1993 PROCESS ADVANCES AT LAC MINERALS LTD. - EST MALARTIC DIVISION

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PROCESS ADVANCES AT LAC MINERALS LTD. - EST MALARTIC DIVISION

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Abstract

Recent process advances at Lac Minerals Est Malartic Division are discussed. The Est Malartic mill treats massive sulphide, copper/gold ore from the Bousquet 2 mine, to produce a gravity gold product, a copper flotation concentrate and a gold precipitate. The mill operation has been improved by the application of Knelson concentrators to replace jigs, the application of Outukumpu high grade tank cells and optimizing the cyanide leach to minimize cyanide consumption.

Introduction

The Est Malartic mill is located near the town of Malartic in the Abitibi Region of north western Quebec. The mill has been in production since the mid-1930's processing mainly clean, free milling gold ore. Active mining from the Est Malartic mine ceased in 1980 and the mill was switched to custom processing of ore from La Mine Doyon located some 45 km to the west. This custom processing arrangement continued until 1987 when the expanded mill at Doyon became available to process all of the Doyon production. In 1987, Lac Minerals commenced development of a second mine at the Bousquet site located 40 km west of Malartic. This mine, Bousquet 2, is the down dip extension of the La Ronde mine developed by Agnico Eagle. During pre-feasibility evaluations, it was determined that modification of the Est Malartic mill to handle this new ore source was economically more attractive than building a new facility on site. Modifications to the mill to process the high copper/gold ore from the Bousquet 2 mine began in late 1989 and the new circuits were commissioned in mid-1990. Modifications included the addition of both gravity and flotation concentration circuits to recover coarse, free gold and the copper. The existing leach circuit continued operating on the copper flotation tailings with some minor modifications to the solution handling systems. An SO_2/air cyanide destruction circuit was also added at this time.

These initial modifications were only the first of several phases designed to convert the mill to copper/gold processing and to raise the throughput from 1350 to 2270 t/d(1500 to 2500 tons per day).

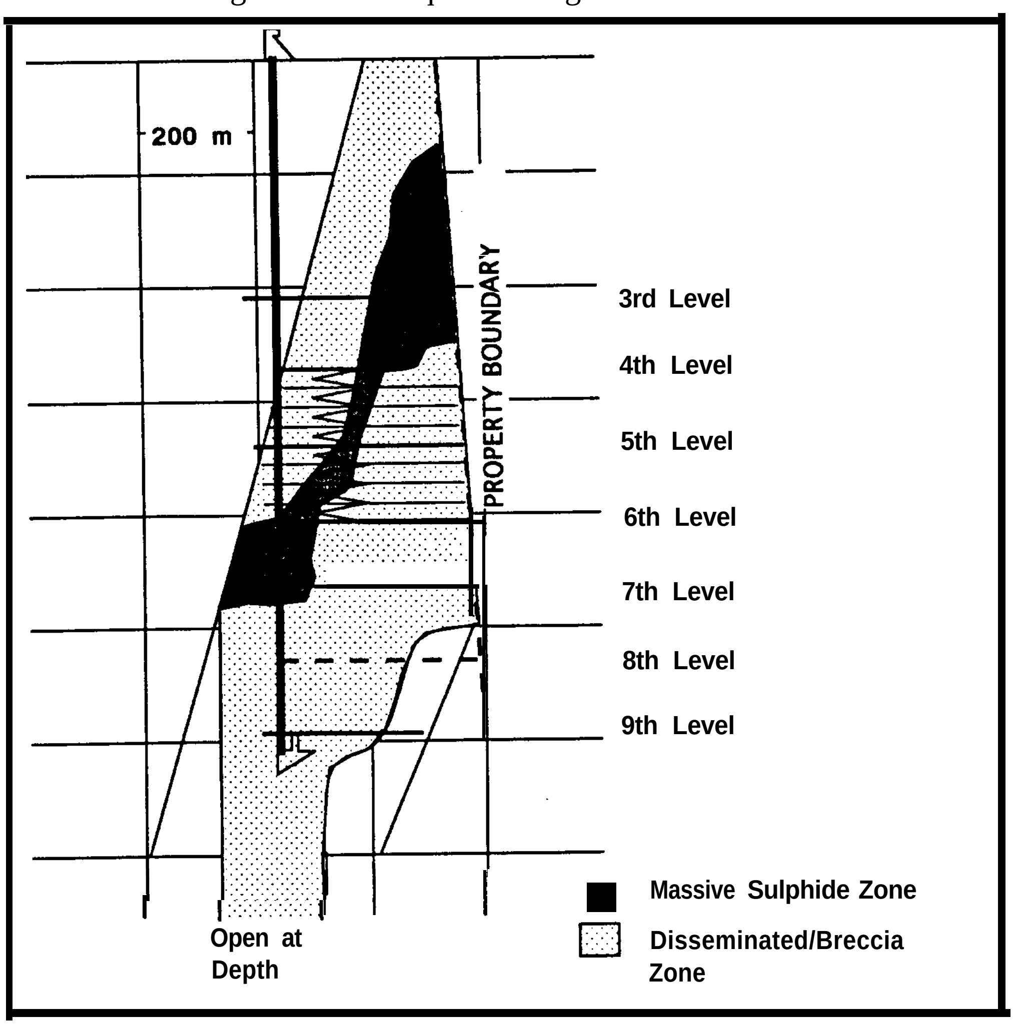
This paper focuses on three aspects of the mill which we consider advances to processing technology.

1)The application of the Knelson concentrator to replace jigs

2)The application of Outokumpu high-grade tank cells to provide increased flotation capacity

3)Improving the leaching performance by optimizing the conditions and reducing the retention time

Figure #l - Bousquet 2 Longitudinal Section



Bousauet 2 Orebody

The main zone of the Bousquet 2 orebody is a massive sulphide zone consisting mainly of pyrite, chalcopyrite and bornite. Parallel zones of brecciated and disseminated sulphides carry lesser amounts of pyrite, copper mineralization and gold. The current ore reserve is 9 million tonnes at 7 g/t (0.2 oz/ton) gold and 0.7 % Cu. Ore from the high grade massive sulphide core has run >2.5 % Cu and carried an ounce per ton of gold. The ore zone plunges at approximately 70° to the west and is nearly vertical north to south. The orebody is accessed via a 5.8 m (19 ft) diameter circular concrete shaft to a depth of 1244 m (4080 ft). The known ore zone extends beyond this depth and will require a shaft deepening program to access the ore below the 9th level. Figure l-Longitudinal Section

Metallurgy

The massive sulphide ore carries significant levels of copper mineralization and therefore dictated a major change in the processing philosophy. The major copper minerals are chalcopyrite, bornite and tennantite. The relative solubility of these minerals in cyanide is shown in Figure 2 and we can see that copper dissolution is a problem, particularly with bornite. The presence of 0.7 to 1.5 % Cu, particularly when predominantly bornite, required the removal of the copper ahead of cyanide leaching to minimize the copper in the leach solution.

Figure #2 - Copper Solubility in Cyanide

Mineral	Formula	Percent Dissolution	
		Leaver & Woolf	Lower & Booth
Azurite	$Cu_3(OH)_2(CO_3)_2$	94.5	91.8
Malachite	$Cu_2(0H)_2CO_3$	90.2	76.9-99.7
Chalcocite	Cu_2S	90.2	54.4-92.6
Covellite	cus		53.2-95.6
Metallic Copper	Cu	90.0	
Cuprite	Cu_20	85.5	82.7-96.6
Bomite	Cu_5FeS_4	70.0	96.0
Enargite	$Cu_3(As,Sb)S4$	65.8	
Tetrahedrite	$Cu_{12}(SB,As)_4S13$	21.9	
Chrysocolla	$CuSi0_3$	11.8	
Chalcopyrite	$CuFeS_2$	5.6	5.8

The removal of copper by flotation, however, created other problems. It was found that the majority (up to 85 %) of the gold reported to the copper concentrate and with the normal 3 month smelter settlement terms, delayed payment for the gold had a negative impact on the cash flow. Also,due to the coarse gold in this concentrate, accurate sampling and therefore payment for the gold was considered a problem. Gravity concentration was thus considered necessary.

Soon after processing of this ore commenced, the sensitivity of the flotation circuit became evident. The flotation recovery is more important in this circuit than normal because of the impact unrecovered copper has on the cyanide consumption and cyanide circuit stability. It was found that beyond a copper tailings grade of 0.2 % Cu, the cyanide circuit becomes unmanageable. In the early days, when the copper head grades were high and erratic, we had to sacrifice concentrate grade to maximize recovery. With the addition of flotation capacity and reduced copper head grades, more normal concentrate grades are now attainable.

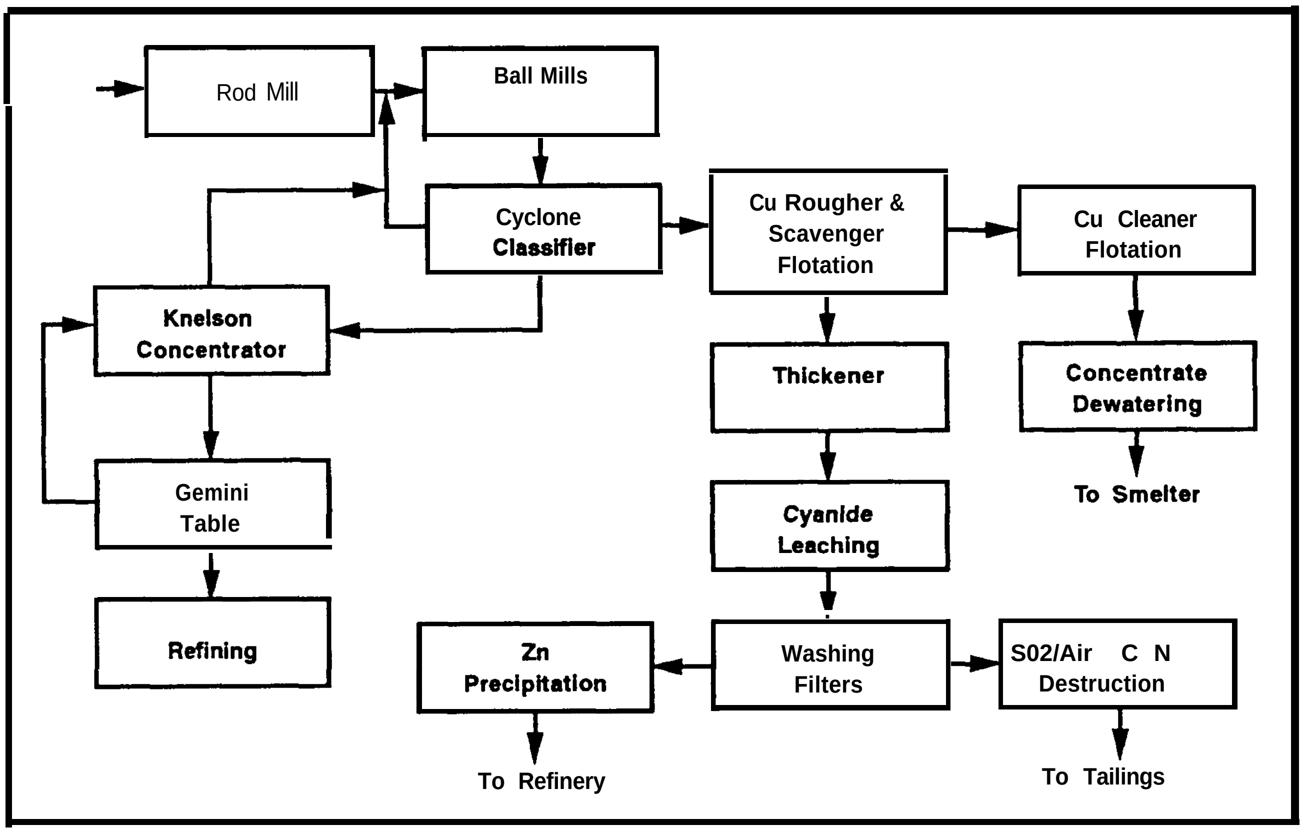
Despite efforts to keep the copper tailings grade as low as possible, cyanide consumptions in the 3-4 kg/t region were common and leach solutions containing up to 1000 ppm of copper were produced. These high copper and cyanide levels over stressed the SO_2/air circuit and resulted in problems controlling the contaminant level in the tailings discharge.

Overview of the Mill

Since it is not the intention of this paper to discuss the complete mill flowsheet, I will only outline the flowsheet in sufficient detail so the reader can understand where the process advances fit in.

Ore is received in 35 tonne rear dump highway trailers which are dumped directly into the crusher feed hopper. The crushing plant is a conventional 3 stage circuit consisting of a jaw and two cone crushers. The -12 mm (1/2 inch) ore is ground in an open circuit rod/closed circuit ball mill circuit to 80% passing $70\mu m$. A portion of the classifying cyclone underflow (notionally equivalent to the new feed rate) is fed to a gravity concentration circuit consisting of two 760 mm (30”) Knelson concentrators operating continuously and two Gemini tables for final upgrading. The gravity circuit tailings are partially dewatered using cyclones before they are returned to the ball mill circuit. The copper flotation circuit consists of a combination of Outokumpu high grade tank cells and conventional cells. The high grade tank cells are used in both roughing and cleaning duties while the conventional cells are used as scavenger and scavenger cleaners. The copper flotation tailing is thickened to 50% solids and cyanide leached for approximately 12 hours in open top agitated vessels. The leaching is followed by solid/liquid separation and pulp washing on drum filters and gold recovery by zinc precipitation. The filter residue is treated in an INCO SO_2/air circuit to remove copper and cyanide prior to discharge to a flooded tailings area. A flooded tailings deposition scheme *was* adopted to eliminate the onset of acidification due to sulphide oxidation.

Figure #3 - Block Flowsheet

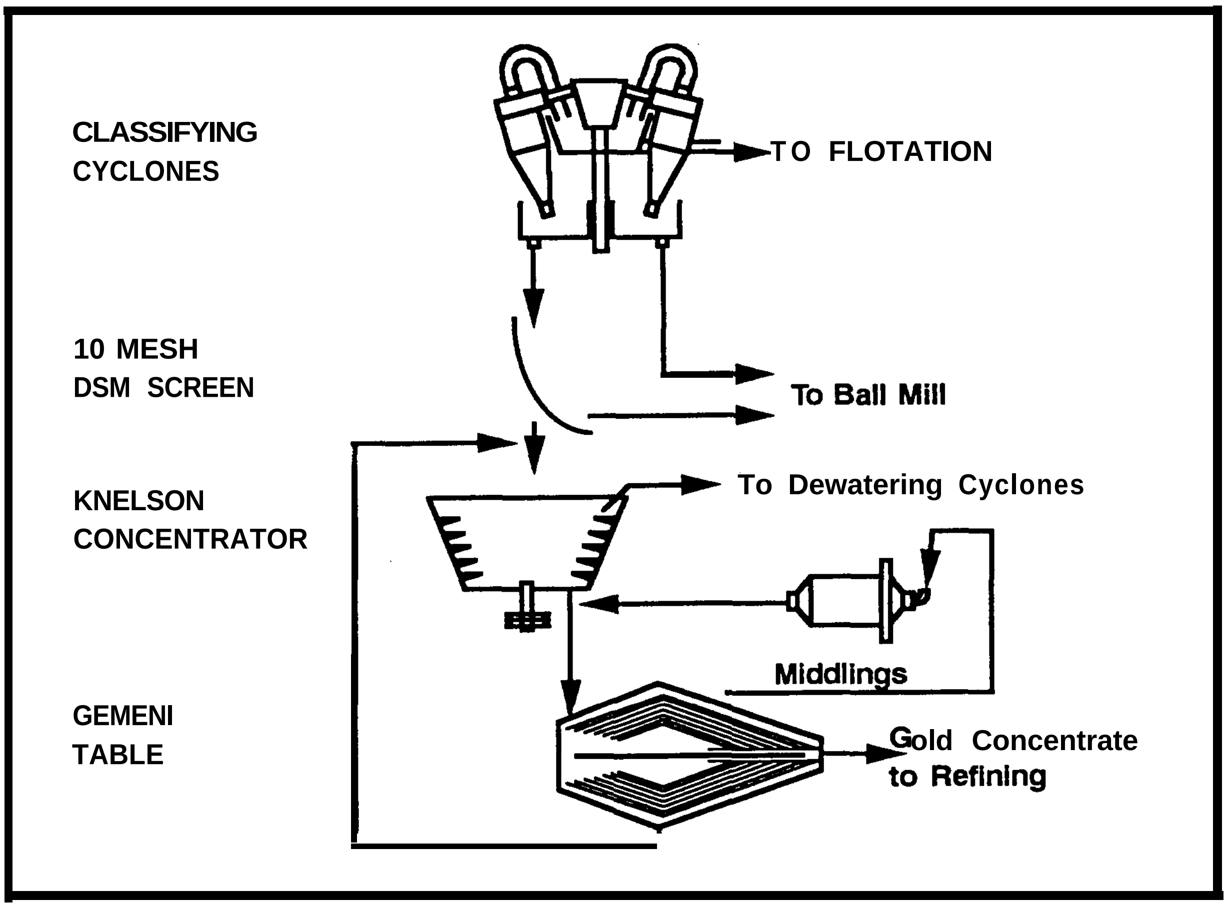


Application of the Knelson Concentrator

The original design of the gravity circuit included conventional jigs and tables. Three different types of jigs were tested including a Pan American style (Propulse), an end diaphragm style (Yuba Richards) and a Denver style. Of these three, only the Denver style jig was able to make the separation between the pyrite and the gold. With both the other two jigs, too much pyrite reported to the hutch and if not pulled hard enough, the hutch would fill with pyrite and sand out. When pulled hard enough to continue operating, the upgrading ratio was too low, 2 or 3:1, and this overloaded the shaking table which followed it. The Denver jig, with appropriate care was able to continue operating. The recovery from the jig was generally low (15-20 %) and the overall recovery even lower. Over the period of a year when the Denver jig was operating reasonably well, the total gold recovered by gravity amounted to 10.8 % of the total gold recovered which left some room for improvement. The recovery on the table was only in the range of 50-60% due to overloading and due to a middling consisting of coarse gold mixed with coarse pyrite that could not be separated on the Wilfley table. No regrind circuit was in place during this time to treat this middling product.

Preliminary tests carried out using a 3” Knelson concentrator indicated a potential to recover substantially more fine gold than the jig. This test was followed up with a short test with a 20” unit which was somewhat inconclusive because of the ore being milled at that time. When a 30” unit became available, it was installed to replace the jig. This unit soon proved its benefit, recovering approximately 40 % of the gold from the head. The other benefit was the high upgrading ratio (1000: 1) which was attainable in a single stage with the Knelson. The jig and table combination from the original installation gave an upgrading ratio of only 200-300 and therefore required more frequent final clean-up.

Figure #4 - Gravity Flowsheet



At the time of writing, we are still struggling with the low recovery or low capacity on the clean-up table. The Wilfley table has been replaced with a Gemini table and a small regrind mill to handle the table middlings will be installed. The latest results of the Gemini testing gave ≈90% gold recovery into a 85% gold concentrate. This result was obtained at a feed rate of only 300 kg/hr rather than the desired 1000 kg/hr. The other disadvantage of the Knelson that is currently in use is the need to manually clean up every two hours. Not only does this require manpower but it introduces a security problem. The operator has access to gold concentrate that runs several hundred oz/ton and therefore we must schedule security coverage when the Knelson is cleaned. This problem will be solved when the new automatic dumping Knelson’s are installed, as no operator is required for this clean up.

The circuit will be set up so that the Knelson concentrate will flow by gravity to a storage hopper with sufficient volume to store a full week of concentrate. On one or two days a week, the operator and security will access the security area and table the concentrate from the Knelson to attain a grade suitable for direct smelting.

Contrary to the jig operation, the Knelson requires very little fine tuning. Good, clean water is required for the backwash but with this provision, there is not much else to do. The jigs on the other hand required adjustment of the water and the stroke and required regular and frequent maintenance of the jig bed. Maintenance requirements for the Knelson have been minimal to date. The Knelson's centrifugal action with no reciprocating parts should result in less maintenance. The only problem encountered to date has been scaling of the water inlet holes in the back of the drum and this relates to the quality of the water. This problem is looked after currently by acid washing the drum, but we will look into antiscalants to prevent the problem.

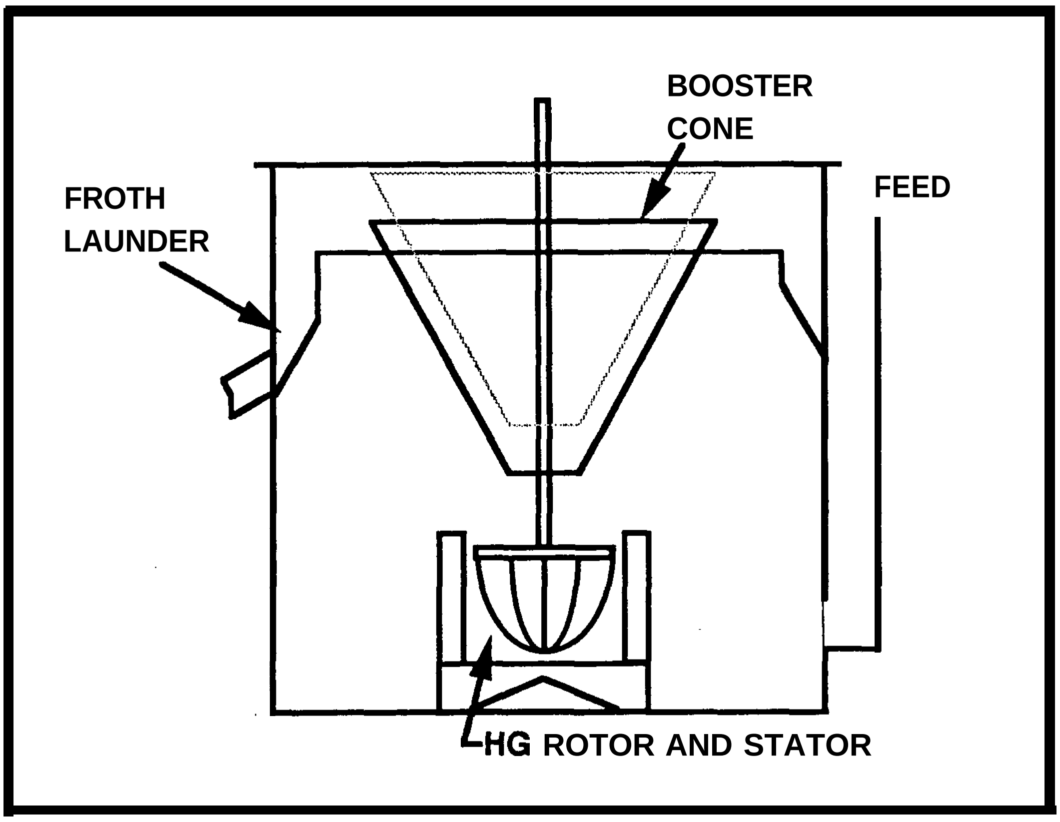
High Grade Tank Cells

The high grade flotation cell is a recent development by Outokumpu designed to produce a higher grade concentrate than produced by a conventional cell. The concept utilizes conventional Outokumpu mechanisms but in a deeper cell with froth crowding to develop a deeper froth layer and therefore better concentrate grade. Outokumpu reports that the high grade cell gives high selectivity, high concentration ratio and high recovery. The concept was only recently applied to tank cells and has proved a viable alternative.

The tank cell consists of a round tank which is slightly higher than its diameter. A conventional Outokumpu mechanism with the configuration and speed matched to the duty requirements is used to agitate the tank. The major difference is the froth crowding which produces a deep and stable froth. Froth crowding is accomplished in the tank cell by adding an inverted cone in the centre surrounding the shaft. The cone thus squeezes the froth by reducing the cross sectional area of the cell as it rises to the cell surface. By squeezing the froth as it rises, a certain amount of entrained water is removed and with it, an amount of hydraulically entrained gangue. Because of the deep and stable froth, froth washing can be utilized if required in this cell. The area required for froth discharge can be varied by the size and elevation of the cone. The opening required is governed by the froth carrying capability, much as the factor used for sizing flotation columns.

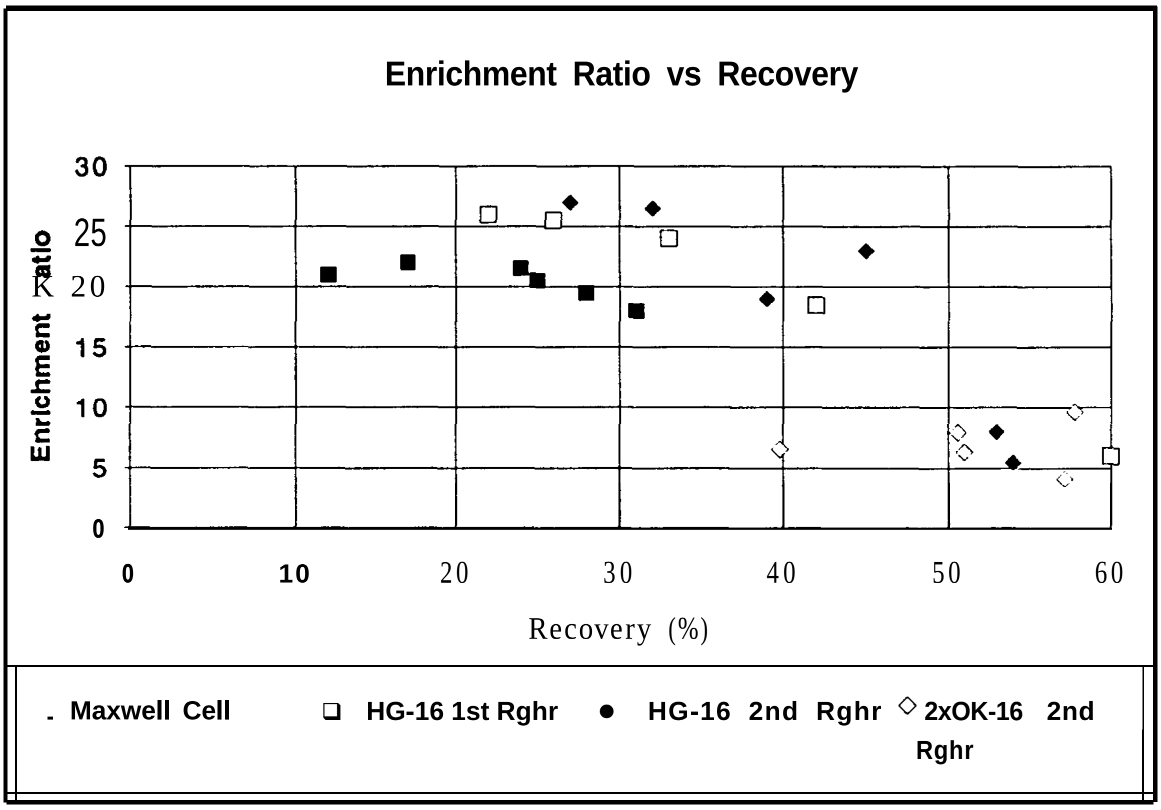
Lac first investigated the use of the high grade concept through on-site pilot plant work done by Minnovex. Minnovex tested columns, HG cells and conventional cells in various combinations and configurations and on various product streams. On the roughing and cleaning duties, the HG cell performed almost as well as the column. The HG cell was preferred because we felt our operators would more easily accept and adapt to the cell because it was physically similar to the cells they were used to. The lack of building height and the potential for scale build-up on the bubble generators in the high lime duty were considered to be problems specific to columns.

Figure #5 - Tank Cell Configuration



The Minnovex pilot plant and plant audit showed that we were short of flotation capacity with the head grades we were seeing at the time. The tank cell was a means to install additional rougher flotation capacity within the space available. Outokumpu agreed to sell us a mechanism and to provide the engineering to retrofit an existing 3 m (10 ft) diameter tank. Thus, we went ahead with the installation to end up with a 20 m^3 (Outokumpu refers to it as an HG-16 Tank cell) tank cell in the rougher flotation circuit. When started, this cell followed a Maxwell MX-12 (38 m^3) tank cell which had also been added. It soon became apparent that the combination was not compatible as the Maxwell cell required far more frother to work at all and this extra frother resulted in excessive froth in the HG tank cell. As a result, the Maxwell cell was only used when the head grades were particularly high and the balance of the time, the HG cell operated as the first rougher. This cell produced final concentrate grade product which therefore relieved the load on the 1st cleaners. The attached figure #6 showing enrichment ratio vs recovery indicates the performance of the HG-16 vs the Maxwell MX-12 and the conventional OK-16's. One must be careful in interpreting this data because the sample periods represent different time frames and different mineralogical mixtures. We believe however that the trend is valid and the HG tank cell has given us an improvement.

Figure #6 - HG-16 Performance

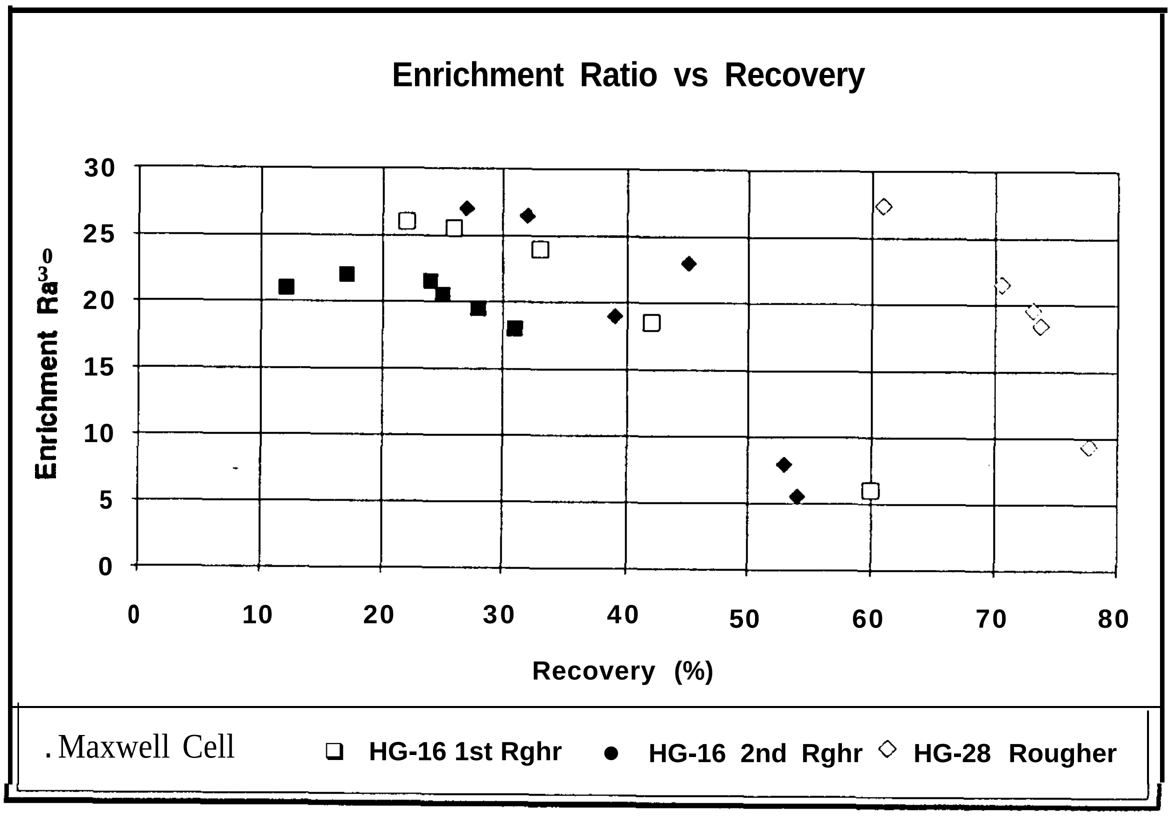


As the mine tonnage increased the mill had to follow and we once again found ourselves short of flotation time. We decided to convert the Maxwell cell to an HG tank cell by replacing the mechanism and adding the tank internals. Outokumpu has designated this cell an OK-28 even though the original tank volume was 34 m^3. The attached figure 7, similar to figure 6, has several results from the HG-28 added. The data for the HG-28 indicates that we have shifted to a new grade recovery curve but it should be remembered that we are looking at unit recoveries only, not global circuit recoveries. During this sampling period we were dealing with considerably lower head grades and a larger proportion of chalcopyrite than previously. However, we don't think there is much doubt that this tank cell in the HG configuration far out preforms the tank prior to conversion.

In terms of operation, the tank cells operate very smoothly even when we have had level control problems due to discharge control valve limitations. The deep froth layer appears to provide a real advantage in stabilizing the operation. Outokumpu has been surprised by the very low air consumption and therefore somewhat higher power consumption.

We are sufficiently convinced by this cell to commit to an additional 3-50 m^3 rougher and scavenger tank cells and 6-5 m^3 cleaner tank cells.

Figure #7 - HG-28 Performance



Improved Leach Performance

Despite best efforts in flotation, some copper still goes to the leach circuit resulting in high cyanide consumption (3-4 kg/t) and highly contaminated solutions (700-1000 ppm Cu). The costs of cyanide and lime were so high that we began questioning the economics of continuing the cyanidation operation.

We addressed the problem from three prospectives as follows:

-a cyanide regeneration circuit to recover copper and recycle cyanide - AVR or other
-improve the gold lotation to the point where cyanidation is no longer economic
-reduce the leach operating cost through reduced consumption of cyanide and SO,.

During 1990 and 1991, we did research and development on various processes to recover cyanide, generally relating to acidification and volatilization of the cyanide and recovery of the copper by precipitation. Sufficient process development was completed to evaluate the economics of this process. This evaluation showed that reducing the cyanide consumption by

0.5 to 0.75 kg/t made the AVR type process uneconomic. We therefore put additional effort into reducing the cyanide consumption rather than committing to a high capital cost cyanide recovery plant.

The first part of this study was to understand the variables that influence the cyanidation and the chemical and physical limitations. These are defined as follows:

- -ore characteristics and associated flotation performance
- -cyanide consumed as CNS (≈1 kg/t) due to high sulphide content
- -rapid consumption of cyanide (50% in initial 30 minutes) and difficult control
- -cyanide consumption is proportional to:
  - -residence time
  - -cyanide concentration in the leach solution
- -gold dissolution kinetics are proportional to:
  - -oxygen concentration
  - -minimum cyanide concentration to prevent re-precipitation (0.65-0.75 kg/t)

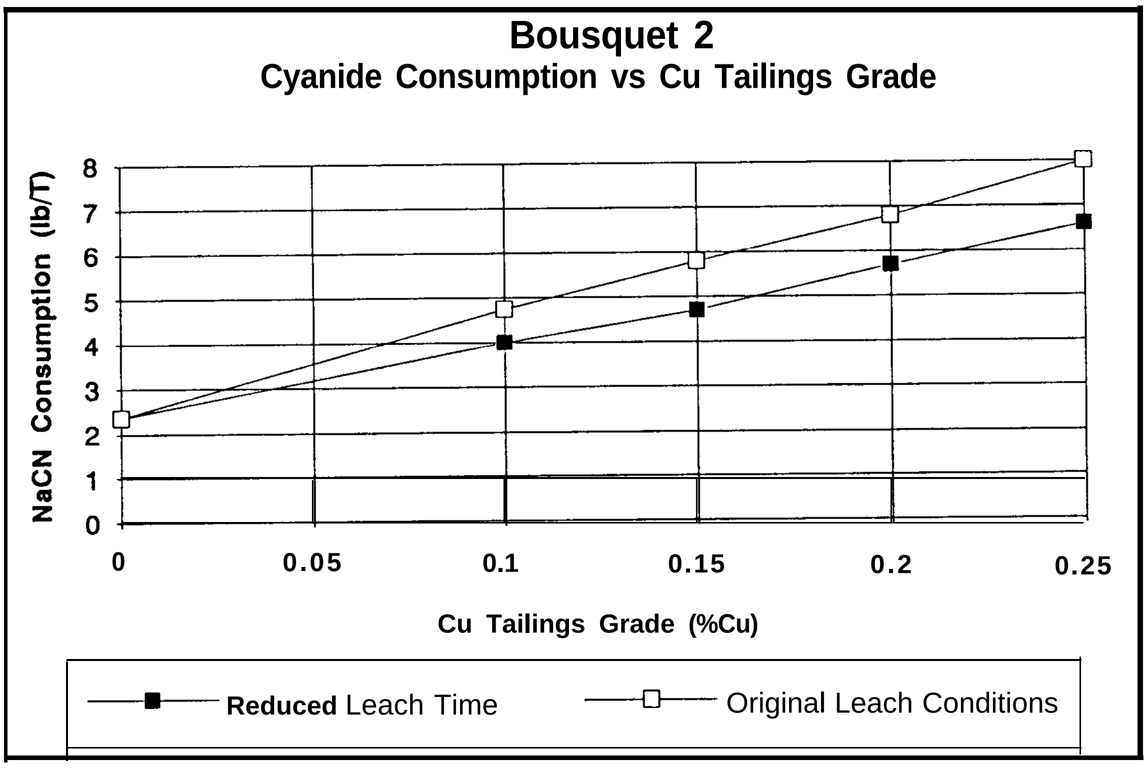
While we are not in control of the ore characteristics, we did have the ability to improve the flotation as noted above. The cyanide consumption as CNS has been addressed by pre-aerating the slurry in one pachucca prior to adding cyanide. However, controlling the cyanide consumption and improving the leach kinetics appeared to be the primary route to optimize the circuit.

When the mill was used for processing La Mine Doyon ore, the leach retention time had been increased to a nominal 72 hours due to the slow leaching of that ore. When we switched to processing Bousquet 2 ore, we realized that we did not require as much retention time and consequently reduced the leach time to ≈48 hours. This was done by eliminating 3 out of 9 pachuccas and 7 agitated leach tanks. During the optimization process, the lab testwork indicated that leaching could be completed in 4-6 hours with the right conditions. The first step in this direction was taken by eliminating the balance of the pachuccas and doing all the leaching in 6 agitated tanks with a combined retention time of about 10 hours. The result of this was a reduction of approximately 0.5 kg/t of cyanide as demonstrated in figure #8 without any increase in the residue grade. There was some initial indication that the cyanide residue grade actually decreased when the leach time was reduced but we cannot confirm this at present.

The following actions are planned to continue this development:

-On-stream cyanide control - As a result of the rapid consumption of cyanide at the beginning and the short retention time, continuous control of the cyanide addition is considered necessary. A Cyanostat system has been ordered and will be installed to measure and eventually control the cyanide addition to 3 or 4 points in the circuit.

Figure #8 - Reduced Cyanide Consumption



-Improve the leach kinetics through the use of oxygen and improved mixing - We have successfully been using liquid oxygen at several of our operations for the past few years. Oxygen improves the leach kinetics and is generally accompanied by a reduction in the cyanide consumption and sometimes an improvement in the gold extraction. It would appear that these benefits are also available at Est Malartic, but that requires replacement of the air lift agitator mechanisms before oxygen use can be started. We have also become convinced that more intense mixing during cyanidation could be a benefit. The gold industry has generally been led down the path of high efficiency, low power input agitators in progressively larger tanks. We are not convinced that this is doing us much good and are considering reversing this trend to go to small, high intensity mixing conditions with much shorter retention times.

-Reduce leach time to the minimum - The laboratory indicated 4-6 hours and thus we will target for this range with some flexibility to add or remove leach time. This short leach time, of course, requires that we maintain optimum leach conditions as mentioned above. Also, with such short retention time, automatic cyanide control is imperative. No longer can we afford to manually titrate the cyanide strength ever 2 to 4 hours and expect to maintain it at a consistent level when the retention time is 4-6 hours.

Conclusion

The treatment of this high copper/gold ore from Lac Minerals Bousquet 2 mine has presented a number of challenges. The problems and solutions discussed here are but a few of those faced at this operation. While these advances are not revolutionary, they have provided some real technical and economic benefits to the operation of the Est Malartic mill.

Acknowledgements

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