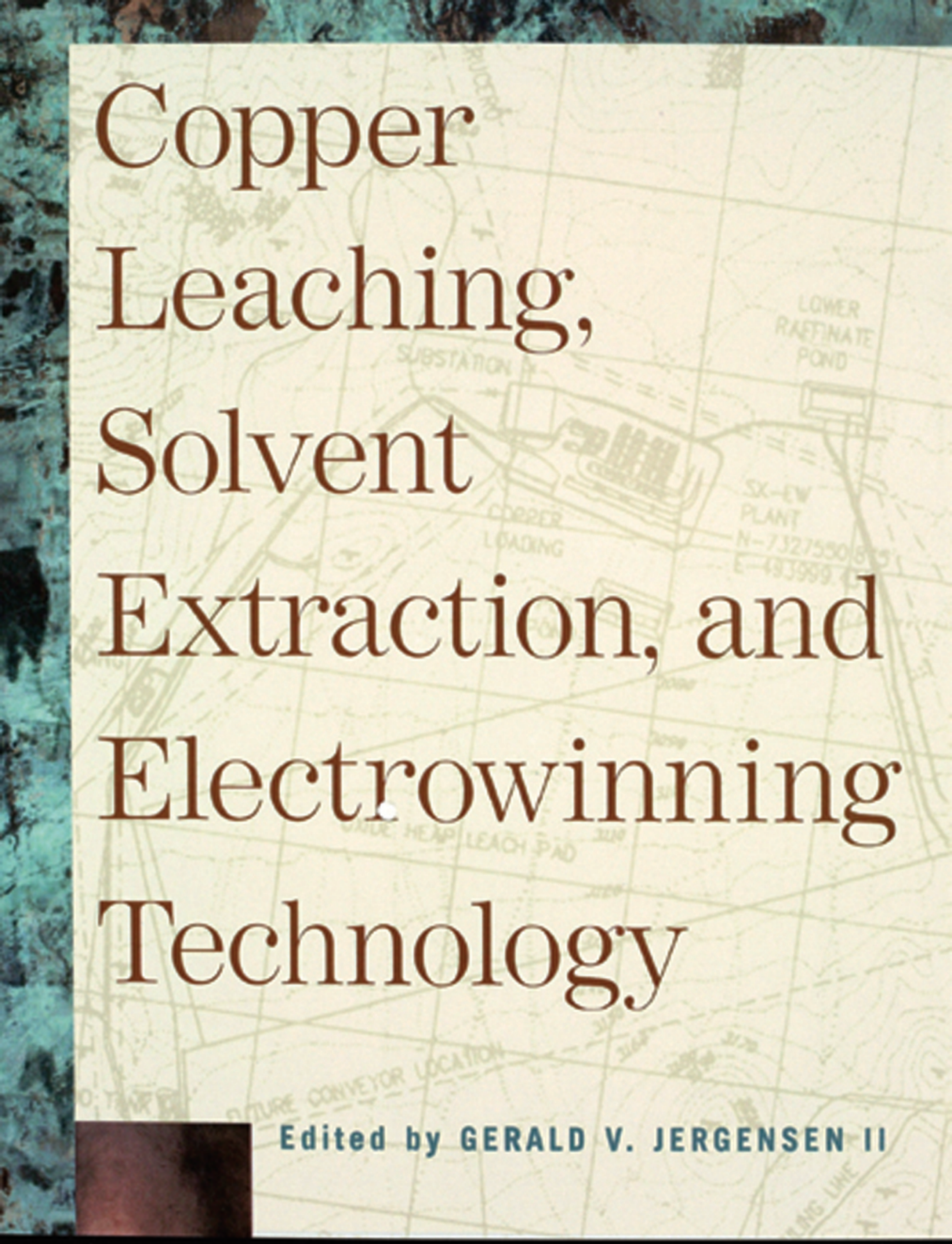

Copper Leaching, Solvent Extraction, and Electrowinning Technology

Edited by GERALD V. JERGENSEN II

SME

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Foreword

This volume is published by the Society for Mining, Metallurgy, and Exploration (SME) in recognition of the growing role of solvent extraction and electrowinning technology in the world copper business. Thirty years ago, the first commercial SX-EW operation was successfully commissioned. Today, this remarkable hydrometallurgical achievement is well established and fills an important niche in our technical ability to extract copper in an efficient and cost-effective way.

The primary objective of this text is to document the present status of the SX-EW business. Although it provides a broad scope of expertise and subject matter, the text is merely an introduction to the business. However, the information does represent a substantial body of historical, scientific, engineering, and commercial information regarding the growth and application of the technology.

Each author and session chairman must be congratulated for committing a significant portion of their crowded agendas to document and share their knowledge and experience. They represent the highly specialized personnel involved in this hydrometallurgical business. Contributors and sponsors include private consultants and engineers, professional consulting firms, engineering and construction companies, chemical suppliers, operators, teachers, and universities.

This book is also dedicated to students of the business, from the campus to the boardroom. We hope that this information will provide a unique foundation for the professional growth of every reader.

Finally, we wish to express our appreciation to SME especially the meetings and publications staff for their assistance and advice about the procedures for managing the production of this book. Also, we gratefully acknowledge the support of SME in bringing this work into print.

Gerald V. Jergensen, II
Joseph M. Keane
Valmer H. Ness

SECTION 1

The Business and Technology of SX-EW

Co-Chairs

Neal Rigby	**Roger Nendick**
Steffen, Robertson, and Kirsten	**Fluor Daniel Inc.**

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Dynamic Modeling of Heap Leaching—Design and Operation Criteria Impact on Production

J. Bartlett* **and F.L. Bazzanella†**

METSIM is being used to model numerous processes in both steady-state and dynamic configuration. One application of dynamic simulation involves the detailed modeling of heap leach operations. One of the more difficult aspects of designing a heap leach operation is developing the design criteria so that all of the various components are coordinated and production forecasts will be met. This presentation provides a comparison of how variations to certain criterion can impact solution flow, solution tenor, and production.

INTRODUCTION

Developing a dynamic model of a heap leach operation requires the definition of certain inputs including:

- Proposed ore delivery (mine plan)
- Extraction curve for various ore types (column test data)
- Total area available for heap construction (site topography)
- Desired production level (resources)

The criteria that must be developed and defined to create a detailed model include:

- Heap construction schedule
- Specific irrigation rate
- Area under leach, PLS volume
- Leaching cycle, rest, leaching, pulse leaching, etc.
- Solution stacking via recirculation of intermediate solution
- Lift height and number of lifts

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* Proware, Tucson, Ariz.

† KD Engineering, Tucson, Ariz.

In many situations, these criteria are determined and a spreadsheet is developed that will calculate the various values for flow rate, solution tenor, and metal production based on average or typical values. If the geometry of the heap is irregular, the heap has more than two or three lifts, the ore schedule varies with different types and grades of ore, or precipitation is significant, the production schedule is often irregular and overstated by the simplified spreadsheet model.

The example presented in this paper is a hypothetical copper heap leach where the assumed data is typical of actual operations but is held constant to avoid undue variation in the output. The example uses a rectangular pad with five lifts. A specific set of criteria is initially defined and the results are generated on a dynamic basis. Two criteria, the heap construction schedule and the leach cycle duration, are then modified to exhibit the change in flow and tenor of PLS solution and the production level of copper.

MODEL DEVELOPMENT

The model development for this example is described briefly in the following paragraphs and includes a summary of the data inputs.

Components

The components used for the model include:

Solid components	Quartz
	Copper metal
	Copper oxide/sulfide—fast dissolving
	Copper oxide/sulfide—slow dissolving
	Copper oxide/sulfide—insoluble
Liquid components	Water
	Copper sulfate
	Sulfuric acid
Gaseous component	Oxygen

The “fast” and “slow” dissolving components are discussed in the section regarding the extraction curves used in the model.

Flowsheet

A typical copper heap leach flowsheet was developed for this example. Ore from a specified source (i.e., mine, crusher, or stockpile) is placed on a heap at a defined rate. Leach solution is delivered, also at a defined rate, and pregnant leach solution (PLS) is produced and collected in a PLS pond. The PLS is then processed in an SX/EW plant where copper cathodes are produced. The raffinate from SX is returned to the leaching circuit.

Other streams, such as rain and evaporation, are not used in this model but can be an important factor and should be included in site-specific models. The dissolved copper inventory in the heap can provide important information relative to the production achieved and should be monitored. The inventory level is influenced by numerous criteria including the extraction curve, heap size, lift height, total heap height, and irrigation rate.

Unit Operations

Modeling of a heap or dump leach operation utilizes various unit operations. Additionally, the heap leach module requires several types of data to clearly define the pad loading and leaching schedule, the pad drainage plan, and the heap leaching cycle. The unit operations used for this model and the associated data are described in the following paragraphs.

Ore Unit Operation (ORE). The ORE unit operation characterizes the ore to be delivered to the heap. This can include any number of ore types, tonnages, grades, and feedrates. For this model, two ore types, each with a distinct extraction curve, are defined. The input for this unit operation includes:

- Total quantity of ore
- Mining rate in tonnes per day
- Operating days per week
- Starting and ending date, if necessary
- Moisture content
- Assays

For heap leaching, the assays include “fast,” “slow,” and “insoluble” fractions.

Heap Leach Unit Operation (HPL). The heap leach unit operation module is the prime unit operation used to simulate a complete heap leaching operation. The input data for this unit operation consist of five major types.

1. HPL defines the heap area, number and size of blocks, lift height, bulk density, final moisture, drainage rate, and drainage distribution method.
   - Area of the heap

     The area is used to determine the rainfall catch basin area.
   - Area of block
   - Lift height
   - Bulk density

     The dry ore capacity of each block is determined by the block area times the lift height times the bulk density.
   - Final moisture
   - Drainage factor

     The final moisture and drainage factor are used to determine the movement of liquor through the heap by assuming that a fraction of the excess moisture drains through each lift each day. For example:

If the moisture after spraying is	.12	12%	
the final moisture is	.08	8%	
the excess moisture is	.04	4%	
the drainage factor is	.75	75%	
Then the moisture leaving in one day	.03	3%	(.04 × .75)

- Drainage method

  The drainage from the heap may be allocated to 1, 2, or 3 output streams. If only one output stream is specified, all drainage reports to that stream. Otherwise the drainage may be allocated based upon flowrate, solution tenor, or spray cycle. These options simulate a daily monitored and controlled heap discharge.

2. HPLA defines the physical configuration of the heap by specifying the levels, block numbers, and fractional filling of the blocks. The sequence in which the data is tabulated determines the sequence in which the heap is constructed. HPLA consists of a large array of data, one line for each block.

   - Level

     The vertical description of the heap is established by specifying a level, usually the elevation, for each lift in the heap. The following table represents a cross section through a heap.

Heap leach vertical section

Level 25			103		
Level 20			103	104	
Level 15		102	103	104	105
Level 10	101	102	103	104	105
Level 5	101	102	103	104	105

     Each lift must be assigned a unique level and the block numbers for each lift/level must be identical to the ones below. All blocks on all levels need not be used depending on the heap configuration.

   - Block number

     A grid in plan view is used to input irregular shaped heaps. The grid must be rectangular and include the entire heap. Each block represents a fixed area, volume, and tonnage. Topographic maps of the proposed heap pad location with the overlying grid structure must be used to generate this table. The block numbers represent the rows and the columns in the grid. Following is an example of the grid layout and numbering system.

     where: R is the grid row
     and C is the grid column

Heap leach plan view grid pattern

R01	R02	R03	R04	—	R0C
—	—	—	—	—	—
301	302	303	304	—	30C
201	202	203	204	—	20C
101	102	103	104	—	10C

- Horizontal loading factor

  Partial filling of blocks will be used to represent edge effects and irregular shapes related to topography. This parameter is used to adjust the solution volume and the ore inventory in each block.

- Vertical loading factor

  The vertical cross-sectional area factor is entered here. METSIM assumes blocks to be rectangular. The volume filling of each block is calculated as the product of the horizontal and vertical loading factors.

This data is used to construct the heap. This construction can then be edited to delete blocks or levels, adjust horizontal and vertical loading, and, if necessary, sort areas into specific construction phases.

3. HPLC defines the leach cycle and spray rate for addition of leach liquor to the heap. Any number of leach cycles with different solutions spray rates and cycle times may be specified. The input parameters required for the leach cycle definition are:
   - Cycle step description
   - Cycle duration
   - Solution type
   - Solution application rate
   - Number of block in cycle
4. HPLD defines the under-heap drainage. This represents the under-heap drainage pattern, natural or constructed. Each block is assigned to a pipe, which is then directed to one of the three output streams to maintain control of heap drainage.
   - Block number
   - Pipe number
   - Fraction of flow
5. HPLS defines the block leaching sequence, area available for leaching, and the starting day for each block. These data use the same heap grid and block numbering system and may be identical to the HPLA array in simple leaches. If a different leaching cycle is required, HPLS must be edited. The leaching schedule must be coordinated with the construction schedule. This is usually simple with rectangular heaps but is often very difficult with irregular heaps and varying mine plans.

PLS Pond and Raffinate Pond

The solution storage ponds are defined in this section. The ponds are specified by size and can be manipulated with special calculation switches to eliminate negative flow during start-up and early operation of the model. It is important that all input and output streams be defined, including input, makeup, rain, inventory, evaporation, discharge, and overflow. The input parameters include dimensions and level controls.

If the actual size of the ponds is not known, a pond volume that will contain at least 24 hours of flow is suggested. The actual pond design is usually defined in the operating permit and reflects the climate and terrain.

Solvent Extraction and Electrowinning

The solvent extraction and electrowinning (SX/EW) process can be modeled in detail if desired. However, in order to simplify the model, a detailed SX/EW process was not used in the example modeled. A simple black box model was utilized to perform the solvent extraction and electrowinning recovery of copper with a defined recovery of 90 percent of the copper contained in the PLS.

MODEL CASE DESCRIPTION

The capability of METSIM and the usefulness in defining operating criteria for heap leach operations is demonstrated by manipulation of specific conditions and recording the changes in production, solution flow, and tenor for the various conditions. In order to evaluate the varying operating parameters, two basic models were established, each having a specific extraction curve. Both models utilized identical heap construction schedules, leaching cycles, and ore delivery. Each base model was then run with a different heap construction sequence to illustrate how the sequence can impact the total cathode production and the timing of the production. In addition, comparisons were made relative to the leaching time used for the model. Detailed data output in graphical form are presented in the figures, while summaries of the models are presented and discussed in the following paragraphs.

The primary variable used for discussion was the sequence of the heap construction. The base case (sequence 1) represents a standard pattern using a 20-cell by 20-cell grid. The heap is constructed and leached row-by-row and lift-by-lift. Each lift is completely filled before the next lift is started. From this base case, two alternative construction/leaching sequences were tested. Each loading sequence is discussed in additional detail in this paper.

Model Conditions—Common

The common conditions for each model are as shown in the following table.

Common criteria—all models

Ore feed rate	10,000	mtpd
Ore grade	1.00	% T Cu
Heap area	810,000	m^2
Block area	2,080	m^2

Common criteria—all models (continued)

Lift height	6	m
Bulk density	1.6	mt/m^3
Leach cycle	100	days
Irrigation rate	0.1667	lpm/m^2
Maximum block leaching	50	
Leaching rate, fast ore	10	%/day
Leach rate, slow ore	1	%/day
SX/EW recovery	90	%

Model Conditions—Specific

The conditions varied and tested are as shown in the following table.

Specific model criteria

Parameter	Oxide Curve	Sulfide Curve
Fast leaching CuO—%	50	10
Slow leaching CuO—%	30	70
Insoluble CuO—%	20	20
Reaction rate—fast	10	10
Reaction rate—slow	1	1
Construction sequence	1, 2 & 3	1, 2 & 3

Three different construction/leaching sequences were used to illustrate the effect of heap configuration on the production schedule. Each construction phase is a set of 5 rows of 20 blocks each. Each phase is represented in the following tables by the numbers from 1 to 20. There are 2,000 blocks in the complete model. In all cases the leaching sequence was the same as the construction sequence. The various sequences are explained as follows:

Sequence 1. This sequence (see Table 1) constructs the heap in a horizontal manner one lift at a time. Each lift is completely filled before the next lift is started.

TABLE 1 Construction sequence no. 1

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
50	17	17	17	17	17	18	18	18	18	18	19	19	19	19	19	20	20	20	20	20
40	13	13	13	13	13	14	14	14	14	14	15	15	15	15	15	16	16	16	16	16
30	9	9	9	9	9	10	10	10	10	10	11	11	11	11	11	12	12	12	12	12
20	5	5	5	5	5	6	6	6	6	6	7	7	7	7	7	8	8	8	8	8
10	1	1	1	1	1	2	2	2	2	2	3	3	3	3	3	4	4	4	4	4

Sequence 2. This sequence (see Table 2) constructs the heap in a vertical manner, 5 rows at a time. Rows 1 to 5 for all 5 lifts, followed by rows 6 to 10 for all 5 lifts, etc.

TABLE 2 Construction sequence no. 2

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
50	5	5	5	5	5	10	10	10	10	10	15	15	15	15	15	20	20	20	20	20
40	4	4	4	4	4	9	9	9	9	9	14	14	14	14	14	19	19	19	19	19
30	3	3	3	3	3	8	8	8	8	8	13	13	13	13	13	18	18	18	18	18
20	2	2	2	2	2	7	7	7	7	7	12	12	12	12	12	17	17	17	17	17
10	1	1	1	1	1	6	6	6	6	6	11	11	11	11	11	16	16	16	16	16

Sequence 3. This sequence constructs the heap in a stairstep manner as illustrated in Table 3.

TABLE 3 Construction sequence no. 3

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
50	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	20	20
40	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	20
30	2	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	19
20	1	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
10	1	1	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18

Cases Evaluated

Several alternative cases were run for 3,000 days and the various data are presented in the following section. Selected results from the runs are displayed graphically and presented in the figures. The cases included herein are summarized in the following table.

Case Designation	Extraction Curve	Construction Sequence	Days Leach
#11100	Oxide	1	100
#12100	Oxide	2	100
#13100	Oxide	3	100
#21100	Sulfide	1	100
#22100	Sulfide	2	100
#23100	Sulfide	3	100
#21075	Sulfide	1	75

Data Summary—Cathode Production

The results of the computer runs are presented graphically in the figures.

Figure 1 displays a comparison of the "oxide" and "sulfide" extraction curves. It should be noted that both curves provide for an ultimate extraction of 80 percent of the total copper contained in the ore. Review of Figure 1 indicates that the oxide curve achieves a copper extraction of 50 percent in 5 days while the sulfide curve requires 18 days to reach the same extraction.

Figure 2 and Figure 3 summarize the daily cathode production resulting from the three alternative construction sequences for both extraction curves. For the oxide ore in Figure 2, the significant difference in the three alternative cases is the production in the first 900

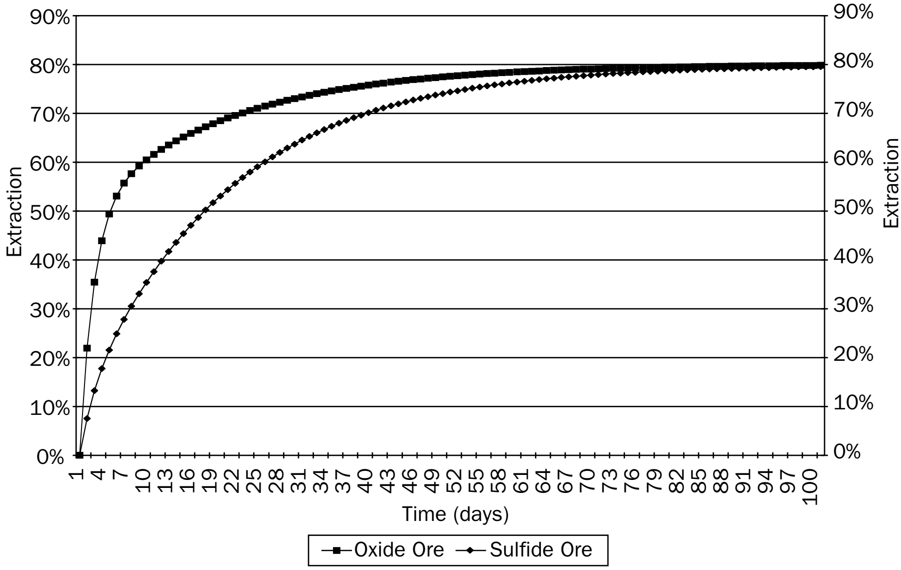

FIGURE 1 Extraction curves

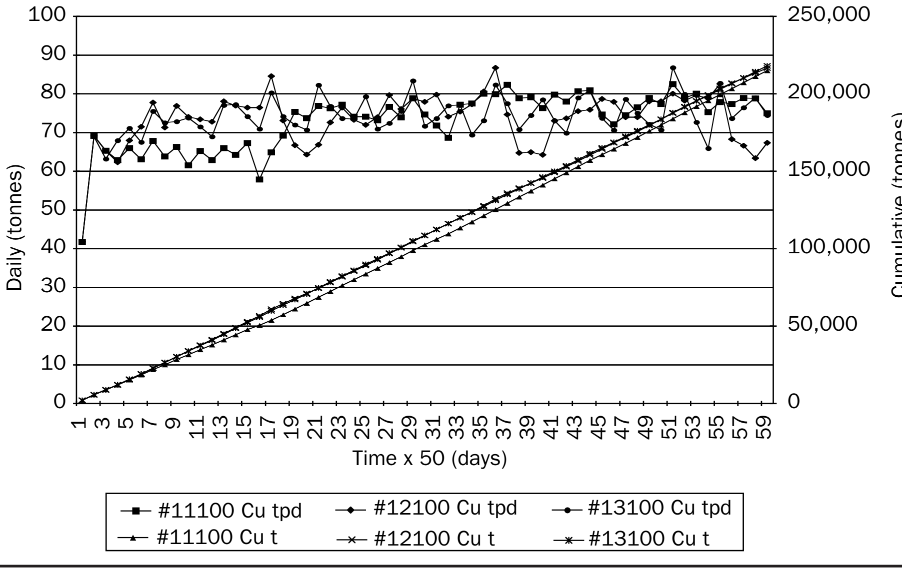

FIGURE 2 Daily and cumulative cathode production oxide extraction curve

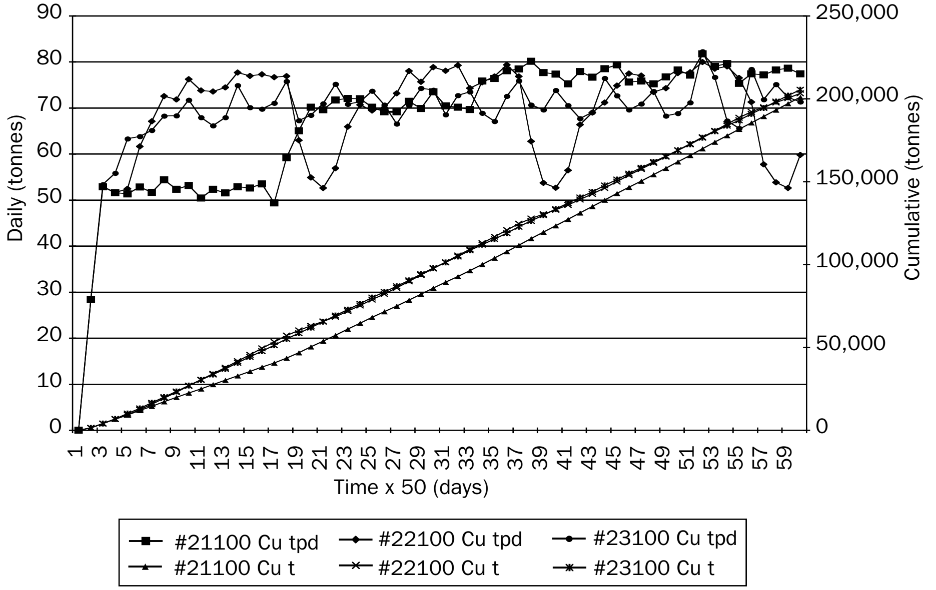

FIGURE 3 Daily and cumulative cathode production sulfide extraction curve

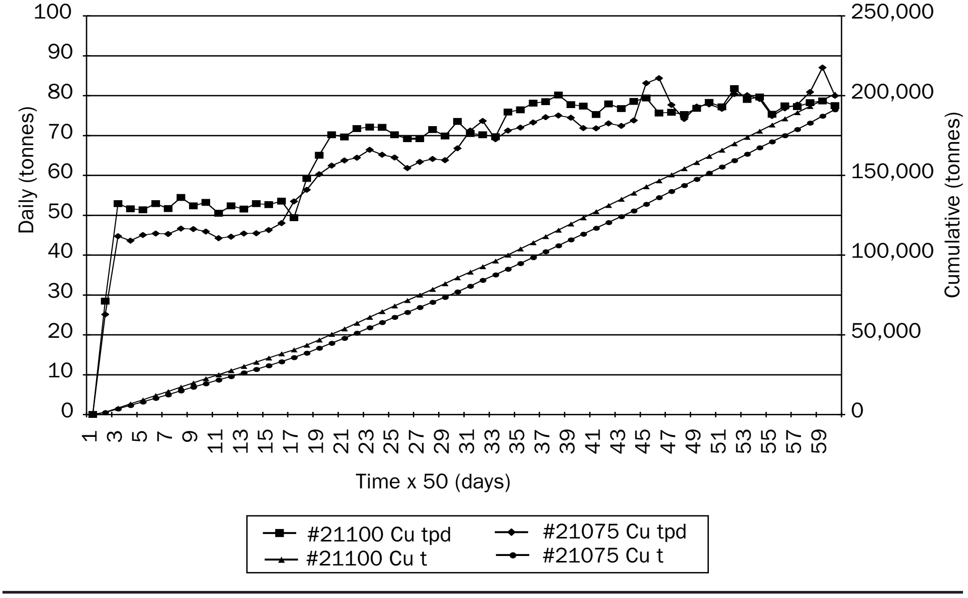

FIGURE 4 Daily cathode production leach cycle comparison

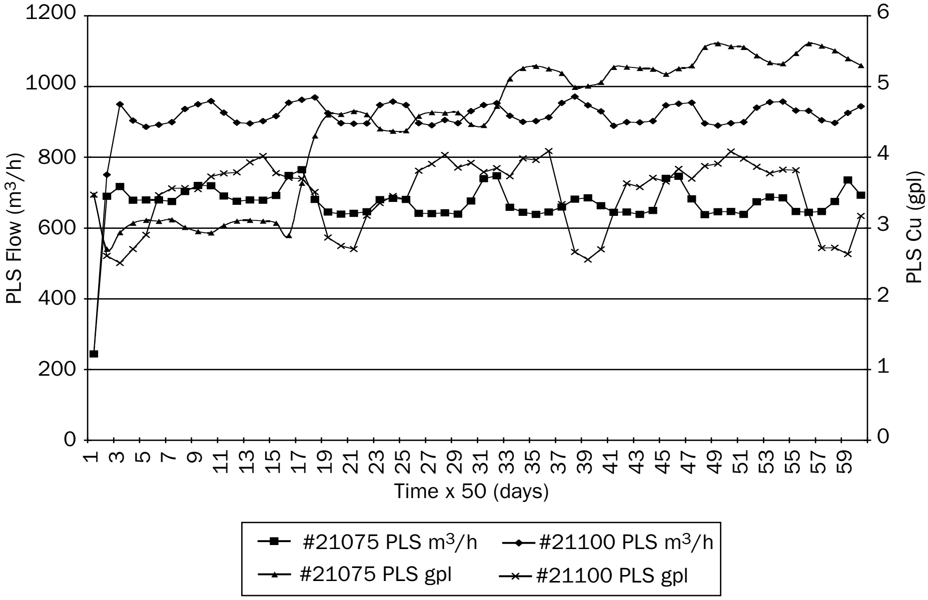

FIGURE 5 PLS flow and copper tenor leach cycle comparison

days of operation. Sequence 1 has an average daily production of 65 tonnes per day while sequences 2 and 3 have an average daily production of about 72 tonnes per day. Sequence 2 has the greatest fluctuation in daily production. For the sulfide ore in Figure 3, the significant difference in the three alternative cases is also the production in the first 900 days of operation. Sequence 1 has an average daily production of 52 tonnes per day, while sequences 2 and 3 have an average daily production of about 72 and 69 tonnes per day, respectively. Again, sequence 2 has the greatest fluctuation in daily production.

The objective of studying alternative construction sequences is an attempt to accelerate the cathode production for ores with slow leaching curves. The sulfide ore curve, where the majority of the copper is extracted as “slow” dissolving copper and high extraction levels are not realized until the second or third lifts are constructed and placed under leach, showed the greatest improvement. The sulfide ore showed a 30% improvement while the oxide ore showed a 10% improvement.

Figure 4 illustrates the difference in cathode production due to reducing the leach time from 100 to 75 days. Figure 5 shows the PLS flowrate and tenor for the same cases as Figure 4. The reduced leach time results in lower PLS flowrates along with higher grades. The net result is a decrease in production. The difference is about 7 tonnes per day. This difference decreases with time as additional lifts are put on leach.

It should be pointed out that at many operating heap leach operations, the actual leach cycle becomes dependent on the rate of heap construction and the volumetric limit of the solvent extraction plant. This leach cycle often becomes shorter than originally planned in order to maintain the higher PLS tenor so that a higher cathode production level can be

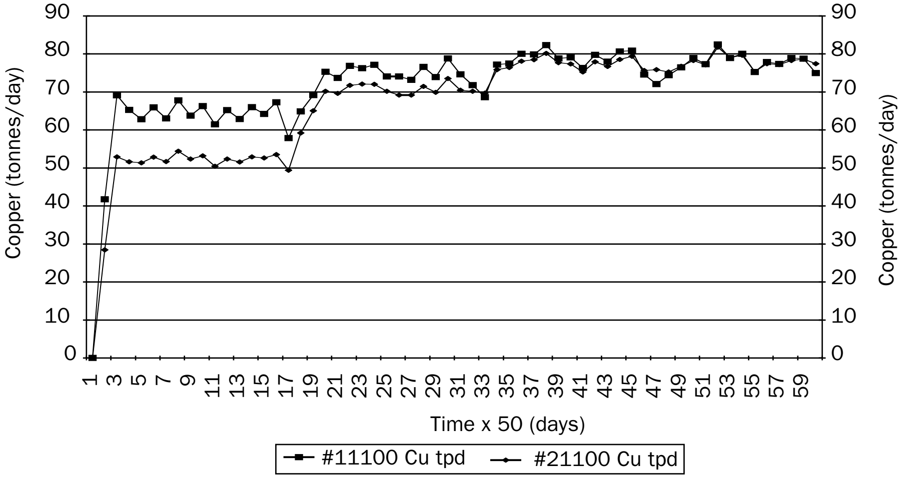

FIGURE 6 Daily and cumulative cathode production oxide to secondary sulfide

achieved with the same flow of PLS solution. At facilities that have limited heap area, a slow extraction curve, and a limited ore production, recycle (stacking) of solution becomes necessary. By construction of a series of models prior to design and construction, the impact of heap construction rate and sequence, leach cycle, solution stacking, and ore production can be evaluated well in advance. Verification of the design criteria through modeling could have a dramatic impact on the plant design, capital cost, and ultimate operating cost. Utilization of models would provide the operator with advance knowledge of the requirements for construction of new pad areas, expansion of the recovery facility, addition of solution recycle systems, and advance information relative to the impact of ore changes on production. This is particularly true if the oxide ore is processed for the first years of production, followed by a mixed or secondary sulfide ore.

In order to demonstrate the impact of leach time, an additional model run was made utilizing the oxide extraction curve and the No. 1 construction sequence. In these cases, the leach cycle was adjusted to 75 days. The impact on the daily cathode production is demonstrated in Figure 5. The cathode production levels were impacted somewhat by this change in operating condition. However, as shown in Figure 6, major impacts are noted in the PLS flow and copper tenor. Several models should be constructed and tested before specific design criteria are selected. Figure 6 demonstrates that for the specific ore being evaluated, a smaller volume of solution could be produced and the desired level of cathode production could be maintained by designing the heap leach operation for a shorter leach cycle.

Figure 7 demonstrates the variation in dissolved copper inventory under different heap construction sequences. The impact of solution stacking can be evaluated and the design engineer can compare intermediate solution recycle to operation with higher volumes of lower grade PLS.

Specific graphical data for all of the model runs are included in the figures. The data include daily values for cathode production, PLS flow, and PLS tenor. The limited data

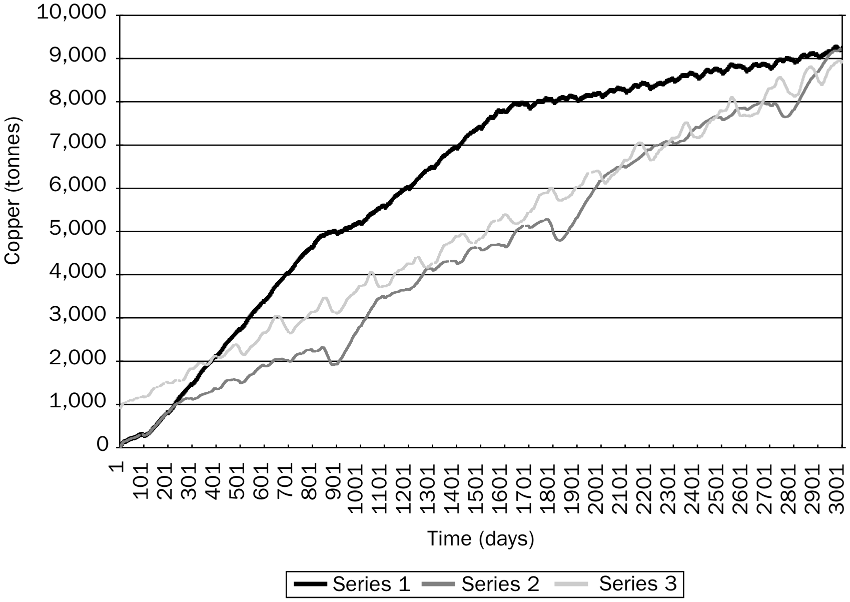

FIGURE 7 Variation in dissolved copper inventory under different heap construction sequences

output was arbitrary to limit the volume of information. Any number of variables can be selected for output. These can include information such as:

- Raffinate acid tenor—gpl
- Dissolved copper inventory—metric tonnes
- Dissolved copper solution inventory tenor—gpl
- Raffinate flow to heap—m^3/hr
- PLS flow—m^3/hr
- PLS tenor—gpl Cu
- Cathode production—tonnes/day
- Cathode production—cumulative tonnes
- Blocks on leach
- Tonnes of copper place on heap
- Raffinate tenor—gpl Cu
- Recovery of total and/or soluble copper—%
- Volume of makeup solution, precipitation, and evaporation

CONCLUSIONS AND OBSERVATIONS

Review of the data and above comments would suggest the following when considering a heap leach operation:

- The flowrate and tenor of PLS, and their effect on production, varies significantly with changes in heap design and leach times. Many other variables have similar effects. These variations need to be determined for proper heap leach design.
- The proposed construction sequence of a heap leach operation should be modeled under varying conditions before selecting a final design.
- The impact of heap design on solution inventory, which has a significant effect on production timing, can be realistically determined by modeling.
- Operating conditions should be evaluated by modeling to ensure that proper design capacities are selected.
- Important information concerning ore schedules, heap construction methods, expansion requirements, and modification to solution distribution and processing systems can be derived from dynamic modeling.
- Production forecasts under a wide variety of operating conditions can be determined by modeling, allowing the operator to plan and implement operational changes to maximize profitability of the project.

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Bioheap Processes—Operational Requirements and Techniques

Corale L. Brierley[*] **and James A. Brierley**[†]

Bioheap leaching of secondary copper ores is commercially practiced at ten operations. Good performance of copper bioheaps under severe climatic and operational conditions underscores the robustness of the technology. Important operational parameters affecting bacterial performance are acid balance, agglomeration and ore-stacking methods, culture addition, nutrients, solution management, aeration, water quality, mineralogy, solution chemistry, toxic constituent management and temperature. Improved heat modeling and better understanding of bacterial development in bioheaps undergoing heating are needed. Copper bioheap leaching is considered an important processing tool by the mining industry because of its low cost, short construction time, operational simplicity, good performance, and environmental advantages.

INTRODUCTION

Copper bioleaching began in the 1950s with dump leaching, a process that releases copper from vast quantities of submarginal grade primary and secondary copper sulfide materials. Today, dump leaching remains a vital process for the copper industry. In the last decade, a coupling of dump leaching, copper oxide heap leaching, and industrial microbiology has yielded a successful process for bioheap leaching secondary copper and sulfidic refractory precious metal ores. Bioheap leaching is a simple yet robust process offering capital and operating cost advantages and environmental benefits. The proliferation of bioheap leach operations (Table 1) during the last five years attests to the good performance and profitability of the technology. This paper aims to provide operational parameters important to successful bioheap leaching.

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[*] Brierley Consultancy LLC, Highlands Ranch, Colo.

[†] Newmont Metallurgical Services, Englewood, Colo.

TABLE 1 Commercial copper bioleach plants

Plant	Size tonnes/day	Years in Operation
Lo Aguirre, Chile	16,000	1980–1996
Gundpowder's Mammoth Mine, Australia	*in-situ*[a]	1991–Present
Mt. Leyshon, Australia	1,370	1992–in closure (1997?)
Cerro Colorado, Chile	16,000	1993–Present
Girilambone, Australia	2,000	1993–Present
Ivan-Zar, Chile	1,500	1994–Present
Quebrada Blanca, Chile	17,300	1994–Present
Andacollo, Chile	10,000	1996–Present
Dos Amigo, Chile	3,000	1996–Present
Zaldivar, Chile	~20,000	1998–Present

[a] ≈1.2 million tonne ore body

BIOHEAP LEACHING PRACTICES

In this section, secondary copper ore bioheap leach practices are addressed in relation to bacterial activity and performance. When appropriate, the differences and similarities between copper bioheap and sulfidic refractory gold bioheap operations are highlighted.

Bioheap Leaching: General Process Description

Mined ore is crushed to an optimum particle size. The crushed ore is mixed with sulfuric acid in an agglomerating device to consolidate the fines with the coarser ore particles and precondition the ore for bacterial development. Water or raffinate is added to optimize the moisture content. In sulfidic refractory gold bioheap leaching, a bacterial culture may be used in the agglomeration step to meet the acid demand of the ore, add moisture and inoculate the bioheap. The agglomerated and preconditioned ore is conveyed to the leach area where it is stacked 6 to 10 m (20 to 33 ft) high on a lined pad or on top of previously leached ore. Plastic piping with ventilation holes are placed on the pad to supply air to the bacteria during leaching. Dynamic (ON/OFF) pads with a coarse gravel bed are always used in bioheap leaching of sulfidic refractory gold ores. The biooxidized gold ore, after rinsing, is removed from the pad, mixed with lime and restacked on a permanent pad for leaching with cyanide or other lixiviant. Both dynamic and permanent pads are used in secondary copper bioheap leaching. Aeration of the bioheap is initiated soon after stacking the agglomerated ore. Low-pressure fans supply air to the ventilation system under the ore. The bioheap is irrigated with raffinate or recycled leach solution at an application rate that does not cause saturation. Pregnant leach solution is collected at the base of the bioheaps and directed to an SX/EW circuit for copper recovery. The raffinate is returned to the bioheap. Leach times vary among the operations but is typically in the 200-day range for secondary copper ores. Copper recoveries also vary with 80 to 85% recoveries achieved at most operations. Two recent book chapters detail the bioheap leaching of secondary copper ores (Schnell 1997) and refractory sulfidic gold ores (Brierley 1997A).

Role of the Bacteria in Bioheaps

Iron- and sulfur-oxidizing bacteria, notably *Thiobacillus ferrooxidans*, *Leptospirillum ferrooxidans*, *Thiobacillus thiooxidans* and other closely related microorganisms, are the principal organisms involved in bioheap leaching. However, moderately thermophilic bacteria, such as *Sulfobacillus* species naturally develop within the bioheap (Brierley 1997B; Shutey-McCann et al. 1997) as exothermic reactions increase the temperature above 40°C. The natural colonization of bioheaps by the extremely thermophilic bacteria (*Acidianus brierleyi* and *Sulfolobus* species) has not been confirmed. Sulfidic refractory gold bioheaps tend to heat as pyrite and arsenopyrite oxidize, liberating precious metals.

There are two viewpoints on the bacterial mechanism in bioleaching. One perspective is that the bacteria directly oxidize the mineral. For secondary copper sulfide leaching, this is represented as follows:

$$Cu_2S + 0.5O_2 + H_2SO_4 \rightarrow CuSO_4 + CuS + H_2O$$

$$CuS + 0.5O_2 + H_2SO_4 \rightarrow CuSO_4 + H_2O$$

The other viewpoint, which is more dominant, is that iron plays the central role in bioleaching. Chemical leaching of secondary copper sulfide minerals by ferric iron is represented accordingly:

$$Cu_2S + Fe^{3+} \rightarrow Cu^{2+} + Fe^{2+} + CuS$$

$$CuS + Fe^{3+} \rightarrow Cu^{2+} + Fe^{2+} + S^\circ$$

In a chemical leach, the principal roles of the iron- and sulfur-oxidizing bacteria are to rapidly oxidize the ferrous iron

$$2Fe^{2+} + 2H^+ + 0.5O_2 \rightarrow 2Fe^{3+} + H_2O$$

and oxidize elemental sulfur to sulfuric acid

$$2S^\circ + 2H_2O + 3O_2 \rightarrow 2H_2SO_4$$

Operational Parameters Critical to Bioheap Leaching

Operational guidelines for effective bioheap leaching are discussed in relation to the factors that are most important to the performance of the leaching bacteria, the kinetics of copper extraction and overall copper recovery.

Crushing and Agglomeration. The optimum particle size for bioheap leaching is determined from laboratory column tests and pilot trials. A range of particle sizes from 3.8 cm (1.5 inches) to 0.9 cm (3⁄8 inch) are evaluated in columns for ultimate copper recovery, leach rates, acid consumption and permeability during the leach cycle. Generally, a particle size of 80% −0.6 cm (−1⁄4 inch) is unacceptable, because the heap becomes too impermeable during leaching. The particle size selected is based on a complex cost/benefit analysis that considers crushing costs (secondary versus tertiary circuits), residence time of the ore on the pad, real estate for pad construction and impoundments and other economic factors.

Crushing equipment that minimizes fines production is best for bioheap leaching. Excessive fines (particles <200 mesh) limit the permeability of the bioheap to air and water. If −200 mesh material constitutes more than 20% of the weight of the ore, fines removal is necessary. On the other hand, some fines are necessary to ensure an adequate residence time of the solution in the bioheap. This allows for a temperature increase in the solution and ensures good oxidation of the ferrous iron by the bacteria.

Dry crushing is desirable, as this allows a controlled amount of moisture in the form of acid, bacterial culture and/or water to be added to the ore during agglomeration. A moisture content between about 7 and 10% provides the best aggregation of the fines to the ore, supplies the bacteria with optimum moisture and ensures good bioheap permeability to air and leach solution.

Acid, water and/or bacterial culture addition to the crushed ore is best accomplished in a drum agglomerator, as this device provides good mixing. If column testwork indicates that fines migration during bioheap leaching reduces permeability, acid-tolerant polymers can also be added during agglomeration.

Ore Stacking and Pad Configurations. Agglomerated ore is conveyed to the pad area and is stacked 6 to 10 m (20 to 33 ft) high using radial or other types of stacking devices. Truck stacking of the agglomerated ore or stacking by any methods involving traffic on the bioheaps is undesirable, because compaction significantly diminishes air and water permeability. ON/OFF, or so-called dynamic pads, are used in the bioheap leaching of sulfidic refractory gold ores. After biooxidation of the sulfides that occlude the gold and rinsing of the biooxidized residue with water to remove cyanicides, the bioheaps are dismantled and the ore is reagglomerated with lime, restacked and leached with cyanide or other gold lixiviant. Dynamic pads can also be used in copper sulfide bioheap leaching. When permanent pads are used, two modes of operation are possible:

1. One option is to stack multiple lifts and allow the leach solution to percolate through all of the lifts, as is done in oxide copper heap leaching. Aeration lines are placed on the top of the lift before stacking the next lift. The principal advantage of this option is that long-term leaching will ensure that maximum copper is recovered from the ore. The disadvantages are that (a) fines may migrate and reduce permeability, causing a hydraulic head buildup in the bioheap and geotechnical failure of the heap, and (b) there is no “bleed” in the system that reduces the build-up of constituents that are potentially toxic to the bacteria.
2. The second permanent pad option is to bioleach each lift and then “take it off line” by compacting the top surface of the leached heap. The PLS is collected at the bottom of each lift. The disadvantage of this option is that maximum copper leaching may not be achieved before the bioheap is taken off-line. The advantages are improved air and water permeability in the bioheap overall, minimized possibility of an undesirable hydraulic head build-up and a bleed mechanism to reduce toxic constituent build-up in the leach solution.

Bacterial Culture. Leaching bacteria, indigenous to the ore and the surrounding environment, will develop naturally in the bioheap when the crushed ore is moistened and acidified with H_2SO_4 to between about 1.5 and 2.5. The initial colonization and development of a good population (around 10^6–10^7 bacteria/g of ore) can require several months. However, once this population develops, some 10^6–10^7 bacteria will be present in the pregnant leach solution. These organisms pass through the SX/EW circuit with a slight loss in numbers and activity. Bacterial numbers in the raffinate are usually 10^4–10^5/ml, which is sufficient for inoculation of the next bioheap. Inoculation occurs when the bioheap is irrigated with the raffinate. The exact amount of time required for the bacteria to infiltrate the entire depth of a 6 m (20 ft) high heap and establish a population of some 10^6–10^7/g of ore is not known but is estimated at 30 to 60 days.

When ON/OFF pads are used, as they are in bioheap leaching of sulfidic refractory gold ore, the residence time of the ore on the pad is critical from an economic standpoint. To hasten the biooxidation process, the ore is inoculated before it is stacked (Brierley 1997A). This lessens the residence time of the ore on the pad by eliminating the time required for the bacteria to filter through the heap and reproduce to the desirable population level. One method of adding bacteria is to culture organisms on ferrous iron in tanks and agglomerate the ore with the culture. This places a large population of active bacteria proximate to the ore particles and distributes the bacteria throughout the entire bioheap. Agglomeration of the ore with the culture adds Fe^{3+}, which enhances chemical leaching, preconditions the ore to the proper pH for bacterial development and provides moisture to consolidate the fine particles onto the larger particles. Ore inoculation is believed to reduce the ore's residence time on the pad by up to two months. If concentrated H_2SO_4 is required to precondition the ore, the inoculation process cannot be accomplished in the same agglomeration device as the acid preconditioning. The high acidity of the concentrated H_2SO_4 is detrimental to the bacteria.

To date, inoculation with bacteria has been confined to the *Thiobacillus* and *Leptospirillum* species. However, bioheap inoculation with different types of organisms may be a sound practice, as more is learned about temperature profiles in bioheaps, the types of bacteria that develop and function under different temperature profiles and the performance of the different organisms.

Acid Balance. The leaching bacteria are acidophilic, requiring a pH of between about 1.3 and 2.3 for optimum activity. This pH range is also required to maintain Fe^{3+} in solution for chemical oxidation. During the column testwork phase, acid consumption and acid production are determined. The amount of acid to be added to the ore for good bioleaching and the amount of acid produced from sulfide oxidation are determined for the particle size that will be used. Carbonate minerals, clays and certain bioleach reactions, including Fe^{2+} oxidation, consume acid. The oxidation of pyrite and some other sulfide minerals is acid generating. Some sulfidic refractory gold ores have a high percentage of sulfides as pyrite with little acid consuming gangue mineralization. With these ores, so much acid can be produced by the oxidation reactions that neutralization of the leach solution may be necessary to maintain a suitable pH for the bacteria.

In copper bioheap leaching, a portion of the total acid required by the ore is added during agglomeration. This amount must be sufficient to precondition the ore for bacterial development and guarantee that Fe^{3+} and Cu^{2+} will remain in solution. Adding too much acid during agglomeration will cause breakdown of gangue materials, which simply consumes acid, thus increasing operating cost without providing any preconditioning benefits. A high quality acid must be used during agglomeration to ensure that no bacterially toxic constituents, such as mercury, are added. The remainder of the acid required is supplied by the raffinate during irrigation, as acid is generated in solvent extraction. This means that, early in the leach cycle of a copper bioheap, the pH varies throughout the bioheap. Acid is consumed near the top portion of the heap, resulting in higher pH values lower in the bioheap. If the pH becomes too high during passage of the solution through the bioheap, iron will precipitate, which would stop chemical leaching of the copper sulfide minerals and vastly diminish bacterial activity. There is a subtle balance that encompasses acid added during agglomeration, effective utilization of the acid in the raffinate, pH control through the depth of the heap and the heap height (Schnell 1997).

Aeration. The leaching bacteria require oxygen, which is the electron acceptor, and carbon dioxide, which is the bacteria's carbon source for cellular components. Carbon dioxide is usually not limiting in bioheaps, because it is produced from the dissolution of carbonate minerals (if present) by the acid. However, oxygen can become limiting. Oxygen requirements in the bioheap are determined theoretically from column test data. The types and quantities of all sulfide minerals oxidized and the time required to oxidize these minerals are ascertained in column tests. The theoretical oxygen requirement is calculated from the oxidation of the sulfide minerals. Although we stated earlier that Fe^{3+} leaching of the copper sulfide minerals is the most important mechanism, oxygen requirements are not based on the bacterial oxidation of Fe^{2+}, but rather on the oxygen requirement for oxidation of the sulfide minerals. An oxygen utilization efficiency factor (usually 10%) is incorporated. The amount of air provided is then calculated based on the amount of sulfide to be oxidized, the efficiency factor, the bioleaching time to achieve the desired metal recovery, and engineering considerations such as elevation above sea level.

The size of the aeration lines under the bioheap, the spacing between the lines and the number and size of the holes in the aeration lines are dependent on the amount of sulfide to be oxidized over time. Higher sulfide contents associated with some sulfidic refractory ores, which oxidize quickly, require larger and more closely spaced air lines with numerous and larger holes. Whereas, many secondary copper sulfide ores, which have lower sulfide content and require longer to oxidize, will have smaller air lines spaced further apart and the lines have fewer and smaller holes. Air is supplied using low-pressure blowers. Sufficient pressure is applied to overcome the pressure drop across the aeration lines and to prevent liquid from entering the air lines (Lancaster and Walsh 1997). Placement of the tubes carrying the air is not an exact science. In sulfidic refractory gold bioheaps, which employ ON/OFF pads, the air lines are located in the protective gravel bed that overlies the HDPE liner. This configuration allows the aeration lines to be undisturbed when the biooxidized ore is removed. For permanent pads, the aeration lines for the first lift can be placed in a protective gravel bed. Thereafter, the aeration lines can be placed directly on top of the lift. When the top surface of the lift has been compacted to prevent solution flow through the bottom lifts, placing the aeration lines on the compacted surface may cause inadvertent flooding of the aeration lines because of their position in the hydraulic head. Positioning the aeration lines in the ore bed above the hydraulic head minimizes the problem of solution in the aeration lines but is operationally difficult.

Water Quality. The bioleaching bacteria are robust; however, certain constituents, namely Cl^- and NO_3^- at elevated concentrations, can be quite detrimental to the bacteria. The manifestation of this toxicity is a reduced rate of Fe^{2+} oxidation. Unfortunately Cl^- and NO_3^- are found in many mining areas around the world. The Cl^- toxicity factor also precludes the use of seawater, which would be desirable in arid, coastal mining operations. In laboratory tests, Fe^{2+} oxidation by *Thiobacillus ferrooxidans* is slowed in the presence of 2.5 g/L NaCl, and 6 g/L NaCl completely inhibits Fe^{2+} oxidation. NO_3^- at 200 mg/L begins to cause inhibition of Fe^{2+} oxidation by *T. ferrooxidans*. Some scientists are now searching for Cl^- and NO_3^- resistant strains of iron oxidizing bacteria that have rapid Fe^{2+} oxidation capability. However, the principal message is that if there is any concern about the quality of the water that will be used in operations, laboratory column tests should be performed using site water to establish and quantify the effect.

Nutrients. The leaching bacteria require about 10 mg/L NH_4^+ and about 30 mg/L of PO_4^{3-} in solution. In bioheap leaching, these nutrients are usually provided initially by adding N-P-K fertilizer. Any fertilizer product must be tested in the lab before use in the field to confirm that the product will support the growth of the bacteria and does not contain any inhibitory substances. Once bioheap leaching is initiated, sufficient NH_4^+ and PO_4^{3-} are dissolved from blasting agents and the rocks to meet the requirements of the bacteria. Indiscriminant addition of nutrients is to be avoided, because it results in undesired precipitation of Fe^{3+} and costly losses of NH_4^+ and PO_4^{3-} as ammonium jarosite and $FePO_4$.

Irrigation and Solution Management. The raffinate, or leach solution, is usually applied to the top surface of the bioheaps using drip irrigation methods. Sprinklers or wobblers can be used in geographic regions with a net positive water balance to evaporate excess water; however, these irrigation methods may cause fine particles to detach from near-surface agglomerates and potentially cause permeability problems from fines migration (Readett et al. 1997). Drip lines can be buried several 10s of cm beneath the surface of the bioheap to minimize evaporative losses and/or minimize freezing problems (Schnell 1997). Solution application is usually about 0.12 $L/min/m^2$ to ensure that bioheap saturation does not occur. Ponding of solution on the bioheap surface is a clear signal that bioheap permeability is poor and that saturation is occurring. If the heap is saturated, oxygen mass transfer to the bacteria is greatly limited. The result is poor oxidation rates, manifested as reduced metal recoveries.

Effluent from the bioheap is recycled in the bioheap leaching of sulfidic refractory gold ores, and this practice can be employed in copper bioheap leaching to increase the copper grade in the PLS. Otherwise, the PLS is directed to the solvent extraction-electrowinning circuit and the raffinate is applied to the copper bioheap.

Irrigation with raffinate during the early stage of chalcocite bioheap leaching is vital to ensure pH conditioning of the bioheap. Remember that not all of the acid to meet the acid demand of the ore is added during agglomeration. This allows use of acid produced during SX-EW to meet the acid demand. Furthermore, irrigation and acidification releases any oxide copper that may be present in the ore.

Rest cycles, during which no solution is applied to the heaps, are employed during bioheap leaching of chalcocite and sulfidic refractory gold ores. The precise benefits of rest cycles are difficult to quantify but are probably related to the following:

1. maximizing the mass transfer of oxygen to the bacteria by limiting the thickness of water films around sulfide mineral particles,
2. optimizing chemical oxidation by building up ferric iron concentrations, and
3. building up heat which enhances oxidation rates.

Rest periods are usually employed in older chalcocite bioheaps after the heap has largely been depleted of leachable copper. Managing rest cycles and irrigation cycles is complicated in copper sulfide bioheap leaching, because of the need to control solution inventories and also ensure a fairly constant PLS grade to the SX circuit.

Iron Dynamics in Bioheaps. Sulfide minerals are oxidized by ferric iron in solution, and the principal role of the bacteria is to reoxidize the ferrous iron resulting from the reaction of the ferric iron with the sulfide minerals. In a well-designed and well-operated bioheap, the bacteria are distributed throughout the heap and have abundant oxygen to function. In such a bioheap, the ferrous iron is instantly oxidized and the ferric iron concentration is relatively constant throughout the bioheap. The concentration of iron in solution is a function of the amount of iron-bearing minerals in the ore, the quantity of these iron-bearing minerals that are degraded during bioleaching, leach solution pH, equilibrium of iron in the leach solution and the solution temperature.

In the bioheap leaching of sulfidic refractory gold ores containing large amounts (e.g., 5–10%) of arsenopyrite and/or pyrite, copious amounts of soluble iron go into solution. Total iron concentrations in solution reach upwards of 20–30 g/L. Prodigious amounts of iron precipitate in the bioheap as iron salts, because the solubility of iron is rapidly exceeded. Precipitation is exacerbated as the bioheap heats from exothermic reactions, because iron solubility is reduced at higher temperatures. Iron precipitation in the bioheap is not problematic as long as heap permeability is retained, sulfide surfaces are not blinded preventing oxidation, and, in refractory gold ore biooxidation, the precipitates do not increase reagent (lime and cyanide) consumption.

Bioheap leach operations, where the chalcocite ores have relatively low amounts of iron-bearing minerals, will operate with quite a low concentration (e.g., 1 to 2 g/L) of total iron in solution. The low iron concentrations are due not only to the small amounts of iron leached from the minerals but also because of precipitation of iron salts in the bioheap. Precipitation appears to be exacerbated by aeration (Schnell 1997). A low iron concentration is not a problem as long as biooxidation of the ferrous iron is rapid so that the ferric:ferrous ratio is high enough to maintain a redox potential higher than +600–650 mv (Ag/AgCl electrode). This redox potential is required for the dissolution of most base metal sulfides. Operating at low solution iron levels is, in fact, desirable in copper leaching, as it is causes fewer problems in the SX-EW circuit.

Toxic Constituents in Bioheap Leaching and Bleed Streams. Recycling of the leach solution undoubtedly concentrates certain constituents that could potentially be toxic to the bacteria. Chalcocite bioheap leach operations and many sulfidic refractory gold bioheaps are designed to continuously recycle the leach solution without purposely having a bleed stream. In effect, there are bleed streams in both types of bioheap operations when ON/OFF pads or permanent pads with inter-lift compaction are employed. Entrained solutions in the materials taken off-line serve as the bleed and probably remove sufficient quantities of potentially toxic constituents to avoid bacterial toxicity problems.

Many microbiologists point to high iron concentrations as being inhibitory to the bacteria and advocate bleed streams to precipitate iron before the treated effluent is recycled back to the bioheap. However, a quick calculation demonstrates that in bioheaps with high concentrations of iron-bearing sulfide minerals, the amount of iron solubilized in a single pass of leach solution through a bioheap can greatly exceed the solubility of iron in solution and the reported toxicity levels of iron. Iron precipitation within the bioheap will occur and this is probably the best place for disposal, because treatment and storage costs of iron precipitates would make many bioheap operations unprofitable. Caustic treatment of a bleed stream, however, is essential, if the amount of acid produced by pyrite oxidation exceeds the tolerance of the bacteria to acidity.

Organic entrainment in the raffinate from solvent extraction probably contributes to the reduced numbers of active bacteria in the raffinate. There is expressed concern about the effect of this organic on bioleaching in the heap. Few studies have addressed this, but bacterial numbers and activity appear to be quite high on the surfaces of bioheaps. This suggests that the organic reagents are quickly and effectively adsorbed to rock surfaces negating any toxic effects, or heterotrophic (i.e., microorganisms that oxidize organic matter for energy) acidophilic microorganisms are degrading the organic reagents.

Temperature. The oxidation of sulfide minerals, arsenopyrite and pyrite in particular, generates heat. Bioheaps, in which arsenopyrite and/or pyrite are specifically oxidized to expose occluded gold, can easily reach internal temperatures approaching 75°C (Brierley 1997B). On the other hand, chalcocite bioheaps depleted in pyrite and operated in cold climates (Schnell 1997) may generate little heat necessitating efforts to add and conserve heat during operation.

Bacterial populations and activity are greatly affected by temperature. When operating conditions are in the 20 to 40°C range, *Thiobacillus* and *Leptospirillum* species will be the predominant leaching organisms. At temperatures less than about 25°C, these organisms will have slower oxidation kinetics, reducing the rate of sulfide oxidation. This is manifested by longer residence time of the bioheap on the pad, negatively affecting capital and operating costs.

If abundant pyrite and/or arsenopyrite are present or these minerals are oxidized rapidly, bioheap temperatures can quickly exceed the uppermost tolerance level of the *Thiobacillus* and *Leptospirillum* species and these organisms will die. At about 35 to 40°C, the naturally occurring moderately thermophilic leaching bacteria, such as *Sulfobacillus* species, will begin to populate the heated zones. These organisms can function up to temperatures approaching 65°C. Extremely thermophilic leaching bacteria, such as the *Sulfolobus* species, *Acidianus brierleyi* and *Metallosphaera sedula*, are known to oxidize sulfur, ferrous iron and mineral sulfides and exist in extreme environments such as volcanic regions and hot, acid springs. However, their existence in bioheaps and dump leach operations has yet to be proven. Also, how the moderately and extremely thermophilic bacteria perform in terms of kinetics and overall metal recoveries has yet to be quantified in bioheap operations.

Bioheap leaching under cold climatic temperatures and high altitudes has been successfully accomplished at Quebrada Blanca (Schnell 1997). However, this endeavor has required some unusual operational efforts including heating the ore, heating the raffinate and ensuring that minimal heat is lost through evaporation. Heat modeling of bioheaps is still in its infancy. This science must be advanced so that a better understanding of what can be operationally expected under a variety of climatic conditions and ore types is predicted with some degree of accuracy.

MONITORING BIOHEAPS

Several techniques can used to monitor bioheaps for performance:

1. PLS is analyzed for pH, redox potential, acidity, total iron concentration and ferrous iron levels. The pH and acidity measurements indicate the extent of acid conditioning the bioheap and provide insight into the oxidation of pyrite. Redox potential, iron concentrations and ferric:ferrous ratios provide information on pyrite and other iron-bearing mineral dissolution and the performance of the iron-oxidizing bacteria.
2. Periodic sampling of ore at different depths and locations throughout the bioleach period to assay for residual copper provides inventory measurements as well as performance information. Other constituents, such as sulfide-sulfur and iron,

should also be assayed to determine the extent of sulfide oxidation, what minerals are oxidizing and the respective rates of oxidation. In refractory sulfidic gold bioheaps, the solids are subject to bottle roll cyanidation to confirm recoveries versus sulfide-sulfur oxidation.

3. Temperature measurements at various depths and locations throughout the bioheap provide information on pyrite oxidation. High temperatures indicate active pyrite oxidation.
4. Oxygen measurements at various depths will indicate whether sufficient aeration is being applied, and, when oxygen measurements are coupled with solids analyses, these data may point to active sulfide mineral oxidation.
5. Oxygen uptake measurements of solid samples, PLS and raffinate using a respirometer provide information on the activity of the bacteria. The principal disadvantage of this technique is the cost of the equipment to perform this test.
6. Bacterial counts can be made with a fair degree of accuracy in liquids, but the data are unreliable for solids. Therefore, indirect measurements, such as oxygen uptake, provide more accurate information.
7. The types and numbers of bacteria on ores and in solutions can be determined using sophisticated molecular methods (Jerez 1997). These methods are just beginning to emerge and performing these tests is relegated to a handful of experts in several laboratories around the world.

The validity of many of these monitoring techniques is predicated on obtaining good head samples and having accurate baseline (i.e., prebioleach) parameters.

CONCLUSIONS

Bioheap leaching of secondary copper ores is commercially applied with a large measure of success at ten operations mostly located in Chile. The continuous operation of Minera Quebrada Blanca since 1994 is particularly noteworthy, because this bioheap performs successfully under harsh conditions for bacteria—average temperature of 5°C (40°F) and an elevation of 4,600 masl (15,092 ft) where the oxygen partial pressure is about one-half atmosphere. Substantial heat losses across the bioheaps at Quebrada Blanca necessitate measures to increase heat in the bioheap and stem evaporation rates. Operating parameters, which benefit the bacterial activity hence kinetics of copper extraction and overall copper recovery, are reasonably well known and applied in chalcocite bioheap leaching. Specific areas requiring definition are heat modeling and bacterial population development and performance with bioheap temperature changes. Despite the comparative novelty of bioheap leaching of secondary copper ores, the mining industry is keen to apply the process. The reasons are (Brierley 1996)

1. good copper recoveries in relation to the level of processing,
2. low capital and operating cost of the process when compared to making and processing concentrates,
3. short construction times,
4. simplicity of operation,
5. robustness of the process under challenging climatic and operational conditions, and
6. environmental benefits realized from no gaseous or aqueous discharges and no tailings dam.

REFERENCES

Brierley, C.L. 1996. Minerals Biooxidation/Bioleaching Technology: Redefining Ore. In *Trends, Technologies, and Case Histories for the Modern Explorationist*, Prospectors and Developers Association, Toronto.

Brierley, J.A. 1997A. Heap Leaching of Gold-Bearing Deposits: Theory and Operational Description. In *Biomining: Theory, Microbes and Industrial Processes,* Edited by D.E. Rawlings, Berlin, Germany: Springer-Verlag and Landes Bioscience.

Brierley, J.A. 1997B. Microbial Populations in Mineral Biooxidation Processes. In *IBS/ BIOMINE 97, Proceedings of the International Biohydrometallurgy Symposium*, PSC1.1–PSC1.10. Glenside, South Australia: Australian Mineral Foundation.

Jerez, C.A. 1997. Molecular Methods for the Identification and Enumeration of Bioleaching Microorganisms. In *Biomining: Theory, Microbes and Industrial Processes,* Edited by D.E. Rawlings, Berlin, Germany: Springer-Verlag and Landes Bioscience.

Readett, D., D. Walsh, and T. Lancaster. 1997. Girilambone Copper Company—Developments in Heap Leaching of Copper Sulphide Ore. Paper presented at the 3rd Annual Copper Hydromet Roundtable '97, November 2–5 at Vancouver, British Columbia.

Schnell, H.A. 1997. Bioleaching of Copper. *Biomining: Theory, Microbes and Industrial Processes*, Edited by D.E. Rawlings, Berlin, Germany: Springer-Verlag and Landes Bioscience.

Lancaster, R., and D. Walsh. 1997. The Development of the Aeration of Copper Sulphide Ore at Girilambone. In *IBS/BIOMINE 97, Proceedings of the International Biohydrometallurgy Symposium*, M5.4.1–M5.4.10. Glenside, South Australia: Australian Mineral Foundation.

Shutey-McCann, M.L., F.P. Sawyer, T. Logan, A.J. Schindler, and R.M. Perry. 1997. Operation of Newmont's Biooxidation Demonstration Facility. In *Global Exploration of Heap Leachable Gold Deposits*, ed. D.M. Hausen, 75–82. Warrendale, PA: The Minerals, Metals and Materials Society.

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Pressure Oxidation of Complex Copper Ores and Concentrates

Rod McElroy* **and Wes Young†**

INTRODUCTION

Background

For over 30 years, various concepts for cost-effective hydrometallurgical treatment of mainstream (i.e., chalcopyrite) copper concentrates have been actively investigated. Many of these have advanced to pilot scale testing and a few to demonstration or small-scale commercial production. To date, the true "hydrometallurgical smelter" has not become a reality, but three current contenders are discussed in this presentation.

Despite the rather broad title, the scope of this presentation is limited to autoclave pressure oxidation ("POx") of copper concentrates (or high-grade ores) containing a major proportion of copper as chalcopyrite, the most common economic copper mineral and the most refractory to hydrometallurgical processing.

There are several examples of commercial atmospheric and pressure oxidation of (primarily) chalcocite-pyrite copper concentrates.[1,2] However, these are very site/deposit specific.

Previous generations of chalcopyrite hydrometallurgical process contenders have included variations on leaching with high chloride brines and nitrate-sulphate solutions. While the CLEAR chloride process[3] was a notable technical (if not economic) success and the Sunshine nitrate-sulphate POx process[4] is a continuing (if small scale) commercial success, the application of hydrometallurgical technology to mainstream copper concentrates remains an elusive goal.

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* Senior Metallurgist, Fluor Daniel Wright Ltd., Vancouver, B.C., Canada.

† Principal Process Engineer, Fluor Daniel Wright Ltd., Vancouver, B.C., Canada.

Incentive for Hydrometallurgical Processing

Smelting is the predominant treatment process for chalcopyrite concentrates. Strategically located facilities, typically with multimodal transport access and annual capacities well in excess of 200,000 metric tonnes per year (mtpy), justify their high capital cost by the economic "muscle" conferred by long economic life and large size compared to most individual mines.

Typically, a start-up mine carries a large load of debt/equity financing, so concentrate treatment as an operating cost is at least superfically attractive. However, especially for remote mine locations, there is definitely no free (or even cheap) lunch! From (for example) a north-central B.C. minesite, direct concentrate charges before the concentrate arrives at a smelter include:

- Road (or road/rail) haulage to ocean port
- Port charges
- Ocean freight
- Handling losses
- Insurance

Typically these total in the $US 45–55/tonne range.

After a 20% concentrate is delivered, the smelter will:

- Pay (eventually) for contained copper, less 1 unit (10 Kg or 5% of total), and precious metals net of contract deductions
- Deduct from payment for impurities in excess of specifications which may be onerous
- Charge $0.08/lb payable copper increasing by $0.01/lb for every $0.10/lb payable price over $0.80/lb. This escalating scale is described as "profit participation" by smelters but in less salubrious terms by concentrate shippers

Additional problems for the concentrate producer arise from:

- Payment delay for transport time of weeks to many months for ice limited ports
- Vulnerability to delays from a transport chain involving (typically) unionized teamsters, rail and longshore workers and maritime hazards

These factors tend to support concentrate producers' on-going quest for a viable minesite process which will put their fate (and more revenue) in their own hands.

Advances in POx and Copper SX-EW Technology

Over the past three decades, minesite production of cathode copper from ore leaching has grown dramatically due to emergence of very robust technology for copper recovery by solvent extraction–electrowinning. In an even shorter timeframe, autoclave POx technology has matured and become dominant in processing of zinc concentrates as well as many refractory gold ores/concentrates containing copper "penalty" elements such as arsenic (As) and antimony (Sb).

These developments provide at least an aura of "proven technology" to support current developers of POx-SX-EW processes for chalcopyrite concentrates.

Refractory gold POx technology provides the basis for the expectation of environmentally stable copper process residues containing noxious elements such as As and Sb, as well as holding out potential for deduction free, high precious metal (PM) recoveries. SX provides a barrier against virtually all impurities which deteriorate copper cathode quality and price.

NEW COPPER POx-SX-EW PROCESSES

The Players

In the past six years, three significantly different POx copper processes have been developed from bench scale scoping tests through to varying levels of continous pilot plant demonstration. The developers are:

- Cominco Engineering Services Ltd. (CESL)
- Placer Dome Inc. (Placer)
- Dynatec (formerly a division of Sherritt, Inc.)

All of these companies are serious players with track records of significant innovation in process metallurgy. Both Placer and CESL have direct or affiliate control of major copper mines/deposits. Since its decoupling from Sherritt, the Metallurgical Technologies Division of Dynatec is a stand-alone developer/vendor of process technology, laboratory, pilot plant and engineering services.

Of the developers, Placer has published by far the most actual test data[5,6,7] and has not hedged its position with patent protection. Dynatec has published a single paper and patented potential key features of its process.[8,9] CESL has taken an extensive patent position and made several presentations/publications which are less than totally revealing.[10,11]

Basis for Process Assessments

Assessments of the developmental processes are based on:

- Published data from developers
- Unpublished Fluor Daniel study data[12]
- Published and prepublication process assessments from various sources[13,14]

It is emphasized that technical and economic feasibility of any/all of the processes is highly site/project specific. Also, it is known that at least two of the process developers have commissioned extensive confidential internal/external engineering studies. Thus, any future quibbles from process developers regarding this presentation will be met with a resolute “no access to relevant data” defense.

Process Descriptions

In overview, the CESL and Dynatec processes are divergent approaches to low severity POx in which a high proportion of sulphide sulphur remains in the elemental (S^o) form in leach residue, thus limiting oxygen consumption which is a major operating cost.

In contrast, Placer’s high severity total oxidation process is an adaption of industrially proven high severity POx technology for refractory gold ores/concentrates.

All of the process developers have reported consideration of copper recovery by direct EW of autoclave pregnant leach solution (PLS). However, the technical feasibility of this

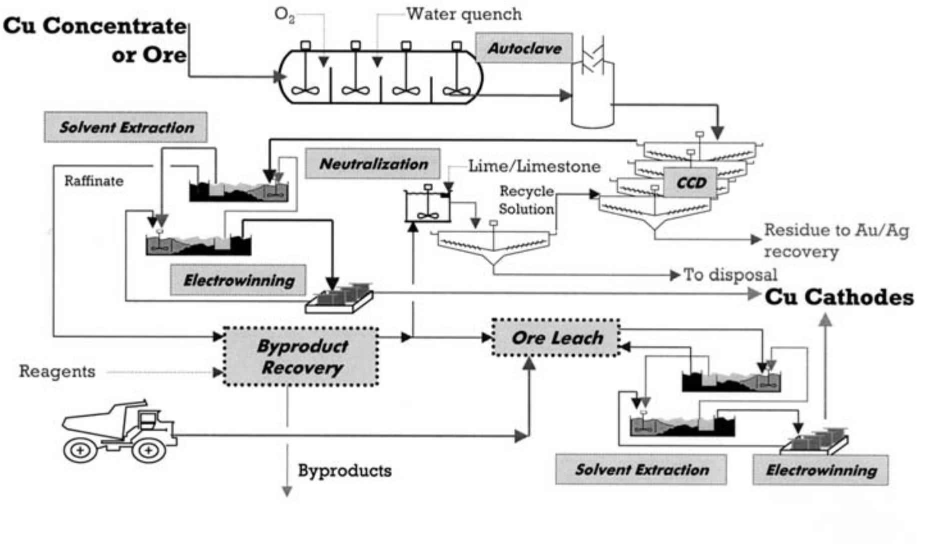

FIGURE 1 Placer's total pressure oxidation process

approach is highly site specific, whereas SX-EW technology is adaptable to most leach solutions. Only Dynatec retains direct EW in its published flowsheet. Interprocess comparisons, an objective of this presentation, are therefore based on copper recovery from PLS and other solutions by SX-EW. This levels the field by (roughly) equalizing post-leach copper recovery and product quality.

Placer Dome—Total Pressure Oxidation. The POx component of Placer's process (see Figure 1) is essentially the same as typical high severity industrial POx of refractory gold ores/concentrates. Reground concentrate is slurried in acidic recycle solution and pumped into the first compartment of an autoclave operated at 210–220°C with oxygen sparged to maintain (about) 700 kPa (100 psig) of overpressure. Near total oxidation of chalcopyrite and pyrite occurs via:

$$2CuFeS_2 + \frac{17}{2}O_2 + 2H_2O \rightarrow 2CuSO_4 \cdot aq + Fe_2O_3 + 2H_2SO_4$$

$$2FeS_2 + \frac{15}{2}O_2 + 4H_2O \rightarrow Fe_2O_3 + 4H_2SO_4$$

Basic iron sulphate (hydronium jarosite) is a significant coproduct which will collect silver (if present) and other monovalent cations via the (generalized) reaction:

$$3FeS_2 + \frac{45}{4}O_2 + \text{(approx) } 7H_2O \rightarrow (H_3O, Na, Ag, \text{etc.})\ Fe_3(SO_4)_2(OH)_6 + 4H_2SO_4$$

For concentrates and high sulphide ores the oxidation is exothermic, so cooling (typically by solution injection) is usually required. "Autothermal" feed is reported[13] to be about 25% solids for a high grade (2.5% Cu) ore containing 11% sulphide sulphur.

Autoclave product PLS is highly acidic (30–60g/L H_2SO_4) and is extensively diluted during CCD residue washing; feed to SX is thus relatively dilute. This minimizes the volume of material for PM (precious metal) recovery by lime/cyanide and maximizes its reactivity. A lime "boil" of POx residue may be justified for silver recovery. Raffinate and bleed streams are neutralized prior to recycle of solution and solids disposal.

A near "ideal" situation for application of total POx would be a site where dilute acid by-product (SX raffinate) could be used to recover copper from a local "oxide" resource, thereby providing an acid credit instead of a process cost for acid neutralization by high calcium material (preferably tailing, but imported limestone may be necessary).

Placer's recent presentation emphasizes that the process is most attractive for deposits with significant PM values in pyrite which are also challenged by penalty impurities and transport costs.[7]

CESL—Chloride Enhanced, Low Severity POx. CESL has experience in low severity POx technology from development and application of zinc concentrate POx processing in the 150–160°C range.

In a pure sulphate electrolyte at these temperatures, and in the absence of an effective dispersant, chalcopyrite reacts slowly[5] with formation of liquid S° "balls" which wet and blind residual sulphide surfaces.

The key innovations in CESL's process are:

- Maintenance of a moderate (~12g/L) chloride content in POx solution
- Control of feed acid content

These two features lead to:

- Much faster and more complete copper dissolution when compared with a pure sulphate electrolyte
- Generation of autoclave PLS at pH 2.5 (ideal for SX)
- Basic copper sulphate as a major secondary copper product which usefully consumes free acid from part of the primary PLS SX raffinate

Chalcopyrite oxidation is typically >96% complete, mainly via:

$$2CuFeS_2 + 2.5O_2 + 2H_2SO_4 \rightarrow 2CuSO_4 \cdot aq + Fe_2O_3 + 2H_2O + 4S^\circ$$

$$2CuFeS_2 + 2.5O_2 + H_2O \rightarrow CuSO_4 \cdot aq + CuSO_4 \cdot 2Cu(OH)_2 + Fe_2O_3 + 4S^\circ$$

Pyrite oxidation is typically incomplete with varying amounts of sulphide being oxidized to S° and SO_4^{-2}.The degree of pyrite oxidation is reported in vague terms and described as ore-specific.

Overall sulphide → sulphate conversion is reportedly controllable to ≤15% for most concentrates.

With reference to Figure 2, main elements of the process sequence are:

- Concentrate grind to "about" $P_{95} = 44\ \mu m$
- Oxygen POx with solution recycle and acid/salt make-up to specified conditions
- Primary liquid-solid separation of leach residue from high-grade PLS, part of which is recycled to POx

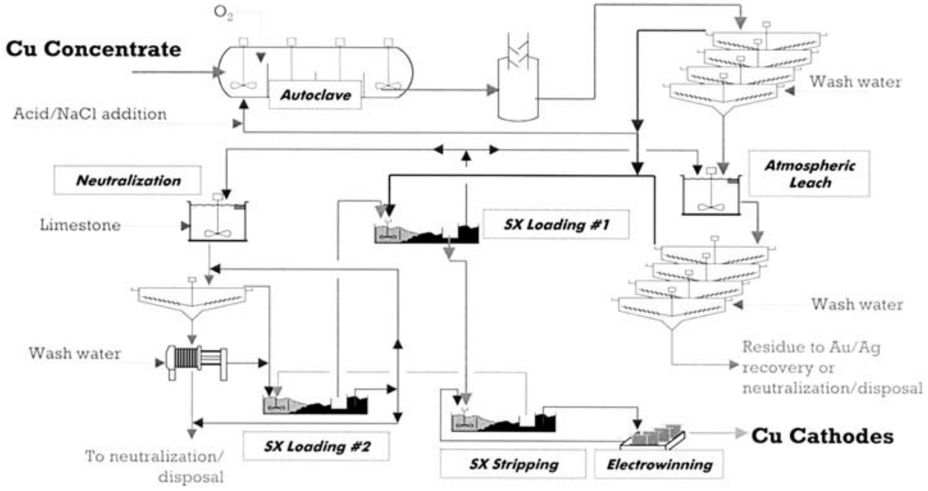

FIGURE 2 CESL's chloride enhanced low severity pressure oxidation process

- Acid leaching and washing of primary residue
- PM recovery (by unspecified method(s), if warranted) from residue prior to stabilization and disposal
- SX recovery of copper using a series configuration which facilitates nearly complete copper extraction from partially neutralized bleed solution before its final neutralization and disposal
- Stripping and EW to produce high quality copper cathode

The authors are not aware of CESL publication(s) on PM recovery technology, costs or performance. However, considering that the secondary leach residue contains a high percentage of S^o, it is safe to infer that PM recovery will be neither as conventional nor as simple as for the Placer process.

Dynatec—Surfactant (Coal) Enhanced Low-Severity POx. Dynatec's process appears to also have roots in zinc pressure leach technology. However, the lignosulphonate and quebracho type surfactants which effectively disperse liquid S° in zinc POx are rapidly degraded in the more aggressive Cu^{+2} $-Fe^{+3}$ $-O_2$ environment of copper POx.

Dynatec has developed technology to use coal at a modest dosage (25Kg/tonne of concentrate) as a more robust "dispersant" to prevent blinding and agglomeration of sulphides by liquid S^o product.

Sulphide oxidation chemistry is similar to that presented for the CESL process with, perhaps, a slightly higher (20% reported) degree of sulphide → sulphate oxidation.

Major unit operations (see Figure 3) include:

- Fine grinding of concentrate (P_{90} ~25 µm)
- POx of fresh feed with coal plus recycled sulphides using copper EW solution (or raffinate) as the aqueous phase

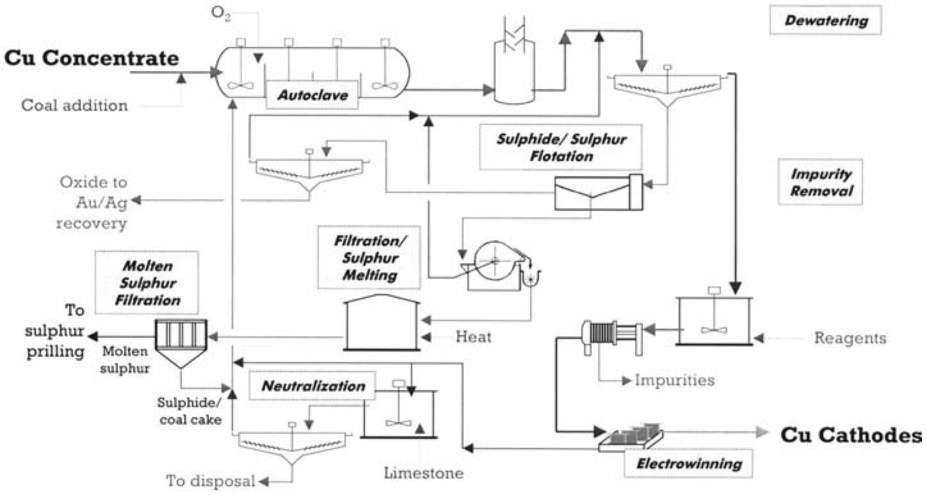

FIGURE 3 Dynatec's surfactant enhanced low severity pressure oxidation process

- Thickening of "conditioned" autoclave product, with underflow passing to sulphur/sulphide flotation
- Filtration of flotation concentrate followed by melting and filtration of sulphur-sulphide residual-coal mixture to produce molten sulphur filtrate and a sulphide-coal cake for recycle to POx
- Washing of flotation tailing (oxide/insoluble residue) followed by PM recovery if warranted and stabilization/disposal
- Copper recovery from PLS by some combination of direct EW and SX-EW
- Bleed stream treatment and waste disposal

Although cyanide consumption data are not reported, Dynatec cites cyanide amenability and carbon in leach testwork in reference to gold-silver recoveries in the 88–91% and 91–96% ranges, respectively, which infers that PM recovery is conventional.[8]

Melt-filtration of sulphur is an extra unit operation (compared to Placer and CESL processes). It can be described as conventional but also as energy intensive and high maintenance. Molten sulphur filter cake will presumably be converted to typical commercial (prill or slate) form but is unlikely to meet commercial purity criteria for numerous parameters including selenium, organic matter, heavy metals and ash content. At best, it might be a give-away product. Otherwise, it may have to be stowed on site in "block" form in the same manner as done at oil refineries.

It can be inferred from the selection of direct EW that high PLS copper grades are practical. However, since the Dynatec process does not incorporate basic copper sulphate coproduction, as the CESL process does, it can also be inferred that hydraulic loading on SX would be intermediate between that of CESL and Placer processes.

Penalty Elements

All processes provide for acceptable rejection of major penalty elements such as arsenic and antimony, provided that Fe:As + Sb ratios are acceptably high.

Cathode Copper Quality

Placer and CESL processes utilize modest variants on commercially proven SX-EW technology which, given good operation, should guarantee acceptable cathode product.

In the absence of documentation, Dynatec's solution purification/direct EW technology must be regarded as unproven.

Copper Recovery

High leach efficiencies and copper recoveries via SX to EW are practical for CESL and Placer processes. From the very limited published information, it is difficult to see how copper will be effectively recovered from Dynatec bleed solution without at least some use of SX technology.

Grinding

Differences in concentrate regrinding specifications are not considered to represent a significant cost variance among the processes.

Pressure Oxidation

The principal difference in pressure oxidation costs between high and low severity processes arises from the higher consumption of oxygen in total oxidation. For chalcopyrite, total oxidation consumes 2.14 Kg of oxygen per Kg of copper versus about 1 Kg/Kg for low severity (20% $S \rightarrow SO_4$) POx.

For a concentrate containing pyrite and chalcopyrite at a 1:1 molar ratio (21% copper on a sulphide-only basis), total POx oxygen consumption balloons to 4 Kg/Kg, whereas, low severity POx oxygen consumption rises modestly to 1.3 Kg/Kg at 20% $S \rightarrow SO_4$.

Assuming a low power cost scenario, and correspondingly low oxygen cost of $33/tonne, shifting to the conventional cost comparison mode of $US/lb copper indicates the following:

	Oxygen Cost, $US/lb Cu	
	Total POx	Selective POx
Chalcopyrite	0.032	0.010
1:1 Chalcopyrite:Pyrite	0.06	0.013

Obviously, in a scenario with relatively high power and oxygen costs (say $66/tonne oxygen), the oxygen cost for total POx of the pyritic concentrate ($0.12/lb Cu or $52.80/tonne concentrate) becomes the dominant operating cost (see "Operating Cost Comparison") and negates a significant fraction of the cost-incentive for on-site processing. High gold content in pyrite would be necessary to justify total POx in this scenario.

Placer has recently reported 98% copper recovery from a continuous pilot campaign at 30 minutes retention time.[7] CESL reports (very limited) batch test data for 1 h retention time, while Dynatec reports 4 h batch retention. Even taking account of the higher vessel pressure and pumping requirements for total POx, these data suggest a significant, but difficult to quantify, capital cost advantage for total POx compared to CESL or, especially, Dynatec.

Copper SX-EW

Due to the necessary dilution of free acid in total POx PLS, it is estimated that the hydraulic loading to SX will be about twice that for the low severity processes. Since mixer settlers are inexpensive on a unit volume basis, the capital cost disadvantage to total POx is not severe. However, organic losses will be roughly proportional to aqueous PLS flow, thereby imposing an operating cost penalty in the range of $0.005/lb Cu on total POx.

Neutralization/Waste Management

Placer publications account explicitly for neutralization costs; whereas, in the authors' opinion, these costs are neglected in documentation of low severity processes.

In an era when acid drainage is almost universally recognized as a potential problem for mining industry residues, disposal of sulphur and unreacted pyrite without equivalent, effective calcium alkalinity would not be permitted except in the context of elaborate (i.e., hazardous waste type) containment and owners' retention of liability, probably in the form of a large closure bond. Thus, neutralization is not necessarily a significant cost differentiator.

In the site-specific case of an acid consuming heap or dump leach, total POx could generate an acid credit of up to $40/tonne or $0.09/lb copper if the alternative acid supply were, for example, market acid supplied via long-distance truck haulage. In such a situation, sulphur credits would also be applicable to low severity processes but would be discounted for costs of converting sulphur and pyrite to useful acid.

Materials of Construction

The Placer and Dynatec processes utilize sulphate solutions which, outside of the autoclave, permit extensive use of stainless steels in wetted metal applications. CESL's (approx) 12 g/L of chloride will require more exotic alloys and/or lined or coated equipment with correspondingly higher (but difficult to quantify in a generic comparison) capital and maintenance costs.

PROVEN STATUS OF PROCESSES

Based on published data and relationship of unit operations to commercial practice, the authors' subjective ranking of the "proven status" of copper POx technologies is:

Placer >> CESL >>> Dynatec

The Placer process has had much less continuous pilot demonstration time, but this is considered to be more than compensated for by extensive commercial operation of high severity POx processes treating refractory gold ores and concentrates. This provides solid assurance of precious metal recovery (and cost estimates) whereas PM recovery from the CESL process residues is not yet documented in the technical literature.

The very small amount of published information is the main basis for the relatively low "proven" coefficient of the Dynatec process.

CAPITAL COST COMPARISON

A recent Placer presentation included a preliminary capital cost estimate of $US 248 million for a facility to produce 100,000 mtpy of cathode copper from a 20% Cu concentrate at an unspecified minesite.[7] At (about) $US 2,500/annual tonne of capacity, this is in line with CESL's published "target" of $2,000–2,500/annual tonne, and competitive with Noranda's[14] order of magnitude capital cost ($2,500–3,000/annual tonne) for the minimum (200,000 mtpy) scale considered to be viable for a new greenfields smelter.

No capital cost data are available for Dynatec technology, but the extra unit operation (sulphur melt-filtration) and longer autoclave retention time do not support much hope for lower costs.

OPERATING COST COMPARISON

Using Placer's recently presented (site unspecified) operating cost estimate[7] as a basis, high and low severity POx-SX-EW process operating costs can be compared as follows:

	$US/lb Copper		
	High Severity Total POx	Low Severity POx*	
Oxygen	0.041	0.012	
Neutralization (mill tailing)	0.006	0.006	
Grinding & Autoclave Agitation	0.018	0.018	(less mass transfer but more agitators)
Maintenance Supplies	0.013	0.019	(scaled)
Salaries/Labour	0.006	0.006	
Total Leach	0.084	0.061	
Total SX-EW	0.109	0.103	(lower reagent loss and mixing)
Waste Management	0.02	0.02	
Cathode Freight & Marketing	0.03	0.03	
TOTAL	0.243	0.214	

*Based on CESL, estimate 10–20% higher for Dynatec.

Thus, for an incremental operating cost of $0.03/lb copper or $13.20/tonne of 20% concentrate (a difference that is well within the probable variance of these preliminary estimates), a potential operator of the total POx process could have the comfort of more proven technology and assured gold-silver recovery.

Recovery of an extra 1.3 grams of gold from fully reacted pyrite could also pay for the differential cost.

Conversely, significantly higher total POx oxygen unit cost or consumption could skew the comparison in favour of low severity POx. In this regard, it may be of interest to note that a 1994 project specific total oxidation study involving the authors of this paper[12] generated an all-in POx, SX-EW, neutralization and disposal cost of $US 0.42/lb Cu for a prospect with both relatively high power cost and oxygen consumption. This was expected to be mitigated by significant gold recovery, avoided cost of a bulk port and improved socioeconomic (i.e., political) impact of extra employment. From the lack of subsequent action, it would appear that the high operating cost was unacceptable.

OVERALL ECONOMIC VIABILITY

Again drawing on Placer's recent presentation[7] their summarized venture analysis for 20% Cu concentrate indicates:

Capital Cost	$US 248 million
Margin between transport and smelter charges and total POx opcost	$US 0.175/lb Cu or $US 38.6 million/year
Payback Period @ 6% interest	8.0 years

Considering the security against extra smelter charges due to (hopefully) copper prices above the $0.80/lb base and escalation of freight, these data indicate a potentially viable venture.

CONCLUSIONS

Potential viability of hydrometallurgical (POx) technology versus toll smelting for new copper projects remains dependent on savings in transport cost and availability of relatively low cost power at the minesite. However, application of a combination of industrially proven processes (i.e., the Placer technology) clearly indicates that appropriately qualified new mines may have a real, cost-effective alternative to contract smelting of their product.

Subject to further confirmation of costs, the low severity processes may offer equivalent or greater savings for appropriate concentrates.

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REFERENCES

1. M.C. Kuhn, N. Arbiter and H. Kling, "Anaconda's Arbiter Process for Copper," CIM Bull., 67(752), (1074), 62–73.
2. W.P.C. Devuysteyn and B.J. Sabacky, "The Escondida Process for Copper Concentrates," *The Paul E. Queneau International Symposium, Extractive Metallurgy of Copper, Nickel and Cobalt, Volume I: Fundamental Aspects*, Edited by R.G. Reddy and R.N. Weisenbach, The Minerals, Metals & Materials Society, 1993.
3. F.W. Schweitzer and R. Livingstone, "Duval's CLEAR Hydrometallurgical Process," paper presented at the AIME Annual Meeting, Texas (February 14–18, 1982).
4. J.B. Ackerman, C.G. Anderson, S.M. Nordwick and L.E. Krys, 1993, "Hydrometallurgy at the Sunshine Metallurgical Complex," *Proceedings of the Fourth International Symposium on Hydrometallurgy*, TMS-SME, Salt Lake City, Utah, August.
5. J.A. King, D.B. Dreisinger and D.A. Knight, "The Total Pressure Oxidation of Copper Concentrates," presented at the CIMM District 6 Meeting, Vancouver, B.C., March 1994.
6. D. Dreisinger and J. King, "Autoclaving of Copper Concentrates," *Proceedings of Copper '95 International, Vol. III, Electrorefining and Hydrometallurgy of Copper.*
7. D. Smith, "The Economics of Autoclaving Chalcopyrite," *Proceedings of New Dimensions in Hydrometallurgy*, October 2, 1998, Fluor Daniel Wright Ltd. (prepublication).
8. K.R. Buban and M.J. Collins, "Acid Pressure Leaching of Chalcopyrite by Dynatec," *Randol Copper Hydrometallurgy Round Table*, Vancouver, B.C., Canada, November 1996.
9. M.J. Collins and D.K. Kofluk, "Hydrometallurgical Process for the Extraction of Copper from Sulphidic Concentrates," U.S. Patent No. 5,730,776, March 24, 1998.
10. D.L. Jones, "CESL Copper Process," *Randol Copper Hydrometallurgy Round Table*, Vancouver, B.C., Canada, November 1996.
11. D.L. Jones, J. Hestrin and R. Moore, "CESL Process for Nickel-Cobalt-Copper Sulphides, Testing in an Integrated Pilot Plant," *Alta '98 Nickel Cobalt Leaching and Hydrometallurgy Forum*, Perth, Australia, May 1998.
12. Anon., Fluor Daniel Wright Ltd., "Preliminary Capital and Operating Cost Estimates for a 200 mtpd Copper Concentrate Pressure Leach Plant and Auxiliary Facilities," Project Report, December 1994.
13. D. Dreisinger, "Autoclave Options for Copper Ores and Concentrates," *Proceedings of New Dimensions in Hydrometallurgy*, October 2, 1998, Fluor Daniel Wright Ltd. (prepublication).
14. J. Peacey, "Economic Comparison Between Pyrometallurgical and Hydrometallurgical Copper Processes," *Proceedings of New Dimensions in Hydrometallurgy*, October 2, 1998, Fluor Daniel Wright Ltd. (prepublication).

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Electrolytic Copper Electrowinning and Solvent Extraction— World Operating Data

Brian Kennedy,[*] William G. Davenport,[†] Jackson Jenkins,[‡] and Tim Robinson[**]

The first survey of world copper electrowinning tankhouse and solvent extraction practices has been carried out. The following pages list the results of over 25 plants around the world. Significant trends are shown to be the installation of stainless steel cathodes, polymer concrete cells and enhanced mixing technologies. The overall consequences of these changes have been improved cathode purities and enhanced SX-EW plant productivities. These published results are from the 1995 survey. The 1998 results will be presented during the SME conference in Denver in March 1999.

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[*] Senior Metallurgical Engineer, Simons Engineering, Inc., Phoenix, Ariz.

[†] Department of Materials Science and Engineering, University of Arizona, Tucson, Ariz.

[‡] Manager Technical Services, Cyprus Sierrita Corporation, Green Valley, Ariz.

[**] CTI Ancor, Dallas, Texas.

TABLE 1A Solvent Extraction Plant Data

	Aberfoyle Resources, Ltd., Gunpowder Division, Queensland, Australia	Girilambone Copper Co., Girilambone, NSW, Australia	Great Australia Mining Co., Cloncurry, Queensland, Australia	Mount Isa Mines Ltd., Mount Isa, Queensland, Australia	Pasminco Metals-BHAS, Port Pirie, South Australia	Compania Minera Carmen de Andacollo, La Serena, Chile
Date of commissioning	1990	1993	1996	1990		1996
Cathode production rate, tonnes per year	13,000	15,600	5,500	3,500	5,000	20,000
number of solvent extraction trains	1	1	1	1	1	1
number of extraction mixer-settlers per train	2	3	2	2	2	2
series or parallel	series	series/parallel	parallel	series	series	2 x 1 series
number of strip mixer-settlers per train	2	1	1	1	2	1
series or parallel	series				series	
washing of organic	organic is washed with slightly acidic sol'n to remove Fe	no	no	no	1 mixer-settler wash with 35 to 40gpl acid	no
FLOWRATES						
pregnant leach solution per train, m^3/h	250	0 to 870	450 (2 x 225)	185	320	827
electrolyte per train, m^3/h	65	0 to 150	44	28	520	166
percent of total electrolyte flow going to SX	26	20				700 recirculation, 166 to scavenger cells
organic between extraction and strip, per train, m^3/h	250	0 to 470	245	205	1,300	827
aqueous-continuous	wash				all	E1
organic continuous	all others		all (to minimize crud)	all (to minimize crud)		E2 and S (design)
PREGNANT LEACH SOLUTION (gpl)						
Cu	4	1.5 to 3	2	2.5 to 3.5	41	3.13
pH	1.9 to 2	1.9 to 2.1	2.4	2.0 to 2.4		1.7 to 2.1
Fe (total)	9	11	0.3	2.5 to 5.0	0.06	3 to 10

TABLE 1A Solvent Extraction Plant Data (continued)

	Aberfoyle Resources, Ltd., Gunpowder Division, Queensland, Australia	Girilambone Copper Co., Girilambone, NSW, Australia	Great Australia Mining Co., Cloncurry, Queensland, Australia	Mount Isa Mines Ltd., Mount Isa, Queensland, Australia	Pasminco Metals-BHAS, Port Pirie, South Australia	Compania Minera Carmen de Andacollo, La Serena, Chile
Fe+++	8	0.4	0.28	1		2 to 7
sulfuric acid	8	2 to 3.5	1.5		20	2.5
Mn	0.28	15 to 20	0.65		0.1	0.2 to 1.2
Cl^-, ppm	8	1,500	150		18 gpl	400 to 600
solids, ppm	20 to 50	50			silica (removed with Polyox coagulant)	<20
RAFFINATE (gpl)						
Cu	0.2	0.1 to 0.5	0.25	0.1	11	0.24
sulfuric acid	8 to 10	4 to 7	3.5		65	7 to 8
pH	1.5	1.5	1.5	1.6 to 1.8		0.8
organic loss to raffinate, ppm	100	10 to 30	20			50–75
organic-from-raffinate recovery method	skimmed from raffinate pond with sobar booms, recovered organic is washed with raffinate	coalescing media at pond entrance and an organic recovery system				Spintek pack in loaded organic tank
extraction efficiency per train, %	E1: 50 to 60 E2: 75 to 80%	90	87			92
extraction mixer retention time, minutes	2.5	2	2	2		2.5
ORGANIC						
Cu concentration in loaded organic, gpl		7 to 9	6.5		24	3 to 5
Cu concentration in barren organic, gpl		2.5 to 3	3			1 to 2
organic depth in settlers, mm		extraction: 180 to 200 strip: 200	400			200 to 300
aqueous depth in settlers, mm		580	500			600
extractant name		LIX 984N	Acorga M5640	LIX 984	Acorga M5640	LIX 964N C

TABLE 1A Solvent Extraction Plant Data (continued)

	Aberfoyle Resources, Ltd., Gunpowder Division, Queensland, Australia	Girilambone Copper Co., Girilambone, NSW, Australia	Great Australia Mining Co., Cloncurry, Queensland, Australia	Mount Isa Mines Ltd., Mount Isa, Queensland, Australia	Pasminco Metals-BHAS, Port Pirie, South Australia	Compania Minera Carmen de Andacollo, La Serena, Chile
solvent name		Shellsol	Shellsol 2046	Shellsol 2046	Shellsol	Escaid 103
extractant concentration in organic, %		13 to 18 (now, 16.8)	13 to 14	5 to 10	20	10 to 12
net copper transference, gpl per % extractant		0.29				0.26 to 0.28
stripping efficiency per train, %		90	95			95
strip mixer retention time, minutes		2.33	2	2		2
phase disengagement time, minutes		1.4	0.5	0.5 to 0.85	<0.75	very rapid, new circuit
organic coalescer		coalescing media in a loaded organic tank which overflows to a second loaded organic tank	none	none		2 hour holding pond with baffles
organic filters		plate and frame filter	no	no	none	Eimco atmospheric dual media filter
crud removal system	diaphragm pumping to E2 picket fence	crud is sucked through a clear pipe from organic/aqueous interface to a crud tank	run organic on floor, crud remains of floor, collect clean organic	sucked from settlers using a portable air diaphragm pump	settler is taken offline and crud is physically drained	by hand or with diaphragm pump
crud treatment system	trying plate and frame filter/clay addition system	clay treatment followed by plate and frame filtration	none	settled in tank unsettled crud is treated with clay	none	holding tank & P&F press with agitated tank for clay & crud treatment
organic cleaning technique	no	filter press	none	none	bentonite clay treatment then filtering	in future
RICH ELECTROLYTE						
Cu, gpl	45 to 50	50	55 to 60	36	46	50
sulfuric acid, gpl	165	170 to 190	150 to 160	175	140	16
organic loss in electrolyte, ppm	<1	<5	50 to 100		8	<50 at S1, <5 at filter effluent

TABLE 1A Solvent Extraction Plant Data (continued)

	Aberfoyle Resources, Ltd., Gunpowder Division, Queensland, Australia	Girilambone Copper Co., Girilambone, NSW, Australia	Great Australia Mining Co., Cloncurry, Queensland, Australia	Mount Isa Mines Ltd., Mount Isa, Queensland, Australia	Pasminco Metals-BHAS, Port Pirie, South Australia	Compania Minera Carmen de Andacollo, La Serena, Chile
electrolyte coalescer		none	none	none		none
electrolyte treatment before tankhouse		Jameson cell followed by Spintek multimedia filters	Spintek dual media	Jameson cell and garnet/anthracite filter	none	Eimco atmospheric filter and Cominco flotation column
SPENT ELECTROLYTE						
Cu, gpl		35 to 40	33 to 38	32	31	35
sulfuric acid, gpl		170 to 190	175 to 185	180	175	170
cobalt, ppm		100 to 120	100 to 150		80	100
DESIGN AND BUILDING MATERIALS						
mixers	316L stainless steel	fiberglass reinforced plastic	fiberglass	fiberglass	mild steel, fiberglass lined	2205 stainless steel
settlers	316L stainless steel	fiberglass reinforced plastic	fiberglass	fiberglass	mild steel, flake lined	HDPE lined
piping	316L stainless steel, HDPE and UPVC	HDPE	HDPE	HDPE		HDPE, PVC & St. steel
tanks		fiberglass reinforced plastic	fiberglass and HDPE	fiberglass/HDPE/ stainless steel		FRP
gratings	fiberglass	FRP	wood	concrete slab		FRP
SETTLERS						
settler dimensions, length x width x depth	18.1 x 6.1 x 1.2	14.55 x 12.3	11 x 5.5 x 1	Krebs 10 x 4 x 1	7.8 x 4.1 x 0.99	21.6 x 21.6 x 1.7
number of picket fences, per settler	1	1	1		1 + 5 up-&-over fences	2
settling velocity, m^3/m^2 per hour	5	5	5.5 to 6.0	4.5		4
MIXERS						
extraction mixers	primary: pump box secondary: axial flow pump (only primary is used)	axial flow pump mixer		Krebs	conventional	Lightnin R320 (primary) 2 secondary with 3 radial blades up, pumping (all variable speed) all used

TABLE 1A Solvent Extraction Plant Data (continued)

	Aberfoyle Resources, Ltd., Gunpowder Division, Queensland, Australia	Girilambone Copper Co., Girilambone, NSW, Australia	Great Australia Mining Co., Cloncurry, Queensland, Australia	Mount Isa Mines Ltd., Mount Isa, Queensland, Australia	Pasminco Metals-BHAS, Port Pirie, South Australia	Compania Minera Carmen de Andacollo, La Serena, Chile
principal extraction mixer	29 kW, 50 RPM	15 kW, 75 RPM				37 kW, 60 to 80 RPM
impeller diameter, m	1.7	1.5				1.6 (305 m/min maximum tip speed)
stripping mixers	primary: pump box secondary: axial flow pump (only primary is used)	axial flow pump mixer		Krebs	conventional	Lightnin R32 (primary), 1 to 3 radial blade (secondary), (all variable speed) all used
principal strip mixer	25 kW, 54 RPM	15 kW, 75 RPM				37 kW, 60 to 80 RPM
impeller diameter, m	1.6	1.5				1.6

TABLE 1B Solvent Extraction Plant Data

	Compania Minera Cerro Colorado Ltda., Cerro Colorado, Chile	CODELCO Chile, El Teniente Division, Colon-Rancagua, Chile	Sociedad Contractual Minera El Abra, El Loa, Calama, Chile	Empresa Minera de Mantos Blancos, Mantos Blancos Mine, Chile	Empresa Minera de Mantos Blancos, Mantoverde Mine, Chile	Compania Minera Quebrada Blanca, Quebrada Blanca, Chile
Date of commissioning		1985	1996	1995	1995	1994
Cathode production rate, tonnes per year	60,000	3,000 to 7,500	225,000 (nominal)	45,000	42,000	75,000
number of solvent extraction trains	3	1	4	1	1	3
number of **extraction** mixer-settlers per train	2	2	2	2	2	2
series or parallel	series	series	series	series	series	series
number of **strip** mixer-settlers per train	1	1	2	2	2	1
series or parallel			series	series	series	
washing of organic	no	no	one stage of organic wash	washing stage for loaded organic	washing stage for loaded organic	no
FLOWRATES						
pregnant leach solution per train, m^3/h	834	612	1,200 (nominal)	800	1,000	1,000

TABLE 1B Solvent Extraction Plant Data (continued)

	Compania Minera Cerro Colorado Ltda., Cerro Colorado, Chile	CODELCO Chile, El Teniente Division, Colon-Rancagua, Chile	Sociedad Contractual Minera El Abra, El Loa, Calama, Chile	Empresa Minera de Mantos Blancos, Mantos Blancos Mine, Chile	Empresa Minera de Mantos Blancos, Mantoverde Mine, Chile	Compania Minera Quebrada Blanca, Quebrada Blanca, Chile
electrolyte per train, m^3/h	169	40	1,100	390	340	240
percent of total electrolyte flow going to SX		13	33			33
organic between extraction and strip, per train, m^3/h	1,000	490	1,400	810	1,050	1,000
aqueous-continuous		all	E1, S2, wash	washing stage	washing stage	E1
organic continuous	all		E2, S1	all others	all others	E2 and S
PREGNANT LEACH SOLUTION (gpl)						
Cu	3.38	0.8 to 1.8	6.25	7	5.5	3
pH	1.5	2.5 to 3.0	1.8	1.4	1.4	1.9
Fe (total)	5.4	0.12 to 0.5	3.26	8.5	12.5	3
Fe+++	1.29	0.1 to 0.3	1.81	0.5	5 to 6	1
sulfuric acid	5	0.04	3.5	6	5 to 8	2
Mn	0.1	0.03	0.2		1.0 to 1.5	0.09
Cl^-, ppm	106	43	1130		500	30
solids, ppm		25	10	80	35 to 40	10
RAFFINATE (gpl)						
Cu	0.34	0.02 to 0.08	0.69	0.5	0.5	0.2
sulfuric acid	9.7	1.2 to 1.5	11.1	16	15 to 16	7.5
pH	1.27	1.5 to 2.0	1.24	0.8	1.1	1.4
organic loss to raffinate, ppm	10	60 to 120	60 to 100	7	80	35
organic-from-raffinate recovery method	no	decanting (after settler)	Pacesetter coalescers			organic which floats to top of raffinate pond is recovered
extraction efficiency per train, %	90	95 to 99	89	93	93	93

TABLE 1B Solvent Extraction Plant Data (continued)

	Compania Minera Cerro Colorado Ltda., Cerro Colorado, Chile	CODELCO Chile, El Teniente Division, Colon-Rancagua, Chile	Sociedad Contractual Minera El Abra, El Loa, Calama, Chile	Empresa Minera de Mantos Blancos, Mantos Blancos Mine, Chile	Empresa Minera de Mantos Blancos, Mantoverde Mine, Chile	Compania Minera Quebrada Blanca, Quebrada Blanca, Chile
extraction mixer retention time, minutes	3.4	3	3	4	3	three mix stages, one minute each (last stage not used)
ORGANIC			PT-5050 MOC 80TD			
Cu concentration in loaded organic, gpl	4.9	2.41	9.5 8.7	9.2	8.5	5.5
Cu concentration in barren organic, gpl	2.35	0.85	3.8 2.2	2.7	3.3	2.4
organic depth in settlers, mm	280	100 to 170 in E, 100 to 120 in S	300 in E, 250 in S	250	320	275
aqueous depth in settlers, mm	420	630 to 700 in E, 680 to 700 in S		620	550	525
extractant name	LIX 84N, LIX 860N	LIX 984	Zeneca PT 5050, Allied MOC 80 TD	Acorga M-5640	Acorga M-5640	LIX 984
solvent name	Solmin 200L	Escaid 103	Phillips Orfam SX12	Escaid 103	Escaid 103	Orform SX 12
extractant concentration in organic, %	12	5.5	PT 30 MOC 18	20	17	11 to 13
net copper transference, gpl per % extractant	0.213	3.0 to 3.2	[PT 0.285 MOC 0.361]	0.325	0.306	0.24 to 0.28
stripping efficiency per train, %	52	90 to 95	A,B 60; C,D 66.5		61	60
strip mixer retention time, minutes	3.4	3	2	2.66	3	2 (2 mix stages, 1 min each)
phase disengagement time, minutes	2	0.75	1 to 1.5	0.75	0.83	2
organic coalescer	none	picket fence	Pacesetter	Four coalescers—2 before washing stage, 2 after washing stage, to recover aqueous	2 coalescers, CODELCO design, to recover aqueous	none
organic filters	no	no	no	no	no	no

TABLE 1B Solvent Extraction Plant Data (continued)

	Compania Minera Cerro Colorado Ltda., Cerro Colorado, Chile	CODELCO Chile, El Teniente Division, Colon-Rancagua, Chile	Sociedad Contractual Minera El Abra, El Loa, Calama, Chile	Empresa Minera de Mantos Blancos, Mantos Blancos Mine, Chile	Empresa Minera de Mantos Blancos, Mantoverde Mine, Chile	Compania Minera Quebrada Blanca, Quebrada Blanca, Chile
crud removal system	diaphragm pneumatic pump	suction by diaphragm pump	diaphragm pump that moves back and forth across back of organic weir	pumping from setter to a launder	pumping from setter to a launder	batch diaphragm pumping
crud treatment system	mechanical rupture tank	pressure filter	conical bottom tanks with mixers	CODELCO/CIMM method	CODELCO/CIMM method	collection tank, centrifuge, solids storage
organic cleaning technique	sparkler filter with activated clay (not used yet)	diatomaceous earth	vertical plate clay filter (used only if needed)	clay treatment	clay treatment	organic is mixed with clay then pressure filtered in Sparkler filter
RICH ELECTROLYTE						
Cu, gpl	50	40 to 47	50	49	45	48
sulfuric acid, gpl	157	140 to 150	150	165	150	150
organic loss in electrolyte, ppm	20	200 to 400	10 to 20	20	20	40
electrolyte coalescer	none	picket fence	Pacesetter	none	none	none
electrolyte treatment before tankhouse	4 Spintek filters, garnet/anthracite	multimedia filter, anthracite, 0.8 m^3, garnet, 1.6 m^3, quartz, 0.54 m^3	Spintek garnet/ anthracite dual media filter	sand filter and Magma column flotation	sand filter and Magma column flotation	2 parallel column flotation cells followed by 4 garnet/anthracite filters
SPENT ELECTROLYTE						
Cu, gpl	35	34 to 38	35	36	30	36
sulfuric acid, gpl	180	155 to 170	165	185	175	170
cobalt, ppm	120	120	120	120	120	100
DESIGN AND BUILDING MATERIALS						
mixers	concrete lined with 316 stainless steel	stainless steel	polymer concrete	concrete lined with FRP	concrete lined with FRP	316L stainless steel
settlers	concrete lined with HDPE	stainless steel	concrete lined with HDPE	concrete lined with HDPE	concrete lined with HDPE	concrete, HDPE lined
piping	HDPE and 316 st.st.	stainless steel/high density PVC	stainless steel, HDPE	HDPE	HDPE	316L and HDPE

TABLE 1B Solvent Extraction Plant Data (continued)

	Compania Minera Cerro Colorado Ltda., Cerro Colorado, Chile	CODELCO Chile, El Teniente Division, Colon-Rancagua, Chile	Sociedad Contractual Minera El Abra, El Loa, Calama, Chile	Empresa Minera de Mantos Blancos, Mantos Blancos Mine, Chile	Empresa Minera de Mantos Blancos, Mantoverde Mine, Chile	Compania Minera Quebrada Blanca, Quebrada Blanca, Chile
tanks	concrete lined with HDPE	stainless steel	fiberglass, stainless steel, HDPE-lined concrete, mild steel	concrete lined with HDPE	concrete lined with HDPE	316L and FRP
gratings	FRP	hormigon sulfurado	Fibergrating	FRP	FRP	FRP
SETTLERS						
settler dimensions, length x width x depth	21 x 22 x 0.91	360 m^3	26.1 x 27.9 x 0.87	27 x 25 x 1.1	27 x 25 x 1.1	23.3 x 24.5 x 0.8
number of picket fences, per settler	2	1 at mixer exit	2	3	3	2
settling velocity, m^3/m^2 per hour	4.4 at maximum flow	620 to 650 m^3/h		3.5	3.4	4
MIXERS						
extraction mixers	pump mixer (auxiliary mixers not operated)	suction and contact impeller plus secondary mixer (all used)	principal mixer, one auxiliary in E2 stage (Lightnin)	Lightnin, curved blades (all used)	Lightnin, curved blades (all used)	Lightnin, R301 (37 kW) & A310 (5 kW), both normally used
principal extraction mixer	60 kW, 13 to 50 RPM (normal, 46)	29 kW, 29 RPM	50 RPM	19 kW, 50 RPM	19 kW, 50 RPM	37 kW, 48 RPM
impeller diameter, m	2.22	2.33		1.85	1.85	2.3 (six blades)
stripping mixers	pump mixer (auxiliary mixers not operated)	suction and contact impeller (all used)	Lightnin (no auxiliary mixer)	Lightnin, curved blades (all used)	Lightnin, curved blades (all used)	Lightnin R301 & A310, both normally used
principal strip mixer	60 kW, 13 to 50 RPM (normal, 46)	29 kW, 29 RPM	S1, 50 RPM, S2, 45 RPM	19 kW, 45 RPM	19 kW, 40 RPM	37 kW, 48 RPM
impeller diameter, m	2.22	2.33		1.85	1.85	2.3 (six blades)

TABLE 1C Solvent Extraction Plant Data

	Compania Minera Zaldivar, Chile	Mexicana de Cobre, S.A. de C.V., Nacozari, Sonora, Mexico	Southern Peru Limited, Toquepala, Tacna, Peru	BHP Copper, Miami, Arizona	BHP Copper, Pinto Valley, Arizona	BHP Copper, San Manuel, Arizona
Date of commissioning	1995	1996	1995	1976	1981	1986
Cathode production rate, tonnes per year	125,000	22,000	39,000	11,800	9,500	55,000

TABLE 1C Solvent Extraction Plant Data (continued)

	Compania Minera Zaldivar, Chile	Mexicana de Cobre, S.A. de C.V., Nacozari, Sonora, Mexico	Southern Peru Limited, Toquepala, Tacna, Peru	BHP Copper, Miami, Arizona	BHP Copper, Pinto Valley, Arizona	BHP Copper, San Manuel, Arizona
number of solvent extraction trains	4	3	3	2	2	4
number of **extraction** mixer-settlers per train	2	2	2	3	2	2
series or parallel	series	series	series	2 series-1 parallel	series	series
number of **strip** mixer-settlers per train	1	2	1	2	1	1
series or parallel		series		series		series
washing of organic	one wash mixer-settler	no	no	no	no	no
FLOWRATES						
pregnant leach solution per train, m^3/h	1,000 to 1,100	630	930	780	820	1,020 to 1,140
electrolyte per train, m^3/h	270	80	92	52	27	90 to 140
percent of total electrolyte flow going to SX	25		23.77	10	10	25 (approx)
organic between extraction and strip, per train, m^3/h	1,000 to 1,100	630	930	380	770	840 to 900
aqueous-continuous	E1 and wash			E1,E2,S2	E1	E1
organic continuous	E2 and S	all	all	E1P,S1	E2, S	E2 and strip
PREGNANT LEACH SOLUTION (gpl)						
Cu	3.2	2.1	1.62	0.9	0.8	1
pH	1.8	2	1.9 to 2.2	1.7	2.1	2.2
Fe (total)	7	1	6	1.8	1.8	1.5
Fe^{+++}	2.5	0.85	5.8	0.9	1.7	0.9
sulfuric acid		0.5	3	2.5	0.9	15
Mn	0.7	0.1	0.4			0.6
Cl^-, ppm	3.5gpl		73			100

TABLE 1C Solvent Extraction Plant Data (continued)

	Compania Minera Zaldivar, Chile	Mexicana de Cobre, S.A. de C.V., Nacozari, Sonora, Mexico	Southern Peru Limited, Toquepala, Tacna, Peru	BHP Copper, Miami, Arizona	BHP Copper, Pinto Valley, Arizona	BHP Copper, San Manuel, Arizona
solids, ppm	30 (also PO_4^-, as 2 to 3 gpl P)	<20	20	<10		40 (also Al, 7gpl and Mg, 8 gpl)
RAFFINATE (gpl)						
Cu	0.2	0.3	0.16	0.06	0.07	0.03
sulfuric acid	7	7	5	5	2.6	17
pH		1.7	1.8	1.6	1.7	1.7
organic loss to raffinate, ppm	30	20	30	40	55	45
organic-from-raffinate recovery method				none		none
extraction efficiency per train, %	94	85	90	92	90	97
extraction mixer retention time, minutes	5	3	3	3	3	3
ORGANIC						
Cu concentration in loaded organic, gpl	7	3.12	2.66	2.9	1.3 to 1.5	2
Cu concentration in barren organic, gpl	3	1.21	1.14	1.2	0.3 to 0.4	0.8
organic depth in settlers, mm	280 to 300	254	292	extraction: 152 strip: 229	127 to 203	extraction: 305 strip: 254
aqueous depth in settlers, mm	400	670	419	extraction: 610 strip: 533	380	508
extractant name	LIX 984NC	LIX 622	LIX 984NC	LIX 984N	LIX 984N	LIX 984
solvent name	Shellsol SX 12	Conosol	Orfom SX-12	Phillips SX-7	Phillips SX-12	Phillips SX-7
extractant concentration in organic, %	13	7.3	6	9.5	3	4.5
net copper transference, gpl per % extractant		0.27 to 0.29	0.26	0.17	0.17	0.3

TABLE 1C Solvent Extraction Plant Data (continued)

	Compania Minera Zaldivar, Chile	Mexicana de Cobre, S.A. de C.V., Nacozari, Sonora, Mexico	Southern Peru Limited, Toquepala, Tacna, Peru	BHP Copper, Miami, Arizona	BHP Copper, Pinto Valley, Arizona	BHP Copper, San Manuel, Arizona
stripping efficiency per train, %	57	95%	92	95 to 99	96 to 99	
strip mixer retention time, minutes	2	1.5	2	3	3	3
phase disengagement time, minutes		0.75	1	<1	<1	0.9
organic coalescer		coalescer column	coalescing boxes with netted walls and HDPE scrap attached to picket fences in E1	none	none	none
organic filters	Disep 2 x 4		bag filter to remove clay from clay-treated organic	no	none	no
crud removal system	manual by centrifuge/ diaphragm pump	suction pump	crud is pumped from aqueous/organic interphase by ODS pumps of clay treatment	interface pumping with wands	external pumping (also set up for flooding from settlers, not now done)	flooding
crud treatment system		packed column	Asarco method. Crud is broken by agitation in organic continuous phase. Organic with over 120 s break time is clay treated with montmorillonite	centrifuge	centrifuge	cone tank
organic cleaning technique	clay treatment + centrifuge + clay filter	clay filter	drycake discharge filter is being tested	none	none	clay treat organic recovered from raffinate pond
RICH ELECTROLYTE						
Cu, gpl	48	45 to 50	45	49	50 to 53	50
sulfuric acid, gpl	156	155	149	165	180	160
organic loss in electrolyte, ppm	15	<10	<20	40	50	<20
electrolyte coalescer	one Comin coalescer per SX train	none	none	column cell packed with Tripak coalescing media	none	air flotation columns

TABLE 1C Solvent Extraction Plant Data (continued)

	Compania Minera Zaldivar, Chile	Mexicana de Cobre, S.A. de C.V., Nacozari, Sonora, Mexico	Southern Peru Limited, Toquepala, Tacna, Peru	BHP Copper, Miami, Arizona	BHP Copper, Pinto Valley, Arizona	BHP Copper, San Manuel, Arizona
electrolyte treatment before tankhouse	four Disep filters per 2 SX trains (8 total)	two Jameson cells in series	three dual media (anthracite/garnet) Spintek filters	garnet/anthracite dual media filter	dual media filter, garnet and anthracite	dual media, anthracite/garnet
SPENT ELECTROLYTE						
Cu, gpl	37.5	30 to 35	30	36	32	38
sulfuric acid, gpl	160	170	170	180	190	170
cobalt, ppm	140	120	110 to 120	120	120 to 150	65
DESIGN AND BUILDING MATERIALS						
mixers	Ti (grade 2) & 316L, 254SMO	stainless steel	316L stainless steel encased in concrete	316L stainless steel	concrete, lined with 316L stainless steel	316 stainless steel
settlers	concrete, HDPE-lined	concrete, polyethylene-lined	concrete, HDPE lined	316L stainless steel	concrete, lined with 316L stainless steel	316 stainless steel
piping	Ti (grade 2) & 316L stainless steel, HDPE	polyethylene	stainless steel and HDPE	316L stainless steel	316L stainless steel	316 stainless steel
tanks	316L	concrete, polyethylene-lined	stainless steel and HDPE-lined concrete	316L stainless steel	316L stainless steel	316 stainless steel
gratings	FRP	FRP	FRP	mild steel	fiberglass	fiberglass
SETTLERS						
settler dimensions, length x width x depth	23.5 x 25 x 0.9	24.5 x 16.9 x 1.25	20 x 24.4 x 1.01	eight: 18.7 x 12.2 x 0.64 two: 18.7 x 12.2 x 0.89	12 x 20.7 x 9.9	18.3 x 24.4
number of picket fences, per settler	2	3	2	1	1 stationary, 3 portable	2
settling velocity, m^3/m^2 per hour	4.2	3.5 at maximum flow	4.4	4.5	7.9	5

TABLE 1C Solvent Extraction Plant Data (continued)

	Compania Minera Zaldivar, Chile	Mexicana de Cobre, S.A. de C.V., Nacozari, Sonora, Mexico	Southern Peru Limited, Toquepala, Tacna, Peru	BHP Copper, Miami, Arizona	BHP Copper, Pinto Valley, Arizona	BHP Copper, San Manuel, Arizona
MIXERS						
extraction mixers	Outokumpu VSF mixers (all used)	primary: double shrouded variable speed 8 backswept blade pump mixer; secondary: variable pitch 3 radial blade axial flow turbine (all used)	pump mixer and one auxiliary are used	Nettco Mark II (auxiliary mixers are not used)	Philadelphia Terro 16 plus auxiliary mixers (used only in warm weather)	Lightnin top shrouded radial turbines with 6 flat blades (auxiliary 3 blade axial flow turbine is used)
principal extraction mixer	9 kW, 45 RPM	22.3 kW, 67 RPM(max)	19 kW, 37 RPM	7.6 kW 50 RPM	19 kW, 42 RPM	22 kW
impeller diameter, m	1.8	1.82	2.29	1.5	2.1	2.13
stripping mixers	Outokumpu VSF mixers (all used)	primary: double shrouded variable speed 8 back-swept blade pump mixer (all used)	pump mixer and auxiliary (both used)	Nettco Mark II (auxiliary mixers are not used)	Philadelphia Terro 16 plus auxiliary mixers (used only in warm weather)	Lightnin top shrouded radial turbines with 6 flat blades (auxiliary 3 blade axial flow turbine is not used)
principal strip mixer	10 kW, 50 RPM	22.3 kW, 67 RPM(max)	19 kW, 41 RPM	7.6 kW 50 RPM	19 kW, 42 RPM	22 kW
impeller diameter, m	1.8	1.82	2.29	1.5	2.1	2.13 primary; 1.7m aux.

TABLE 1D Solvent Extraction Plant Data

	Burro Chief (Phelps Dodge), Santa Rita, New Mexico	Burro Chief Copper Co. (Phelps Dodge), Tyrone New Mexico	Phelps Dodge Morenci, Central SX Plant, Morenci, Arizona	Phelps Dodge Morenci, Metcalf SX Plant, Morenci, Arizona	Phelps Dodge Morenci, Modoc SX Plant, Morenci, Arizona	Phelps Dodge Morenci, Southwest SX Plant, Morenci, Arizona	Zambia CCM Tailings Leach Plant, Chingola, Zambia
Date of commissioning	1988	1984	1987	1989	1992	1988 (closed July 1995)	1974
Cathode production rate, tonnes per year	62,000	68,000	47,000	94,000	94,000	8,000	110,000
number of solvent extraction trains	2	2	1	2	2	1	4
number of **extraction** mixer-settlers per train	2	1E x 2E	3	3	3	3	3

TABLE 1D Solvent Extraction Plant Data (continued)

	Burro Chief (Phelps Dodge), Santa Rita, New Mexico	Burro Chief Copper Co. (Phelps Dodge), Tyrone New Mexico	Phelps Dodge Morenci, Central SX Plant, Morenci, Arizona	Phelps Dodge Morenci, Metcalf SX Plant, Morenci, Arizona	Phelps Dodge Morenci, Modoc SX Plant, Morenci, Arizona	Phelps Dodge Morenci, Southwest SX Plant, Morenci, Arizona	Zambia CCM Tailings Leach Plant, Chingola, Zambia
series or parallel	series	parallel x series	1 x 2	1 x 2	1 x 2	1 x 2	2 series; 2 series/parallel
number of **strip** mixer-settlers per train	1	1	1	1	1	1	2
series or parallel							series
washing of organic	plan to install coalescer	no	no	no	no	no	no
FLOWRATES							
pregnant leach solution per train, m^3/h	1360	2725	2950	2700	3070	1360	800
electrolyte per train, m^3/h	250	250	390	480	550	70	200 (plus recycle)
percent of total electrolyte flow going to SX	26	16.4	16	39	45	3	100
organic between extraction and strip, per train, m^3/h	1250	1250	1520	1400	1600	900	880
aqueous-continuous	extraction	extraction	E1	E1	E1		Extraction
organic continuous	strip	strip	all other	all other	all other	all	Strip
PREGNANT LEACH SOLUTION (gpl)							
Cu	3.2	1.6	2.22	2.3	2.24	1.31	5
pH	2	1.9	1.51	1.63	1.66	1.91	1.8 to 2.0
Fe (total)	13.9	2.3	11.6	6.53	6.31	3.53	0.4
Fe^{+++}	5.8	2			6.14	3.52	
sulfuric acid	2.6	3	4.4		2.84		0.5 to 1
Mn	1.8	1.2	0.86	0.591	0.595	0.537	0.2
Cl^-, ppm			10	10	10	10	
solids, ppm	5		10 (also humic acid and other organic acids)	10 (also humic acid and other organic acids)	10 (also humic acid and other organic acids)	10	50

TABLE 1D Solvent Extraction Plant Data (continued)

	Burro Chief (Phelps Dodge), Santa Rita, New Mexico	Burro Chief Copper Co. (Phelps Dodge), Tyrone New Mexico	Phelps Dodge Morenci, Central SX Plant, Morenci, Arizona	Phelps Dodge Morenci, Metcalf SX Plant, Morenci, Arizona	Phelps Dodge Morenci, Modoc SX Plant, Morenci, Arizona	Phelps Dodge Morenci, Southwest SX Plant, Morenci, Arizona	Zambia CCM Tailings Leach Plant, Chingola, Zambia
RAFFINATE (gpl)							
Cu	0.2	0.11	0.27	0.34	0.33	0.25	0.2
sulfuric acid	8	6.5			6		6
pH	1.6	1.6	1.3	1.35	1.36	1.71	1.2
organic loss to raffinate, ppm	50	220 (total), 80 (net)	100	100	100	100	30
organic-from-raffinate recovery method	planning to install	4 Jameson cells, 1,363 m^3/hour each plus small after settlers prior to raffinate pond	floated in organic recovery tanks; also 2 lengths of plastic grating in the settlers	floated in organic recovery tanks; also 2 lengths of plastic grating in the settlers	floated in organic recovery tanks; also 2 lengths of plastic grating in the settlers		
extraction efficiency per train, %	95	93	87	86	88	70	96
extraction mixer retention time, minutes	2	1.8	2.5 to 3	2.5 to 3	3	4	3
ORGANIC							
Cu concentration in loaded organic, gpl	5.1	6.7	4.21	5.75	6.64	1.2	8.5
Cu concentration in barren organic, gpl	2	3.2	0.8	1.82	1.75	0.29	3
organic depth in settlers, mm	279	330	extraction: 254 strip: 279	305	305	E: 356 S: 305	200
aqueous depth in settlers, mm	483		extraction: 457 strip: 432	406	406	E: 355 S: 406	350
extractant name	Acorga 5640	Acorga M5640	LIX 860/84	LIX 860/84	LIX 860/84	LIX 860/84	LIX 984N, LIX 984 and PT5050
solvent name	Phillips SX-7	Phillips SX-7	Phillips SX-7	Phillips SX-7	Phillips SX-7	Phillips SX-7	Escaid 100 and Shellsol
extractant concentration in organic, %	10 by volume	13	11	12.5	12.4	2.5	18
net copper transference, gpl per % extractant	0.3	0.26 to 0.3	0.34	0.31	0.32	0.17	0.266

TABLE 1D Solvent Extraction Plant Data (continued)

	Burro Chief (Phelps Dodge), Santa Rita, New Mexico	Burro Chief Copper Co. (Phelps Dodge), Tyrone New Mexico	Phelps Dodge Morenci, Central SX Plant, Morenci, Arizona	Phelps Dodge Morenci, Metcalf SX Plant, Morenci, Arizona	Phelps Dodge Morenci, Modoc SX Plant, Morenci, Arizona	Phelps Dodge Morenci, Southwest SX Plant, Morenci, Arizona	Zambia CCM Tailings Leach Plant, Chingola, Zambia
stripping efficiency per train, %	95	95	80	69	74	76	65
strip mixer retention time, minutes	1.8	1.8	3	3	3	4	3
phase disengagement time, minutes	0.85	1.17	0.85	0.85	0.85	0.85	1.2 to 1.5
organic coalescer	none	none	aqueous is removed from organic on the E1's by two rows of water sprinklers that knock down aqueous	aqueous is removed from organic on the E1's by two rows of water sprinklers that knock down aqueous	none	none	none
organic filters	no	no	none	none	no	no	no
crud removal system	interface pumping using wands and diaphragm pumps	interface pumping	settler dumping and interface pumping	settler dumping and interface pumping	settler dumping and interface pumping	settler dumping	organic and aqueous are drained from the settlers; settled crud is washed & drained to crud ponds
crud treatment system	crud is dumped to a pond and sprinkled with water to settle solids; organic is skimmed, clay treated and pumped through 25 and 10 micro-meter sock filters	clay treatment of crud; filter through a polypropylene sock filter	wash crud with diluent to recover organic (2:1 diluent:crud)	wash crud with diluent to recover organic (2:1 diluent:crud)	wash crud with diluent to recover organic (2:1 diluent:crud, centrifuge	water washing	centrifuge
organic cleaning technique	clay treated	none	water washing and clay treatments	water washing and clay treatments	water washing and clay treatments, centrifuge	water washing	no
RICH ELECTROLYTE							
Cu, gpl	50	50	46.1	44	47.1	46.7	50
sulfuric acid, gpl	157	160	197	204	196.5	183.9	160

TABLE 1D Solvent Extraction Plant Data (continued)

	Burro Chief (Phelps Dodge), Santa Rita, New Mexico	Burro Chief Copper Co. (Phelps Dodge), Tyrone New Mexico	Phelps Dodge Morenci, Central SX Plant, Morenci, Arizona	Phelps Dodge Morenci, Metcalf SX Plant, Morenci, Arizona	Phelps Dodge Morenci, Modoc SX Plant, Morenci, Arizona	Phelps Dodge Morenci, Southwest SX Plant, Morenci, Arizona	Zambia CCM Tailings Leach Plant, Chingola, Zambia
organic loss in electrolyte, ppm		10 to 20	20	20	20	20	<30
electrolyte coalescer	none	five 1.2m dia. x 6m high column flotation cells		organic is floated from rich booster tank	organic is floated from rich booster tank	Jameson cell and filters	organic flotation
electrolyte treatment before tankhouse	four 3m3/min Disep packed bed anthracite filters	Disep garnet/ anthracite filters	2 Jameson cells and a 4m diameter anthracite/garnet filters	2 Jameson cells and a 4m diameter anthracite/garnet filter	2 Jameson cells and a 4m diameter anthracite/garnet filter	4m diameter anthracite/garnet filters	no
SPENT ELECTROLYTE							
Cu, gpl	35	34	33.3	33.3	36	38.9	35
sulfuric acid, gpl	182	180	212	222	212.8	208.2	180
cobalt, ppm	180	180	166	166	166	162	200 to 300
DESIGN AND BUILDING MATERIALS							
mixers	stainless steel	stainless steel	concrete with 6mm 316L stainless steel liner	concrete with 6mm 316L stainless steel liner	concrete with 6mm 316L stainless steel liner	concrete, epoxy-lined	concrete
settlers	stainless steel	stainless steel	316L stainless steel	316L stainless steel	316L stainless steel	concrete, epoxy-lined	concrete
piping	stainless steel	stainless steel	316L stainless steel	316L stainless steel	316L stainless steel	316L stainless steel	stainless steel
tanks	stainless steel	stainless steel	316L stainless steel	316L stainless steel	316L stainless steel	316L stainless steel	stainless steel, mild steel and lead lined
gratings	fiberglass	fiberglass	Sea Fab fiberglass	Sea Fab fiberglass	Sea Fab fiberglass	Sea Fab fiberglass	concrete/crush stones
SETTLERS							
settler dimensions, length x width x depth	24.4 x 24.4 x 1	27 x 24 x 1	22 x 25 x 1	22 x 25 x 1	22 x 25 x 1	15 x 15 x 0.9	Ext: 36.4 x 12.3 x 0.76 Strip: 27.0 x 12.4 x 0.76
number of picket fences, per settler	E2, S: 2; E1: 3	extraction:3; strip:2	2	2	2	1	1

TABLE 1D Solvent Extraction Plant Data (continued)

	Burro Chief (Phelps Dodge), Santa Rita, New Mexico	Burro Chief Copper Co. (Phelps Dodge), Tyrone New Mexico	Phelps Dodge Morenci, Central SX Plant, Morenci, Arizona	Phelps Dodge Morenci, Metcalf SX Plant, Morenci, Arizona	Phelps Dodge Morenci, Modoc SX Plant, Morenci, Arizona	Phelps Dodge Morenci, Southwest SX Plant, Morenci, Arizona	Zambia CCM Tailings Leach Plant, Chingola, Zambia
settling velocity, m^3/m^2 per hour	5.3	4	5.5	5.5	5.5		Ext: 3.7, Strip 3.2
MIXERS							
extraction mixers	Lightnin pump mixers (auxiliaries are not used)	Lightnin one primary pump mixer two auxiliary mixers (all in use)	Lightnin straight-vaned pump impellers (series), Philadelphia Mixers curve vaned impellers (parallels). All are used.	Lightnin straight-vaned pump impellers (series), Philadelphia Mixers curve vaned impellers (parallels). All are used.	Lightnin straight-vaned pump impellers (series), Philadelphia Mixers curve vaned impellers (parallels). All are used.	Lightnin straight-vaned pump impellers (all used)	pump mixers (all used)
principal extraction mixer	37 kW 36 to 46 RPM	29.8 kW; 42 to 46 RPM	17.6 kW, 51 RPM	17.6 kW, 51 RPM	17.6 kW, 51 RPM	17.6 kW, 51 RPM	37 kW, 43 RPM
impeller diameter, m	2.18	2.17	2.18	2.18	2.18	2.18	2.7
stripping mixers	Lightnin pump mixers (auxiliary [1] is used)	Lightnin one primary pump mixer one auxiliary mixer (all in use)	Lightnin straight-vaned pump impellers (series), Philadelphia Mixers curve vaned impellers (parallels). All are used.	Lightnin straight-vaned pump impellers (all used)	Lightnin straight-vaned pump impellers (all used)	Lightnin straight-vaned pump impellers (all used)	pump mixers (all used)
principal strip mixer	37 kW 42 RPM	29.8 kW; 42 to 46 RPM	17.6 kW, 51 RPM	17.6 kW, 51 RPM	17.6 kW, 51 RPM	17.6 kW, 51 RPM	37 kW, 45 RPM
impeller diameter, m	2.18	2.17	2.18	2.18	2.18	2.18	2.7

TABLE 2A Electrowinning Tankhouse Data

	Aberfoyle Resources, Ltd. Gunpowder Division, Queensland, Australia	Girilambone Copper Co. Girilambone, NSW Australia	Great Australia Mining Co., Cloncurry, Queensland, Australia	Mount Isa Mines Ltd., Mount Isa, Queensland Australia	Pasminco Metals-BHAS Port Pirie, South Australia	Compania Minera Carmen de Andacollo, La Serena, Chile
Cathode capacity, tonnes/year	13,000	18,000	5,500	3,500	5,000	20,000
1994 cathode, tonnes	6,440	12,000 to 13,000 (1995)			3,500 (4,000 in 95/96)	startup at end of 1996
ANODES						
composition Pb %	98	98	yes		94	
Sn %	1	1.75	yes			
Ca %	0.6	0.1	yes			
other		0.005 to 0.02% Al			Sb, 6	
rolled or cast	rolled	rolled	rolled	rolled	cast	rolled (not cross-rolled)
life before replacement, years	blades, 3–4; bars, 6–8	5 to 10	life of mine			new
blade l x w x t, mm	1060 x 920 x 6	1114 x 920		1120 x 920 x 6	1000 x 905 x 10	
hanger bar material	copper	4–5%Sb/Pb over solid Cu			copper	copper
bar/blade connection	lead solder	8% Sb/Pb burn	lead solder	lead solder	cast antimonial lead	RSR type
hung symmetrically or asymmetrically	symmetrically	asymmetrically	asymmetrically	asymmetrically	symmetrically	symmetrically
anode/cathode spacers: what/where	3 poly buttons	three buttons near blade bottom, two PVC full blade-length hairpins	buttons, one in center, two on bottom	cast polymer concrete is used as capping board between cells	nylon buttons, two at bottom of anode, one slightly above midpoint	polymer resin, tight tolerance ±2 to 4 mm
CATHODES						
plating time, days	7	7	7	7	7	7
stainless steel blanks or electrorefinery or electrowon copper starting sheets	stainless steel	stainless steel	stainless steel	stainless steel	titanium blanks	stainless steel
Copper Starting Sheets						
length x width x thickness, mm						
mass after plating, kg						
loading method						

TABLE 2A Electrowinning Tankhouse Data (continued)

	Aberfoyle Resources, Ltd. Gunpowder Division, Queensland, Australia	Girilambone Copper Co. Girilambone, NSW Australia	Great Australia Mining Co., Cloncurry, Queensland, Australia	Mount Isa Mines Ltd., Mount Isa, Queensland Australia	Pasminco Metals-BHAS Port Pirie, South Australia	Compania Minera Carmen de Andacollo, La Serena, Chile
Stainless Steel Blanks						
ISA or Kidd	ISA	ISA	ISA	ISA		Kidd
length x width x thickness, mm	1290 x 1042 x 3.25	1290 x 1026 x 3	1044 x 1288 x 3.2	1336 x 1042 x 3.25		1060 x 1000
type of stainless steel	316L	316L	316L	316L	titanium	316L
side edging	CPVC	CRL-ABS, Quaona O/O ABS	PVDC	CPVC	acrylic/PVC	Schiebler
bottom edging	wax	wax (changing to CPVC Quaona & Quaona corner pieces)	PVDC	wax	none	Kidd groove
hanger bar						
I-beam or rectangular	hollow rectangular	solid & hollow rectangular	hollow rectangular	hollow rectangular		CAID
material	304L stainless steel plated with approximately 2mm of Cu	copper plated stainless and solid copper	copper-coated stainless steel	copper coated 304L stainless steel		copper
height, width, length, mm	48 x 35 x 1400	38 x 25 x 1405		48 x 35 x 1410		CAID Standard
joint between bar & blade	stitch welded	Cu-coated stitch weld	weld	Cu-plated stainless steel weld		welded
mass of Cu plate from each side of blank, kg	40-80, grade dependent	40 to 60	45 to 55	45	35 to 40	60
stripping method, machine or manual	manual	manual	manual	manual	manual in wash tank	machine
machine manufacturer	Mesco/ISA	CRL/ISA	CRL/ISA	Mesco/ISA		AISCO/Kidd
water temperature, °C	76	80, dropping to 45 with no wax	60	80	room temperature	70 to 80
mother blanks/hour	60	70 to 80	40	60		60
features						
washing	yes	yes	no	yes		yes
stripping		yes	yes	no		yes

TABLE 2A Electrowinning Tankhouse Data (continued)

	Aberfoyle Resources, Ltd. Gunpowder Division, Queensland, Australia	Girilambone Copper Co. Girilambone, NSW Australia	Great Australia Mining Co., Cloncurry, Queensland, Australia	Mount Isa Mines Ltd., Mount Isa, Queensland Australia	Pasminco Metals-BHAS Port Pirie, South Australia	Compania Minera Carmen de Andacollo, La Serena, Chile
stacking	yes	no	yes	no		yes
sampling	no	no	no	no		no (drilled)
weighing	no	no	no	no		forklift to scale
strapping	no	no	no	no		hand pneumatic
CELLS						
number	74	92	32	26	30	84
length x width x depth, m	3.4 x 1.19 x 1.28	3.43 x 1.2 x 1.51		3.14 x 1.45 x 1.5	4.7 x 1.2 x 1.32	4.6 x 1.25 x 1.45
construction material	concrete	polymer concrete	polymer concrete	concrete	concrete	polymer concrete
lining material	paraliner	fiberglass		HDPE	paraliner	fiberglass
capping board material	copolymer poly	fibrous resin	epoxy composite	HDPE sheet and timber or cast polymer concrete cathode/ anode spacers		polymer resin
intercell busbar type	dogbone	triangular	triangular	triangular	dogbone	triangular
x-sectional area, mm^2	2475	450		545		
current density, A/mm^2	7	50				750
anodes, cathodes/cell	34,33	34,33	34,33	31, 30	38,37 (41, 40 max)	46,45
cathode-cathode center-line distance, mm	95	95	95	95	105	95
cell inspection system	infrared camera	infrared	infrared and current	infrared		infrared gun
inspections per day	1	1	6	3		3
mist suppression system						
balls	no	in overflow boxes	no	no	no	yes, 19mm
BB's	yes	yes	yes	yes	no	no
ventilation	no	open-sided tankhouse	no	no	no	no
FC 100	no		no	no	yes	no
FC 1100	no		no	no	no	yes

TABLE 2A Electrowinning Tankhouse Data (continued)

	Aberfoyle Resources, Ltd. Gunpowder Division, Queensland, Australia	Girilambone Copper Co. Girilambone, NSW Australia	Great Australia Mining Co., Cloncurry, Queensland, Australia	Mount Isa Mines Ltd., Mount Isa, Queensland Australia	Pasminco Metals-BHAS Port Pirie, South Australia	Compania Minera Carmen de Andacollo, La Serena, Chile
other						
cell cleaned after (days)	365 to 547	720	at least one year	180	180	90
ELECTRICAL						
current per cell, amps	18,000	22,000	21,900		16,000 to 17,000	
cathode current density, amps/m^2	280	280 to 310	305	150 to 200	240 to 260	280
cathode current efficiency, %	87 to 95	89 to 90	93		92	90 to 93
anode-cathode voltage, v	2	2.07	2 to 2.1			2.1
total AC electrical energy consumption (MWh per tonne of cathode)	2.1 to 2.3	2				2
ELECTROLYTE						
total volume in plant, m^3	485	750	330		700	1600
flow of electrolyte to solvent extraction, m^3/day	1560	3600	1056	672	750	4000
electrolyte treatment before entering tankhouse	Jameson cell then dual media filter	Jameson cell, then three Spintek multimedia filters, then 17 polishing cells	Spintek dual media filter plus scavenger cells	Jameson cell, dual media filter	from solvent extraction to tankhouse via Jameson cell	column flotation followed by dual media filter; Co adjusted to 100 ppm, Cl to 15-20
electrolyte flow in and out of each cell, m^3/min	0.12	0.13	0.13	0.09	0.06	0.148
flow control system	diaphragm or ball valves in offtakes from header	manual ball valve on cell feed line - no outlet control	valve on inlet, open outlet	manually operated valve	none	globe and ball valve on cell feed line
electrolyte distribution system and hole size, mm	32mm PVC inlet discharges into 100mm vertical PVC pipe (no manifolds)	via 50mm line into PVC cell distribution manifold	soon to install manifolds in each cell	single entry point at end of cell	feed pipe to one end of cell	manifold with 10mm hole spaced at each cathode
overflow pipe diameter, mm	100	110	75			100

TABLE 2A Electrowinning Tankhouse Data (continued)

	Aberfoyle Resources, Ltd. Gunpowder Division, Queensland, Australia	Girilambone Copper Co. Girilambone, NSW Australia	Great Australia Mining Co., Cloncurry, Queensland, Australia	Mount Isa Mines Ltd., Mount Isa, Queensland Australia	Pasminco Metals-BHAS Port Pirie, South Australia	Compania Minera Carmen de Andacollo, La Serena, Chile
entering cells						
Cu, gpl	39	40	55		36	two cascade system with scavengers: 50 gpl Cu to scavengers; 42 gpl Cu to cascade A; 38 to 40 gpl Cu to cascade B
H_2SO_4, gpl	170	175	160		170	
temperature, °C	45	42	35 to 40			40 to 45
temperature control method	heat exchanger	heat exchanger	none			heat exchanger and trim heater
leaving cells						
Cu, gpl	35	37	35		31	3 to 4 gpl drop
H_2SO_4, gpl	180 to 185	185	185		180	170
temperature, °C	48	46	35 to 40			
other electrolyte components						
Fe^{++}, gpl	3 to 5 total Fe	0.37	0.3			<2 total Fe
Fe^{+++}, gpl		0.84	1.2			
Co^{++}, ppm	175	100	100			100
Cl^-, ppm	8	25	<20		80 to 100	15 to 20
other, ppm		Mn, 11	Mn, 120			
guar addition, g per tonne of cathode	none	150	none	none	300	not yet
cobalt addition, g per tonne of cathode	to maintain 175 ppm	to maintain 100 ppm	yes	to maintain 80 ppm	320	to 100 ppm in electrolyte added as sol'n

TABLE 2A Electrowinning Tankhouse Data (continued)

	Aberfoyle Resources, Ltd. Gunpowder Division, Queensland, Australia	Girilambone Copper Co. Girilambone, NSW Australia	Great Australia Mining Co., Cloncurry, Queensland, Australia	Mount Isa Mines Ltd., Mount Isa, Queensland Australia	Pasminco Metals-BHAS Port Pirie, South Australia	Compania Minera Carmen de Andacollo, La Serena, Chile
maximum Mn level		Mn oxidized in tankhouse. It started to degrade SX organic. Mn was >80 ppm and Fe was <0.5 gpl in electrolyte	carryover yes, but now under control			add ferrous sulfate when Mn concentration is greater than 200 ppm in electrolyte
electrolyte bleed from tankhouse, m^3/day	28	25 to 30 backwash from Spintek filters	nil, use filter backwash as control	12	3 to 4	50 to 100
bleed to control	stable Fe (in cell output)	stable Fe and Cl^- levels	Mn and Cl^-	Mn and Cl^-	as required by Zn plant	Mn
destination	pregnant leach solution	E2 organic launder or S1 picket fence			to purify zinc sulfate solution	E1 in solvent extraction plant
CATHODE COMPOSITION						
Pb	<1	<0.1	<1		<15	<1
S	0.5	2	5 to 10		<15	<3
Fe	1.4	1.5	<2		<2	<2
Mn	<0.1	<0.1	<0.1		0.1	<0.1
Co	<0.1	<0.1	<0.1		0.1	<1
Ni	<0.1	<0.1	<0.1		<0.5	<0.3
other			all <0.1			tot. LME impurities, <10

TABLE 2B Electrowinning Tankhouse Data

	Compania Cerro Colorado Ltda., Cerro Colorado, Chile	CODELCO Chile, El Teniente Division, Colon-Rancagua, Chile	Sociedad Contractual Minera El Abra, El Loa, Calama, Chile	Empresa Minera de Mantos Blancos, Mantos Blancos Mine, Chile	Empresa Minera de Mantos Blancos, Mantoverde Mine, Chile	Compania Minera Quebrada Blanca, Chile
Cathode capacity, tonnes/year	60,000	3000 to 7500	225,000 (nominal)	45,000	42,000	82,500
1994 cathode, tonnes	21,500	5000		new	new	46,350 (1995)

TABLE 2B Electrowinning Tankhouse Data (continued)

	Compania Cerro Colorado Ltda., Cerro Colorado, Chile	CODELCO Chile, El Teniente Division, Colon-Rancagua, Chile	Sociedad Contractual Minera El Abra, El Loa, Calama, Chile	Empresa Minera de Mantos Blancos, Mantos Blancos Mine, Chile	Empresa Minera de Mantos Blancos, Mantoverde Mine, Chile	Compania Minera Quebrada Blanca, Chile
ANODES						
composition Pb %		98.9	98.4	98.63	98.63	
Sn %	1.25	1	1.5	1.3	1.3	1.5
Ca %	0.6	0.08	0.08	0.07	0.07	0.065
other	0.0005 to 0.002% Al		Al, 0.02%			
rolled or cast	rolled RSR	rolled	rolled	rolled	rolled	rolled (RSR)
life before replacement, years		5	new	new	new	8+
blade l x w x t, mm	950 x 940 x 6	950 x 1143 x 9	1093 x 940 x 6	1087 x 940 x 6	1087 x 940 x 6	1053 x 940 x 6
hanger bar material	copper (6mm Sb-Pb)	lead coated copper	copper, lead covered	copper	copper	Pb plated Cu bar
bar/blade connection	welded		welded	tin welding	tin welding	welded
hung symmetrically or asymmetrically	symmetrically	symmetrically	asymmetrically	asymmetrically	asymmetrically	symmetrically
anode/cathode spacers: what/where	proprietary	5cm	two conical insulation buttons at bottom of anode	two buttons and 1 polypropylene hairpin	two buttons and 1 polypropylene hairpin	yo-yo's at corners, PVC center spacer plus hairpin
CATHODES						
plating time, days	7	8 to 12	7	7	7	7
stainless steel blanks or electrorefinery or electrowon copper starting sheets	stainless steel	electrowon starting sheets	stainless steel	stainless steel	stainless steel	stainless steel
Copper Starting Sheets						
length x width x thickness, mm		1000 x 1000 x 5				
mass after plating, kg		70				
loading method						
Stainless Steel Blanks						
ISA or Kidd	ISA		Kidd	Kidd	Kidd	ISA
length x width x thickness, mm	1001 x 1000 x 3.25		1300 x 1050 x 3	1000 x 1000 x 3	1000 x 1000 x 3	1180 x 1036 x 3
type of stainless steel	316L		316L	316L	316L	316L

TABLE 2B Electrowinning Tankhouse Data (continued)

	Compania Cerro Colorado Ltda., Cerro Colorado, Chile	CODELCO Chile, El Teniente Division, Colon-Rancagua, Chile	Sociedad Contractual Minera El Abra, El Loa, Calama, Chile	Empresa Minera de Mantos Blancos, Mantos Blancos Mine, Chile	Empresa Minera de Mantos Blancos, Mantoverde Mine, Chile	Compania Minera Quebrada Blanca, Chile
side edging	CPVC		Scheibler	Scheibler polypropylene	Scheibler polypropylene	C-PVC (Quadra)
bottom edging	wax		V notch	V notch	U notch	wax
hanger bar						
I-beam or rectangular	rectangular		rectangular	rectangular	rectangular	rectangular
material	316L stainless steel, copper coated		solid copper	solid copper	solid copper	solid copper
height, width, length, mm	43 x 30 x 1450		38 x 25 x 1430	38 x 25.4 x 1450	38 x 25.4 x 1450	38 x 25 x 1378
joint between bar & blade	SS MIG welded, MIM process		solder (weld)	copper welding	copper welding	copper welded
mass of Cu plate from each side of blank, kg	40		50	42.5	42.5	45
stripping method, machine or manual	machine		machine	machine	machine	machine
machine manufacturer	Wenmec/ISA		Aisco/Kidd	Aisco/Kidd	Aisco/Kidd	Wenmec/ISA
water temperature, °C	hot		two stage washing	65	65	80
mother blanks/hour	182		500 (nominal)	300	300	360
features						
washing	yes		yes	yes	yes	yes
stripping	yes		yes	yes	yes	yes
stacking	yes		yes	yes	yes	yes
sampling	no (separate scale)		yes	no	no	no
weighing	no (separate scale)		yes	yes	yes	yes
strapping	no (separate scale)		yes	no	no	yes
CELLS						
number	224		600	164	156	264
length x width x depth, m	6.17 x 1.27 x 1.42	4.8 x 1.21 x 1.25	6.57 x 1.30 x 1.40	6.17 x 1.27 x 1.41	6.17 x 1.27 x 1.41	6.1 x 1.24 x 1.36

TABLE 2B Electrowinning Tankhouse Data (continued)

	Compania Cerro Colorado Ltda., Cerro Colorado, Chile	CODELCO Chile, El Teniente Division, Colon-Rancagua, Chile	Sociedad Contractual Minera El Abra, El Loa, Calama, Chile	Empresa Minera de Mantos Blancos, Mantos Blancos Mine, Chile	Empresa Minera de Mantos Blancos, Mantoverde Mine, Chile	Compania Minera Quebrada Blanca, Chile
construction material	pre-cast polymer concrete	concrete	polymer concrete	polymer concrete	polymer concrete	polymer concrete
lining material	none	paraliner	none	FRP	FRP	none
capping board material	polymer concrete	PVC	polymer concrete	polymer concrete	polymer concrete	polymer concrete
intercell busbar type	triangular	triangular	triangular	triangular	triangular	dogbone
x-sectional area, mm^2	25mm side			629	629	2900
current density, A/mm^2				0.8	0.8	11.4
anodes, cathodes/cell	61, 60	43,42	67, 66	58,57	58,57	61,60
cathode-cathode center-line distance, mm	95	110	95	100	100	95
cell inspection system	laser temperature detector	gaussmeter	infrared scanner (manual)	gaussmeter	gaussmeter	infrared
inspections per day		3	2	3	2	1
mist suppression system						
balls	no	yes	yes	yes	yes	yes
BB's	yes	no	no	no	no	yes
ventilation	yes	no	yes	cross-flow ventilation	cross-flow ventilation	yes
FC 100	no	no	no	no	no	none
FC 1100	no	no	yes	no	no	none
other						
cell cleaned after (days)	60	90	100	70	60	130
ELECTRICAL						
current per cell, amps	29,000		36,000 to 40,000	30,000	30,000	33,000
cathode current density, $amps/m^2$	242	90 to 150	270 to 300	260	260	275
cathode current efficiency, %	93	90 to 92	95	95	95	91
anode-cathode voltage, v	1.86	1.7	2	1.6	1.6	1.9

TABLE 2B Electrowinning Tankhouse Data (continued)

	Compania Cerro Colorado Ltda., Cerro Colorado, Chile	CODELCO Chile, El Teniente Division, Colon-Rancagua, Chile	Sociedad Contractual Minera El Abra, El Loa, Calama, Chile	Empresa Minera de Mantos Blancos, Mantos Blancos Mine, Chile	Empresa Minera de Mantos Blancos, Mantoverde Mine, Chile	Compania Minera Quebrada Blanca, Chile
total AC electrical energy consumption (MWh per tonne of cathode)	1.79			2.07	1.9 (for plating)	1.95
ELECTROLYTE						
total volume in plant, m^3	2700	850		3400	3500	4000
flow of electrolyte to solvent extraction, m^3/day	4060	960	1600 to 1800	9400	8400	16,000
electrolyte treatment before entering tankhouse	Cominco columns and garnet/anthracite filters	garnet, quartz, anthracite filter	electrolyte coalescer (Pacesetter), electrolyte filters (Spintek)	flotation columns (Magma design) and garnet/anthracite filter	flotation columns (Magma design) and garnet/anthracite filter	column cells and U.S. Filters (garnet/anthracite) for organic removal
electrolyte flow in and out of each cell, m^3/min	0.19		0.22	0.2	0.23	0.18
flow control system	flow control valve	flowmeter	manual	manual valve	manual valve	manual valves and visual control system
electrolyte distribution system and hole size, mm	ring distributor, 190 six mm holes	via a feed manifold independently to each cell	distribution manifold with 8mm holes	pipe distributor, 50 mm hole	pipe distributor, 50 mm hole	38mm diameter manifold with 3mm holes every 42mm along sides of cell
overflow pipe diameter, mm	101.6	150		100	100	128 (schedule 40 PVC)
entering cells						
Cu, gpl	38	45	40	39	36	48
H_2SO_4, gpl	176	150 to 155	160	180	160	150
temperature, °C	50	28 to 32		45	40	40+
temperature control method	heat interchange	electronic sensor	heated water and heat exchangers	heat exchanger	heat exchanger	boilers/heat exchangers
leaving cells						
Cu, gpl	35	32 to 36	36	36	33	36
H_2SO_4, gpl	180	155 to 160	166	186	165	170

TABLE 2B Electrowinning Tankhouse Data (continued)

	Compania Cerro Colorado Ltda., Cerro Colorado, Chile	CODELCO Chile, El Teniente Division, Colon-Rancagua, Chile	Sociedad Contractual Minera El Abra, El Loa, Calama, Chile	Empresa Minera de Mantos Blancos, Mantos Blancos Mine, Chile	Empresa Minera de Mantos Blancos, Mantoverde Mine, Chile	Compania Minera Quebrada Blanca, Chile
temperature, °C	50	36 to 39	41 to 46	45	45	45
other electrolyte components						
Fe^{++}, gpl	0.52		0.16	1	0.7	0.5
Fe^{+++}, gpl	1.08	500 to 800 ppm	0.6	0.8	0.4	1
Co^{++}, ppm	120	110 to 140 ppm	120	120	120	100
Cl^{-}, ppm	5	20 to 35 ppm	10	18+L19	12	15
other, ppm	Mn, 18	Mn, 2; Si, 20 to 50; Mg, 30 to 60	Mn, 8.5		Mn, 40	Mn, 20
guar addition, g per tonne of cathode	290	300	210	200	220	300
cobalt addition, g per tonne of cathode	350	630	to maintain 120 ppm	800	to 120 ppm in electrolyte	600 cobalt sulfate
maximum Mn level	no, Mn was important before change from strip organic tank to load organic tank				only had Mn problems in pilot plant (rapid anode corrosion), not in production plant	
electrolyte bleed from tankhouse, m^3/day	15			80	intermittently	80
bleed to control			mainly Cl^{-} at this time	Cl^{-} to below 25 ppm	Mn or Cl^{-}	control Fe to <2gpl
destination	raffinate pond		SX wash stage (for pH control) or E1	washing stage in solvent extraction plant	E1 in solvent extraction plant	E2 in solvent extraction plant
CATHODE COMPOSITION						
Pb	2	0 to 5	<1	1	1	1 to 2
S	5	0 to 15	<6	10	12	4 to 6
Fe	1.5	0 to 7	<2	1	1	2 to 3
Mn	<0.4	0 to 1	<3	<1	<1	0.1
Co	<1.0	0 to 1		1	<1	<1
Ni	<0.8	0 to 1	<0.8	<1	<1	0.3
other		Te: 0-2, Sb: 0-4, Cd: 0-1	O2, 89 ppm (50 to 200)			

TABLE 2C Electrowinning Tankhouse Data

	Compania Minera Zaldivar, Chile	Hellenic Copper Mines Ltd., Skouriotissa, Cyprus	Mexicana de Cobre, S.A. de C.V., Nacozari, Sonora, Mexico	Southern Peru Limited, Toquepala, Tacna, Peru	Impala Platinum Ltd., Springs, South Africa (1996 data)	Impala Platinum Ltd., Springs, South Africa (proposed new plant)
Cathode capacity, tonnes/year	125,000	8000	22,000	39,000	7500	10,000
1994 cathode, tonnes	22,500 (6 months)	new	new	new	5400	
ANODES						
composition Pb %			98.4	98.54	94	
Sn %		1.75	1.5	1.3		1.25
Ca %		0.075	0.075	0.07		0.06
other			Al, 0.0125%	0.0009	Sb, 6%	
rolled or cast	rolled	rolled	rolled	rolled	cast	rolled
life before replacement, years	new	new	4 to 5	5	3	not known
blade l x w x t, mm		1092 x 940 x 6	1100 x 860 x 6	1171 x 950 x 6.4	1172 x 995 x 10	1160 x 956 x 9
hanger bar material	copper, tin plated	copper	copper	Cu with Pb alloy coating	copper	copper
bar/blade connection	budweld	welded	welded		bar is tinned and molded into anode	cast in Pb and burned onto blade
hung symmetrically or asymmetrically	symmetrically	asymmetrically	symmetrically	asymmetrically	asymmetrically	symmetrically
anode/cathode spacers: what/where	two buttons on bottom corners	two polypropylene insulator cones, 100mm from bottom and edge	injection molded individual chairs of fiber-filled polycarbonate	3 buttons: 2 at bottom corners and one central, 533mm up from bottom	3 polypropylene hairpins hung over each anode	
CATHODES						
plating time, days	7 to 8	7	7 to 8	7	7	7+
stainless steel blanks or electrorefinery or electrowon copper starting sheets	stainless steel	stainless steel	stainless steel	stainless steel	electrowon starter sheets produced on Ti blanks	stainless steel
Copper Starting Sheets						
length x width x thickness, mm					1200 x 1080 x 1	
mass after plating, kg					60	

TABLE 2C Electrowinning Tankhouse Data (continued)

	Compania Minera Zaldivar, Chile	Hellenic Copper Mines Ltd., Skouriotissa, Cyprus	Mexicana de Cobre, S.A. de C.V., Nacozari, Sonora, Mexico	Southern Peru Limited, Toquepala, Tacna, Peru	Impala Platinum Ltd., Springs, South Africa (1996 data)	Impala Platinum Ltd., Springs, South Africa (proposed new plant)
loading method					starting sheet loops stitched to starting sheet and hung over bars	
Stainless Steel Blanks						
ISA or Kidd	ISA	ISA	ISA	ISA		ISA
length x width x thickness, mm	1100 x 1026 x 3.25	1102 x 1000 x 3.25	928 x 1148 x 3	1245 x 1036 x 3.25		1230 x 1070 x 3
type of stainless steel	316L, Polarit	316L	316L	316L		316L
side edging	CPVC	ABS	dual durometer edge	CPVC		ABS
bottom edging	wax, BarecoR	ABS edge strip	durometer edge strip	wax		ABS
hanger bar						
I-beam or rectangular	hollow rectangular	hollow rectangular	rectangular	hollow rectangular		hollow rectangular
material	copper, tin plated	stainless steel, copper coated	solid copper	stainless steel with copper coating		316L, with 2.5mm Cu plate
height, width, length, mm	43 x 30 x 1330	43 x 30 x 1368	40 x 25 x 1220	45 x 30 x 1450		38 x 25 x 1340
joint between bar & blade	weld	weld	Cu welded	stitch weld		welded
mass of Cu plate from each side of blank, kg	38 to 42	54	47	54 to 56		45
stripping method, machine or manual	machine	semi-automatic	machine	machine		machine
machine manufacturer	Wenmec/ISA	T.M. Engineers/ISA	Mesco/ISA	Wenmec		CRL/ISA
water temperature, °C	70 to 75°C	70	60 to 80	75		75
mother blanks/hour	400	60	144	250		100
features						
washing	yes	yes	yes	yes, automatic		yes
stripping	yes	yes	yes	yes, automatic		yes
stacking	yes	yes	yes	yes, automatic		yes

TABLE 2C Electrowinning Tankhouse Data (continued)

	Compania Minera Zaldivar, Chile	Hellenic Copper Mines Ltd., Skouriotissa, Cyprus	Mexicana de Cobre, S.A. de C.V., Nacozari, Sonora, Mexico	Southern Peru Limited, Toquepala, Tacna, Peru	Impala Platinum Ltd., Springs, South Africa (1996 data)	Impala Platinum Ltd., Springs, South Africa (proposed new plant)
sampling	yes	no	no	no, manual		yes
weighing	yes	yes	yes	yes, automatic		no
strapping	yes	yes	no	no, manual		no
CELLS						
number	368	50	94	122	76 with 16 more being installed	92
length x width x depth, m	6.72 x 1.24 x 1.37	3.48 x 1.25 x 1.5	6.02 x 1.61 x 1.12	6.17 x 1.27 x 1.40	4.17 x 1.22 x 1.49	4.17 x 1.22 x 1.49
construction material	polymer concrete	concrete	monolithic polymer concrete	polymer concrete	Bekaplast/concrete	Bekaplast/concrete
lining material	none	paraliner	none	none	Bekaplast/ polypropylene (14 paraliners remain)	Bekaplast/ polypropylene
capping board material	polymer concrete	FRP	fiber filled polycarbonate Lexan	polymer concrete	rubber	polypropylene
intercell busbar type	dogbone	triangular	dog bone	triangular		dogbone
x-sectional area, mm^2		440	1400	629	1825	2025
current density, A/mm^2		1.2	1			
anodes, cathodes/cell	61,60 (64,63 maximum)	31,30	55,54	61, 60	37,36	39,38
cathode-cathode center-line distance, mm	120	100	102	95.25	111	104
cell inspection system	infrared camera	infrared camera and gaussmeter	infrared camera, hand held	infrared scanner	infrared	infrared
inspections per day	1	1	3	2	2	2?
mist suppression system						
balls	19mm	yes	yes	no	Allplas	Allplas
BB's	none	no	no	yes	no	no
ventilation	yes	no	yes	yes, crossflow	no	no
FC 100	no	no	no	no	no	no

TABLE 2C Electrowinning Tankhouse Data (continued)

	Compania Minera Zaldivar, Chile	Hellenic Copper Mines Ltd., Skouriotissa, Cyprus	Mexicana de Cobre, S.A. de C.V., Nacozari, Sonora, Mexico	Southern Peru Limited, Toquepala, Tacna, Peru	Impala Platinum Ltd., Springs, South Africa (1996 data)	Impala Platinum Ltd., Springs, South Africa (proposed new plant)
FC 1100	yes	no	no	being tested	no	no
other						
cell cleaned after (days)	60 to 90 (1996)	1 year	6 months	2 months	8 weeks	8 weeks
ELECTRICAL						
current per cell, amps		18,150	35,000 maximum	30,600	10,000 to 14,000 17,000 maximum	10,000 to 14,000 17,000 maximum
cathode current density, amps/m^2		275	305 maximum	255	120 to 150 (180 max.)	160 maximum
cathode current efficiency, %		90	90	94	85	more than 85
anode-cathode voltage, v	1.9	2.5	1.8 to 2.0	1.96	1.7	
total AC electrical energy consumption (MWh per tonne of cathode)	1.9	2	3.47 including leach	2.5		
ELECTROLYTE						as with present plant
total volume in plant, m^3	8000	450	3300	2400	780	
flow of electrolyte to solvent extraction, m^3/day	26,000	1600	5800	6600	350	
electrolyte treatment before entering tankhouse	Cominco air flotation columns (1 per SX train); Disep sand/ anthracite filters (4 per 2 SX trains)	dual media filter to recover organic and remove solids	through two Jameson cells	through two column flotation cells and three dual media filters	contacted with 90°C sulfurous acid to precipitate Se & Te, then filtered twice	
electrolyte flow in and out of each cell, m^3/min	0.2	0.106	0.13; 0.21 scavengers	0.16		
flow control system	manual ball valve for each cell	valves	V notch outlet weir into overflow box	PVC inlet valves and discharge weirs	single entry point	
electrolyte distribution system and hole size, mm	manifold with two x 60 holes, 8mm diameter	2 distribution pipes with thirty 6mm dia. holes every 100mm, 30° from vertical	manifolds on each side of cell bottom, with 6mm diameter holes between cathode pairs	53mm PVC rectangular loop manifold with 6.3mm holes every 46mm		
overflow pipe diameter, mm		110 O.D.	152	152		

TABLE 2C Electrowinning Tankhouse Data (continued)

	Compania Minera Zaldivar, Chile	Hellenic Copper Mines Ltd., Skouriotissa, Cyprus	Mexicana de Cobre, S.A. de C.V., Nacozari, Sonora, Mexico	Southern Peru Limited, Toquepala, Tacna, Peru	Impala Platinum Ltd., Springs, South Africa (1996 data)	Impala Platinum Ltd., Springs, South Africa (proposed new plant)
entering cells						
Cu, gpl	40	45	45	33	65 to 80	
H_2SO_4, gpl	156	156	160	165	20 to 30	
temperature, °C	42 to 44	42		46	60	
temperature control method	hot water heat exchange in cell feed	boiler and heat exchanger	two heat exchangers	electrolyte/electrolyte heat exchanger	steam-heated annular heat exchangers	
leaving cells						
Cu, gpl	37.5	42	30	30	18 to 23	
H_2SO_4, gpl	160	160	180	170	90 to 110	
temperature, °C	42 to 44	42		47	60	
other electrolyte components						
Fe^{++}, gpl	0.17	0.2	0.3	0.1	total Fe, 1	
Fe^{+++}, gpl	1.5	1.3	1	1.5		
Co^{++}, ppm	140	100	120	110 to 120	300	
Cl^-, ppm	<5	22	0	<10	0	
other, ppm			Mn, 80	Mn, 60 to 90	As, 200; Ni, 22 to 35 gpl; Pb, 20 to 30; Se, 1 to 2; Te, 1 to 4	
guar addition, g per tonne of cathode	250	not yet	120	270	none	
cobalt addition, g per tonne of cathode	800	1000 g CoSO4:7H2O	to 120 ppm	630	present in feed	
maximum Mn level	no proven problems		Mn is a problem when the Fe concentration is low (<0.5 gpl)			
electrolyte bleed from tankhouse, m^3/day	300	50	none	96 to 108	all spent electrolyte is recycled to leach	
bleed to control		Cl^-		Fe level to 1.5–1.6 gpl		
destination	raffinate pond	pregnant leach solution feed pond		raffinate pond		

TABLE 2C Electrowinning Tankhouse Data (continued)

	Compania Minera Zaldivar, Chile	Hellenic Copper Mines Ltd., Skouriotissa, Cyprus	Mexicana de Cobre, S.A. de C.V., Nacozari, Sonora, Mexico	Southern Peru Limited, Toquepala, Tacna, Peru	Impala Platinum Ltd., Springs, South Africa (1996 data)	Impala Platinum Ltd., Springs, South Africa (proposed new plant)
CATHODE COMPOSITION					1995 averages	
Pb	2	<0.1	1.3	<1	56	
S	4	1.5	6.57	<5	210	
Fe	<2	3	1.17	<1	16	
Mn	<1	<0.1	0.155	<1		
Co	<1	<0.1	0.208	<1		
Ni	<1	<0.1	0.783	<1	90	
other			As, 0.051; B, 0.047; Cd, 0.159; Cr, 0.446; Sb, 0.305; Se 0.277; Sn, 0.202; Te, 0.225; Zn, 0.549		O2, 700; Se, 55; Te, 47	

TABLE 2D Electrowinning Tankhouse Data

	BHP Copper, Miami, Arizona	BHP Copper, Pinto Valley, Arizona	BHP Copper, San Manuel, Arizona	Burro Chief Copper Co. (Phelps Dodge), Santa Rita, New Mexico	Burro Chief Copper Co. (Phelps Dodge), Tyrone, New Mexico	Phelps Dodge Morenci (Arizona) Central Tankhouse
Cathode capacity, tonnes/year	11,800	9500	55,000	68,000	74,000	172,000
1994 cathode, tonnes	11,000	9000	55,000	61,000	69,000	172,000
ANODES						
composition Pb %	81 to 94	81 to 94	98	90 to 99	98.82	98.5
Sn %	1 to 2	1 to 2	1.35	0 to 10	1.1	1
Ca %	0.05 to 0.08	0.05 to 0.08	0.065		0.08	0.5
other	5 to 15%Cu	Cu, 5 to 15%				
rolled or cast	rolled	rolled	rolled	rolled	rolled	rolled
life before replacement, years	5	5	8 to 10	5 to 7	5 (approx.)	7+
blade l x w x t, mm	1270 x 813 x 6.4	1270 x 813 x 6.35	1162 x 953 x 6.35	1270 x 965 x 6.35	1168 x 914 x 6.35	1073 x 889 x 6.3
hanger bar material	copper	copper	copper	copper	copper	copper

TABLE 2D Electrowinning Tankhouse Data (continued)

	BHP Copper, Miami, Arizona	BHP Copper, Pinto Valley, Arizona	BHP Copper, San Manuel, Arizona	Burro Chief Copper Co. (Phelps Dodge), Santa Rita, New Mexico	Burro Chief Copper Co. (Phelps Dodge), Tyrone, New Mexico	Phelps Dodge Morenci (Arizona) Central Tankhouse
bar/blade connection	inlaid	inlaid	slotted bar with solder	welded	slotted hanger bar and welded	soldered
hung symmetrically or asymmetrically	asymmetrically	asymmetrically	asymmetrically	asymmetrically	asymmetrically	symmetrically
anode/cathode spacers: what/where	19 mm hairpin center, 2 25 mm buttons, bottom corner	four PVC hairpins	1 hairpin center, 2 buttons corner	50 mm center vinyl ester capboards	19 mm PVC pipe hairpin	vinyl ester cap blocks, 102 mm anode to anode spacing
CATHODES						
plating time, days	7	5	7	7	7	7
stainless steel blanks or electrorefinery or electrowon copper starting sheets	stainless steel	electrowon starting sheets	stainless steel	electrorefinery copper starting sheets	electrorefinery copper starting sheets	electrorefinery copper starting sheets
Copper Starting Sheets						
length x width x thickness, mm		1067 x 889 x 1.6		953 x 953 x 1.6	953 x 953 x 1.6	953 x 940 x 1.6
mass after plating, kg		82		109	102	92
loading method		ten at a time from strong back with overhead crane		20 at a time, by crane	starter sheet loops on rectangular copper hanger bar	single punch Cu loops, starting sheet hangs 235 mm below hanger bar
Stainless Steel Blanks						
ISA or Kidd	ISA		ISA			
length x width x thickness, mm	1422 x 914 x 3.2		990 x 990			
type of stainless steel	316L		316L			
side edging	CPVC		CPVC			
bottom edging	wax		wax			
hanger bar						
I-beam or rectangular	both		both			
material	copper		I-beam Cu-coated stainless; rectangular - solid Cu			

TABLE 2D Electrowinning Tankhouse Data (continued)

	BHP Copper, Miami, Arizona	BHP Copper, Pinto Valley, Arizona	BHP Copper, San Manuel, Arizona	Burro Chief Copper Co. (Phelps Dodge), Santa Rita, New Mexico	Burro Chief Copper Co. (Phelps Dodge), Tyrone, New Mexico	Phelps Dodge Morenci (Arizona) Central Tankhouse
height, width, length, mm	1232 x 51 x 19					
joint between bar & blade	welded		welded			
mass of Cu plate from each side of blank, kg	50		43			
stripping method, machine or manual	hi-pressure steam		machine			
machine manufacturer	Wenmec		Mesco			
water temperature, °C	110		75			
mother blanks/hour	63					
features						
washing	yes		yes			
stripping	no		yes			
stacking	yes		yes			
sampling	no		no			
weighing	no		yes			
strapping	no		no			
CELLS						
number	102	8 starter, 52 commercial	188	188	224	548
length x width x depth, m	2.64 x 1.09 x 1.63	3.51 x 1.09 x 1.63	6.17 x 1.27 x 1.14	3.86 x 1.17 x 1.36	6.76 x 1.17 x 1.35	6.55 x 1.22 x 1.22
construction material	concrete	concrete	concrete	28 polymer concrete, 160 concrete	concrete and polycrete	376 concrete 172 polymer concrete
lining material	paraliner	paraliner	PVC	PVC in concrete cells	paraliner in concrete cells	paraliner in concrete cells
capping board material	polycarbonate	PVC		vinyl ester	vinyl ester	vinyl ester
intercell busbar type	triangular	triangular	triangular	triangular	triangular	dogbone
x-sectional area, mm^2	628	629	323	988	856	150
current density, A/mm^2	21.15	35		36.4	42.1	

TABLE 2D Electrowinning Tankhouse Data (continued)

	BHP Copper, Miami, Arizona	BHP Copper, Pinto Valley, Arizona	BHP Copper, San Manuel, Arizona	Burro Chief Copper Co. (Phelps Dodge), Santa Rita, New Mexico	Burro Chief Copper Co. (Phelps Dodge), Tyrone, New Mexico	Phelps Dodge Morenci (Arizona) Central Tankhouse
anodes, cathodes/cell	22,21	31,30	61,60	61, 60	61, 60	64,63
cathode-cathode center-line distance, mm	101.6	101.6	95.25	101.6	101.6	101.6
cell inspection system	infrared scanner	infrared scanner	infrared scanner	gaussmeter	gaussmeter and infrared	gaussmeter, IR, water
inspections per day	6	5	3	1	1	1
mist suppression system						
balls	yes	yes	no	yes	9.5 and 19mm	no
BB's	no	no	yes	yes	no	yes (beads)
ventilation	no	no	yes	no	no	yes
FC 100	no	no	yes	no	no	no
FC 1100	no	no	no	yes	yes	yes
other					water sprays	
cell cleaned after (days)	102	180	60	365	180	1 to 2 years
ELECTRICAL						
current per cell, amps	10,000 to 13,300	15,000 to 22,000	36,000	36,000	34,500 (average)	30,000 on two sections, 40,000 on three sections
cathode current density, amps/m^2	237 to 312	246 to 370	323	330	317	315 nominal, 350 max
cathode current efficiency, %	93	91	93 to 94	90	91	90
anode-cathode voltage, v	1.86	2.2	2	2	2.05	2.02
total AC electrical energy consumption (MWh per tonne of cathode)	2.279	2.65 (SXEW) 5.12 including leach	1.8	2.8	1.1	4.85
ELECTROLYTE						
total volume in plant, m^3	760	570	3800	2500	3400	8000
flow of electrolyte to solvent extraction, m^3/day	2500	1400	8540	12,000	12,000	35,000

TABLE 2D Electrowinning Tankhouse Data (continued)

	BHP Copper, Miami, Arizona	BHP Copper, Pinto Valley, Arizona	BHP Copper, San Manuel, Arizona	Burro Chief Copper Co. (Phelps Dodge), Santa Rita, New Mexico	Burro Chief Copper Co. (Phelps Dodge), Tyrone, New Mexico	Phelps Dodge Morenci (Arizona) Central Tankhouse
electrolyte treatment before entering tankhouse	column cell, Disep dual media filter	dual media filter	air flotation, sand filters	column flotation followed by anthracite Disep filters	column cells for organic removal then garnet-anthracite Disep filters	settling tank skim, Jameson cell, packed media filters
electrolyte flow in and out of each cell, m^3/min	0.07	0.099	0.19	0.189	0.208	0.13
flow control system	25 mm PVC inlet valves	106 mm header to 25 mm cell feed line	valve	valve in, adjustable level drain out	none	constant
electrolyte distribution system and hole size, mm	25 mm PVC manifold with 5.6 mm holes angled in 45°	manifold around cell perimeter with 5.6mm holes angled in 45°	manifold with 60 six mm holes	none	50 mm inlet header drops vertically to 150 mm from cell bottom	none
overflow pipe diameter, mm	51	76	–150	102	102	168
entering cells						
Cu, gpl	38 to 40	33 to 35	42	38	38	37
H_2SO_4, gpl	165 to 180	160 to 170	160	180	180	200
temperature, °C	32 to 41	41 to 46	43	49	38	42
temperature control method	boiler	heat exchangers	heat exchange and steam addition	plate heat exchangers	heat exchangers	heat exchangers
leaving cells						
Cu, gpl	36	31 to 33	40	35	34	32
H_2SO_4, gpl	180	178 to 180	163	182	185	210
temperature, °C	32 to 41	41 to 46	44 to 46	53	40	45
other electrolyte components						
Fe^{++}, gpl				1.9 (total Fe)	0.35	0.6
Fe^{+++}, gpl	1.5	3.5	1.5		1.45	2.4
Co^{++}, ppm	120	120	65	180	200	170
Cl^-, ppm	15 to 28	36	16	30	28	27

TABLE 2D Electrowinning Tankhouse Data (continued)

	BHP Copper, Miami, Arizona	BHP Copper, Pinto Valley, Arizona	BHP Copper, San Manuel, Arizona	Burro Chief Copper Co. (Phelps Dodge), Santa Rita, New Mexico	Burro Chief Copper Co. (Phelps Dodge), Tyrone, New Mexico	Phelps Dodge Morenci (Arizona) Central Tankhouse
other, ppm		Mn, 147		Mn, 100, solids 4	Mn, 120	135
guar addition, g per tonne of cathode	300	250	265	237	200	none
cobalt addition, g per tonne of cathode	120	240	to 65 ppm in electrolyte	263	300	310
maximum Mn level	none		kept below 30 ppm		400 ppm levels attacked anode, causing Pb flakes to entrain in cathode. Mn+7 caused phase instability problems in SX circuit.	
electrolyte bleed from tankhouse, m^3/day	196	44	190	136	325	164
bleed to control	Fe^{+++} to < 1.5 gpl	stable ferric level	Fe, Mn	stable Fe and Mn	stable Fe and Mn	stable Fe and Mn levels
destination	E1 mixer settler	E1 in solvent extraction plant	E1 mixer settler	raffinate pond	pregnant leach solution feed pond	raffinate pond
CATHODE COMPOSITION						
Pb	<1	<5	<1	2.4	1 to 2	1 to 2
S	<10	<15	1 to 2	8.6	6 to 10	6 to 8
Fe	<1	<10	1 to 2		2 to 3	0.6 to 1.6
Mn						
Co						0.1
Ni			<1			
other						

TABLE 2E Electrowinning Tankhouse Data

	Phelps Dodge Morenci (Arizona) Southside Tankhouse	Zambia Cons. Cu Mining Chambishi Cobalt Plant Copper Tankhouse	Zambia CCM Nkana Cobalt Plant Copper Tankhouse	Zambia CCM Tailings Leach Plant, Chingola Old Tankhouse	Zambia CCM Tailings Leach Plant, Chingola New Tankhouse
Cathode capacity, tonnes/year	82,000	15,000	13,400	110	thousand
1994 cathode, tonnes	new	12,000	14,300	84	thousand
ANODES					
composition Pb %	98.5	94	94 to 96	94	94
Sn %	1				
Ca %	0.5				
other		Sb, 6%	Sb, 4 to 6%	Sb, 6%	Sb, 6%
rolled or cast	rolled	cast	cast	cast	cast
life before replacement, years	7	3	3	3	3
blade l x w x t, mm	1073 x 889 x 6.3	1120 x 860 x 15		1120 x 880 x 12	1120 x 880 x 12
hanger bar material	copper	copper	copper	anode copper	anode copper
bar/blade connection	soldered	blade cast on hanger bar	blade cast on hanger bar	blade cast on hanger bar	blade cast on hanger bar
hung symmetrically or asymmetrically	symmetrically		asymmetrically	symmetrically	symmetrically
anode/cathode spacers: what/where	vinyl ester cap blocks, 100mm anode to anode spacing	rubber capping boards along length of cell	rubber capping boards along the length of the cell	polypropylene insulator strips, 3 each side of anode	polypropylene insulator strips, 3 each side of anode
CATHODES					
plating time, days		4 to 5	5	4 to 10	4 to 10
stainless steel blanks or electrorefinery or electrowon copper starting sheets	stainless steel	electrorefinery copper starting sheets	electrorefinery copper starting sheets	electrorefinery copper starting sheets	electrorefinery copper starting sheets
Copper Starting Sheets					
length x width x thickness, mm		960 x 960 x 1	1000 x 1000 x 1	950 x 950 x 1	950 x 950 x 1
mass after plating, kg		30 to 35	45 to 48	45 to 50	45 to 50
loading method		put by men manually after hooking	put by men manually after hooking	put by men manually after hooking	put by men manually after hooking
Stainless Steel Blanks					

TABLE 2E Electrowinning Tankhouse Data (continued)

	Phelps Dodge Morenci (Arizona) Southside Tankhouse	Zambia Cons. Cu Mining Chambishi Cobalt Plant Copper Tankhouse	Zambia CCM Nkana Cobalt Plant Copper Tankhouse	Zambia CCM Tailings Leach Plant, Chingola Old Tankhouse	Zambia CCM Tailings Leach Plant, Chingola New Tankhouse
ISA or Kidd	ISA				
length x width x thickness, mm	1245 x 990 x 3				
type of stainless steel	316L				
side edging	CPVC dual durometer				
bottom edging	wax				
hanger bar					
I-beam or rectangular	I-beam				
material	304L stainless				
height, width, length, mm	51 x 29 x 1292				
joint between bar & blade	stainless steel stitch weld				
mass of Cu plate from each side of blank, kg	50				
stripping method, machine or manual	machine				
machine manufacturer	Wenmec				
water temperature, °C	70				
mother blanks/hour	420				
features					
washing	yes				
stripping	yes				
stacking	yes				
sampling	no				
weighing	yes				
strapping	yes				

TABLE 2E Electrowinning Tankhouse Data (continued)

	Phelps Dodge Morenci (Arizona) Southside Tankhouse	Zambia Cons. Cu Mining Chambishi Cobalt Plant Copper Tankhouse	Zambia CCM Nkana Cobalt Plant Copper Tankhouse	Zambia CCM Tailings Leach Plant, Chingola Old Tankhouse	Zambia CCM Tailings Leach Plant, Chingola New Tankhouse
CELLS					
number	220	290	80EW, 78Strip	800	320
length x width x depth, m	6.71 long	4.50 x 1.16 x 1.16	EW 6.56 x 1.06 x 1.09 Strip 3.5 x 1.0 x 1.43	4.37 x 1.09 x 1.32	6.52 x 1.09 x 1.32
construction material	CTI polymer concrete	precast reinforced concrete	precast reinforced concrete	precast reinforced concrete	precast reinforced concrete
lining material	none	lead	lead	PVC and lead	PVC and lead
capping board material	CTI cap block	rubber	rubber/wood	nitrile rubber	nitrile rubber
intercell busbar type	dogbone	rectangular/square	triangular	triangular	triangular
x-sectional area, mm^2	150	175	692	25mm x 25mm x 25mm	50mm x 50mm x 50mm
current density, A/mm^2		62.8 to 74.2	8.67 to 34.68	15 to 20	15 to 20
anodes, cathodes/cell	64,63	33,32; 49,48; 31,30	EW 49,48; Strip 27,26	41,40	61,61
cathode-cathode center-line distance, mm	100	130	100	100	100
cell inspection system	water	manual with sticks	manual with sticks	physical	physical
inspections per day	1	1	2	6	6
mist suppression system					
balls	yes	no	no	polypropylene	polypropylene
BB's	yes (beads)	no	no	no	no
ventilation	no	no	yes	no	no
FC 100	no	no	no	no	no
FC 1100	yes	no	no	no	no
other		Dowfax	Dowfax		
cell cleaned after (days)	2 years	30	EW 40, Strip 9 to 10	90	90
ELECTRICAL					
current per cell, amps	40,000 avg, 45,000 max	11,000 to 13,000	EW 12,000 to 24,000; strip 6,000 to 13,000	6,000 to 12,000	21,000 to 36,000
cathode current density, amps/m^2	320	140 to 204	EW 139-280, St 69-265	100 to 150	100 to 220

TABLE 2E Electrowinning Tankhouse Data (continued)

	Phelps Dodge Morenci (Arizona) Southside Tankhouse	Zambia Cons. Cu Mining Chambishi Cobalt Plant Copper Tankhouse	Zambia CCM Nkana Cobalt Plant Copper Tankhouse	Zambia CCM Tailings Leach Plant, Chingola Old Tankhouse	Zambia CCM Tailings Leach Plant, Chingola New Tankhouse
cathode current efficiency, %	87	70 to 80	EW 70 to 75, Stp 45-66	87 to 90	87 to 90
anode-cathode voltage, v	2.1	2.4 to 2.7	2.0 to 2.5	1.8 to 2.3	1.8 to 2.3
total AC electrical energy consumption (MWh per tonne of cathode)	3.53	1.5 to 4.2	EW 2.6, Strip 3.5	2	2
ELECTROLYTE					
total volume in plant, m^3	15,000	17,000 to 18,000	3000 to 3500	5000	to 6000
flow of electrolyte to solvent extraction, m^3/day	58,500				
electrolyte treatment before entering tankhouse	tank skim, Jameson cell, garnet-sand-anthracite filter	clarification by thickeners	clarification by thickener	two organic removal	flotation columns, series
electrolyte flow in and out of each cell, m^3/min	0.15	0.0166	EW 0.016, St. 0.016-3.5	0.02 to 0.03	0.08 to 0.32
flow control system	commercial electrolyte constant speed pump	valve adjustment	manual valve adjustment		
electrolyte distribution system and hole size, mm	none	single end feed			
overflow pipe diameter, mm	168	50	50	100	100
entering cells					
Cu, gpl	38	20 to 35	20 to 35	45 to 50	45 to 50
H_2SO_4, gpl	200	20 to 25	15 to 30	150 to 170	150 to 170
temperature, °C	43	20 to 30	50 to 60	25 to 35	25 to 35
temperature control method	boiler & heat exchanger	none	none	none	none
leaving cells					
Cu, gpl	34	15 to 25	0.5	30 to 35	30 to 35
H_2SO_4, gpl	205	50 to 60	50 to 60	180 to 200	180 to 200
temperature, °C	45 to 46	60	60 to 70	35 to 45	35 to 45
other electrolyte components					

TABLE 2E Electrowinning Tankhouse Data (continued)

	Phelps Dodge Morenci (Arizona) Southside Tankhouse	Zambia Cons. Cu Mining Chambishi Cobalt Plant Copper Tankhouse	Zambia CCM Nkana Cobalt Plant Copper Tankhouse	Zambia CCM Tailings Leach Plant, Chingola Old Tankhouse	Zambia CCM Tailings Leach Plant, Chingola New Tankhouse
Fe^{++}, gpl	0.6	Total Fe, 1.5 to 3.0	1.5 to 3.0	0.18	0.18
Fe^{+++}, gpl	2.4		<0.5		
Co^{++}, ppm	170	10 to 15 gpl	10 to 15	200 to 300	200 to 300
Cl^{-}, ppm	30	15		30 to 50	30 to 50
other, ppm	Mn, 130			Total Fe, 200 to 800; Ca, 50 to 550; Mn, 30 to 400	Total Fe, 200 to 800; Ca, 50 to 550; Mn, 30 to 400
guar addition, g per tonne of cathode	none	none	none	none	none
cobalt addition, g per tonne of cathode	325	none	none	none	none
maximum Mn level	above 180 ppm excessive small anode flake is produce, necessitating more frequent washing				
electrolyte bleed from tankhouse, m^3/day	885	720 to 1200	1200 to 1400	2	0
bleed to control	stable Fe level	push Co to Co circuit	to recover Co in Co plant	desludge	of cells
destination	raffinate	cobalt plant	cobalt plant		
CATHODE COMPOSITION					
Pb	1 to 2	60	43	<10	<10
S	4 to 6	34	12	<5	<5
Fe	0.6 to 1.4	2	27	<10	<10
Mn			2	0.6	0.6
Co	0.1	2	46	0.4	0.4
Ni		<1	1.8	1	1
other		Bi <1; Sb <1; Se 5	Bi 0.8; Sb 1.0; Se 25	Bi<1; Sb<1; Se<1	Bi<1; Sb<1; Se<1

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Economic Considerations for SX/EW Operations

Toni L. Wallis* **and Gregory F. Chlumsky***

Since the 1960s, solvent extraction/electrowinning (SX/EW) technology, developed for the relatively low cost processing of copper resources, has evolved rapidly and attracted a large share of investment capital, particularly in South America and Australia. This has led to a significant structural change in the copper supply to the world market. The following presentation is based in large part on the study produced by Pincock, Allen and Holt, titled "Review of Current and Potential SX/EW Copper Producers," and on ongoing research to update the contained information.

In 1980, SX/EW copper production contributed approximately 2.0 percent of the total copper mine production, and by 1998, this had increased to approximately 22 percent of total production (see Figure 1). Pincock, Allen & Holt predicts that by the year 2003, SX/EW production of copper could reach 30 percent of total copper production.

The improvement in the technology of production of SX/EW copper over the last 10 years has resulted in the downward pressure on supply costs and has attracted new investment from small- to medium-size mining companies who could not previously compete in the copper business because of the large capital investment required. Major proponents of the SX/EW process claim that the technology has now developed to the point where the copper cathode produced is superior in quality to that produced by the traditional smelting and refining methods.

Hydrometallurgy is the process of using an aqueous solvent to dissolve or leach minerals for ores. In the SX/EW process, the ore is leached usually with sulfuric acid, then the copper is extracted from the acid leach with a solvent, and the copper is subsequently electrolytically refined. Mineralogy is the key to applying SX/EW techniques, and the original ore mineralogy of a deposit determines the ultimate recovery from the leach process. The hydrometallurgy process allows for the recovery of copper from low-grade copper-oxide ores which, because of their chemical composition, are not suitable for concentration by flotation. Leaching can be enhanced by promoting the growth of naturally occurring bacteria which can speed up the oxidation process.

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* Pincock, Allen and Holt, Lakewood, Colo.

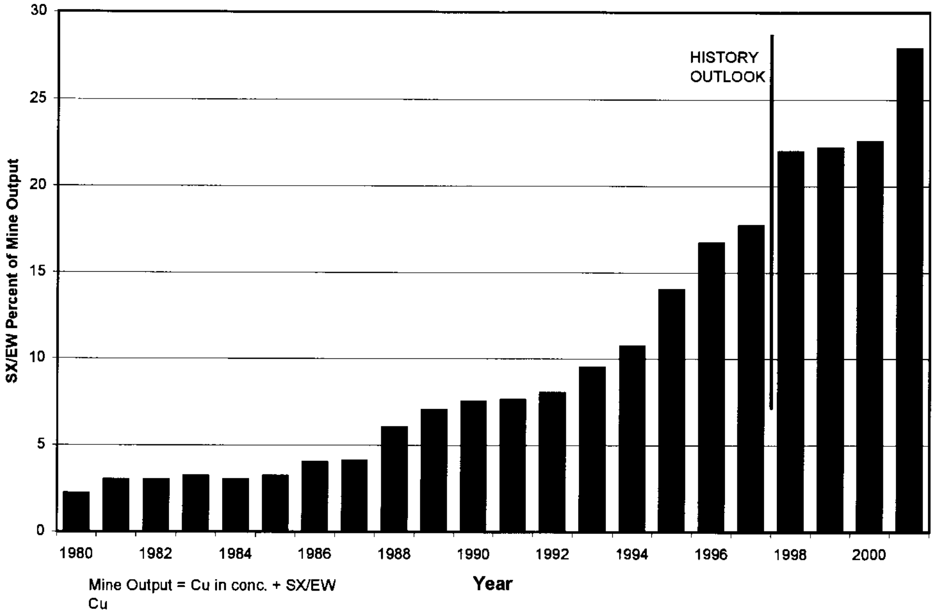

FIGURE 1 SX/EW contribution to mine production—1980–2001

Leaching of pure oxide minerals is the usual scenario; however, some sulfide and mixed ores can be leached. A new innovation in the hydrometallurgical processing of copper is the CESL copper process developed by Cominco Ltd. In this autoclave processing treatment, the copper sulfides are oxidized in an autoclave forming soluble copper and sulfuric acid. The copper is extracted with a solvent and electrolytically refined. Costs are reported to be at US$0.09 per pound versus US$0.24 per pound for conventional smelting and refining.

Unlike the traditional pyrometallurgical recovery of copper from sulfides by smelting, the hydrometallurgical process does not involve any heat transfer, or the possibility of environmental contaminants. The recovery of copper by the hydrometallurgical process of SX/EW occurs in a closed circuit recovery process where wastes and impurities are recycled and recovered. Copper production by the SX/EW method helps control the environmental impact of the mining activities, because operations are exhaust-free and clean solutions are obtained at each stage of the purification process. Transportation of material by pipelines makes handling easier and reduces the risk of pollution.

The great advantage of acid leaching and SX/EW recovery of copper is its low costs. SX/EW production of copper from low-grade material is relatively low cost because no heat transfer is required, processing is on site, and no transportation is required to a refinery. In 1997, the average direct operating cost was US$0.43 per pound of copper. This includes US$0.08 cents for mining and handling, US$0.15 cents for leaching, US$0.18 cents for SX/EW recovery, and US$0.02 cents for general and administrative charges. For copper produced by concentration, smelting, and refining, the costs are around US$0.70 per pound. In 1997, the average direct operating costs were US$0.08 cents for mining and handling, US$0.36 cents for milling, US$0.24 cents for smelting and refining, and US$0.02 cents for general and administrative.

The investment associated with developing a new SX/EW plant averages about US$2,300 per tonne of annual cathode capacity. Expansions of existing facilities average US$1,400 per tonne. In comparison, the capital costs for conventional milling operations ranges from US $3,700 to US$4,500 per annual tonne of copper ore produced, and these figures do not include the capital investment for smelters.

In comparing cost estimates for mill copper ore and SX/EW copper production, in 1997 production costs for SX/EW production are $0.43 which is about $0.09 lower than mill ore. The projection for year 2000 indicates only a marginal increase in costs for SX/EW production. As more of the SX/EW projects come on stream as stand-alone projects, the average costs of this type of production will probably increase since the projects will have to absorb mining and overhead costs. Projects currently being developed are predominantly in South America, and their ore grades are much higher than in other locations.

One advantage that mill copper ore has over SX/EW copper production is that of by-product credits for other contained metals. By-products of copper mining that can be separated out or alternatively contained in the concentrate are one way of improving the cost of production on the balance sheet. Gold and silver can be present in the concentrate, increasing the value of the product for the purchaser, who can in turn separate the precious metal component. Molybdenum is also a by-product of copper production and is produced as a concentrate and sold separately.

Copper operations can only compete on a cost basis since the final product is the same. In the current hard economic times for the copper industry, large copper operations have to scale down their operating costs. The future cost of producing copper by SX/EW is of great significance to the growth of the copper industry. New SX/EW production is at the bottom of the cost curve and has driven out the higher-cost marginal producers. The low-cost, high-volume production provides a buffer on any sustained price increases, and the goal for all producers will be to lower costs in order to stay competitive.

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Checklist for Environmental Compliance

Patricia Acker[*]

INTRODUCTION

Environmental resources are protected from undue degradation under various laws, regulations, and guidelines. There is a broad range of environmental requirements that have evolved in developed countries pertaining to major industrial activities. Many developing countries are now adopting these standards and guidelines or developing their own. Financial institutions also require compliance with international standards and guidelines as a requirement for project financing. Many international mining companies use environmental regulations from their home countries or have developed internal standards to guide sustainable development in countries lacking comprehensive standards.

Contributions to the environmental planning process may be required from specialists in geology, mining, metallurgy, chemistry, hydrology, and engineering, as well as from specialists in the environmental sciences, human relations, finance, law, and management. Each discipline may be asked to contribute to or comment upon studies from other disciplines. This symposium is directed at the metallurgical aspects of the copper SX-EW business. In reviewing this environmental checklist, the reader (more than likely a metallurgist—nobody's perfect) should consider how his or her expertise may be relevant (aside from the obvious process issues).

For most projects, environmental activities will represent a substantial investment in time and money throughout the duration of the development, operation, closure, and postclosure phases of a project. The costs for site visits and technical and management services should be anticipated, estimated, monitored, and controlled to the fullest extent possible from the start. This project checklist is intended to define the broad scope of environmental studies, planning efforts, and management actions that may be required for a modern-day mining project.

A general list of references of world standards for environmental issues is presented in Table 1.

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[*] Senior Scientist, SRK Consulting, Denver, Colo.

TABLE 1 References for world standards for environmental compliance

Environmental Impact Assessment		
World Bank	Operational Directive 4.01 "Environmental Assessment" October, 1991	Mandates an environmental assessment for all projects seeking financing that may have significant impacts on the environment.
Air Quality		
World Bank/IFC	Environment, Health and Safety Guidelines, Mining and Milling—Open Pit, August, 1995	Numeric guidelines for ambient and workplace air quality at open pit and milling operations.
World Bank/IFC	Environment, Health and Safety Guidelines, Thermal Power Plants, October, 1996	Numeric guidelines for ambient air quality measured outside the project property and for concentrations of contaminants emitted from the stacks of thermal power plants.
Environmental Noise		
World Bank Group	Pollution Prevention and Abatement Handbook, September, 1997	Numeric guidelines for ambient noise limits measured at noise receptors located outside the project property boundary.
World Bank/IFC	Environmental Guidelines—General September, 1995	Numeric guidelines for ambient noise limits measured at noise receptors located outside the project property boundary.
Water Pollution/Wastewater Discharge		
World Bank/IFC	Environmental Guidelines—General September, 1995	Numeric guidelines for effluent discharges of process wastewater, domestic sewage, contaminated stormwater and runoff, and leachate from a solid waste disposal site.
World Bank Group	Pollution Prevention and Abatement Handbook, September, 1997	Numeric guidelines for base metal mines.
World Bank/IFC	Environment, Health and Safety Guidelines, Mining and Milling—Open Pit, August, 1995	Numeric guidelines for effluent discharges from tailings impoundments, mine drainage, sedimentation basins, sewage systems, and stormwater discharge.
Water Quality		
World Bank/IFC	Health and Safety Guidelines—General, September, 1995	Recommendations for the use of the World Health Organization drinking water guidelines for project water supplies.
World Health Organization	Guidelines for Drinking Water Quality, Second Edition, 1993	Numeric drinking water guidelines based on human health and aesthetic quality.

TABLE 1 References for world standards for environmental compliance (continued)

Soils/Reclamation		
World Bank/IFC	Environment, Health and Safety Guidelines, Mining and Milling—Open Pit, August, 1995	Erosion and sediments control and reclamation plan guidelines.
Flora and Fauna		
World Bank	Operational Directive 4.04 “Wildlands: Their Protection and Management in Economic Development”	Conservation of endangered plant and animal species, critical habitats and protected areas, wetlands, wildlands, and tropical forests.
World Bank/IFC	Environment, Health and Safety Guidelines, Mining and Milling—Open Pit, August, 1995	Requirements for project sponsors to prepare and implement a mine reclamation plan. Recommendations for use of native vegetation to sustain a productive ecosystem.
Socioeconomics/Land Use		
World Bank/IFC	Environment, Health and Safety Guidelines, Mining and Milling—Open Pit, August, 1995	Requirements that a project sponsor provide information on social issues related to the project as part of the environmental information report.
Cultural Resources		
World Bank	Operational Directive 4.25, “Management of Cultural Property in Bank-Financed Projects”	Protection of archaeological sites, historic monuments, and historic settlements.
Health and Safety		
World Bank/IFC	Health and Safety Guidelines—General, September, 1995	Recommends the use of standard international worker safety practices, including numeric values for workplace ambient air quality and noise level limits, safety procedures for confined spaces, general health and safety practices, drinking water criteria, employee training, health monitoring, record keeping, and accident reporting procedures.
World Bank/IFC	Environment, Health and Safety Guidelines, Mining and Milling—Open Pit, August, 1995	Recommends health and safety practices specific to liquid effluents, ambient air, erosion, reclamation, sewage, solid waste, sewage, solid waste, and hazardous materials.

TABLE 1 References for world standards for environmental compliance (continued)

Solid and Hazardous Waste		
World Bank Group	Pollution Prevention and Abatement Handbook, September, 1995	Recommendations for tailings impoundment design and operational and postclosure management.
World Bank/IFC	Environment, Health and Safety Guidelines, Mining and Milling—Open Pit, August, 1995	Defines handling, storage, and disposal methods for hazardous materials, tailings, sewage sludge, and other solid wastes from open pit mines and milling operations. Numeric guidelines for liquid effluent discharges from tailings impoundments, mine drainage, sedimentation basins, sewage systems, and storm water drainage.
World Bank/IFC	Health and Safety Guidelines—General, September, 1995	Recommendations for the use of standard industry practices for hazardous materials including labeling, storage, record keeping, spill and fire prevention, and emergency planning.

ENVIRONMENTAL IMPACT ASSESSMENT

Most countries require the preparation and submittal of an environmental impact assessment. Environmental baseline studies form the basis for the impact assessment. Environmental impact assessments review the potential environmental impacts of a proposed project and identify mitigation strategies to alleviate potential significant impacts. The assessment procedure defines the regulatory environment and the baseline conditions at the site. This assessment often includes environmental and socioeconomic baseline studies. The final result of this process is the definition of an environmental plan which, in turn, provides the decision-base for the granting of permits and approvals to proceed with the project.

ENVIRONMENTAL AND SOCIOECONOMIC BASELINE STUDY

A substantial up-front investment in baseline environmental information will be needed to successfully move a proposed project through the permitting phases and into construction and operation. A thoroughly documented project database will usually pay dividends in not only identifying concerns, but in more rapidly developing workable solutions and in negotiating and expediting agreements early in the project. A comprehensive database will also be helpful in resolving unforeseen problems or conflicts later in the project.

Independent environmental studies are usually conducted to provide baseline reference information for leaching, solvent extraction, and electrowinning copper projects. The objectives of this work include:

- Identification and quantification of preexisting and ongoing environmental issues that may be present on and near the project site and that are potentially related to mining activities

- Delineation of areas that may be unsuitable for future disturbance or project development due to environmental constraints
- Development of sufficient information to form a basis for evaluating a new or expanded mining operation with respect to acceptable environmental standards
- Provision of practical interpretation and application of company, local, national, and international rules, regulations, and permitting requirements that may apply to the site and to the project
- Estimation of costs for complying with environmental laws, permits, and project agreements and costs for mitigating existing and future environmental problems

Essential topics for a baseline study include:

- *Regional Site Characteristics.* A description of the regional and local history and geology of the project area is required to provide supporting information for evaluating mine waste and surface and groundwater resources in the project area. The seismicity of the site and the presence of any natural geologic hazards such as the potential for landslides and mudslides will also be summarized. A description of the topographic features of the site provides background information for surface water hydrology and information required for evaluating the potential impacts to visual resources.
- *Meteorology and Air Quality.* A comprehensive meteorological evaluation of the region and the project site is needed to assess the current conditions in the region and for compiling the necessary data for an air quality modeling study. Local and regional climatological data will be summarized in the report for general site characterization purposes and for air quality modeling purposes.
- *Noise.* Typical mining, processing, and associated operations produce noise that may disturb off-site receptors (i.e., structures, animal populations, and local residents) and on-site workers and facilities. Noise will be generated from drilling, blasting, mobile equipment, generators, pumps, and other equipment as well as from various human activities on-site or adjacent to the site.
- *Surface Water Resources.* The objective of a surface water baseline collection program is to establish background and preacquisition (also preproject) surface water quality, flow dynamics, and current appropriations and use allocations.
- *Groundwater Resources.* The objective of the groundwater baseline review is to establish existing preproject groundwater quality and flow conditions as well as current appropriations and use allocations.
- *Soils.* Baseline information concerning the soils of the project site will be obtained through soils mapping and sample collection to provide information on soils characteristics. A representative selection of soil samples will be tested and characterized to determine future use during reclamation.
- *Mine Wastes.* Baseline information required to characterize mine waste is collected from drill samples with assistance from exploration and mine geologists to ensure representative samples and data. Similar programs collect samples for metallurgical testing. Samples are tested for total metals and sulfides and acid-base accounting tests (static columns for acid generation, etc.) and for leachable metal content.

- *Flora and Fauna.* A field study is usually conducted to identify vegetation and wildlife in the project area. A map identifying various plant communities and important habitat or ecological communities is prepared to analyze existing and future impacts to each community. An aquatic and terrestrial wildlife survey of the project area is also carried out to identify rare, unique, and endangered species.
- *Socioeconomics.* A baseline socioeconomic analysis is developed to define the local economy, local demographic conditions, housing, and public infrastructure. The analysis will describe social and cultural conditions in the area and will also describe the potential impact of project activities (improved roads and other infrastructure) within the area.
- *Land Use.* A land use map for the area is prepared that will define the locations of rock dumps, leach facilities, plant facilities, and related ancillary buildings and infrastructure. The site plans are then related to buildings, agriculture, forests, and open lands. Land values are also quantified.
- *Cultural Resources.* The baseline cultural (historical and archaeological) inventory presents information to establish the presence or absence of significant cultural sites within the potential disturbance areas.

ENVIRONMENTAL IMPACT ASSESSMENT REPORT

An Environmental Impact Assessment (EIA) Report is prepared from these baseline studies. The purpose of the EIA is to identify potential environmental liabilities and to provide guidance to minimize the potential for adverse affects of the mining and processing operations on the environment. The recommended components of the report vary depending upon the agencies involved. Following is the recommendation from the World Bank for the components of an EIA:

- Executive summary
- Policy, legal, and administrative framework
- Project description
- Baseline data
- Environmental impacts
- Analysis of alternatives
- Mitigation plan
- Environmental management and training plan
- Environmental monitoring plan
- Supporting documentation
    - —References
    - —Database (species lists, laboratory data sheets, etc.)

The report should be concise and focus upon significant environmental issues. It should be written in layman's terms as much as possible as it is addressed to lending institutions, permitting agencies, and the general public. The resulting impacts detailed by the plan, whether adverse or beneficial, will form the basis for the environmental management plan.

ENVIRONMENTAL MANAGEMENT PLAN

Compliance with environmental regulations and mitigation of environmental impacts is possible through ongoing environmental management during project operations. The environmental management plan generally becomes the core agreement upon which environmental permits are agreed upon and issued. The following general categories will be defined and implemented during the term of the project and beyond:

- Environmental Management Plan
- Overall Site Monitoring Plan
- Surface Water Management Plan
- Health/Hygiene and Safety Plan (HASP)
- Spill Prevention, Control, and Countermeasure Plan (SPCC)
- Emergency Response Plan (ERP)
- Waste Management Plan
- Mine Waste Management Plan
- Explosives Management Plan
- Wastewater Treatment and Management Plan
- Reclamation and Closure Plan
- Postreclamation Monitoring Plan

The primary goal of environmental monitoring is to determine the construction and operational activities that may cause significant environmental changes, to minimize the potential for unanticipated impacts, and to evaluate the effectiveness of the proposed countermeasures to minimize impacts. The plan will also provide the proposed monitoring protocols to detect environmental changes. If significant adverse environmental changes develop as a result of project operations, then appropriate remedial measures or corrective actions will be implemented to minimize adverse effects.

The monitoring and management plans will include standards for the mine and plant facilities, waste dumps, access roads, water quality management, air quality management, reclamation, and health and safety. Reclamation plans will be implemented whenever the project assumes a stand-by condition or reaches closure.

ENVIRONMENTAL COMPLIANCE AND ENVIRONMENTAL MANAGEMENT

A modern-day copper leaching, solvent extraction, and electrowinning project must comply with increasingly complex and comprehensive rules and regulations governing the mining and processing operations. Potential environmental impacts can often be minimized through standard mitigation measures and by proper environmental management.

TYPICAL MINING PERMITS

Permit requirements for mining projects vary from state to state and from country to country. All new projects and expansions of existing projects will likely encounter new permitting requirements. The following permits are typically required for mining projects in developed countries.

- Surface mining permit/reclamation plan
- Wastewater discharge permit
- Stormwater discharge permit
- Land application of process water permits
- Permit authorizing construction in wetlands
- Stream channel authorization permit
- Air quality permit
- Dam safety permit
- Solid waste disposal facility permit (landfill)
- Hazardous waste permit

In addition to these environmental permits, there are generally numerous permits related to construction, transportation of equipment to and from the site, operations and licenses, importation and licenses, and so on. This regulatory environment may change with local, regional, national, and international political trends. This checklist, therefore, should only serve as an advisory. Be prepared and be committed to explain to all interested parties and to the interested public, to the clearest extent possible, the risks and benefits of the project.

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Managing Zaldivar— Placer Dome's Approach to Project Management

M.A. Creaney,[*] J.J. Thompson,[*] and T.J. Smolik[*]

INTRODUCTION

As the competition increases globally between the major mining companies, the impetus continues to build mines at the lowest cost with the shortest construction schedule. Placer Dome Inc. (PDI) has met that challenge by building mines around the world at record pace.

The strategic advantage that Placer Dome Inc. has in developing its own mines is well recognized. Every aspect of development, from exploration, resource modeling, reserve determinations, metallurgical evaluations, feasibility studies, through the design construction process and into operations are all managed in-house by Placer Dome personnel.

The main aspects of PDI's mine development process are structured so that they are not discrete activities but continuing sequences. As an example, the operation management group are very involved in the feasibility study and participate in the process of "buying into" the concepts that are developed. When subsequent production decisions are made, the key design/construction/operating personnel are on board and the targets have been set to build the mine in the most expedient manner possible (see Figure 1).

This paper will describe the three main processes which Placer Dome's project development division (PDD) utilizes in building mines. Those processes will demonstrate their relationships with exploration, strategic development and operations in carrying out the mine development aspects of our company's business. The paper will also describe how the Zaldivar SX/EW was organized from a design and construction perspective, including the associated assets for developing water and power supplies.

EVALUATIONS GROUP

The ability to carry out evaluations and feasibility studies in-house is one of the project development division's greatest strengths. This strength is based on teamwork and collaboration with the resource estimation group of the exploration division and support from other divisions within Placer Dome Inc.

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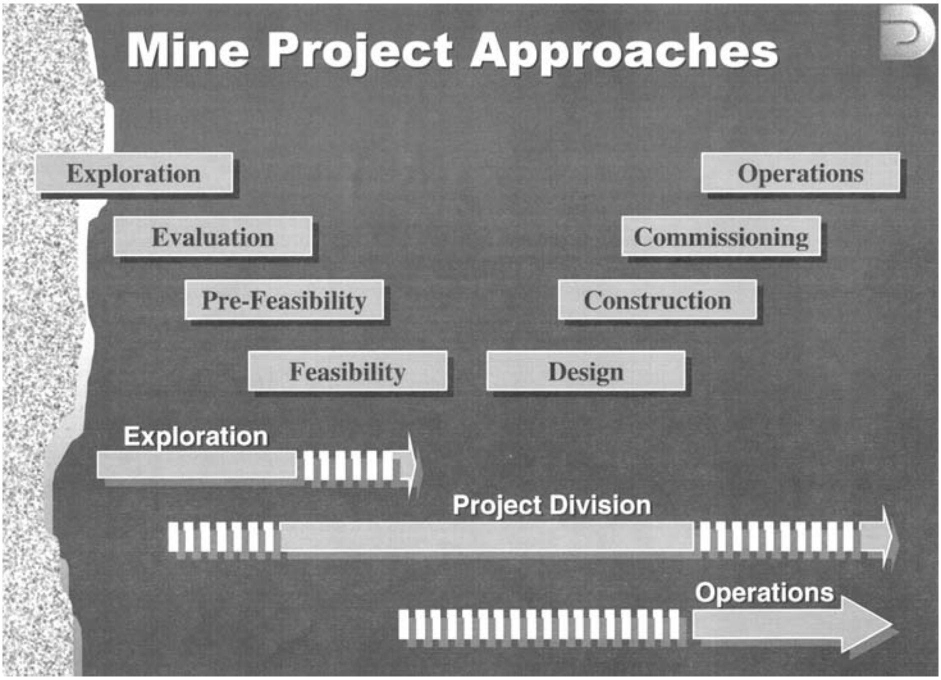
[*] Placer Dome, Inc., Vancouver, B.C.



Mine Project Approaches

Exploration

Evaluation

Pre-Feasibility

Feasibility

Design

Construction

Commissioning

Operations

Exploration

Project Division

Operations

FIGURE 1 **Mine project approaches**

One of the first actions is the selection of a strong leader to manage a feasibility study, paramount to the success of the study and subsequent mine development. The feasibility study manager must have experience with previous feasibility studies and bring an extensive knowledge base to the project. The manager works with a team of up to 20 people, including site geologists, consultants and all Vancouver-based office engineering disciplines within our division.

To complete the team, a mine manager is appointed who then selects his operating superintendents. The operational involvement at an early feasibility stage establishes a plan which carries through the study phase, into design, construction and operations. The operators' participation develops ownership because they've been involved throughout the process and are aligned to execute the study action plan when the investment decision is made.

Additional benefits result because our employees are given the opportunity to enhance their knowledge and career goals. Staff are encouraged to transfer between our division into the design and construction groups, as well as externally to the exploration division and our operating mines. This deliberate movement of staff benefits both the employee and the company as the knowledge and experience will be transferred into other divisions of our company.

Our expertise and procedures have also been transferred to smaller project development teams located in offices in Sydney, Australia, and Santiago, Chile. These separate project development groups contribute significantly to Placer Dome's overall evaluation process and complement the expertise of PDD's evaluation team through collaboration and knowledge sharing.

One significant benefit of PDI's in-house evaluation process is the ability to respond quickly to opportunities. In today's world, with the competition for world class ore bodies, we can rapidly draw on decades of evaluation and mine development experience. This knowledge base was acquired from previous feasibility studies and through the construction of 30 mines and expansions since the 1960s.

The number of projects has given us another powerful tool with the development of a capital cost database. This large database allows us to use updated information to produce capital cost estimates for new evaluations or feasibility studies using conceptual flowsheets, new equipment lists and general layouts.

Another procedure used in our evaluation analysis is a "risk analysis" that we carry out to detect "fatal flaws." This analysis could detect flaws or opportunities in the quality of a resource/reserve or environmental issues involving acid rock generation or tailing disposal concepts.

The Research Centre, functioning as part of our technology group, is an extremely important component together with Placer Dome Inc.'s exploration division and our evaluations group. The Research Centre supports exploration activities either by assaying core samples from the project or by providing a check assay program for samples assayed elsewhere. Concurrently, the metallurgists initiate a test program to determine at a very early stage how the ore will respond metallurgically, and also determine if there are any major problems to consider prior to developing the metallurgical flowsheet.

The Research Centre can also set up sample preparation laboratories at our major exploration sites, train local personnel in sample preparation and perform our own QA/QC at these facilities. The Research Centre has developed a reputation for excellence through its high standards in assaying and metallurgical testwork and has become ISO9002 certified.

The Research Centre is also part of the metallurgical team, completing important testwork on development projects. The laboratory has its own pilot plant facilitates with capabilities for flotation, autoclaving and subsequent leaching of ores or concentrates. Extensive column leach tests have been done for heap leaching operations and we have also developed in-house expertise in autoclaving. Additionally an autoclave pilot plant, including an SX/EW facility, is available for doing research on concentrates or complex refractory gold ores.

DESIGN AND CONSTRUCTION

Design

The design and construction group of the project development division includes metallurgical, electrical, civil and mechanical engineers. The group develops conceptual flowsheets, feasibility concepts for the processing and ancillary facilities and manages the detailed design of the facilities once approval has been given to commence a mining project.

Similar to the evaluation group within our division, there is an overlap of duties as our involvement includes a strong working relationship not only with the feasibility group but also with construction and operational activities of our business.

The conceptual engineering is undertaken by individuals who have extensive experience in plant operations, maintenance and plant design. The design group is responsible for the metallurgical flowsheets, the electrical power systems and all other ancillary facilities and processes. The group also works closely with the environmental engineering group of our sustainable development division, and other environmental and geotechnical specialists, to ensure that the mine design incorporates the highest environmental standards.

When a production decision is made, specific teams are assigned to manage the project's detailed design aspects. A project coordinator is chosen and a core team is assembled comprising a metallurgist, a civil/mechanical engineer, an electrical/instrumentation engineer and a purchasing agent. Often the operations, mill and maintenance superintendents are also assigned to work with the design team.

Depending on the project size, we may have one or two major consultants supplemented by other specialists who have specific expertise in power generation, high voltage transmission lines, tailing dam design or water sourcing from borefields. The design consultants work efforts are coordinated by PDD's design team to produce timely, effective engineering results. Our division has also developed in-house design standards and specifications to ensure that Dome's facilities reflect the experience gained on several projects.

During the later construction stages, and the commissioning phase, the design group continues its involvement in the project with specific site assignments. Again, the focus in managing the plant commissioning is to maintain the strong partnership between project development and operating personnel. A senior member of the project development division is typically assigned the responsibility for managing the plant commissioning. This individual will manage the plant commissioning efforts and facilitate any required changes.

Construction

The construction group is responsible for the construction of all Placer Dome Inc.'s major projects and also contributes to the development of most major feasibility studies. Through the years we have developed an experienced and adaptable core team of people able to set up construction projects and build mines anywhere in the world. This group includes an international collection of Placer Dome people who have built mines in Papua, New Guinea; Australia, Chile, the United States and Canada.

With appropriate standards and systems to support the projects, we can act immediately upon project approval and complete the projects quickly and effectively. Placer Dome's mine building success was demonstrated in 1997 with the completion of two new mines in Canada and the United States within a time period of 12 months.

The construction team has developed standards that are now used on all PDI projects. These standard building blocks are:

- Standard contract packages
- Detailed systems and procedures
- Project management information system
- Project implementation programme

Standard Contract. The standard contract is one of the most significant issues in PDD's mine-building success. For many years, various forms of contract documents were used based on recommendations from consultants involved in the individual projects. Using different forms of contracts was a major disadvantage in that there was no continuity in the management of contracts. Contractual experience gained on one project was not used effectively to improve the next project. Additionally, contractual disputes were adding to the uncertainty of meeting budgets, even after firm-priced contracts were awarded. Mr. Gordon Leck of Gordon E. Leck & Associates from Brisbane, Australia, developed a standard contract document which was introduced in 1989. The contract:

- Avoids or resolves problems on a timely basis and protects the interest of Placer Dome Inc.
- Sets out an explanation of the tendering and contractual requirements such that the document also serves as a policy and procedures manual for contractors
- Standardizes all forms required for tendering and administration of the contract
- Enables use of the contract on any project worldwide with minimal alteration for various legal requirements

Systems and Procedures. The division has also developed a standard systems and procedures manual for PDD's construction projects to ensure quality and consistency in all aspects of the project. The manual allows us to maintain a high degree of internal control within our organization, even in countries where principles and practices vary greatly.

The systems and procedures manual is divided into five volumes. Each volume addresses a branch of the construction organization and is designed to be a "stand-alone" document with a minimum of cross-referencing between volumes. The volumes cover:

- Project management
- Construction management
- Budget and cost control
- Purchasing
- Logistics, administration and accounting

Utilizing these procedures, an employee with experience at other Placer Dome projects can expect the same basic procedures on any project. Our system improves with every project as procedures continue to be streamlined to improve efficiencies and incorporate technology.

Project Management Information System. The project management information system (PMIS) is the automated reporting system and control used for all projects. The system provides automated functions for project cost control and reporting, procurement, expediting, warehousing, and accounting. The system consists of several modules, each with a specific function. The modules are integrated into a main selection menu, which makes the integration transparent to the users.

The primary reporting document is the budget status report where information is compiled and summarized to give the current financial status of the project. This information, in conjunction with schedule information and input from management personnel, forms the basis of the project manager's assessment of project status.

Project Implementation Programme. The project implementation programme was developed based on our experience with numerous projects. This programme lists everything that needs to be done in a sequential order. This programme starts four months before a project is expected to be approved and runs three months into the project. The programme serves as a guide for the project manager and addresses a wide range of activities required for a project such as:

- Project agreements
- Project and regional specifics
- Permits and licenses

- Early design coordination
- Finance and administration
- Contract tendering and administration
- Purchasing and logistics
- Early construction activities
- Personnel mobilization

We have three main training tools to assist in the successful start-up of projects:

- Project action plan
- Team building
- Training programmes

The project action plan and project schedule are developed during an ESP (engineered systematic planning) seminar. This process was developed by Mr. Terry Lee of the Simons Mining Group in Vancouver, who leads the sessions. Key personnel from the design consultants, project management and operations attend these sessions. The action plan is a high level report which addresses issues in a broad perspective as they relate to the long-term development of the project.

At the start of the project, key personnel attend a team building seminar to develop goals and commitments for the new project. Personnel from the Vancouver design coordination team, project management team, as well as operations, participate in this exercise.

Training is a major focus during the first three months of construction. Training is coordinated by the Vancouver office using personnel with experience from several of our projects. During the first weeks on the project site, construction employees spend considerable time learning about systems and our methods for building mines. They are required to attend several induction and training sessions to become familiar with policies and procedures. The induction programme is tailored for personnel from management, field supervisors and clerical staff. Sessions are generally one hour in length. They cover a wide range of subjects, from a warehouse overview for field supervisors, to detailed sessions for those managing individual departments. Training for project management system users is conducted in small groups on an individual basis.

The strong construction role of our division has led to tight control of the costs and schedule and, at the same time, provides enough flexibility to be innovative when the need arises. It has also given us a closer working relationship for our clients, the operating groups. We are committed to a transparent delineation between operating and construction forces and recognize that the project is not a success if the mine is not a success, once construction is complete. The operational personnel are an integral part of the construction team and share the responsibility of ensuring construction success.

On the last two projects at Pipeline and Musselwhite, a stronger awareness of our environmental responsibility as a construction group was developed. We also focused on improving the construction safety record which was evident by both projects achieving one million man-hours of work without a lost time accident.

ZALDIVAR—OVERVIEW

The Zaldivar Mine is located in the 2nd Region of Chile approximately 1,000 kilometers to the north of Santiago and 196 kilometers to the southeast of Antofagasta in the Atacama Desert at an altitude of 3,100 meters. Please see site map Figure 2.

FIGURE 2 **Zaldivar site map**

The Atacama is one of the driest parts of the world with precipitation averaging 4.5 mm per year and evaporation in the order of 3.2 meters annually. The extremes of temperature vary between −7°C in winter and +25°C in summer, with an average temperature of +10°C.

The prevailing winds generally blow from the east and northeast at a velocity of 20 to 22 kilometers/hr, with occasional gusts up to 70 kilometers/hr resulting from fronts generated by a local phenomena known as the "Bolivian Winter."

The seismic zone is UBC Zone 4 with one major earthquake, which measured at 6.5 on the Richter scale, experienced during the course of construction activities.

Mining

Two separate ore zones are scheduled to be mined during the 19-year mine life. A lower grade mine area (Pinta Verde) was mined initially to provide oxidized ore to the facility while the higher grade main zone was stripped of waste. The initial ore reserves at Zaldivar included 320 million tonnes of copper ore averaging 1.07% copper with a waste:ore strip ratio of 2.75:1. The ore mineralogy in this orebody is porphyritic in nature with the chief copper mineralization being chalcocite and brochantite. A deeper, primary sulphide orebody exists which may be mined in the future.

Run-of-mine ore is crushed, conveyed and placed into lifts for leach solution application and eventual copper recovery. During the crushing process, a fines fraction of the ore is removed and treated by flotation for recovery of additional copper values.

Coarse Crushing

The crushing plant was designed to crush 40,000 tpd of copper ore. This crushing plant consists of one Nordberg 1,370 mm × 1,800 mm gyratory primary crusher and two Nordberg MP 1000 secondary crushers. The secondary crushed product reports to a 10,000 tonne live coarse ore stockpile.

Fine Crushing

Tertiary crushing is accomplished with four Nordberg WF800 water flush crushers. Design tonnages in excess of 40,000 tpd were achieved after the construction of a prescreening plant which removed large quantities of product-size material. This facility removed the fines from the tertiary crusher feed and conveyed them directly to the leach pad feed conveyor.

The product of the tertiary crushers is transported to the leach pads by two overland conveyors. The crushed ore is spread on the leach pads by either a RAHCO mobile stacker for oxide ore or a Krupp bridge stacker with portable grasshopper conveyors for sulphide ore.

Solvent Extraction/Electrowinning

The leach solutions from the leach pads are collected in a lined pond from which the decant flows by gravity to the solvent extraction plant. From the solvent extraction plant, electrolyte is fed to the electrowinning plant which is designed to produce 125,000 tonnes of electrowon cathode copper each year. The cathode product is then freighted by rail from the site to the port of Antofagasta.

Water Supply

Water for the mine is piped from the Negrillar borefield located approximately 100 kilometers from the mine. The pipeline is a combination of 24″ and 26″ carbon steel pipe which was installed in less than 9 months. There are six, multistage submersible pumps used for water sourcing. The system is designed to pump water quantities of 400 L/sec with the capability of expansion to 500 L/sec.

Power Supply

Power for the mine is supplied from the El Crucero Substation located 240 kilometers to the northeast of the plantsite. The transmission service is provided in a dual circuit 220 kV line supplying power to both Zaldivar and Escondida. It was constructed on schedule in a period of 11 months.

Access Road

Access to the mine is via a 176-kilometer arterial road linking the port city of Antofagasta to the mine. A section of the road (94 kilometers) was constructed as part of the Zaldivar Project. There is also access to the mine by aircraft, with an airstrip constructed by Zaldivar for Escondida and located near the Escondida Mines.

Operations Camp

The Zaldivar permanent mines operations camp provides accommodation for a maximum of 846 personnel and contractors. The mine roster has been arranged so that only 50% of the operation beds are utilized at any one time.

Zaldivar Design

Due to the size of the project and the specialized design areas incorporated therein, it was decided to have the design done in both Chile and Canada using the expertise available in both countries. The design done offshore in Canada included:

- Fluor Daniel Wright: solvent extraction plant, electrowinning plant, tank farm heap leach conveying and stacking systems
- Bateman Engineering: primary/secondary and tertiary crushing facilities design
- Conveyor Dynamics: Overland conveyor control and braking system design
- J.A. Kahne: electrical and instrumental for the crushing, conveying and SX/EW plant
- Fransen Engineering: electrical and instrumentation for the Negrillar water system and the raffinate system
- Ian Hayward Int.: El Crucero to Zaldivar 220 kV transmission line

The onshore design in Chile included:

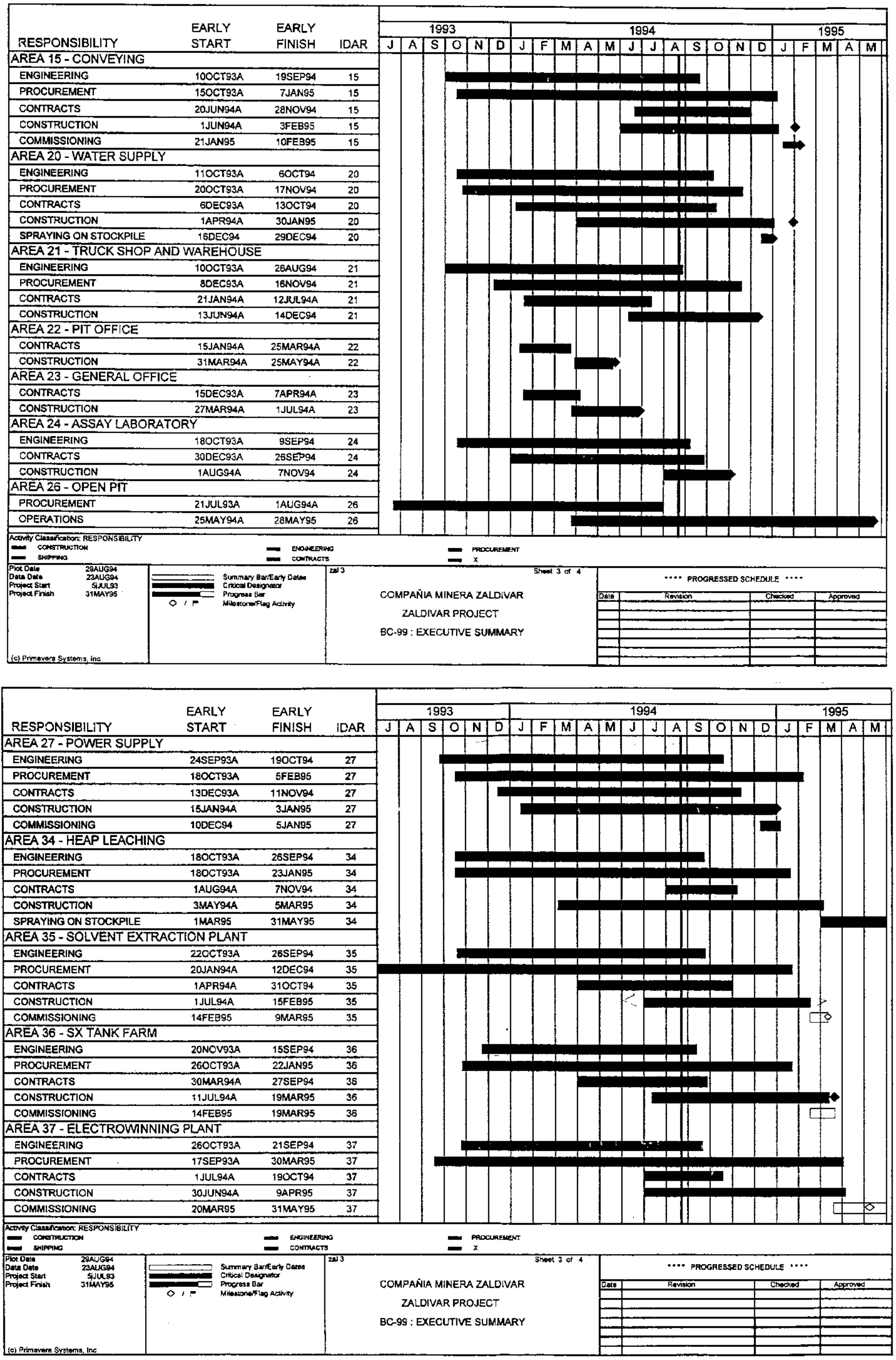
- Indepro: (a 100% subsidiary of Outokompu Oy of Finland) designed the construction water supply system, El Crucero and Zaldivar substations and provided miscellaneous structural and electrical design
- E.C. Rowe: tailing system in association with Bruce Geotechnical in Vancouver
- Brotec: La Negra–Zaldivar access road
- Esedei: site power distribution
- Juan Rayo: Negrillar water system mechanical and hydraulic design

This may appear to be a large group to manage and coordinate; however, it was accomplished through close cooperation between the site construction group and the design management teams located in Vancouver and Santiago.

Construction

The plant and all infrastructure were constructed from September 1993 to February 1995 followed by a five-month commissioning period. The project schedule overview is presented in Figure 3.

The bulk of construction activities were completed on schedule in a period of 18 months. During this time, construction accommodations ranged from a low of 700 to a peak of 3,500 personnel housed in camps at Zaldivar, Negrillar, Baquedano and El Crucero.



RESPONSIBILITY	EARLY START	EARLY FINISH	IDAR
AREA 15 - CONVEYING			
ENGINEERING	10OCT93A	19SEP94	15
PROCUREMENT	15OCT93A	7JAN95	15
CONTRACTS	20JUN94A	28NOV94	15
CONSTRUCTION	1JUN94A	3FEB95	15
COMMISSIONING	21JAN95	10FEB95	15
AREA 20 - WATER SUPPLY			
ENGINEERING	11OCT93A	6OCT94	20
PROCUREMENT	20OCT93A	17NOV94	20
CONTRACTS	6DEC93A	13OCT94	20
CONSTRUCTION	1APR94A	30JAN95	20
SPRAYING ON STOCKPILE	16DEC94	29DEC94	20
AREA 21 - TRUCK SHOP AND WAREHOUSE			
ENGINEERING	10OCT93A	26AUG94	21
PROCUREMENT	8DEC93A	16NOV94	21
CONTRACTS	21JAN94A	12JUL94A	21
CONSTRUCTION	13JUN94A	14DEC94	21
AREA 22 - PIT OFFICE			
CONTRACTS	15JAN94A	25MAR94A	22
CONSTRUCTION	31MAR94A	25MAY94A	22
AREA 23 - GENERAL OFFICE			
CONTRACTS	15DEC93A	7APR94A	23
CONSTRUCTION	27MAR94A	1JUL94A	23
AREA 24 - ASSAY LABORATORY			
ENGINEERING	18OCT93A	9SEP94	24
CONTRACTS	30DEC93A	26SEP94	24
CONSTRUCTION	1AUG94A	7NOV94	24
AREA 26 - OPEN PIT			
PROCUREMENT	21JUL93A	1AUG94A	26
OPERATIONS	25MAY94A	28MAY95	26

Activity Classification: RESPONSIBILITY
CONSTRUCTION
SHIPPING
ENGINEERING
CONTRACTS
PROCUREMENT
X

Plot Date 29AUG94
Data Date 23AUG94
Project Start 5JUL93
Project Finish 31MAY95

Summary Bar/Early Dates
Critical Designator
Progress Bar
◇ / ▸ Milestone/Flag Activity

zal 3
Sheet 3 of 4

COMPAÑIA MINERA ZALDIVAR
ZALDIVAR PROJECT
BC-99 : EXECUTIVE SUMMARY

**** PROGRESSED SCHEDULE ****

Date	Revision	Checked	Approved

(c) Primavera Systems, Inc

RESPONSIBILITY	EARLY START	EARLY FINISH	IDAR
AREA 27 - POWER SUPPLY			
ENGINEERING	24SEP93A	19OCT94	27
PROCUREMENT	18OCT93A	5FEB95	27
CONTRACTS	13DEC93A	11NOV94	27
CONSTRUCTION	15JAN94A	3JAN95	27
COMMISSIONING	10DEC94	5JAN95	27
AREA 34 - HEAP LEACHING			
ENGINEERING	18OCT93A	26SEP94	34
PROCUREMENT	18OCT93A	23JAN95	34
CONTRACTS	1AUG94A	7NOV94	34
CONSTRUCTION	3MAY94A	5MAR95	34
SPRAYING ON STOCKPILE	1MAR95	31MAY95	34
AREA 35 - SOLVENT EXTRACTION PLANT			
ENGINEERING	22OCT93A	26SEP94	35
PROCUREMENT	20JAN94A	12DEC94	35
CONTRACTS	1APR94A	31OCT94	35
CONSTRUCTION	1JUL94A	15FEB95	35
COMMISSIONING	14FEB95	9MAR95	35
AREA 36 - SX TANK FARM			
ENGINEERING	20NOV93A	15SEP94	36
PROCUREMENT	26OCT93A	22JAN95	36
CONTRACTS	30MAR94A	27SEP94	36
CONSTRUCTION	11JUL94A	19MAR95	36
COMMISSIONING	14FEB95	19MAR95	36
AREA 37 - ELECTROWINNING PLANT			
ENGINEERING	26OCT93A	21SEP94	37
PROCUREMENT	17SEP93A	30MAR95	37
CONTRACTS	1JUL94A	19OCT94	37
CONSTRUCTION	30JUN94A	9APR95	37
COMMISSIONING	20MAR95	31MAY95	37

Activity Classification: RESPONSIBILITY
CONSTRUCTION
SHIPPING
ENGINEERING
CONTRACTS
PROCUREMENT
X

Plot Date 29AUG94
Data Date 23AUG94
Project Start 5JUL93
Project Finish 31MAY95

Summary Bar/Early Dates
Critical Designator
Progress Bar
◇ / ▸ Milestone/Flag Activity

zal 3
Sheet 3 of 4

COMPAÑIA MINERA ZALDIVAR
ZALDIVAR PROJECT
BC-99 : EXECUTIVE SUMMARY

**** PROGRESSED SCHEDULE ****

Date	Revision	Checked	Approved

(c) Primavera Systems, Inc

FIGURE 3 Project schedule overview

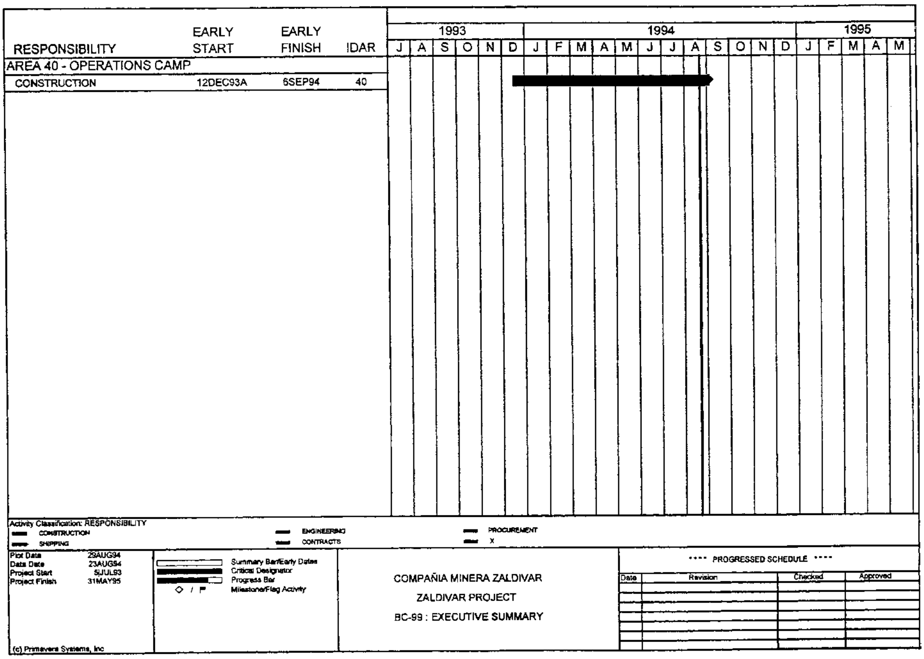


FIGURE 3 **Project schedule overview (continued)**

Additionally, there were contractor camps located at Domeyko, Neurara, Pampa Union, Baquedano and La Negra to support the water line and powerline construction.

The project achieved a significant milestone in accomplishing 9,285,691 site hours worked without a fatal accident on the mine site. During the course of the project, however, the contractor building the mine access road suffered two fatalities.

Earthworks on the Zaldivar sites comprised 4,500,000 m^3 of excavation and 1,980,000 m^3 of backfill. The concrete batch plant on site produced 50,000 m^3 of concrete throughout construction, with aggregates sourced offsite, from La Negra near Antofagasta.

The project was completed at a cost of US $574.1 million which was $13.6 million under the total budget of US $587.7 million including the US $43.4 contingency. An overview report of the 10 most significant over/under cost items is shown in Figure 4. Additional unbudgeted capital costs were experienced with project improvements requiring a rail connection to site and expanded acid pumping systems.

Due to the vast distances over which the borefield pipeline, 220 kV transmission line and project access road were built, the project was constructed almost as four individual projects.

1. The immediate project site including the crushing/conveying/leaching and copper producing facility
2. The 220 kV powerline
3. The Negrillar borefield and 100 kilometers pipeline
4. The 70-kilometer minesite access road

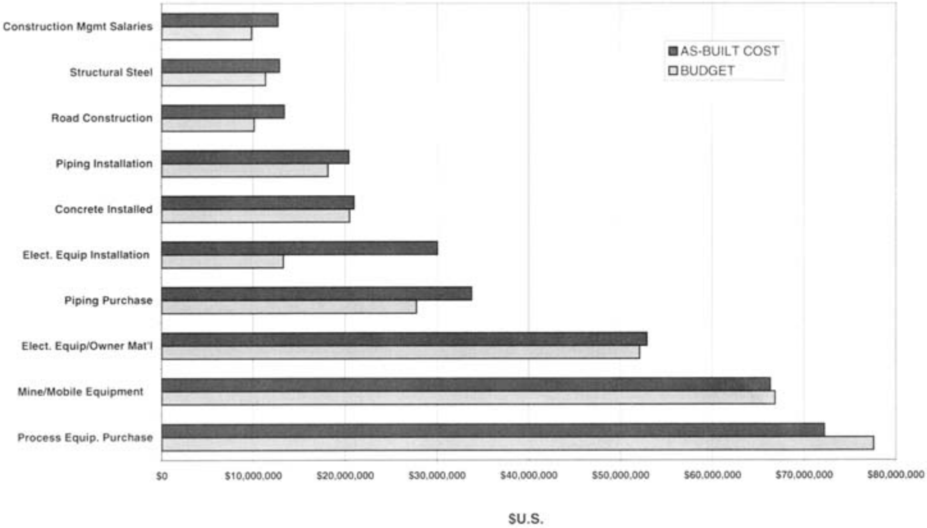


FIGURE 4 Zaldivar Pareto

The construction contract methodology allowed the powerline and water supply system to be built on a lump sum basis, with the mine access road being successfully built as a design construct contract.

The process plant was a combination of lump sum, cost reimbursable and unit price contract determined by the construction schedule and was dependent upon the status of the associated design completion.

One of the key project successes was the close relationship that existed between the operation and construction group. This was evident in the infrastructure support allowing the preproduction mining to meet its monthly targets as quickly as possible. This support included erecting the open pit mining equipment as well as a “fast-track” approach to having the operation administration building and assay laboratory functional at a very early date.

There was significant support to the project provided by our administration groups in Santiago. They provided much of the customs approval and duty exoneration requirements of the project as well as the accounting and treasury functions necessary to manage a cash flow of up to US $45 million per month.

The equipment and supply logistics were very significant with purchasing offices in Vancouver for most process and mining equipment purchases. Additionally purchasing was done on site for leach supplies and from Santiago for sources of other materials such as specialty pipe and tankage. The logistics effort required two people to manage freight at the Port of Antofagasta and subsequent truck freight to site. During October 1998, a peak of 513 truck loads of materials/equipment were delivered at an average rate of 17 loads/day. Additionally the railway from Antofagasta was used to deliver the 100 kilometers of 24-inch and 26-inch diameter pipe required for the pipeline.

One of our key advantages was having two purchasing groups who could tender internationally and get the best pricing/delivery available from international suppliers.

The core group of personnel who have worked on previous Placer Dome Inc. projects adapted well to the Chilean culture and trained many Chileans in our construction methodology.

ZALDIVAR SX/EW PLANT

Concepts

The concepts and layouts for the heap leaching, solvent extraction and electrowinning tank farm at Zaldivar were incorporated into the Zaldivar Feasibility Study completed in 1994 with contributions from PDD's evaluation group, Outokompu Oy of Finland and Fluor Daniel's Vancouver office.

During the feasibility study process, Fluor Daniel personnel worked closely with PDD's design and construction groups to reach alignment with regard to construction methods, selection of materials and contract and procurement philosophy as well as the project schedule.

Fluor Daniel was later selected to supply the detailed design of the heap leach conveying/stacking as well as the SX/EW facilities. The design took eight months and was completed on schedule.

General arrangements of the overall plantsite as well as the Zaldivar SX/EW plant are attached as Figures 5 and 6, respectively.

Procurement

The major equipment selected included Outokompu's patented mixer settlers and cathode stripping machines. The 368 electrowinning cells were manufactured in Santiago and were installed prior to the erection of the building. The tank farm tanks were built locally with stainless steel supplied out of Chile by the installation contractor. The reverse osmosis plant was designed with the key components supplied by Hydropro. Anodes were supplied from Royston Lead Plc. and cathodes from MIM Technology.

Construction Technologies (SX/EW)

The close-working relationship between Fluor Daniel and PDD, for example, enabled the design to take advantage of simple forming techniques for the electrowinning cells which allowed for the maximum rate of advance in concrete slab and support placement.

The cells were installed from one end of the 400-meter-long building to the other end with the sequence of building steel following in the same pattern.

The pipe rack layout in the SX, EW, and tank farm areas allowed efficient pipe installation activities to be carried out minimizing placement time and costs.

Construction Management (SX/EW)

There were approximately 18 major contracts awarded over the 11-month construction period.

Two civil contractors accomplished most of the civil works with two other contractors erecting the SX and EW buildings. There were two piping contractors and two electrical

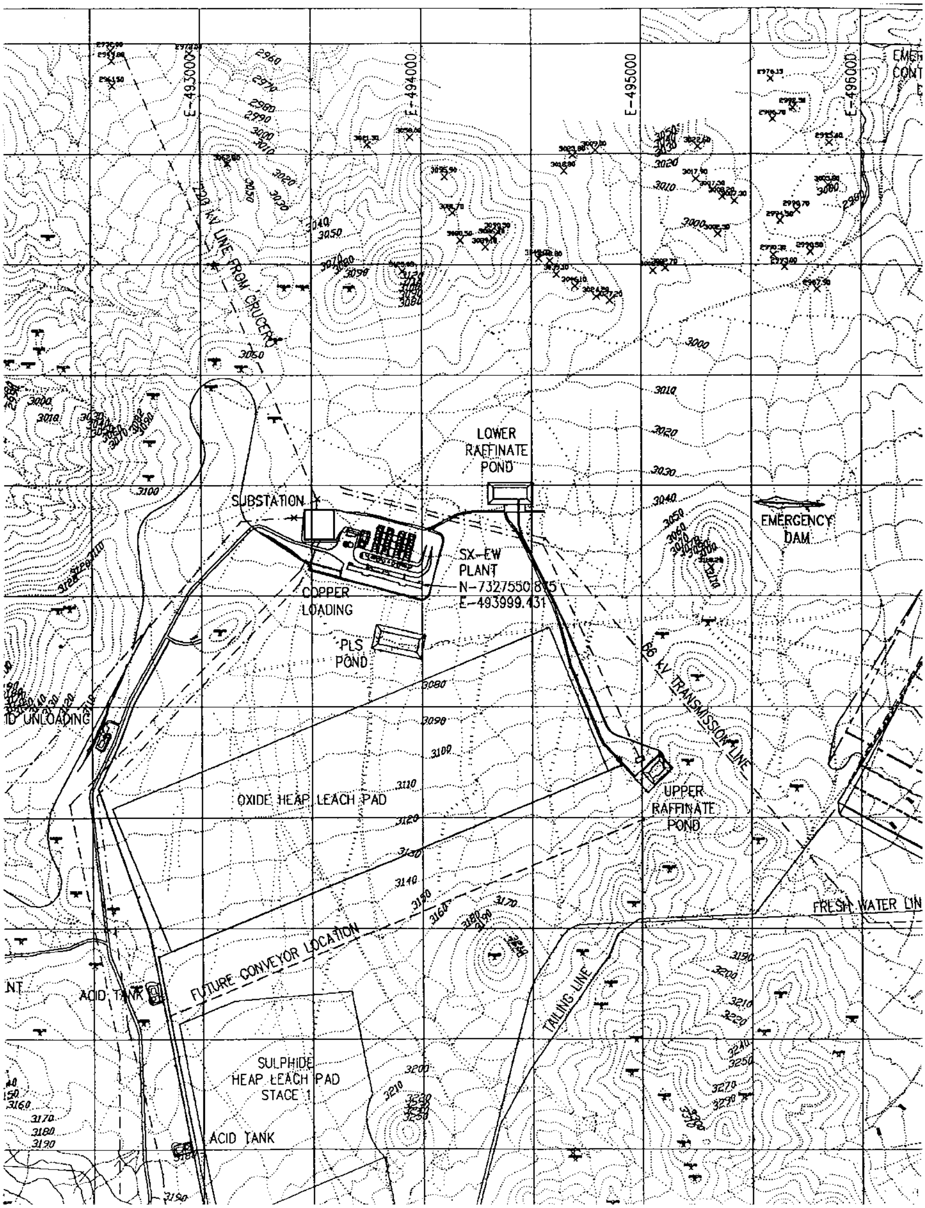


FIGURE 5 Zaldivar SX/EW plantsite

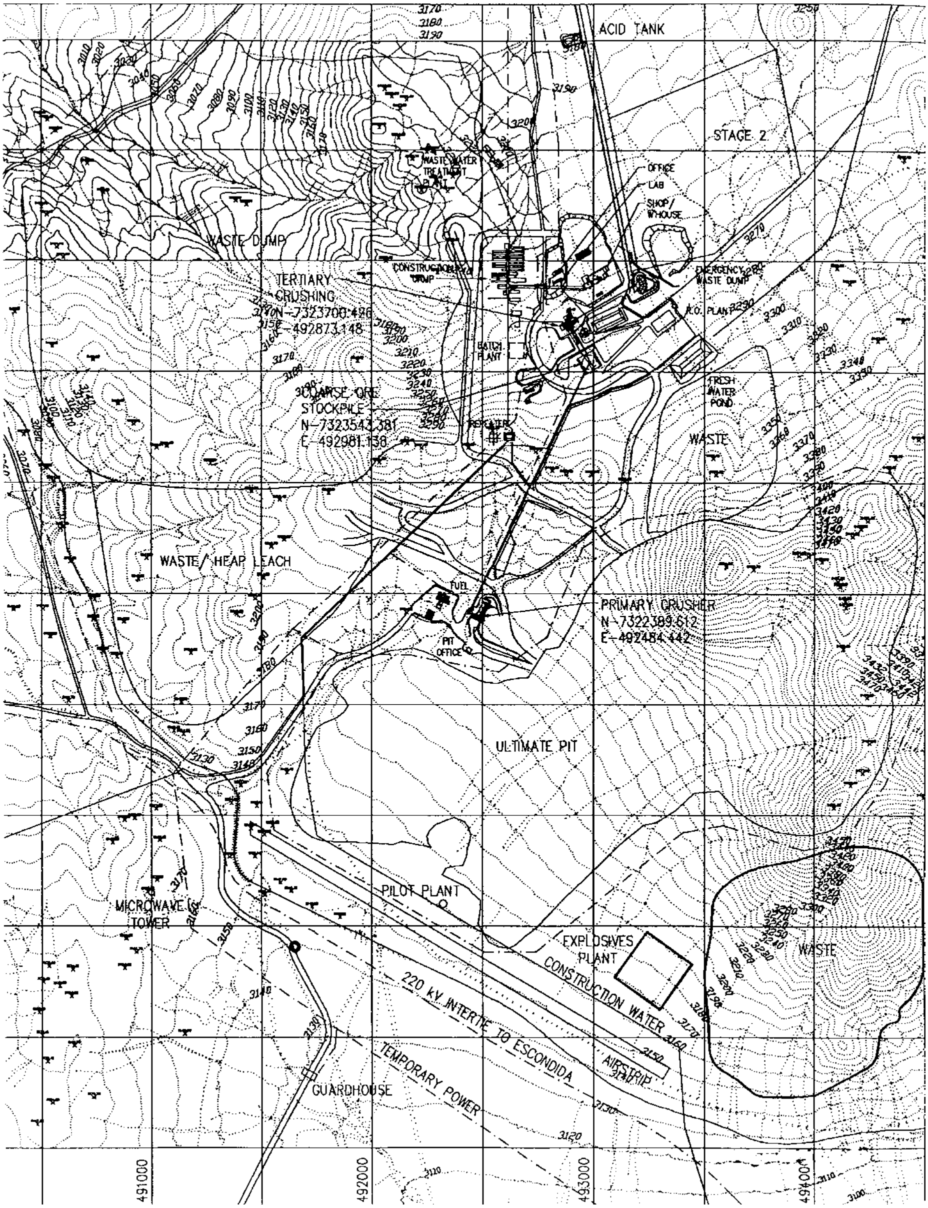


FIGURE 5 Zaldivar SX/EW plantsite (continued)

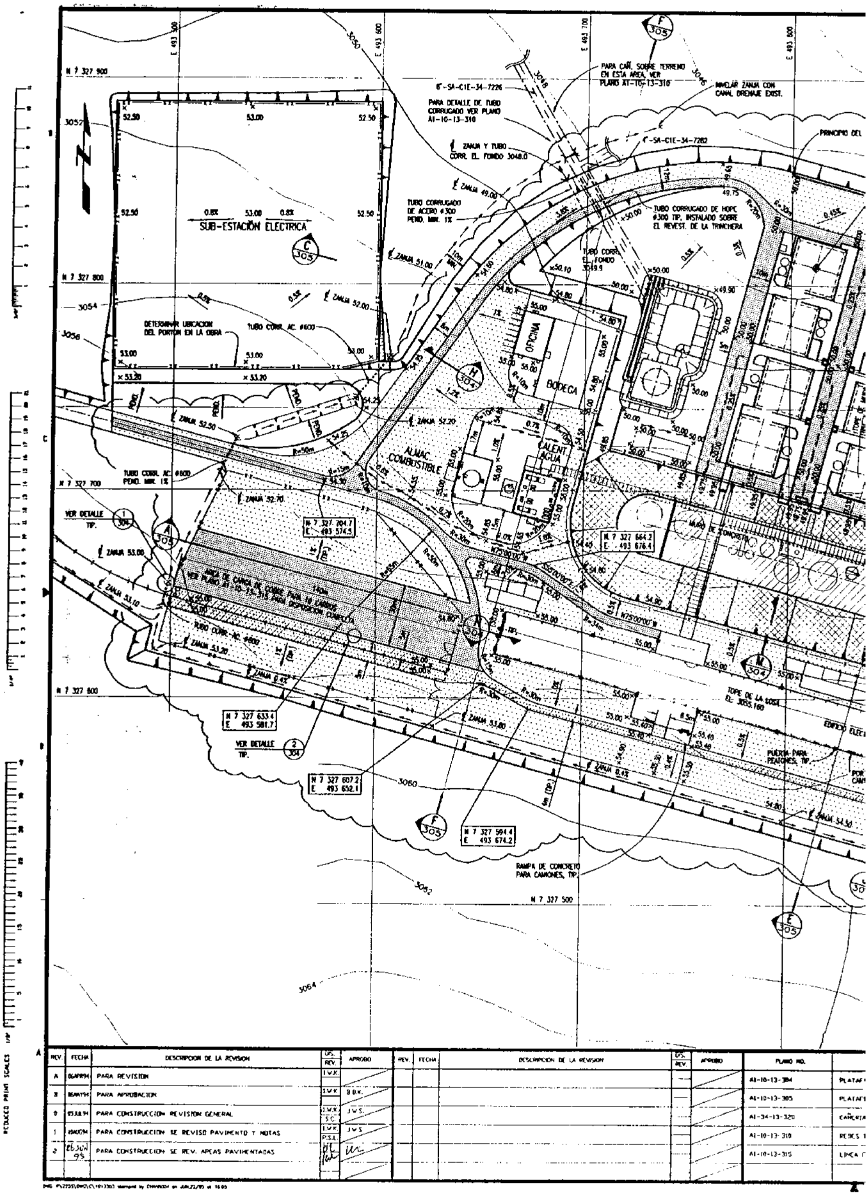


FIGURE 6 Zaldivar SX/EW plantsite

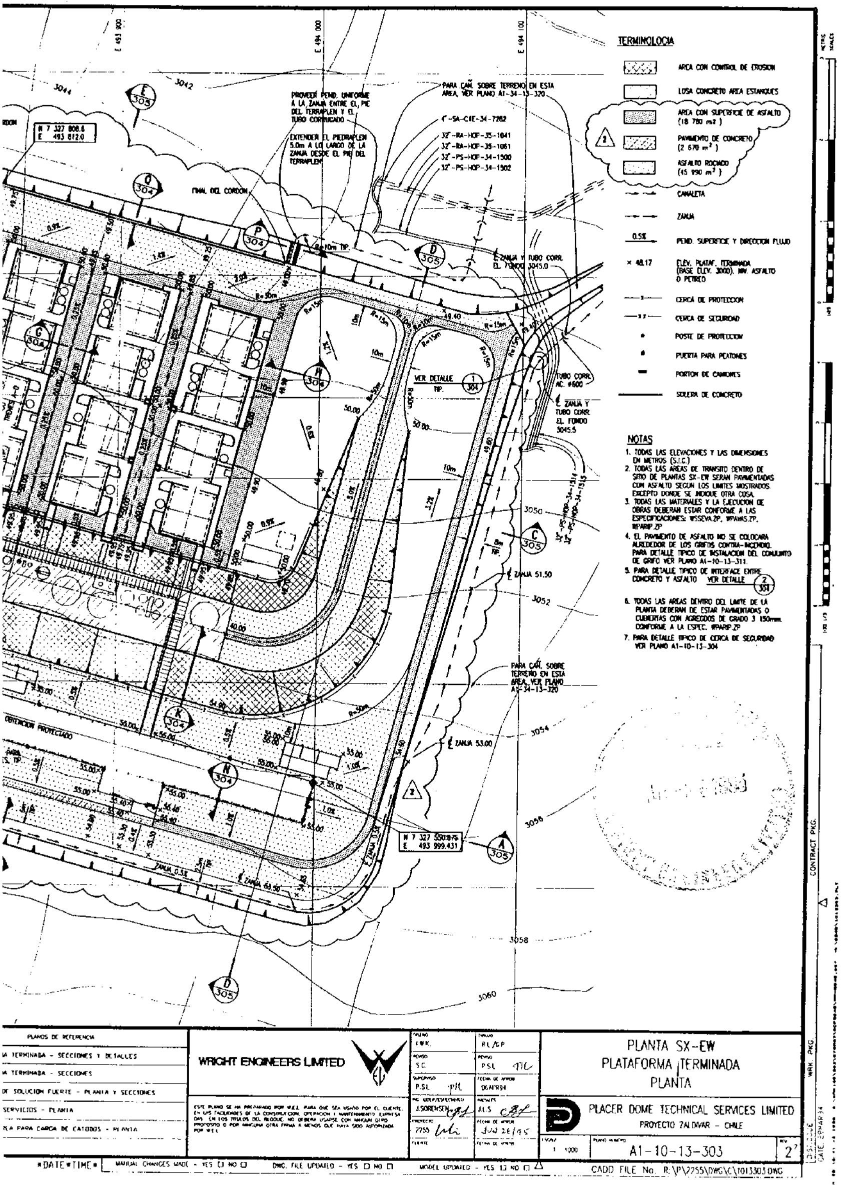


FIGURE 6 Zaldivar SX/EW plantsite (continued)

contractors. The contract philosophy developed was to have at least two contractors of the same discipline on a site so that we could select the most appropriate for additional work based upon their present performance and unit prices.

The selection of the right contractors is critical to a project's success. Considerable time was spent in the pre-award session to ensure that the contractors understood how to execute the work and have the appropriate equipment and resources available to accomplish their work on a timely basis.

Budget

The combined budgets excluding the indirect costs of design and construction management for the heap leach, SX, plant, tank farm and electrical facilities was US $117.32 million versus the actual cost of US $132.49 million. The approximate US $15 million overrun was due to:

- Underestimating the cost of leach pad preparation and the quantities of work involved
- A slight overrun in equipment purchases, notably the mixer settlers, which was offset and justified on the basis of predicted lower reagent consumption during the operational period
- Piping in the tank farm underestimated and funds not budgeted for piping insulation, shotcrete in tankfarm area and increased quantities of asphalt paving around the SX/EW plant

Commissioning

The SX/EW plant was designed as four identical segments and was commissioned in the same sequence as constructed.

Once sufficient ore was placed on the oxide pad with an average 8.0 meter lift height, or approximately 500,000 tonnes, the heap leach spray system was commissioned and leaching of the ore commenced. It took three days for solution to report to the pregnant solution pond and several days later before the copper concentrations in the solvent extraction plant reached 2.6 g/l, which allowed electrowinning operations to start.

The commissioning team included PDD's design staff from Vancouver, some construction personnel as well as CMZ operators and management staff. Outokompu complemented the team by providing personnel for commissioning expertise. In addition, technical support was received from the equipment vendors.

Some problem areas were encountered during the final construction and commissioning stages of the SX/EW areas. These areas included (1) leaks in some of the solvent extraction liners, (2) an undersized boiler system for colder winter temperatures and (3) leaks in two large SX/EW stainless steel tanks.

As deficiencies were noted during commissioning, they were corrected by the construction SWAT team under the direction of the commissioning personnel. In summary, the success of the Zaldivar SX/EW plants was due to nature of the partnerships developed by all participants through the feasibility study, design, construction and plant commissioning.

ACKNOWLEDGMENTS

The contributions of many individuals and firms are acknowledged with special thanks to the following for the successful completion of the Zaldivar project:

- Outokompu Oy of Finland
- Dedicated Zaldivar CMZ personnel
- Placer Dome Inc. employees
- Placer Dome Latin America employees
- Placer Dome project development personnel
- Placer Dome technical services employees
- Design firm and contractor representatives

SECTION 2

Theory and Practice of Copper Leaching

Co-Chairs

J. Keane
KD Engineering, Inc.

R. Dorey
Steffen, Robertson, and Kirsten

. . . .

Practical Aspects Associated with Evaluation of a Copper Heap Leach Project

E. Iasillo[*] **and W.J. Schlitt**[†]

Definition of leach parameters required for development of a copper heap leach flowsheet is driven by mineralogy. This may include oxide, supergene and/or hypogene ores. Baseline testwork should be conducted on representative ore. The effects of solution pH, ORP, ferric iron tenor, dissolved oxygen and crush size upon leach kinetics, final extraction and acid consumption in such tests are discussed. The primary objective of the testwork is to develop criteria for design and construction of a facility. Selection and design of the leach system (surface or in-situ) are based on such parameters as recovery, acid requirements, crush size, lift height, flowrate and PLS grade. Various engineering aspects are also discussed.

OVERVIEW

The fundamentals of leaching include three phases: testing, engineering design and operations. The first phase involves the testing of representative ore samples to obtain basic design data. Key parameters to be measured are the extraction rate curve for copper dissolution and a corresponding rate curve for acid consumption. These are determined as functions of variables that include ore grade and type, crush size distribution, agglomerating conditions (if required), leach solution parameters (e.g., pH, ferric content and ORP), irrigation rate, temperature and column/heap height.

Engineering requires translation of the test results from the metallurgical study into the design of the leaching and recovery facilities. In addition to planning the leach system itself, the engineer must estimate the flowrate and copper tenor of the pregnant leach solution (PLS). These are required to establish the design base and size for both the SX and the EW plants. The former is primarily dependent on the flowrate, while the latter is dependent only on the amount of copper recovered from the PLS. In considering the leach facilities design, the engineer must first decide on the type of system—usually a

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[*] METCON Research Inc., Tucson, Ariz.

[†] Kvaerner Metals, San Ramon, Calif.

multilift permanent heap or a single-lift, on-off pad. The next decisions involve the equipment for agglomeration and ore placement followed by the selection of the solution distribution and collection system. Other decisions include the active and leach cycle times so that total pad requirements can be established, the ramp-up period to achieve full (design) production and estimates of capital and operating costs.

The third phase is not the focus of this paper but involves the practical operation of the leach and recovery facilities. This combines operation, control and maintenance of the commercial system. The same unit operations must be addressed as were considered in the design—crushing, agglomeration, ore stacking and solution management. However, the operator must also interface with mine planning and ore control to ensure that adequate tons of material are delivered for leaching. In the end, the operator is responsible for seeing that the budgeted amount of copper moves through the system and that the final product meets required quality levels. Normally, this would be LME Grade A copper, or equivalent.

A method leading to project failure is to develop the basic criteria in a vacuum, with little or no interaction between the testing personnel, the operating personnel and the design engineer. If there is no communication, the metallurgist will conduct a test program that provides accurate results but overlooks key parameters needed to design and engineer the system. The engineer will then prepare a design that looks great, at least on paper. However, the operator will find that the plant is difficult to control and maintain, so that production shortfalls plague the operation right from day one.

The answer is a team approach. Here the engineer and operator will have input to the metallurgical test program. Then the metallurgist and operator will help the engineer with interpretation of the test results and establishment of the plant design features, especially those associated with process control and maintenance. In a sense, the operator will be responsible to meet production. However, the metallurgist and engineer should at least participate in the start-up and shake-down of the plant to help solve the inevitable problems and to be sure that it is operating as expected. Since slow leaching ores may require extended leach cycles (a year or more) to ramp up to full output, the metallurgist and engineer may not stay involved to this point, at least on a full-time basis.

As a guide to the team approach, the first two phases are addressed by the appropriate author in the following sections.

METALLURGICAL TEST PROGRAM FOR COPPER LEACHING PROJECTS

by Eugenio Iasillo

Development of the flowsheet and design criteria for a copper leaching operation is driven by the mineralogy of the resource under study. In order to develop defensible and meaningful metallurgical data, mineralogical data should be developed on a selected suite of samples deemed representative of the different ore types identified in the resource. The different types of materials should then be classified as accurately as possible into oxide, supergene and/or hypogene ores. Different analytical techniques may be used to make a preliminary assessment of the mineralogy present in each type of material, but these analytical techniques are not intended to replace mineralogical examination. Analytical results should be viewed as a complement to a mineralogical study as well as a quick and efficient way of identifying ore zones within the resource. During the mineralogical examination and initial phase of the testwork, special attention should be placed in determining the host rock components and their ability to consume acid. A preliminary evaluation of leaching characteristics is particularly important when treating sulfide copper ores.

The leaching characteristics of copper sulfide minerals vary, and these relative differences allow their identification and quantification using analytical techniques as discussed herein.

Solubility of Copper Minerals

Most copper minerals are soluble to various degrees in sulfuric acid solutions. In order to make a preliminary assessment in quantifying the most soluble portion of the copper contained in the material under evaluation, a pulverized sample is contacted with a dilute sulfuric acid solution. It should be noted that some of these minerals are soluble in cyanide solutions as well. The dissolution behavior exhibited by the copper minerals may be used as a basis for development of analytical techniques that provide a semiquantative determination of oxide and sulfide components in the ore. These analytical techniques are usually applied in a sequential fashion to determine the amounts of leachable oxide, supergene and hypogene mineralization. Table 1 (Parkison and Bhappu 1995) outlines the solubility at room temperature of various copper minerals including native copper.

As seen in the table, not all oxide copper minerals are readily soluble in sulfuric acid. It is also important to note that secondary and primary sulfide minerals are partially soluble in acid.

The solubility of copper minerals in cyanide solutions was studied (Leaver and Woolf 1931) and published in American Cyanamid Company Mineral Dressing Notes, Number 23, June 1958.

TABLE 1 Dissolution of various copper minerals in sulfuric acid and sodium cyanide solutions

Mineral	Composition	Approximate Dissolution in Sulfuric Acid Solution	Approximate Dissolution Sodium Cyanide Solution
Oxides			
Atacamite	$Cu_2Cl(OH)_3$	100	100
Azurite	$2CuCO_3Cu(OH)_2$	100	100
Cuprite	Cu_2O	70	100
Chrysocolla	$CuSiO_32(H_2O)$	100	45
Malachite	$CuCO_3Cu(OH)_2$	100	100
Native Copper	Cu	5	100
Tenorite	CuO	100	100
Secondary Sulfides			
Chalcocite	Cu_2S	3	100
Covellite	CuS	5	100
Primary Sulfides			
Bornite	Cu_5FeS_4	2	100
Chalcopyrite	$CuFeS_2$	2	7

NOTE: Samples are generally finely ground (–150 mesh) and reaction time is generally one hour or less.

TABLE 2 Solubility of copper minerals in NaCN solutions

		Percent Total Copper Dissolved	
Mineral	Composition	At 23°C	At 45°C
Azurite	$2CuCO_3 \cdot Cu(OH)_2$	94.5	100.0
Malachite	$CuCO_3 \cdot Cu(OH)_2$	90.2	100.0
Cuprite	Cu_2O	85.5	100.0
Chrysocolla	$CuSiO_3$	11.8	15.7
Chalcocite	Cu_2S	90.2	100.0
Chalcopyrite	$CuFeS_2$	5.6	8.2
Bornite	$FeS \cdot 2Cu_2S \cdot CuS$	70.0	100.0
Enargite	$3\ CuS \cdot As_2S_5$	65.8	75.1
Tetrahedrite	$4\ Cu_2S \cdot Sb_2S_3$	21.9	43.7
Metallic Copper	Cu	90.0	100.0

Table 2 outlines the solubility of various copper minerals in 0.099 percent NaCN solutions at two different temperatures and 24-hour retention time. The ratio of solution to solids was 10 to 1.

The best approach in quantifying oxide, secondary and primary copper components of an ore is to utilize a sequential type of leaching technique. This technique should be designed to render the copper mineralization soluble in solutions that selectively dissolve the copper species in a sequential fashion.

Metallurgical Study

Metallurgical data related to the leaching characteristics of materials under study may be developed using a stepwise approach.

Initially, samples under study should be subjected to a sequential type of copper analysis as well as to total copper analysis. This type of information has proven useful in the design of a metallurgical study and selection of samples representing various types of mineralization. These analyses will also help define mineralized zones within the ore body and the potential for copper recovery by extractive metallurgy.

Bottle Roll Leach

Once it has been determined that the material under study contains sufficient leachable copper to merit further investigation, a dynamic type of leach test is usually performed. These tests are sometimes conducted during the operating phase of the project to determine if the mined material is suitable for processing. There are many ways to conduct dynamic leach tests commonly known in the industry as “bottle roll tests.” The main variables to be considered are pulp density, acid concentration, ferric concentration, retention time and temperature. Agitation should not be considered a variable at this stage. It has been demonstrated that a bottle roll apparatus will provide results that are reproducible and most closely simulate the agitation obtained in an agitated tank at an industrial scale. On a bottle roll apparatus all the material remains in continuous motion and a pulp air interface is created for aeration of the pulp.

The feed material used in a typical bottle roll leach test is usually crushed to 10 mesh or pulverized to minus 150 mesh and contacted with a leach solution containing sulfuric acid and ferric iron. An excess of acid may be added initially to ensure a pH below 2.0, or periodic acid additions may be made throughout the leach cycle.

A copper extraction curve may be developed by conducting accurate and careful sampling. The overall copper extraction and the maximum sulfuric acid consumption to be expected from the material are determined. These data are necessary for definition of leaching parameters for column testwork and optimization of the number of column tests required for development of design criteria for the leach operation.

Column Leach Testing

The static column leach test program may be divided into three phases. The phase descriptions and the main variables studied in each phase are included below.

1. Baseline data (minicolumns)
    - Acid cure and no-cure
    - Acid cure dosage
    - Acid cure time
2. Open cycle (full-size columns 8 inch Ø × 6 ft)
    - Acid cure dosage (method)
    - Crush size (extraction by screen fraction)
    - Irrigation rate
    - Lixiviant acid concentration
3. Locked cycle
    - Lift height
    - Cure versus no-cure
    - Temperature

Sample Procurement

It is recommended that the entity conducting the metallurgical study work collaborate with the exploration and operations personnel to generate a suitable amount of material for testing at an appropriate particle size distribution. Once the samples are produced, it is imperative that a sample preparation protocol be agreed upon in order to produce composite samples for the evaluation that are representative of various ore types identified in the ore body. Composite samples deemed representative of the initial phases of production for the project would also be suitable for evaluation. The sample preparation procedure utilized for production of test charges should ensure that results from column tests are reproducible.

In general, it is not advisable to classify the material to be studied by copper grade (assay) only. The effect of optimized leaching parameters should be evaluated on all the ore types identified in the ore body. If possible, water from the site should be used in the metallurgical evaluation. The water available at a mine site is, in many cases, significantly different to tap water.

Acid Cure Technology

The sulfuric acid required for leaching should be added in a cure pretreatment step. The acid addition methodology, acid cure dosage and rest parameters must be evaluated by column leach experimentation. Preliminary acid consumption data developed in the bottle roll leach tests are useful in defining acid cure parameters. The mineralogy usually determines the length of the rest or cure period. The rest period tends to be relatively short for oxide mineralization but could be as long as 2 weeks for copper sulfide mineralization. For testing purposes, evaluation of rest periods in 5-day increments provides a trend that will indicate the optimum rest period following acid cure pretreatment.

Crush Size

The parameter that often has the largest impact on overall copper extraction is crush size. The optimum size for leaching is defined by conducting column tests on carefully prepared column test charges representing size distributions typically obtained from industrial scale crushing equipment.

Flowrate

Copper extraction is nearly always a function of the flowrate utilized for leaching. The optimum flowrate is the one that provides the highest copper extraction at the lowest possible flowrate. At least two flowrates should be studied.

Lift Height

Lift height is of utmost importance in the design since it will affect the metallurgy, the stability of the material under leach, the pad area requirements and the solutions inventory. At least two different lift heights should be evaluated under optimized leaching conditions and, preferably, under a locked cycle type of leaching regime.

Locked Cycle Leach

A column study conducted in locked cycle dictates that all the pregnant solution produced will be processed using solvent extraction and the raffinate produced will be recycled as leach solution. Acid, if required, may be added prior to recycling a batch of raffinate.

Under a locked cycle leaching regime, the chemistry and reactions taking place in the heap and solvent extraction are continuously accounted for. The acid consumed in the test is equivalent to the amount of acid added to the system in the acid cure and into the raffinate acid make-up. This acid consumption is commonly known as “gangue acid consumption.” As implied by the name, it is the acid consumed by the gangue minerals present in the ore. An acid inventory is sometimes accumulated in the unused portions of raffinate and wash volumes produced over the last days of the leach cycle. In this case, a credit may be applied to the overall gangue acid consumption at the completion of the test.

Other Aspects

Petrographic analysis should be conducted on a composite sample from the feed and leach residue from at least one test column conducted on each ore type. It may also be necessary to conduct petrographic analyses on the individual screen fractions to determine the reasons associated with the refractory nature of copper mineralization present in the residue.

Sequential copper assays on feed and leached residues produced from selected column leach testing should be conducted as well. The main objective in conducting these analyses is to develop a copper analysis methodology that will allow operations personnel to make a preliminary assessment of the ore content of leachable oxide, secondary and primary sulfide copper minerals.

A combination of petrographic work, metallurgical testing and analytical data should provide useful information for the mine plan and definition of different geologic and mineralogic zones and related leachable copper components.

Effect of Leaching Parameters

Accepted reaction mechanisms that take place during dissolution of various copper minerals are as follows:

Oxide. The leaching reactions for various copper oxide minerals and native copper as outlined in the 1996 SME Short Course Copper Heap Leach notes are as follows:.

Malachite	$CuCO_3 \cdot Cu(OH)_2 + 4H^+ \rightarrow 2Cu^{2+} + CO_2 + 3H_2O$
Brochantite	$CuSO_4 \cdot 3Cu(OH)_2 + 6H^+ \rightarrow 4Cu^{2+} + SO_4^{2-} + 6H_2O$
Antlerite	$CuSO_4 \cdot 2Cu(OH)_2 + 4H^+ \rightarrow 3Cu^{2+} + SO_4^{2-} + 4H_2O$
Atacamite	$3CuO \cdot CuCl_2 \cdot 3H_2O + 6H^+ \rightarrow 4Cu^{2+} + 2Cl^- + 6H_2O$
Azurite	$2CuCO_3 \cdot Cu(OH)_2 + 6H^+ \rightarrow 3Cu^{2+} \rightarrow 2CO_2 + 4H_2O$
Chrysocolla	$CuSiO_3 \cdot 2H_2O + 2H^+ \rightarrow Cu^{2+} + SiO_2 + 3H_2O$
Cuprite	$Cu_2O + 2H^+ \rightarrow Cu^{2+} + Cu^\circ + H_2O$
Tenorite	$CuO \cdot 2H^+ \rightarrow Cu^{2+} + H_2O$
Native Copper	$Cu^\circ \cdot 2H^+ + \tfrac{1}{2}\,O_2 \rightarrow Cu^{2+} + H_2O$

Most oxide copper minerals dissolve readily in sulfuric acid solutions. Dissolution of elemental copper, which is an intermediate product in the dissolution of cuprite, requires oxygen as well.

Supergene. The leaching reactions for chalcocite and covellite and the oxidation reactions for ferrous sulfate and elemental sulfur are presented below.

(1) $Cu_2S + H_2SO_4 + \tfrac{1}{2}O_2 \rightarrow CuSO_4 + CuS + H_2O$
(2) $Cu_2S + Fe_2(SO_4)_3 \rightarrow CuSO_4 + CuS + 2FeSO_4$
(3) $CuS + Fe_2(SO_4)_3 \rightarrow CuSO_4 + 2FeSO_4 + S^\circ$
(4) $CuS + 2O_2 \rightarrow CuSO_4$
(5) $2FeSO_4 + H_2SO_4 + \tfrac{1}{2}O_2 \rightarrow Fe_2(SO_4)_3 + H_2O$
(6) $S^\circ + H_2O + 1.5O_2 \rightarrow H_2SO_4$

Dissolution of chalcocite requires 2 stages. The first stage (a combination of reactions 1 and 2) occurs rapidly while the second stage (reactions 3 and 4) occurs at a slower rate. It should be noted that dissolution of covellite is enhanced by an increase in ferric iron and oxygen concentration. Bacteria have an indirect impact in these reactions in promoting iron and sulfur oxidation in a heap leach process.

Hypogene. A number of important copper sulfide minerals and dissolution reactions are outlined below. Brown and Sullivan (1934) and Sullivan (1933) studied the dissolution of chalcopyrite, enargite and bornite in ferric sulfate solutions.

Chalcopyrite	$CuFeS_2 + 4Fe^{3+} \rightarrow Cu^{2+} + 5Fe^{2+} + 2S^\circ$
	$CuFeS_2 + 4Fe^{3+} + 2H_2O + 3O_2 \rightarrow Cu^{2+} + 5Fe^{2+} + 2H_2SO_4$

Bornite	$Cu_5FeS_4 + 12Fe^{3+} \rightarrow 5Cu^{2+} + 13Fe^{2+} + 4S^\circ$
Enargite	$4Cu_3AsS_4 + 35O_2 + 10H_2O \rightarrow 12CuSO_4 + 4H_3AsO_4 + 4H_2SO_4$

The dissolution of hypogene mineralization is influenced by the same factors mentioned for supergene mineralization. It is evident from the reactions above that the presence of ferric and oxygen in the lixiviant will enhance the rate of copper dissolution. If the rate controlling factor is diffusion, the presence of the ferric and/or oxygen will not have an impact on leach kinetics.

A side effect from these reactions is acid generation and sometimes dissolution of elements such as arsenic and antimony which can act as bactericides. However, iron and sulfur oxidizing bacteria are adaptable to adverse leaching conditions.

Copper bearing materials exhibit differences in leaching characteristics. The leaching characteristics are influenced mainly by the mineralogical make up of the ore. Other factors that influence the leaching kinetics and acid consumption in a heap or dump leach process are as follows:

- Temperature (ambient and heap)
- Heap humidity and aeration
- PLS pH and potential (ORP)
- Bacterial activity

The mineralogy of the ore will affect the leaching time required to achieve an acceptable level of copper extraction. As can be seen in Figure 1, the leach time of oxide, supergene and hypogene mineralization may be measured in weeks, months and years, respectively. These leach times will directly impact the leach cycle in actual operation.

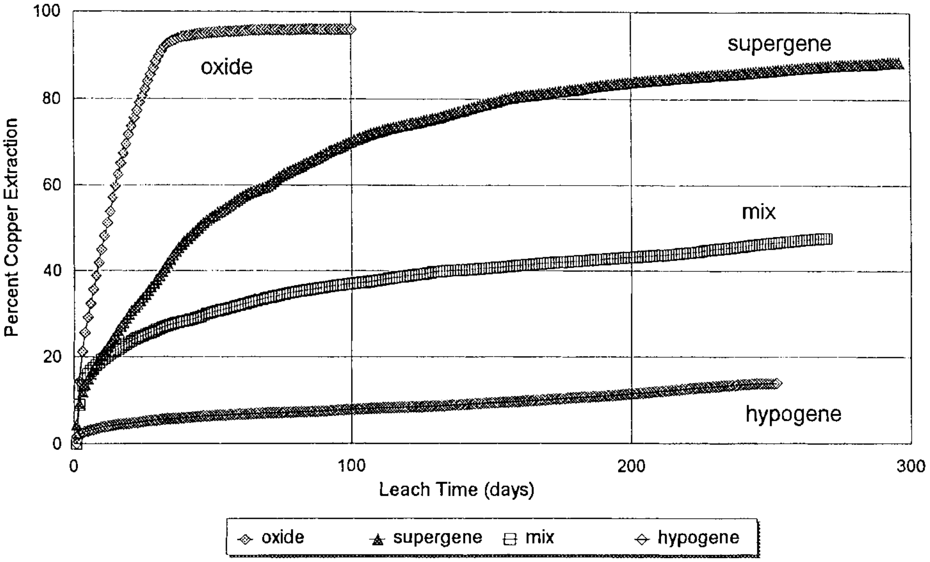


FIGURE 1 Column leach tests conducted in open cycle on material crushed to –3⁄8 inch. Lixiviant used consisted of 10 gpl sulfuric acid and 3 gpl ferric.

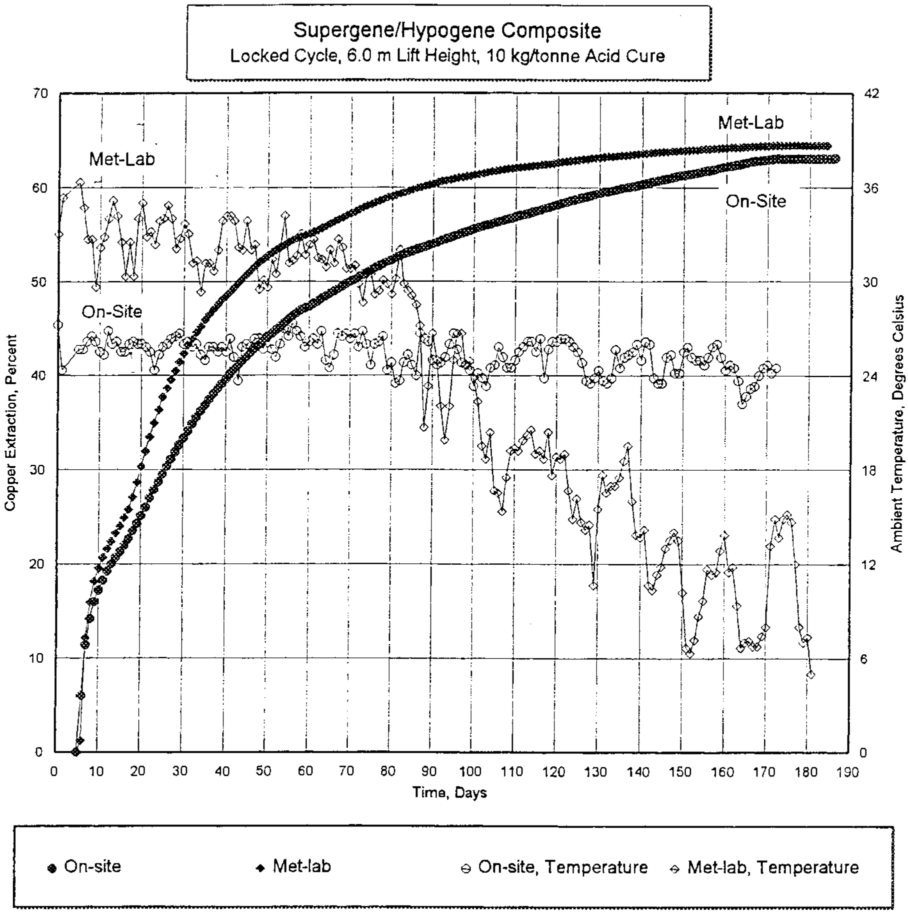


FIGURE 2 Column leach tests conducted at two different locations (on-site and Met-Lab). Different climatic condition and test site elevations.

In general, acid consumption in leaching copper sulfide mineralization will tend to be lower than for oxide copper ores. This is mainly due to acid generating characteristics associated with leaching copper sulfide mineralization.

Temperature. The impact of temperature can be significant in leaching copper sulfide minerals. Most ongoing reactions are exothermic and the material in the heap is insulated by the material under leach. If climatic data is available from the site, the temperature conditions may be evaluated in the laboratory. Figure 2 exhibits the copper extraction kinetics of two column tests conducted on-site and at a laboratory where the outside temperature was higher for the first half of the leach cycle while the temperature on-site was relatively constant.

Differences in copper extraction in excess of 5 percent were observed between these columns. The temperature in one of the columns declined below the temperature of the column conducted on-site but the final results were equivalent after a 180-day leach cycle.

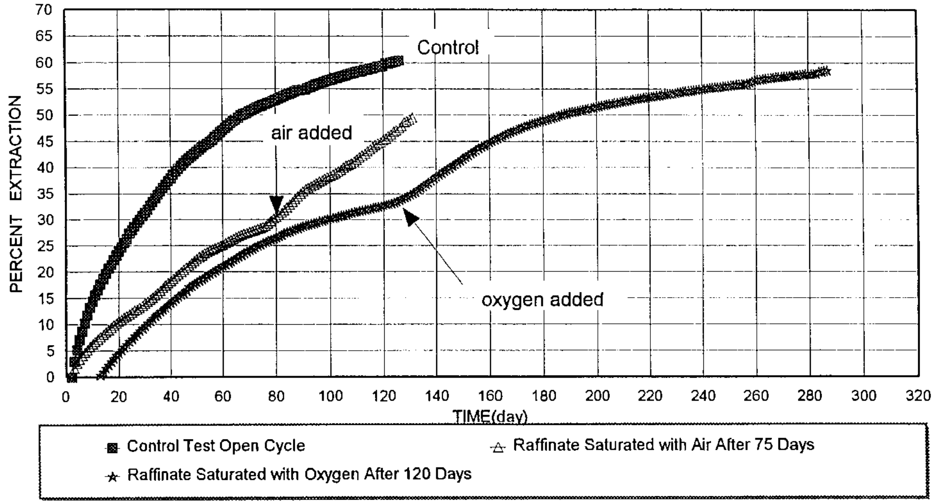


FIGURE 3 Column tests were conducted with a low dissolved oxygen content in the lixiviant induced by sparing nitrogen in the leach solution. Air and oxygen were sparged into the leach solution to evaluate the effect of enhanced aeration.

Heap Aeration. The oxygen required in the dissolution reactions is provided by proper aeration of the heap. The driving force for natural aeration is the difference between the ambient and internal heap temperatures. The operating criteria of the heap must ensure that the continuous phase in the heap is air. Flooding and ponding on top and within the heap will impede oxygen flow thus causing dissolution reactions to slow or cease completely. A lack of oxygen will also affect bacterial activity. Figure 3 shows a series of experiments conducted to evaluate the impact of poor and enhanced aeration in column testing.

The data indicate that natural aeration usually provides sufficient oxygen to effect copper dissolution. Abnormal aeration conditions may be corrected by forced aeration of the heap.

PLS pH and Potential (ORP). One of the most important parameters to be monitored and controlled in copper leaching is the pH of the PLS. In order to ensure that copper and iron will remain in solution, the pH of the PLS must be maintained between 1.8 and 2.2. Sulfuric acid addition is usually made during a cure pretreatment step to ensure that sufficient acid has been added to acidify the material under leach and the maintained pH between acceptable limits. Figure 4 shows data developed in a series of mini-column tests to determine the amount of acid to be added to a cure pretreatment.

The data generated indicate that minimum acid cure dosage for the ore being evaluated required a 10 kg per tonne of the PLS and 5- to 15-day cure to ensure that the PLS pH will be below 2.2 from leach day one.

In leaching copper sulfides, the solution potential of the PLS should be monitored on a daily basis. The solution potential (ORP) in Figure 5 showed an increase after approximately 3 weeks of leaching. A sudden increase in ferric iron concentration caused an increase in solution potential. This phenomenon occurs naturally and is a positive

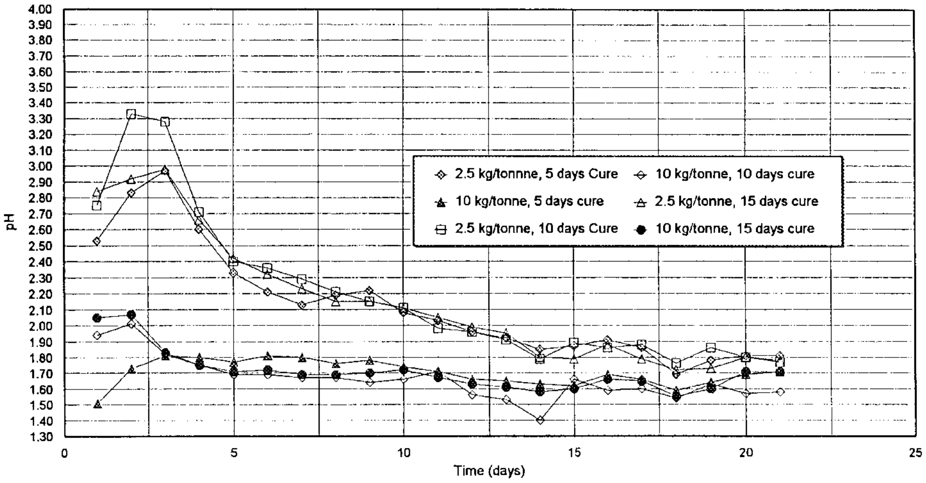
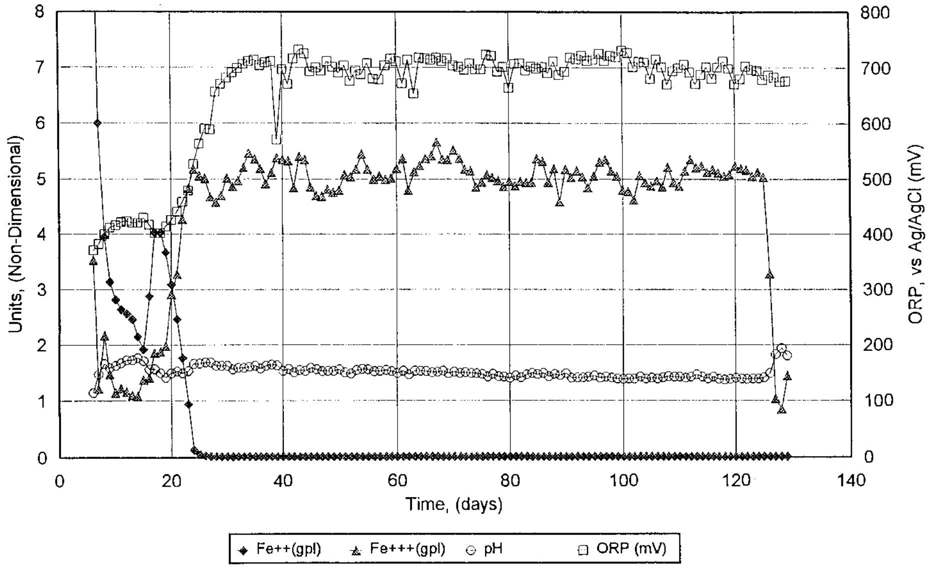


FIGURE 4 Minicolumn studies conducted to develop baseline data for optimization of cure parameters



ORP was measured using a platinum electrode with a Ag/AgCl reference electrode.

FIGURE 5 Column test conducted in locked cycle on a mixture of supergene and hypogene ore

indication of bacterial activity. In order to corroborate and evaluate the extent of this activity, bacterial population should be evaluated. Most Probable Number (MPN) enumeration of bacteria should be conducted on PLS samples at the beginning of the test and after the indications of bacterial activity are detected. When conducting MPN analyses, positive and negative control counts are necessary.

References

Brown, S.L. and Sullivan, J.D. 1934. “Dissolution of Various Copper Minerals,” Washington, D.C., U.S. Bureau of Mines, RI–3228.

Leaver and Woolf. 1931. “Copper and Zinc in Cyanidation, Sulphide-Acid Precipitation,” Tech. Paper 494, U.S. Bureau of Mines.

Parkison, G.A. and R.B. Bhappu. 1995. “The Sequential Copper Analysis Method—Geological, Mineralogical and Metallurgical Implications,” SME Annual Meeting, Denver, Colo.

Sullivan, J.D. 1933. “Chemical and Physical Features of Copper Leaching,” *Trans. Am. Inst. Min. Metall.*, 106:515–546.

DESIGNING AND ENGINEERING THE LEACH SYSTEM

by W. Joseph Schlitt

Introduction

As the metallurgists are completing their testwork program, the engineers can begin interpreting the results and developing the design criteria for the leach system and recovery plant. The risk here is in the ore reserve estimate and extrapolation of leach performance to the commercial operation. With 30 years of development, the design of the solvent extraction (SX) and electrowinning (EW) facilities has become straightforward. Design throughput and the cathode output will most likely be achieved, but only if the expected ore grade and leach rate are met.

The important design considerations are discussed in the following section. These include the impact of the ore reserve, selection of the type of leach system, and establishment of the leaching parameters and cycle times. These provide the bases for sizing the SX and EW operations. Another important activity is to prepare forecasts of copper production and acid consumption, both during the start-up mode and under steady-state conditions.

Design Considerations

Ore Reserve and Mine Plan. The ore reserve and mine plan are as important to the engineer as the metallurgical testwork parameters. The key issue is not the definition of the geologic resource, but the mineable tonnage and grade. If possible, the mine plan should pull high-grade material forward so it is mined early in the operation. This will improve the net present value of the project by minimizing mining costs while still achieving design cathode output. Thus, the capital cost of the project will be amortized as quickly as possible.

Another important consideration is whether or not the mining rate and/or recoverable grade vary with time. If changes are expected, then steps may be taken to minimize production shortfalls in years when the tonnage or grade decline. For example, it may be possible to maintain production by boosting the mining rate to compensate for lower grade ore. However, if the mining rate is increased significantly, provision for some type of intermediate leach solution (ILS) recycle may become necessary. Otherwise, leaching more ore could boost the leach solution flow rate above the design capacity of the SX plant.

If a reasonably constant amount of copper cannot be delivered to the recovery plant, then the engineer is left with two options. One is to build-in excess plant capacity that is only used for short periods—an approach which drives up the capital cost of the facilities.

The second option is to allow some copper to build up as circulating inventory during periods when the recoverable amount of copper is high. Then this inventory is pulled back down when the ore grade or mining rate drops. This option reduces the plant capital cost but also reduces the peak output or sales revenue that can be generated.

Sizing the Recovery Plant. The next step is usually to set the sizes of the SX and EW facilities. This sizing is normally done on the basis of expected steady-state throughput. It is important to realize that the size of the EW plant is primarily driven by the maximum cathode output, with secondary consideration given to current density. Once selected, these parameters set the number of cells and the electrolyte flowrate. In turn, these set the size of pipelines, tankage, heat exchangers, electrolyte filtration equipment and the like. On the other hand, the SX circuit configuration will be set by the expected copper tenor of the pregnant leach solution (PLS). The size of the individual mixer–settlers will then be determined almost exclusively by the flowrates being processed. The mix boxes will be sized to give the desired retention times at the expected flow rates. Likewise, settlers will be sized on the basis of specific throughput (gpm/ft^2 or m^3/h/m^2).

Steady-state cathode output will be determined by multiplying the ore mining rate times the ore grade times a recovery factor. The recovery factor is derived from the ultimate level of extraction achieved during metallurgical testing. Quite often this recovery factor involves some discounting of the leach results obtained under the nearly perfect conditions occurring in column tests. The discount may take the form of either a reduced level of recovery or a longer leach cycle to achieve the same level of recovery.

The size of the discount is something of a judgment call. It depends on how closely column parameters match expected leach conditions. The availability of corroborating results from one or more test heaps will also lend credibility to the estimate of any discount. If the crush size distribution, column height, cure conditions and irrigation rate used for the columns all match those expected in the operation, then the deduction or leach cycle extension will be small, say five percent or less. However, if significant differences are expected between the test work and operations, then the deduction may have to be increased to 10 percent or more. Examples of expected problems might include a coarser crush size or some compacted layers in the commercial heaps. An empirical approach to arriving at this deduction has been described by Schlitt (1997).

Another consideration that is often overlooked is the build-up of copper in solution inventory within the heap system. For a permanent multilift leach arrangement, not all of the copper dissolved from the ore is immediately recovered as cathode. As much as 5 percent of this copper may be held up in the solution retained within the expanding heaps. Some additional copper will also be carried as circulating inventory. The latter is associated with the fact that only about 90 to 92 percent of the copper in the PLS is extracted in the SX operation. The remainder is recycled back to the leach system in the raffinate. The copper in solution inventory is not lost permanently but is retained until the heap is given a final rinse, usually at the end of the operation.

The PLS flowrate to SX is determined by the flow of raffinate and make-up water sent to the heaps for leaching. Adjustments are then made for appropriate losses. The biggest of these is evaporation. The type of spray system affects evaporation. Sprays (Wobbler- or impact-type) may have 5 to 6 percent evaporation, versus only about 2 percent for drip liners or emitters. If evaporation varies seasonally, say from 5 percent in the winter to 15 percent in a hot, dry summer, the smaller loss should be used for sizing plant throughput. Ore soak can represent another major loss of water. This is the amount of moisture that must be added to initiate outflow from a leach heap or dump and is typically about

10 percent by weight. The loss to ore soak will depend on the moisture content of the ore being emplaced for leaching. If the ore is dry, the loss will be high. However, ore moistened during agglomeration may have a very low water demand to achieve drainage.

The estimation of the PLS tenor is quite straightforward once cathode output and the PLS flowrate are known. The net pickup of copper is the cathode production per unit time divided by the PLS flow per the same unit time. This figure is then added to the copper content of the raffinate to give the total PLS tenor.

Selection of the Leach System. One of the key design decisions is the type of leach system to be employed. Presently, most projects opt for a permanent multilift heap. This involves periodic expansion of the lined pad area and/or progressive stacking of one lift atop another. These systems are not without their problems, however. One is the size and cost of the leach pad which must ultimately support the entire tonnage that will be mined. The pad must be properly constructed and covered with a membrane liner that does not fail (leak) even when subjected to differential settlement or the loading imposed by ore stacked several hundred feet (100 meters or more) high. Another problem is the cost and complexity of a conveying system that can stack to such heights. A third problem, one mentioned above, is the accumulation of soluble copper as inventory within the heap.

The permanent multilift approach does have one overriding advantage for ores that have a slow leaching component. As new lifts are added, leach solution will still percolate down into the underlying material and continue extracting small amounts of copper. This maximizes the ultimate level of recovery of copper from the resource without the need for any rehandling of the ore.

An alternative that is suitable for fast leaching ores, say less than 100 days, is a single lift, on-off pad. This would be used mainly for oxides or mixed ores. The approach does require double handling of the ore, once to lay it down (usually with a stacker) and again to remove it (usually with a front-end loader or other type of excavator). The cost of rehandling will be offset by the small size and low cost of the reusable leach pad. This need only support a single lift of ore and will only be sized to hold ore during a relatively short leach cycle. In addition, the spent ore can be given a final rinse with raffinate or fresh water to remove any remaining soluble copper. Thus, there is virtually no loss of soluble copper or any inventory holdup, other than the copper recirculating in the SX raffinate.

In most cases, the washed leach residue can be off-loaded to an unlined spoil area, e.g., atop old tailings or within an abandoned pit. If a lined residue area is required, then much of the advantage would be lost. The residue area would have to be large enough to accommodate the entire tonnage that will be mined and leached. Thus, it would be similar to the permanent multilift leach pad in terms of size and cost.

Other types of leach systems may be appropriate for special situations. One is an in-place arrangement for broken material such as block caved remnant ore, other collapsed underground workings or the rubblized walls of abandoned pits. This approach has worked well at a number of underground mines including Bingham Canyon, Ray, Miami and San Manuel and in abandoned pits at Silver Bell and Mineral Park. True in situ leaching of undisturbed copper ore is another possibility. However, such efforts have been more notable for their degrees of failure than their successes. Only BHP at San Manuel seems to be doing well with in situ leaching. Here the ore is apparently fairly permeable. This is probably due to the weathered nature of the oxide ore. The permeability may have been further enhanced by fracturing from the open pit mining activities that preceded the development of the in situ well field along the old benches.

In the past, vat and agitation leaching have also been used, but these have now been abandoned in favor of various forms of heap leaching. This decision is undoubtedly driven by the high costs and complexities of the agitated leach circuits. Using Twin Buttes as an example, the plant involved the following postcrushing steps: grinding, agitation leaching, washing and liquid–solid separations, including a polishing step to minimize carryover of solids into SX (Hopkins and Lynch, 1985). Today, economics appear to favor a low-cost heap leach approach, even if recovery is somewhat lower and the leach rate is slower.

Selection of Leach Parameters. Selection of the various leach system parameters comes largely from the results of the metallurgical testwork program. These should have included a range of crush size distributions that replicated the discharge from various crushers. Then the crush size is selected so that it represents the optimum balance between copper recovery and the added cost of finer crushing. Likewise, testing should have identified the optimum agglomeration and acidulation conditions. Then value engineering would be applied to select the best method of achieving the desired conditions. This could range from drum or belt agglomeration of ore prior to stacking to acidulation of the ore after it has been placed on the heap. In-place acidulation can be achieved by soaking the ore with acidified raffinate, spent electrolyte, or acidified water.

Lift height is another parameter that should have been tested with tall columns. The intent is to use the highest lift height that is consistent with good solution-rock contact. This will minimize the pad area and thus the costs. When leaching at a given irrigation rate, increasing the lift height will also reduce the total area under leach. This reduces the heap outflow rate while maximizing the copper tenor in the PLS. Both will benefit the downstream SX plant.

Decisions must also be made about options for solution management. These include selection of the optimum irrigation rate. This should minimize the amount of water pumped per unit of copper recovered. However, the rate should not be so low that soluble copper is not washed out of the heap effectively. The need for leach/rest cycles should also be assessed as a way of maximizing recovery from well-leached ore while still maintaining a reasonable copper tenor. This can be achieved with an ILS recirculation system. Here raffinate is advanced to older heap areas with slow recovery. The resulting outflows are then advanced to fresher ore to build copper tenor and reduce the final PLS flow back to SX. The added cost of the ILS pumping and piping system will likely be offset by increased copper recovery and the savings from a smaller SX facility.

A final consideration should be the leach cycle time. This is not simply the time needed to leach the desired amount of copper from the ore during the column tests. The cycle must start with the time needed to stack the ore, cure it as necessary and install the spray piping or drip lines. Any rest cycles should be included in the active leach time. Finally, the time needed to rinse and drain the ore should be included. The drain down should be long enough for the ore to dry to the point where it can be off-loaded or support the stacking equipment needed to add another lift of ore. These nonleaching activities could easily add five or six weeks to the required leach cycle.

Forecasting Copper Production and Acid Consumption. With a conventional milling operation, full recovery is achieved almost as soon as the ore leaves the pit. In a heap leach operation, the ramp-up time is much longer. Just getting the ore out of the mine and onto the heap with piping in place can take two or three weeks. Then with slow leaching ore, that block of material could be contributing copper and consuming (or generating) acid over a period of several months to two or three years. Thus, ramp-up to steady-state cathode output will be a slow affair.

Often the design engineer will be called upon to provide a forecast of both copper production and acid consumption. An effective way of doing this is to start with the metallurgical data and develop a composite copper extraction rate curve and acid consumption curve for the ore. These should include whatever adjustments are necessary to go from laboratory-based curves to those that reflect the various operating constraints discussed above. The curves are divided into monthly increments. Next, the mine plan is used to determine the ore grade and tonnage associated with monthly mine output. Spreadsheets are then developed using the monthly blocks of ore and each month's increment of production and acid consumption. As more and more blocks of ore are added to the matrices, a point is reached where the first blocks mined are no longer contributing to production or acid consumption.

Once the matrices are complete for the life of the operation, the gross copper production and acid consumption from all ore blocks contributing during a particular month are summed. The copper figures are adjusted for losses to the circulating and in-heaps inventories. The result gives the net cathode production that can be realistically expected as production ramps up to steady-state operation.

Adjustments must also be made to the gross acid consumption figures. These all take the form of credits. One of these is the acid generated at the anode during electrowinning. This generates 1.54 units of sulfuric acid per unit of copper plated per the following reaction:

$$H_2O \rightarrow 2H^+ + 2e^- + \tfrac{1}{2}O_{2(gas)}$$

The normal electrolyte bleed also contributes acid, as the acid in the bleed ends up reporting to the raffinate. The bleed rate will be site-specific but typically contributes about 0.25 units of acid per unit of copper plated. The final acid credit is any acid generated by pyrite oxidation in the heaps or dumps. As discussed by Templeton and Schlitt (1997), the cumulative effect of these credits may be sufficient to make many sulfide leach operations net acid generators, rather than net acid consumers.

Conclusions

The keys to good design are an accurate mine plan and results from metallurgical testing that can be extrapolated to give accurate projections of copper recovery and acid consumption or generation. Given this information, the design engineer should be able to work with other members of the project team to provide an optimized leach system and properly sized SX-EW operation. The resulting plant is one that produces the budgeted amount of copper and is easy to operate and maintain.

References

Hopkins, W.R., and A.J. Lynch. 1985. "Anamax oxide plant." In *SME Mineral Processing Handbook*, ed. N.L. Weiss, 14D-2 to 14D-5, New York, AIME.

Schlitt, W.J. 1997. "Proposed method for estimating leach recovery from coarse ores." *M&MPJ*, 14:50–53.

Templeton, J.H. and W.J. Schlitt. 1997. "Method for determining the sulfuric acid balance in copper-leach systems." *M&MPJ,* 14:1–7.

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The Good, the Bad, and the Ugly Lessons Learned in the Design and Construction of Heap Leach Pads

Allan J. Breitenbach, M.S., P.E.[*]

Heap leach operations provide a low-cost method for recovering gold, silver, and copper metals from low-grade ore materials. The processed ore materials are stacked in controlled lifts on a lined pad facility for heap leaching. A controlled barren solution is applied to the heap ore surface by sprinkler or drip emitter irrigation, and ore metals are leached into the solution for collection and extraction in plant process facilities. The leach pad and solution collection facilities are lined with impervious materials for containment of all process solutions. The liner system typically includes a composite geomembrane and clayey soil liner with a protective drain cover fill to maximize solution recovery and protect the underlying groundwater conditions.

This paper discusses the types of pad liner systems currently used by the mining industry and the "good, bad, and ugly" lessons learned in engineering practice over the last ten years in the design and construction of heap leach pads.

INTRODUCTION

The heap leaching process for precious and base metals has developed into a low-cost method for recovering copper, gold, and silver from large and small deposits of low-grade ore. The earliest known heap leaching process with chemicals is attributed to Spanish miners percolating acid solutions through large heaps of oxide copper ore on the banks of the Rio Tinto around 1752 (Van Zyl, et al. editors, 1988). In 1969 the U.S. Bureau of Mines researchers proposed the use of dilute alkaline cyanide solutions as a low-cost means for recovering gold values from low-grade ore containing micron-sized particles of gold (Miners News, 1991). The modern-day heap leaching process typically involves the percolation of leach solutions through controlled lifts of stacked low-grade ore in a heap for chemical extraction of precious and base metals into solution. The

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[*] Senior Project Manager, Agra Earth and Environmental, Lakewood, Colo.

enriched (pregnant) solution is collected by gravity flow in lined basins, sumps, and ponds for processing and recovery of the target metals (copper, gold, and silver). The barren solution from the recovery facilities is recirculated to the heap, in addition to chemical reagents and makeup water, for containment of all process solutions in an enclosed and environmentally sound system. This involves the use of low-permeability liner systems beneath the heap leach stacks to maximize solution recovery and protect the underlying groundwater conditions.

This technical paper discusses the typical heap leach liner systems currently used by mining companies and some of the lessons learned in preventing heap leach pad failures. The most successful heap leach pad liner system from past precious metal operations generally included an underlying low-permeability soil bedding fill, a geomembrane liner, and an overlying protective drain cover fill.

TYPES OF LINERS

The four basic types of liners most commonly used in lining heap leach facilities include (1) clay liners, (2) geomembrane liners, (3) rubberized asphalt liners, and (4) composite liners. The clay liners include low-permeability natural clayey foundation and compacted fill materials as well as manufactured geotextile and bentonite clay sheets. The geomembrane liners commonly include manufactured sheets of relatively impervious polyethylene and polyvinyl chloride materials. The rubberized asphalt liners are a mixture of compacted rubber compound and asphalt materials for typical use in reusable leach pads to accommodate heavy equipment loading and unloading of the leached and spent ore from the lined pad surface. The composite liner is a combination of clayey soils underlying and in direct contact with manufactured synthetic liners as a composite liner system. The composite liner is the most practical and environmentally sound liner system in use today. Each type of liner is discussed in more detail below.

Clay Liners

The three basic types of clay liners for solution containment include (1) natural low-permeability clayey soils, (2) pervious soils amended with bentonite treatment to create a low-permeability soil liner, and (3) manufactured geosynthetic clay liners (GCL). The natural clayey soils have generally been used as the primary or secondary heap liner system, as discussed later in this section. The bentonite soil treatment and GCL liners have seen limited use as heap leach liners mainly due to the lower cost to develop natural clay borrow areas in the vicinity of the project site. The GCL materials have been used primarily in landfills and ponds since 1982 and consist of dry bentonite clay soil enclosed within sheets of woven and nonwoven needle-punched geotextiles. The permeability of a wetted GCL sheet is of the order of 1×10^{-8} cm/sec or less under a few feet of confining soil cover (Estornell and Daniel, 1992).

The natural clayey foundation soils or surficial clayey soils compacted to moderate density typically have been used in the past as primary liners for copper heap leach operations (acidic sulfuric solution). A compacted clayey soil liner in combination with manufactured geomembrane liners typically have been used in the gold and silver heap leach operations (alkaline cyanide solution). The compacted clayey soil liners provide an effective impervious barrier beneath the ore heaps considering the affects of a reduction in soil permeability (hydraulic conductivity) under increasing heap loads and the plastic deformation or self-healing behavior of clayey soils during potential foundation settlement movements.

The hydraulic conductivity of clayey soil liners can vary depending on the in-place moisture content, density, exposure time to climatic conditions, and cover fill loads; however, the general range is between 1×10^{-6} and 1×10^{-8} cm/sec. Past laboratory permeability test results on clayey soil liner samples under simulated heap loads to 100 ft (30 m) generally show an order of magnitude reduction in permeability due to clay liner consolidation. It should be noted that geotechnical engineers classify soils with a hydraulic conductivity in the 1×10^{-5} to 10^{-7} cm/sec range as very low permeability material, and practically impervious below 1×10^{-7} cm/sec (Terzaghi and Peck, 1967).

In addition to the low-permeability compaction and consolidation characteristics of clayey soils, the relative past success of clay-lined basins for containment of valley copper heap leach operations may also be attributed in part to the apparent chemical reaction of clays to acidic solutions. Past laboratory testing by others has shown that acidic liquids, when permeated through clays, can result in decreases in the hydraulic conductivity of the clay (Bowders, 1987; Peterson and Gee, 1984). Particles in the uppermost surface of the clayey soil liner dissolve when contacted by the leachate. As the leachate further permeates into the clayey soil liner, additional dissolutioning results in raising the pH of the solution, thereby causing some precipitation of the dissolved particles. These particles in turn cause clogging of voids in the clay, which result in a reduced hydraulic conductivity of the soil liner (Gordon, et al., 1989). Thus, the clayey soil liner can essentially become the primary liner under increased ore heap loads and potentially can develop an impervious surface boundary to acidic copper leach solutions.

Geomembrane Liners

The geomembrane liners have been used extensively in gold and silver heap leach operations since the late 1970s and in recent times for copper heap leaching. A properly installed and protected geomembrane liner can be roughly equivalent to more than 100 ft (30 m) of low-permeability clayey soil.

The polyethylene and polyvinyl chloride materials were first developed in the 1930s and 1940s for use as packaging and molding materials or potable water liners (Koerner, 1990). The polyvinyl chloride (PVC) liners were first on the market for industrial use in the 1950s (U.S. Bureau of Reclamation water canal projects in the western U.S.). The ultraviolet (sunlight) resistant or oil and gas resistant counterpart to regular PVC liners are sometimes designated as UV-PVC or OGR-PVC liners. The high- and medium-density polyethylene (HDPE/MDPE) liners were developed in West Germany by the mid 1960s for dam embankment facing and impoundment liners. The chlorinated and chlorosulfonated (CPE/CSPE) liners generally were developed in the mid 1970s as pond liners. The CSPE liners are better known as Hypalon. The very low, very flexible low, and linear low density polyethylene (VLDPE/VFDPE/LLDPE) liners were developed in the late 1980s primarily for landfills and heap leach pads. These low density polyethylene liners are sometimes refered to as LDPE liners for simplification purposes, since the material elongation and interface friction strength properties are similar. A relatively new polypropylene (PPE) liner has come on the market in the early 1990s initially as a process pond cover liner and currently is being used as a pad liner on several cold weather heap leach projects. Each of the liner systems can be successfully designed for heap leach operations; however, the three most common types of liners used in the past for heap leach pads include the PVC/UV-PVC/OGR-PVC, HDPE/MDPE, and VLDPE/VFDPE/LLDPE geomembrane liners.

Rubberized Asphalt Liners

The rubberized asphalt liners have seen limited use beneath permanent heap leach pads generally because of the relatively brittle nature of asphalt liners with time and under high heap loads. These liners are typically used on small-scale reusable pads for equipment traffic loading and unloading and require a relatively thick compacted granular soil or bedrock foundation for firm support to prevent differential liner movement. A secondary liner system is generally constructed beneath the asphalt liner to contain any leakage from cracks in the asphalt layer.

Composite Liners

The composite liner is the state-of-the-art liner system for solution containment and generally includes a low-permeability clayey soil liner in direct contact with an overlying geomembrane liner as a composite liner system. The geomembrane liner permeabilities are estimated to be on the order of 2×10^{-11} cm/sec or less (U.S. Federal Register, 1987). However, pinhole leakage can occur in geomembrane liners during manufacturing, transportation, installation, and subsequent heap placement operations. Thus, it is important to place a low-permeability clayey soil in direct contact with the geomembrane liner. This assumes a low hydraulic head is maintained above the liner with adequate cover fill drainage, as discussed in the design section of this paper.

PAD LINER DESIGN

The primary purpose of the pad liner design is to accommodate the placement and leaching of ore in controlled lifts and prevent the loss of pad and pond process solutions from the lined facilities for both economic and environmental reasons. The geomembrane liner must function for the life of the facilities considering all likely design, construction, and operational factors such as the local topography, geology/seismicity, climatic wet seasons, and other site-specific conditions. This section focuses on typical engineering design aspects of a composite liner system to prevent leakage and minimize the risk of pad liner failures, including the design of the underlying bedding fill, geomembrane liner, and overlying cover fill materials.

Underliner Bedding Fill Design

The underlying fine-grained bedding fill provides a secondary containment barrier for leach solutions and also protects the overlying geomembrane liner from subsurface rock puncture. The bedding fill design generally includes the following: (1) a fine-grained low-permeability soil with a maximum minus ¾-inch rock size, (2) a moisture content within optimum to two percent dry of optimum moisture content (ASTM D–698), (3) a compacted firm and smooth surface, and (4) a top surface graded to drain to solution collection systems for positive gravity drainage.

Limiting the bedding material maximum rock size to less than ¾-inch will minimize the risk of geomembrane rock puncture. The geomembrane sheets are generally dragged across the compacted bedding fill surface during deployment, which may loosen and roll the bigger rock particles on top of the bedding surface. These rock protrusions are more susceptible to puncture beneath the liner than the placement of more coarse rock materials above the liner surface. Some selective finer borrow excavating, rock screening, or hand picking may be necessary to maintain recommended minus ¾-inch maximum rock sizes on the final soil liner surface prior to smooth steel-drum or rubber-tired roller compaction for a firm and smooth final soil liner surface in preparation for geomembrane liner placement.

The hydraulic conductivity of clayey soils has been demonstrated in laboratory permeability tests to decrease to the lowest value when placed wet of optimum moisture content (Hermann and Elsbury, 1987). However, high plasticity clayey soils placed wet of optimum moisture content have a tendency to desiccate and crack, even when covered by geomembrane liners, compared to clayey soils placed at or dry of optimum moisture (Daniel and Wu, 1993). Desiccation cracks may not heal sufficiently under heap loads and potentially can compromise the clayey soil liner integrity in preventing solution leakage. Another reason for placing clayey soil materials at or within two percent dry of optimum moisture content (ASTM D–698) is that potential rutting and bearing capacity failure during rapid loading in overly wet clayey soils can create harmful stresses in the overlying geomembrane liner.

Geomembrane Liner Design

A geomembrane liner beneath heap leach stacks provides a primary containment barrier for leach solutions, with adequate underlying bedding fill and overlying drain cover fill protection, and proper liner selection and installation. As discussed earlier, the three basic types of geomembrane liners used for heap leach pads include PVC, HDPE, and LDPE liners, and recently PPE liners. Each type of liner is chemically resistant to the acidic and alkaline solutions used in the heap leach operations (Koerner, 1990).

The leach pad geomembrane liner selection must consider all engineering, construction, and operational aspects of the project for the most effective overall liner to prevent leakage. Factors in design affecting the selection of the most suitable geomembrane liner at each project site include (1) the liner type; (2) the liner thickness (40-mil = 1-mm thickness); (3) the initial loading conditions from the cover fill or first ore lift; (4) the final loading conditions from the maximum heap height load; (5) an adequate engineering construction quality assurance (CQA) program for acceptable bedding soil, geomembrane liner, and cover fill materials as well as seam tests for strength and water tightness; and (6) a suggested geomembrane liner evaluation to select the most economical and functional liner to accommodate site specific conditions.

An overall liner evaluation for selection typically compares the pros and cons of the various geomembrane liners used in past heap leach operations from a combined design, construction, and operation point of view (Breitenbach, 1992). Some of the more important engineering aspects in liner selection include geomembrane liner resistance to rock puncture, adequate liner friction strengths for slope stability, elongation capacity to withstand foundation settlements under high heap loads, and long-term exposure to climatic conditions. Some of the more important construction and operational aspects in liner selection may include maximizing storage in shipping containers for trucking and overseas transport, minimizing unloading and handling at remote sites, ease in liner installation (deploying, seaming, testing, and covering), and suitable liner anchorage and pad access grades for loading the ore in controlled lifts without damaging the liner from dozer, conveyor, or truck traffic. The more important evaluation factors at each site can be appropriately factored to select the optimum geomembrane liner type and thickness to suit site-specific conditions.

Overliner Cover Fill Design

The general design requirements for overlying liner cover fills consider several aspects including (1) the maximum rock particle size at the liner surface contact, (2) the rock particle shape, (3) the cover fill lift thickness, (4) adequate solution drainage, (5) available or planned equipment for placement and spreading, and (6) the type of geomembrane liner for withstanding initial dynamic construction traffic live loads and final heap dead loads.

The cover fill is sometimes incorporated within the first ore lift for finely crushed ore (minus one-inch nominal rock sizes) or agglomerated silty and clayey ore materials. Larger rock sizes are possible in the cover fill depending on site-specific conditions and the geomembrane liner type and thickness. Sometimes a geofabric cover with adequate friction strength can be used above the more flexible liners to place primary crush ore materials on the pad surface. The finer ore materials acceptable for cover fill in the first ore lift are generally placed by truck/dozer or radial stacker/dozer operations.

Minimum cover fill lift thicknesses and maximum rock sizes are generally specified by an experienced engineer, based on past performance, and verified during construction. Construction quality assurance personnel under the supervision of the engineer, or construction quality control (CQC) mine personnel, are responsible for maintaining acceptable cover fill operations, including construction test fills if deemed appropriate by the engineer.

The cover fill ideally is more pervious than the overlying ore lifts for maintaining low hydraulic heads on the liner system to minimize the risk of leakage and allow quick metallurgical recovery of solutions for operational control. Corrugated and perforated drain pipes are typically placed on the geomembrane liner surface in low- or flat-lying areas to supplement cover fill drainage, particularly on slopes less than a five-percent grade and along major pad cell or valley drainages.

It is suggested that all geomembrane liners include a cover fill for protection from biological, climatic, and man-made conditions. The flexible liners especially require a cover fill for limited exposure time periods of preferably less than a year after installation, unless properly protected (white coated surfaces, laminated HDPE/LDPE/HDPE sheets, ultraviolet (UV) resistant PVC, etc.). The crystalline polyethylene liners on large pad areas are subject to stress cracking from temperature changes, folding, and wind movement. Less crystalline polyethylene liners and regular polyvinyl chloride liners may lose flexibility with time from exposure.

SELECTED MAJOR ENGINEERING CONCERNS

The large-scale existing and planned copper heap leach pads and some of the recent gold heap leach pads have estimated operating time periods of 10 to 15 years compared to the more typical past gold and silver heap leach operational time periods of 3 to 5 years. Therefore, this author believes the long-term durability of the geomembrane liner is the most important liner design consideration for high fills, which includes geomembrane liner stability and exposure concerns for compatibility with site-specific conditions. Other major concerns (chemical compatibility and rock puncture) can easily be tested and verified in the laboratory under simulated worst-case conditions. Due to the elongation properties of the flexible geomembrane liners, foundation settlement has been a lesser concern to date, but may need to be considered in the design of site grading foundation fills to maintain positive gravity drainage beneath the high fill loads. A discussion on selected liner stability and exposure concerns follows.

Pad Liner Stability

The geomembrane liner surface generally has lower friction strength characteristics compared to the overlying fill and underlying foundation materials, thus the liner system requires special design and operational considerations to prevent instability. In a review of published information and second-hand information from clients and colleagues, this author is aware of 12 known pad slope failures on geomembrane-lined heap leach pads in North America, South America, and Australia (Breitenbach, 1997), as well as a large lined landfill slope failure in California (Mitchell et al., 1990). These geomembrane-related liner slope failures were caused by several factors including the following most probable single or multiple causes concerning leach pads:

- placement of the first ore lift on pad grades steeper than 5 percent,
- placement of the first ore lift higher than 40 feet (12 meters) in vertical height at the natural angle of repose,
- placement of the first ore lift in the down-gradient pad liner direction,
- placement of the ore heap on a weak foundation,
- placement of geotextiles in direct contact with the geomembrane liner surface for very low friction strengths at the interface,
- heap loading on poor to moderately compacted valley berm fills in which some berm settlement and subsequent heap movement may have reduced the stacked ore friction strengths from the higher peak strength before movement to a lower residual strength along the slide surface,
- placement of high plasticity clayey bedding materials wet of optimum moisture content causing high pore pressure (low strength) conditions or slow-movement plasticity creep failure with heap loading, and
- placement of ore lifts on snow or frozen ice sheets at the geomembrane liner surface.

Adequately designed site grading fills by an experienced engineer in the critical down-gradient toe areas of the heaps (high toe berm fill or grades flatter than 5 percent) in combination with liner selection and experienced operator heap stacking operations would have eliminated most of the known pad slope failures. It is interesting for operators to note that the majority of the slope failures occurred during placement of the first ore lift, mostly before leaching. Engineers may want to note that the majority of the geomembrane-lined slope failures occurred on the more rigid geomembrane liners. Textured geomembrane liner sheets are frequently being designed and constructed in current practice to improve on the smooth liner surface friction strength. It is worth repeating that these geomembrane-lined slope failures could have been avoided with proper design, construction, and stacking of ore.

This list of probable causes for known slope failures does not include other types of geomembrane liner failures related to natural causes such as stress-cracking from climate exposure (Paradise Peak in Nevada), potential snow avalanche damage (Summitville in Colorado), or wind storm damage (Ridgeway in South Carolina). It also does not include the more common man-made geomembrane leakage failures such as excessive foot traffic in exposed areas, poor quality liner materials and installation, or inadequate engineering inspection of the bedding fill, geomembrane liner, and cover fill operations.

Pad Liner Exposure

Geomembrane pad liners of all types are recommended to be covered with an overlying cover fill for adequate long-term protection from exposure. The various geomembrane liner exposure problems may include ultraviolet (sunlight) sensitivity, thermal expansion and contraction with temperature changes, high wind and snow movement, equipment and foot traffic punctures, potential blast fly rock from the mine, and oversized rock roll-out onto the liner surface during rocky haul road ramp construction and first ore lift placement. The geomembrane liners are relatively thin membranes and need all the cover fill protection the owner can afford. It is difficult to damage the geomembrane liner after it has been properly covered by a few feet of suitable cover fill material.

In some cases such as the side valley walls of a leach pad basin, the geomembrane liner may be constructed in phases to avoid more than 1 to 3 years of exposure, until the relatively level ore stack lifts cover the liner. The time of acceptable or allowable exposure can be verified by conventional (real-time) outdoor exposure tests or outdoor accelerated weathering tests for flexible membrane liners.

Another exposure problem is the potential development of high tensile forces in geomembrane liners exposed on long steep slopes or in large open areas where liner installation occurs in the summer but remains exposed until covered in cooler seasons. The latter case has been observed on one project in Nevada in which a geomembrane liner was being pulled out of a backfilled, 2-foot-deep anchor trench on a flat pad liner surface. Multiple anchor trenches may be required on steep slopes for the high expansion and contraction liner types where heated liners can expand down-gradient but must overcome gravity and soil friction to contract back up-gradient when cooled.

CONCLUSIONS

This paper presented a general discussion of the various types of pad liners used in the mining industry, and the major engineering design concepts and concerns for geomembrane-lined systems beneath high fill structures such as gold and copper heap leach pads. These concepts and concerns may also apply to landfills and tailings impoundments, but for purposes of this article were focused on heap leach pads.

The three most common types of geomembrane liners used for heap leach pads include PVC/UV-PVC/OGR-PVC, HDPE/MDPE, and VLDPE/VFDPE/LLDPE liners. The geomembrane liner must function for the life of the facilities considering all pertinent site-specific conditions as well as planned construction and operation conditions. Some engineering judgment is required in the overall geomembrane liner selection and design to economically utilize on-site materials as much as practical in the liner system. The engineering judgment is based in part on past liner experiences and in part on appropriate laboratory geomembrane liner and project site material tests.

Each type of liner has been successfully used in heap leach operations with proper liner selection, testing, design, installation, and heap stacking operations. Thus, the success of liner projects can be directly attributed to an experienced team effort by the design engineer, followed by the liner fabricator/installer and construction testing personnel, and finishing with the operator of the facilities.

ACKNOWLEDGMENTS

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REFERENCES

Bowders, J.J. Jr. and Daniel, D.E. (December, 1987); "Hydraulic Conductivity of Compacted Clay to Dilute Organic Chemicals," *Journal of Geotechnical Engineering*, ASCE, Vol. 113, No. 12, pp. 1432 to 1448.

Breitenbach, A.J. (November, 1992); "Comparison of Geomembrane Liners for Heap Leach Pads," *Geotechnical Fabrics Report*, Vol. 10, No. 8, pp. 41 to 45.

Breitenbach, A.J. (March, 1997); "Geomembrane Pad Liner Failures Under High Heap Fill Loads," *Geosynthetics '97 Conference*, Industrial Fabrics Association International (IFAI), Long Beach, California, Mining Session, Volume 2, pp. 1045 to 1062.

Daniel, D.E. and Wu, Y. (February, 1993); "Compacted Clay Liners and Covers for Arid Sites," *Journal of Geotechnical Engineering*, ASCE, Vol. 119, No. 2, pp. 223 to 237.

Estornell, P. and Daniel, D.E. (October, 1992); "Hydraulic Conductivity of Three Geosynthetic Clay Liners," *Journal of Geotechnical Engineering*, ASCE, Vol. 118, No. 10, pp. 1592 to 1606.

Gordon, M.E., Huebrer, P.M. and Miazga, T.J., (August, 1989); "Hydraulic Conductivity of Three Landfill Clay Liners," *Journal of Geotechnical Engineering*, ASCE, Vol. 115, No. 8, (Paper 23786).

Hermann, J.G. and Elsbury, B.R. (1987); "Influential Factors in Soil Liner Construction for Waste Disposal Facilities," *Geotechnical Practice for Waste Disposal '87*, R. Woods, ed., ASCE, pp. 522 to 536.

Koerner, R.M. (1990), "Designing with Geosynthetics," Prentice Hall, New Jersey, 2nd Edition, Section 5, pp. 387 to 389.

Mitchell, J.K., Seed, R.B., and Seed, H.B. (1990); "Kettleman Hills Waste Landfill Slope Failure, Volume I: Linear Systems Properties," *Geotechnical Engineering Journal*, ASCE, 116(4), pp. 647 to 668.

Peterson, S.R. and Gee, G.W. (1984); "Interactions Between Acidic Solutions and Clay Liners: Permeability and Neutralization," ASTM, STP 874, Philadelphia, Pennsylvania, pp. 229 to 245.

Terzaghi, K. and Peck, R.B. (1967); "Soil Mechanics in Engineering Practice," 2nd Edition, John Wiley and Sons, New York.

U.S. Federal Register (April, 1987); "Synthetic Liner Permeabilities," Vol. 52, No. 74.

Van Zyl, D., Hutchison, I., and Kiel, J., editors (1988); "Introduction to Evaluation, Design, and Operation of Precious Metal Heap Leaching Projects," Chapter 1, Society of Mining Engineers, Littleton, Colorado.

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Design Strategies for Heap Leach in Extreme Environments

Richard Frechette[*]

Physical environments associated with mine projects are very diverse and include some of the most extreme conditions found in the world. Environmental controls and strategies for heap leaching systems must take into account the extreme conditions particular to a given site in order to successfully permit, operate and close the project. Extreme conditions that are considered in this paper include cold weather, high rainfall, desert (extremely low rainfall), high altitude and large earthquakes. Design aspects addressed include water balance, heap configuration, solution management, solution storage options, caps and covers.

INTRODUCTION

Improper design of a leach pad to accommodate the environment, or operation in a manner that is not compatible with the design, have the potential to result in reduced efficiency of the operation, unanticipated additional cost to maintain operation, direct and indirect costs in the event of a noncompliance situation, remediation costs in the event of a failure and possible premature termination of operations. In the case of a heap leach facility sited in an extreme environment, any of these possible outcomes must be acknowledged and addressed. The context for discussion in this paper regarding extreme environment is limited to climatic extremes, adverse terrain and severe earthquake potential.

CHALLENGES OF CLIMATE AND TERRAIN

The following site characteristics are typically troublesome with respect to design, construction and/or operation of heap leach facilities:

- Cold weather
- Net precipitation
- Net evaporation

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[*] SRK Consulting, Lakewood, Colo.

- Steep terrain
- Extreme seismicity

In extreme cold weather, difficulties in maintaining solution circulation due to freezing can occur and process ponds can freeze up.

At a net precipitation site, water balance problems can develop that require discharge of excess solution or could result in an uncontrolled release of process solution.

Conversely, at a net evaporation site, water shortages can result in expensive measures to obtain adequate water supply or the water deficit can result in severe restrictions on operating that can have substantially adverse effects on the project.

Steep terrain can make access for construction very difficult and result in slope stability problems that could endanger the operation of the facility.

Similarly, large earthquakes can have an adverse impact on stability of a heap leach facility and potentially cause failure of the containment system, creating environmental impacts and interrupting operation of the facility.

SOLUTIONS AND EXAMPLES

Cold Weather

In extreme cold weather, solution management can be severe from both the application perspective as well as storage/recovery. Under these conditions, specific measures may be required to enable the continuous recycle of process solution and prevent freeze-up of the system. Insulation of the application system as well as the solution storage and processing circuit pipework may be necessary.

One means of insulating the application system pipework is achieved by placing a sufficient cover layer of ore over the drip irrigation network. The required layer thickness is a function of the anticipated ambient temperature and thermal properties of the ore and circulating solution. Heating of the solution as it exits the process circuit and is reapplied to the heap may assist in the thermal balance. At a recently constructed heap leach project in the Alaskan interior, waste heat was used to elevate the temperature of the barren solution as it was recirculated through the buried emitters to the heap during winter leaching operations.

The pregnant solution reservoir may be insulated in a similar manner via in-heap solution storage (see Figure 1). This requires specific design considerations relating to the project water balance and ore characteristics associated with moisture retention properties. Ore placement in the solution storage basin must be carefully coordinated with the mine plan and the availability of the appropriate ore material. Placement of material that is too fine or lacks sufficient durability may contribute not only to the loss of circulation rate and reduced process flow but also may result in a reduced solution storage capacity that could create a compliance issue. In the event of extreme snowpack or rainfall, a reduced capacity may present an increased risk of discharge.

High Rainfall (net precip)

In an extreme net precipitation environment, solution management requirements sometimes include creative means to deplete excess water accumulated in the system and the provision for treatment and discharge of solution to prevent breaching of storage facilities.

FIGURE 1 **Ore placement at in-heap solution storage facility (Alaska heap leach project)**

The most efficient means of "depleting" excess solution is via take-up in the ore as it is initially wetted following placement on the pad. This typically represents the single largest consumption of water on a heap leaching site and can entail up to 300 gpm of loss or more at a large heap leach operation. This loss is relatively consistent as long as ore is being mined and placed on the leach pad. At preparation for closure, as mining is reduced and ore depleted, other means of solution balance must be instituted to avoid excess buildup and the eventual need to discharge.

Sizing of ponds and sequencing the use of ponds in a net precip environment is also important. By providing large enough ponds, and a contingency pond sized for a specified event, the need to discharge can be diminished and potentially eliminated. This requires development of a comprehensive site water balance that accurately accounts for all significant inflows and losses and allows estimating seasonal storage requirements that must be accommodated. Planning for discharge must entail forethought to provide appropriate treatment to allow discharge prior to the onset of conditions that could produce an uncontrolled release of untreated effluent. Hydraulic connection of ponds using passive mechanisms such as open channel spillways and sequencing of ponds or establishing consistent pond crest and spillway inlet elevations will ensure that the maximum volume is engaged prior to initiating a discharge if cumulative capacity limits are exceeded.

The use of temporary or permanent covers to divert rainfall from the solution collection system has been employed during the rainy season to reduce excess inflow to the process ponds at an operation in Panama that receives 2 meters of rainfall annually. The covers are low-thickness geomembranes that are installed over the outer slope faces of the heap and drain to double-lined ditches that direct the rainfall runoff to an overflow pond, bypassing the process ponds (see Figure 2). Drip emitters are installed under the liner covers and are used to leach the heap sideslopes. The delivery lines are exposed to the

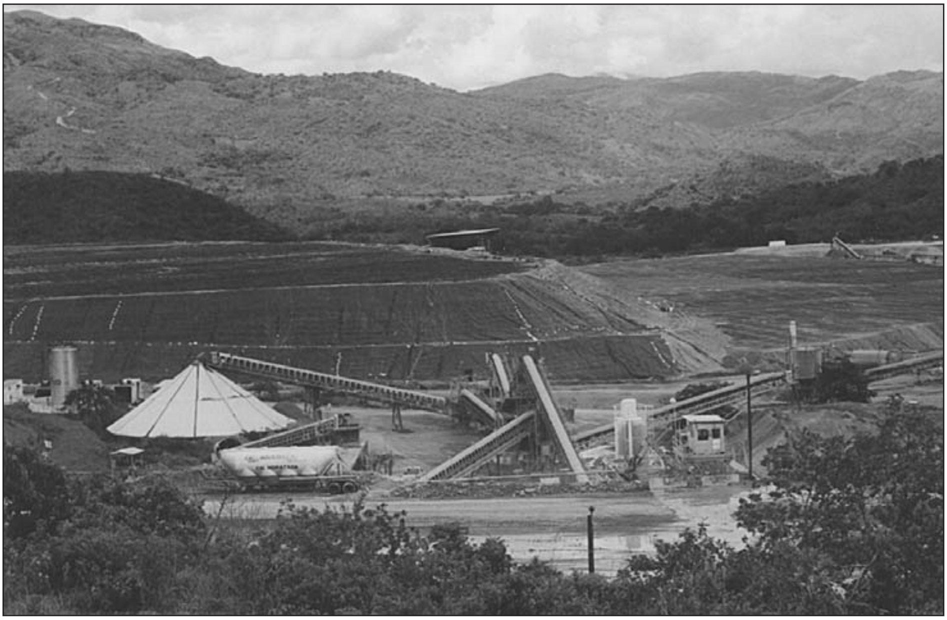

FIGURE 2 Heap side-slope cover in high rainfall area (Central America heap leach project)

point where headers split off and connect to emitters underneath the liner cover. Leachate from all areas of the heap, including beneath the covered slopes, is transported between the double-ditch liners to the process ponds for metal recovery.

Low Rainfall (net evap)

The opposite concern exists for projects in areas of net evaporation, particularly where make-up water is in short supply. In this case, the need exists to conserve water through such means as the use of low evaporation application systems, covered ponds to reduce solar evaporation and scheduling leaching or other water use for nonpeak evaporative hours, e.g., nondaytime hours. It is rare for these measures to conserve sufficient quantities of water to eliminate the water deficit in a net evaporation environment, but the water savings can be significant. If water supply is plentiful despite the net evaporation condition, the measures identified may not be important and may, in fact, represent incremental project costs.

On the positive side, in such an environment, sizing of the process/storage ponds can be typically smaller than their counterparts in the net precipitation environment, resulting in lower capital cost. In addition, the risk of an uncontrolled discharge or the potential need for treatment is generally alleviated. These savings have the potential to more than offset the costs associated with increased water supply needs.

High Altitude and Steep Terrain

High altitude leaching is being conducted at an increasing number of properties in mineralized regions of the Rocky Mountains in North America and the Andes mountain range of South America. Factors directly related to high altitude include reduced performance of machinery and personnel that can adversely affect project efficiency. Indirect

FIGURE 3 **Substantial regrading, use of textured liner and intermediate anchor trenches at a steep leach pad site in Central America**

factors can include extreme cold temperatures and severe terrain. Cold temperatures were addressed previously. Severe terrain can adversely impact construction progress and affect stability and associated design parameters that can increase capital costs for the project. Increased costs could result from specialized liner selection and/or additional earthwork requirements to produce acceptable construction conditions and stability factors. Textured liner that affords an increased friction angle relative to smooth liner is priced at a premium over its smooth counterpart. Earthwork on a steep site is not only difficult but typically results in a large earthwork effort to create even modest improvements in terrain due to slope angle differentials. A combination of earthwork improvements and liner considerations is typically used to optimize cost and produce acceptable stability (see Figure 3).

Large Earthquakes

In seismically active areas, the potential exists for frequent earthquakes and large-magnitude earthquakes. Strong ground motion has the potential to cause movement of natural and man-made earthen features and structures. Movement can result in failure of a leach pad and loss of containment of process solution. This can result in a one-time event that impacts the environment or an ongoing situation where processing time is lost due to facility disruption, damage and repair requirements. Mitigation strategies include designing for a specified ground motion with no movement of the facility for a reasonably expected site acceleration or, in the case of extreme seismic conditions, designing for tolerable movement under maximum credible earthquake ground accelerations. Allowable movement is typically a finite quantity that can be experienced by the heap without incurring liner damage or disruption of solution management facilities. Deformation analyses may require the use of sophisticated finite element software and is typically more costly and time-consuming than basic limit equilibrium slope stability analysis techniques.

CONCLUSIONS

Preventive measures to deal with extreme conditions can sometimes be costly but, if identified and implemented properly, are usually much less expensive than remedial costs associated with facility repair or modification, or the costs associated with reduced recovery efficiencies or premature project closure. Indirect costs associated with project failures that result in environmental impacts include damage to the credibility of the mining industry and its engineering consultants. This damage may be very difficult to quantify but sometimes even more difficult to repair.

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Parallels in the Development of Copper and Gold Heap Leaching Technology

Jock McGregor* **and Dirk Van Zyl**†

The principles of copper dump leaching have been understood and applied in Europe and Asia for over 400 years. Heap leaching of gold ores has been done since the early 1970s in the United States and is now commonly practiced on an international basis. Copper heap leaching, i.e., crushed and/or agglomerated ore placed on lined facilities with careful management of solutions, has developed during the last 15 years. This paper explores the parallels, and differences, in the development and implementation of copper and gold heap leaching technology. Ore testing, site development, operations and closure issues are used as a basis for the discussions. Applications of bacterial leaching to both copper and gold leaching are also discussed.

INTRODUCTION

This paper presents a general overview of the major parallels and differences in the development and implementation of copper and gold heap leaching technology. It is aimed at the engineers who are more familiar with the techniques applied to one metal rather than the other. Some of the differences are obvious and some are not so obvious, but the continuing cross fertilization of techniques and ideas are important for the continued development of efficient and cost-effective extraction techniques for low-grade ores.

BACKGROUND

The principles of copper dump leaching have been understood and applied in Europe and Asia for over 400 years. Heap leaching on the other hand is a relatively recent innovation (Van Zyl, et al., 1988). For the purpose of this paper, heap leaching is defined as leaching of prepared ore placed on a lined heap leach pad with an engineered solution

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* Vista Gold Corporation, Denver, Colo.

† TRC Companies, Englewood, Colo.

collection system. This definition purposely excludes run-of-mine (ROM) leaching of both copper and gold ores that will be referred to as dump leaching in this text.

In the late 1960s and early 1970s, at a time when the federal government understood the value of a strong domestic natural resource industry and funded the U.S. Bureau of Mines, the Bureau's research offices both in Salt Lake City and Reno were working together and with private sector companies to develop heap leaching of low-grade gold ores in the western United States (Heinen, et al., 1978). The first commercial applications of gold heap leaching were at Carlin, Nevada (Hiskey, 1983), and Cortez, Nevada, in the early 1970s and the first integrated heap leaching/carbon recovery plant was at the Cortez Gold Acres site in the mid-1970s (McQuiston and Shoemaker, 1975).

The first copper leach project was Rancher's Bluebird Project in Arizona, which integrated copper dump leaching with a solvent extraction recovery plant and started operations in 1968 (Miller, 1967). While not falling within the definition of heap leaching used in this paper, this project is generally considered to be the beginning of the growth in copper heap leach technology.

The introduction of solvent extraction technology as an alternative to cementation with iron for copper recovery, and the introduction of carbon adsorption technology as an alternative to zinc precipitation for gold recovery, gave momentum to the use of heap leaching as an economic extraction process.

ORE CHARACTERISTICS

Ore suitable for heap leaching must exhibit certain characteristics:

- It must contain sufficient recoverable mineral values to justify the cost of ore preparation.
- Ore preparation must be able to expose the metal values so contact with the leach liquor can occur.
- The ore either naturally, or through preparation by agglomeration, must have the physical strength and stability to allow percolation of the leach solution through the heap.
- The ore must be chemically suitable so it does not consume or destroy the leaching agents or adsorbs solubilized metal values.

Gold

Gold generally occurs as elemental gold or alloyed with silver as electrum. Leaching is almost exclusively carried out using a diluted alkali cyanide solution. While the initial impetus to develop heap leaching focused on limestone and dolomitic siltstone-type ores containing submicron particles of gold (Heinen, et al., 1978), a wide range of gold ores are currently processed by heap leaching techniques. Both oxide and sulfide ore can be readily leached provided the gold in sulfidic ores is only associated with and not locked in the sulfide matrix. Sulfides such as enargite ($Cu_3 As S_4$) that are soluble in cyanide can be detrimental, however, as they increase cyanide consumption. Other components of the ore that consume cyanide (cyanicides), adsorb oxygen, reduce the pH of the solution, or adsorb gold cyanide complexes ("preg robbers"), such as some naturally occurring carbonaceous materials and clays, can all have a detrimental effect on gold recovery by heap leaching. In general, gold ores containing around 0.04 ounces of gold per ton of ore are suitable candidates for heap leaching, though many factors such as strip ratio, crush size, recovery, economies of scale and location influence the grade that can be economically leached.

Copper

Copper, as well as occurring in its native form, occurs in a wide variety of both oxide and sulfide minerals. Leaching of oxide copper is done in an acidic environment with a dilute solution of sulfuric acid, and any components of the ore which are basic will adversely effect leaching and acid consumption. Unlike gold, the presence of sulfides can be beneficial as the slow leaching of sulfides generates acid and enhances the leaching of the oxide copper. Some copper sulfides are readily leachable, but the mechanism is different as it is now generally agreed that copper sulfide leaching is accomplished by the ferric ion in solution and leaching is enhanced by the presence of bacteria which catalyze the reaction by converting ferrous to ferric ions.

In general, copper ore of 0.4% and above are suitable candidates for heap leaching, but again many factors go into the equation of what constitutes an economic heap leachable orebody. Detailed exploration of this topic is beyond the scope of this paper.

ORE TESTING

For both copper and gold ores, it is important to properly test the ores before committing to the costs of a full-scale operation. Laboratory and pilot testing normally consists of the following steps:

1. *Mineralogical and Petrographic Examinations.* By relatively inexpensive examination of polished sections of ore samples, the type and location of valuable minerals in the ore can be determined and the gangue minerals can be characterized. This information can then be used to structure a more meaningful and focused metallurgical test program.
2. *Bottle Roll Tests.* Bottle roll tests are used to get a “feel” for the ore’s metallurgical response and compare different types of ore and spatial variables within an orebody. Bottle rolls can highlight obvious problems and indicate the maximum reagent consumptions and extractions that can be expected.
3. *Column and Pilot Heap Tests.* These tests are generally run on a variety of rock sizes and conditions to determine rate of recovery and ultimate recovery at different crush sizes, as well as to check out other possible operating conditions such as agglomeration, reagent strength and ferric or acid cure techniques in the case of copper ores.

   Columns should be of a large enough diameter to avoid serious channeling or sidewall effects and tall enough so that a reasonable sample volume is tested. Often there is a trade off between column size and sample availability due to the high cost of gathering metallurgical samples.

   Finally, several large-diameter, full-height columns or a pilot heap should be tested to confirm the selected leach parameters and get a more realistic measure of reagent consumption and recovery rate. If access is available to an operating mine, large-scale tests are easier than shipping large tonnages to a laboratory. This is particularly true for a pilot heap test.

In the case of copper, more and more projects are evaluated for leaching both oxide ores and secondary sulfide and chalcocite-rich copper deposits. As a result, it is increasingly important to determine, during analyses and testing, the relative abundance of acid soluble (or oxide) copper, “leachable” sulfide copper and “insoluble” copper such as chalcopyrite. Sequential analytical techniques have been developed which use the acid and cyanide solubility of different copper minerals to determine the relative type of copper in the ore (Parkison and Bhappu, 1995). This technique is very effective but should be used in conjunction with mineralogical studies because the presence of certain minerals such as enargite, which is soluble in cyanide but does not leach in a heap environment, can create false results if it is not identified.

SITE SELECTION AND DEVELOPMENT

Heap leaching, by the definition used in this paper, requires some form of ore preparation, which means the ore must be crushed and placed on a conveyor. Once on a conveyor, the ore can be transported relatively cheaply to a suitable site for leaching. Suitable heap leach sites need to have a slope to allow leach solution to flow to a collection point but cannot be so steep that heap stability is impaired or that they are impractical to line with low permeability liners. Other considerations are generally based on factors such as depth to groundwater, methods for heap building, etc.

Heap leach pads fall into three categories:

1. *Valley-Fill Pad.* A suitable valley is selected, preferably with a limited upstream catchment area, a berm is erected across the valley and the pad is prepared behind the berm.
2. *Single-Use Pad.* A suitable area is selected and a pad is built with the correct slopes, etc., to channel solution to a collection point.

For both the valley-fill and single-use pad systems, a wide variety of designs have been used. Some have ponds internal to the pad, some are external. Some have multiple ponds, etc. Generally topography, real estate availability and environmental considerations dictate the design.

Valley-fill leach pads are used at a number of gold mines. Crushed ore, which is stable under large overburden loads, allows a sufficient pore space for the storage and pumping of pregnant solutions. No copper valley-fill leach pads have been constructed to date. The potential chemical weathering of copper ores during leaching (discussed later) must be carefully considered in the case of valley-fill leach pads for copper. Extensive weathering during ore testing is a clear indicator that a valley-fill leach pad will not be feasible.

Site development consists of clearing and grubbing the planned area and installing the liner system to allow collection of the leach liquor without loss to the environment, both for ecological and financial reasons. Liner systems vary greatly from naturally occurring clays or soils modified with bentonite to layers of synthetic materials (geomembranes) such as PVC, HDPE, LDPE, or LLDPE. Geomembranes are available in various thicknesses, surface roughness, etc.

The tendency has been to become more and more sophisticated in the design of lining systems. These can include multiple layers and leachate collection systems. There is a danger in “overkill” in such designs in that heap-leaching technology ceases to remain cost effective for low-grade ores.

3. *Reusable Pad.* The third type of pad is the reusable pad in which ore is leached, rinsed and then removed. One of the first pads of this kind was installed at Round Mountain, Nevada (McQuiston and Shoemaker, 1980), where the same method is still in use today. Another version with a sophisticated feed arrangement was installed at Goldfield’s Ortiz Project in New Mexico (Hickson, 1981) and used

> effectively until the mine ceased production. The stacking system was then purchased and shipped to Bolivia for a second lease on life at the Kori Kollo mine before the main leach plant was constructed in 1992.
>
> Reusable pads are generally made of multiple layers of asphalt and rubberized asphalt placed over a structural-bearing layer. The asphalt, as well as providing low permeability layers, provides good wear resistance.
>
> Reusable pads are also common in the copper industry, particularly in Chile where the recently developed projects of El Abra, Radomiro Tomic and Cerro Colorado all use reusable pads.

Finally, in the development of a leach pad, it is important to provide internal berms and recovery piping to maintain a low static head on the liner but also for more versatility in solution management. Careful design evaluations must be done to keep these systems simple as well as effective.

ORE PREPARATION

Crushing requirements for heap leachable ores depend on ore characteristics and crush sizes varying from single-stage crushing (±150 mm) to four-stage crushing (±6 mm). There appears to be a major difference in copper ores occurring in North America and those in South America and Australia. In North America, the copper minerals tend to occur on fractures and are leachable at relatively coarse sizes, while the South American and Australian ores tend to be more highly silicified and disseminated requiring finer crushing for heap leaching (Keane, 1998). Generally, the only difference in crushing between copper and gold industries is that copper operations tend to be significantly larger.

Other types of ore preparation do vary greatly however. For many gold ores, agglomeration is essential to promote good permeability in the heap. The U.S. Bureau of Mines developed agglomeration as a technique for gold ores in the seventies (Heinen, et al., 1979). It entails adding a binder and moisture to the ore and mechanically tumbling the wetted mixture so that fine particles and clay adhere to the coarser particles or onto each other (as in the case of agglomerated tailings). Initial agglomeration work used 5 to 10 lbs. of Portland cement per dry ton of feed but, subsequently, other binders were introduced such as lime, fly ash and various long chain polymers. Lime and cement are the cheapest and preferred agglomeration aides; however, they also modify the pH of the ore. Generally, the amount of cement or lime used provides a pH in the preferred operating range 10.5–11.5. Occasionally the optimum amount of lime or cement for agglomeration produces too high a pH for optimum leaching and then the ability to separate agglomeration from pH control by the use of polymers becomes more desirable. Cyanide can also be added during the agglomeration step and this enhances the leach rate as gold dissolution starts immediately and does not have to wait for the application of spray solution (McClelland and Eisele, 1982). Care must be taken, however, because if the ore is acidic or needs to be buffered by lime or cement prior to cyanide addition, cyanide addition to the agglomerate step will increase cyanide consumption. Agglomeration is performed using pugmills, drums or a series of transfer points on conveyor belts.

In copper, natural agglomerating aids such as lime or cement cannot be used because leaching is done in an acidic environment and this has eliminated the use of some clayey copper ores from heap leaching. Recently, however, long chain polymer binders for use in acid environments have been developed (Scheffel, 1998). Other preparation techniques for copper heap leaching that do not have a parallel in gold leaching are the use of acid or ferric cures. Oxide ore can be acid cured prior to leaching by agglomerating with concentrated sulfuric acid. This increases recovery rate and reduces overall acid consumption. The ferric cure (curing with strong ferric sulphate solution), which was first used commercially by Inspiration Copper at Miami, Arizona (now Cyprus Miami) in 1980 (Bilson, 1989), has a similar effect on ore containing leachable sulfides. Agglomeration of copper ores with acid has the added advantage of limiting segregation of fine and coarser particles during placement.

HEAP CONSTRUCTION

Generally, once ore is on a conveyor, it is most cost effective to conveyor stack the ore on the heap. The geometry of the heap as dictated by site-specific conditions may not allow this, however, and then the material is transferred to trucks for truck dumping. There are numerous ways of building heaps with truck or conveyor stacking, but the main aim is to develop a construction method that minimizes cost and minimizes the impact of the building method on the permeability of the heap. The finer the material or the more essential agglomeration is to an ore, the more important it is to stack it properly. Compaction can significantly reduce the permeability of materials. Lift heights for crushed material tend to vary significantly from 2 to 10 meters in thickness. The thinner the lift the quicker the extraction rate but the more rapidly it is necessary to advance across the pad surface. The optimum heap height is normally a balance between technical and economic considerations. It is also imperative to rip the leached surface between lifts on permanent pads to enhance percolation through the boundary layers. Several copper operations in Chile add intermediate liners and collection systems between lifts because of permeability and pregnant liquor grade considerations.

A final point in heap construction is that when a project is first started up it is at its most vulnerable with respect to operating problems and meteoritic events. The thickness of the first lift tends to be irregular as the underlying ground contours are smoothed out for subsequent lifts. Also on a local level, the changes in pad slope may channel solution flows through areas where the conveyor or trucks are stacking fresh ore. Activities on the heaps are also closest to the liner, which makes it more vulnerable to damage. Finally, with large amounts of empty pad areas and very little material to act as a “sponge,” the operation is particularly vulnerable to heavy rainstorms, and solution management is critical.

LEACHATE APPLICATION AND BEHAVIOR

Solution application rates in copper leaching are extremely variable with solution application rates as low as 0.001 gpm per ft^2 to as high as 0.08 gpm per ft^2 in commercial applications in Arizona. The low end is constrained by the need to contact fresh leach solutions with the ore and provide an excess of solution to get a solution outflow from the heap. The upper end is often constrained by the permeability of the ore, but if that is not the rate determining factor, then pump capacity and/or the process plant capacity become the limiting factors. In gold leaching, a target solution application rate of between 0.002–0.005 gpm per ft^2 is usually used.

Sprinklers, wobblers, drip feeders on surface, or buried, are the most common ways to provide a uniform application of solution to the heaps. The selection of the type of system depends on the ore characteristics and the location of the project. For example, under freezing conditions or areas of high evaporation, buried drip systems are often used. In high rainfall areas, high evaporation sprinklers can be used to help manage the water balance.

As the leach solution percolates down through the heap, ideally all the particle surfaces become wetted and leaching of the metal values take place. In a typical gold ore heap leach with a head grade of 0.04 oz Au per ton of ore and a 75% recovery, the cyanide solution dissolves just under 1 part per million (ppm) gold. In most ores, there are other trace metals that will also dissolve but generally cyanide is a metal specific lixiviant that dissolves the precious material values and not much of the rest of the rock.

The situation in copper leaching is entirely different. Acid is a nonspecific lixiviant and copper values to be leached are in the thousands of ppm. As a result, there is significant mass transfer between the ore and the leach solution and recirculating leach solutions can contain over 250 gpl total dissolved solids. This increases both the density and viscosity of the leach liquor and can influence leach performance adversely, a condition which is difficult to duplicate in laboratory testwork. Also, because there is significant dissolution of rock in copper leaching, the particle size distribution in the heap leach tends to change with time leading to a substantial gain in the fine fractions. This can effect permeability and shear strength and can also expose new surfaces for leaching in ore where copper is generally found on fractures.

Scaling, both in the solution distribution systems and the pads, can create solution flow problems in both gold and copper leaching. In gold heap leaching, the addition of a suitable antiscalant is used to control scale build up.

LEACHATE COLLECTION AND PLANT DESIGN CONSIDERATIONS

As mentioned previously, the ability to separate solutions from different sections of the pad has significant advantage in managing the heap. “Preg building,” when intermediate solutions are applied to the heap several times before being fed to the recovery plant, is common practice in both gold and copper.

Single or multiple collection ponds in, or external to, the heap can be used. One often-overlooked aspect of solution collection design involves solution clarity. In gold heap leaching where the gold is recovered by carbon adsorption, solution clarity is not particularly important. In both zinc precipitation recovery circuits for gold and solvent extraction plants for copper, however, pregnant solution turbidity can create major problems. Generally, heap leach solutions flowing into a pregnant pond are clear because the solution has percolated through the heap, which acts as a large sand filler. However, during rainstorms, fines often wash off the side slopes of the heaps. If these fines are not settled out before the process plant, filters can blind rapidly or serious “crud” problems can be created in the solvent-extraction settlers.

HDPE is used extensively for piping in both types of plants, but in gold plants, mild steel can be used for tankage, piping and valves. In copper extraction plants, because of the acid environment, stainless steel (normally 316L) has to be used on high pressure piping in leach systems tankage, etc. This tends to make the plant construction more expensive. In recent years, the use of HDPE lining and polymer concrete has replaced stainless steel in many applications. Fiberglass is also used extensively for tankage, etc.

Brass is attacked by both cyanide and acidic solution and should be avoided as should any galvanized materials or reducing metals (brass, copper, zinc or aluminum). These metals in contact with acid solution containing arsenic can generate arsine gas which is significantly more toxic than hydrogen cyanide (Briggs, 1995).

ENVIRONMENTAL CONTROLS

Synthetic liner materials are resistant over a large pH range so that the same liner materials can be used for copper and gold heap leach pad construction. Clay materials can be more sensitive to low pH than to high pH solutions. Site-specific testing should be done to evaluate possible changes in permeability of clays when subjected to the specific leach solutions.

Most metals are soluble at low pH. The result is that acid leaching for copper is much more aggressive in terms of metal dissolution than alkaline cyanide leaching for precious metals. The higher pH cyanide leaching results in more selective leaching of metals. Table 1 shows some typical concentrations for selected constituents in raffinate at a copper leach facility and gold heap effluent at a heap leach facility near closure. The metal concentrations are much higher in the copper raffinate than in the gold heap leach effluent.

The relative aggressiveness of the leachate affects the design of environmental controls and the remaining environmental impacts from heap leach facilities. Cyanide is a very toxic substance, and its release into a stream will have immediate impacts. However, cyanide also dissipates relatively quickly, and the remaining metals may be less of a concern. The higher metal concentrations in copper leach solution could have a longer lasting impact on the environment if released.

The relative aggressiveness of the leachate also plays a dominant role in the closure of copper and gold heap leach facilities. It has been shown that cyanide concentrations in heap leach effluent can be reduced through rinsing or the use of cyanide reducing bacteria, often to levels low enough for discharge (Smith and Mudder, 1991). The same is not true in copper leach facilities. Catalan, et al. (1998) reports on the rinsing of leached oxide copper ore with water, lime solution and soda ash. They concluded that "long-term treatment of the heap seepage seems inevitable."

High metal concentrations in the effluent remain an ongoing concern at copper heap leach facilities. The project site climate is therefore the ultimate control in closing copper heap leach facilities. In extremely dry climates, such as the Chilean desert, very little to no effluent is expected, while in the tropics ongoing treatment of seepage will be inevitable.

FUTURE ISSUES

In this paper, the most commonly used heap leaching scenarios have been discussed, but, as time goes by, more and more variations of the same basic themes are being developed to leach specific ores. Other site-specific issues may also result in variations of the same theme.

In gold, agglomeration with pulp allows the heap leaching of a high-grade portion of the ore to achieve mill-type recoveries without the expense of leach tanks and tailings disposal facilities (McGregor and McClelland, 1989). This technology has successfully been applied commercially by Viceroy at their Castle Mountain mine (Zaebst, 1994).

Gold ores made refractory by encapsulation in sulfides, particularly pyrite and arsenopyrite, have been made accessible to leaching by using bacteria to oxidize some or all of the sulfides and exposing the precious metal values. This technology, while it has been used successfully in tank leaching of high-grade ores (Dew, et al., 1997), has been adopted by Newmont (Shutey-McCann, et al., 1997) for the treatment of low-grade ores by heap leaching. If this technology is successful, it will turn many millions of tons of material, now classified as waste, into ore. Pilot tests are currently underway at Kori Kollo in Bolivia on bacterial oxidation of low-grade sulfide ores at high altitudes.

TABLE 1 **Typical copper leach raffinate and gold heap leach effluent characteristics**

Analyte	Raffinate (mg/L filtered)	Gold Heap Leach Effluent (mg/L)
pH	1.7	9.4
Acidity	65,100	
Alkalinity	<3.4	143
Chloride	52	
Fluoride	100	
Nitrate (N)	17	55.6
Nitrite (N)	<10	12.2
Sulfate	46,000	1,850
Calcium	550	
Magnesium	3,300	
Potassium	<2.0	
Sodium	11	
Aluminum	4,500	
Antimony	<0.10	
Arsenic	0.25	0.45
Barium	<5.0	
Beryllium	1.2	
Boron	<50	
Cadmium	1.7	
Chromium	0.32	
Cobalt	20	
Copper	87	10.3
Iron	1,300	
Lithium	1,000	
Lead	1.4	
Manganese	750	
Mercury	<0.0002	0.164
Molybdenum	<0.50	
Nickel	7.6	
Selenium	0.23	0.41
Silver	0.037	0.31
Strontium	<0.50	
Thallium	<5.0	
Tin	<0.50	
Vanadium	<0.050	
Zinc	110	
TDS	88,500	

In copper, some major heap leach operations have been built in the last few years, such as El Abra in Chile which leaches oxide copper ores (Hickson, 1996) and Quebrada Blanca which leaches sulfide copper ores (Schnell, 1997). Several operations have buried pipes under their heaps and injected air to promote bacterial oxidation, which requires large quantities of oxygen (Holle, 1996; Schnell, 1997). This promotes more rapid bacterial oxidation.

A major technology thrust in the copper industry is to extend leachability of sulfide copper to chalcopyrite, and several research institutions have developed promising technology to accomplish this. Both BHP (Neira, et al., 1997) and Mintec (Vancas and Cornejo, 1995) have developed small-scale pilot plants to test ferric generators where the catalyzing reaction of converting ferrous to ferric iron by bacteria is done in a separate unit process outside the heap where conditions can be optimized for bacterial activity. Tests using these ferric generators have demonstrated some spectacular results.

In Giralambone in Australia, where there is a mixture of oxide and leachable sulfide ores, it has been found that a recovery rate of sulfide ores is dramatically lower if the ores are stacked in the winter as opposed to the summer months. This illustrates the sensitivity of bacterial activity to temperature (Holle, 1996). Since the introduction of forced aeration, however, this difference has been insignificant.

Copper and gold augment leach technology aims to recover both gold and copper values from ore where copper values are too high for conventional gold heap leaching to be economic (Younts, 1995). Still other operators have developed their own technology for treating mixed ores. At Mt. Leyshon in Australia, copper and gold were leached sequentially in a double heap leach system (Ellis, 1994).

Heap leaching has come a long way since the 1970s and is constantly evolving. On the positive side, more and more applications are being developed to exploit reserves or portions of reserves that are subeconomic by conventional milling techniques. On the negative side, liner and other environmental control systems are becoming increasingly sophisticated and expensive. These systems are often justified because their inclusion results in a lower risk of failure. It is important to understand, however, that the cost of failure is not only a function of the environmental consequences, which in many cases can be minor to negligible. The true cost of failure is extremely high in financial, social, political and public relations terms for the company unfortunate enough to suffer a failure. Including carefully designed elements to reduce the chances of a failure to a minimum is warranted as long as the overall costs are contained.

REFERENCES

Bilson, E.A. 1989. “Hydrometallurgy at Cyprus Leaching Operations—Review,” Arizona Section SME Section, Hydrometallurgy Division, May 20.

Briggs, A. 1996. “Practical Implementation of the Bacterial Oxidation Process for Refractory Gold Ores,” Australian Mineral Foundation/TMS Short Course on Bioleaching, Las Vegas, Nev., February 11–12.

Catalan, L.J.J., M.G. Li, and G. Comeau. 1998. “Rinsing Copper Oxide Heaps Leached with Sulfuric Acid: A Pilot Study,” Tailings and Mine Waste ’98, A.A. Balkema, Rotterdam, pp. 881–893.

Dew, D.W., E.N. Lawson, and J.L. Broadhurst. 1997. "The BIOX® Process for Biooxidation of Gold-Bearing Ores or Concentrates," In: *Biomining: Theory, Microbes and Industrial Processes,* D.E. Rawlings (Ed.), Springer-Verlag and Landes Bioscience Publishers, Berlin, Germany, pp. 45–80.

Ellis, S.J. 1994. "Bacterial Copper Heap Leach Followed by Heap Leach Recovery of Gold at Mt. Leyshon Gold Mine," Biomine '94, Perth, Western Australia, September 19–20.

Heinen, H.J., D.G. Peterson, and R.E. Lindstrom. 1978. "Processing Gold Ores Using Heap Leach—Carbon Adsorption Methods," U.S. Bureau of Mines, IC 8770.

Heinen, H.J., G.E. McClelland, and R.E. Lindstrom. 1979. "Enhancing Percolation Rates in Heap Leaching of Gold-Silver Ores," U.S. Bureau of Mines, RI 8388.

Hickson, R.J. 1981. "Heap Leaching Practices at Ortiz Gold Mine, Santa Fe County, New Mexico," SME–AIME Fall Meeting, Denver, Colo., November 18–20.

Hickson, R.J. 1996. "El Abra, World's Largest SX/EW Mine on Track to Join Copper Mining Elite in 1997," SME Annual Meeting, Phoenix, Ariz., March 11–14.

Hiskey, B. (Editor). 1983. "Heap and Dump Leaching Practice," SME Fall Meeting, Salt Lake City, Utah, October 19–21.

Holle, H. 1996. "Process Innovations at Girilambone Copper Company's Heap Leach/SX/EW Operation," Randol 2nd Annual Copper Hydromet Roundtable, Vancouver, B.C., November 12–15.

Keane, J. 1998. Private Communication, President, KD Engineering, Tucson, Ariz.

McClelland, G.E., and J.A. Eisele. 1982. "Improvement in Heap Leaching to Recover Silver and Gold from Low-Grade Resources," U.S. Bureau of Mine Report of Investigations, RI 8612.

McGregor, R.J., and G.E. McClelland. 1989. "Agglomeration With Pulp," Randol Gold Conference, Sacramento, Calif.

McQuiston, F.W., Jr., and R.S. Shoemaker. 1975. "Gold and Silver Cyanidation Plant Practice," Vol. I. SME. Monograph.

McQuiston, F.W., Jr., and R.S. Shoemaker. 1980. "Gold and Silver Cyanidation Plant Practice," Vol. II. SME. Monograph.

Miller, A. 1967. "Heap Leaching Copper Ore at Ranchers Bluebird Mine, Miami, Arizona," SME Fall Meeting, Las Vegas, Nev., Sept. 6–8.

Neira, M.P., T.D. McWater, C. Flores, and R. Patel. 1997. "Bio-Reactor Development for Ferrous to Ferric Conversion," SME Annual Meeting, Denver, Colo., February 24–27.

Parkison, G.A. and R.B. Bhappu. 1995. "The Sequential Copper Analysis Method—Geological, Mineralogical, and Metallurgical Implications," SME Annual Meeting, Denver, Colo., March 6–9.

Scheffel, R. 1998. Private Communication, Consultant, Denver, Colo., October.

Schnell, H.A. 1997. "Bioleaching of Copper." In: *Biomining: Theory, Microbes and Industrial Processes,* D.E. Rawlings (Ed.), Springer-Verlag and Landes Bioscience Publishers, Berlin, Germany, pp. 21–43.

Shutey-McCann, M.L., F.P. Sawyer, T. Logan, A.J. Schindler, and R.M. Perry. 1997. "Operation of Newmont's Biooxidation Demonstration Facility," In: *Global Exploration of Heap Leachable Gold Deposits,* D.M. Hausen (Ed.), The Minerals, Metals and Materials Society, Warrendale, Penn., pp. 75–82.

Smith, A. and T. Mudder. 1991. *The Chemistry and Treatment of Cyanidation Wastes.* Mining Journal Books, London, England, 345 pp.

Vancas, M.F., and R. Cornejo. 1995. Minbac^{TM} Technology Bio-Oxidation Process for Copper, 1st Randol Copper Hydromet Roundtable, Vancouver, B.C., October 1–4.

Van Zyl, D.J.A., I.P.G. Hutchison, and J.E. Kiel (eds.). 1988. *Introduction to Evaluation, Design and Operation of Precious Metal Heap Leaching Projects,* SME, 372 pp.
Younts, W.E. 1995. "Commercial Applicability of Augment's Copper-Gold Cyanidation Technology," 1st Randol Copper Hydromet Roundtable, Vancouver, B.C., October 1–4.
Zaebst, Robert J. 1994. "Pulp Agglomeration at the Castle Mountain Mine," 100th Annual Northwest Mining Association Convention, Spokane, Wash., Nov. 29–Dec. 2.

SECTION 3

Theory and Practice of Tankhouse Operations

Co-Chairs

R. Pariani	**M. Kuhn**
Wenmec Systems, Inc.	**Western States Engineering**

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Principles and Practical Considerations of Copper Electrorefining and Electrowinning

J. Brent Hiskey[*]

Electrolytic processes have been of critical importance in the recovery and refining of copper for over a century. At present there are approximately 92 electrolytic copper refineries in the world producing about 11 million tonnes per year of high quality cathode. Copper electrowinning continues to grow with the expansion of leaching activities. Electrowon cathode quality has improved dramatically in recent years and in some instances exceeds that produced by electrorefining. In the United States approximately 30% of all copper production is by electrowinning. Fundamental considerations are discussed in order to define the framework for copper electrorefining and electrowinning. Recent technological improvements are reviewed for each of these processes.

INTRODUCTION

Copper has been important to society for millennia and is currently a critical material in advanced communications and microelectronics. At the beginning of the 19th century, the significance of copper took on new meaning. Practical and theoretical work involving the nature and application of electricity were progressing at fast pace. Many pioneering efforts contributed to this field; however, the contributions of Michael Faraday (~1830) are especially notable. He provided the foundations in electromagnetic theory which led to the “dynamo” and the large-scale generation of electricity. A continuous electrical generator was first invented by Gramme in the late 1860s. This invention established a marriage between electrical energy and copper refining (winning) that remains in force today. High-purity copper is an essential part of the production, transmission, and utilization of electricity. Coupled with this is the need for large quantities of electrical energy to produce bulk amounts of refined copper. In other words, the two industries are dependent on each other and in fact have experienced parallel growth.

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[*] Department of Materials Science and Engineering, University of Arizona, Tucson, Ariz.

Elkington is credited with the first patent (1865) for a process to refine crude copper. The first electrolytic copper refinery based on his discovery was erected near Pembrey, Wales, in 1869.[1] A copper refinery following this example was constructed in Hamburg, Germany, by Norddeutsche Affinerie in the late 1870s. The first electrolytic copper refinery in the United States was located in Newark, New Jersey. This plant was built in 1883 by the Balback Smelting and Refining Co. Near the start of the 20th century there were ten electrolytic copper refineries in the U.S. producing approximately 250,000 mt per year. There are currently 8 refineries operating in the U.S. with an overall capacity of nearly 2×10^6 mt per year. Worldwide there are an estimated 92 electrolytic copper refineries producing about 11×10^6 mt per year.[2]

Copper electrowinning (EW) using inert lead anodes was instituted early on as a method of purifying electrorefining (ER) electrolytes. These so-called liberator cells remain the main form of electrolyte purification in many plants today. The Carmichael Process (1903) was one of the first attempts to recovery copper from a leach liquor.[3] However, the first large-scale copper electrowinning from leach solution was the New Cornelia Copper Co. complex at Ajo, Arizona. This plant was completed on May 1, 1917, and employed electrowinning in conjunction with vat leaching. Pregnant leach solution (PLS) containing 39 g l^{-1} Cu and 22 g l^{-1} H_2SO_4 and impurities (Fe, Al, Mg, Ca, Mn, and Cl) was introduced directly into the electrolytic circuit. This EW operation was followed shortly by vat leaching–EW plants constructed by Inspiration Consolidated Copper Company at Miami, Arizona, and Chile Copper Company at Chuquicamata, Chile.

Modern copper electrowinning cannot be discussed without acknowledging the contributions of solvent extraction. Solvent extraction can produce copper electrolytes essentially free of unwanted contaminants. In addition, solvent extraction is an interface technology that links relatively rigid electrowinning conditions with a variety of leaching situations. Since the first stand-alone leaching/SX/EW plant located at the Bluebird Mine, there has been an amazing growth in production of copper by electrowinning. U.S. electrowon copper production during the last three decades is shown in Figure 1.

Electrowon copper production in the U.S. during 1997 was 579,000 mt of a total production of 2,060,000 mt.[4]

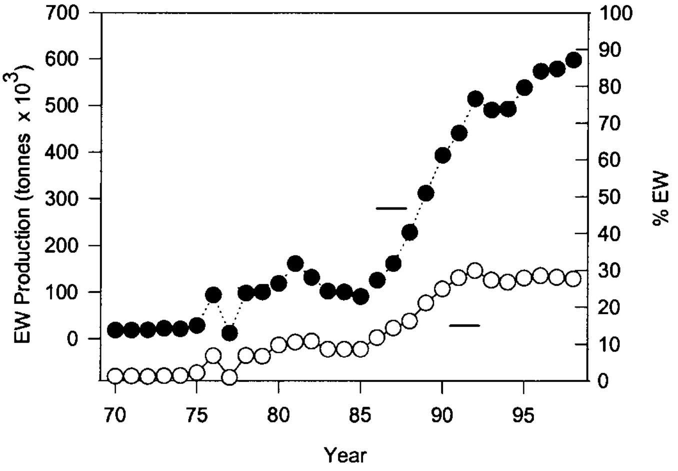

FIGURE 1 Electrowon copper production and percent of total copper production since 1970

The objective of this paper is to examine the fundamental principles of electrochemical operations used for the recovery of copper and to review recent and emerging technologies in copper electrorefining and electrowinning.

FUNDAMENTAL PRINCIPLES

Electrorefining

The basic principles and theories fundamental to copper refining practice were well known 100 years ago. Key elements of the technology such as operating conditions, tanks and cell house configuration, and electrode considerations remain about the same.

The overall cell reaction for the electrorefining of copper is as follows:

$$\mathrm{Cu}_{(\text{impure})} = \mathrm{Cu}_{(\text{pure})} \tag{1}$$

Impure copper dissolves anodically releasing copper ions in the vicinity of the anode. The theoretical cell potential is 0 V; however, to overcome certain kinetic constraints and to sustain the refining reaction, a cell potential must be applied. This is normally in the range of 0.3 V which is necessary to overcome the overvoltages associated with each electrode and the solution resistance. The overvoltage on an individual electrode represents the total of all types of polarization, usually one type predominates. The sum for charge transfer (activation), mass transfer (concentration), and resistance polarization is shown below

$$\eta_T = \eta_A + \eta_C + \eta_\Omega \tag{2}$$

where η_A, η_C, and η_Ω are overvoltages associated with activation polarization, concentration polarization, and *IR* contribution, respectively. In copper ER, approximately 80% of the polarization is attributed to the IR drop in the electrolyte. An additional 5% is related to the various electrical contacts.[5]

Depending on pH, anions present, and certain kinetic factors, copper dissolution takes place in two stages:

$$\mathrm{Cu}_{(\text{impure})} \rightarrow \mathrm{Cu}^+ + e^- \quad \textit{fast} \tag{3}$$

$$\mathrm{Cu}^+ \rightarrow \mathrm{Cu}^{2+} + e^- \quad \textit{slow} \tag{4}$$

Mattsson and Bockris found that for pure copper at high current density, the conversion from cuprous ion to cupric ion is the rate-determining step.[6] The cupric ions produced at the anode are transported to the cathode surface primarily by convective forces in the bulk electrolyte and finally diffusion through the boundary film. The cathodic reduction of copper occurs by the following reaction

$$\mathrm{Cu}^{2+} + 2e^- \rightarrow \mathrm{Cu}^\circ \tag{5}$$

As with the anodic processes, the exact mechanistic details of this reaction are very complex. The Tafel slopes for anodic dissolution and cathodic deposition of copper were found to be unsymmetric by Mattsson and Bockris.[6]

anodic dissolution	*b = 30–40 mV per decade*
cathodic deposition	*b = 120 mV per decade*

Important aspects of the anodic processes associated with copper electrorefining will be discussed in terms of reactions, behavior of impurities constituents, and anode passivation.

TABLE 1 Possible species present in anode slimes

Group VA (As, Sb, Bi)	Group VIA (O, S, Se, Te)	Group IB (Cu, Ag, Au)	Group IB (Sn, Pb)	Other Elements (Fe, Ni, Zn)
As_2O_3	Cu° Ag°	Se° Au°	SnO_2	Fe_2O_3 Fe_3O_4
Sb_2O_3	Cu_2Se Ag_2Se	Cu_2S	$Sn(OH)_2 \cdot SO_4$	NiO
Bi_2O_3	$CuAgSe$	$(Ag,Cu)_2Se$	$PbSO_4$	ZnO
$SbAsO_4$	$AgTe$ $(Ag,Au)Te_2$	$Cu_2(Se,Te)$	$x PbO \cdot As_2O_5$	$CuFeO_2$ $NiFe_2O_4$
$BiAsO_4$	CuO Cu_2O	$CuSO_4 \cdot 5H_2O$	$y PbO \cdot Sb_2O_5$	$3Cu_2O \cdot 4NiO \cdot Sb_2O_5$
	$Cu_3(SbO_4)_2$	$Cu_3(AsO_4)_2$	$Pb(Sb,As)O_4$	$(Cu,Ni)SO_4 \cdot 5H_2O$
	$CuSeO_3 \cdot 5H_2O$			SiO_2

Anode Reactions. Possible anode reaction during copper electrorefining are summarized as follows:

- Copper dissolution

$$Cu^{\circ} \rightarrow Cu^{2+} + 2e^{-} \tag{6}$$

- Oxygen evolution

$$H_2O \rightarrow 2H^{+} + \tfrac{1}{2}O_2 + 2e^{-} \tag{7}$$

- Formation of metal oxides

$$Cu^{\circ} + H_2O \rightarrow CuO + 2H^{+} + 2e^{-} \tag{8}$$

- Oxidation of ions in solution

$$AsO_2^{-} + 2H_2O \rightarrow AsO_4^{3-} + 4H^{+} + 2e^{-} \tag{9}$$

- Oxidation of addition agents

$$2CS(NH_2)_2 \rightarrow [(NH_2)_2C{-}S{-}S{-}C(NH_2)_2]^{2+} + 2e^{-} \tag{10}$$

Reaction (6) is the only reaction desired in electrorefining, while reactions (7) and (8) are side processes which need to be avoided. The reactions encountered will depend on the chemistry of the anode and electrolyte.

Behavior of Impurities. When copper dissolves, impurities contained in the anode are continuously released from the copper matrix and grain boundary structures. These impurities are released either without chemical change, as in the case of $Cu_{2-x}Se$ inclusions, or possible by chemical and electrochemical reactions, as in the case of As_2O_3 and As. Metals more reactive than copper, such as nickel and iron, report as dissolved species. Insoluble phases forming the anode slimes will remain attached to the electrode surface or fall to the bottom of the refining cell as a sludge (anode mud). Possible species present in anode slimes are summarized in Table 1.

Anode Passivation. Anode passivation, frequently encountered in commercial copper electrorefining operations, causes a reduction in refinery output and an increase in power consumption. In fundamental terms, passivation is a situation when the rate of dissolution of the anode becomes infinitesimally small, or, in other words, the overpotential required to maintain a constant rate of dissolution increases dramatically. The passivation of copper anodes results from the formation of inhibiting surface films. The most significant anode films associated with passivation are the slimes, copper sulfate, and copper oxide. Slimes inhibit diffusion and cause copper sulfate saturation and precipitation at the electrode surface and within the pore spaces of the slimes layer. Ultimate passivation occurs when the anodic potential increases to a point where copper oxide is stabilized.

The overall passivation response are controlled by a number of parameters including (1) anode composition, (2) electrolyte composition, (3) current density, (4) electrolyte temperature, and (5) additives.

Electrowinning

The overall cell reaction for the electrowinning of copper from a sulfate electrolyte is as follows:

$$CuSO_4 + H_2O \rightarrow Cu + H_2SO_4 + \tfrac{1}{2}O_2 \qquad (11)$$

In the process, copper metal is deposited, sulfuric acid is generated, and oxygen evolved. It is convenient to discuss the electrowinning chemistry in terms of the respective cathodic and anodic processes.

The Cathodic Process. The cathodic reduction of copper occurs by the following reaction:

$$Cu^{2+} + 2e^- \rightarrow Cu^o \qquad (5)$$

At 65°C, the standard electrode potential for this reaction is 0.34 V. By applying the Nernst equation to the equilibrium shown in reaction (5) the following expression is produced

$$E = 0.34 + 0.034 \log a_{Cu^{2+}} \qquad (12)$$

where E is the equilibrium electrode potential and $a_{Cu^{2+}}$ is the activity of the cupric ions in the electrolyte. The equilibrium electrode potential has been measured at about 0.25 V (SHE) for pure copper in an electrolyte containing 40 g/l Cu^{2+} and 160 g/l H_2SO_4 at 65°C.

In concentrated electrolytes, the maximum limiting current density (i_l) is determined by diffusion through the boundary film and is usually expressed as

$$i_l = zFD\frac{C_B}{\delta_N} \qquad (13)$$

where z is the number of electrons transferred, F is the Faraday constant, D is the diffusion coefficient, C_B is the bulk solution concentration, and δ_N is the boundary layer thickness. The value of the limiting current density is approximately 450 A m^{-2} for a solution containing 45 g/l Cu^{2+} and under static conditions (i.e., $\delta_N = 0.03$ cm). Most copper electrorefining and electrowinning plants operate at a current density much lower than the limiting value, since deposit structure becomes nodular, porous and even powdery at high current density. Poor cathode quality results with these types of deposit morphologies.

Figure 2 shows the idealized polarization curve for the deposition of copper. Also included on this figure is the polarization curve for the reduction of ferric ion to ferrous in solution at ~10 g/l Fe^{3+} (i.e., ~60 A m^{-2}). Copper deposition can only proceed when cathodic currents are greater than i_l (Fe^{3+}). At current densities lower than i_l (Fe^{3+}) copper cathodes will redissolve. It is desirable to maintain the "iron current" as low a fraction of the total current density as possible. At first glance this can be achieved by two

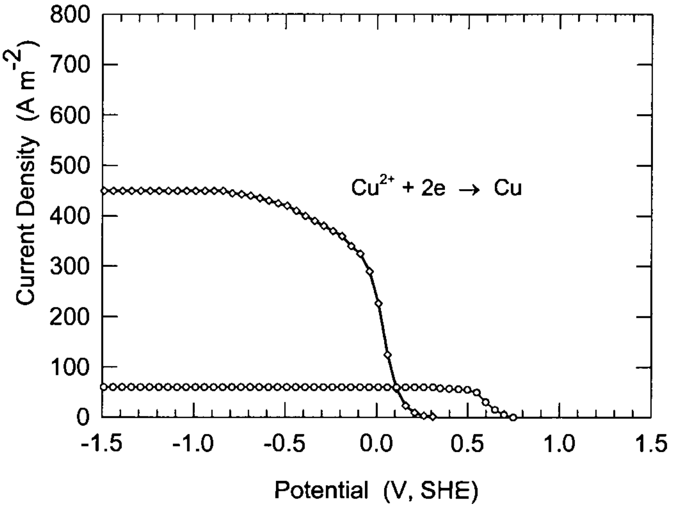

FIGURE 2 Hypothetical polarization curves for copper and ferric ion reduction

approaches: (1) by increasing the total current density and (2) by decreasing the concentrations of Fe^{3+} and thus lowering i_l (Fe^{3+}).

The first approach has certain limitations. Operating at extremely high current densities can, as stated above, cause poor deposit quality. In addition, higher current densities cause a proportional increase in i_l (Fe^{3+}) due to increased evolution of oxygen (agitation). The only sure way of improving copper electrowinning current efficiencies is by keeping the iron concentration in the electrolyte as low as possible. Highly selective SX reagents, combined with good plant practice, have helped copper electrowinning achieve very high efficiencies.

The Anodic Process. Since the anode serves as the positive electrode which completes the electrolytic circuit, it must be formed from a material that will be a good conductor. Further it must be chemically inert towards the electrolyte. The fundamental anodic process occurring during the electrowinning of copper is the decomposition of water according to the following reaction:

$$H_2O \rightarrow 2H^+ + \tfrac{1}{2}O_2 + 2e^- \qquad (14)$$

At 65°C, the standard electrode potential for this reaction is 1.20 V. The Nernst expression for reaction (14) when oxygen is liberated at 1 atm is

$$E = 1.20 - 0.067\,pH \qquad (15)$$

and the equilibrium potential is about 1.16 V. The reversible cell potential for the overall copper electrowinning reaction is $E_{cell} = 0.91$ V. In practice, the actual operating voltage is about twice this value (i.e., ~2 V). This difference is attributed to the irreversibility of the system and is explained in terms of electrochemical overvoltages. In electrowinning, the anodic reaction (i.e., oxygen discharge) is highly irreversible. As a result of this irreversibility, high overvoltages are required for the liberation of oxygen from water. Lead, the most popular anode material for copper electrowinning from sulfate electrolytes, is characterized by one of the highest oxygen overpotentials known, values approaching as high as 1 V.

INNOVATIVE TRENDS IN ELECTROREFINING

General

The production of copper by electrorefining has improved substantially in the last 100 years. Improvements in current efficiency, materials handling, electrolyte purification, and automation and process control have led to considerable increases in production, particularly in the last several decades. One way of examining advances in productivity is to compare operating data for plants in terms of annual tonnes per cell. Figure 3 shows 1960 and 1995 data for several plants. These older refineries all show a marked increase in productivity from 1960 to 1995. The refinery built in 1994 at Miami, Arizona, and operated by Cyprus Mineral Co. has utilized a significant amount of new technology. This refinery achieves an impressive 330 to 380 annual tonnes per cell.[2]

Some of the recent developments that have led to increases in production and noticeable improvements in cathode quality have been reusable cathodes, electrolyte purification, monitoring of additives, period current reversal, and anode preparation machines.

Reusable Cathodes. The use of copper starter sheets dates back to the early periods of electrorefining. These thin sheets can create difficulties due to warping or bending which can result in short circuiting. Embossing, lapping, and mechanical pressing after 24 to 48 hours of deposition are ways of attempting to ensure straight cathodes. In recent years, permanent cathode blanks have eliminated the need for these steps. Copper is electrodeposited on a mother plate of titanium or stainless steel for a suitable time, typically 7 to 14 days. The copper is then removed usually by a mechanical stripping machine. Following a visual inspection, the starter sheets are returned to electrorefining cells.

Stainless steel technology has been growing in popularity compared to titanium because of the significantly lower initial capital expenditure. There are two different stainless steel technologies, the ISA Process[7] and the Kidd Process.[8] The major difference between the methods relates to the bottom of the cathode. In addition to side-edge strips, the ISA Process uses wax to prevent copper deposition and thus produces two sheets (one on each side of the stainless steel mother plates). The Kidd Process leaves the bottom exposed, which creates two connected sheets. The absence of wax or other plating inhibitors in the Kidd Process avoids bottom edge contamination of the cathode.

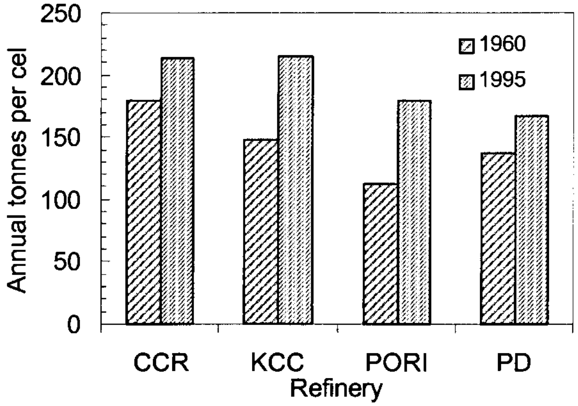

FIGURE 3 Copper electrorefining productivity data for 1960 and 1995

The major advantage of stainless steel technology over traditional starter sheets are the avoidance of starter sheet manufacturing and improved verticality, greater current densities, and purer cathodes. In a 1987 review of copper refining tank house data, Scholen found the ISA Process installed at only 2 operations.[9] In the 1991 review, the ISA Process was noted at 4 operations.[10] The 1995 Scholen and Davenport indicated that approximately one-fourth of the refineries surveyed had adopted stainless steel cathode technology, 5 using the ISA Process and 6 using the Kidd Process.[2]

Pariani listed the following advantages for the full deposit stripping system:[11]

- Completely eliminates the need for conventional starter sheet stripping and preparation plus associated handling
- Results in a labor reduction due to low frequency of electrical shorts normally attributed to bent sheets
- Facilitates a tighter electrode spacing
- Eliminates cathode loop corrosion and minimizes variations in cell electrolyte level
- Provides operating flexibility due to elimination of sheet production which may occur on weekends at some refineries

Electrolyte Purification. While various strategies to remove impurities from the electrolyte have been practiced for many decades, there has been some recent advances in this area. Traditionally, the copper industry depended on liberator cells followed by crystallization to control impurities in the main electrolyte circuit. In the liberator cells, the electrolyte is decopperized in a series of cascading electrowinning cells. Electrowinning removes copper, and the Group VB elements (As, Sb, and Bi). The first electrowinning stage usually deposits a fairly pure copper product. The second and sometimes third electrowinning stage removes Cu, As, Sb, and Bi. Following electrowinning, the solution is heated to evaporate water and thus crystallize nickel, cobalt, and iron as sulfate salts.

The removal of As, Sb, and Bi from bleed streams using solvent extraction and chelating resins has received considerable attention during the last twenty five years. Hoey et al. indicated that solvent extraction can be utilized for As removal with a final salable product of bicupric arsenate, which is a precursor for chrome copper arsenate, a lumber preservative.[7] Tallert and King described an electrolyte purification circuit installed at Asarco's Amarillo Copper Refinery.[12] This plant employees ion exchange technology and a novel eluant for stripping the ion exchange resin. A simplified flowsheet is shown in Figure 4. A strong chloride salt solution (5 M) with a small amount of HCl (0.5 M) serves at the eluant in this process. The elute is neutralized with caustic which precipitates oxides and oxychlorides of antimony and bismuth. These precipitates are filtered and removed for further treatment or disposal.

Dreisinger and Scholey have described an ion exchange procedure for removing antimony and bismuth using phosphonic type resins.[13] The overall efficiency of the process is improved by contacting the electrolyte with a packed bed of copper wire turnings. This step accomplishes reduction of certain species like Fe^{3+} which can reduce the loading capacity of the resin. In addition, this technique can be used to control the chloride concentration of the electrolyte according to the following reaction:

$$Cu^{2+} + Cu + 2Cl^{-} \rightarrow 2CuCl_{(s)} \downarrow \tag{16}$$

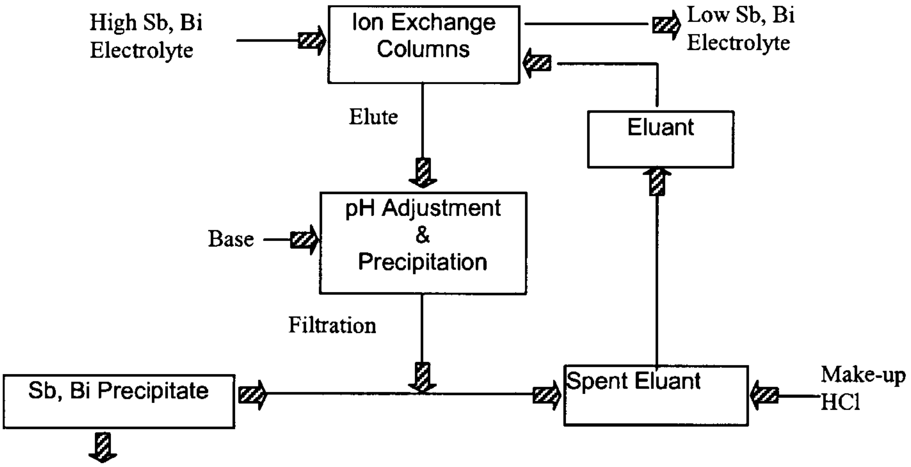

FIGURE 4 Simplified flowsheet for Asarco's Amarillo copper refinery electrolyte purification circuit (after Tallert and King)

Additive Monitoring. Cathode deposits obtained from acidified copper sulfate solutions without additives are often soft and coarsely crystalline or nodular. Under certain conditions, the deposit can be dendritic and even powdery. Such deposits are apt to entrain electrolyte and anode residues, resulting in cathodes of less than desirable purity. Glue, thiourea, and chloride are common additives used in modern tankhouses to produce smooth, dense, coherent deposits. These additives effect the electrocrystallization of copper in various ways: glue (protein colloid) functions as a leveling agent and thiourea and chloride ions act as grain refiners. Recent research has shown that chloride appears to activate the polarization caused by glue.[14] The need for these and possibly other additives (i.e., Avitone A, Tembind, Orzan, and Magnafloc) ensures the proper deposit structure and morphology. While additives have been used for many decades, the monitoring and automatic control of reagent addition has only recently been employed.

The two most notable systems are the Reatrol system[15] developed by Asarco for monitoring thiourea and the CollaMat system[16] developed by Norddeutsche Affinierie for glue. The Reatrol method employs differential pulse polarography to determine the active concentration of thiourea with a dropping Hg electrode. This technique is claimed to have a detection limit of 100 ppb. The use of Reatrol at the Asarco Amarillo refinery has resulted in a 5% increase in current efficiency, reduced shorts, and improved precious metal recovery because of less slimes occlusion.[12] A disadvantage of the Reatrol process is that it involves off-line instrumentation and analysis. CollaMat is a proprietary method that provides continuous on-line real-time analysis of glue in the electrolyte. Other techniques have been developed by CODELCO, Inco, Kennecott, and Noranda for measuring addition agents. However, there remains a need for further design and development of robust and sensitive sensors that can tolerate the frequently harsh conditions of copper refining.

Periodic Current Reversal. Application of period current reversal (PCR) in copper refining has been practiced for about three decades. Its significance is still disputed; however, it is generally conceded that it allows the use of higher current densities without anode passivation and significant deterioration of cathode quality. The PCR approach involves applying a forward current for a set length of time followed by a short current reversal (i.e., cathode momentarily becomes anode and vice versa). The reversal current period is typically 5% of the forward period (ratio $t_f/t_r = 20/1$), and occurs every 30 to 300 s.

Electrodes are temporarily depolarized or depassivated during the reversal period. The cathode undergoes anodic dissolution which occurs mainly at dendritic sites and at nodular protrusions. At the same time, the anode becomes cathodic and copper is depleted from the boundary layer region thus avoiding the formation of passivating surface films like $CuSO_4 \cdot 5H_2O$ and Cu_2O.

Current efficiency needs to be redefined for systems employing PCR.

$$CE_{PCR} = \frac{t_f - t_r}{t_f + t_r} \cdot CE_{MAX} \tag{17}$$

where CE_{PCR} and CE_{MAX} are the current efficiencies for PCR and maximum possible, respectively; and t_f and t_r are the forward and reverse periods, respectively. This results in higher energy costs which is a disadvantage of PCR refining. According to the survey by Schloen and Davenport,[2] PCR is in use at 11 refineries; however, the acceptance of this technology is not growing.

Anode Preparation Machines. Copper anodes produced pyrometallurgically can weigh between 150 and 400 kg and are typically about 1 m^2 and 5 cm thick. Anode casting is usually accomplished with moulds on a wheel. Alternately, continuous casting machines such as the Hazelett and Contilanod systems are used to manufacture anodes that are geometrically accurate and have repeatable lug shape. Adoption of continuous casting systems does not appear to be growing; however, the implementation of automatic anode weighing during casting and anode preparation machines does appear to be growing in popularity. In the 1995 review by Schloen and Davenport,[2] anode preparation machines were found in about half of the refineries surveyed.

Anode preparation steps that help ensure the physical quality of anodes complement efforts to achieve straight and uniform cathodes (i.e., stainless steel technology). Anode preparation systems perform several functions such as:

- Rejecting out-of-spec anodes
- Straightening the body and lugs in the vertical plane
- Aligning the lugs in the horizontal plane
- Milling the lug underside to achieve good electrical contact
- Providing uniform anode spacing

These operations are normally accomplished in a sequence of stations. The AISCO system is based on a carousel configuration that has the potential of incorporating 8 stations. Normally, the AISCO machine contains a station for each of the five functions listed above and a sixth for rejection anodes. Figure 5 shows a schematic of the AISCO anode preparation machine.

WENMEC manufactures an anode preparation machine that accomplishes similar operations in a linear configuration which requires more floor space. Bosscher and Berends indicate that production rates of 500 anode h^{-1} can be achieved with the AISCO system.[17] The WENMEC unit achieves comparable production rates.

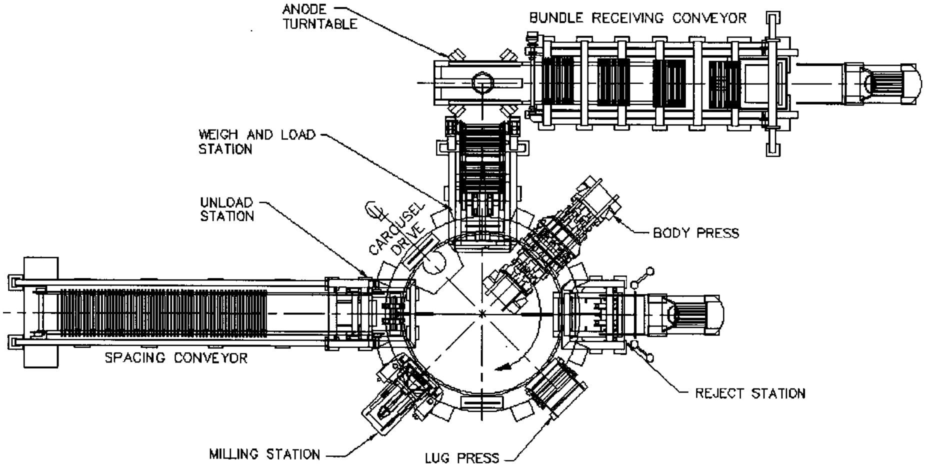

FIGURE 5 Schematic of AISCO's anode preparation system (courtesy of AISCO Systems Inc.)

Maintaining proper anode weight minimizes electrical shorts and reduces anode scrap. Controlling dimensional uniformity, alignment, and spacing ensures uniform dissolution and minimizes shorts. Other benefits of automated anode preparation machines are:

- Reduced manpower
- Higher current efficiency
- Improved safety

INNOVATIVE TRENDS IN ELECTROWINNING

General

In the early days of SX-EW there was some concern about the purity of copper cathodes produced by this technology and that EW cathodes would require further refining. Frequently there were problems with lead contamination as a result of corrosion and spalling of lead-antimony anodes, some contamination by entrained organic, and higher sulfur levels caused by entrapped electrolyte. Current practice has eliminated these problems and produces high quality copper cathodes that can compete for any market in the world. It is interesting to examine the improvements in EW cathode quality. The chemical analysis of electrowon copper from some early plants is compared with that produced from modern plants is shown in Table 2.

COMEX and LME certification of EW cathodes is being achieved by a number of plants around the world.

Several areas of innovation that have contributed to the success of copper electrowinning technology are discussed as follows. They include permanent cathode technology, anode materials development, electrolyte additives, electrolyte circulation, mist control, cell construction, and plant automation. Permanent cathode technology was discussed in the section of electrorefining and will only be briefly mentioned here.

TABLE 2 Analysis of electrowon copper cathodes

Electrowinning Plant	Impurities (ppm)						
	Pb	S	Fe	Mn	Co	Ni	Other
Bagdad (1971)	3	5–10	4			4	Ag 1, As 1, Bi 0.3, Sb 2, Sn 2, Te 2
Chambishi (1978)	5		2			1	Bi 1, As 0.2, Se 20
BHP San Manuel	<1	1–2	1–2			<1	
Girilambone	<0.1	2	1.6	<0.1	<0.1	<0.1	
PD, Morenci	1–2	6–8	0.6–1.6		0.1		
Zaldivar	2	4	<2	<1	<1	<1	
LME Grade A	5	15					<49

Permanent Cathode Technology. In addition to the five advantages previously cited for permanent cathode technology, the permanent cathode system eliminates the safety hazards associated with starter sheet preparation and materials handling. Jenkins and Eamon estimate that there is a 60% reduction in EW workforce with the use of permanent cathode technology.[18] A sampling of 12 of the newest EW facilities in the world indicates that all are using permanent cathode technology. Seventy-five percent have adopted the ISA Process, with the remaining using the Kidd Process.[19]

Anode Materials. Prengaman described the development of insoluble lead anodes for copper electrowinning.[20] Rolled and cast Pb-Ca-Sn anodes have replaced the early Pb-Sb anodes. It has been observed that Pb-Sr alloys exhibit similar electrochemical behavior as the Pb-Ca alloys. Cast anodes suffer from severe corrosion while rolled anodes are not handicapped by corrosion problems. In general, the rolled Pb-Ca-Sn anode offer greater design flexibility, better mechanical properties, corrosion resistance, and long service.

The composition of Pb-Ca-Sn anodes ranges from 1.00 to 1.75% Sn and 0.05 to 0.50% Ca. Some anodes in use contain small amounts of Al (0.0005 to 0.02%). The grain structure of the cold-rolled Pb-Ca-Sn alloy is very fine in comparison to as-cast alloys which contain very large dendritic grains. The cold-rolled anodes produces a very fine less adherent corrosion product than does a cast anode. These fine Pb particles can create a slimes problem in the cell, which may necessitate more frequent cell cleaning. However, superior dimensional stability of the roll anodes more than compensates for the problems associated with anode corrosion products.

Dimensionally stable anodes (DSA) widely used in the chlor-alkali industry have been suggested and tested for copper electrowinning. DSA are commonly fabricated from titanium coated with platinum or ruthenium oxide. As a result they are chemically very stable and do not introduce impurities into the cell. DSA have a pronounced depolarization effect on the oxygen discharge reaction. The reduction in anodic overpotential is shown in Figure 6.

At a current density of 300 A m^{-2} the anodic overpotential is reduced by 360 mV. This represents a potential reduction of about 20% in power cost. For a power cost of 5¢/kWh, the savings in power consumption would be about 2¢/kg of copper produced. In summary, the advantages of DSA are:

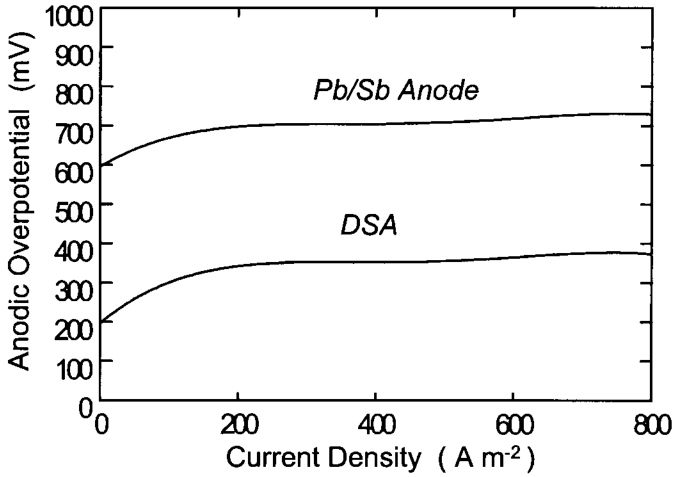

FIGURE 6 Anodic overpotentials of insoluble Pb/Sb anode and DSA in copper electrolytes

- Reduced power cost (lower anodic overpotential, fewer shorts)
- Reduced cell cleaning
- Purer cathodes

Dimensionally stable anodes are very expensive due to the precious metals used in manufacturing. DSA also require periodic recoating. Dimensionally stable electrodes have been developed using conductive electrocatalytic coatings on sintered ceramic materials. These electrodes exhibit excellent chemical and electrochemical stability and have considerable economic advantage.

Diamond Shamrock Technologies S.A. and Impianti Electrochimici S.p.A. have devoted considerable effort towards the development of new electrodes for electrochemical processes.[21–23] Some of these have been tested as the De Nora anode for copper electrowinning. Design aspects for these electrodes have incorporated composite materials and unconventional substrate arrangements. De Nora described lead anodes that exhibited extraordinarily low oxygen overvoltages which could provide considerable energy savings in electrowinning.[22] Figure 4 shows the reduction in anode potential for 107 g/l H_2SO_4, 400 A m^{-2} and 35–40°C. There is a noticeable drop in anodic potential especially in the first 8 h of polarization.

Modern copper electrowinning practice employs cold rolled anodes. The typical life of these anodes ranges from 5 to 10 years. These anodes have low cost, greater design flexibility, excellent mechanical properties, corrosion resistance, and long service.

Electrolyte Additives. As noted in the section on copper electrorefining, a wide range of colloidal or reducible substances has been used in electrolysis to improve the structure and quality of the deposit. Introducing colloidal addition agents leads to the production of smooth dense fine-grained deposits. Crystal size is decreased when these additives are adsorbed on the cathode surface. Copper electrorefining has employed numerous addition ages such as glue, thiourea, alkyl sulfonates, lignosulfonates, and other under trade names like Avitone, Casein, Lignone, and Orzan.[2] On the other hand,

TABLE 3 Inhibiting effects of Fe, Ni, Co on the corrosion of lead anodes (0.4% As and 10% Sb)

System	Corrosion Rate (mm y^{-1})
H_2SO_4	2.6
10 g l^{-1} Fe	0.9
8 g l^{-1} Ni	0.3
0.05 g l^{-1} Co	0.07

copper EW has avoided the used of colloidal additives because of the potential problems associated with the formation of crud in solvent extraction.

Many newer copper EW tankhouses are operating at current densities of over 300 A m^{-2} and there is always the push to raise current density as a means of increasing production without changing the footprint of the plant. Pushing the current density envelop has the potential of causing severe nodulation problems and producing poor quality cathodes. Common leveling agents such as glue have not been considered because of phase separation problems in SX. Tyroler et al. reported on the use of a high molecular weight guar gum derivative (Jaguar Plus) as a leveling agent for copper electrowinning plant at the Copper Cliff Copper Refinery.[24] Jenkins and Eamon recently described the use of a guar gum additive (Guartec) to control deposit morphology and quality while operating at current densities of 320 A m^{-2}.[18] This reagent has been successfully used for a number of years at San Manuel without any adverse impact on SX performance. In fact, it is reported that guar is currently used at over half of the EW tankhouses surveyed by Robinson, Davenport, and Jenkins.[19] For those plants using guar, the average dosage is 240 g t^{-1} of cathode.

It is been recognized for some time that small additions of cobalt to the copper electrolyte result in decreased corrosion of Pb anodes and a reduction in Pb contamination of the cathode. Anderson et al. examined the corrosion rate of 0.5% As and 10% Sb Pb anode in 180 g l^{-1} H_2SO_4 at 28°C using 258 A m^{-2}.[25] They found iron, nickel, and cobalt to have an inhibiting effect on anode corrosion rate. However, as shown in Table 3, relatively high levels of iron and nickel were required.

By comparison, cobalt is more effective even at extremely low concentrations. Cobalt concentrations of 100–120 ppm are very common for copper EW electrolytes. Additions of cobalt to the electrolyte is not the total answer for controlling Pb contamination of cathode copper. Cobalt does decrease the corrosion rate of Pb anodes; however, achieving very low Pb levels requires careful control of operating parameters and diligent maintenance of the system.

Gendron et al. have shown that additions of Co in copper electrowinning depolarizes the anode.[26] The oxygen discharge reaction is catalyzed according to the following proposed mechanism:

$$Co^{2+} = Co^{3+} + e \tag{18}$$

$$4Co^{3+} + 2H_2O = 4Co^{2+} + 4H^+ + O_2 \tag{19}$$

Reaction (18) has a standard potential of 1.84 V. The cobaltic ion oxidizes water by reaction (19) which is catalyzed by the presence of PbO_2. Figure 7 illustrates the dependence of anode potential as a function of cobalt concentration in the electrolyte. Shown are the steady state potential for 300 A m^{-2}. Increasing the cobalt concentration decreases the anodic potential by about 46 mV per decade.

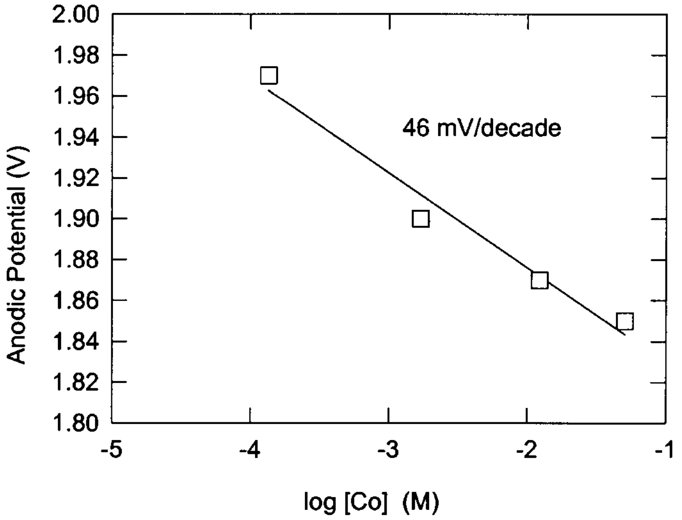

FIGURE 7 Reduction in anodic potential as a function of cobalt concentration

Electrolyte Circulation. Jenkin and Eamon discussed the operating performance of an electrolyte distribution system developed for the San Manuel EW tankhouse.[18] This system was found to be effective while operating at current densities of approximately 320 A m^{-2}. The electrolyte distribution manifold introduced electrolyte into the cell at a flow rate of 0.115 m hr^{-1} (0.047 gpm ft^{-2} of planting surface). The original design of the plant was for electrolyte flow rates of 0.081 m hr^{-1} (0.033 gpm ft^{-2}).

The manifold cell was shown to provide uniform temperatures across the cell and with depth in the cell. Average copper concentration was also extremely uniform throughout the manifold cell. The nonmanifold system exhibited noticeable variations in both temperature and copper concentration.

It was concluded that the electrolyte circulation system provides the following:

- High purity cathodes (< 1–2 ppm Pb) at 320 Am^{-2}
- Improved cathode appearance to an acceptable level with only minor modulation in the bottom ⅓ of the cathode
- Current efficiency equaling or exceeding nonmanifold standard current density cell

Approximately one half of the plants in the most recent survey indicate the use of electrolyte distribution manifolds.

Mist Control. Electrowinning of copper results in the liberation of oxygen bubbles at the anode surface. When these bubbles rupture at the electrolyte-air interface, an acid mist is released to the atmosphere. Tyroler and co-workers reported that the objective of mist control measures is to maintain the sulfuric acid concentration in the tankhouse below 1 mg m^{-3}.[24] A recent article by Davis and Hopkins discusses developments in electrometallurgical tankhouse environmental control.[27] Exposure limits for the TLV–TWA (threshold limit value – time weighted average) and TLV–STEL (threshold limit value – short term exposure limit) for H_2SO_4 were 1 and 3 mg m^{-3}, respectively. Factors that influence acid aerosol production rates in copper EW are site elevation, current density, barometric pressure, ambient temperature, anode conditions, electrolyte temperature, and electrolyte composition.

The control of copper EW tankhouse environments can be divided into two types of strategies. The first deals with source control and includes such measures as hoods and covers, mechanical interference devices, surfactants, and devices that trap or collect gas bubbles at the anode. The most common source control measures are mechanical interference devices and surfactants. In terms of mechanical interference, layers of plastic beads and balls floating on the electrolyte are the most common. Plastic balls and beads create a surface where the oxygen bubbles can burst less violently and also provide a protective layer on the electrolyte surface which depresses mist evolution. On the other hand, surfactants stabilize the oxygen bubbles into a foam layer on the electrolyte surface. This foam blanket traps escaping sulfuric acid droplets. A major difficulty with foam is maintaining a uniform thickness and covering on each cell. If the layer is too thin, bare spots may appear and mist is allowed to escape into the tankhouse environment. On the other hand, when the foam layer becomes too thick, it may come into contact with the bus bar system and interfere with the electrical contacts. The second strategy deals with secondary control measures which includes mechanically assisted ventilation systems. Davis and Hopkins describe a "Close Capture" design that limits the acid mist to within 75 mm of the top of the electrolytic cell.[27] This maintains the air quality of the breathing zone and prevents corrosion of the tankhouse structure.

Cell Construction. Concrete cells have been used in electrolytic operations for years. To protect the concrete, liners must be incorporated into the cell design and construction. Originally lead was used as a lining material. The liners would become oxidized and would frequently leak. There are potential health hazards associated with lead. Furthermore, lead liner systems required a skillful "lead burner" for fabrication and repair. For a substantial period, lead was replaced with other liner materials and systems such as PVC liners, fiber-reinforce plastic liners, and HDPE liners.

In recent years the trend has been towards the use of polymer concrete cells. In fact, polymer concrete cells are installed in all of the new EW tankhouses. These cells are of unitized construction and require no liner or buffer sheets. These are constructed from a modified vinylester resin with corrosion resistant fillers and a nonconductive rebar as reinforcement. Precise casting technologies produce cells that have no seams or bonded joints and the cells can incorporate integrally cast overflow and decant system. This class of material has excellent mechanical properties and cells can be constructed with thinner and strong cell walls (i.e., 6.4 cm polymer concrete versus 12.7–15.2 cm conventional). Sloped nonskid bottoms are possible with these materials and fabrication techniques. This feature aids in the removal of slime particles from these cells.

REFERENCES

1. Ulke, T., *Modern Electrolytic Copper Refining,* Wiley, New York, 1903.
2. Schloen, J.M. and Davenport, W.G., "Electrolytic copper refining—world tankhouse operating data," *Proc. Copper 95 Int. Conf., Vol III—Electrorefining and Hydrometallurgy of Copper,* W.C. Cooper et al. eds., TMS-CIM, Warrendale, PA, 1995, pp. 3–25.
3. Hofman, H.O., *Metallurgy of Copper,* McGraw-Hill Book Co., Inc., New York, 1914.
4. U.S.G.S., Mineral Commodity Summaries: http://minerals.er.usgs.gov/minerals/pubs/mcs

5. Tseidler, *Metallurgy of Copper and Nickel.* Jerusalem: U.S. Department of the Interior and the National Science Foundation, 1964.
6. Mattsson, E. and Bockris, J.O.M., "Galvanostatic studies of the kinetics of deposition and dissolution of copper + copper sulfate system," *Transactions of the Faraday Society,* Vol. 55, 1959, pp. 1586–1601.
7. Hoey, D.W., et al., "Modern tank house design and practices at copper refineries pty. ltd.," *The Electrorefining and Electrowinning of Copper,* Hoffmann, J.E. et al. eds., TMS, Warrendale, PA, 1987, pp. 271–287.
8. Kemp, D.J., Matwijenko, O., and Scott, J.D., "Development of the Kidd Process permanent stainless steel cathode technology at Falconbridge Limited, Kidd Creek Division," *Proc. Copper 91 Int. Conf., Vol III—Hydrometallurgy and Electrometallurgy of Copper,* W.C. Cooper et al. eds., Pergamon Press, New York, 1991, pp. 529–539.
9. Schloen, J.H., "Electrolytic copper refining tank room data," *The Electrorefining and Electrowinning of Copper,* Hoffmann, J.E. et al. eds., TMS, Warrendale, PA, 1987, pp. 3–18.
10. Schloen, J.H., "Electrolytic copper refining—tank room data," *Proc. Copper 91 Int. Conf., Vol III—Hydrometallurgy and Electrometallurgy of Copper,* W.C. Cooper et al. eds., Pergamon Press, New York, 1991, pp. 491–507.
11. Pariani, R.L., "Mechanization of full deposit stripping in electrowinning," *The Electrorefining and Electrowinning of Copper,* Hoffmann, J.E. et al. eds., TMS, Warrendale, PA, 1987, pp. 449–455.
12. Tallert, H.E. and King, M.G., "Twenty years of ongoing improvement at the Amarillo Copper Refinery," *Proc. Copper 95 Int. Conf., Vol III—Electrorefining and Hydrometallurgy of Copper,* W.C. Cooper et al. eds., TMS-CIM, Warrendale, PA, 1995, pp. 81–91.
13. Dreisinger, D.B. and Scholey, "Ion exchange removal of antimony and bismuth from copper refinery electrolytes," *Proc. Copper 95 Int. Conf., Vol III—Electrorefining and Hydrometallurgy of Copper,* W.C. Cooper et al. eds., TMS-CIM, Warrendale, PA, 1995, pp. 305–314.
14. Collins, D.W. and Hiskey, J.B., unpublished results, 1998.
15. Goffman, M. and Jordan, T.L., U.S. Patent No. 4,511,443, April 16, 1985.
16. Langner, B., et al., U.S. Patent No. 4,834,842, May 30, 1989.
17. Bosscher, J. and Berends, W., "Advances in copper tankhouse automation," *Proceedings of the International Symposium on Electrometallurgy Plant Practice,* Claessen, P.L. and Harris, G.B. eds, Pergamon Press, New York, 1990, pp. 271–279.
18. Jenkins, J.G. and Eamon, M.A., "Plant practices and innovations at Magma Copper Company's San Manuel SX-EW Plant," *Proceedings of the International Symposium on Electrometallurgy Plant Practice,* Claessen, P.L. and Harris, G.B. eds, Pergamon Press, New York, 1990, pp. 41–56.
19. Robinson, T., Davenport, W.G., and Jenkins, Jackson, "Electrolytic copper electrowinning and solvent extraction world operating data," to be published.
20. Prengaman, R.D., "New insoluble lead anodes for copper electrowinning," *The Electrorefining and Electrowinning of Copper,* Hoffmann, J.E. et al. eds., TMS, Warrendale, PA, 1987, pp. 457–467.
21. Nidola, A. and De Nora, O., "Catalytic anodes based on lead and their use," Patent, Ger. Offen. DE 3,405,059, 1984, 13 pp.
22. DeNora, V., Nidola, A., and Bianchi, G., "Mechanically and chemically resistant long-lasting electrodes," Patent, Can. CA 1,147,699, 1983, 32 pp.
23. Nidola, A., "Electrode for use in electrochemical processes," Patent, PCT Int. Appl. WO 86 03,790, 9184, 39 pp.

24. Tyroler, P.M. et al., “Copper electrowinning at INCO’s refinery,” *The Electrorefining and Electrowinning of Copper,* Hoffmann, J.E. et al. eds., TMS, Warrendale, PA, 1987, pp. 421–435.
25. Anderson, T.N., Adamson, D.L., and Richards, K.J., “The corrosion of lead anodes in copper electrowinning,” *Met. Trans.,* Vol. 5, 1974, 1345–1349.
26. Gendron, A.S., Ettle, V.A., Shinichiro, A., “Effect of cobalt added to electrolyte on corrosion rate of Pb-Sb anodes in copper electrowinning,” *Can. Met. Quart.,* Vol 14, No 1, 1975, pp. 59–61.
27. Davis, J.A. and Hopkins, W.R., “Recent developments in electrometallurgical tankhouse environmental control,” *CIM Bull.,* Vol 87, No 981, 1994, pp. 86–94.

Eighty-five Years of Operation at Cyprus Miami Mining Corporation

Samuel C. Yang* **and Kenneth H. Larson***

Cyprus Miami Mining Corporation, a subsidiary of Cyprus Climax Metals Company, is located in the historic Globe/Miami Arizona mining district and has been in operation since the 1910s. Some of today's hydrometallurgical facilities started operation from that time era and have experienced numerous changes, improvements, successes and misfortunes. This paper chronicles the past and present hydrometallurgical activities at the site and includes a selection of historical and present day photographs of the facility.

HISTORICAL PERSPECTIVE

Beginnings of the Globe/Miami Mining District

Cyprus Climax Metals Company purchased the Inspiration Consolidated Copper Company (Inspiration) property in 1988 and renamed it Cyprus Miami Mining Corporation (CMMC). The property is located in the historic Globe/Miami Arizona mining district that has had over a century of mining activity. Since the start of commercial mining, approximately 3.8 million tons of copper has been produced at the site.

Commercial mining activities in the Globe/Miami area originally started with gold prospecting in the 1860s. The area was very rugged and remote for the early miners. Nonetheless, many placer gold prospecting parties were sent to this region to find the precious metal. There was very little success in the gold ventures of that era.

In the 1870s, the mining focus switched from gold to silver. The district had a high abundance of exposed silver and was compared in value to the famous Comstock lode in Nevada. Numerous successful silver mining camps, including the Silver Queen, the Silver King, the mines in Richmond basin and around McMillenville, sprung up through the region during this period. The silver rush, however, did not last long and by the 1880s was nearly over.

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* Cyprus Miami Mining Corp., Claypool, Ariz.

The recovery of copper then became the predominant mining activity in the district from the 1880s and onward. During the silver rush, massive copper deposits were known to be available just below the ground. The early silver miners, however, did not capitalize on the copper metal value due to the lure of the abundant silver. In the early 1880s, silver prices dropped due to high worldwide inventories and copper prices began to increase. With the increased copper prices, the copper recovery efforts increased dramatically.

Some well-known copper operations such as the Old Dominion Mine, the Copper Hills, the Black Warrior and the Miami Copper Company rose from the high desert floor from the late 1870s through the 1920s. Included in this growth of copper mining was the Inspiration Mine.

Inspiration

The first mining activity at the Inspiration Mine, where CMMC is presently located, began in 1908 with the sinking of the Woodson tunnel. Other mining claims near the Inspiration Mine were purchased between 1908 and 1911 to form the Inspiration Consolidated Copper Company (ICCC) in 1912. Low grade copper sulfide deposits (high grade by today's standards), about two to three percent copper, were ICCC's target. With exception of about four years during the Great Depression, ICCC/CMMC have continued to operate these mines and plants at high production levels.

CHANGES TO MEET THE ECONOMY

Several innovative mining and processing methods were successfully utilized at ICCC to recover copper. All of the methods, described below, kept high copper productivity and positive cash flow in focus.

Mining

Inspiration, concurrently with Ray Mines, developed and implemented the undercut block caving mining method between 1911 and 1912. Block caving became the primary method for underground porphyry copper mining due to the potential for high production volumes and lower operating costs.

During the mid-1940s, open pit or "opencut" mining was started at Inspiration due to several favorable advantages compared to block caving. Advantages included safer working conditions, visible selectivity of ore grade and higher productivity (pounds of copper per manhour). Inspiration also used a combined opencut/block cave transfer method of mining in the 1950s. In this method, ore collected from the open pit mine was transferred from the pit area by diesel powered trucks to existing haulageways in block caved areas and then trammed to the original ore hoisting shafts that serviced the underground mine. This novel use of new technology (open pit mining and surface haulage trucks) and existing technology (block cave, underground haulage and ore hoisting shafts) reduced capital cost requirements while maintaining high production rates at the mine.

Plant Operations and Leaching

In 1915, ICCC patented the flotation cell and became the first company in the United States to utilize the flotation process commercially to recover sulfide copper. Copper recovery rates with the flotation process were approximately 80% compared to gravity methods which had recoveries in the 65% range. During the same time period, the nearby International Smelting and Refining Company completed their large copper smelter, making it convenient for Inspiration to process their copper concentrates.

Sulfide copper processing through flotation was the prime copper recovery method at ICCC until the mid-1920s. It was known at the time that a significant quantity of oxide copper was present in the ore and in 1926, leaching of the oxide and soluble sulfides began. The original leaching process utilized a sulfuric acid/ferric sulfate leach in vats, followed by direct electrowinning of the pregnant leach solution. Leach copper became the primary copper recovery method at the site from 1926 through 1955.

In 1955, ore mineralogy changed and the “dual process” for copper extraction was developed by ICCC. The dual process included various combinations of sulfuric acid/ferric sulfate leach to recover the oxide copper minerals, followed by application of standard milling/froth flotation on the leach residue to recover the sulfide mineralogy.

In the dual process, crushed ore was leached in a vat leach operation with a ferric sulfate/sulfuric acid solution. Due to the presence of the ferric sulfate, sulfide mineralogy in the form of chalcocite was effectively leached along with oxide copper. The pregnant leach solution was treated with direct electrowinning at the tankhouse to produce copper cathodes. The solid leach residue was removed from the vats, neutralized, then sent to the mill for the recovery of the sulfide copper. Several other intermediate steps were implemented in conjunction with the dual process to optimize copper recovery. The dual process continued at ICCC from 1955 until 1979. Changes in the ore mineralogy and new technological advances forced the change from the dual process.

ICCC also utilized “in-situ” leaching from 1950 through 1974. In this process, acid leach solution was introduced into the caved area overburden for the recovery of oxide copper. After leaching, the copper in the pregnant leach solution was precipitated on scrap iron.

Heap leaching began at ICCC in 1962. At that time, a waste rock dump, No. 5 dump, was treated with acidic leach solutions, and the pregnant leach solutions were again processed through iron cementation. The OxHide Mine, started in 1967, was one of the first “mine for leach” operations and used the heap or dump leach technology. In 1969, another waste rock dump that contained some leachable sulfide ore, No. 9 dump, was also activated as a leach dump.

About ten years after implementation of the dump leach technology, ICCC installed a 4,000 gpm solvent extraction plant for purification of the pregnant leach solution prior to electrowinning at the existing tankhouse. Shortly thereafter, the development of the ICCC patented high acid/high ferric “ferric cure” process began. High total copper recoveries of 75% and greater were achieved with this process. The ferric cure process includes exposing the ore to a solution containing ~200 g/l acid and 5 to 8 g/l ferric iron, followed by a rest period of ~5 days prior to rinsing of the treated ore.

Due to further changes in ore mineralogy, the mill/concentrator facility was shut down in 1986 and the ferric cure/solvent extraction/electrowinning method of copper recovery became the mainstay of processing the ore taken from the ICCC mines from 1986 through today.

In 1988, Cyprus Climax purchased the Inspiration property from Anglo American and renamed it Cyprus Miami Mining Corporation. Innovations continued to be utilized throughout the property to keep CMMC operating profitably.

Mining

In 1993, long range planning forecasts identified an optimum mine/dump leach/SX/EW production rate of 158MM pounds of EW cathode per year from the CMMC operation. To achieve this, seven new 240-ton haul trucks, two P & H 53 yard 4100 shovels, and one track dozer were purchased. A long range mining plan was developed to place ore at the optimum production rate on the existing dumps.

Dump Leach/SX

Leach solution and ore dump production planning is controlled to maximize recovery and reduce material handling costs. In the 1980s, dump stacking was instituted as a means of increasing production without increasing flowrates at the SX plant. Raffinate was applied to the older and lower grade PLS dumps. The off solution from these dumps was applied to the higher grade dumps. The leach solution from the higher grade stacked dumps was pumped to the aqueous feed pond at the SX plant. This strategy was effective in increasing production with very low capital cost and continued through 1995.

The use of triple stacking techniques allowed the mine to produce 100MM pounds in 1987. While the stacking of ore enabled the processing of higher grade leach solution without increasing flowrate through the SX plant, it also increased the volume of leach solution and, therefore, the copper contained in dump inventories. Estimates made at that time calculated 28 million pounds of copper were held in the dumps.

The decision was made in 1987 to build additional SX capacity to increase flows and recover approximately 16 million pounds from the dump inventory. Several process designs were considered, including series/parallel. The most attractive alternative was to build a second train identical to the first. With this option there was very little production downtime, no design risk, and the engineering was minimal. In August of 1988, construction of B train was completed and brought the SX plant flow capacity to 8,500 gpm.

In 1994, the limited leach pad space on the existing dumps became a concern at CMMC. Installation of new, lined leach pads was not an attractive option due to the high capital cost. Instead, beginning in the summer of 1995, the “double lift rinse” (DLR) practice of leaching was implemented at CMMC. In this procedure, a leach pad that has been ferric cured but not leached was immediately overlaid with another leach pad that has been ferric cured. When the “top” pad completed the cure, the leach solution of raffinate was applied to the top pad and allowed to percolate through the two pads. There were no copper recovery losses noted with the DLR process. The DLR, which continues to be used today, has allowed CMMC to delay construction of new lined pads.

As the leach dumps grew in size with each year of operation, the inventory of copper bearing PLS solution in the dumps continued to grow as well. Higher flowrates through the leach dumps and the SX plant were again required to recover the PLS inventory. In the 1996 and 1997, the two trains at SX were converted to series/parallel operation. CMMC employees completed nearly all the work, and downtime required to implement the new mode of operation was minimal. The flow capacity through the plant currently stands at 17,500 gpm.

Another novel idea recently (1998) implemented in the CMMC leach system is the “capping” of existing leach dumps with a semipermeable soil liner so that copper recovered from new ore placed on top of the soil liner could be "quickly returned" to the SX/EW plant for conversion into cathodes. Without the soil liner cap, leach solutions must percolate completely through the existing leach dumps, some of which are 500 feet high, and may take up to 6 months before being recovered at the SX/EW plant.

These soil liners, of course, must be placed only on leach dumps that have been thoroughly rinsed with raffinate to recover the majority of the copper in solution inventory. If complete rinsing is not achieved, large amounts of copper could potentially be left in the leach dump and never be recovered. To counteract this potential, CMMC is testing what is called an "underliner rinse system." The underliner rinse system consists of placing raffinate distribution lines under a soil liner cap mentioned earlier. Raffinate is pumped through the distribution lines to flush out any remaining copper bearing solution in the leach dump that is under the soil liner. The underliner rinse system along with the soil liner achieves two things: (1) recovery of copper that may be under the soil liner and (2) acceleration of the recovery of new copper that has been placed on top of the liner.

Electrowinning

The tankhouse facility began operation at CMMC in 1926. A devastating fire that occurred in 1944 nearly destroyed the entire tankhouse. Today, much of the 1940s vintage equipment still exists and is used daily at the tankhouse.

In 1994, the increased production from the mine/dump leach/SX and the modernization of the CMMC smelter (Inspiration purchased the adjacent International Smelting and Refining Company in 1960) required the tankhouse to be upgraded as well. Prior to the expansion, the tankhouse operated with 140 electrowinning cells to process the copper recovered from the leaching operation. There were also 80 electrorefining cells (installed in 1960) in operation at the tankhouse to process anodes produced from the CMMC smelter.

A new (October 1994) electrorefinery was constructed to accept the higher anode production capacity smelter operation, which uses the proprietary ISA smelt technology, implemented in 1993. The new electrorefinery uses the most technologically advanced equipment to produce up to 206,000 tons of refined cathode per year. Construction of the electrorefinery allowed the conversion of the 80 electrorefining cells at the tankhouse to electrowinning cells. This conversion brought the tankhouse electrowinning capacity to over 100,000 tons per year of electrowon cathode.

Other Installations

Several other installations at the CMMC site have been implemented since the time of mine inception. These installations included the following:

- SCR designed continuous cast rod plant (1968)
- Double contact acid plant; first smelter to meet stringent air quality standards (1974)
- Elkem electric furnace at the smelter (1974)
- Holistic erosion control of old mill tailings (1995)

CYPRUS MIAMI FUTURE

The future continues to look good for CMMC. Domestic mine SEC reserves currently are estimated at approximately 200 million tons of ore. This places the life of the mine through the year 2006.

With low copper prices forecast for the next several years, CMMC's focus to remain competitive is to achieve the continuous improvement goals in safety, environment, quality, productivity, cost control and employee relations and development. These goals will be achieved through utilization of a recently initiated corporate wide quality management system called QUEST 21 (Quality Enrichment and Success Through the 21st century). QUEST 21 follows the basic guidelines of ISO 9000 and QS 9000 and will help maximize profitability at the site.

APPENDIX—CYPRUS MIAMI MINING CORPORATION SOLVENT EXTRACTION AND TANKHOUSE GENERAL OPERATING DATA

Solvent Extraction

General

Production rates

Nominal flowrate	17,500 gpm
Nominal PLS grade	2.77 g/l
Nominal raffinate grade	0.15 g/l
Nominal production	200,000 Mlbs/Y

Design

Configuration	2 trains Series/parallel 2-series extract **X** 1-parallel extract **X** 1-strip for each train
Electrolyte filtration	8 dual media filters (anthracite/garnet)
Raffinate organic recovery	6 WEMCO Pacesetter coalescing units

Operation

O:A	1:1 extraction stages; 3:1 strip stage
Reagent strength	18%
Extraction efficiency	94%
Raffinate organic entrainment	<60 ppm
Electrolyte organic entrainment	<10 ppm

Tankhouse

General

EW capacity	200,000 Mlbs/Y
Normal pull days/yr.	208 days
Normal starter sheet pull days/yr.	365 days
Cathode pull cycle commercial	14 days
Cathode pull cycle stripper	24 hours

Anodes

Weight	320 lbs
Nominal effective anode dimensions	
Width	37½″
Length	46½″
Thickness	¼″

Starter Sheet Blanks

Material titanium

Dimensions	
Width	42″
Length	44″
Thickness	(3 mm) 0.118″

Hanger bar is copper clad steel core 1¼″ square × 57½″

EW Anodes

Analysis	
Lead	98.585%
Calcium	0.65%
Tin	1.35%

(Rolled sheet with copper hanger bar 110 ETP rolled; avg. life of 3–5 years)
Insulator is a 2 piece PVC strip, ¾″ base, tapered 1½″ high and 39″ long.
Four insulator strips per anode.

Arrangement—Electrolytic Cells

Number of cells total	220
Number of stripper	20
Number of commercial (includes 36 "swing" tanks)	200

EW tanks in two stages:

Stage 1 (incoming from SX) consists of OTH cells of 120 commercial and 20 starter sheet with feed piping to groups of 20 cells. Cells are arranged in pairs with common center wall.

Stage 2 (overflow from Stage 1) consists of NTH cells all 80 in commercial production. Cells are arranged in sections of 20 with common walls.

Cell dimensions

	NTH	OTH Comm.	OTH Stripper
Width	4′	4′	4′
Length	35′6″	34′8″	35′6″
Depth	4′7⅛″	4′6″	4′3″

Cell construction	Concrete (steel reinforced)
Cell lining	PVC
Number of anodes	
Commercial	90
Stripper	92
Cathode spacing	
Commercial	4½″
Stripper	4½″

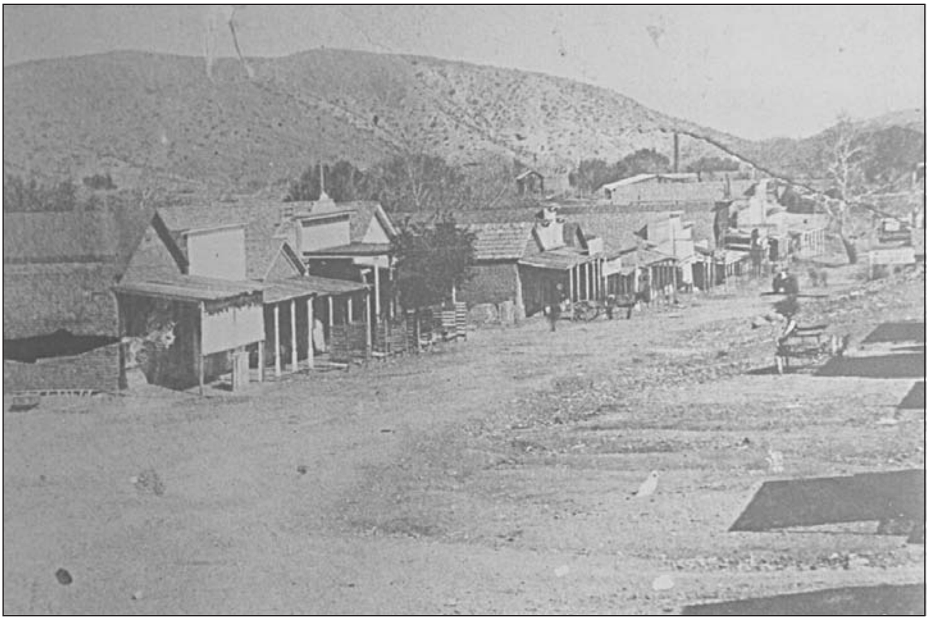
Cathode dimensions (starter sheet)	
Width	41¾″
Length	43″
Thickness	0.65 mm
Avg. weight	16 lbs
Current per cell (NTH & OTH)	50,000 amps
Current density	
Commercial	16.6 amps/ft^2
Stripper	21.2 amps/ft^2
Voltage	
Commercial	1.8 volts
Stripper	1.9 volts
Electrical connection	Cells/series; electrodes/parallel
Finished cathode weights	
EW	165–335 lbs.
Stripper	15–16 lbs.

Electrolyte

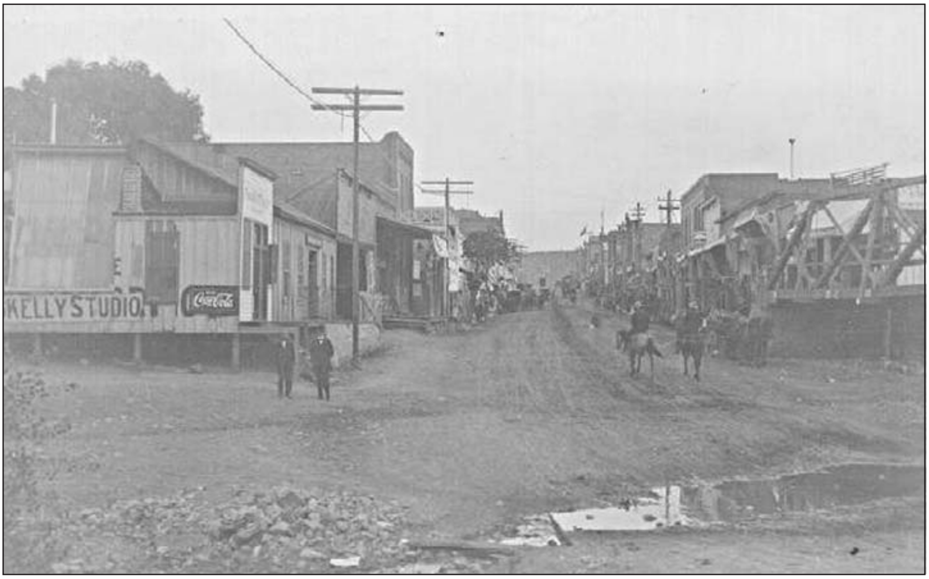
Circulation per cell	(Stage 1)	24–40 gpm
	(Stage 2)	38–63 gpm
g/l copper at overflow	(Stage 1)	35–40 g/l
	(Stage 2)	30–36 g/l
g/l H_2SO_4		180–200 g/l
Cell overflow temperature		38–45°C
Guar (Guartec EW)		0.5 lbs/ton
FC–1100 (Demister)		100–150 mg/l

Cathode Analysis

Element	ppm
Ag	1.0
Fe	1.0
Se	0.5
Zn	<1.0
Sn	<1.0
Ni	<1.0
Bi	<0.1
Sb	<1.0
As	<1.0
Te	<1.0
Cd	<1.0
Pb	<5.0
S	<10.0



Downtown Globe, Arizona—1890



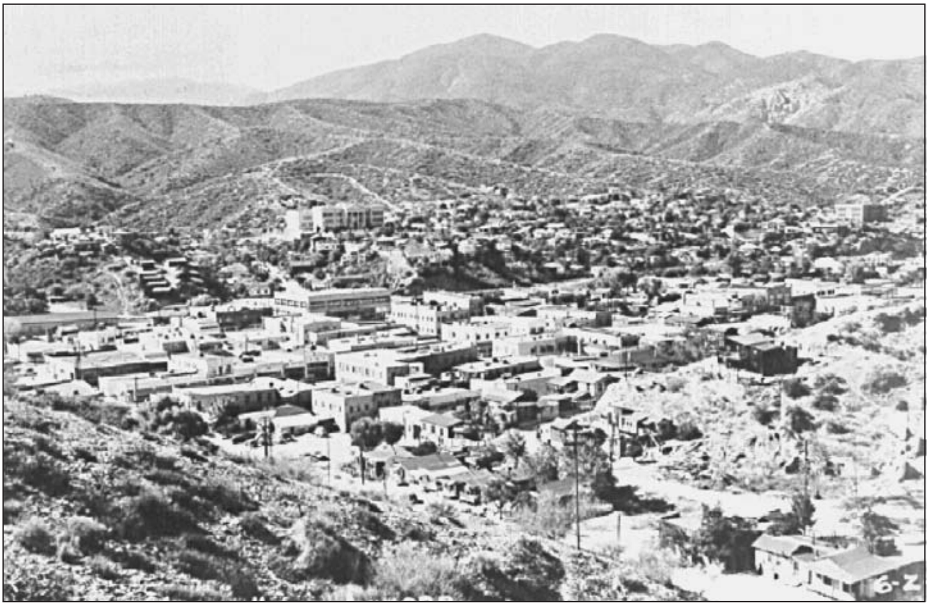
Globe, Arizona—1904, after the flood

Globe Arizona—the flood of 1904

Inspiration townsite—1915

Globe, Arizona, labor strike—1917

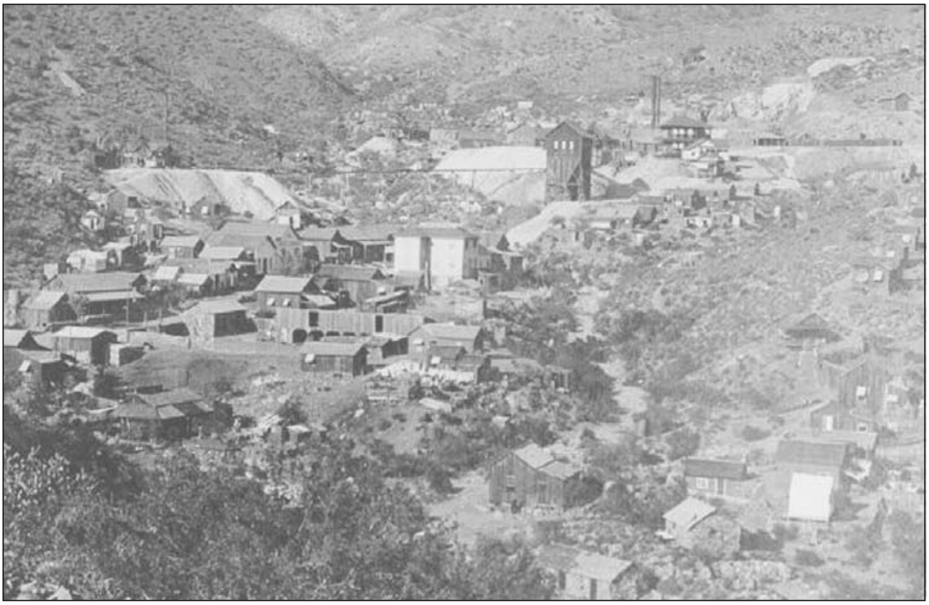
Downtown Globe, Arizona—1920



View of Miami, Arizona—1920

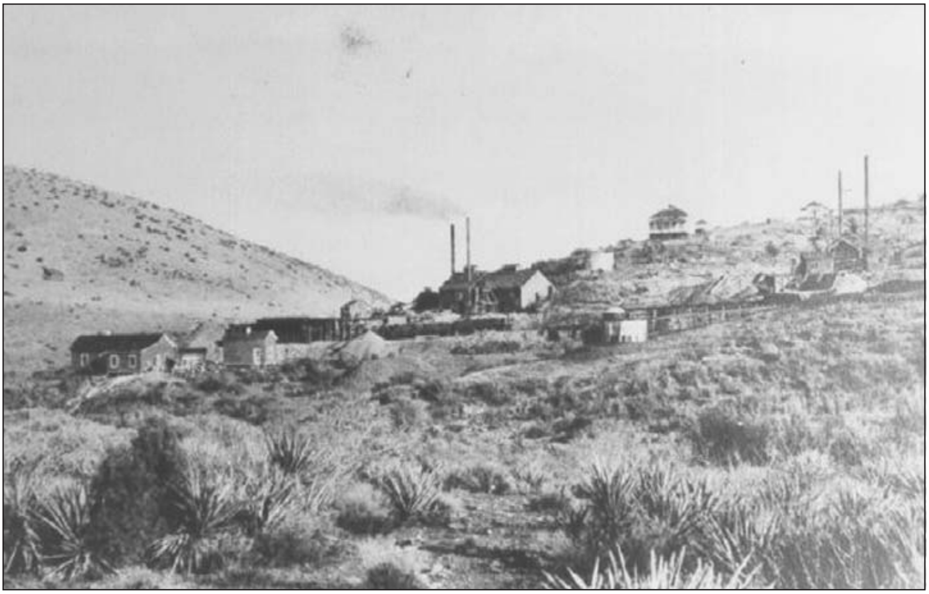
Bullion Plaza, Miami, Arizona—1928

SILVER MINING ACTIVITY



Silver King, Arizona—1880

COPPER MINING ACTIVITY



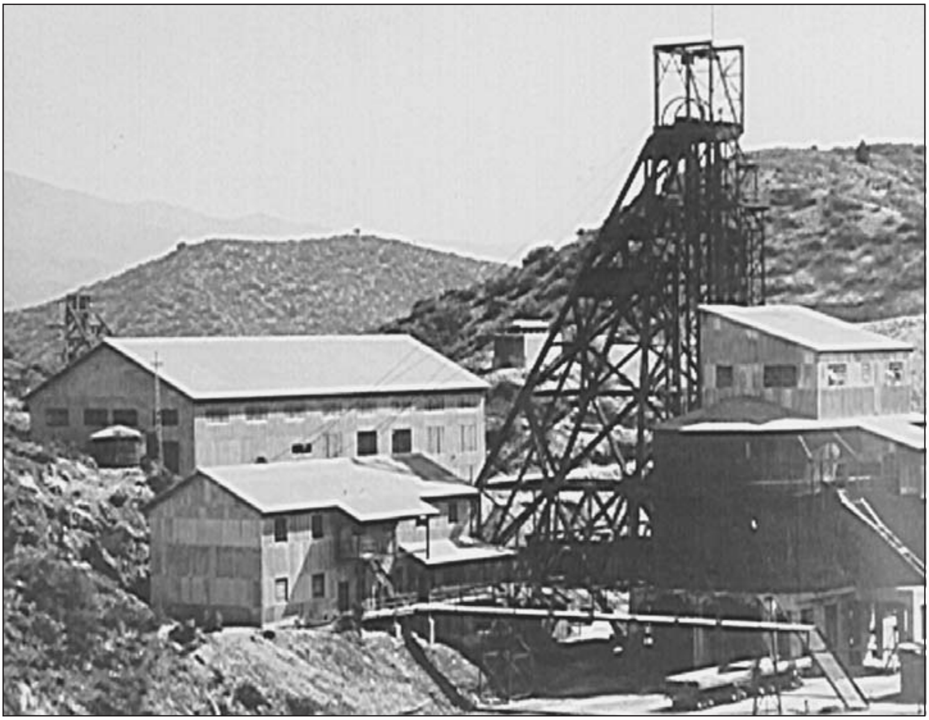
Copper Hills Mine, Globe, Arizona—1890

Black Warrior Mine, Miami, Arizona—1896

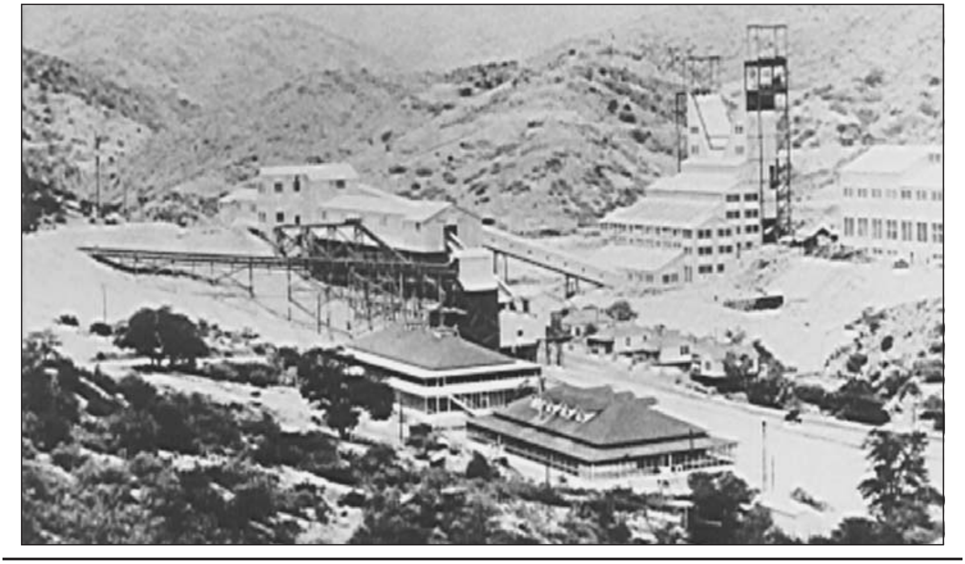
Old Dominion Mine, Globe, Arizona—1910

Miami Copper Mill, Miami, Arizona—1915

Shift change, unknown mine—1920



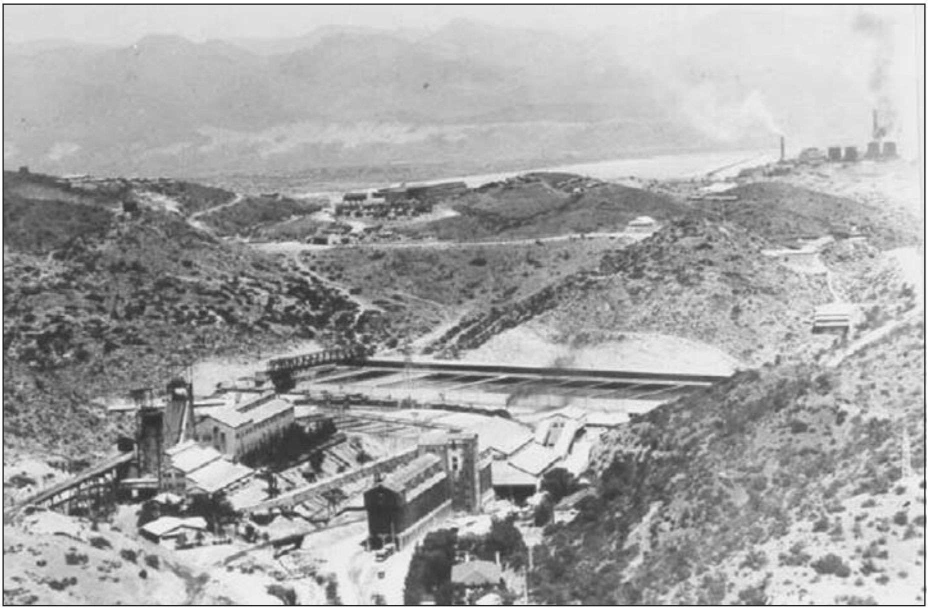
Inspiration live oak shaft—1926



Inspiration fine crushing plant—1915

Inspiration mill/concentrator—1915

Mine development, Inspiration, Arizona—1920



View of Inspiration operations—1929

Inspiration crushing/leach and electrowinning plant—1930–1940

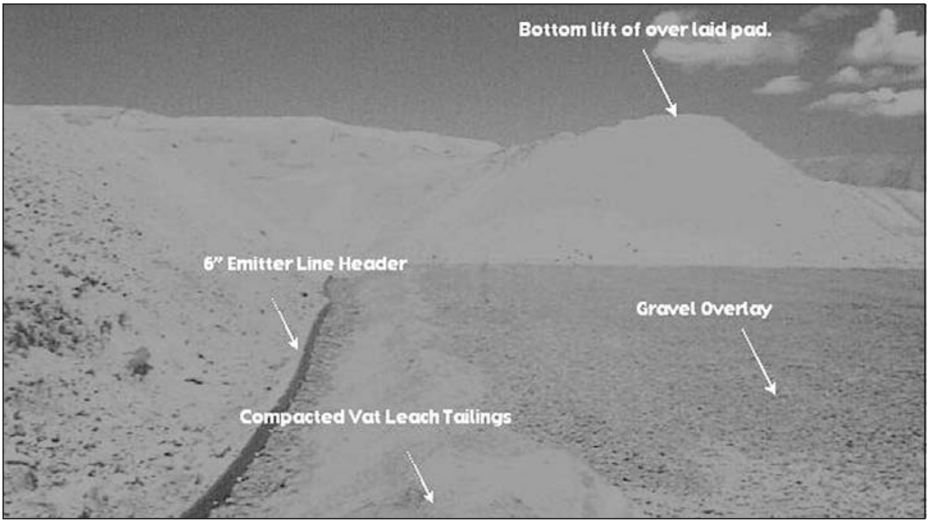
Inspiration vat leach and electrowinning plant—1930–1940

International Smelting, Miami, Arizona—post-1950

Cyprus Miami Mining Corporation, view of mine/dump and leach/SX/EW

Cyprus Miami Mining Corporation mine operations

Cyprus Miami Mining Corporation dump leach operations

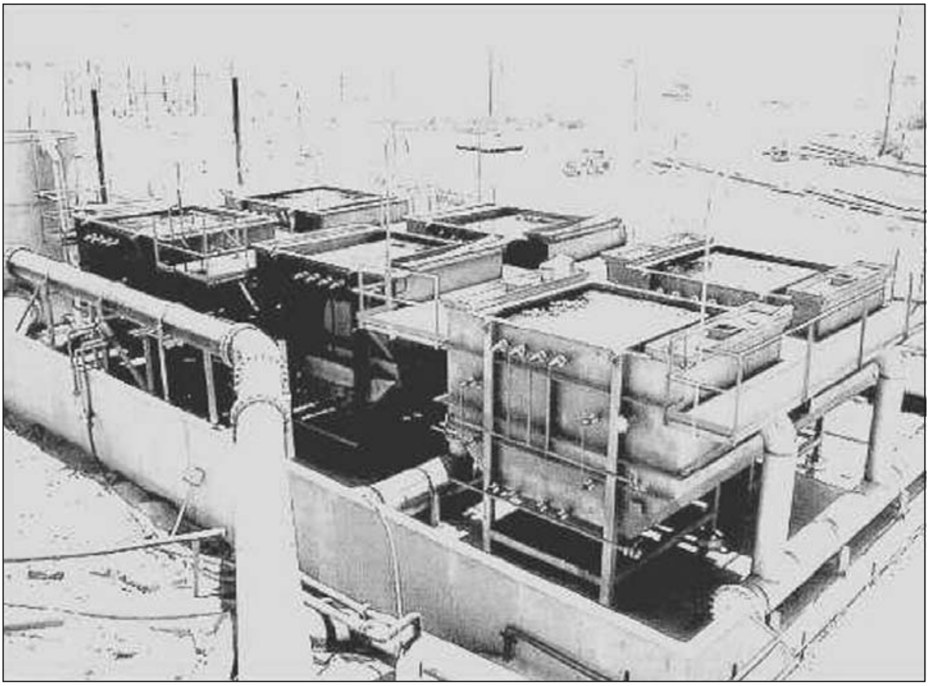


Cyprus Miami Mining Corporation under liner leach system

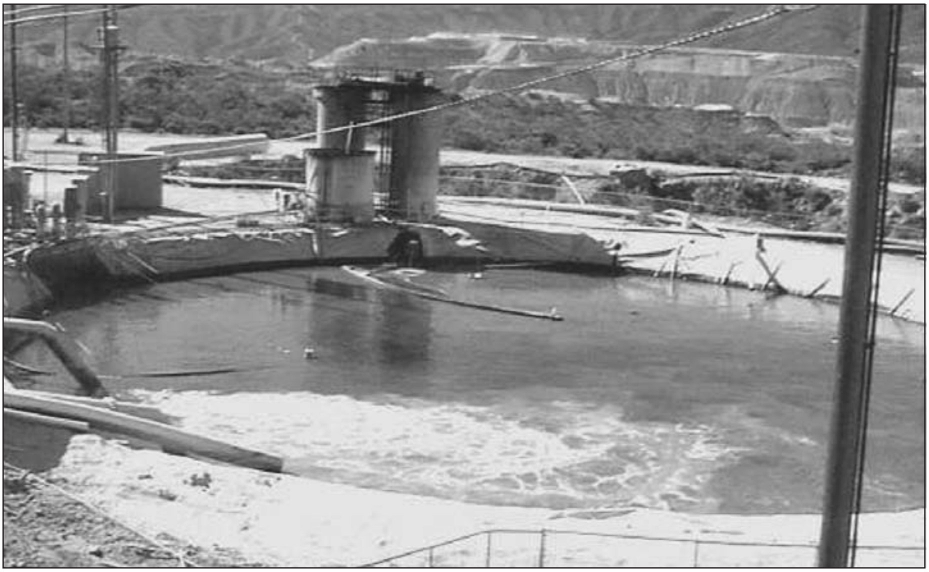
Cyprus Miami Mining Corporation 28 dump PLS reservoir

Cyprus Miami Mining Corporation SX plant mixer/settlers

Cyprus Miami Mining Corporation SX plant tank farm

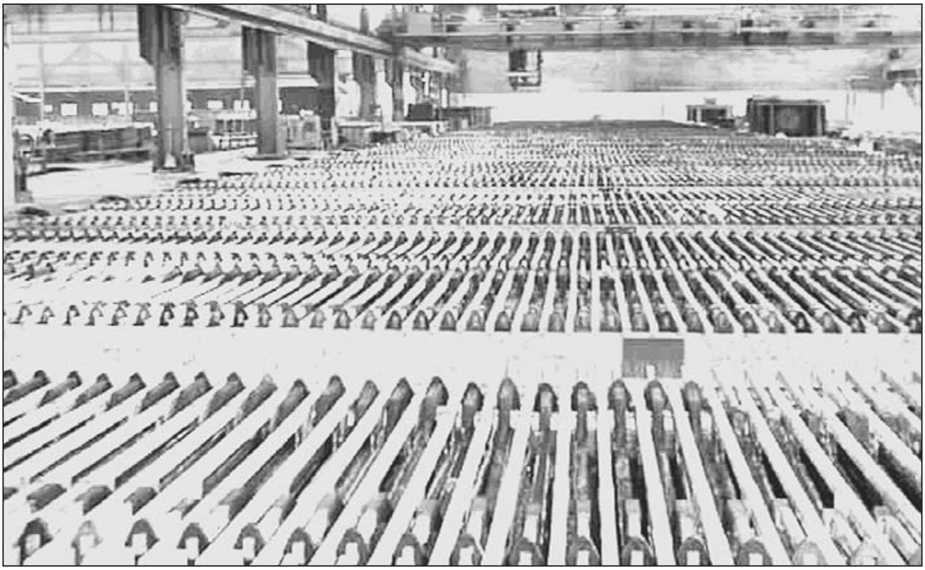


Cyprus Miami Mining Corporation SX plant coalescers



Cyprus Miami Mining Corporation SX plant raffinate pond

Cyprus Miami Mining Corporation EW tankhouse facility

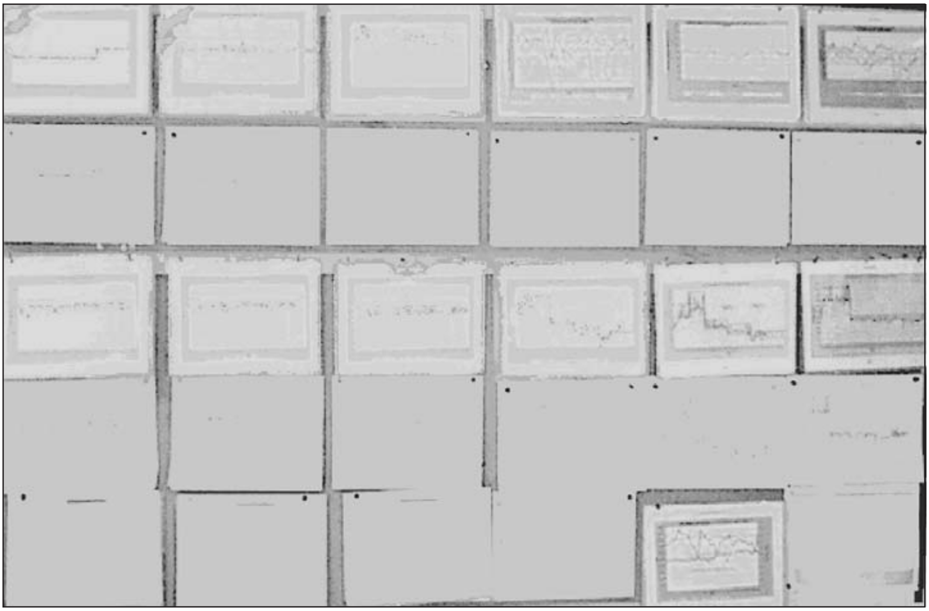


Cyprus Miami Mining Corporation EW tankhouse

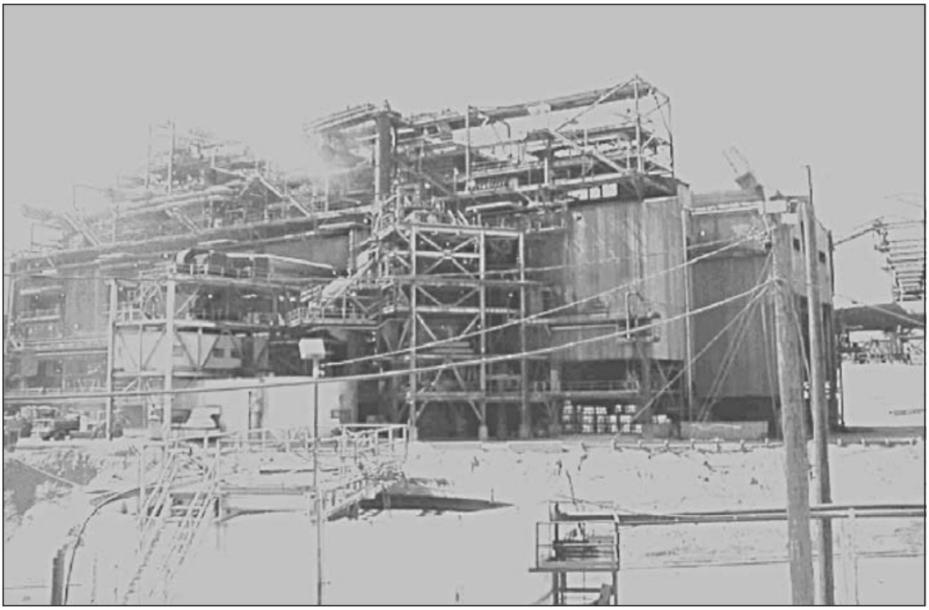
Cyprus Miami Mining Corporation EW cathodes

Cyprus Miami Mining Corporation surface tension measurement

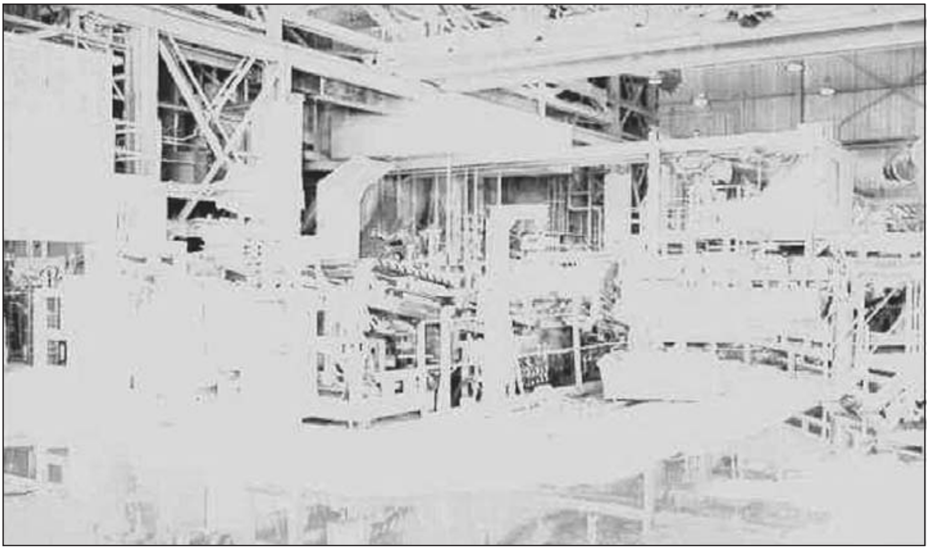
Cyprus Miami Mining Corporation employee involvement



Cyprus Miami Mining Corporation EW SPC control charts



Cyprus Miami Mining Corporation smelter operations



Cyprus Miami Mining Corporation smelter anode casting

Cyprus Miami Mining Corporation new electrorefinery

Cyprus Miami Mining Corporation rod plant

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Do's and Don'ts of Tankhouse Design and Operation

Chris L. Pfalzgraff[*]

The successful performance of an electrowinning tankhouse relies on both the initial design and execution as well as the subsequent operating practice. Some of the more important elements incorporated into the initial design of a tankhouse include electrolyte quality and distribution, electrical distribution, material handling, materials of construction, acid mist control and maintenance philosophy. However, tankhouse performance is dependent on the operational practice of inspection, control and correction of parameters which go outside the normal operating range.

The paper will discuss these issues with an emphasis on the criteria which are critical to plant performance and those which improve operator efficiency. The ultimate success of the plant depends on a joint effort between the operating/maintenance personnel and the designer.

INTRODUCTION

Copper electrowinning has a long history, with substantial production starting early in this century. The introduction of solvent extraction in the 1960s changed the image of electrowon copper within the industry, as it became possible to consistently produce high quality cathode. While the basics of the electrowinning process have not changed, the design of the plants and their operation have improved significantly in recent years.

The main areas to be addressed in the design of an electrowinning tankhouse are:

- Electrolyte quality
- Electrolyte distribution
- Electrical distribution
- Material handling
- Materials of construction
- Acid mist control

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* Bechtel Mining and Metals, Greenwood Village, Colo.

Except for the materials of construction, the same factors are important to the operation of a copper electrowinning tankhouse.

TANKHOUSE DESIGN ELEMENTS

Electrolyte Quality

Electrolyte quality involves many factors such as electrolyte chemistry, electrolyte temperature, eH and organic and solids entrainment.

Electrolyte chemistry is critical to the production of high quality cathodes. The major components, copper and sulfuric acid, are controlled by the plant operators to maintain the desired conditions in the electrowinning cells. The mass balance for the solvent extraction/electrowinning plant must also account for these components. In addition, the concentration of impurities can affect current efficiency, cathode quality, materials of construction and anode life. The impurities of most concern in a modern copper electrowinning tankhouse are iron, chloride and manganese. There is some chemical iron transfer in solvent extraction, but the majority of the impurities are transferred by entrainment of pregnant leach solution in the loaded organic.

The current efficiency is mainly affected by the concentration of iron in the electrolyte and the oxidation/reduction of the ferrous/ferric couple at the anode and the cathode. Minimizing the concentration of iron reduces the consumption of current by this reaction.

The use of stainless steel for cathode mother blanks requires that the chloride concentration be kept within limits which will not corrode or etch the blade. Related to this is control of the eH to prevent damage to the stainless steel cathode blade due to the evolution of chlorine gas.

If the concentration of manganese in the electrolyte is not controlled, it will deposit on the anodes as manganese dioxide and spall as well as passivate the anode. These chips will carry some lead from the anode surface, leading to increased lead contamination of the cathode product. Equally important is the ratio of iron to manganese in the electrolyte. Because the iron is more reactive, the reduction of ferric to ferrous will take precedence over the deposition of manganese dioxide.

Consistent control of the electrolyte temperature is important to produce dense, high-quality cathode copper. In most operations, the excess heat generated in the electrowinning process is used to heat the incoming rich electrolyte to maintain the electrolyte temperature in the cells. Under extreme conditions of low pregnant leach solution temperature or cold climate, auxiliary heat may be required to maintain an electrolyte temperature of approximately 50°C in the cells. Auxiliary heating capacity is also beneficial for the initial startup of the electrowinning tankhouse allowing more rapid ramp up to design production of high-quality copper.

Entrainment of solids or organic in the electrolyte is detrimental to the cathode quality. Solids will report as impurity in the cathode and can also lead to dendritic growth on the cathode surface by providing nucleation sites. Organic in the electrowinning cells will lead to “burning” of the cathode copper resulting in a porous growth which will be contaminated. These factors make it extremely important to clean the rich electrolyte before it reports to the electrowinning circuit. This cleaning can be accomplished in a number of ways, but the most common techniques are column flotation and sand filters. Proper backwash and maintenance of this equipment is essential to maintain electrolyte quality in the electrowinning circuit.

Additives to the electrolyte, such as cobalt and guar, must be monitored and controlled for greater efficiency and higher cathode quality. The cobalt in the electrolyte stabilizes the lead oxide layer on the anode to decrease the likelihood that it will become detached and contaminate the cathode product. Typical cobalt concentrations in the electrolyte are 30 to 120 parts per million. Guar is used as a leveling agent to produce smoother, more dense cathodes. Consumption of guar is proportional to cathode production, approximately 500 grams per ton.

Electrolyte Distribution

Two factors must be considered in the distribution of electrolyte in electrowinning. The first is the distribution to the cells and the second is the distribution within the cells. The flow of electrolyte must be evenly distributed to each cell which can be accomplished by proper sizing of the header pipe and the feed lines to the individual cells. During operation, it is important to visually monitor the distribution from cell to cell in order to produce high quality cathodes in all cells. Cells with low electrolyte flow will produce less dense cathode copper with higher impurities, especially sulfur.

The flow distribution of electrolyte within each cell is also important. Generally this is accomplished by using a header in the bottom of the cell with holes drilled to direct the electrolyte flow between the anodes and cathodes. The use of this type of distribution system increases the velocity of electrolyte across the face of the cathode plate allowing electrowinning operation at higher current densities. Many of the newer plants worldwide are operating at current densities in excess of 300 amps per square meter. It is important to keep this header clean to ensure uniform distribution along the full length of the cell.

Electrical Distribution

The two main goals of electrical distribution in a copper electrowinning tankhouse are to (1) eliminate, or at least minimize, stray currents and (2) uniformly distribute the current within each cell in order to maintain high current efficiencies.

The two major routes for stray currents around the electrowinning cells are (1) conductive materials of construction and (2) electrolyte conducting to ground or other facilities. Generally, the cells, piping and work platforms are constructed from nonconductive materials, i.e., plastics of one form or another. The standard for copper electrowinning cell construction is polymer concrete which is both nonconductive and nonreactive with the electrolyte. Piping is generally PVC, FRP or HDPE inside the tankhouse. The work platforms are generally FRP for both the support structure and the grating. The introduction of electrolyte into the cell line should be at the electrical null point. This will minimize the flow of current from the cell line to external facilities or ground.

The second major route for stray current is through electrolyte leaks. The cells are electrically insulated from ground and any electrolyte dripping from the cells can provide a path for stray currents around the insulation. The most effective way to combat this problem is to maintain good housekeeping practices, especially washing around and below the cells. It is therefore necessary for the designer to provide access for this housekeeping.

The electrical distribution within each cell depends on good contacts between the electrodes and the intercell bus bar as well as eliminating shorts within the cells. It is essential that the operator check each electrode on a daily basis for any shorts or poor contacts. There are a number of methods which have been used successfully such as infrared scanners, ammeters, etc. The important thing is that a method be adopted and followed rigorously every day. Good housekeeping, which includes daily washing of the electrode contacts, is important to maintaining uniform current distribution within each cell.

Material Handling

The coordination of the tankhouse crane and the cathode stripping machine is critical to the harvest of the cathode copper. A time and motion study can be very helpful in determining the required crane speeds and the required rate of cathode stripping. The tankhouse harvest pattern should be designed to give approximately the same crane travel each day.

In the design and construction of the tankhouse, it is critical to have proper alignment of the electrowinning cells, the capboards and the electrodes. The harvesting cycle time can be greatly affected by poor alignment because of the additional time required to attach the cathodes with the crane and reinsert the cathodes in the cells.

Materials of Construction

In a copper electrowinning tankhouse, the basic guidelines are that the materials of construction are both resistant to corrosion and nonconductive for all equipment and structures around the cells as explained in the section on electrical distribution. The building structure and the stripping machine must also be resistant to corrosion due to the acid mist generated in the electrowinning cells. The most common materials of construction for equipment, piping and structures are PVC, HDPE, FRP and austenitic stainless steel. The building structure is generally limited to mild steel with a protective coating due to the cost. Wood has also been used as an even lower cost alternative. During the design stage, the materials of construction must be evaluated for meeting the criteria of corrosion resistance and nonconductivity as well as cost.

Acid Mist Control

Acid mist in the tankhouse is created when the oxygen bubbles generated at the anodes burst at the electrolyte surface. The goal of acid mist control is twofold: to provide a suitable work environment inside the tankhouse building for the operators and to minimize the corrosion to the building and equipment inside the tankhouse. There are three basic methods to control the acid mist which can be used individually or in combination. These three methods are minimizing the escape of acid mist into the atmosphere by coalescence of the oxygen bubbles generated at the anode, trapping the acid mist within the cell for removal and air movement or ventilation control through the tankhouse.

Coalescing the bubbles is accomplished by providing a tortuous path with beads or spheres floating on the electrolyte surface in the cell or using a surfactant to allow the bubbles to coalesce at the surface. The two alternatives have been used in combination with some success.

Trapping the acid mist within the cell requires that the cells be covered and the gases extracted from the cell. This can be accomplished by using a seal below the level of the electrode hanger bars to trap the acid mist and extracting the gases through the cell outlet. Units have been built with the seal being made of solid PVC or brushes attached to the anodes with separate covers for each end of the cell. This method has proved successful for keeping the acid mist out of the tankhouse atmosphere, but some mechanical problems remain.

Ventilation control is used in combination with either of the other two methods. The amount of air movement required depends on how much acid mist escapes from the cells.

TANKHOUSE OPERATING ELEMENTS

Operation and Maintenance Philosophy

The degree of automation and materials of construction are heavily dependent on the operation and maintenance philosophy of the plant owner. It is generally a case of trading capital cost for operating and maintenance costs. For instance, utility piping for water and compressed air in the tankhouse can be installed in mild steel with a protective coating which will require maintenance. On the other hand, stainless steel piping will be more expensive to install but will be maintenance free.

The most important factors for a safe, efficient electrowinning operation are training of the operators and good housekeeping. Operator training should include the safety aspects of each task required in the tankhouse, including crane operation, cell cleaning, stripping machine operation and reagent control. The importance of paying attention to the details will result in higher cathode production and cathode quality. For example, detecting shorts or poor electrode contacts on a daily basis will give higher current efficiency, reducing the cost of copper production.

Good housekeeping will reduce cathode contamination for higher cathode quality. Two activities which will also contribute to higher copper quality are careful washing of the cathodes to remove all traces of electrolyte and regular cleaning of the cells to remove the lead sludge.

CONCLUSIONS

The design and operation of a copper electrowinning tankhouse are closely related and the operation and maintenance philosophy must be incorporated in the design. The tradeoffs to consider are between capital costs and operating and maintenance costs. The level of automation and the materials of construction are two good examples of this concept. A higher level of automation will increase the capital cost and require more skilled maintenance but can significantly decrease the number of operators required while providing better control of the plant operation. Corrosion resistant materials will have a longer life and lower maintenance requirements than coated materials but will increase the capital cost.

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Evolution of Cathode Quality at Phelps Dodge Mining Company

Ted Maki[*]

Phelps Dodge Mining Company has gone from zero cathode production in 1984 to 125,700 tonnes (277 million pounds) of questionable quality cathode in 1988 to an estimated 385,600 tonnes (850 million pounds) of quality cathode in 1998. This paper will look at operational changes and initiatives implemented during this period of explosive growth that resulted in significant improvements to cathode quality.

HISTORY

Phelps Dodge made the decision to invest in solution extraction/electrowinning technology with the start-up of facilities in Tyrone in 1984. SX-EW facilities were then brought on line at Morenci in 1987 and Chino in 1988. Several expansions occurred at all three sites over the following years and continue at Morenci today. Cathode copper now comprises approximately 54% of total domestic copper production.

With the initial capital spent on these facilities, the emphasis was rapid return on investments. This translated solely to copper production while cathode quality was not a major concern. While production continues to be a priority, cathode quality is now of significant importance to all operations. Figure 1 outlines production and associated cathode quality. Note that in the early years of production some of the figures are estimates, which further highlights the lack of emphasis on cathode quality at that time.

The focus on cathode quality materialized along with the continuing need to increase operating efficiencies and reduce costs. In most instances, efficiency and cost improvements led to better cathode quality. For example, elimination of short circuits in the tankhouse not only led to better current efficiencies and power savings but also resulted in a better quality cathode.

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[*] Hydrometallurgical Manager, Chino Mines Company, Hurley, N.M.

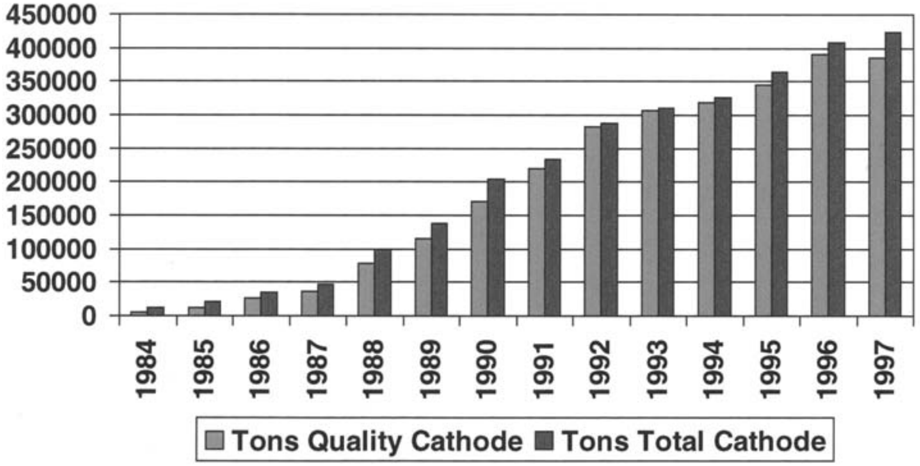

FIGURE 1 PDMC cathode production

Production, particularly at low copper prices, is still the driving business factor. Without maximizing production, the divisor is low and unit costs increase. Phelps Dodge has the benefit of owning its own rod mills and refineries, which turn the lower quality cathode into a high quality salable product. In this case, the refining costs are higher but are overcome by volume of copper produced. Fortunately, in most instances, cathode quality can still be maintained at high production rates.

CATHODE QUALITY ENHANCEMENTS

Cathode quality enhancements have been implemented by focusing on electrolyte quality and chemistry, procedural changes and physical/mechanical improvements. As demands of production often distract operating personnel from concentrating on quality initiatives, one of the key decisions made was to create the position of “quality manager” at all of the sites. The quality manager interprets quality data and makes suggestions for improvement and interfaces with customers, technicians, production managers and quality managers at the sister properties. The quality manager has also taken the lead role in COMEX certification and ISO registration.

Table 1 illustrates enhancements that resulted in a positive impact on cathode quality. Note that these enhancements may not be in place at all of the properties.

RESULTS

Enhancements have provided for more consistent and improved cathode quality. The company has reduced downstream refining costs or does not take pricing penalties for refining of poor quality cathode, employees take more pride in producing quality cathode and costs and efficiencies improve. Internal customers, the Norwich and El Paso rod mills, experience less upsets and reduced blending requirements. Upsets due to poor cathode quality are now rare. Phelps Dodge’s El Paso refinery also benefits as the reduction in poor quality cathode frees space for additional production. External customers have higher confidence in the product received and produce higher quality finished products.

TABLE 1 Cathode quality enhancements

Enhancement	Purpose	Impact on quality
Organic wash stages/coalescers	Remove entrained PLS from organic prior to stripping	Reduced manganese transfer in electrolyte thereby slowing anode corrosion rate and lead contamination
Gunk filter presses	Remove gunk from SX circuit without fear of losing organic	Reduced manganese carryover into electrolyte as above
Temperature control with heat exchangers/boilers	Increase electrolyte temperature	Better plating characteristics, reduced contamination from lead and sulfur
Electrolyte filters/aeration cells	Remove entrained organic from electrolyte	Eliminated staining of cathodes with organic and lead contamination
Stainless steel technology (Morenci)	Reusable mother blanks, reduction in manpower	Improved plating characteristics, reduced lead and sulfur contamination, more visually appealing
Cell level control	Eliminate top rope on cathodes	More visually appealing, reduced lead and sulfur contamination
Guar/thiourea addition	Better plating characteristics	More visually appealing, reduced lead and sulfur contamination
Rectifier shut down and start-up procedures (ramping)	Reduce flux in cells	Reduced lead contamination
Short correction with Gauss meters	Less nodulation and short circuits	More visually appealing, reduced lead and sulfur contamination
Cell cleaning/anode washing schedules	Remove lead from system	Reduced lead contamination
Quality rectifier shutdowns	Reduce lead levels	Reduced lead contamination
Minimum lean electrolyte grades	Improve plating characteristics	More visually appealing, less lead and sulfur contamination
Optimum acid concentration in electrolyte	Reduce sulfur contamination	Reduced sulfur contamination
Visual grading	Downgrade cathode due to appearance	More visually appealing
Replacement of beads with balls and mist agent	Reduce beads embedded in cathode	More visually appealing, reduced contamination
ISO registration	Refined and consistent procedures	More consistent cathode quality
COMEX certification	Certify cathode quality	Customer assurance of quality cathode
Pilot plant	Improve existing process and test for future	Assisted in enhancements above
100% runs at rod mills	Verify sampling and assaying accuracy	Customer assurance of quality and confidence
100% run at wire mill	Prove electrowon cathode can be turned directly into final product	Customer confidence
Tests of electrowon and electrorefined cathode at rod mill	Prove that electrowon cathode is equal to electrorefined cathode	Customer assurance of quality
Installation of PIMS (plant information management system)	Increase control of plant inputs and outputs	More consistent cathode quality
On-line chemical analysis	Increase control of plant variables	More consistent cathode quality
Quality managers hired	Sole purpose to monitor and improve cathode quality and interface with customers	Improved cathode quality and customer relationships

Integrating cathode quality into production was initially a struggle. Today it is a part of business. Efforts continue to improve the quality of cathode produced. Improvements thus far have resulted in increased customer satisfaction, cost savings and increased operating efficiencies.

SECTION 4

Theory and Practice of Solvent Extraction

Co-Chairs

P. Thompsen
Fluor Daniel Inc.

V. Ness
Kilborn SNC Lavalin, Inc.

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Morenci: Making the Most of a World Class Resource

R. Hunter White[*]

Recognized as one of the world's premier copper ore bodies, the Morenci District deposit represents an exceptionally flexible resource with mineralization amenable to processing via conventional concentration and smelting techniques as well as emergent hydrometallurgical technologies. This paper briefly reviews Morenci's progression from traditional sulphide copper ore treatment methods to its greater reliance on SX-EW technology and subsequent emergence as the single largest producer of electrowon copper cathode. Challenges associated with the integration of leaching and sulphide operations are explored as well as the management and operational issues which result when the definition of "ore" must be critically examined.

INTRODUCTION

In open-pit operation since 1937, Phelps Dodge's Morenci Mining District clearly stands out as a world class copper deposit. At the end of 1997, the operation boasted a Phelps Dodge lifetime metal production of 19.7 billion pounds, with an estimated 15.9 billion pounds still remaining in the proven and probable reserves and an additional 16.0 billion pounds contained in other mineral reserves within the district. Currently positioned as the world's number one producer of electrowon copper cathode, Phelps Dodge Morenci, Inc. (PDMI) is well positioned to enter the next century but realizes that the mining and metallurgical processes as well as the management and operating philosophies must continue to evolve as the operation enters its seventh decade.

Perhaps the most significant shift in metallurgical focus to date occurred in 1987 when Morenci began a fundamental transformation from reliance on conventional copper sulphide processes to adoption of newly emerging solution extraction-electrowinning (SX-EW) technology. Although proven successful at other properties, including its sister operation at Tyrone, New Mexico, SX-EW operations represented a departure from classical metallurgical processes and certainly an element of calculated risk. Confident of success, Morenci completed construction of its first SX-EW plants just as a depressed copper market began to recover.

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[*] General Manager—Operations, Phelps Dodge Morenci, Inc., Morenci, Ariz.

Unlike many southwestern copper deposits, the vast majority of Morenci's sulphide resource occurs as chalcocite rather than chalcopyrite, permitting beneficiation by either conventional or hydrometallurgical methods. At an average mining rate exceeding 800,000 short tons per day, the Morenci resource continues to provide sulphide ore to two concentrators as well as oxide and low-grade, leachable sulphide ores, which are deposited via haul trucks on approximately 15 to 20 million square feet of active stockpile area. Virtually no material is mined as waste. Rather, material containing less than 0.1% total copper is used as a base in developing additional stockpile surface area.

Challenges still exist with the integration of leaching and conventional operations, particularly when conducted on such an expansive scale. Tremendous efforts must be made to coordinate mine and stockpile plans with the logistics of haulage and ore quality. Without the presence of a dedicated, skilled and highly motivated workforce, the past, present and future accomplishments at PDMI would not be possible. While investments in emerging technology have provided an avenue for success, profound changes in management philosophy have also been undertaken.

For more than two years, Phelps Dodge has been working to change its culture, allowing all employees to best utilize their talents and unique knowledge, in large part, by the creation of teams and fostering an environment in which teamwork can thrive. In an atmosphere where seemingly continuous expansion is coupled with continuous improvement, teamwork among all operating divisions and levels is crucial to continued success.

HISTORY

In 1865, as the end of the Civil War drew near, volunteer soldiers from the Union Army stationed in California discovered the first mineralized outcroppings in the Morenci Mining District. Initially operated by the Longfellow Copper Company between 1872 and 1882, the ore grade averaged 20% copper and was mined via underground methods. Phelps Dodge entered the Morenci District in 1881, when it purchased an interest in the Detroit Mining Company. Mining in the district continued under the direction of several companies until 1921, at which time Phelps Dodge became the sole operator.

All mining ceased in 1932 and was not reinstated until 1937 when development of the Morenci open pit began. Following an approximate two-year stripping campaign, ore production started in 1939. The onset of U.S. involvement in World War II precipitated construction of the Morenci Concentrator Original Side in 1942 and was rapidly followed by a mirror-image expansion of the facility in 1943. Construction of the Metcalf Concentrator in 1974 again provided an increase in concentrate production, as shown in Figure 1.

Throughout Phelps Dodge's history of operation, a relatively significant quantity of copper was produced by iron precipitation. Large, low-grade stockpiles produced from historic mining operations and stripping of the Morenci pit provided an immense source of inexpensive copper. Indeed, from 1937 to the end of its era in 1989, the precipitation plants yielded nearly 1.1 billion pounds of saleable copper.

Just as these low-grade stockpiles furnished a feed stock for the precipitation plants, they were also the basis for PDMI's entry into the emerging technology of SX-EW. Somewhat surprisingly, Phelps Dodge did not elect to begin its SX-EW production at Morenci. Rather, the first facilities were constructed at Tyrone, New Mexico, in 1984. This initial excursion into a relatively new technology proved so successful that a decision was made to expand EW cathode production to Morenci.

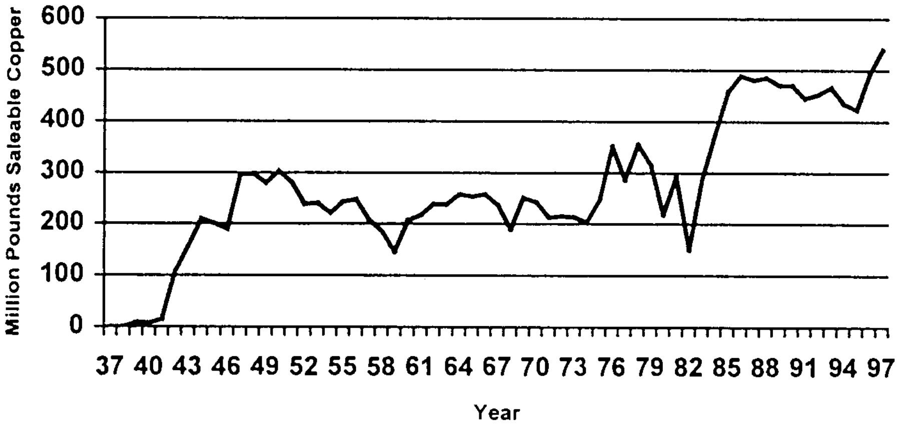

FIGURE 1 **PDMI concentrator production**

PRODUCTION PHILOSOPHY

The evolution from conventional copper sulphide processes to adoption of SX-EW technology requires a fundamental shift in the mindset of a mining operation. A paradigm shift must occur as the long-accepted definition of "ore" must be critically examined and reshaped. Oxide ore, which has historically been considered untreatable, has now become a prime resource. Sulphide materials assaying below concentrator cut-off grade are no longer "a poor substitute for ore" but a newly discovered, recoverable resource.

Although SX-EW operations began at PDMI in 1987, challenges still exist with the integration of leaching and conventional operations. The existing stockpiles allowed rapid entry into SX-EW without disruption of the concentrator operations. As more copper is recovered via hydrometallurgical methods, however, the balancing act required to coordinate mine and stockpile plans with the logistics of haulage and ore quality becomes increasingly more complex. Without the presence of a dedicated, skilled and highly motivated workforce, such accomplishments would not be possible. While investments in emerging technology have provided an avenue for success, profound changes in management philosophy have also been undertaken.

At PDMI, change never ceases. And change is never comfortable. In such a dynamic system, it is crucial that management foster an environment conducive to rapid transition, global optimization and continuous improvement. Teamwork is a key element. In a team environment, individuals develop and work together to achieve growth and operating excellence. But teamwork alone is only a part of the equation. Without frequent, honest and open communication, it is impossible to establish clear goals which are consistent with corporate business objectives, or to develop an environment of trust. Current and past practices must be critically reviewed. Innovations and departures from standard practice must also be encouraged.

Perhaps one of the most important lessons learned, however, is that it is acceptable to say, "I don't know. Can you help me?"

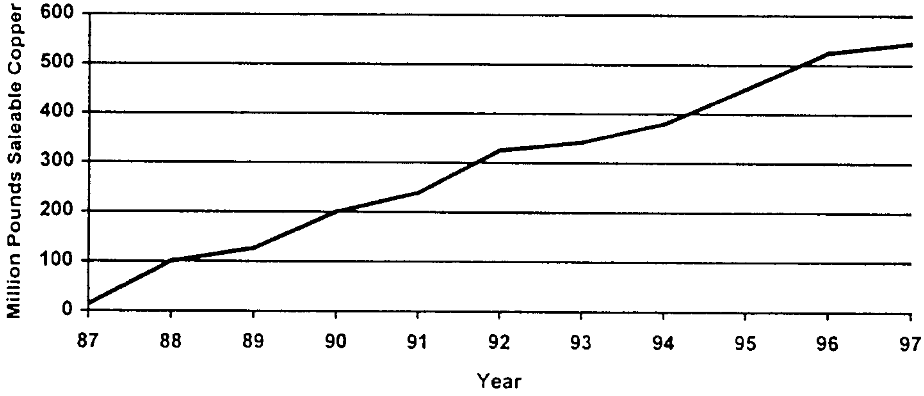

FIGURE 2 PDMI EW production

CONTINUOUS IMPROVEMENT

The hydrometallurgical operations at PDMI clearly represent continuous expansion coupled with continuous improvement. From 1987 to the present, production has risen from 14 million pounds of cathode per annum to more than 540 million, as depicted in Figure 2.

Cementation to SX-EW

From historic times until 1989, copper cementation, or iron precipitation, was used at PDMI as a means of exploiting existing low-grade stockpiles. Smelters were always within close proximity, providing easy access for further treatment of the precipitated copper. From 1937 until closure of the operation in 1989, approximately 1.1 billion pounds of saleable copper was produced. As demonstrated in Figure 3, the contribution of this operation was substantial in certain years.

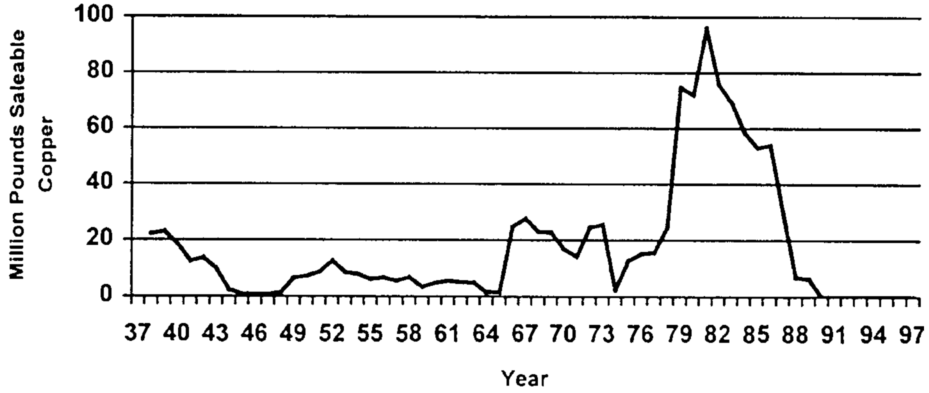

FIGURE 3 PDMI copper cementation plant production

A severe downturn in the copper market during the mid-1980s demanded that less expensive methods of production be considered. The early SX-EW venture at Tyrone set the stage for adoption of the emergent technology at Morenci. In 1986, the decision was made to begin construction of what is now the largest hydrometallurgical operation in the world. While metal production through copper precipitation continued through 1989, closure of the Morenci smelter in December 1984 combined with increased cathode production brought a close to this era.

Evolution of Plants I through IV

The PDMI hydrometallurgical division has undergone four expansions since its inception. Because the mineralogy within the district will permit beneficiation via conventional concentration and smelting or hydrometallurgy, PDMI has been able to rapidly advance the production of low-cost, high-quality copper cathode.

The Beginning. In September 1987, PDMI's first three SX plants—Metcalf, Southwest and Central—were commissioned. A single tankhouse capable of producing 100 million pounds of cathode per annum was also constructed at the Central SX facility. Designed to accept solutions not only from the resident plant, Central EW receives solutions from the two satellite installations as well. Selection of the satellite locations was strategic as well as dictated by topography. As the Morenci District is located in reasonably rugged terrain, areas which do not require significant civil site preparation are scarce. Additionally, the historic stockpile systems were situated in distinctly, and farflung, locations. Rather than pump pregnant leach solution (PLS) over long distances, it was decided to transfer a significantly lower volume of rich electrolyte (RE).

Located seven miles to the northwest from the Central SX-EW in close proximity to the old Metcalf mine and townsite, Metcalf was designed for a 12,000 gallon per minute (gpm) flow capacity. Previously unleached stockpiles located in the northern portion of the district provided a readily available feed stock. This installation was originally designed as two separate trains, running in standard 2 × 1 series configuration.

As its name implies, the Southwest facility was positioned approximately five miles to the southwest of the Central facility. Also constructed for a 12,000 gpm flow, this plant obtained its feedstock from stockpiles located nearby which had previously been used in the production of cementation copper. The plant design was a single train, however, employing a then innovative, state-of-the-art Krebs technology.

The Central SX was constructed as a single train in a conventional 2 × 1 configuration, with a design flow of 6,000 gpm. Located adjacent to the Central tankhouse, PLS was obtained from the Lower Chase Creek stockpile system. The Central tankhouse, as previously mentioned, was capable of producing cathode at a rate of 100 million pounds per annum. The original plant consisted of two sections, containing a total of 188 cells, and was designed for starter sheet technology.

Expansion I. In February 1989, Expansion I was completed, doubling production capacity to 200 million annual pounds. The Central electrowinning tankhouse capacity was expanded through the addition of 188 cells which brought the total to 376. At this time automated cathode handling and starter sheet fabrication equipment were also added to the Central tankhouse. The Central SX plant flow capacity was expanded via addition of a parallel circuit, yielding a 1 × 2 × 1 configuration. PLS to support this expansion was obtained solely from increasing the surface area under leach.

Expansion II. Expansion II was based on a mine-for-leach ore body, the Northwest Extension, which proved to contain significant quantities of oxide and mixed oxide-sulphide material when opened. This expansion was conducted in three separate phases: (1) addition of 172 cells to the Central tankhouse bringing the total to 548, (2) upgrade of the solution transfer system, and (3) construction of the Modoc SX plant. Modoc, located approximately three miles west of the Central facility, was constructed in a manner similar to Metcalf with a design flowrate of 12,000 gpm. This 1992 expansion provided a design production increase to 348 million annual pounds.

Expansion III: Southside. Although mining in the Morenci Pit was discontinued in early 1995, a rim of mineralization still existed along the southern border, outside the area previously designated for concentrator suitable material. The Southside ore body, combined with a higher mining rate in the Metcalf pit, prompted PDMI's third expansion. In July 1995, a second tankhouse capable of plating 152 million annual pounds was commissioned. Dubbed Southside, the tankhouse took advantage of the newer ISA stainless steel mother blank technology.

Now at a design plating capacity of 500 million pounds, it was also necessary to build additional raffinate pumping capacity as well as double the flow capacity of the Metcalf and Modoc SX plants. Parallel extractors were added to each of the four trains, thus providing a 1 × 2 × 1 circuit configuration and doubling the flow capability of each plant.

Expansion of the Metcalf and Modoc SX plants permitted closure of the Southwest SX facility. Originally designed to provide approximately five to six years of service, the plant had reached the end of its useful service life. The stockpiles used as its feed source had been previously leached for cementation and as mining had moved to more northerly regions of the district, they no longer received new ore because of adverse haulage requirements.

Expansion IV: Southwest. In 1997, work was initiated on a fourth, two-phase expansion to increase SX-EW production capacity to a design rate of 550 million annual pounds.

A total rebuild of the Southwest SX facility, using state-of-the-art mixing technology and plant design was initiated in late 1997. Commissioned in June 1998, the new "Son of Southwest" has incorporated Outokumpu VSF mixing technology, a loaded organic wash stage and a gunk filter system. Design flow capacity is 6,000 gpm. At present, the plant is operating at 6,500 gpm and has been tested at 10,000 gpm. PLS is obtained from the existing Southwest stockpile system. In late 1998, the new Stargo Canyon stockpile system will be initiated, increasing the PLS tenor and subsequent production. Although currently limited to one train, the Southwest facility has the potential to expand to multiple trains.

Conversion of the Central tankhouse from copper starter sheet technology to ISA stainless steel mother blank technology is ongoing at the time of this writing. The original two sections containing 188 cells are slated for completion in late November 1998. The three expansion sections are scheduled to be completed in March 1999.

COMEX and ISO Certification. In November 1995, barely four months after commissioning of the Southside tankhouse, PDMI began the process to obtain COMEX certification for this cathode. Listed as PDSS, certification was obtained on April 23, 1998.

Quality assurance is also critical not only to the production but also the marketing of EW cathode. In September 1997, the PDMI EW department obtained ISO 9003 certification. A second audit, this time for certification of ISO 9002, is slated for September 1998.

Advances in Stockpile Construction and Management. Without improvements in stockpile construction and management, sustaining copper production would be difficult, if not impossible. Originally constructed as monolithic structures, stockpiles are now built in low lifts, 20 to 30 feet in height. Premium grade leach ore, assaying greater than 0.3% total copper, is placed in lower lifts to provide conditions more conducive to faster and more complete copper extraction.

Solution is currently applied to all stockpiles using drip emitters. Full-scale tests conducted on run-of-mine ore by the Phelps Dodge Process Technology Center have strongly implied that drip emitters provide substantially better wetting than sprinklers. Again, this application method creates a condition in which the ore can be more thoroughly wetted and increases the probability of improved copper recovery.

Although not fully exploited, implementation of extended rest cycles on some sulphide stockpiles has proved efficacious. It is surmised that by allowing a stockpile system to rest, the resident bacteria are provided an opportunity to “regroup” and continue the leaching process.

HYDROMETALLURGICAL OPERATIONS SUPPORT

Investing in Our People

Without question, the most important resource at Morenci is its people. As found in most companies, PDMI operated with a classic, hierarchical structure until early 1993 when concepts of participative management were introduced. Basic principles were embraced and project teams formed, slowly at first and then with increasing enthusiasm. In early April 1997, a salary conversion plan was presented to all employees, with an ultimate goal of evolving to a completely salaried workforce by January 1, 2000. PDMI is still firmly committed to achieving this goal and has formed an oversight group to assist, support and certify the teaming effort.

Guidelines were established early in the conversion process to define expectations and provide a level playing field. Eleven criteria were developed and must be met before a team can achieve salaried status. Additionally, all teams are required to develop goals that are written, measurable and within the team’s control. These goals must also be in line with respective division goals and be an improvement over past performance. Each team must demonstrate the ability to perform together for at least six months prior to petitioning the oversight group for salaried status.

It is essential that teams seeking conversion from hourly to salaried status develop an attitude and ethic espousing high performance and promoting the best interest of PDMI and Phelps Dodge Mining Company as a whole. The teams’ efforts must be channeled in the most efficient manner to promote safety, ensure quality, reduce costs, increase productivity and enhance long-term job security. By meeting these requirements, teams will create a workforce that will meet PDMI’s business objectives and strengthen its position as an industry leader.

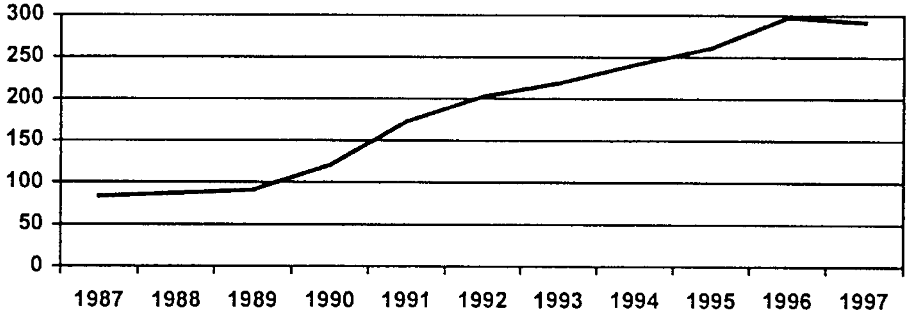

FIGURE 4 **PDMI mining rate, million short tons per annum**

While there have been challenges and difficulties experienced in this very fundamental alteration of its organizational structure, PDMI and its workforce have realized many benefits. One of the most striking examples of teamwork began on December 23, 1997, with the collapse of a 185-foot section of PDMI's overland conveyor that transports ore from the mine to both the Morenci and Metcalf concentrators. In just thirty days, a new reinforced conveyor truss was designed, fabricated and installed. To accomplish the rapid reconstruction, however, employees cancelled holiday vacations, rearranged their schedules and worked 3,400 hours of overtime. Without a team process already in place, it is doubtful that this stunning accomplishment could have been achieved.

Mining Effort

Clearly, the Hydrometallurgical Division did not solely achieve the rapid production increases experienced at PDMI. Without the cooperation and support of the Mine Division, the dramatic production expansion via SX-EW realized since 1987 would not have been possible. As shown in Figure 4, the mining rate also increased significantly to support expansion of the stockpile system. Indeed, during the time period considered in Figure 4, the mine achieved a world record production of 1.3 million short tons of combined sulphide and leach ores in one twenty-four hour day.

Achieving and maintaining superior performance is an ongoing challenge, however. During its nearly seventy-year history, the mining effort and methods at PDMI have undergone significant change. Mine production originally relied on rail haulage, with the limited use of small haul trucks for development mining. Improvements in haul truck design combined with an expanding and ever-deepening pit and increased demand for leach stockpile construction dictated that alternative haulage methods again be examined. Conversion from ore rail haulage to a fleet of 170- and 190-ton trucks was completed in 1989. The mine currently utilizes a fleet of 85 haul trucks consisting of 240-, 255-, and 320-ton capacity units.

Referring again to Figure 4, it must be noted that mine production peaked in 1996 and then declined slightly. It soon became apparent that the organizational structure, work processes and support systems had not kept pace with the Mine Division's rapid growth. Indeed, this situation was perceived to represent a substantial risk, potentially jeopardizing PDMI's ability to easily survive the current downturn in copper prices and successfully compete in the global market. In May 1998, PDMI initiated a Mine Excellence Program to address and resolve these critical issues.

A cross-functional team of about thirty individuals, including operators, mechanics and professional staff was formed and commissioned to undertake a four-month evaluation. Through concerted effort, this team built upon past successes to develop a plan to achieve superior internal and external customer service and dramatically reduce unit costs. It is important to note that the creation of a new vision was not isolated solely to these individuals. Indeed, in a three-day conference held in June 1998, approximately 200 individuals met to discuss perceptions, challenges and expectations. Members of the surrounding community, vendors and employees were provided an opportunity to openly express their concerns and hopes for the future of PDMI.

The study's findings were presented to PDMI management and employees in late August 1998, with implementation initiated in September of the same year. At the time of this writing, it is not possible to comment on the evaluation's impact. It is clear, however, that the recommendations were distilled from the ideas of many. It is also evident from a declaration in the team's vision statement that a certain degree of success has already been achieved: *We make a difference.*

CURRENT STATUS

As previously noted, PDMI now operates four SX plants and two EW tankhouses. A sampling of key operating statistics and production records are presented in Tables 1 and 2.

TABLE 1 **PDMI hydrometallurgical division average operating statistics**

Million square feet under leach	15.5
Application rate, sulphide stockpiles—gpm/ft^2	0.0025
Application rate, oxide stockpiles—gpm/ft^2	0.0040
PLS flow, gallons per minute	61,500
PLS tenor, grams copper per liter	2.60
Rich electrolyte flow, gallons per minute	9,500–10,000
Current density, amperes per square foot	30–32

TABLE 2 **PDMI hydrometallurgical division records**

Achievement	Record	Date
Copper extracted through SX (lbs. per day—all plants)	2,035,423	Sep. 25, 1997
Copper extracted through SX (lbs. per day—single plant)	945,390	Sep. 25, 1997
Copper plated through EW (lbs. per day—all plants)	1,754,990	Sep. 24, 1997
Copper plated through EW (lbs. per day—single plant)	1,206,992	Sep. 24, 1997
Copper plated through EW (lbs. per month—all plants)	50,249,696	July 1997
Copper plated through EW (lbs. per month—single plant)	34,701,642	July 1997

FUTURE OPPORTUNITIES AND CHALLENGES

In an increasingly competitive world market, PDMI is well positioned to meet the challenges of the next century. The Morenci deposit, containing over 15.9 billion pounds of copper in proven and probable reserves, remains one of the world's most impressive resources. Opportunities and advances gained from tapping into the knowledge and initiative of the workforce through participative management and reliance on teamwork will only enhance the viability and vitality of the operation.

At present, the concentrator and hydrometallurgical divisions produce approximately equal proportions of PDMI's 1.1 billion pound per annum copper production. It is quite possible, however, that this balance may shift more heavily to cathode than concentrate within the next few years. Improvements in stockpile construction, solution application and other leach practices will allow more complete and efficient copper extraction, providing an even greater source of low cost metal.

While common practice in South America, the crushing of leach ore followed by stockpile construction with conveyors and stackers in still uncommon in the North American copper industry. A study initiated at Morenci in 1997 strongly implies that reducing leach material from run of mine to minus one-half inch in size may provide copper extractions equaling those obtained from concentrating and smelting, at a significantly lower cost. The potential for transfiguring PDMI into a solely mine-for-leach facility cannot be discounted.

The commitment to teams and to transforming the way employees work together will continue to move forward despite the depressed copper market. Phelps Dodge believes that the investments in teamwork, while not capital intensive, can yield substantial returns and are critical to weathering the downturn in the market and to positioning Phelps Dodge for future success. The commitment to teamwork speaks to the confidence of the people at PDMI as well as all other locations and to their ability to provide a competitive advantage.

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Practical Aspects of Copper Solvent Extraction from Acidic Leach Liquors

John R. Spence* **and Matthew D. Soderstrom†**

Copper solvent extraction from acidic leach liquors has been used for approximately 30 years. Since the first commercial operation at Ranchers Bluebird in 1969, some 15 to 20% of the world's copper production is now generated by leach, solvent extraction, and electrowinning. As new equipment is introduced and as each operation encounters new challenges with variation in ore and operating parameters, optimization of the technology still continues. This paper will describe fundamental, practical issues necessary to optimize both the metallurgical and physical parameters of the solvent extraction circuit.

INTRODUCTION

There is a substantial volume of literature in relation to the use of solvent extraction for the recovery of copper. Much of this data covers technical aspects, selection of the correct reagent, design improvements, etc., but there is little practical information on how to operate and optimize a copper solvent extraction circuit. The aim of this paper is to try and redress this.

The authors would like to thank David Readett, Straits Resources, for his contribution to this paper.

HISTORICAL

Copper solvent extraction from acidic leach liquors has had a short yet hectic history going back only 30 years. Since its introduction and the initial resistance of many copper producers, its acceptance as an industry standard has been outstanding. Since the first commercial operation at Ranchers Bluebird in 1969, world production has increased exponentially.

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* Technical Supervisor, Zeneca Specialties, Acorga Metal Extraction Products, Phoenix, Ariz.

† Project Manager, Zeneca Specialties, Acorga Metal Extraction Products, Blackley, Manchester, United Kingdom.

There is now some 15 to 20% of the world's primary production derived from SX-EW. 1996 saw the commissioning of the largest single greenfield SX-EW plant at El Abra, which has the capacity to produce 225,000 tpa of LME grade A copper.

CHEMISTRY

Fortunately, for the metallurgist, the basic chemistry for copper solvent extraction is summarized by one simple equation which needs to be reversed depending on whether the extraction or stripping stage is being described.

Extraction

$$Cu^{++}_{(aq)} + 2RH_{(org)} \rightarrow R_2Cu_{(org)} + 2H^+_{(aq)}$$

where:
$Cu^{++}_{(aq)}$ is copper in PLS solution
$2RH_{(org)}$ is the extractant, i.e., stripped organic
$R_2Cu_{(org)}$ is the copper/extractant complex, i.e., loaded organic
$2H^+_{(aq)}$ is acid in raffinate solution

Extraction of copper occurs when pregnant leach solution PLS containing copper is mixed with stripped organic solution containing the extractant. The extractant releases its protons and coordinates with the copper so that the amount of copper in the aqueous solution decreases, the acid level increases and the copper is transferred to the organic phase as the copper/extractant complex.

Stripping

The stripping stage for the solvent extraction of copper can be represented by:

$$R_2Cu_{(org)} + 2H^+_{(aq)} \rightarrow Cu^{++}_{(aq)} + 2RH_{(org)}$$

Where:
$R_2Cu_{(org)}$ is the copper complex, i.e., loaded organic
$2H^+_{(aq)}$ is the acid in the spent electrolyte solution
$Cu^{++}_{(aq)}$ is the copper in the advance electrolyte solution
$2RH_{(org)}$ is the stripped copper complex

Stripping of copper occurs when a strongly acidic solution (e.g., spent electrolyte at 180 gpl acid) is mixed with the loaded organic copper complex. The complex releases its copper and takes on acid so that the level of copper in the aqueous solution (electrolyte) increases and the acid level decreases as copper transfers out of the organic phase and is replaced by acid.

Based on the combination of extraction and stripping, a typical modern copper SX flowsheet will consist of two extraction stages and a single stripping stage.

The Acorga products are synthetic organic chemicals specifically built to be highly selective for copper over the other metal ions in the PLS. By basically only extracting copper and then allowing the loaded organic to separate from the aqueous phase (raffinate), the purification stage of solvent extraction has been achieved. The organic is insoluble in aqueous so the phases should always separate. Since the organic is lighter than the PLS (i.e., approximately 0.85 kg/l cf. 1.10 kg/l) the organic will always float on top of the

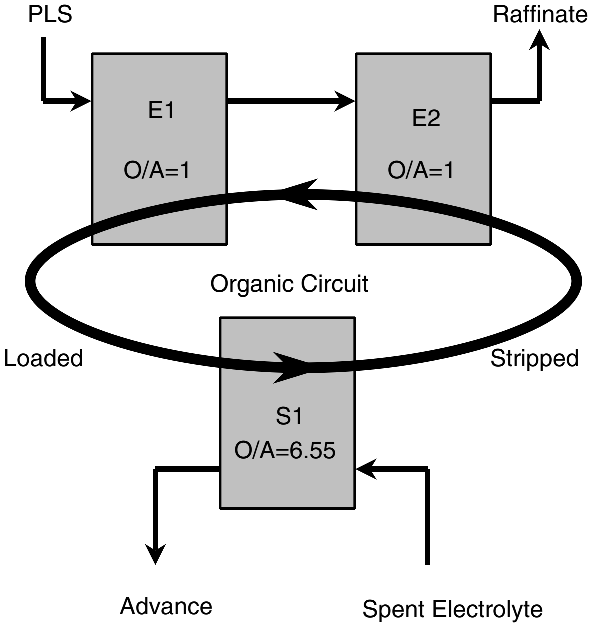


FIGURE 1 Typical SX flowsheet

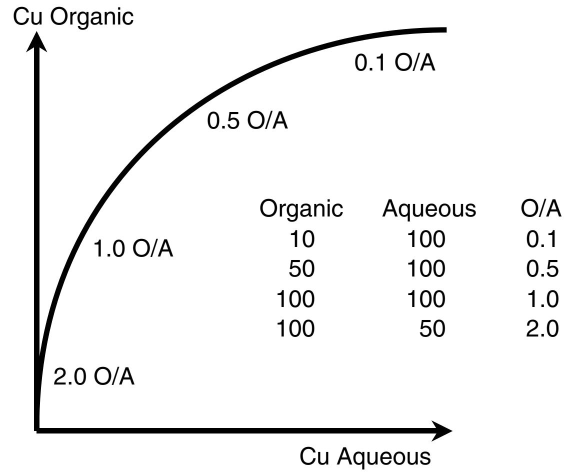
aqueous. To assist in separation from the aqueous phase, Acorga products are diluted with a hydrocarbon-based solvent. This is an inert kerosene-based hydrocarbon which simply acts as a carrier medium for the extractant. This carrier medium is normally called diluent. A typical flowsheet is shown in Figure 1.

Isotherms

Optimization of a typical flowsheet can be described as maximizing the copper transfer in the organic phase from PLS to the electrolyte, ensuring that the absolute minimum of other unwanted metal ions are transferred. Because of the nature of the continuously recycling organic flow, it is necessary to optimize both the extraction and stripping sections.

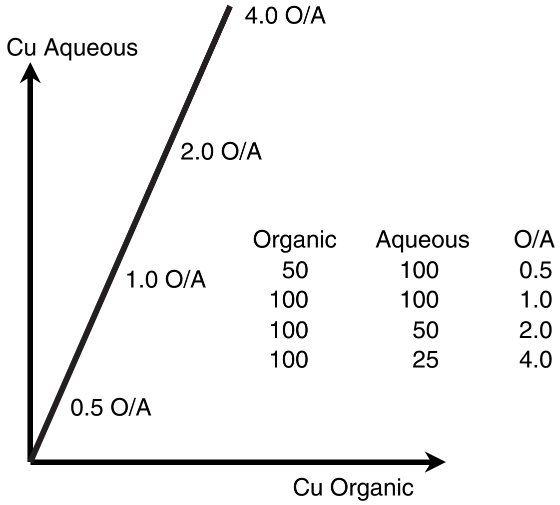
In order to analyze the performance of a solvent extraction circuit and then conduct optimization it is necessary to determine the equilibrium isotherms for the specific system in question.

Extraction Isotherm. The extraction isotherm defines the maximum amount of copper which may be removed from the PLS for each organic to aqueous volumetric ratio (O/A ratio). Plant organic (stripped of copper) is mixed with PLS at various O/A ratios. The organic and aqueous are separated and the copper concentration in each analyzed. The data is then graphed, organic copper concentration on the Y axis and aqueous copper concentration on the X axis. A typical isotherm is illustrated in Figure 2.



Organic	Aqueous	O/A
10	100	0.1
50	100	0.5
100	100	1.0
100	50	2.0

FIGURE 2 Extract isotherm



Organic	Aqueous	O/A
50	100	0.5
100	100	1.0
100	50	2.0
100	25	4.0

FIGURE 3 Strip isotherm

Strip Isotherm. The strip isotherm defines the maximum amount of copper which may be removed from the organic for each O/A ratio. Plant organic (loaded with copper) is mixed with lean electrolyte at various O/A ratios. The organic and aqueous are separated and the copper concentration in each phase is analyzed. The data is graphed, aqueous on the Y axis and organic on the X axis. A typical isotherm is given in Figure 3.

McCabe Thiele Diagrams

Based on the isotherms generated, McCabe Thiele diagrams can then be developed using the specific circuit configuration and parameters. The strip O/A ratio is equal to the

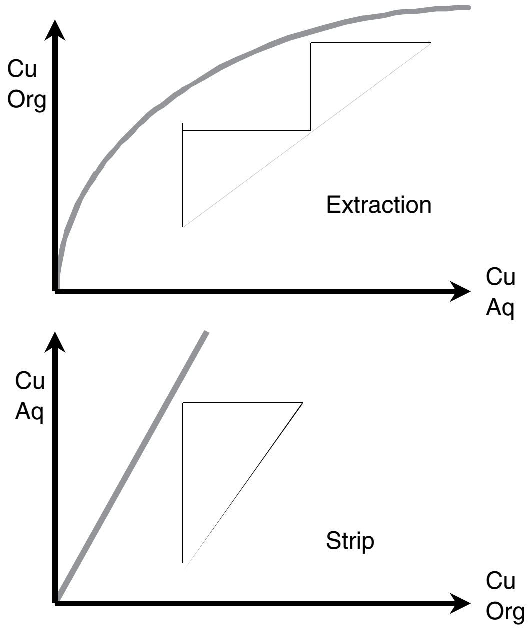


FIGURE 4 McCabe Thiele diagrams

slope of the operating line while in extract 1/m is equal to the operating line. The extraction (two stages) and stripping (one stage) diagrams are shown in Figure 6 and combining the results generates a circuit mass balance (Figure 4).

OPTIMIZATION

Extraction

The optimization of extraction relies on the adjustment of factors which increase the copper transfer from the aqueous phase to the organic phase. The chemical equilibrium equation for the extraction part of the process (displayed earlier) provides the clues to the major variables:

1. PLS copper mass flow (i.e., concentration and flow)
2. PLS acid mass flow
3. Stripped organic mass flow
4. Approach to equilibrium in the mixing stage (practically this means stage efficiency)

In a normal operating environment the PLS concentration is difficult to control and flow is normally kept at a maximum. Based on this, the parameters that can be altered are then:

- **(a)** Increase the extractant concentration (v/o). This raises the extract isotherm. At equivalent flowrates, the PLS is contacted with more extractant due to the concentration increase. As a result, more copper can be transferred into the organic. Figure 6 illustrates this point.

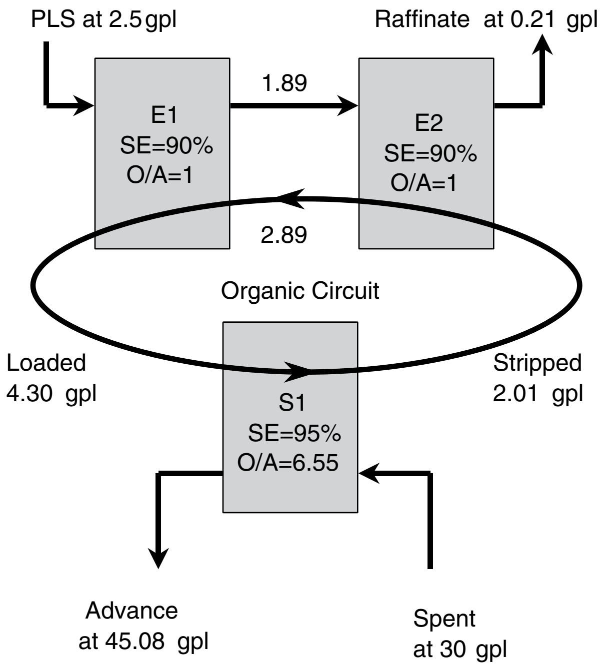


FIGURE 5 **Typical SX flowsheet**

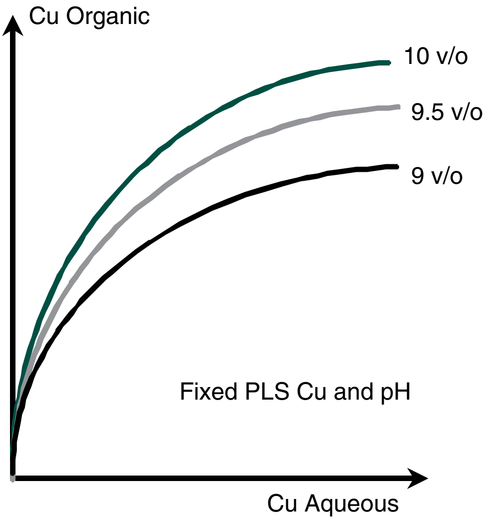


FIGURE 6 **Extract isotherm—effect of reagent concentration**

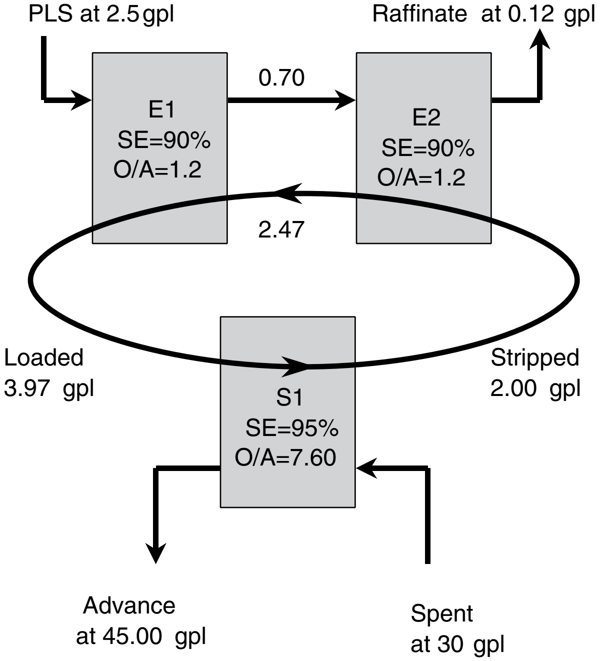


FIGURE 7 Extraction O/A increased to 1.2

(b) Increase the stripped organic flow, i.e., increase the ratio of organic to aqueous thereby changing the slope of the McCabe Thiele operating line. This achieves the same outcome as (a) in allowing the PLS to contact more extractant and hence increase the copper transfer. When the PLS copper concentration is variable and there is a need to optimize copper extraction, then it is preferable to adjust the organic flow rather than change the organic concentration. Increasing the extractant concentration incurs significant cost compared to the cost associated with increasing organic flow. Also decreasing extractant concentration (via dilution) can be limited by circuit storage capacity and running a circuit with excess extractant can lead to excessive chemical transfer of iron. Figure 7 illustrates the effect of changing O/A ratio from 1/1 (Figure 5) to 1.2/1. (Note the increase in copper recovery.) N.B. Increasing organic flow must be batched by increasing spent electrolyte flow in S1 to maintain the same metallurgical balance.

(c) Decrease the PLS acidity. By decreasing the PLS acidity, the equilibrium extraction reaction is shifted to the right resulting in an increased copper extraction potential. Figure 8 shows the effect of decreasing PLS pH (increasing acidity) at a fixed reagent and copper concentration.

(d) Increase stage efficiency (Figure 9). The stage efficiency is an indicator of the physical operation of a mixer. It can be affected by reaction kinetics (this can be temperature related) or surfactants (introduced or generated via manganese degradation of the extractant). The vast majority of effects are physical or mechanical. Typical stage efficiencies for multistage mixers are 90–95% and in strip 95–100%. For a given mixer residence time, the more staging a mixer has, the higher

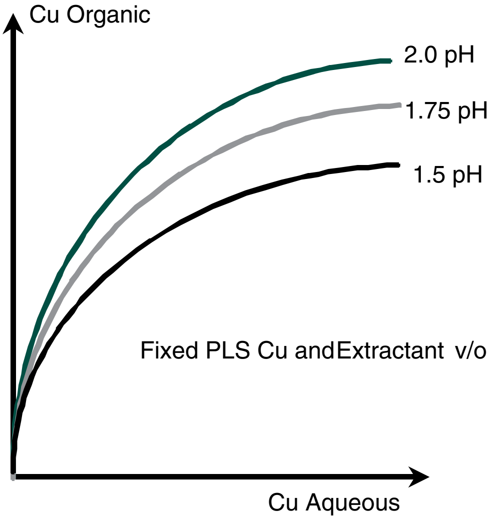


FIGURE 8 **Extract isotherm—effect of PLS pH (acid concentration)**

and more consistent its stage efficiency is. Single stage mixers have lower efficiencies and are more prone to fluctuation. If dispersion is present in the settler, the continuing chemical transfer in the settler may mask poor stage efficiencies. Up to 20% of the mass transfer can take place in the settler but only under conditions of poor mixer function.

Stripping

The optimization of stripping relies on the adjustment of factors which increase the copper transfer from the organic phase to the aqueous electrolyte phase. In operational terms the strip is optimized to produce the lowest copper loading in the stripped organic. The simplified chemical equilibrium equation illustrated earlier again provides the clues to the major variables:

1. Loaded organic extractant mass flow
2. Electrolyte copper mass flow
3. Electrolyte acid mass flow
4. Approach to equilibrium in the mixing stage (stage efficiency)

In a normal operating environment the following changes can be made:

1. Decrease O/A ratio. This decrease can be achieved by increasing the electrolyte (aqueous) flow rate. A decrease in the organic flow rate can achieve the same outcome; however, although the stripping will be more efficient, the total mass transfer of copper to electrolyte will be reduced. The decreased O/A ratio has the effect

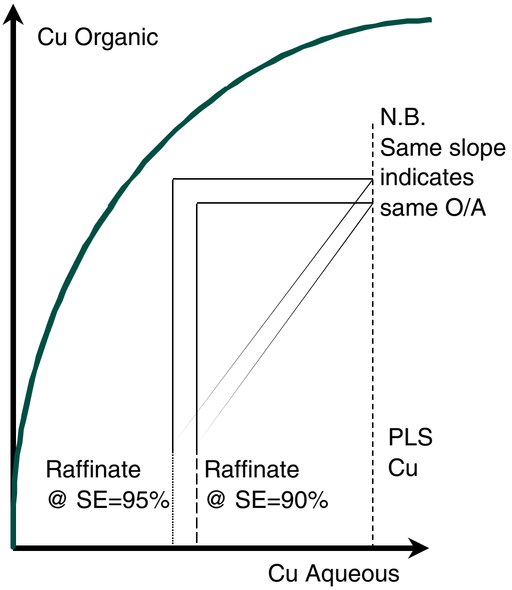


FIGURE 9 **Extract isotherm—effect of increased stage efficiency**

of changing the McCabe Thiele as the organic is contacted with more acid resulting in improved stripping. The stripped organic (copper) is lowered by allowing the McCabe Thiele further "under" the isotherm. This is highlighted in Figure 10.

2. Decrease copper concentration (in spent with the same electrolyte delta copper). As there is less copper to compete against the acid in the stripping equilibrium reaction, the equilibrium shifts to the right thereby increasing the stripping efficiency. The strip isotherm shifts down and to the left as the spent electrolyte copper decreases. At the same delta copper the McCabe Thiele slides further under and to the left of the isotherm, lowering the stripped organic concentration. Figure 11 illustrates the effect of copper concentration variance on strip isotherm.
3. Increase acid concentration in spent electrolyte. The strip isotherm shifts further to the left decreasing the allowable stripped organic. The organic transfer capacity is increased again as a consequence in a shift of the stripping equilibrium. This approach can be obviously sensitive to fluctuation, and the benefit does decrease with increasing acid concentration. One should also bear in mind that acid strength has a direct effect on reagent degradation so there is a balance here. The effect of acid concentration on the strip isotherms is shown in Figure 12.
4. Increase stage efficiency. As illustrated previously (Figure 11), an overall increase in recovery will result from higher stage efficiency.

Circuit

By combining the optimum extraction and stripping conditions, it is then possible to ensure a fully optimized circuit for the plant in question. Each of the optimization factors discussed cause differing degrees of process improvement but taken cumulatively can

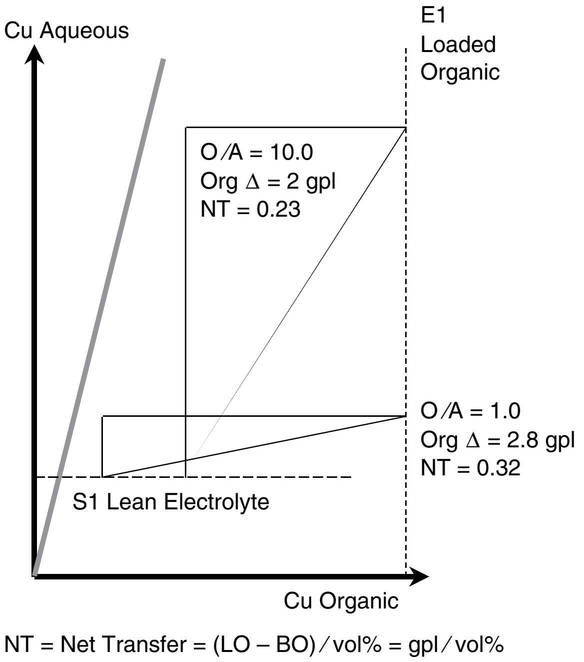


FIGURE 10 Strip isotherm—effect of decreasing O/A on organic transfer

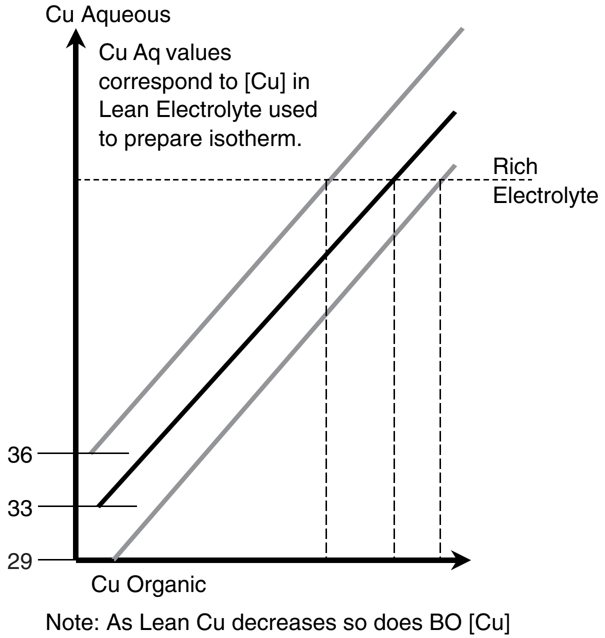


FIGURE 11 Strip isotherm—effect of differing lean Cu concentration

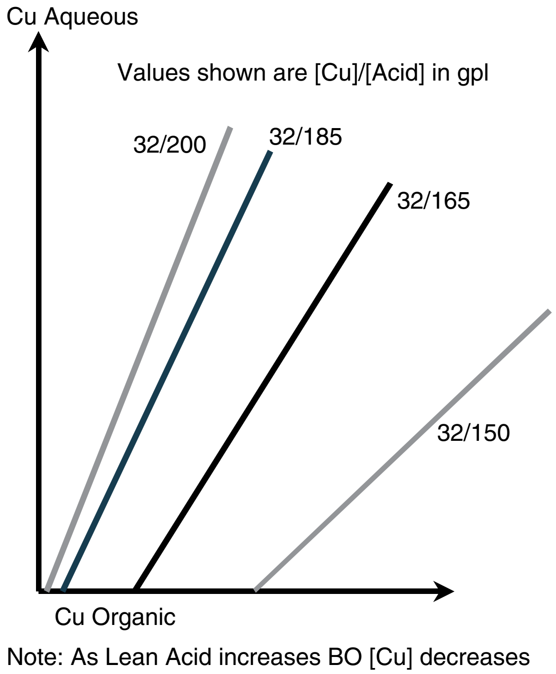


FIGURE 12 **Comparative strip isotherms—effect of differing lean acid concentration**

result in drastic improvement in overall production. A summary of factors which maximize the organic transfer of copper, and hence copper production, is shown in Figure 13.

Mass Balance

This refers to an "accounting process" much like balancing a checkbook. Money in must equal money out. If money in and out are not equal then there may be an unrecorded loss or gain, or the total dollars will increase or decrease. A mass balance substitutes a metal concentration or flow for dollars. This is the mass balance for SX. There are also balances for leaching and for electrowinning which must all equalize for the solution concentrations to be stable.

Assuming the circuit shown in Figure 14 and a PLS flow of 500 m^3/hr the extortion mass balance is:

Copper transfer from PLS = PLS flow × (PLS [Cu])
= 500 × (2.5 – 0.12)
= 1,189 kg/hr

Copper transfer to organic = Organic flow × [loaded organic (Cu) – stripped organic (Cu)]

(NB Organic flow = PLS flow × O/A) = 500 × (4.3 – 1.92)
= 1,189 kg/hr

In a plant situation, a third check on the mass balance is based on acid concentration as for every gram of Cu extracted 1.54 grams of acid is generated.

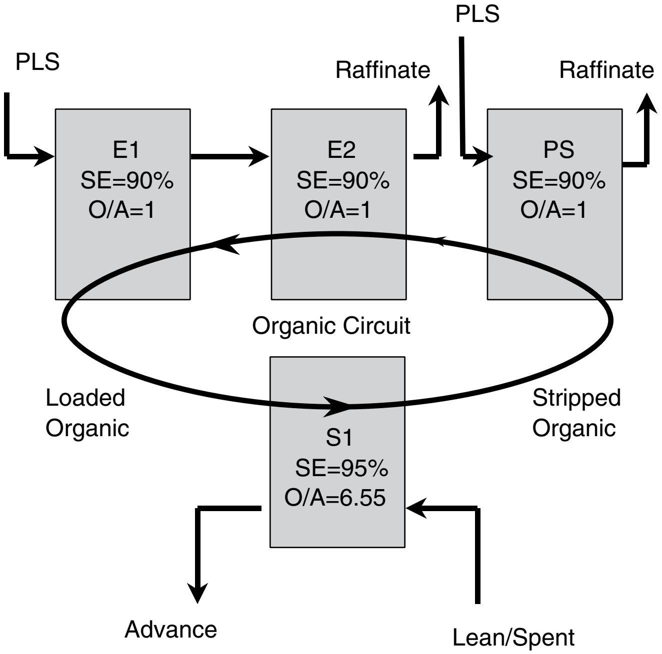
Extraction

- Increase Stage Efficiency
- Increase Extractant v/o
- Decrease O/A Ratio by increased PLS Flow
- Decrease PLS Acidity

Strip

- Increase Stage Efficiency
  - Use recycle for good mixing
- Decrease O/A Ratio
  - Increase Cu in Lean with same Advance
  - Decrease Cu in Lean with same delta Cu
- Increase Lean Electrolyte Acid/Cu Ratio

FIGURE 13 **Factors promoting maximum organic transfer**



Note extractant concentration is approximately double a 2+1

FIGURE 14 **Typical series parallel SX flowsheet**

Acid transfer to raffinate = PLS flow [raff (acid) – PLS (acid)].

Copper transfer from PLS + acid transfer to raffinate/1.54.

A similar balance can be conducted for the stripping stage.

Copper transfer to electrolyte = electrolyte flow [advance electrolyte (Cu) – spent electrolyte (Cu)]

(NB electrolyte flow = organic flow – O/A) = 79.2 × (45 – 30)
= 1,189 kg/hr

This balance can also be checked by the change in acid concentration in electrolyte. Mass balance of Fe transfer to S1 via entrainment and chemical means are also very useful to demonstrate the relative significance of each. The mass balance can also be an indicator of mechanical problems such as holes or leaks in the E1 organic launder.

CIRCUIT CONFIGURATION

Series Parallel

In a number of operations around the world, there has been a need to increase production over design or maintain production with a PLS copper concentration lower than the initial design (e.g., Tyrone, Girilambone). In both circumstances, it has been found a cost-effective option to install an additional mixer settler(s) or to convert from an existing 2 + 2 configuration and change the circuit from series to series parallel.

One of the first companies to take advantage of this form of plant layout in order to expand production at lower cost was Phelps Dodge Corporation at their Tyrone plant in Arizona. The plant started production in 1984 with a production capacity of 39,000 tpa copper. During 1988, Tyrone was able to produce 49,000 tpa copper by expanding the existing plant through addition of two new mixer settler units, one added to each stream of the existing plant. The original plant treated 1,820 m^3/hr of dump leach solution. By adding the extra mixer settler units to each stream, the throughput was increased to 3,640 m^3/hr. The extra flowrates to the dumps reduced the copper tenor in solution, but production was nevertheless increased by an extra 26% at low cost. Others have added the series parallel option to expand production, such as Cyprus Miami and Phelps Dodge at Morenci.

In summary, series parallel offers a way of increasing production at low cost. Recovery is normally reduced, production increased and reagent concentration requirement is higher in the circuit. Figure 14 illustrates a series parallel configuration and is an example of increasing production with the same copper tenor.

Assuming a PLS flow of 500 m^3/hr to extraction stage 1 (E–1) and 5,000 m^3/hr to extraction stage 2 (E–2), then the increase in production achieved over a conventional circuit at 500 m^3/hr is highlighted in the following calculation:

Series parallel production = 500 (2.5 – 0.47) + 500 (2.5 – 0.05)
= 2,240 kg/hr Cu

Conventional circuit = 500 (2.5 – 0.12)
= 1,189 kg/hr Cu

Increased production $= \frac{2,240}{1,189} = 188\%$

Wash Stage

Physical Impurity Transfer. The major source of impurity transfer from the extraction stage is via aqueous entrainment in the loaded organic discharging from E1. Even running this stage aqueous continuous is often insufficient to control the impurity transfer especially when the impurity levels in the PLS are very high (i.e., chloride above 100 ppm or iron above 5 gpl). In these cases, it is common to add a wash stage between extraction and stripping.

In the wash stage, the loaded organic is mixed with clean water, acidified with electrolyte bleed. Acidification is necessary for good phase separation. This mixing allows for the entrained aqueous in the organic to be washed out and significantly diluted (typically by 10–1,000 times). If any aqueous entrainment leaves the wash stage, the levels of impurities are such that they will have a negligible effect on the rest of the circuit.

An alternative to this is the use of a loaded organic coalescer arrangement. Different types have been incorporated in some Codelco plants (e.g., Radamiro Tomic), in Girilambone, and in Phelps Dodge. This acts to mechanically reduce the concentration of aqueous entrainment prior to the loaded organic being pumped to the stripping stage.

Chemical Impurity Transfer. Factors affecting chemical iron transfer include the copper/iron ratio in the PLS, the degree of maximum loading of the organic (crowding effects) and the PLS acidity. Installation of a suitable wash stage utilizing acidified wash solution containing some copper allows for the iron to be preferentially stripped, and the excess copper present in the wash water can then be extracted onto the extractant. This can achieve reductions in chemical iron transfer of 50 to 80%.

It has been noted that series parallel configurations appear to give poorer selectivity, hence increased iron transfer.

MIXING

Mixer Continuities

The mixing of the organic and aqueous can be done in two ways, depending on the relative volume of each in the mixer:

1. Aqueous continuous
2. Organic continuous

Under normal conditions, if there is more organic than there is aqueous (i.e., an O:A of 1.1 to 1), then the mixing will be organic continuous and vice versa.

The term organic continuous means that in the mixer there will be a matrix of organic and dispersed through it in small discrete bubbles will be the aqueous phase. For aqueous continuous there is a matrix of aqueous with bubbles of organic through it.

The relative volume of the phases in the mixers can be changed in two ways:

1. Increasing the actual pumped flow rate of one phase. This will result in a change in the extraction or stripping chemistry as it will change relative amounts of reactive components present (organic, acid or copper).
2. Recycle one phase back into the mixer. This will not change the extraction or stripping chemistry as the components recycled have already reacted in that mixer previously. (Note: Recycle, particularly in strip, can improve the stage efficiency relative to very high O/A ratios by reducing stratification of aqueous droplets.)
3. Use of surfactant to stabilize aqueous continuous mix.

Under normal conditions, the mixer continuity does not effect the overall chemistry of the extraction process.

The mixer continuity is important, however, as it is the major controlling factor for the entrainment. It should be noted also that in practice, although it is possible to maintain a desired mixer continuity at an O:A of 1:1, it is normal to maintain continuities by ensuring there is always about 10% excess volume of the continuous phase.

Entrainment

Entrainment is a word used to describe the situation where a quantity of one phase, say the aqueous phase, remains in the organic phase when the organic phase exits the settler. This is called aqueous entrainment in organic. Organic entrainment in aqueous is when quantities of organic remain with the aqueous phase as it exits the settler.

The following relationship applies for mixer continuities:

Mixer Continuity	Effect on Entrainment
Organic continuous	Minimizes organic entrainment in aqueous
	Maximizes aqueous entrainment in organic
Aqueous continuous	Minimizes aqueous entrainment in organic
	Maximizes organic entrainment in aqueous

When deciding what continuity a mixer should be operated at, the impact of the entrainment exiting the settler must be considered.

E1 aqueous discharges into E2 mixer where any organic entrainment gets recontacted with organic in the mixer and should not cause any problems. E1 loaded organic discharges into the loaded organic system. Any aqueous entrainment will contain the impurities which the solvent extraction is trying to remove the copper from. The entrained aqueous flowing with the loaded organic will take the impurities with it. If the entrainment has not been removed by the time the loaded organic enters the stripping stage, then the impurities, such as iron, chloride and manganese, will also enter the electrolyte and can cause serious process problems both in solvent extraction and electrowinning. It is therefore normally necessary to run E1 aqueous continuous so that the aqueous entrainment in organic is minimized.

E2 aqueous entrainment in the organic is not normally a problem as the organic discharges into E1 where the aqueous is remixed with PLS. However, the aqueous discharge from E2, the raffinate, leaves the circuit and reports to the raffinate pond. Any organic entrainment leaves the circuit and reports to the raffinate pond. Any organic entrainment leaving with the aqueous will be lost from the solvent extraction circuit and will end up floating on the surface of the raffinate pond. Considerable evaporation of the diluent will occur in the pond together with degradation of the extractant. This material is difficult to remove from the surface of the pond and once recovered it must be treated prior to reintroduction into the solvent extraction circuit. As a consequence, E2 is normally run in an organic continuous mode to minimize organic entrainment.

In the stripping stage, the electrolyte leaving S1 is strong electrolyte which is the feed stream to electrowinning. Organic entrainment entering the electrowinning creates very serious operational problems, is difficult to recover and again can lead to the degradation of the organic. Consequently S1 is normally run in organic continuity.

S1 organic, stripped organic, discharges into the E2 mixer. Any aqueous entrainment contained in the organic represents a loss of copper and acid from the electrolyte circuit and can create extraction problems in E2 due to excess acid. However, this is normally seen as a secondary problem to organic entrainment in electrolyte and also the loss of electrolyte can be utilized as part of the controlled electrolyte bleed that is normally required.

These rules apply to “normal ideal operating” conditions which often do not prevail in an industrial plant. Under conditions of excessive phase disengagement times, it may be necessary to run aqueous continuous, or if there is significant crud being generated, it may be necessary to run organic continuous such that the crud compacts to the aqueous organic interface in the settler.

Tip Speeds

High mixer tip speed can lead to increased entrainment due to excessive shear as well as increase propensity for crud generation and air entrainment (aqueous continuous). The usual defining requirement for impeller tip speed is to maintain a given mixer pumping capacity and corresponding solution flowrate. Once this requirement is met however, tip speed, especially in auxiliary mixers, should be maintained as low as possible while maintaining the dispersed phases. Maximum considered 400 m/min.

Low tip speed can result in insufficient pumping capacity in the primary mixer which backs up solutions in the organic and/or aqueous weirs. When this occurs, the height of the weir is no longer controlling the hydraulic balance in the settler. This can cause very significant flowrate fluctuations. Tip speed must remain high enough to maintain a cascade of flowing solution over the weir. Low tip speed reduces entrainment. In the extreme can effect mixer stage efficiency.

SETTLERS

After the aqueous and organic phases have been mixed and the extraction or stripping reactions are complete, related to the defined mixer residence time, then the two phases will separate normally as the organic is insoluble in the aqueous phase. Because the organic is lighter than both the PLS and electrolyte (i.e., approx. 1.10 kg/l and 1.2 kg/l, respectively), the organic will be in the upper layer.

Phase Disengagement Time

The time that is required for the organic and aqueous phase to separate is called the phase disengagement time. The phase disengagement time is important as the settler size is designed such that there is sufficient residence time to allow the two phases to separate. If there is insufficient time in the settler for separation to occur (i.e., the phase disengagement time is greater than the settler residence time), then excessive aqueous and organic entrainment can occur.

New organic has a phase disengagement time of 30–60 seconds, while typical plant organics have phase disengagement times of 30–120 seconds. When the phases have been mixed aqueous continuous, the phase disengagement time will tend to be more rapid than when the phases have been mixed in organic continuity. A number of materials, which if present during the mixing, will result in increasing phase disengagement times. Examples of these materials are entrained air; chemical surfactants such as soaps, flocculants and degreasers; degraded organic products; colloidal material; over mixing. In an organic continuous mix, the presence of solids may destabilize the mix, causing the phase continuity to spontaneously change to aqueous continuous. However, solids may also significantly decrease phase disengagement time in organic continuous mixing.

In extreme cases, phase disengagement times in excess of 300 seconds may result and remedial action must be taken to stop the material of concern entering the circuit and it may also be necessary to treat the plant organic.

When phase disengagement problems do occur, it may be necessary to slow the flowrate through the settler (i.e., decrease the plant flow rate). This decreased flow rate allows for an increase in the settler residence time and therefore more time for the two phases to separate before they exit the settler. In less extreme cases, it may only necessitate the volume (indicated by the level) of organic or aqueous in the settler to be changed by utilizing the settler aqueous weir. In some instances, it may be beneficial to change (flip) the mixers running organic continuous so that they run aqueous continuous thus giving improved disengagement times.

Clay Treatment

It has become standard practice in industry to clay treat organic which has extended phase break times, has been contaminated by surfactant or has been recovered from the raffinate pond. The clay acts to remove contaminants and restores the phase disengagement times back to similar levels shown by fresh organic.

Settler Depths

The depth of organic in the settler can be altered by adjustment of the settler aqueous overflow weir. As the aqueous overflow level is increased, the organic depth decreases and vice versa. Plants normally attempt to operate with minimum organic depth in order to minimize the organic inventory. The shallower the depth, however, the greater the difference in solution velocity between the organic and aqueous phase. Absolute velocity in the settler should not exceed 3 m/min.

Shallow organic depth results in:

1. Increased risk of aqueous entrainment in organic
2. More rapid coalescence
3. Minimized organic inventory

A deep organic inventory results in:

1. Increased risk of organic entrainment in aqueous
2. Slower coalescence
3. Better ability to control crud

One frequently overlooked aspect of settler depth is to optimize the control of entrained solution by increasing the depth of the opposite phase. Hence, if iron transfer via A in O entrainment in E1 to S1 is significant, the organic depth in E1 could be increased, allowing far longer settler retention times with less carryover of A in O entrainment. Similarly, increased E2 aqueous depths would decrease O in A entrainment in solution leaving E2.

CRUD

When solids, colloidal material or chemical precipitates enter a solvent extraction circuit, there is a strong likelihood that these materials when mixed with organic and aqueous will form a stable material called crud. This crud contains varying quantities of organic and aqueous stabilized by very small quantities of the solids mentioned (i.e., normally less than 5% by weight). The type of crud formed and where it will appear in a circuit is difficult to predict and is very site specific.

The problem with crud is that it can:

1. Retain organic and reduce the effective amount of organic available for extraction and stripping
2. Build up in a settler and result in decreased settler capacity
3. Migrate through the circuit and transfer impurities from the extraction stage to the stripping stage
4. Lead to excessive entrainment

The following comments are only generalities about the forms of some crud and how best to treat these crud problems.

When running mixers in aqueous continuity, the generated crud tends to float throughout the organic phase due to entrained air. In the settler, the crud can then transfer over the organic weir and into the next solvent extraction stage. By changing the mixer to organic continuity, the crud tends to either compact at the aqueous/organic interface in the settler or wash out with the aqueous. Crud generated from suspended solids generally builds up in the extraction stages while crud generated from colloidal material or chemical precipitates builds up in S1.

Crud Management

Various crud management techniques have been developed over the years and have generally proved to be very effective. For example, a South American mine uses a portable compressed air driven pump to suck crud from the settlers on a regular basis. This material is then sent to a centrifuge for organic recovery and the solids for disposal. Maintenance of the centrifuge proved to be problematic and organic recovery was rarely more than 80%. An alternative approach was developed in which clean plant organic was added to an air agitated pachuca and the crud added to this, taking care to maintain organic phase continuity throughout the agitation process. It was found that when the agitation was stopped, the separation of phases was clean and rapid with the solids forming a compact layer in the aqueous phase. The organic trapped in the crud was released to the clean organic phase and could be fed back to the circuit. The solids could be easily disposed of. A North American company has further refined the technique and claims organic recovery as high as 98%. Girilambone use plate and frame filters or sparkle filters for crud management; Codelco has their own developed crud management.

FUTURE TRENDS

It is felt that future trends in solvent extraction plant management and optimization will increasingly rely on computerized techniques. It can be envisaged that, in the near future, plants will have the ability to use on-line analysis of feed solutions, this data being fed to a proprietary software package such as Minchem/MEUM for automatic isotherm generation, and McCabe Thiele constructions to the design parameters. Such a plant will have its performance regularly monitored and compared with design so that automatic adjustments can be made to maintain the plant performance as close to optimal as possible. In this way, maximum financial benefits will be realized. All the abovementioned techniques are available now. It is really a question of putting together the elements and testing the overall package.

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Evolutionary Development of Solvent Extraction Reagents: Real-Life Experiences

Gary A. Kordosky,[*] Robert B. Sudderth,[*] and Michael J. Virnig[*]

For a variety of reasons, it is rare that the first solvent extraction reagent developed for the extraction of a certain metal under a given set of conditions remains the reagent of choice for that specific application over an extended period of time. Three examples are presented which detail evolutionary changes in solvent extraction reagents for a specific application. Example 1 discusses the development of oxime reagents for the recovery of copper from sulfuric acid leach solutions. Example 2 details the development of oximes and betadiketones for the extraction of copper from ammonia leach solutions, while example 3 traces the development of the oxime reagent used by Queensland Nickel for their nickel solvent extraction plant.

INTRODUCTION

Metal recovery by solvent extraction is a well-known technique that is practiced commercially for the recovery of copper, uranium, nickel, cobalt, palladium, vanadium, molybdenum and other metals. Solvent extraction reagents for a specific application are developed in two ways:

1. Any and all available reagents and/or combinations of reagents that might work for a specific application are tested, and the most promising candidates are developed.
2. When no known reagent or combination of reagents has the required properties for a specific application, a new molecule must be developed and often further modified or a new combination of reagents must be found.

It is rare indeed when the first reagent developed for a specific application remains the reagent of choice for an extended period of time. One such reagent is Alamine[†] 336 which was developed in the mid 1950s specifically for the extraction of uranium from sulfuric acid leach solutions and is still today, over 40 years later, the reagent of choice for this application.

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[*] Henkel Corporation, Tucson, Ariz.

[†] Alamine is a registered trademark of the Henkel Group.

There are a variety of reasons why the first reagent developed for a specific application does not often remain the reagent of choice for an extended period of time, among them: not fully understanding the requirements of the application, incomplete testing, the availability of raw materials, changes in the specific application, economics, advances in reagent manufacturing, the quality of the final metal product and regulatory changes.

This paper uses the term reagent of choice to reflect the assumption that the reagent user will select the reagent that he or she believes will result in the best overall economics for the process. This means that the reagent of choice may not always have the best metallurgical performance properties but, rather, the optimum combination of performance properties and economics. While this paper makes the previous assumption, it is the experience of the authors that this assumption is not always true. People making a reagent choice do not always consider all the performance properties of a particular reagent for two main reasons: either they do not fully understand the process and therefore cannot accurately assess how certain reagent properties will affect the process, or, more often, because certain performance properties cannot be accurately assessed or measured in the laboratory or in a pilot plant, they can only be accurately measured in the commercial plant and testing in a commercial plant poses a significant risk.

To be commercially successful a reagent must:

- Extract the desired metal(s) selectively from the metal-containing solution
- Be strippable into a solution from which eventual metal recovery can take place
- Be stable to the circuit conditions so it can be recycled many times
- Be nonflammable, nontoxic, noncarcinogenic, etc.
- Be soluble in an inexpensive organic diluent or be able to function as the diluent
- Transfer metal at a rate fast enough to allow for the use of economical mixing times
- Not promote stable emulsions
- Not transfer deleterious species from strip to extraction
- Have an acceptable cost

However, it is important to realize that in many cases commercially successful reagents are not the best with respect to all properties listed above, rather they possess a good balance of these properties.

The development of solvent extraction reagents, from the first go around to the present day reagent(s) of choice, for the following specific commercially practiced applications are discussed:

- Copper extraction from dilute sulfuric acid leach solutions
- Copper extraction from ammonia leach solutions
- Nickel extraction from a dilute ammonia leach solution followed by stripping with concentrated ammonia solution

COPPER EXTRACTION FROM DILUTE SULFURIC ACID LEACH SOLUTIONS

This is the largest application of solvent extraction for metal recovery with over 2 million tonnes of installed capacity for copper cathode production. It is fitting then that the development of reagents for this application has had more twists and turns than the development of a reagent or reagents for any other application. In 1960, building on the

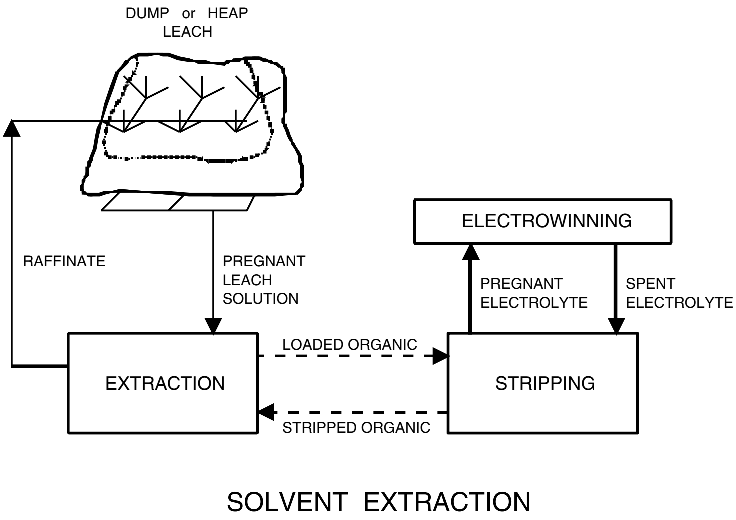
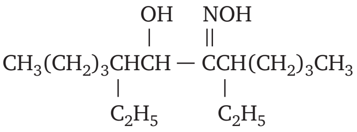


FIGURE 1 **General flowsheet for leach–solvent extraction–electrowinning**



$$CH_3(CH_2)_3\underset{\displaystyle C_2H_5}{\underset{|}{C}}H\overset{\displaystyle OH}{\overset{|}{C}}H - \overset{\displaystyle NOH}{\overset{\|}{C}}\underset{\displaystyle C_2H_5}{\underset{|}{C}}H(CH_2)_3CH_3$$

FIGURE 2 **Structure of the active extractant in LIX 63**

success of Alamine 336, the Minerals Development Group of General Mills Chemicals identified the extraction of copper from dilute sulfuric acid leach solutions as a potential large scale application of solvent extraction and began the search for a molecule which would extract copper selectively in the presence of iron from dilute sulfuric acid leach solutions according to the general flowsheet given in Figure 1 (House 1981).

The first reagent developed by General Mills for this application (LIX* 63, Figure 2) possessed many of the required properties needed to be successful as a copper solvent extraction reagent in that it was specific for copper, compatible with the stripping process, stable, water insoluble, affordable and safe to use (Swanson and Agers 1964). However, the reagent did not extract copper well below pH ~3.5, and, as a result, it was not a suitable reagent for the general flowsheet given in Figure 1.

Development work continued and in late 1965 the solvent extraction reagent LIX 64 was introduced (Agers et al. 1965). This reagent was composed of LIX 65 (Figure 3) with a catalytic amount of LIX 63 added to increase metal transfer kinetics. This reagent became the first commercial copper solvent extraction reagent when in March 1968 Ranchers Exploration and Development Corporation brought their new copper SX-EW plant on stream (Power 1970).

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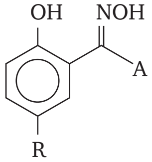
* LIX is a registered trademark of the Henkel Group.



Reagent	R	A
LIX 65N	$C_{12}H_{25}$	C_6H_5
LIX 65N	C_9H_{19}	C_6H_5
LIX 84-I, SME 529	C_9H_{19}	CH_3
LIX 860-I	$C_{12}H_{25}$	H
P-1, LIX 860N-I	C_9H_{19}	H

FIGURE 3 General structure of oxime extractants used for copper recovery

While LIX 64 had the required properties to be used commercially as a copper extractant (see flowsheet given in Figure 1), the reagent had significant limitations with respect to extractive strength, metal transfer kinetics, copper/iron selectivity, viscosity and phase separation. As a result LIX 64 could only be used with a narrow range of leach solution under well-controlled conditions. Interestingly, several of the limitations of the reagent resulted from a lack of high quality starting material and a manufacturing process that needed further development and refinement. This is not surprising. These were the early days of development and the impact of impurities in the reagent on many of the required reagent properties such as copper/iron selectivity, entrainment and phase separation was not fully appreciated or well understood.

LIX 64 remained the copper solvent extraction reagent of choice for only about six months when in November 1968 LIX 64N was added as makeup to the Ranchers plant. LIX 64N had significantly improved properties over LIX 64, including greater extractive strength, faster kinetics, faster phase separation, lower entrainment, increased copper/iron selectivity and lower viscosity. These improved properties, which broadened both the range of copper leach solutions which could be treated and the conditions under which these leach liquors could be treated (Dement and Merigold 1970), resulted as much from improved purity of starting materials and improved reagent manufacturing techniques as from the structural change of a C_{12} to a C_9 alkyl side chain.

While LIX 64N, which contains LIX 65N (Figure 3) as the active extractant with a catalytic amount of LIX 63, remained the reagent of choice for about 10 years, it still had some limitations with respect to kinetics, copper/iron selectivity and the pH and copper content of leach liquors which could be economically treated.

At the time of the development of LIX 64 and LIX 64N, General Mills Chemicals was well aware that the A group in Figure 3 could also be either CH_3 or H instead of C_6H_5, but General Mills Chemicals decided not to include these two potential extractants in the claims of their patent application. Had they included these two potential extractants in the claims, the reagent development story which follows would have been very different. It is likely that reagents based on the active extractants in SME 529 and/or Acorga P–1 discussed later would not have been developed at that time by the companies that developed them even though they might have been developed at a later date by General Mills or Henkel.

Why did General Mills not claim the molecules where A in Figure 3 is either CH_3 or H? At the time of the development of these reagents, General Mills Chemicals had surveyed the industry and found that a major concern of the copper companies was the stability of the copper solvent extraction reagent under the conditions in which it would be used. The molecule where A is C_6H_5 was thought to be much more stable than the molecules where A is H or CH_3, so General Mills chose to claim only that particular class of molecules (House 1996). In fact, General Mills was right, the molecule where A = C_6H_5 is more stable; however, the molecules where A = H or CH_3 are sufficiently stable under the conditions of copper SX so that what General Mills Chemicals believed to be a major concern in the mid to late 1960s was a minor issue 10 years later.

In 1968, Ashland Chemical Company introduced Kelex* 100 as a reagent which would fill the perceived need for a copper solvent extraction reagent that could more effectively treat leach solutions having a lower pH and/or a higher copper content than LIX 64N could treat these leach solutions (Ashland 1968). Kelex 120 followed in 1970 (Hartlage and Cronberg 1973). Along with these reagents, Ashland also introduced the idea of using varying amounts of modifier, either nonylphenol or tridecanol, to facilitate the stripping of copper from strong copper extractants. While neither Kelex 100 or Kelex 120 was ever used commercially for copper recovery from dilute sulfuric acid leach solutions, the idea of using modifiers to alter the extraction/stripping equilibrium of a copper extractant represented a significant advance in reagent development which has resulted in improved reagents still in use today.

In the mid 1970s, Shell International Chemicals brought SME 529 (Figure 3) to the market as an alternative to LIX 64N (van der Zeeuw 1975). This reagent found only limited commercial use mainly because the poor properties of the side products from the manufacturing process overwhelmed the good properties of the active extractant. Had Shell International Chemicals at the time produced a highly purified SME 529, the reagent would likely have found wide commercial acceptance and Shell might still be a supplier today.

In 1974, Birch reported that P–1† (Figure 3) from Acorga had very good properties including the ability to extract copper very efficiently, rapid copper transfer kinetics and excellent copper/iron selectivity (Birch 1974). However, the reagent was such a strong copper extractant that it required more than 225 g/l H_2SO_4 for efficient stripping, an acid concentration which is not compatible with the plating of high quality copper under currently acceptable tankhouse conditions.

Three years later, Acorga introduced the P–5000 series of reagents (Tumilty et al. 1977; Tumilty et al. 1979). In introducing this series of reagents Acorga not only applied Ashland's idea of using nonylphenol to facilitate copper stripping to the strong copper extractant P–1, Acorga also took the use of modifiers one step further by combining the strong copper extractant P–1 with varying amounts of nonylphenol to give a series of reagents where the extractive strength could be tailored for any given leach solution or set of conditions. Copper leach liquors having a higher copper concentration and/or a lower pH could now be effectively treated with only two extraction stages even at low temperatures. It is interesting to speculate on whether Acorga would have combined P–1 with nonylphenol had Ashland not discovered that molecules such as nonylphenol and

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* Kelex 100 and Kelex 120 are no longer produced by Ashland Chemical Company. SME 529 is no longer produced by Shell Chemical Company.

† Acorga P-1 is produced by Zeneca Chemical Specialties.

tridecanol altered the copper extraction/stripping equilibrium of the active copper extractant in Kelex reagents.

Unfortunately, the presence of nonylphenol in the reagent contributed to increased reagent degradation, higher organic phase viscosity and higher entrainment and, in addition, nonylphenol was not compatible with certain materials of construction. Reagent users now had to evaluate the benefits of using the P–5000 series of reagents, such as stronger copper extraction, faster kinetics and less staging, against the drawbacks of using the P–5000 series of reagents such as higher viscosity of the organic phase, increased entrainment, greater crud generation and higher reagent consumption per tonne copper produced when compared to the use of LIX 64N. It is interesting to see how the market passed judgment on this new series of reagents. The reagent Acorga P–5100, which had the lowest modifier to oxime ratio in the P–5000 series, is still used today. In this case, the industry has judged that the benefits of using this reagent outweigh the drawbacks. However, Acorga P–5200 and Acorga P–5300, which had higher modifier to oxime ratios than Acorga P–5100, were not widely used and are no longer part of the Acorga line of reagents. The industry has judged that the benefits of using these reagents did not outweigh the drawbacks with respect to entrainment, viscosity, etc.

Acorga either did not mix P–1 with the then available ketoxime reagents, LIX 64N and SME 529, or if they mixed these reagents, they did not recognize the benefits of these mixtures. In either case, this was a missed opportunity as discussed later.

In 1979, Henkel introduced the tridecanol modified aldoxime reagent LIX 622 which in 1981 became the first tridecanol modified copper extractant to be used commercially.

In 1983, Henkel reported that patented mixtures of strong aldoxime extractants such as LIX 860-I (Figure 3) with the moderately strong ketoxime extractants such as LIX 64N and SME 529, gives reagents the fast kinetics and good copper extraction characteristics of the aldoximes combined with the stability and good physical performance of the ketoximes, all without the detrimental properties of modifiers (Kordosky et al. 1983). The addition of the pure aldoxime extractant LIX 860-I to existing copper SX plants using LIX 64N or SME 529 afforded these plants an easy way to upgrade their plant performance and flexibility if needed or desired without the negative properties of the modifiers. Aldoxime/ketoxime mixtures have proven to be a significant advance in copper solvent extraction reagents, and today these mixtures produce more copper than any other single reagent system.

Why were the aldoxime/ketoxime reagents mixtures developed? At the time these mixtures were developed, the copper industry was changing—copper companies were willing to pay more in operating costs in order to save on capital, and leaching techniques were producing leach solutions containing greater amounts of copper at lower pH. These changes in the industry meant that stronger copper extractants having faster kinetics were required and that LIX 64N as a reagent by itself would soon be obsolete. Henkel needed a way to keep LIX 64N as a viable reagent because they had a plant which produced it and a significant inventory of the product in stock so they tested mixtures of LIX 64N with aldoximes and to their surprise found that these mixtures performed much better than expected. The aldoxime/ketoxime mixtures were quickly introduced and made available.

In 1984, Henkel purchased the manufacturing and applications technology and associated patent rights to SME 529 from Shell International Chemical Company. In late 1986, using new manufacturing techniques, Henkel began producing LIX 84-I (Figure 3), a highly purified and improved version of the old SME 529. This reagent replaced

TABLE 1 Trends in reagent properties

Property	1965	1970	Late 1970s	Today
Extractive strength	Moderate	Moderate	Strong	Tailored
Cu/Fe selectivity	Fair	Good	Good	Excellent
Kinetics	Slow	Moderate	Fast	Fast
Stability	Excellent	Excellent	Good	Very Good
Crud generation	Moderate	Good	Moderate	Good
Versatility	Poor	Marginal	Good	Excellent

both LIX 64N and LIX 65N in the Henkel line of reagents and led to mixtures of aldoximes with LIX 84-I.

In 1986, the use of high molecular weight, highly branched alcohols and esters as modifiers was reported (Dalton et al. 1986). While the paper was very optimistic about the higher molecular weight alcohols as modifiers, in practice these modifiers caused very high entrainment in operating plants and they are not used today. Ester modified reagents, however, are widely used and are particularly noted for their exceptional copper/iron selectivity and very good stability. Unfortunately, with many leach liquors, the use of ester modified aldoxime reagents often results in greater entrainment and crud generation than does the use of nonmodified reagents.

Copper Solvent Extraction Reagents Today

The improvements in copper solvent extraction reagent properties previously detailed is summarized in Table 1. These improvements have come about for four reasons:

1. New molecules have been developed such as the active extractants in SME 529, P–1 and LIX 860-I.
2. Better manufacturing processes have resulted in cleaner reagents; LIX 84-I compared to SME 529 is a good example.
3. Reagent properties can be manipulated; aldoxime blends with modifiers or with ketoximes are examples.
4. The technical and economic success of leach-solvent extraction-electrowinning plants led to new leaching techniques which in turn forced reagent suppliers to develop new reagents to effectively treat leach solutions resulting from these new leaching techniques.

There are two distinct classes of modern copper extractants: ketoximes, where A in Figure 3 is a methyl (CH_3) group; and aldoximes, where A is a hydrogen (H). A general comparison of their extraction characteristics as well mixtures of the two is given in Table 2.

The availability of pure aldoxime extractants, such as LIX 860-I and LIX 860N-I, modified aldoxime reagents such as Acorga PT5050, Acorga M5640 and LIX 622N among others, the pure ketoxime reagent LIX 84-I and various aldoxime/ketoxime nonmodified mixtures, gives the operators of copper solvent extraction plants a wide choice of reagents from which to choose and choose they do. While some plants use a single reagent such as a modified aldoxime or a premixed aldoxime/ketoxime blend, many other plants purchase the aldoxime and ketoxime components separately and then blend a reagent best suited to their plant conditions. In fact, Chuquicamata has developed a

TABLE 2 Comparison of properties for reagents based on ketoximes, modified aldoximes and ketoxime/aldoxime mixtures

Property	Ketoxime	Aldoxime	Mixture
Extractive strength	Moderate	Strong	Customized
Stripping	Very Good	Good	Customized
Cu/Fe selectivity	Excellent	Excellent	Excellent
Copper kinetics	Very Good	Very Fast	Fast
Phase separation	Fast	Fast	Fast
Stability	Very Good	Very Good*	Very Good
Crud generation†	Low	Variable	Low

*Dependent upon the particular modifier used.
†Dependent upon the leach liquor and modifier.

special four-component blend for their copper SX plants; plants which operate very well under difficult conditions (Pincheira, et al., 1997).

Since ketoxime reagents are moderately strong copper extractants, they operate best when the pH of the leach solution is relatively high (above pH 1.8), the solution is relatively warm (20°C or higher) and the amount of acid wanted in the tankhouse is relatively low (less than 160 g/l H_2SO_4). These reagents strip well even with one strip stage.

Modified aldoxime reagents have excellent metallurgical properties even at low temperatures, low pH's or when the copper content of the leach liquor is very high and high copper recovery is needed. However, in most situations modified aldoximes cause higher entrainment and/or greater crud generation resulting in increased carryover of impurities to the electrolyte and higher reagent consumption per tonne copper produced.

While the nonmodified aldoxime–ketoxime mixtures also operate well at lower pH's, lower temperatures and with leach liquors having a high copper content without the higher entrainment and crud generation problems of the modified aldoximes, the mixtures are not quite as effective as the modified aldoximes with respect to copper extraction. For this reason, there are and will continue to be situations where the plant operator must make a choice between the better metallurgical properties of the modified aldoximes and the superior physical properties of the nonmodified aldoxime/ketoxime mixtures.

Summary

The development of copper solvent extraction reagents is an example of reagent development where a new molecule had to be designed and synthesized and where the first molecule offered (LIX 63) had some of the needed properties. However, an understanding of the chemistry of the extractant led directly to the design and synthesis of a second new molecule (the active extractant is LIX 65) which had the required properties to be used commercially. Since then there have been many evolutionary improvements in copper solvent extraction reagents. The authors classify these improvements as evolutionary for two reasons:

1. The basic structure of the extractants is the same as the basic structure for LIX 65 even though certain parts of the molecule are changed.
2. Some of the improvements in reagent performance are due to better manufacturing processes and purification techniques.

Today there is not "a reagent of choice" for the extraction of copper from sulfuric acid leach liquors, but many good reagents and/or blends of reagents from which to choose.

COPPER EXTRACTION FROM AMMONIA LEACH SOLUTIONS

Ammonia as a leachant for copper from a variety of sources has been well known for a long time even though commercial applications have been limited (Benedict 1917; Forward 1953; *Eng. & Min. J.* 1973). When General Mills found that LIX 63 did not extract copper well from dilute sulfuric acid leach solution, they turned their attention to the extraction of copper from ammonia solutions with LIX 63 as a way to demonstrate the solvent extraction process to the copper industry (Swanson and Agers 1964). At the time LIX 63 was considered a good extractant for copper from ammonia solutions, but it was never used commercially because of stripping problems and also because of the development of the more versatile and less expensive LIX 64N. LIX 64N did find limited commercial application for the extraction of copper from ammonia, most notably in the Arbiter Process (*Eng. & Min. J.* 1973) and in the Imperial Smelting Process for leaching lead dross (Hopkin et al. 1983). While LIX 64N extracted copper very rapidly and very strongly from ammonia solutions, its properties in this application were not optimum, particularly with respect to ammonia transfer, which was somewhat greater than desired, and copper transfer, which was limited to about 12 g/l copper. SME 529 behaved much the same. Thus there existed a need for a reagent for the extraction of copper from ammonia solutions which would overcome the shortcomings of the hydroxyoximes.

In 1975 LIX 54, a betadiketone reagent designed specifically for the extraction of copper from ammonia leach solutions (Figure 4), was brought to the market (General Mills Chemicals 1975). This reagent has outstanding properties for this application: it transfers over 30 g/l Cu and remains highly fluid, ammonia loading on the reagent is very low and the reagent can be stripped with relatively low acid concentrations. In fact, pregnant leach solutions having 50 g/l Cu and only about 10 g/l free sulfuric acid can be generated giving the operator great flexibility in the production of the final copper product. On the negative side, betadiketones are weaker copper extractants than hydroxyoximes and for this reason they will not extract more than about 50 to 60 g/l Cu from an ammonia solution.

Unexpectedly, it has recently been found that LIX 54 will react with ammonia at pH above ~9.5 to give surface active molecules which appear to increase entrainment of aqueous in the loaded organic (with the resulting carryover of ammonia to the wash stage) and to decrease copper stripping kinetics. Henkel in response to this problem studied the reaction of LIX 54 with ammonia and then designed a new extractant molecule to overcome the problem. In the laboratory, this new extractant, called XI–57,

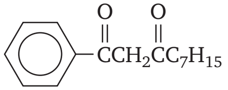

FIGURE 4 Chemical structure of the active extractant in LIX 54

TABLE 3 General properties of betadiketone and ketoxime copper extractants in the ammonia leach system

Property	LIX 54–100	LIX 84
Extractive strength	Moderate	Strong
Ease of stripping	Excellent	Good
Viscosity	Low	Moderate
Ammonia loading	Very low	Low
Selectivity of copper	Very good	Very good
Kinetics	Fast	Fast
Stability	Good	Good
Phase separation	Very good	Good
Copper complex solubility	Very high	Very high

displays many of the good properties of LIX 54, but it does not react with ammonia in the manner of LIX 54. After four months of pilot testing with an actual ammonia leach solution there is no indication the reagent has reacted with ammonia, ammonia carryover to wash has been low and stripping kinetics have remained rapid.

To date, only ketoxime and betadiketone reagents have been used commercially for the extraction of copper from ammonia leach solutions. The general properties of these two classes of reagents are summarized in Table 3.

Betadiketone extractants are the reagent of choice in most applications because of their high copper net transfer, low ammonia loading, ease of stripping and overall smooth operating characteristics. LIX 84-I is used only when there is a need to extract more than about 50 g/l Cu from an ammonia solution or when a high percentage of copper must be recovered from a solution having a high free ammonia content.

Summary

The first reagents used for the extraction of copper from ammonia solutions (hydroxyoximes) represent the situation where a reagent developed for one application, in this case the extraction of copper from dilute sulfuric acid, is adapted for use in another application and de facto becomes the reagent of choice because there is no other reagent which works reasonably well. The development of the betadiketone reagent, LIX 54, represents the case where a reagent is developed specifically for a given application to give better overall performance than the existing reagent of choice. The development of XI–57 is an example where a specific problem with the reagent of choice has been identified and a new molecule has been designed and synthesized to overcome the specific problem. The history of the development of a "reagent of choice" for this application appears to be a lot of work for a reagent which may never be all that widely used, not because the reagent lacks good properties, but rather because the leaching system has limitations when compared to other leaching systems for copper. One wonders if less effort would have been put into developing a reagent for this system if the limitations of the leaching system had been well known 25 years ago. Then again, the availability of an excellent reagent for this application may spur research and development which will overcome the limitations of the ammonia leaching system.

NICKEL EXTRACTION FROM A DILUTE AMMONIA LEACH SOLUTION FOLLOWED BY STRIPPING WITH A CONCENTRATED AMMONIA SOLUTION

Typical chelating metal extractants extract many base metals depending on the conditions of the extraction. For example, nickel extraction with a solvent extraction reagent developed for the extraction of copper was first commercially practiced by S.E.C. in 1972 (Eliasen and Edmunds 1974). In this plant, nickel was extracted from an ammoniacal solution and stripped from the extractant with sulfuric acid to give a solution from which nickel was direct electrowon.

Queensland Nickel operates a nickel production plant based on the Caron Process for treating nickel laterite ores. This plant produced $NiCO_3$ as an intermediate product by distillation of a relatively dilute ammoniacal nickel carbonate leach solution. The $NiCO_3$ intermediate product was further processed to give NiO and Ni metal as final products. A mixed cobalt/nickel sulfide product was also produced and sold. In order to reduce production costs, produce higher purity products and because the market for mixed cobalt/nickel sulfide product was going to disappear, Queensland Nickel developed a flowsheet incorporating solvent extraction for nickel with the special requirement that the solvent extraction process produce a highly concentrated pure solution of Ni in concentrated aqueous ammoniacal carbonate (Reid and Price 1993).

Queensland Nickel conceptualized the flowsheet in 1985 and immediately began to test many solvent extraction reagents then commercially available, including the oxime copper extractant LIX 64N and the betadiketone copper extractant LIX 54, in an effort to find a reagent that fit the chemistry of the Queensland Nickel Process. On the one hand, LIX 64N and other oxime copper extraction reagents proved to be such strong nickel extractants that they did not strip adequately under the conditions of the Queensland Nickel flowsheet. LIX 54 on the other hand stripped very well, but it was a weaker Ni extractant than desired. Still, LIX 54 provided the best balance of properties so it was piloted for about six months. LIX 54 performed adequately in the pilot plant with respect to the metallurgy of the process; however, under the relatively harsh stripping conditions, LIX 54 degraded too rapidly to be economical. At that point, the lack of a reagent having all the required properties nearly killed the project.

The properties required for a reagent to be successful in the Queensland Nickel process follow:

- Extract nickel strongly and selectively from a dilute ammonia leach solution
- Be strippable with an ammonia solution having ~220 g/l ammonia and ~200 g/l CO_2 such that a pregnant strip solution having about 75 to 80 g/l Ni can be produced
- Be stable to the relatively harsh stripping conditions
- Have low entrainment so that impurities, particularly sulfate ions, are not carried by entrainment from the leach solution to the strip solution
- Be commercially available at a reasonable cost

At the time this was a formidable challenge considering the large number of reagents Queensland Nickel had already tested and the fact that reagent producers had been consulted about the reagent properties needed and asked to submit samples for testing. Henkel returned to the lab knowing that an oxime reagent which would strip more readily than typical oxime reagents was needed. Several modified versions of LIX 84-I were tested. Overall the work was not encouraging, but, for unknown reasons, the Henkel personnel doing the work decided a certain formulation had to work so they abandoned the lab and went straight to a small on-site continuous pilot plant test. To their surprise and delight, the first formulation tried in the small pilot plant for this application worked quite well and within about a week a reagent formulation had been defined and the small continuous pilot plant was giving the desired nickel extraction while producing the needed nickel pregnant strip solution. Queensland Nickel quickly piloted the experimental reagent on a larger scale for an extended period of time, then designed and built a nickel SX plant which has been running for more than 10 years (Reid and Price 1993). The plant initially used LIX 87QN, a tridecanol modified ketoxime; however, the ability of the plant to produce a concentrated ammonia strip solution improved and Queensland Nickel now uses a less modified version of LIX 87QN which better fits the new stripping conditions.

Summary

This is a case where a reagent originally designed for the extraction of a certain metal (copper) has been modified to extract and strip a different metal (nickel) under a set of unique and unusual extraction and stripping conditions and, additionally, where an understanding of the chemistry of the extractant has allowed a further modification of the extractant.

REFERENCES

Agers, D.W., J.E. House, R.R. Swanson, and J.L. Drobnick, 1965, "A New Reagent for the Liquid Ion Exchange of Copper," *Mining Engineering,* December, pp. 76–80.

Ashland Chemical Company, 1968, Preliminary Technical Data Sheet, "Copper Solvent Extraction Reagents."

Benedict, C.H., 1917, "Ammonia Leaching of Calumet Tails," *Eng. & Min. J.,* Vol. 104, pp. 43–48.

Birch, C.P., 1974, "The Evaluation of the New Copper Extractant P–1," *Proceedings of the International Solvent Extraction Conference,* ISEC 74, Lyon, September.

Dalton, R.F., K.J. Severs, and G. Stephens, 1986, "Advances in Solvent Extraction of Copper by Optimized Use of Modifiers," *Proceedings for IMM's Mining Latin America Conference, Chile,* November.

Dement, E.R., and C.R. Merigold, 1970, "A Progress Report on the Liquid Ion Exchange of Copper," AIME Annual Meeting, Denver, February.

Eliasen, R.D., and E. Edmonds Jr., 1974, "The S.E.C. Nickel Process," CIM Bulletin, February 1974, 82–86.

Eng. & Min. J., 1973, "In Clean-Air Copper Production Arbiter Process is First off the Mark," February, pp. 74–75.

Forward, F.A., 1953, "Ammonia Pressure Leach Process for Recovering Nickel, Copper and Cobalt From Sheritt Gordon Nickel Sulfide Concentrate," Trans. CIMM, Vol. LVI, pp. 363–370.

General Mills Chemicals, 1975, "Preliminary Report: LIX 54—A New Reagent for Metal Extractions from Ammoniacal Solutions," Tucson, Arizona.

Hartlage, J.A., and A.D. Cronberg, 1973, "Chemical and Physical Factors to be Evaluated Upon Designing a Kelex Extraction System," CIM Conference of Metallurgists, Quebec City, August.

Hopkin, W., et al., 1973, “The Copper Recovery Plants of Commonwealth Smelting Limited and Mitsui Mining and Smelting Co. Limited,” *Hydrometallurgy: Research and Plant Practice,* K. Osseao-Asare and J.D. Miller (Eds.), Warrandale, PA: AIME Annual Meeting, pp. 985–999.

House, J.E., 1981, “The Development of the LIX Reagents,” 1981 Gaudin Lecture, AIME Annual Meeting, Chicago, February.

House, J.E., 1996, Private conversation.

Kordosky, G.A., J.M. Sierakoski, and J.E. House, 1983, “The LIX 860 Series: Unmodified Copper Extraction Reagents,” *Proceedings of the International Solvent Extraction Conference,* ISEC 83, (Denver CO, American Institute of Chemical Engineers), pp. 191–192.

Pincheira, A., Alvarez, G., Garcia, M., Reghezza, A., and Vergara, J., 1997, “Planta SX-EW de Chuquicamata,” LIX Reagent Users Conference, Iquique, Chile, Oct. 20–23.

Power, K.L., 1970, “Operation of the First Commercial Liquid Ion Exchange and Electrowinning Plant,” AIME Annual Meeting, Denver, February.

Reid, J.G., and M.J. Price, 1993, “Separation and Recovery of Nickel and Cobalt in Ammoniacal Sytems—Process Development,” *Proceedings of the International Solvent Extraction Conference,* ISEC 93 (York, England), SCI, London, 1, 159–166.

Reid, J.G., and M.J. Price, 1993, “Ammoniacal Solvent Extraction at Queensland Nickel—Process Installation and Operation,” *Proceedings of the International Solvent Extraction Conference,* ISEC 93 (York, England), SCI, London, 1, 225–231.

Swanson, R.R., and D.W. Agers, 1964, “A New Reagent for the Extraction of Copper,” AIME Annual Meeting, New York, February.

Tumilty, J.A., R.F. Dalton, and J.P. Massam, 1977, “The Acorga P–5000 Series: A Novel Range of Solvent Extraction Reagents for Copper,” The Institution of Mining and Metallurgy, London, Advances in Extractive Metallurgy.

Tumilty, J.A., G.W. Seward, and J.P. Massam, 1979, “The Acorga P–5000 Series in the Solvent Extraction of Copper: Performance Characteristics and Implications for Plant Performance,” *Proceedings of the International Solvent Extraction Conference,* ISEC 77, CIM Special Volume 21.

van der Zeeuw, A.J., 1975, Symposium on Hydrometallurgy of the Institution of Chemical Engineers, Manchester, Institution of Chemical Engineers, Symposium Series No. 42, p. 161.

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Do's and Don'ts of SX Plant Design and Operation

J.M. Keane,[*] V.H. Ness,[†] and G.V. Jergensen[‡]

INTRODUCTION

A first look at a solvent extraction circuit seems to reveal very little. Plant layouts are usually nice and tidy—there seems to be very little bustle of operators and the process equipment isn't obviously turning, gurgling, rattling, or shaking. It's like being on a movie set, or watching paint dry or observing events in geological time. Even in theory, the extraction concepts seem beguilingly simple. In practice, however, the application of solvent extraction technology depends upon faithfully implementing and monitoring a myriad of interactive details.

This article offers a checklist to consider the many nuances in understanding, designing, and managing the extraction process. Be forewarned, however, that this checklist is neither all-inclusive nor entirely site-specific.

PROJECT MANAGEMENT

A leaching, solvent extraction, and electrowinning project depends upon the management and planning skills of a large group of process specialists and engineers. The coordination of these skills also requires "heads-up" management and experience in the human factors part of the business. A few reminders for managers include:

- Carefully define project objectives. Money spent on up-front planning to establish goals, design criteria, specifications, personnel qualifications, budgets, and schedules will send the message to all contributors as to the ultimate intent of the project. Vague directives will create drift, importune politics, indecision, and avoidable mistakes. Be as specific as possible.

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[*] President, KD Engineering, Inc., Tucson, Ariz.

[†] Senior Process Engineer, Kilborn SNC Lavalin, Inc., Denver, Colo.

[‡] Senior Associate, SRK Consultants, Denver, Colo.

- Carefully define the responsibilities of all participants. Spend the time to communicate these reponsibilities. Create the human environment that rewards teamwork and performance.
- Keep records. Detailed records of performance will provide the database essential to solving unforeseen problems.

PROCESS

The SX process is a conceptually simple yet elegant concept. Still, there are many details that must be considered to successfully apply the process concepts.

- Do run thorough lab scale tests to develop design criteria. Simple circuits may require only very basic testing done by reagent vendors. More complex circuits may require detailed testing. Sometimes, pilot scale test heaps complete with solvent extraction and electrowinning circuits are necessary to define the project or to observe and understand current operations.
- Be sure to sample frequently and thoroughly.
- Don't ignore reality. Testwork should be performed with water from the anticipated project source. This is especially true for projects located in arid environments that often have "briney" make-up water.
- It is important to fully define equilibrium levels of deleterious ions (chloride, bromide, iodide, manganese, nitrate, and molybdenum) in recirculating leach solutions and electrolyte.
- Assume that process conditions WILL vary. Provide flexibility to adjust reagents, flow rates, and recycle rates. Be sure to consider the probable duration and frequency of upset conditions.
- Do consider possible changes in ambient conditions, especially ambient temperatures and process solution temperatures. The impact of the dry and wet seasons should always be anticipated. Substantial variation in the grade of pregnant leach solutions and changes in suspended solids should be expected.
- Develop working models of the process and frequently compare the model to laboratory and actual conditions.

REAGENT HANDLING AND STORAGE

The solvent extraction process utilizes a number of reagents that must be stored and handled with great care. Both facilities and reagent handling techniques should be carefully designed. Some suggestions include:

- Provide reagent preparation facilities that include adequate storage and reasonable access for mixing and storage.
- Monitor reagent concentrations and quality at least on a daily basis.
- Concentrated sulfuric acid is dense and abrasive and reacts strongly with many common substances in the normal plant environment, including water. Specialized storage facilities should be designed with this in mind.

- Provide for addition of strong acid into a large and turbulent flow for adequate mixing and dispersal. Don't add acid to ponds even though ponds provide a very large sink for heat dispersal.
- Always check new deliveries of extractant and diluent for quality before accepting shipment.

MATERIALS OF CONSTRUCTION

The process solutions utilized in solvent extraction are generally very aggressive and require care in selection of the materials of construction for trouble-free operating experience. Some considerations include:

- Carefully specify and chose materials of construction from concrete to piping and valves to tank and pond liners. Particular attention must be paid to areas that may be subjected to acid spills such as the concentrated acid handling areas and the tankhouse.
- Don't shy from "exotic" materials. Stainless steel liners in mixers and settlers are usually the most cost-effective option. As well, the ruggedness and strength of metal when encountering large pipe wrenches and other mechanized maintenance devices may be the difference between on-line productivity and downtime.
- However, do consider HDPE or similar materials for ponds, pad linings, launder linings, tanks, and reaction vessels, but be sure to carefully monitor the specifications, specify the vendors, and monitor construction and operating requirements.
- Be sure to implement a thorough QC/QA program during construction. Carefully inspect the condition of materials and equipment at every opportunity.

PIPING

While the mixer-settler is the heart of the operation, piping represents the interconnecting arteries and delivery systems. Apply care and judgment when designing, operating, and maintaining process piping.

- Carefully specify pumping requirements, especially for organic flows and aqueous/organic mixtures.
- While bends, elbows, valves, and other possible constrictions are needed to minimize (or optimize) layouts and operating flexibility, simplify piping as much as possible.
- Also be sure to consider possible siphon effects and entrance and discharge losses in the piping systems. Also be sure to determine when and where vents are required, especially in long pipe runs.
- HDPE is acceptable for many piping applications. However, do consider structural support, impact resistance, and strength requirements for process piping. Also consider pipe strength where elevated temperatures may be encountered.
- Minimize solution flow velocities but be sure to provide adequate head to maintain minimum velocities without pumping wherever possible. Pipe velocity concerns apply mainly to interstage piping where the pump-mixer is providing the pumping head. Elsewhere, conventional velocities may apply.
- Long runs of plastic pipe for organic phase flows should have provisions for grounding to prevent build-up of static electricity.

- Also ensure that electrolyte piping to the tank house is arranged to minimize electrical losses to ground and to short circuits.
- Provide drains for all tankage as well as low points in exposed process piping.
- Arrange for peer reviews of the piping and other plant layouts to ensure that the designs are operable and maintainable.

LAYOUT

The efficiency of the solvent extraction process depends upon the ability to handle large flows of solution containing relatively low metal values, particularly with the pregnant leach solution. As with reagent handling, materials selection, and piping design, the plant layouts must consider the aggressive nature of the process fluids.

- Utilize gravity to the fullest extent possible. However, carefully arrange gravity and suction lines to avoid drawing air into the lines. Two phase chemistry is difficult enough without adding a third stage.
- Design for maintenance. Provide adequate area and elevation for access and maintenance of tanks, platforms, mechanical equipment, and piping.
- Minimize the severity of pumping requirements, especially for aqueous/organic mixtures.
- Protect facilities from damage by mobile equipment, operator carelessness, and ambient extremes.
- Provide adequate containment in the event of inadvertent spills or material failures.
- Be sure to allow for possible modifications to flowsheet to accommodate greater flows or changes in process circuit from parallel to series operations.
- Provide adequate lighting for observation of operations and for maintenance and control.
- Minimize possible sources of solids contamination. Cover large tanks and minimize open-top tanks. Also be aware of the seasonal effects on solution quality.
- Carefully design a crud-removal system. Also consider means of preventing solids ingress such as adding noninterfering flocculents to pregnant leach solutions and PLS storage ponds.
- Carefully design settlers to maintain even flow with adequate time for phase disengagement.
- Use a zoned layout to segregate areas that require organic handling. This will simplify provisions for fire protection.
- Specify cost-effective buildings and covers. There are few structural requirements and large crane capacity for overhead lifts are not severe for most tanks, settling vessels, pumps, and mixers. As well, enclosures can be economically constructed around tank farms for cold ambient conditions. Conversely, covers and shade can be provided to protect against hot desert sunshine.

EQUIPMENT SELECTION

The equipment for most solvent extraction processes is relatively simple, the major items being pumps and mixers. The other important items are the filter system and the electrolyte heat exchanger. Generally, conservative equipment selection does not substantially compound capital cost requirements. However, trade-off studies are always suggested. Other considerations include:

- Be conservative in equipment selection, but know what your design factors are. Distinguish between factors for design and availability but don't compound factors for unlikely events (unless cost is no object).
- Consider electrical hazards of organic solvents and vapors. TEFC and chemical resistant motors are a must.
- Use direct coupled drives wherever possible.
- Carefully specify filters and other means to maintain the highest solution qualities possible.

INSTRUMENTATION AND CONTROL

Most simple SX plants are designed for normal operation with two or three people per shift, one of whom is the control room operator who would also oversee the related operations upstream and downstream of the SX plant. The control philosophy should reflect the complexity and the size of the operation.

- Provide adequate instrumentation for information to advise operators of plant conditions.
- Carefully evaluate economics of automation and the extent of automation.
- Consider providing automated sampling to monitor process conditions.

OPERATIONS

Like all operations, efficiency is a direct result of well-trained and conscientious operators with a commitment to attention to detail.

- Do provide thorough and continuing training for all operating staff. Notice to managers: Be there when you can. You may learn something. You will also send a message that all aspects of the operation can always be improved. Be part of the solution.
- Continually communicate the objectives of the operation. Goals are important, they are especially effective when the operating staff understands them and believes in them.
- Do decentralize decision-making, as quick and decisive action may sometimes be required.
- Do hold frequent staff meetings to consider operating trends and problems. Meetings should always be considered as training opportunities for all personnel.
- Make safety the highest priority. The plant contains organic solvents and flammable vapors which must always be considered in the design of the fire suppression and lightning protection systems and in emergency response plans.

LABORATORY

Operating conditions in a solvent extraction process usually change very slowly. Trends in PLS grade, organic degradation, and crud formation generally occur over several shifts, days, weeks, and even months.

- Provide a well-equipped laboratory to monitor reagent conditions, process responses, and plant performance.
- Run daily shake-flask tests on actual pregnant solutions. Daily tests are mainly for tracking of the extraction performance and quality of the organic phase. Daily tests should measure peak loading capacity, barren organic metal tenor, and phase separation characteristics.
- Be sure that laboratory personnel adequately understand the process, the need for quality sampling, the analytical technique, and the need for timely information.
- Develop comprehensive written procedures of analytical methods and protocols. Enforce them diligently.

Apply these comments and suggestions with common sense and good judgment. Remember that every operation is unique.

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Plant Description and Operation of Heap Leaching, Solvent Extraction, and Electrowinning of Copper at Minera Michilla

Freddy Aroca[*]

CATHODE PROJECT, STAGES I AND II

Minera Michilla S.A. was formed on December 31, 1993, by the fusion of the existing Compañía Minera Lince Ltda. and Compañía Minera Carolina de Michilla.

The cathode project, stages I and II, belongs to Minera Michilla S.A., which is part of the Luksic Group of Companies, and is located in the Michilla District of the II Region of Chile, 110 kilometers north of the city of Antofagasta, 80 Km south of Tocopilla, and 15 Km from the Pacific Coast, in front of the small Michilla Bay.

The project was developed in two stages: the first stage was commissioned in January 1992 and the second stage was commissioned in September 1994.

The realization of the cathode project required a total investment of US $130,000,000 with US $63,230,000 invested in the first stage and US $67,014,000 in the second stage.

All development and engineering work carried out on stage I and stage II of the Cathode Project was done entirely by Chilean companies and professionals. Stage I required 120,000 man-hours, and stage II required 111,000 man-hours.

First Stage

The engineering project was initiated, in mid-1988, to develop the open cast mine, with crushing plant, agglomeration plant, heap leaching, and solvent extraction, followed by electrowinning, with the final objective of producing copper cathodes. The initial development of the open pit, construction of the processing plant and installation of machinery began in early 1991.

The plant was designed to produce 19,700 tons per annum of fine copper in the form of cathodes.

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[*] Superintendent, Metallurgical Research, Antofagasta, Chile.

The Development of Lince Mine. It was decided to exploit the mine as an open cast mine using benches with a height of 10 meters. Development was done using 6½″ drilling equipment, and blasting with ANFO. Front-end loaders with a capacity of 13 cubic yards and trucks with a capacity of 45 tons were used in the charging and transport processes.

The extraction rate was 1,600,000 tons per annum, with a copper grade of 1.56%. The mineral was composed of 90% copper in the oxidized state.

5,200,000 tons of overburden was removed, previous to the exploitation of the mineral.

Crushing Plant. The crusher was designed in 3 stages. The primary crusher is a jaw crusher of 48″ × 60″ with a capacity of 600 tons/hour, which produces material with a maximum size of 8″. The mineral is transported by conveyor belt and deposited in stockpiles of 10,000 tons. The secondary crusher is a hidrocone 1260, with a diameter of 5″, which reduces material from the primary crusher to a final size of less than 2″. The tertiary crushing system is made up of three, 5″ hidrocone 460 crushers working with three vibrating screens in a closed circuit to produce a final product, 100% less than ⅜″. The secondary and tertiary crushing systems have a combined capacity of 320 tons per hour.

Agglomeration Plant. The –⅜″ mineral particle are stored in 2 silos, each with a capacity of 3,200 tons, which feed a 2,4 meter diameter agglomerating drum with a capacity of 320 tons/hour. The agglomeration process is carried out by adding 35/40 Kg/ton of concentrate sulfuric acid mixed with 6% by weight of seawater. The agglomerate produced is discharged into 25-ton trucks for transport to the heap leaching plant.

Heap Leaching Plant. A self-propelled unit, manufactured in Chile, receives the agglomerate, which is discharged directly via a hopper to the stacker, which then forms heaps 4 meters high. The heaps are irrigated, by spraying with rafinate returned from the solvent extraction plant. The heaps are permanent structures and remain intact during the irrigation process which lasts one month. During the leaching process, the leached liquor passes to a pregnant liquor storage pond, with a capacity of 20,000 m^3.

Solvent Extraction Plant (SX). The copper contained in the low-acid, copper-rich liquid produced from leaching is extracted from the weak acid solution using an organic reagent, which reacts in a very specific manner with copper. The operation is carried out in two stages, in two mixing/decantation tanks.

When the organic copper compound is formed, it is brought in contact with a highly acid solution (150 grams H^+ per liter), and all the copper in the organic layer passes into the highly acid solution. This process of returning copper to an acidic solution is called re-extraction, and the solution formed has a high acid and copper content and is used as electrolyte in the next processing stage of electrowinning. The organic layer stripped of copper is then returned for use again in the extraction process, thus completing a closed-loop circuit.

The solution, after passing through two extraction processes, still has a residual amount of copper and acidity, so it is recirculated to the heap leaching plant for use in the irrigation of the heaps. This solution is called rafinate.

The major contaminant in the process is chloride, which is contained in the seawater used in agglomeration, and also in the mineral, which is mostly composed of atacamite. The chloride can reach concentrations of 50,000 ppm during the equilibrium of the solutions during extraction. To eliminate the chloride, pure demineralized water is used to wash the organic layer which is loaded with copper. This process is carried out in a mixing/decantation tank.

Electrowinning. The recovery of the copper dissolved in the concentrated electrolyte received from the solvent extraction plant is achieved by passing an electric current from an anode composed of an insoluble lead alloy to a cathode made of a thin sheet of pure copper, called the starter sheet, which is produced in the same plant using titanium anodes.

The tank house contains 90 electrolytic cells, which are manufactured in Chile. The cells are constructed of polymeric concrete, and nine of these cells are used to produce starter sheets. The current density used in the electrowinning process is 200 A/m^2.

Two, 15,000 A, 200 V rectifiers are used to supply the continuous current, required to carry out the process of copper deposition on the cathodes.

Utilities.

Electric Power. A transmission line of 110 kV stretches 74 Km from the power plant at Mejillones to the Lince Plant. The system also includes a power delivery substation at Mejillones and a reception substation at Lince. The power distributed to the various unit is 23 kV.

Sea Water. A pumping system has been installed at the coast, with a pumping capacity of 3100 m^3/day, to pump sea water to the Lince Plant, which is situated 835 above sea level.

Process Water. The company imported a demineralization plant, which uses state-of-the-art technology to produce ultrapure water from sea water (< 4 ppm total dissolved solids) for use in electrowinning, washing of the organic layer in S/X and drinking water. The demineralization system uses evaporation under vacuum and produces 500 m^3/day.

Second Stage

General Aspects. Immediately after completing stage I, studies were commenced on the implementation of stage II. With stage II in mind, several of the installations in stage I had been designed with dimensions much greater than required, where the additional amount of money required was not significant. An analysis was made of the various alternatives of increased capacity, and, in view of the success of the first stage, the decision was made to increase cathode production from 19,700 tons/year to 50,000 tons/year. To achieve this increase in production, the plant would need to treat 3,500,000 tons/year of mineral with a grade of 1,8% Cu.

The decision was taken to start the expansion work on October 1, 1992.

As part of the total expansion, it was decided to carry out a partial expansion, amounting to 15% of the total, which would involve very little investment and would be quickly completed. This part of the second stage was commissioned in March 1993, increasing the production capacity of cathodes to 22,700 tons/year. Consequently, the implementation of the rest of stage II would mean an increase in the production of cathodes to 27,300 tons/year.

Mining. In addition to developing the open pit at Lince, the decision was made to develop Susana Mine, which is an underground mine, and to continue exploiting several small mines in the district using contractors to supply additional oxide minerals to Michilla. An evaluation showed the mines had sufficient reserves of mineral of average grade to last 7½ years.

It is also well known that there is an excellent potential for further mining developments in the district, which could increase the life of the mines for several more years.

Crushing Plant. To process the maximum tonnage of our own mineral, the decision was made to enlarge the existing crushing plant, changing the existing secondary crushing system for another of greater capacity. A Nordberg HP 700 was chosen for the new secondary crushing system, together with a fourth vibrating screen, to bring the capacity for the secondary crusher to the required level of 440 tons/hour.

In addition, it was decided that a new, separate plant should be constructed to process exclusively ore delivered by contractors exploiting small mines. This plant will have a capacity of 150 tons/hour and will include automatic weighing and sampling systems to control and pay for all ore delivered. This new plant will also operate with three stages: primary crushing, using a 25″× 50″jaw crusher with a grizzly feeder at a capacity of 300 tons/hour; secondary crushing, using a hidrocone 1260 with the same capacity; and tertiary crushing using a 5½″short head. The primary and secondary crushers have been designed with greater capacity to minimize the waiting time of contractor trucks.

Agglomeration Plant. A second agglomerating drum, with the same dimensions and capacity as the original drum, was installed.

Heap Leaching Plant. A second self-propelled stacker with a 400 tons/hour capacity was purchased and put into the system.

The leached solutions obtained from the second group of heaps are routed to a system of two storage ponds of copper rich liquor linked with the original storage pond.

Solvent Extraction Plant. The new plant has a circuit completely independent of the original plant and has the same characteristics, except that the mixing/decantation tanks have greater capacity and have incorporated some new features which have improved their efficiency. The most important of the new features are:

- An increase in the number of separating barriers in the separating tanks to obtain a better separation between the organic and aqueous solutions.
- A new design of the organic recirculation tank to give a longer residence time and separate the last traces of entrained aqueous solution, which contains chlorine.

Electrowinning Plant. The plant contains 104 cells with the same dimensions and characteristics as the original plant. In the initial partial enlargement process, of 15%, 14 additional cells were added. In the cathode deposition process, a new process was introduced. Permanent stainless steel cathodes, 3,2 mm thick, which deposit copper on both sides of the cathode, were used instead of the starter sheets used in the original plant. The new process has the additional advantage that the copper plates formed on both sides of the cathode can be easily stripped using a new automated/continuous stripping machine which has been installed as part of the new plant.

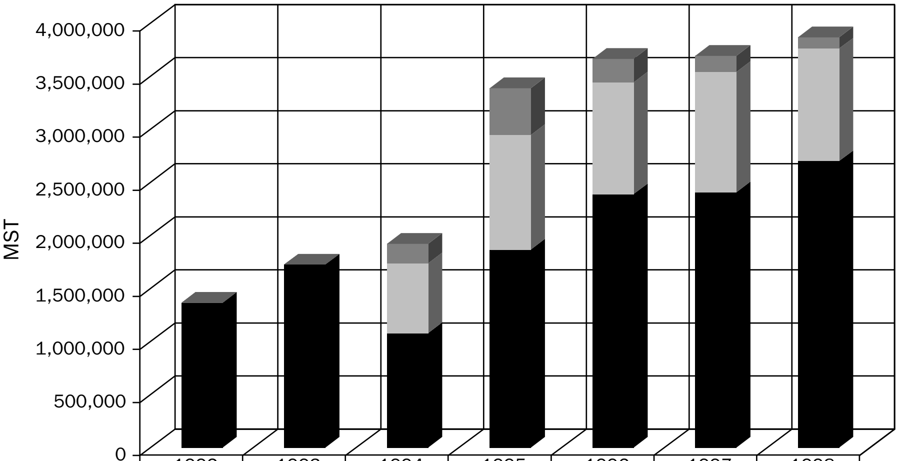
Utilities.

Electric Power. The existing transmission line has sufficient capacity to supply the additional energy required by the second stage, with the result that the second stage only required the installation of an additional substation with the same capacity as the first stage substation.

Sea Water. A second pipeline and pumping system was installed to double the capacity of the system.

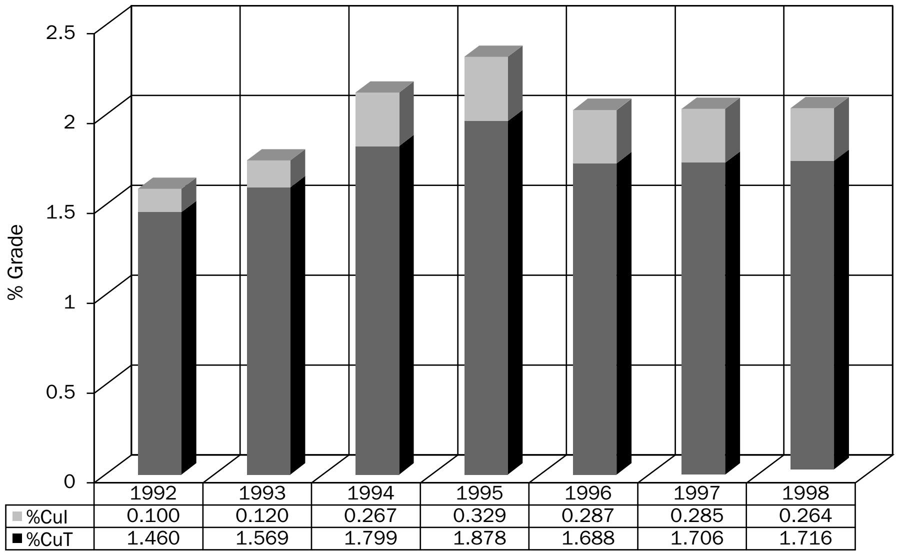
Demineralized Water. A second demineralization plant was purchased with the same characteristics as the original.

Results Obtained



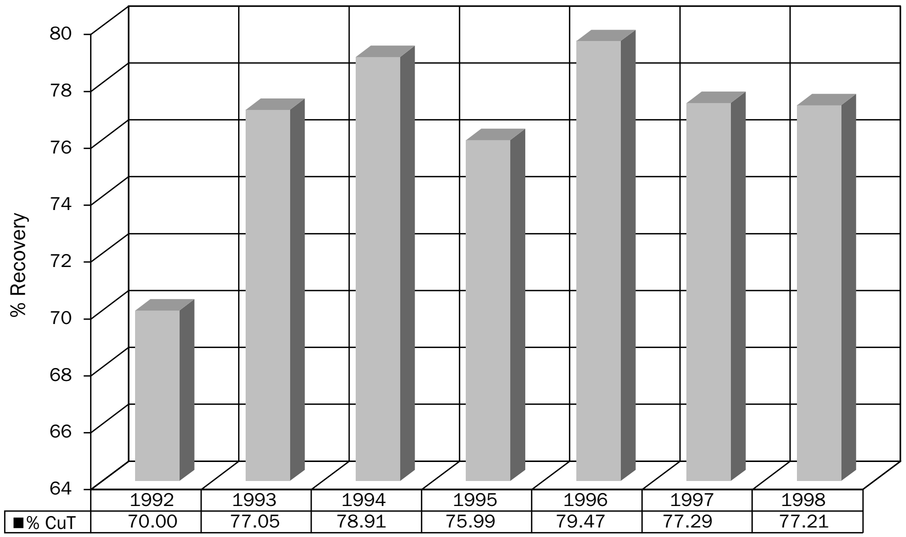
	1992	1993	1994	1995	1996	1997	1998
▒ Estefania	0	0	184,009	439,716	221,590	152,714	105,949
░ Oxido Terceros	0	0	656,170	1,080,214	1,052,292	1,133,890	1,061,041
■ Oxido Lince	1,365,664	1,725,401	1,080,107	1,866,484	2,389,374	2,406,274	2,699,778

Tons processed—cathode plant



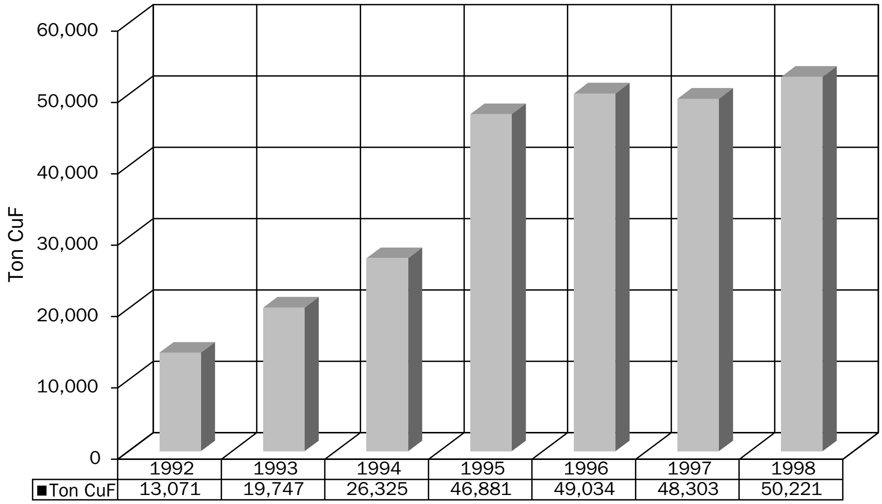
	1992	1993	1994	1995	1996	1997	1998
■ %CuI	0.100	0.120	0.267	0.329	0.287	0.285	0.264
■ %CuT	1.460	1.569	1.799	1.878	1.688	1.706	1.716

Grade of mineral processed—cathode plant



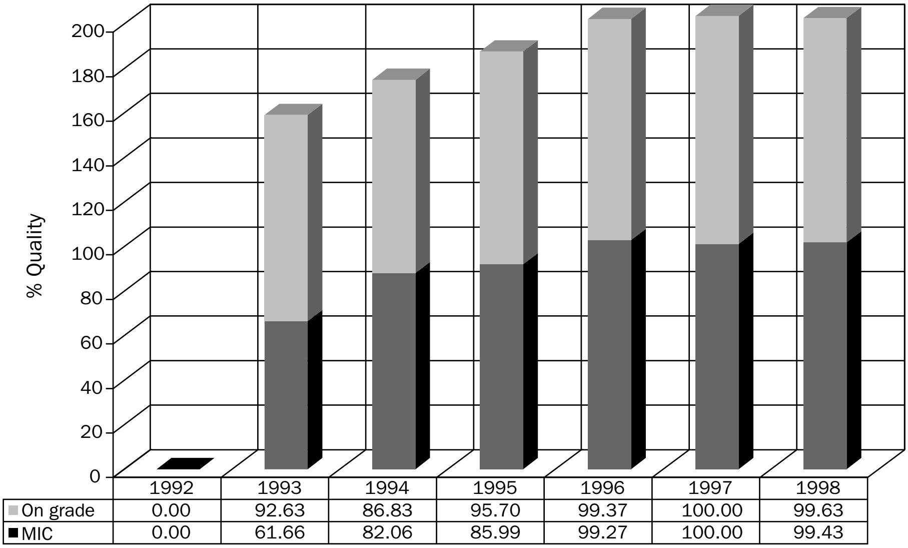
	1992	1993	1994	1995	1996	1997	1998
■ % CuT	70.00	77.05	78.91	75.99	79.47	77.29	77.21

Total copper recovery—cathode plant



	1992	1993	1994	1995	1996	1997	1998
■Ton CuF	13,071	19,747	26,325	46,881	49,034	48,303	50,221

Cathode production



	1992	1993	1994	1995	1996	1997	1998
■On grade	0.00	92.63	86.83	95.70	99.37	100.00	99.63
■MIC	0.00	61.66	82.06	85.99	99.27	100.00	99.43

Cathode quality

Chloride Distribution Study

The object of this exercise is to determine the causes and sources in the metallurgical process which cause the accumulation of chlorine in the commercial circuit and to develop new metallurgical techniques to control this increase of chloride in the circuit. An analyses was carried out on all the variables present in the process to determine the reason for the increase in chloride content and identify the specific points in the commercial circuit where this occurs. The possibility of oxidizing chloride ions to chlorine gas using treatment with ozone was also studied. The object of these studies was to reduce the consumption of cobalt in the cells and minimize the loss of electrolyte.

The study was designed to evaluate the comportment of chloride in all the metallurgical processes: in spent electrolyte, in copper rich electrolyte, commercial electrolyte and R.A.L.

The principal conclusion of this study was that the cause of the increase and accumulation of chloride in the commercial circuit was principally due to the electrochemical conditions which exist in the tank house and not to entrainment of aqueous solution in the organic layer or to inefficient operation of the washing stage.

Definition of the Problem. During the extraction stage, the copper rich organic layer is contaminated with chloride, producing a contamination equivalent to 6 ppm chloride in the organic layer. The washing stage, using demineralized water, is only successful in reducing the chloride level to 1,5 ppm. In the later re-extraction stage of the process, the chloride in the organic layer is transferred to the aqueous spent electrolyte layer, which produces a volumetric increase of chloride in the spent electrolyte to 6 ppm.

On the other hand, the oxidation action at the anode oxidizes chlorine in solution to chlorine gas, and this routine part of the electrochemical process eliminates 95% of the chloride remaining from the washing stage. Discarding the residual electrolyte eliminates the 5% of the original chloride which still remains.

Objectives. A metallurgical study was carried out with the objective of optimizing operations in the SX/EW Plant. The specific objectives of the study were:

1. To develop a method to measure the entrainment of chloride in the circuit and enable us to prepare a metallurgical balance of chloride in the circuit
2. To determine the metallurgical causes which produce the accumulation of chloride in the commercial circuit

Justification for the Study. The advantages of using the new proposed procedure to eliminate chloride, by oxidizing chloride in the electrolyte to chlorine gas, can be summed up in the following points:

1. The reduction of chloride concentration in the electrolyte to give the optimum metallurgical balance for the plant
2. To provide a new alternative for the control of chloride in the circuit
3. A reduction in the amount of electrolyte discarded
4. A reduction in the consumption of cobalt, due to reductions achieved in point no. 3
5. A reduction in the consumption of demineralized water
6. Increased plant capacity

TABLE 1

Process stage	No. 1 Plant	No. 2 Plant
RAL	264	348
EP	39	16
AS 2	47	20
AS 1	47	20
ECC	39	13
OC	6	5
OCL	1	1
OS 1	0	0
OS 2	0	0

Choloride Balance in the SX/EW Circuit. Samples were taken from the organic and electrolyte phases at each stage of the process to determine the chloride profile in the system. However, it was first necessary to develop an analytical method to determine the chloride content in the organic phase.

Details of the analytical procedure developed for this purpose is given in the Appendix to this paper. The results from the sampling exercise are given in Table 1.

Criteria Used in the Chloride Balance. The metallurgical balance of chloride must take into consideration the electrowinning stage, since chloride is eliminated at the anode by the oxidation of chloride to chlorine gas. Analytical measurements have shown that there is a difference of between 1 and 2 ppm chloride in the electrolyte entering the commercial cells and the electrolyte leaving after electrolyses.

$$\text{Difference (ppm)} = F_{ecc} \cdot (Ch_{ecc} - Cl_{scc}) + F_{asi} \cdot (CL_{ecp} - Cl_{scp})$$

$$\text{Difference (ppm)} = F_{ep} \cdot (Cl_{as} - Cl_{ep})$$

Abbreviations used

RAL = Recirculating washing water
EP = Spent electrolyte
DEP = Discarded spent electrolyte
AS2 = Electrolyte from S2
ASI = Copper rich electrolyte
ECC = Electrolyte in commercial cells
OC = Copper rich organic layer
OCL = Copper rich organic layer, washed with demineralized water
OSI = Partially stripped organic layer
OS2 = Totally stripped organic layer
EECC = Electrolyte entering commercial cells
ESCC = Electrolyte leaving commercial cells
EECP = Electrolyte entering splitting cells
SR = Copper rich solution
F = Flow
CL = Chloride

The concentration of chloride in the copper rich solution is 50,000 ppm.

Choride balance in No. 1 plant:

$$F_{OC} \cdot Cl_{OC} = F_{RAL} \cdot Cl_{RAL} + F_{EP} \cdot (Cl_{AS1} - Cl_{EP}) + F_{DEP}\,Cl_{EP}$$

(1) $F_{OC} \cdot Cl_{OC}$ = 8400 LPM · 6 PPM = 50,400 mg Cl/min

(2) $F_{RAL} \cdot Cl_{RAL}$ = 130 LPM · 264 PPM = 34,320 mg Cl/min = 64.6 %

(3) $F_{EP} \cdot (Cl_{AS1} - Cl_{EP})$ = 2200 LPM · (47 − 39 PPM) = 17,600 mg Cl/min = 33.2 %

(4) $F_{DEP}\,Cl_{EP}$ = 30 LPM · 39 PPM = 1,170 mg Cl/min = 2.2 %

(1) ~SUMA (2 + 3 + 4) = 53,090 mg Cl/min = 100 %

Equation 1 represent the mass flow of chloride contained in the copper rich organic phase, which corresponds to 50,400 mg/min. On calculating the mass flow of chloride in the discarded solutions in equations 2, 3 and 4, a mass flow of 53,090 mg/min was obtained. A comparison of the standard deviation between the two methods gave a value of 5%, which is acceptable.

As a result, to determine the chloride balance in the plant, we only have to solve equations 1, 2, 3 and 4.

Equation 2 corresponds to the mass flow of chloride removed during the washing stage. This corresponds to 64.6% of the total mass flow and represents the stripping efficiency of the washing stage. Under normal, stable operating conditions, the efficiency is stable.

Equation 3 corresponds to the mass flow of chloride removed in the electrowinning tank house, that is, it corresponds to the amount of chloride removed by oxidation at the anode. This amounts to 33.2% of the total mass flow. This value varies according to the electrochemical conditions in the tank house, such as current, voltage and acid concentration. For this reason, for the same level of entrainment, with chloride constant, a reduction in the oxidation of chloride in the tank house means an increase of chloride in the electrolyte.

$$\text{Difference (ppm)} = F_{ECC} \cdot (CL_{ECC} - Cl_{SCC}) + F_{AS1} \cdot (CL_{ECP} - CL_{SCP}) \qquad \text{(a)}$$

$$\text{Difference (ppm)} = F_{EP} \cdot (CL_{AS1} - Cl_{EP}) \qquad \text{(b)}$$

Equalizing (a) and (b), and considering $(Cl_{ecc} - Cl_{scc}) = (Cl_{ecp} - Cl_{scp})$, we obtain:

$$(CL_{ECC} - Cl_{SCC}) = [F_{EP}/(F_{ECC} + F_{AS1})] \cdot (CL_{AS1} - Cl_{EP})$$

$$(CL_{ECC} - Cl_{SCC}) = 0.85 \text{ ppm}$$

Equation 4 corresponds to the mass flow of chloride in the electrolyte discarded. This corresponds to 2.2% of the total mass flow of chloride. Therefore, the effect on the total mass balance of chloride of the discarded electrolyte is minimal. For this reason, discarding electrolyte is a very ineffective method for reducing chloride in the electrolyte.

The total mass flow obtained by adding the values in equations 3 and 4 corresponds to the mass flow of chloride, which could not be extracted in the washing stage with demin water.

To calculate the amount of entrained aqueous solution, the general balance equation should be used:

$F_{entrainment} \cdot Cl_{sr} = F_{ral} \cdot Cl_{ral} + F_{ep} \cdot (Cl_{asi} - Cl_{ep}) + F_{dep}\, Cl_{ep}$

$F_{entrainment} \cdot 50{,}000 \text{ ppm} = 53{,}090 \text{ mg/min}$

$F_{entrainment} = 1.06 \text{ lt/min}$

Aqueous Entrainment (ppm) $= F_{entrainment} \cdot P\text{sol} \cdot 10^6 / \text{F oc}$

$= 159$ ppm aqueous

P sol $= 1{,}25$ grms/cc

Appendix

The following is a list of the analytical procedures used by Minera Michilla's quality control department to measure chloride in minerals and plant liquors.

Method 1—The Indirect Determination of Chloride in Solution by Atomic Absorption Spectrophotometry.

Scope of Method. This method specifies a procedure for the indirect determination of chloride in solutions and can be applied to solutions containing chloride concentrations up to 500 ppm.

Principle of Process. The chloride present in plant liquors is reacted with standard silver nitrate solution to produce a precipitate of silver chloride, leaving an excess of unreacted silver, which is then measured by atomic absorption, using an air acetylene flame, where the absorbance of silver in the solution is compared with the absorbance obtained from standard silver solution of known concentration.

Apparatus and Materials.

- Reagents: All reagents used in the analysis are analytical grade reagents, and all water used is pure distilled water.
  1. 65% nitric acid, s.g 1.40
  2. Silver matrix matching solution with 1.0 grm/lt of copper
  3. Silver matched calibration solution (100 ug/ml)—prepare by diluting the copper matrix solution with 10 times the volume of silver nitrate solution prepared in no. 4, then add sufficient analar Hcl to give a 1% solution
  4. Analytical grade silver nitrate (5,000 ppm AG)—weigh 158 grms into a beaker, dissolve in chloride-free distilled water, and dilute to 20 litres with distilled water
- Materials: Normal laboratory equipment. All materials used in the analytical technique must be carefully washed and dried in a drying oven and allowed to cool at room temperature.
- Equipment: Atomic absorption spectrophotometer fitted with air/acetylene burner and a silver hollow cathode lamp.

Calibration Solution. Add 0, 2.5, 5.0 ml of stock silver nitrate solution (concentration 100 ug/ml Ag) to a series of 100 ml volumetric flasks using a pipette, and then add 10 ml of conc HNO_3 to each flask, make up to the mark with distilled water and shake to mix. The solutions obtained contain 0, 2.5, 5.0 ug/ml Ag and are used to calibrate the A/A spectrophotometer.

Procedure. Add 2 ml of conc HNO_3 to a 100 ml volumetric flask and 10 ml of 5000 ppm silver nitrate, then add 10 ml of the test solution using a pipette, dilute to the mark with chloride free distilled water and shake to mix. Filter, using a whatman No. 42 paper into a 50 ml beaker. Pipette 2 ml of this solution into a clean 250 ml volumetric flask, dilute to the mark with distilled water and shake thoroughly to mix. Aspirate the solution in the A/A and read the absorbance obtained for silver. In the case of RAL solutions, use an aliquot of 10 ml, and follow the same procedure described above.

Setting Up the Spectrophotometer. Switch on the instrument, and install the silver hollow cathode lamp, using the procedures recommended by the manufacturer to obtain the optimum sensitivity and stability of emission at 10 mA. Light the air/acetylene flame, and adjust the flows, as recommended by the manufacturer to obtain an oxidizing flame, and wait a few minutes to allow the burner flame temperature to stabilize. Use the calibration solution with the highest Ag concentration to optimize all the conditions of the equipment.

- Measuring Condition for Silver:

Wave length	328.1
Slit	0.7
Oxidant	Air
Fuel	Acetylene
Type of flame	Oxidizing
Linear range	0–4 ppm

Reading of Samples. Adjust the zero on the instrument, and calibrate with the matched solutions, according to the concentration expected in the sample. Once the equipment has been calibrated, aspirate the test samples and note the absorbancies obtained. Aspirate a standard solution of known silver concentration after every five test samples to ensure the instrument remains stable and that the calibration does not change.

Expression of Results.

$$\text{Ppm Cl} = 500 \text{ Reading} \times \text{F.D.} \times 0.3287$$

$$\begin{array}{l} \text{F.G.} = 0.3287 = \text{Ag} \times \text{AgCl} \times \text{Cl} \\ \qquad\qquad\qquad\qquad\;\; \text{Ag} \quad\; \text{AgCl} \end{array}$$

If a dilution is used, multiply by the corresponding factor.

Method 2—Determination of Chloride in Solution Using the Volhard Volumetric Method.

Scope. This method is applicable to the determination of high concentration of chlorides in solution and is used for solutions containing more than 10 g/l chloride.

Principle. Chloride is determined by taking an aliquot of the test solution and titrating it volumetrically with ammonium thiocyanate solution to determine the amount of excess silver present.

Apparatus and Materials.

- Reagents: All reagents used in the analysis are of analytical grade, and all solutions are prepared with distilled water free from chloride.

1. Analar concentrated nitric acid
2. Analar silver nitrate
3. Silver nitrate solution (approximately 0.1 N $Ag+$)—weigh 85 grms of analar silver nitrate and transfer to a beaker. Dissolve in distilled water free of chlorides, then transfer to a 5-litre volumetric flask, make up to the mark with distilled water, and shake to mix. Standardize the solution using standard titrisol sodium chloride solution.
4. Titrisol NaCl 1.5421 gr/lt
5. 2% analar potassium chromate solution
6. 6% analar ferric sulfate solution
7. Ammonium thiocyanate solution (approximately 0.1 N)—dissolve 38 gr of analar ammonium thiocyanate in a beaker with distilled water, then dilute to the mark with distilled water in a 5-litre volumetric flask, shake to mix, then standardize using 0.1 N $AgNO_3$ solution

- Materials: normal equipment used in analytical laboratories. All equipment must be carefully cleaned and used with care to prevent contamination.

Procedure. Pipette 10 ml of test solution into a 100 ml volumetric flask, dilute to the mark with distilled water, and shake to mix. Pipette 2 ml of this solution into a 250 ml titrating flask, and pipette 10 ml of 0,1 N standard $AgNO_3$, then add 1 ml of ferric indicator, and titrate with ammonium thiocyanate. If the solution turns red immediately on titration, this means that a more dilute solution must be prepared. Titrate until an orange end point is obtained.

Expression of Results.

$$\text{gr/lt Cl}^- = \frac{(T_{Ag} \cdot V_{Ag}) - (T_{scn-} \cdot G_{scn-})}{A} \times 35.453$$

Where:

T_{Ag} = Silver nitrate, titre

V_{Ag} = Volume of AgNO3 used up

T_{scn} = Ammonium thiocyanate titre

V_{scn} = Volume of thiocyanate used up

A = Sample aliquot

Method 3—Gravimetric Determination of Chloride in Minerals.

Scope. This method is applicable to the determination of low levels of chloride in minerals and can measure concentrations in excess of 10 ug/ml.

Principle. The chloride in the mineral is put into solution by agitating with dilute sulfuric acid for 60 minutes. The solution obtained is then filtered, and a portion of the filtrate is taken for the chloride determination by precipitation with silver nitrate to remove all the chloride from the test solution in the form of a precipitate of silver chloride.

Apparatus and Materials.

- Reagents: All reagents used in the analysis are of analytical grade, and all solutions are prepared with distilled water free from chloride.
  1. Analar conc sulfuric acid 95/97%, sg 1.84
  2. Approximately 1 N sulfuric acid solution—dilute conc sulfuric acid with distilled water to give a solution of approximately 49 g/lt
  3. Analar silver nitrate
  4. 20,000 ug/ml Ag+ solution of silver nitrate—weigh exactly 158 grms of analar silver nitrate and dissolve in a beaker with distilled water, then transfer to a volumetric flask of 5 litres, dilute to the mark with distilled water and shake to mix
- Materials: normal equipment used in analytical laboratories. Care should be taken in the preparation and use of these materials to prevent contamination.

Procedure. Weigh exactly 10 grms of test samples, to the nearest 0.1 mg, and transfer it to a 250 ml beaker covered with a watch glass. Add 100 ml of approximately 1 N H_2SO_4. Leach with constant agitation for 60 minutes using a shaker. Remove the flask from the shaker, filter into a beaker, and then pipette 25 ml of the filtrate into a 250 ml beaker, add 5 ml of 20,000 ug/ml Ag+ solution to the beaker using a pipette, 50 ml of chloride free distilled water, and 5 ml of nitric acid. Boil the beaker gently until all the silver chloride has precipitated. The precipitate is white in color. Remove the beaker from the hot plate and allow to cool. Filter the solution under vacuum using a glass Gooch filter crucible with 0.45 micron porosity, which has been previously weighed. Wash the precipitate three times with cold distilled water, and then place the glass crucible in a drying oven at 105°C for 2 hours. Remove the crucible from the oven, and place in a dessicator to cool. When the Gooch crucible is cool, remove it carefully from the dessicator. Weigh accurately to obtain a constant weight, then calculate the chloride content of the test sample.

Expression of Results.

$$\%Cl^{-} = \frac{P_2 - P_1}{P} \times 0.247368(F.G.) \times 100$$

Where:

$P_{test\ sample}$ = Weight of test sample (25 ml of solution is equivalent to 2.5 gr of mineral)

$P_{initial}$ = Initial weight of Gooch crucible

P_{final} = Final weight of Gooch crucible + precipitate

F.G. = Gravimetric factor

$$F.G. = \frac{Cl^{-}}{AgCl} = \frac{35.453}{143.321}$$

Method 4—Determination of Total Chloride in Mineral Using the Volhart Volumetric Method.

Scope. This method is applicable to the determination of chlorides in high concentrations in minerals. The method is used to detect chlorides in solutions with concentrations in excess of 10 grms/litre.

Principle. The chloride in the mineral sample is put into solution by hot leaching with conc analar nitric acid. The solution obtained is filtered, and an aliquot of the filtrate is then taken and transferred to a titrating flask, and titrated with standard thiocyanate solution to determine the excess silver present.

Apparatus and Materials.

- Reagents: Only analytical grade reagents are used in the analysis, and all solutions are prepared using chloride-free distilled water.
  1. Analar conc. nitric acid
  2. Analar silver nitrate
  3. Approx. 0.1 N silver nitrate solution—weigh exactly 85 grms of analar silver nitrate, and dissolve in a beaker with chloride-free distilled water, then transfer the solution to a 5-litre volumetric flask, make up to the mark with distilled water, and shake to mix
  4. Titrisol NaCl 1.5421 gr/lt Cl^-
  5. 2% by volume analar potasium chromate solution—Weigh an equivalent weight of the salt to give a 2% by volume solution of potassium chromate
  6. 6% by volume analar ferric sulfate solution—weigh an equivalent weight of the salt to give a 6% by volume solution of ferric sulfate
  7. Approximately 0.1 N ammonium thiocyanate—weigh 38 grms of analar ammonium thiocyanate, and dissolve in a beaker with chloride-free distilled water, then transfer to a 5-litre volumetric flask, make up to the mark with distilled water, and shake to mix. Standardize the solution using 0.1 N standard $AgNO_3$.
- Materials: normal materials used in an analytical laboratory. All materials must be cleaned and handled with care to prevent contamination.

Procedure. Pipette 10 ml of the test solution into a 100 ml volumetric flask, make up to the mark with chloride-free distilled water, and shake to mix. Pipette 2 ml of solution from the flask into a 250 ml titrating flask, add 10 ml of standardized silver nitrate solution with a pipette, then add 1 ml of ferric indicator, and titrate with standard ammonium thiocyanate solution. If the titration produces an immediate red coloration, a more dilute solution must be prepared from the test solution. Titrate the solution to an orange end point.

Expression of Results.

$$\text{gr/lt Cl}^{-} = \frac{(T_{Ag} \cdot V_{Ag}) - (T_{scn-} \cdot G_{scn-})}{\text{A.M.}} \times 354.53$$

Where:

T_{Ag} = Silver nitrate titre

V_{Ag} = Volume of silver nitrate used up

T_{scn} = Ammonium thiocyanate titre

V_{scn} = Volume of ammonium thiocyanate used up

M = Weight of sample

. . . .

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Note: *f* indicates figure, *t* indicates table.