% and a product ash content of 10.9%. The corresponding ash and sulfur rejection values were 88.4% and 41.7%, respectively. It may be noted that the proportion of organic and inorganic sulfur present in the G-Cell feed was nearly 50:50, which may explain the relatively low sulfur rejection values obtained during this investigation.

Careful examination of the ideal performance curve (in Figure 6 a) produced by the release analysis reveals the possibility of increasing the combustible recovery values up to nearly 80% at a reasonable product ash content of nearly 10%. However, the best combustible recovery values obtained from the G-Cell tests without any tailings recirculation was below 70% at a reasonable product ash content in spite of using significantly higher collector and frother dosages. To further improve combustible
Figure 6. Metallurgical performance data obtained from G-Cell experiments in comparison the ideal flotation performance obtained from Release analysis.
recovery values, additional tests were conducted by recirculating a portion of the tailings. Although, combustible recovery values were increased up to 78.7% and 86.1%, for 25% and 40% tailings recirculation tests, respectively, the corresponding product ash content were above 20%, which is considered excessively high. Based on this observation, it is inferred that a 2-stage rougher-scavenger type of circuit is required for G-Cell operation to obtain high mass yield and combustible recovery values without sacrificing product quality.

Parametric Study:

Empirical models were developed for important process responses, including combustible recovery, product ash and sulfur rejection, achieved from the G-Cell as a function of the aforementioned four key process parameters. Step-wise regression analysis was conducted to include significant main factors and factor interactions in the model equations. A few of the extreme response data, considered as outliers, had to be eliminated while conducting the regression analysis to develop suitable quadratic models with reasonably high $R^2$ values of around 0.85. The model equations are described as follows.

$$\sqrt{CR} = 7.24 - 0.046A + 0.628B + 0.293C + 0.602D + 0.670AB - 0.379BC - 0.745BD - 0.601CD$$ [1]

$$\frac{1}{PA} = 0.113 - 0.032A - 0.003B + 0.003D + 0.028BD - 0.019AD$$ [2]

$$SR = 67.0 - 0.140A - 2.385B - 8.798C - 1.127D + 11.5A^2 - 8.57AC + 13.1BD$$ [3]

where, $CR$, $PA$ and $SR$ represent combustible recovery, product ash and sulfur rejection, respectively. $A$, $B$, $C$ and $D$ represent feed flow rate, frother concentration, collector dosage and wash water rate, respectively in coded terms, which are described by the following expressions.

$$A = \frac{\text{Feed Rate (L/Min)} - 90}{10}$$

$$B = \frac{\text{Frother (ppm)} - 15}{5}$$

$$C = \frac{\text{Collector (kg/tonne)} - 0.6}{0.2}$$

$$D = \frac{\text{Wash Water (L/Min)} - 6.5}{1.5}$$

The main factor effects and interaction effects, which were found to be statistically significant at $\alpha$ value of 0.05 have been included in the above equations to describe the three process responses. By coding the factors in the aforementioned manner, a normalized range, i.e., -1 to +1 is established for each factor so that it is easier to examine the relative effects of operating parameters on the process responses. For example, it would be safe to conclude from Equation 1 that the most influencing main factors for
combustible recovery include frother concentration (B) and wash water rate (D) whereas, the most important factor interactions include BD, AB and CD. Feed rate (A), collector dosage (C), and factor interaction (BC), although statistically significant, have relatively little effect on combustible recovery response.

Although, the above three equations help determine the relative importance of factor main effects and interaction effects on the process responses, the nature of the effect cannot be assessed by examining only these equations. For example, the positive sign of the term D in Equation 1 may appear to indicate that the combustible recovery increases with an increase in wash water rate; however because the CD interaction has a negative sign, it is difficult to ascertain the exact nature of the relationship without a graphical analysis. Therefore, graphical illustrations (Figures 7 through 9) have been utilized to further analyze the parametric effects on the three key process responses.

**Combustible recovery:** Figures 7 (a) through (d) help study the effect of various key process parameters on combustible recovery response in detail. As shown in Figure 7 (a), combustible recovery obtained from the G-Cell increases from 62% to 83% with an increase in feed rate from 80 L/Min to 100 L/Min at the highest level of frother concentration, i.e., 20 ppm. On the other hand, combustible recovery decreases from 43% to 27% for the aforementioned change in feed rate at the lowest level of frother concentration, i.e., 10 ppm. With the aforementioned increase in feed rate, the feed pressure at the venturi is increased from nearly 1.7 Bar (25 psi) to 2.4 Bar (35 psi). Higher feed pressure promotes better shearing action in the downcomer and thus helps produce more fine bubbles in the presence of sufficient amount of frother solution. As a result, more clean coal particles get a chance to be attached to the air bubbles and be carried over to the product launder, thereby increasing combustible recovery. However, in absence of sufficient amount of frother, air bubbles are relatively larger and less of the total bubble surface area is available for clean coal particles. This phenomenon is believed to be causing the decrease in combustible recovery with an increase in feed rate at the lowest frother concentration.

A considerable increase in combustible recovery from 25% to nearly 70% is indicated in Figure 7 (b) due to an increase in frother concentration from 10 ppm to 20 ppm at the lowest collector dosage of 0.4 kg/tonne. An increase of a relatively smaller magnitude is indicated at the highest collector dosage of 0.8 kg/tonne. Since frothers are known to have collecting properties, a portion of the frother is believed to be acting as collectors for coal particles due to an apparently insufficient collector dosage. Understandably, this phenomenon may affect bubble generation process the most at the lowest frother concentration resulting in the lowest combustible recovery. As the frother concentration is increased, the proportion of frother used in the bubble generation processes increases producing finer and more air bubbles. As a result, combustible recovery increases significantly to nearly 70%, which is also nearly the same recovery value achieved with high dosage of collector at the highest frother concentration. It appears that the high collector dosage of 0.8 kg/tonne provides sufficient amount of collector required for enhancing the hydrophobicity of the coal surfaces so that 100% of the frother solution is used in the bubble generation and froth stabilization process.
Figure 7: The parametric effects on combustible recovery obtained from the G-Cell

Figures 7 (c) and (d) illustrate the interaction effects of frother-wash water and collector-wash water. It appears that the mobility of the froth is significantly improved by increasing the wash water level from a low of 5 L/Min to a high of 8 L/Min resulting in near 33% increase in combustible recovery at a frother concentration of 10 ppm. On the other hand, wash water rate appears to have little effect on combustible recovery at the highest frother concentration of 20 ppm indicating a significant frother-wash water interaction. It is quite reasonable to expect an increase in combustible recovery with increasing collector dosage, as shown in Figure 7 (d), especially at low wash water rate.
However, this increase in combustible recovery, which results from the increasing recovery of weakly hydrophobic coal particles due to increasing collector dosages, is reversed at high wash water rate. High wash water rate is believed to force the detachment of weakly attached particles from the froth zone of the G-Cell to its pulp zone and thus, reduce combustible recovery at high collector dosage.

**Product ash content:** As indicated by Equation 2, the main factors A, B and D as well as the factor interactions AD and BD were found to be statistically significant for the product ash response. The nature of these factor main effects and interaction effects are illustrated in Figures 8 (a) and (b). The increased froth mobility due to an increase in wash water rate not only increases combustible recovery, as explained in the previous paragraph, but also allows more water to report to the flotation concentrate launder. Higher amounts of feed water in the concentrate launder result in higher product ash content. An increase in frother concentration from 10 ppm to 20 ppm leads to production of more and finer air bubbles. It is well understood that when availability of bubble surface area is limited, only the more hydrophobic (which are usually the lesser ash content) coal particles are captured by the air bubbles and recovered to the product launder. As more and more bubble surface area becomes available with an increase in frother concentration, coal particles of relatively higher ash content are captured and carried over to the product launder, thus increasing the product ash content. In addition, more air bubble carry-over to the concentrate launder also means more entrainment of feed water to the product. These two phenomena may be causing an increase in product ash content as a function of increasing frother concentration at the lowest water rate of 5 L/Min. However, this increasing trend of product ash content appears to be reversed at high wash water rate of 8 L/Min. It is believed that high wash water rate tends to minimize the entrainment of feed water to product launder. Thus, the product ash content is reduced with an increasing frother concentration at high wash water rate.

As illustrated in Figure 8 (b), product ash content increases as a function of feed rate irrespective of the wash water addition rate. However, the increase is more significant at higher wash water rate. At the lowest feed rate, high wash water rate is effective in reducing the product ash content to as low as 6.0%. However, with the increase in feed flow rate to 100 L/Min, the pulp/froth interface in the G-Cell rises since the tailings discharge is maintained at a constant level of 12 cm from the top of the internal product launder. This phenomenon reduces the amount of froth drainage and thus, increases the amount of water recovered to the product launder. Increased amount of feed water to the product launder results in increased product ash content. In addition, it also appears that higher volumetric flow of wash water also further raises the pulp/froth interface inside the G-Cell causing higher product ash content.

**Sulfur rejection:** As indicated in Equation (3), all four factor main effects as well as AC and BD interaction effects were found to be statistically significant for sulfur rejection response. In addition, it is also clear that sulfur rejection has a quadratic relationship with feed flow rate. The exact nature of these parameter relationships are illustrated in Figures 9 (a) and (b). It may be noted that with an increase in the feed flow rate, the tangential feed inlet velocity inside the separation vessel of the G-Cell also increases resulting in a
Figure 8. The parametric effects on product ash content obtained from G-Cell

Figure 9. The parametric effects on sulfur rejection obtained from G-Cell
centrifugal field of higher magnitude. Therefore, the coal pyrite particles, which are believed to be weakly hydrophobic, tend to be detached from the air bubbles and rejected to the tailings stream under the action of increased centrifugal force at a feed flow rate of 90 L/Min or higher. The effect of centrifugal force on the rejection of coal pyrite particles appears to be minimal at lower feed rates. However, at high collector dosage, the hydrophobicity of the pyrite particle is sufficiently enhanced, a fact which has been observed by other investigators (Olson and Aplan, 1984), to prevent their detachment even under the action of high centrifugal field. On the other hand, as described earlier, the pulp/froth interface rises in the G-Cell with an increasing feed flow rate. This reduces bubble coalescence and increases the recovery of coal pyrite particles along with a majority of the coal combustibles in the presence of sufficient amount of frother. This phenomenon is believed be the cause of a significant decrease in sulfur rejection with an increase in feed rate at the highest level of collector dosage.

As illustrated in Figure 9 (b), the sulfur rejection response increases significantly from 55% to more than 75% with an increase in frother concentration from 10 ppm to 20 ppm at the highest level of wash water. It is believed that the high wash water rate of 8 L/min is sufficient to force the weakly attached coal pyrite particles to drop back to the pulp but allow the more hydrophobic combustibles to remain attached to the air bubbles. Therefore, sulfur rejection increases although combustible rejection decreases due to greater availability of bubble surface area with an increase in frother concentration. This finding agrees well with the previous explanation for the increase in combustible recovery as a function of increasing frother concentration. However, lower wash water rate does not appear to have sufficient force to detach any weakly attached particles from the air bubbles. Therefore, recovery of coal pyrite particles and also organic sulfur increases as a function of increasing frother concentration and thus causes the total sulfur rejection to decrease as shown in Figure 9 (b).

Process Optimization:

A statistical response surface methodology (Box, Hunter and Hunter, 1978) was pursued to identify an optimum experimental region to achieve a desired set of target responses. The empirical models developed for the three key process responses were utilized to generate response surface contours using a commercially available statistical software package. Response surface contour plots generated over the entire range of parameter values for all four operating parameters investigated provided insight into the feasibility of achieving a specific target response value. Then, by overlaying contour plots for each response, an appropriate experimental region was identified to simultaneously achieve each response target. Based on the model predictions about optimum achievable performances, a set of response variables, such as minimum combustible recovery of 70%, maximum product ash content of 13% and minimum sulfur rejection of 50%, was selected as the target performance. Using the aforementioned software package, an appropriate experimental region was located that is shown as the shaded (dark) area in the overlay plot of Figure 10. As indicated, the suitable experimental condition to achieve the selected set of response targets is described by a feed rate in the range of 97 L/Min to 100 L/Min, a frother concentration in the range of 17.5 ppm to nearly 18.5 ppm, a collector
dosage of 0.8 kg/tonne and a wash water rate of 5.0 L/Min. It may be noted that the identified experimental region is not necessarily the only experimental region that can produce the desired response targets. Other similar experimental regions may be found to produce the desired response targets. These experimental regions may be identified by investigating the entire range of other process parameters, which are maintained at a constant level for the above analysis.

Additional G-Cell Tests:

During the experimental program, it was noticed that the feed ash content of the plant cyclone overflow stream occasionally rose to nearly 60%. It was desired to investigate the performance of G-Cell at such extreme conditions by conducting a series of experiments. As shown in Figure 11, the product ash content remained at reasonably low levels until the point that the mass yield value was below 30% and the combustible recovery value was below 65%. Attempts to increase mass yield from the G-Cell by reducing the froth depth below 10 cm resulted in extremely high product ash content, i.e., above 18%.
Figure 11. G-Cell performance obtained while treating coal of extremely high ash contents (nearly 60%)

Task 4: SBF and G-Cell Circuit Testing

Dewatering performance of the SBF technology was thoroughly investigated as a part of a past project (Mohanty, 2003). However, several more series of SBF tests were conducted as a part of this research task to select a suitable flocculant type and dosage. In addition to the polymer-based flocculants, a few cellulose-based dewatering aids were also investigated in an attempt to reducing the moisture content of the SBF product. However, the product moisture content (surface moisture) could not be reduced below 30 to 35% apparently due to the fine size of the flotation concentrate (indicated in Figure 12). Similar observation was made during the previous study (Mohanty, 2003) on dewatering of flotation product alone; it was found that the dewatering of a combined spiral and flotation product enabled reduction of moisture content to below 20%. However, it was not possible to investigate this issue during the present investigation due to technical limitations at the test-site.

Prior to the long-term circuit test, it was desired to investigate the maximum volumetric and mass throughput of the 0.6 meter wide SBF unit tested during this investigation. Without having any significant control on feed solid content of the G-Cell product, the feed flow rate to the SBF was gradually increased from 12 L/Min to a maximum of 36 L/Min to increase the feed solids rate reporting to the SBF unit. The belt speed had to be increased from 0.7 m/min to nearly 1.8 m/min to prevent any spillage (overflow) from the belt with the increase of feed flow rate to the SBF unit. As a result, moisture content of
Figure 12. Particle size consist of the G-Cell feed material and the SBF product obtained during this investigation.

Figure 13. Feed and product throughput versus product moisture content relationship developed during this investigation.
the SBF product increased from nearly 33% to 37.6% as shown in Figure 13. The corresponding increase in product throughput was from 0.24 to 0.74 tph.

Upon establishing the optimum operating conditions for both G-Cell and SBF units, it was desired to investigate the sensitivity of the G-Cell and SBF circuit performance to the continuously fluctuating feed characteristics in a plant environment. Both units were operated for a total of eighteen hours on three consecutive days at their optimum operating conditions. Test samples were collected around the circuit at intervals of nearly two hours; in total eleven sets of samples were collected over the three-day period. Figure 14 (a) indicates the variation in product ash and product moisture contents obtained from the G-Cell and SBF circuit as a function of fluctuating feed (to G-Cell) quality. As shown, there was an increase in product ash content from 9.43% to 12.2% with an increase in feed ash content of 43.7% to 58.2%; however product moisture content varied in an arbitrary manner in the range of 37.4% to 42.4%. As expected, there was a constant decrease in circuit mass yield from 39.2% to 31.3% with the aforementioned change in feed ash content. The change is combustible recovery was more arbitrary in the range of 67.2% to 61.7%, as shown in Figure 14 (b).

The data obtained from the aforementioned eleven tests were statistically analyzed to establish a 95% confidence interval for the plant feed quality and the corresponding performance parameters of the G-Cell and SBF circuit. The 95% confidence interval for the true mean ash content of the plant cyclone overflow stream was predicted to be between 47.6% and 55.3%. The corresponding lower and upper limits of the true mean mass yield, product ash and product moisture content were 33.0% and 37.1%, 10.2% and 11.6%, 38.3% and 40.7%, respectively. The relatively high moisture content of the product was attributed to the extremely fine size consist of the G-Cell concentrate. As discussed previously, the moisture content of SBF product could be significantly lowered by adding coarser clean coal (for example, spiral product) to the G-Cell concentrate before the dewatering process.
Figure 14. Test data obtained from circuit tests conducted over a period of 18 hours on 3 days to investigate the sensitivity of the G-Cell and SBF circuit to the fluctuating plant environment.
Task 5: Economic Analysis

An economic analysis was conducted based on technical information obtained from this study, general information about coal preparation plants operating in the US and cost information obtained from equipment manufacturers. The objective was to determine the payback period and fine coal recovery cost for a G-Cell and SBF fine coal circuit suitable for an average size coal preparation plant in the US. The assumed plant capacity was 700 tph of run-of-mine coal. Assuming 5% of the run-of-mine coal is of a particle size finer than 150 micron, the G-Cell feed rate was estimated to be 35 tph and nearly 2500 gpm (based on an average feed solid content of 5%). Based on information from Maelgwin Mineral Services (Imohf, 2005), it is estimated that a G-Cell having a diameter of 3.6 meter (Model G36) will be suitable for treating 2500 gpm of coal slurry. In addition, two G-Cell units will be required to operate in a rougher-scavenger circuit to obtain high mass yield (nearly 60% for an assumed feed ash content of 40%), as discussed previously in this report. Based on a past discussion (Buisman 2003) with PSS, it is estimated that two full-size SBF units will be required to dewater nearly 21 tph (on a dry basis) of G-Cell concentrate of a particle size finer than 150 micron.

Capital and Installation Cost:

The capital cost for the G-Cell units are as follows:

- Cells (2): $70,000
- Aerators (2): $24,000
- Feed distributors (2): $16,000
- Feed Pumps (2): $86,000
- Starters and frequency controllers (2): $90,000

Total capital cost for the two G-Cell units is estimated to be $286,000. The capital cost for two full-scale SBF units (with ancillary equipment) is estimated to be $400,000. Thus, the estimated total capital cost for the complete circuit is $686,000. A minimum of 100% of the capital cost is usually considered as the installation cost by the plant manufacturers (Jain, 1999). Thus, the total capital and installation cost for the circuit is estimated to be $1.37 million. Considering a life of 15 years and discounting at a rate of 12%, the annualized capital plus installation cost is estimated to be $201,000 (=0.1468*$1.37 million).

Operating Cost:

Operating costs for the G-Cell include primarily the costs of power, fuel oil and frother. At a rate of $0.05/kWh, the power cost for operating the feed pumps is estimated to be $7.5/hour. The estimated fuel oil cost is $2.0/hour (@1 lb/ton and $0.06/lb) and frother cost is $16.0/hour (@20 ppm and $0.65/lb). Thus, the total operating cost for the two G-Cell units is estimated to be $25.5/hour. Considering 5,000 work-hours/year, the annual operating cost for the two G-Cell units is estimated to be $127,500.
The annual operating costs for the two SBF full-scale units are listed as follows:

- Belt replacement cost: $20,000
- Rerubbering of rollers: $4,000
- Bearing cost: $2,000
- Flocculant cost: $100,000
- Power cost: $70,000

Thus, the total operating cost for the SBF units is estimated to be $196,000. This means the total annual operating cost for the SBF and G-Cell circuit is $323,500.

Revenue and Payback Period:

A hypothetical coal preparation plant or a slurry pond recovery system, which cleans nearly 35 tph by the G-Cell and SBF circuit, may produce 28 tph (which is equivalent to 21 tph of dry coal) of clean coal having a moisture content of nearly 30%. Assuming a coal selling price of $25/ton and 5,000 work-hour/year, the additional coal production will increase the annual revenue by $3.5 million. As estimated above, the annual cost for operating the G-Cell and SBF circuit is $323,500, which results in an annual net income of $3.18 million. Considering the estimated total capital cost of $1.37 million, a payback period of 5.2 months is expected for the proposed G-Cell and SBF circuit.

Based on previous calculations, the total annualized capital plus installation plus operating cost for the G-Cell and SBF circuit is estimated to be $524,500. Based on 21 tph of dry clean coal achievable from the circuit, the additional clean coal produced per year is estimated to be 105,000 ton. This gives a fine coal recovery cost of nearly $5.00/ton of clean coal and nearly $3.00/ton of raw coal based on 35 tph of raw coal treated by the circuit. In other words, this means that the coal companies lose $20 (= $25- $5) for every ton of fine clean coal lost to the refuse ponds. Commercialization of the proposed fine coal circuit may stop this colossal financial loss for the coal companies operating in Illinois and elsewhere and make them more profitable.

CONCLUSIONS AND RECOMMENDATIONS

- An improved design of a pneumatic flotation technology, known as Imhoflot G-Cell, has been successfully tested and optimized for fine coal cleaning application at a coal preparation plant in Illinois. The best G-Cell performance obtained from cleaning a nominally minus 150 micron size coal stream having an average ash content of nearly 45% may be described by a combustible recovery of nearly 70% at an ash content of nearly 11%.

- The recovery-grade data generated from the G-Cell compared satisfactorily with the ideal performance curve generated from a standard release analysis. However, combustible recovery value above 70% was not achievable from a single stage G-Cell operation without resorting to partial tailings-recirculation, in which case the product
ash contents were unreasonably high. This implies that a two-stage (rougher-scavenger) operation may be required to achieve high mass yield and combustible recovery from G-Cell operation. This hypothesis should be tested in a future study.

- Excellent sulfur rejection values, as high as 70%, were achievable from the G-Cell at fairly high combustible recovery values. The centrifugal field created in the froth-separation vessel of the G-Cell by the tangential entry of the downcomer discharge is believed to force-detach the weakly hydrophobic coal pyrite particles from the air bubbles and thus help achieve excellent rejection of pyritic sulfur.

- A detailed parametric study indicated the major role the parameter interaction effects play in influencing the key process responses such as, combustible recovery, product ash content and sulfur rejection. The frother-feed rate, collector-frother, frother-wash water and collector-wash water interactions were found to be significant in affecting combustible recovery response. Frother-wash water and feed rate-wash water were found to be significant for product ash response, whereas feed rate-collector and frother-wash water interactions were found to be significant for sulfur rejection response. Among all four operating parameters studied, frother concentration in the range of 10 ppm to 20 ppm was found to have the maximum influence on all three key responses.

- Solid recovery of more than 99% was achieved from the SBF dewatering of G-Cell concentrate; however the lowest moisture (surface moisture) content achieved was nearly 30%. This relatively high product moisture content was attributed to the fine size consist of the G-Cell concentrate. It is expected to achieve product moisture content in the range of 20% by dewatering a mixture of spiral product and flotation product instead of flotation product alone. This fact was verified in a past study (Mohanty, 2003).

- The data obtained from a long-term test were statistically analyzed to establish 95% confidence intervals for the plant feed quality and the corresponding performance parameters of the G-Cell and SBF circuit. The 95% confidence interval for the true mean ash content of the plant cyclone overflow stream was predicted to be between 47.6% and 55.3%. The corresponding lower and upper limits of the true mean mass yield, product ash and product moisture content were 33.0% and 37.1%, 10.2% and 11.6%, 38.3% and 40.7%, respectively.

- An economic analysis conducted as a part of this investigation estimates the payback period for a fine coal recovery circuit consisting of a G-Cell and SBF unit having a feed throughput capacity of nearly 2,500 gpm and 35 tph to be less than 6 months. The cost of fine coal recovery using the proposed circuit was estimated to be nearly $5/ton of dry clean coal and $3/ton of dry raw coal. The coal companies in Illinois should realize that they lose $20 per each ton of clean coal discarded to the refuse ponds. For an average size coal preparation plant treating 700 tph, this loss could amount to more than $2.5 million each year.
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APPENDIX

Photographs of the G-Cell and SBF circuit demonstrated at the Creek Paum Mine of Knight Hawk Coal Company