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HEAP LEACHING WITH FINE GRINDING AND AGGLOMERATING AT POTOSI - BOLIVIA

Presented at GoldTECH 4 – Reno, Nevada, USA – September1990

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ABSTRACT

In 1984, work was started to determine the viability of heap leaching silver-containing waste dumps from Cerro Rico near Potosi, Bolivia, at an elevation of 4300 meters (14,100 ft.). The result today is an unusual heap leach utilizing fine crushing, grinding, and agglomeration to process 1000 tonnes/day of ore containing 200 g/tonne silver. The unique metallurgical, economic, and political factors discussed led to the decision to build a heap leach rather than a conventional agitation leach plant. Capital and operating costs are given along with some of the uncommon process design and equipment problems encountered while developing the operation.

INTRODUCTION

In 1984, Compañia Minera del Sur S.A. (Comsur) approached Kappes, Cassiday & Associates (KCA) regarding an idea to leach waste dumps from Cerro Rico de Potosi in Bolivia.

Cerro Rico is considered to be the world's largest known silver deposit. Records show that mining began in 1545 and that more than 30,000 tonnes of silver have been produced from the mountain. During all of these years of production, tremendous mine waste dumps have been accumulating on the slopes of the mountain. Comsur obtained control of approximately one million tonnes of these dumps.

KCA initiated a metallurgical testing program and then designed a heap leach to recover 75% of the 200 grams of silver per tonne contained in the waste dumps.

Comsur set up a subsidiary company to build and operate the heap leach, Compañia Minera Concepcion (Comco). Beginning in 1987, Comco, with assistance from KCA, built a 400 tonne/day heap leach operation approximately 6 km south of Cerro Rico de Potosi at an elevation of 4300 m (14,100 ft). This operation was improved and expanded over the next two years and is presently treating 1000 tonnes/day. The operation is one of the most complicated heap leach operations in the world due to the unique metallurgical requirements necessary to attain 75% recovery. Protecting the environment in the heap leach area was an important consideration. With this in mind, the leaching facilities were designed and built to the same standards as operations in the United States.

METALURGICAL TESTING

Three test programs were completed on samples of silver-bearing material from the Cerro Rico waste dumps. The program included 25 column leach tests and 15 agitated leach tests. Silver was the primary mineral of interest. The minor amount of tin in the material was not considered economically recoverable at the time, and at best, only trace amounts of gold were present.

The samples tested varied in ore size and silver content, with head assays of the samples ranging from 153 to 517 grams silver/tonne. Tin content averaged 0.5%. The purpose of the test programs was to determine the amenability of the materials to cyanide processing for silver recovery, especially regarding low-cost production methods such as heap or vat leaching. Tin recovery was not tested, although future tin recovery was considered in the final selection of optimum grind size.

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Table 1. Silver Recovery at Various Crush Sizes		
TEST TYPE	CRUSH SIZE	RECOVERY
Column	75 mm	15%
Column	25 mm	44%
Column	6 mm	64%
Column	2.36 mm	66%
Column	600 micron	84%
Column	150 micron	87%
Bottle Roll	600 micron	80%
Bottle Roll	150 micron	85%
Bottle Roll	53 micron	89%

The testing methods included cyanide centrifuge tube tests on pulverized material, bottle roll or agitated beaker tests, drip column leach tests, and flooded column (vat-type) leach tests.

The recovery of silver from the samples is highly dependent on the crushed size of the material tested. In column testing, a significant increase in recovery occurs as crush size is decreased from 75 mm to 25 mm to 6 mm. The effect continues at finer crushed size down to 150 micron (100 mesh). Bottle roll test results also indicate the dependence of silver recovery on crushed size. Table 1 shows recoveries at various crush sizes for column and bottle roll tests.

The Potosi ores leached very slowly. At the finer crush sizes, column test recoveries of 50 to 70 percent of contained silver were attained within 10 to 20 days, but attaining recovery levels of 75 to 80 percent required significant additional leaching time (up to 120 days total in laboratory tests).

Cyanide consumptions were projected to be 1.5 to 2.0 kg/tonne of ore if the leach solutions were maintained at 1.0 gpl NaCN. Cement used for agglomeration provided pH control.

Agglomeration of the Potosi dump material was required to maintain proper percolation characteristics in a vat or heap leach situation. A series of column tests were set up that optimized cement requirements at 10 kg/tonne of ore for material crushed to 600 micron (28 mesh).

The metallurgical basis for the project design included grinding to 600 micron (28 mesh) and leaching with 1.0 gpl NaCN solution for at least 120 days. Under these conditions, an ore of average Potosi dump grade (200 g/tonne silver) yields 80% silver recovery in laboratory tests and a projected 75% recovery in a production heap.

PROJECT DESIGN

Options

Reducing the ore to minus 600 micron (28 mesh) was considered too fine for conventional crushing, and therefore some type of grinding would be required. Several process options were considered during the early stages of the project development. These options were:

1. Crush, wet grind, agitation leach
2. Crush, wet grind, dewater/dry, agglomerate, heap leach
3. Crush, dry grind, agglomerate, heap leach

Option 1 – Agitated Leaching

Conventional agitated leaching would have allowed finer grinding, but only slightly increased recovery. However, the capital cost of the plant would be significantly higher and operating costs would be higher.

Another factor considered was the political and economic stability of Bolivia. In the year the project was conceived, inflation in Bolivia averaged 16,000%. As agitated leach plant would take longer to design and build and have a longer payback period. This would have increased the risk of the project.

After reviewing the economics and other risk factors, it became apparent that a heap leach would be a better alternative if the technical problems could be overcome.

Option 2 – Heap Leach With Wet Grinding

Wet grinding is the most logical method to reduce the dump material from the crushing plant to 600 micron (28 mesh) or finer for leaching. There is, however, a very significant problem in trying to dewater or dry the resulting slurry to 12 to 15 percent moisture to allow for proper agglomeration of the material prior to stacking.

It was decided that the advantages of wet grinding were not worth the added cost and problems associated with trying to dewater or dry the ground material.

Option 3 – Heap Leach with Dry Grinding

After reviewing the alternatives, this option was chosen as the most logical choice for this particular project. Dry grinding is relatively common in other industries, but not in metals mining. However, it was believed that the problems were not insurmountable and that the costs associated with dry grinding were not excessive.

ORIGINAL PLANT DESCRIPTION

The initial plant design consisted of the following basic elements:

1. Production rate of 500 tonnes/day
2. Conventional three stage closed circuit crushing to minus 9.5 mm (3/8 inch)
3. Dry grinding in closed circuit to minus 600 microns (28 mesh)
4. Agglomeration with 10 kg/tonne cement
5. Conveyor stacking to a height of 6 m
6. Three stage, countercurrent leaching with 1 gpl NaCN solution at an application rate of 0.30 l/min-m^2 (0.0073 gpm/ft^2
7. Zinc precipitation for recovery of silver from solution

Mining & Transport

The ore from various dumps on Cerro Rico was loaded into 10, 12, and 25 tonne capacity trucks and hauled to the heap leach facility. Comsur operates a single 25 tonne Caterpillar D25C articulated haul truck, while the smaller trucks are operated by contractors. It was not possible to use larger trucks due to the narrow roads on the mountain, and the need to use public roads for part of the haul.

Crushing

The original crushing plant was a standard three-stage circuit using a jaw crusher, two cone crushers (one standard and one shorthead), and a double deck screen. Final crusher product is discharged to a 500 tonne fine ore shortage bin ahead of the grinding circuit.

Dry Grinding

The dry grinding portion of the design was based on a Bond grindability index of 11.3 kw/hr. A used 2.4 meter diameter by 2.9 meter long (8.0 ft x 9.5 ft) Denver ball mill with a grate discharge was available from another Comsur operation. The mill was designed to work with a 78% critical speed and a 42.5% ball charge. Also, the design was based on having the moisture content of the feed at one percent or less.

The existing Denver ball mill was modified for dry grinding service by the addition of a water cooling system for the bearings and a modified grate discharge system.

Three double-deck 1.5m by 3 m (5.0 ft x 10.0) Rotex screens were used in closed circuit with the ball mill.

An 8500 m^3 per hour (5000 cfm) baghouse-type dust collection system was specified for the grinding plant.

Agglomeration and Stacking

The minus 600 micron (28 mesh) material from the grinding plant is fed through a surge bin with a variable speed feeder to a 2 meter diameter by 7 meter long agglomerating drum. Cement is fed from a silo onto the drum feed belt.

Barren solution with 1 gpl NaCN is used to agglomerate the material to a final moisture content of 15 to 17 percent.

The agglomerated material is conveyed by a series of fixed and mobile conveyors to a 24 meter radial stacker with a 4.5 meter extendable tip conveyor. This system builds heap segments 47 meters wide by 6 meters high.

Leaching

The system is designed as a continuously expanding permanent heap. The leach pad is divided into modules with a capacity of 67500 tonnes each. This allows the separation of solutions from different parts of the heap. The system is designed for three 45 day countercurrent leach cycles followed by a fresh water wash cycle.

The leach pad is constructed with a 0.75 mm (30 mil) PVC pad liner installed on top of a compacted soil base. 50 mm perforated drain pipe is installed on 6 meter centers to insure good drainage of solution from the heap. A 0.5 meter layer of crushed, screened ore is used as a protective pad cover and drainage layer on top of the PVC liner in the areas where the conveyor stacking system operates.

The solution application rate is 0.30 l/min-m^2 (0.0073 gpm/ft^2). Solution is applied with no. 11 Senninger Wobblers on a 6 x 9 meter spacing. This results in a flowrate of 90 m^3/hr in each of the three leach cycle.

The heaps are covered with shadecloth to protect the fine agglomerates from degradation due to the impact of sprinkler droplets.

Solution Recovery

Silver is recovered from the leach solutions in a conventional zinc precipitation plant. The plant was purchased used in the United States. It consists of two 46 m^2 (500 ft^2) U.S. Filter Auto-Jet type pressure filters for clarification, a 1.5 m diameter by 4.6 m high deaeration tower, and three 0.9 m by 0.9 m plate and frame filters for precipitation. Each plate and frame filter has approximately 37 m^2 (400 ft^2) of filtering area. The recovery plant has a maximum capacity of 135 m^3/hr (600 gpm).

Silver precipitate from the filter presses is mixed with fluxes and smelted in a potless tilting furnace to produce a silver doré product.

Construction

All of the construction was done by Comco. KCA supervised the installation of the agglomeration and stacking equipment along with the leach pad, ponds, solution collection system, recovery plant, and solution distribution system.

On-site construction was started in June 1987 and start-up began in June 1988. Construction was delayed to some extent due to delays in obtaining some equipment and also due to a later than normal rain season.

Start-Up

Crushing, grinding, agglomerating, and stacking began in June 1988. The first sprinklers were started late in July and the recovery plant was started in early August.

The principle problems encountered during startup of the operation were related to the dry grinding circuit. Initially, there were problems in getting the feed into the ball mill. This was solved by redesigning the feed chutes and adding a compressed air jet to the chute to help move material into the mill.

As the feed to the ball mill was increased, the discharges of material from the mill became the limiting factor in throughput. It is believed that this problem is due to the sizing and spacing of the grate openings, as well as some material buildup on the grate openings due to the moisture in the ore.

It was also discovered that the moisture content of the ore was greater than that indicated in the original sampling and testing of the waste dumps. Original testing indicated one percent moisture or less, while the average during startup was actually two to three percent. This moisture caused a number of problems throughout the grinding plant. First, the heat generated in the mill drove the moisture out of the ore and into the air, saturating the air. When this warm, moisture saturated air was pulled into the dust collection ducting and into the bucket elevator and Rotex screens, it cooled and caused moisture to condense on the piping, screens, etc. This combined with the ore to form mud in these areas and caused significant problems with the plugging of ventilation ducts and the dust collection filter. This problem was partially cured by separating the ball mill ventilation from the rest of the dust collection circuit and increasing the airflow through the mill.

Original Plant Performance

The problems in the dry grinding circuited limited average plant throughput to approximately 400 tonnes/day. Due to the reduced ball mill throughput, the average size distribution was slightly finer than designed and the recirculating load was only about 10%. Because of the minimal recirculating load, the grinding circuit was changed to open circuit and the Rotex screens were used as scalping screens ahead of the ball mill. This change eliminated the bucket elevator which had caused some problems due to spillage and high maintenance requirements. The operators also made the decision to change from 600 micron (28 mesh) to 850 micron (20 mesh) screens on the Rotex scalping screens in order to help throughput.

The leaching was somewhat slower than originally predicted. This was at least partially due to less leach solution being applied to the heaps in a given time period for each tonne of ore stacked. The high elevation and resulting decrease in oxygen content in solution had no apparent influence on the leach rate. It is interesting to note that when the recovery results are plotted as percent recovery versus tonnes solution per tonne ore, the curve matches very closely with the laboratory testing completed in KCA's laboratory in Sparks, Nevada.

Overall, the heap permeability is quite good; although, even with shadecloth protecting the top of the heap, after several months of leaching some agglomerate degradation due to the impact of sprinkler droplets was noticed. Because of this, after each leach cycle for a given heap module, the shadecloth is removed and the top 0.2 m of the heap is manually turned over to break up the layer of fines that builds up on the surface.

Later in the operation, a test was made using ore crushed and screened to 2.36 mm x 19 mm as a protective cover on top of the heap. This worked much better than the shadecloth and no turning of the top layer of the heap was required.

The zinc precipitation plant operated very efficiently from the beginning of operations. Pregnant solution grades were in the range of 30 to 50 ppm silver. The ratio of zinc to silver precipitated was approximately 0.75 to 1. The precipitate grade was over 90% silver and the doré was over 98% silver.

Cyanide consumption is approximately 0.35 kg/tonne, which is considerably less than the original projections of 1.5 to 2.0 kg/tonne.

Plant Expansion

In November, 1988, a proposal was made by KCA to double the plant capacity by the installation of a fine crushing circuit. The agglomerating, leaching and recovery sections of the plant all had sufficient excess capacity to handle up to 1000 tonnes/day with only minor changes.

The basic concept was to crush the ore to 100 percent passing 2.36 to 1.70 mm (8 to 10 mesh) using a Barmac impact crusher. It was reasoned that crushing to this fine size would result in a final crusher product which was at least 50% minus 850 micron (20 mesh). This allows 50% of the ore to be scalped prior to the ball mill and sent directly to the agglomeration surge bin. This would approximately double the throughput with the addition of the new crushing equipment and one extra Rotex screen in the grinding section.

Samples of the ore were tested by REMco, the U.S. representative for Barmac. These tests indicated that the ore could be successfully crushed using the Barmac and that closed circuit crushing to 2.36 mm (8 mesh) would give the desired product size distribution.

Starting in January, 1989, KCA designed the new crushing circuit and also designed a complete dust collection system for the new crushing circuit. The new system consists of 2-stage, closed-circuit crushing to minus 19 mm using an Allis-Chalmers Model 60120 Type Rt jaw crusher and an Allis-Chalmers 10 – 500 Hydrocone, with the existing 2 m x 5 m double deck screen.

The minus 19 mm product from this circuit then goes to an Allis-Chalmers 2.4 m x 6 m double deck Model ST Ripl-Flo screen. The minus 2.36 mm (8 mesh) product from the screen is directed to the fine ore bin ahead of the grinding plant, while the plus 2.36 mm material goes to a Barmac Mark II Rotopactor. The Barmac product is returned to the 2.4 m x 6.0 m screen.

A portable crushing plant was used during installation of the new crushing equipment to minimize production disruptions. The new crushing plant started operating in August, 1989.

Performance

The average amount of ore stacked increased from approximately 400 tonnes/day to 830 tonnes/ day after startup of the new crushing circuit. The crusher product was 50% minus 850 micron (20 mesh) as expected.

The Barmac crusher is performing quite well in providing the required size reduction and throughput; however, the ore is highly abrasive and some of the Barmac wear parts require frequent replacement. This situation is being improved through some fine tuning of the Barmac and investigation of some alternative wear materials. Even with the extra Barmac wear and the fine crushing, total crushing costs are only US$ 0.88/tonne. Overall, the operators are quite happy with the new crushing system.

Table 2. Capital Costs

ORIGINAL PROJECT (400 TONNES/DAY)		US$ 1000's
Haul Road and Equipment		500
Crushing Plant		650
Grinding Plant		800
Agglomerating and Stacking Equipment		460
Leach Systems - Pads/Ponds/Piping		620
Zinc Precipitation Recovery Plant		450
Power and Water Systems		200
Office, Laboratory, Wharehouse		270
Vehicles		90
Engineering and Construction Management		510
Working Capital		400
	US$	**4950**
EXPANSIONS TO 1000 TONNES/DAY		**US$ 1000's**
New Jaw and Cone Crushers		550
Barmac System and Screen		300
Grinding Modifications, Rotex Screener		110
Dust Collection System		150
Engineering and Construction Management		100
	US$	**1210**
TOTAL CAPITAL COST	**US$**	**6160**

Wet Grinding

After the successful startup of the new crushing circuit, Comco investigated the possibility of changing to a wet grind at a relatively high slurry density. The basic idea is that the 50% of the total crusher output, which is milled, could be milled at a high percent solids and then recombined with the other dry 50% of the crusher product in the agglomerating drum to result in a moisture content compatible with good agglomeration.

Laboratory testing indicated that wet grinding could be successfully accomplished at slurry densities as high as 78% solids. This would result in a moisture content of approximately 13% after the slurry is combined with the dry, minus 850 micron (20 mesh) crusher product.

The largest potential problem with the change to wet grinding was the need to maintain very close control of the slurry density. The benefits are increased throughput and reduced dust control problems.

In April, 1990, the milling circuit was changed to wet grinding. Barren solution with 1 gpl NaCN was used in the mill. Slurry density was controlled manually, and some problems had been encountered while trying to control the moisture content in the agglomerator. In general, however, the system is working well and has allowed the plant throughput to be increased to an average 1000 tonnes/day.

A study is[was] underway investigating the possibility of adding an automatic slurry density control system to the milling circuit.

CAPITAL & OPERATING COSTS

Table 2 shows the capital costs for the project, including the expansion. Much of the project funding was obtained through a loan from the IFC. Financing delayed the project development by about one year.

Table 3 shows the operating costs at 1000 tonnes per day. The lower cyanide consumption has resulted in significantly lower operating costs than originally projected.

CONCLUSION

The Comco operation demonstrates that, in the proper circumstance, a heap leach with fine crushing and grinding can be a successful alternative to more conventional agitated leaching. Comco is possibly going to the extreme by including the grinding, however, the general trend in heap leaching is to crush finer for better recovery. Newer crushing technology makes fine crushing possible and many times the increased recovery justifies the additional crushing costs.

Heap leaching technology has similarly progressed to the point that, with proper agglomeration, the risks of having an impermeable heap are greatly reduced when dealing with finely crushed and clayey ores.

Comco has been successful because of the attention paid to important details during testing, design, construction, and particularly operation. Also, they have been quick to react when unexpected problems arose and have been willing to try new ideas to improve the operation.

ACKNOWLEDGEMENTS

The authors wish to thank the management of Compania Minera del Sur and Compania Minera Concepcion for permission to print this paper.

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Table 3. Operating Costs

OPERATING COSTS (1000 TONNES/DAY)		US$/TONNE
Mining and Transport		$ 1.56
Crushing		$ 0.90
Grinding		$ 0.89
Agglomerating/Stacking		$ 1.15
Leaching (Initial Pads Were Capitalized)		$ 1.49
Recovery and Refinery		$ 0.58
Site G & A		$ 0.63
	US	**$ 7.20**