Ammonia Leaching of Oxide Copper Ores by Samin Limited at Burra SA

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STAGE 1

Synopsis

The importance of the Burra Burra mine to the development of South Australia has been well documented by others, in particular a noted historian Ian Auhl (1). This major high grade oxide copper resource was mined during the period 1845 to 1877 and the mine was commonly referred to as the Monster Mine. During this period 234,648 tonnes of ore was mined at a grade of 22% copper.

The mine remained closed (apart from two minor unsuccessful attempts at revival at the turn of the century and in the 1920's) until Samin Limited was formed in 1969. The opportunity presented itself when Broken Hill South Ltd relinquished the leases they held over the mine and their then Chief Geologist Mr B R Lewis resigned and with acquisition of the Burra leases formed Samin Limited in 1969, with Mr E O Myers and Mr A O Thomas. Poseidon Limited acquired 100% ownership of Samin soon after formation and Lewis and Myers became Joint Managing Directors and Thomas Company Secretary of Poseidon Limited.

It was initially envisaged that the copper recovery method would involve an acid leach and cementation process with the sale of the product copper cement to a smelter. This route was soon rejected due to the presence of a high proportion of acid consuming host rocks. The Australian Mineral Development Laboratories were commissioned to develop a treatment route under the guidance of their Director, Mr N Draper. Mr Draper joined Samin in 1970 as Chief Metallurgist and Director and following the takeover became Chief Metallurgist and Associate Director of Poseidon.

The options of acid leaching and flotation of the ore were soon rejected due to high acid requirements and low copper recoveries, respectively. The ammonia leaching route was selected for development and the process that subsequently evolved (Stage 1) was developed from reports on the Kennecott Copper Corporation's operation at Kennecott in Alaska in the early 1900s and the nickel Nicaro operation in Cuba.

1.0 Introduction

Plate 1 shows the layout of the initial Stage 1 operation and it's relationship to the township of Burra. The main features in the photograph are:

- Central is the early open cut surface development of the Burra Burra mine.
- Below centre is the mine workshop and vehicle holding yard.
- Bottom central and progressing downwards are:
  - Crusher and covered crushed ore storage.
  - Log washer and spiral classifier desliming plant.
  - Murray and Raw water storage tanks.
  - Ore feed conveyers and five low pressure cylindrical leach vessels.
  - Process liquor tank farm and distillation area.
  - Control room, boiler house, store and maintenance workshop.

PLATE 1 Aerial view of Burra Mine and Processing Plant, 1972. PIRSA Photo T010375

  - Copper oxide filtering drying and packaging shed.
  - Bunker C fuel oil storage tanks.
  - Employee change rooms.
  - Administration Office

- Right/middle is the process reject slimes dam.
- Below the slimes dam are waste dumps and wall to the bottom of the photo.
- Below the main waste dump are the leached ore tailings contained by waste walls.
- Above the mine is the township of Burra.
- Left is the township of Burra North.
- The vacant area between the two towns is the site where the smelter used during the 19th century operations was located.

The Stage 1 process operated from early 1971 to October 1973, when it was shut down for the commissioning of the whole of ore treatment process (Stage 2). The writer saw the last 3 months of the Stage 1 operation and the following description is based on observations at the time, verbal discussions with the initial operating crew and reports on the operations. The plant was constructed and commissioned under the guidance of the Manager of Operations, Mr. H J Rich and the Maintenance Engineer Mr. N van Buuren.

2.0 Stage 1 Operations

As stated at the outset, the Stage 1 operation was developed and designed by the Australian Mineral Development Laboratories (AMDEL). In essence it was a direct vat leach process of a deslimed crushed ore developed from operational reports for the Kennecott operation in Alaska during the early 20th century. The Stage 1 plant flow sheet is shown in the Appendix. The technology was developmental with no similar process in operation in the world and understandably numerous problems were encountered. This report discusses and describes the operation and the problems encountered along with variations from the original design.

2.1 Mining & Mineralogy

The simplified geology of the resource (Fig. 1) has been reported by others (2). The Burra ore copper minerals were principally malachite ($CuCO_3.Cu(OH)_2$), azurite ($2CuCO_3.Cu(OH)_2$), and chrysocolla ($CuSiO_3.2H_2O$), with a cuprite (Cu_2O) proportion increasing at depth. Laboratory leaching tests defined that the major copper minerals, except chrysocolla, were highly soluble in ammoniacal liquors. Chrysocolla was insoluble along with all the remaining gangue materials. It was therefore essential that Stage 1 ore mining avoided the high chrysocolla areas in the mine. Alternative processes were being investigated at the time for the future treatment of chrysocolla and fine ores in montmorrillonite clay hosted gangue.

The mining operation was carried out under contract by Roche Bros. under the supervision of Samin geologists. The open cut was developed in 10m benches from

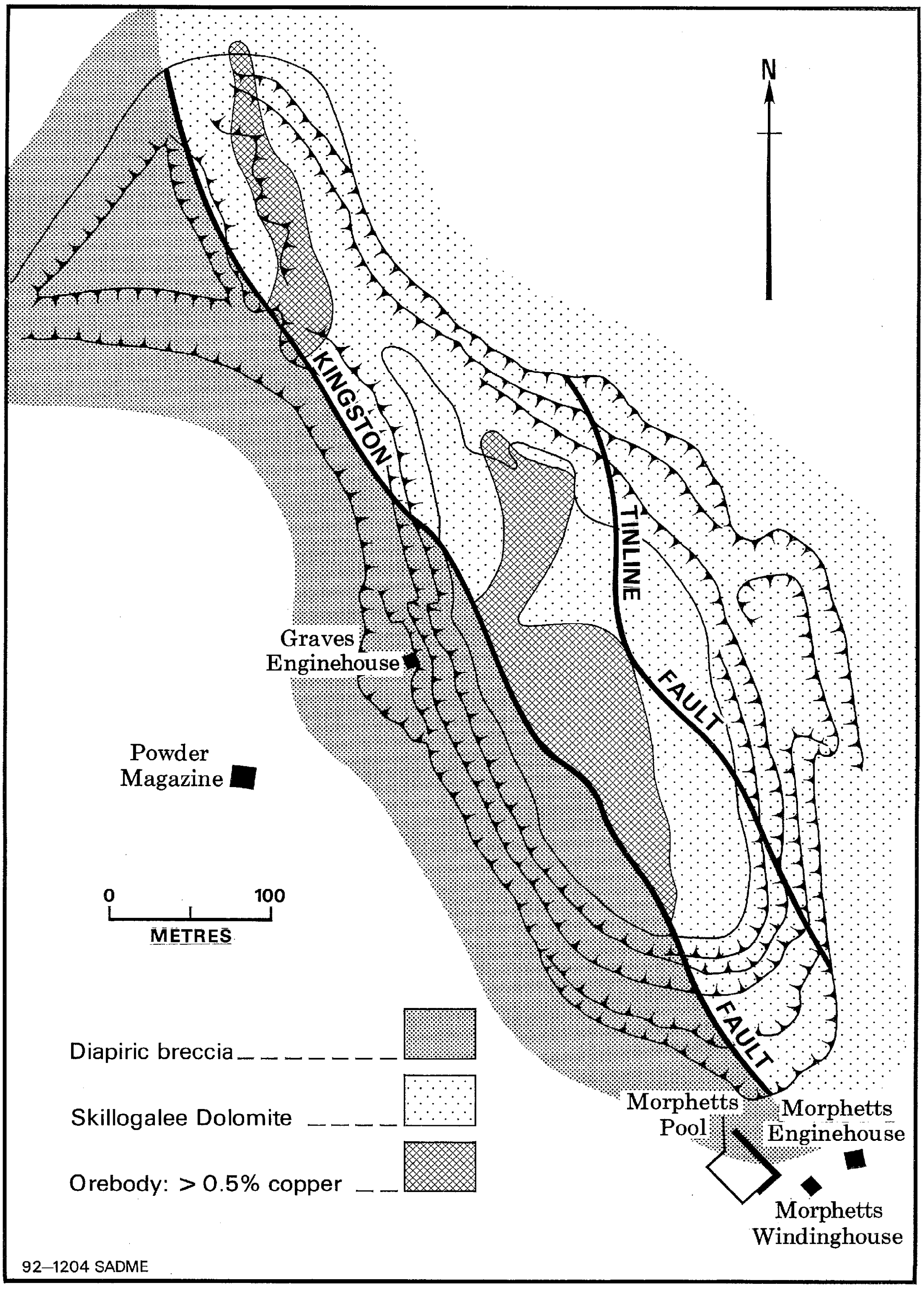

Figure 1 Simplified geological plan of the Burra Mine

512m RL to a depth of 90 meters. A Marion 111M diesel powered face shovel was used throughout the 11 year mining period with ore haulage in 35 tonne EUCLID trucks. Ore was mainly free dig, but most waste required drill and blast. Mining was managed on a 10 hour day and 6 days per week basis.

2.2 Crushing

The initial circuit saw ore from the mine delivered to the crusher at a nominal size of minus 1000mm and truck dumped into a 50 tonne hopper. It was extracted from the hopper using a variable speed 12 ft Linkbelt vibrating feeder, passed over an 8ft by 4 ft Linkbelt scalping screen fitted with a 21/2 inch screen with the oversize passing to a single toggle 42 inch by 30 inch Jacques Jaw crusher. The primary crushed ore was screened on a double deck primary screen fitted with 25mm upper and 10mm lower decks with the plus 10mm ore passing to a Hazemag Type AP4BR impact crusher. Ore delivered to the secondary crusher passed under an electromagnet where numerous metal objects, such as miners picks from the previous mining period, were collected. The secondary crushed ore was screened on a single deck 10mm secondary screen with the oversize returning to the secondary crusher. The minus 10mm ore was conveyed to a 4500 tonne open stockpile. Crushed ore weight was recorded on a belt weightometer and ore samples were taken at this location for metallurgical accounting.

Variations from the original crusher design over time were:

- The crushed ore size was increased from -10 to -12.5mm at an early stage to overcome lost throughput with increasing ore moisture.
- Original design ore moisture was 4 percent at which levels crushing rates of 350 tonnes per hour were achievable. However, ore moisture gradually increased to 8 to 12 percent and at these levels crushing rates reduced to 160 tonnes per hour. Winter moisture of 12 to 18 percent was common with subsequent crushing rates decreasing to less than 100 tonnes per hour.
- The original wire screen decks were replaced with rubber to alleviate blinding.
- Had the design feed moisture been correctly defined a pan feeder would have been selected for the 50 tonne hopper rather than the vibrating option. The vibrating action caused ore bridging in the hopper.
- Rubber liners on the vibrating feeder tray replaced the original steel liners to reduce wear and ore build-up.
- Timber from old mine workings became an increasing problem when it peaked at an estimated 6000 tonnes per annum. A conveyer was installed to the scalping screen where wooden beams weighing up to 200kg were removed by mechanical means. Hand picking of smaller timber was also required from the primary screen feed belt.
- Hopper ore bridging and timber resulted in delays to the mining equipment and subsequent friction between the two work forces. Ore was finally stockpiled in front of the crusher and trammed to the crusher feed hopper using a 966 FEL at the contractors expense.
- The open crushed ore stockpile was covered after the first winter to minimise the impact of rain increasing ore moisture and thus causing further materials handling problems.

- The live stockpile capacity rapidly decreased with time and high moisture and a 920 FEL was required to maintain steady feed rates from the stockpile. Periodically the stockpile was totally depleted and replaced to increase the live capacity.

2.3 Wet Ore Classification

During laboratory leach testwork it was shown that the whole of the ore would not percolate at a sufficient rate to permit vat leaching due to the clay content. It was therefore necessary to deslime the ore prior to leaching. 250 tonnes of deslimed ore was processed daily.

Ore from the crushed ore stockpile was extracted through one of four tunnel vibrating feeders and conveyed to a Warman Log Washer at the start of the wet classification circuit. (Due to the vibration compaction effect at the stockpile the vibrating feeders were finally replaced with variable speed extraction belts.) Density of the log washer pulp was maintained in the 30 to 40 percent solids range to suspend accretions and clay lumps evenly in the pulp to assist mechanical disintegration. The product exiting the log washer was diluted with water to 10 to 15 percent solids and the slurry piped to the feed box of a 1.5m diameter spiral classifier. The operational objective for the spiral classifier was to produce a course product for vat leaching with a nominal size of plus 52 mesh with the minus 52 mesh slimes fraction pumped to the slimes tails dam for future treatment in Stage 2. (Reclaimed water from the tails dam was piped to the reclaim water dam for re-use. Make-up raw water was pumped from the Bon Accord Shaft). The deslimed ore was belt conveyed to the top of the leach tanks and with the use of a tripper car the ore was dropped into the appropriate leach tank via a swirling disc distributor. Ore conveyed to the leach tanks was weighed and sampled for assaying and metallurgical accounting. The slimes slurry was also sampled for accounting purposes.

The classification section was operated on a day shift basis and produced 250 tonnes of deslimed ore daily. The ore split between coarse and slimes was approximately 55:45, respectively, which gave a daily crushed ore requirement of 450 tonnes. During placement of ore into the leach tanks raw water was pumped into the tank base pipe manifold and overflowed through the overflow pipe and gravity drained to the reclaim water dam. This further removed slimes from the ore. On achieving a 250 dmt leach ore charge the classification section was shut down, water pumping into the base of the leach tank ceased and the ore allowed to drain. Operators would then enter the tank and level the surface by hand with shovels. After levelling the ore and completion of water drainage the tank was sealed ready for leaching.

Stage 1 Leaching Statistics

Ore Leached	dmt	196,239
Leach Ore Head Grade	% Cu	1.35
Slimes Discarded	%	45.5
Copper Leach Recovery	%	65.5 (Budget 70%)
Slimes Discarded	dmt	163,829
Slimes Grade	%Cu	0.90

2.4 Leaching

The leach tanks (5) were 22 feet in diameter with a total height of 40 feet. The volume of each tank was approximate 58,000 gallons. The tanks had an operational design pressure rating of +5 to -0.5 psig and were fitted with an hydraulic box at the base for pump discharging of the leached solids to the tailings dam.

The initial leach design called for a counter current leach through the 5 tanks and the process was instrumented for simplicity to control all valve sequencing. The design leach cycle was as follows:

1. Solids filling - 8hrs
2. Flushing with water to remove fines - 16hrs
3. Leach liquor fill - 4hrs
4. Liquor recycle - 12hrs
5. Pregnant liquor to storage - 4hrs
6. Liquor in - 4hrs
7. Liquor transfer - 4hrs
8. Liquor recycle - 16hrs
9. Liquor transfer - 4hrs
10. Liquor transfer - 4hrs
11. Liquor recycle - 16hrs
12. Make-up liquor in - 4hrs
13. Leach liquor drain - 4hrs
14. Tailings wash and drain - 7hrs
15. Tailings steam wash and drain - 7hrs
16. Solids discharge to tails dam - 6hrs

The initial leach liquor was defined as a freshly prepared ammonia solution.

The capital equipment was installed according to design, but was simplified early in the commissioning period due to unexpected chemical and physical problems that arose. These principal problems were:

- The initial leach liquor was a 100g/l NH_3 solution prepared using River Murray mains water and anhydrous ammonia. No copper was leached until the importance of carbon dioxide in the leach liquor was realised. Six tonnes of liquid CO_2 were purchased and introduced to the circuit. Copper leaching commenced immediately and no further carbon dioxide purchases were required as on-going needs were supplied from the carbonates in the copper mineralisation.
- The leach liquors became saturated with CO_2 and solid ammonium carbonate froze in vapour lines at temperatures below the 50 to 60^oC range.
- Leach flows relied on pump curve design and no leach flow meters were installed. Lines could be blocked with ammonium carbonate or worse still valves seized with ammonium carbonate. Seized ball valves would have their shafts sheared by the actuator so visually the actuator was not telling the operator the position of the valve. Hence liquors were pumped every which way.

- Ammonium carbonate also froze in the tank nipples for the pressure transmitters and thus pressure indication was unreliable.

The problems encountered during commissioning were addressed as follows:

- The leach cycle was changed to:
    - Day 1 - Load ore into the leach vessel, drain the water and add leach liquor.
    - Day 2, 3 & 4 - Recirculate leach liquor.
    - Day 5 - Drain pregnant liquor, wash and steam ore and discharge solids to tails.
- Countercurrent leaching and remote valve sequencing was abandoned.
- All interconnecting tank pipework was removed.
- Reverted to manual control which required the operator to activate valves.
- Steam bleeds were fitted at all pressure transmitters to eliminate ammonium carbonate freezing.

The typical leach chemical reaction can be defined as follows.

$$CuCO_3.Cu(OH)_2 + (NH_4)_2CO_3 + 6NH_4OH = 2Cu(NH_3)_4CO_3 + 8H_2O$$

After draining and isolating the leach tanks 12,000 gallons of ammoniacal leach liquor were added. The lids on the tank were bolted shut and recirculation of liquors from top to bottom commenced and continued for 3 days. After the fourth day drain valves were opened on the base of the tank and the copper enriched solution of grades up to 40g/l Cu drained to the 40,000 gallon pregnant liquor storage tank. Approximately 14,000 gallons of pregnant liquor was produced daily. As the draining liquor disappeared below the solid surface in the leach tank 2,000 gallons of Hot Barrens (hot water ex the evaporators at the conclusion of distillation) were added to the top of the tank and saturated steam injection to the surface was introduced at a rate of 1,500 lbs/hr. The steam while condensing would push pregnant liquor and the plug of water to the bottom of the leach tank until it finally broke through at the base. The hot water and steam recovered soluble copper, ammonia and carbon dioxide. The steam on break-through was drawn through a three plate bubble cap with a vacuum pump and into the condensing and gas scrubbing train associated with the evaporators. Break-through steaming was allowed to flow for approximately one hour.

After steaming the solids were ready for discharging to the tailings dam. This was accomplished by opening a high pressure water jet to the hydraulic box at the base of the leach tank cone and combining this water with raw water and pumping the slurry with a 60 HP 6/4 Warman slurry pump through 800 metres of six inch steel Victaulic tailings line. Due to high wear the steel line was gradually replaced with HDPE Victaulic pipes. The schedule was designed to have the solids discharged from the tank by 0700hrs each day when delivery of new leach ore would commence.

Other important design aspects observed during this phase of the process development were:

- Generally the copper tetramine solutions (both cuprous and cupric) were not corrosive to mild steel and the plant was constructed largely using this material. However, at higher temperatures (+70^oC) and at lower copper concentrations the liquors were corrosive to mild steel and 316SS became the recommended material of construction for these liquors. The cause was not well appreciated in this stage of the development, but was later shown to be related to high chloride levels in the recirculated ammoniacal process liquors.
- Solution entering tanks through the side wall showed wall corrosion where the liquor ran down the walls. This corrosion was overcome by extending the pipe 6 inches through the wall into the tank so that the solution would fall on the tank liquor surface.
- Pipework rapidly built up with a calcium/magnesium scale forcing periodic line replacement. The cause was later shown to be associated with the highly variable equilibrium solubility levels for these elements at varying ammonia concentrations. It remained a problem for the duration of Stage 1, but diminished markedly in Stage 2.

2.5 Distillation

Distillation of the process liquors was the method used to produce cupric oxide and recover the ammonia and carbon dioxide for return to leaching. Four shell insulated evaporator vessels were installed with the pipework designed to have three vessels operating in series at all times. The fourth vessel was either filling or emptying. Distillation was therefore a batch operation in relation to the process liquor treatment, but was continuous for distillation and ammonia recovery. The heat source was saturated steam at 60 psig. The typical chemical reactions were:

$$Cu(NH_3)_4CO_3 + 4H_2O + heat = CuCO_3 + 4NH_4OH$$
$$CuCO_3 + heat = CuO + CO_2$$
$$2NH_4OH + CO_2 = (NH_4)_2CO_3 + H_2O$$

2.5.1 Initial Process Design

The initial design called for four 12 ft diameter by 35 ft high evaporators with a vessel design operating pressure rating of 70 psi. They were vertical vessels of approximately 16,000 gallon capacity with a conical base. Steam was introduced through the side of the conical base to an internal manifold with downward facing 25mm nozzles. Steam was metered to the first evaporator at rates in the range 5,000 to 8,000 lbs/hr. The off-gases from the top of the first evaporator passed to the manifold in the second evaporator who's off-gases passed to the third evaporator. The fourth evaporator was on standby and was introduced into the third position, when steaming of the first in line evaporator was complete.

Gases from the third evaporator passed through two condensers in series. The first condenser was used to preheat liquor for the next evaporator charge from 50^oC to 90^oC. The second condenser was a design 20 million Btu/hr tube and shell unit using cooling water in closed circuit with a cooling tower. The condensers were backed with four gas scrubbers in series. The initial scrubber used recirculated cooled condenser condensate and the remaining three operated with recirculating flows of 110 gal/min with a total fresh water input of one gal/min. The recirculating scrubber liquor flow was cooled using plate heat exchangers serviced by water from a cooling water tower. Scrubber input water was cooled Barrens liquor.

Condensate from the evaporators, liquor from the gas scrubbers and wash water from the leach tanks and Barrens were combined to prepare recirculated leach liquor in a 40,000 gallon storage tank. Anhydrous ammonia vapour was bled into the tank through a sparge pipe to make-up for ammonia losses and maintain leach liquor strengths at 100 g/l ammonia. Carbon dioxide make-up was supplied from the oxide copper minerals. All tanks in the tank farm area were vented through the scrubbers.

Steam injection continued into the first evaporator until no ammonia could be smelt in the liquor sampled. The evaporator was then isolated, vented to atmosphere and steam injection was switched to the evaporator previously second in line and the fourth evaporator was brought into the third in line position. The contents of the evaporator were then discharged to a thickener. The overflow of the thickener was pumped to the 40,000 gallon Barrens storage tank for internal process use with excess overflowing to the raw water reclaim pond. The underflow of the thickener was pumped to a vacuum drum filter for solid (cupric oxide)/liquid separation. The initial design called for the sale of the wet cupric oxide (concentrate) to a smelter.

2.5.2 Process Design Changes

Changes to the original design were numerous. Stage 1 was a pilot plant for a final whole of ore treatment plant - Stage 2. Crushing and distillation equipment was envisaged to be suitable for both Stages. Most problems were recognised during Stage 1 operations, but were not always fully understood and rectified until operating well into Stage 2. The major design changes were:

2.5.2.1 Evaporator Steam Injection

The elaborate steam distribution ring in the base of the evaporator soon blocked with dense cupric oxide scale. Steam was also not injected close enough to the base of the evaporator leaving a stagnant “cool” liquid zone where copper carbonate formed. The initial product produced was therefore a copper carbonate/copper oxide mixture with a lower concentrate copper grade. The problem was overcome with the installation of a centrally located six inch Schedule 80 steam spear extending from the top to within 12 inches of the base of the evaporators. This simple installation proved to give a close approximation to the theoretical liquid/vapour phase equilibrium.

2.5.2.2 Evaporator Scale

Scaling in distillation systems in other metallurgical operations has proved to be an operational problem requiring routine shutdowns for cleaning. The Samin designed evaporator experienced minor wall accretions and occasional scale blockages in the cone, but the frequency did not warrant redesign considerations. The units operated for 13 years before internal shell scale cleaning was undertaken. The reason for low scale formation at Samin is believed to be a result of low calcium and magnesium levels in the make-up process Barrens water.

2.5.2.3 Evaporator Steam and Vapour Piping

The initial pipes were rigidly mounted insulated mild steel lines. Pipe line expansion and contraction caused frequent seizure of the stainless steel remote operated 100mm ball valves controlling evaporator sequencing. This led to the installation in the pipe work of stainless steel expansion joints at each valve. There were occasional valve seizures after this alteration, which were readily identified on the evaporator temperature profile chart and rectified.

The evaporators were installed in the open and with the cycling process temperatures, and winter temperatures as low as $0^{o}C$, some vapour condensation occurred. Corrosion in the base of the pipe work was observed and although not severe it resulted in a maintenance and future material of construction review to adopt 316SS for process vapour lines.

2.5.2.4 Condensers

The original evaporator off-gas condenser cooling system had several process and design deficiencies that required resolution to avoid on-going process inefficiencies. There were initially two shell and tube condenser bundles, the former being a smaller unit to preheat the pregnant liquor prior to distillation and the second a 20 million Btu/hr water cooled unit for vapour condensing.

The unit installed for pregnant liquor preheating was bypassed and abandoned early in the commissioning program. Continuous flow through the exchanger could not be practically maintained to preheat the liquor for a batch processing system. Stagnant liquor at times remained in the tubes and boiled by the vapours ex the evaporators. This caused pipe blockages with scale. The tube bundle was removed.

The main condenser had two problems to be addressed namely, maintaining vapour temperatures above $60^{o}C$ to avoid ammonium carbonate freezing and avoiding scaling in the water side of the condenser tube bundle. Actions in this area were as follows:

- With a "cold" evaporator last in line the off-gas vapour temperature would be low and there was always the risk of totally blocking the condenser with solid ammonium carbonate. If this happened the only way to clear the blockage was to apply live steam to the bundle and wait for it to break through the blockage. Operational methods of control tried were:
  - The cooling water flow was reduced to reduce vapour cooling, but this was detrimental for water tube scaling.
  - The cooling water temperature was allowed to rise by manually stopping and starting the cooling tower fans. This was partially successful.
  - Finally the condenser tube bundle was continually irrigated with liquor from the Condensate Liquor Tank, which continually dissolved any solid ammonium carbonate. The cooling tower fans were then only stopped under extreme temperature conditions.

- Scaling on the water side of the condenser tubes was the major problem and took the longest time to resolve. The local plant raw water was high in calcium and magnesium salts with a TDS of 1600 ppm. The original water treatment in the cooling towers was tannin based only for oxygen corrosion protection. Action taken in this area to finally eliminate lost operating time due to cooling water scaling were:

  - With varying water temperatures in the tube bundle calcium and magnesium scaling forced regular plant shutdowns for acid washing. This was effective until heat loads increased with increasing production.
  - Higher temperatures resulted in denser scale which was more refractory to acid washing. High pressure water jet cleaning of individual tubes was introduced which had a good success rate. The down time for pressure cleaning was approximately 16hrs.
  - With the high down time associated with pressure jet cleaning a second tube bundle was purchased for change over and cleaning during normal operating time.
  - The tube bundle was of the hair pin design and cleaning of refractory scale in the hair pin was difficult. Catoleum water treatment personnel were invited to assist in selecting a change in the chemical dosage scale control. A two step dosage procedure was selected which allowed precipitation of salts to continue. Precipitated salts remained in suspension and did not adhere to the tube surface. Solids would settle to the floor of the cooling tower.
  - The tube bundle head was designed for eight water passes, which presented the same cooling water to the condenser 8 times before it was returned to the cooling tower. The head was modified to operate as a four pass condenser.
  - Tubes continued to block with soft precipitate which was easily removed with water jets.
  - Examination of the cooling water pumping system revealed that the water velocity in the tube bundle was too low for transporting solids. The cooling water system was redesigned, which called for an 800 gal/min water pump to transport solid precipitates compared to the 180 gal/min pump originally installed. A new fibre glass cooling tower was purchased to handle the increased cooling water flow and the head of the condenser tube bundle was changed again for a two pass cooling water system.

No further problems were encountered with the condenser cooling system.

2.5.2.5 Gas Scrubbers

The gas scrubbers (4) were vertical towers 4 ft diameter and 24 ft high packed with Raschig rings, individual plate heat exchangers, piped for 110 gal/min recirculating water flow and fitted with top and bottom automatic water level control. One gal/min make-up water was added to the last scrubber, which progressed automatically through to the third scrubber where it was bled off to the 40,000 gallon recycle leach liquor tank. The fourth scrubber used the condensate ex the evaporators for scrubbing the vapours. Each scrubber was fitted with it’s own pump and plate heat exchanger and all four were serviced by one cooling water tower. Action taken in this area to overcome deficiencies were:

- All scrubbers were originally manufactured from mild steel. The temperature and ammonia strength were higher in Scrubber 4 and with the high solution flows corrosion was high. The scrubber was replaced with a 316SS unit.
- The 316SS plates in the heat exchanger of Scrubber 4 suffered corrosion pin holes due most likely to chloride attack. The plates were replaced with titanium plates.
- Scrubbers 3 & 4 would flood at higher than normal temperatures, which resulted in an operational delay to re-establish normal control. This was a result of insufficient head on the scrubber pumps to meet their NPSH requirements at these operating temperatures. This problem was never rectified due to the down time required to lift the four scrubbers and to re-engineer the full system.

2.5.2.6 Tank Farm

All major process tanks were fabricated from mild steel and were each of 40,000 gallon storage capacity. Two 8,500 gallon mild steel tanks for preheated pregnant liquor storage and condensate ex the evaporators were also installed. The Barrens Tank developed corrosion holes, (generally in welded areas) and was cement lined within two years for protection. The Barrens pipe lines also corroded in weld areas, but in general were protected by a calcium/magnesium/copper oxide scale. 316SS was adopted for new lines and maintenance repairs. The condensate ex evaporator storage tank suffered heavy corrosion and was replaced with a 316SS fabricated tank. The leach liquor storage tank suffered minor corrosion along weld lines and was finally cement lined after six years of operation. This is the only mild steel that experienced corrosion with the “lower” ammoniacal (60 to 100 g/l NH3) liquors and is thought to have been a result of lower copper levels during Stage 1 operation. The preheated pregnant liquor storage tank was retained as an 8,000 gallon pregnant liquor volume measuring tank for each evaporator charge for metallurgical accounting.

Tanks in the tank farm were all vented to the gas scrubbers through insulated mild steel ducts. All ducts blocked with ammonium carbonate and were successfully replaced with 316SS uninsulated ducts fitted with hot barren liquor flushing lines. It became a routine operational function to flush the ducts to maintain clear vents.

2.5.2.7 Cupric Oxide Thickening, Filtration and Drying

The 20 ft diameter Dorr Oliver thickener was fabricated from mild steel and operated for 13 years without major incident. Corrosion leaks developed in the side walls at the solid/liquid interface, which were repaired with rolled steel plate covering the area. A stainless steel screen box was fitted prior to the thickener feed well to remove pieces of evaporator scale which would periodically block the inlet to the underflow pump. After 10 years operation the thickener rakes were replaced with 316SS after inspection revealed major corrosion of the original mild steel rakes. The original thickener mechanical rake lifting mechanism was replaced with hydraulic jacks.

The initially installed filter, a 3 ft diameter by 1 ft face Eimco vacuum drum, was operationally unacceptable. The cupric oxide was dense and with a high SG would settle rapidly in the filter bath. The rakes could not maintain the oxide in suspension to be filtered and the solids load would finally "bog" the rakes. The unit was replaced with a 6 ft x 4 ft x 2 ft vacuum pan filter. The pan filter overcame the original bottle neck caused by the drum filter, but it was manually operated and labour intensive. As production increased this filter also became a bottle neck. At this point during Stage 1 Samin engineers designed their own purpose built continuous filter which proved successful. The filter was a 6ft diameter x 1ft long vacuum drum with internal slurry feed and internal circumferential vacuum pads. Approximately 4 ft diameter of one side of the drum was open for slurry feed and solid discharge. Excess slurry feed would overflow return to the thickener.

The original cupric oxide was sun dried and packaged for sale to the ER&S Smelter at Port Kembla, NSW. The grade of this product would not have exceeded 70%Cu due to moisture and carbonate content. The value of the product as a chemical was soon recognised and the installation of a dryer became necessary to improve quality. A 4ft diameter by 21 ft long rotary dryer with an externally fired single diesel burner located at the feed end, was installed and operating in 1972 with the oxide packaged at the hot discharge end in steel drums. The product specification was lifted to 77% Cu. The dryer capacity was a nominal 2.0 tonnes per hour. Further improvements to the dryer during Stage 2 were:

- With the arrival of natural gas in the area the dryer was converted to gas with three sidewall burners.
- The original mild steel dryer shell was replaced with heat resistant stainless steel.

2.5.2.8 Purchased Materials

To supplement copper oxide production during Stage 1 copper materials in the form of scrap copper and copper cement were purchased and leached. Scrap copper was purchased from scrap merchants in Adelaide and cement purchased from Israel and local tributers who produced cement by acid leaching of ore from numerous copper outcrops in the area followed by cementation on scrap iron. The following equations show the chemical reactions for leaching.

$$2Cu + (NH_4)_2CO_3 + 2NH_4OH + 1/2O_2 = Cu_2(NH_3)_4CO_3 + 3H_2O$$
$$Cu_2(NH_3)_4CO_3 + (NH_4)_2CO_3 + 2NH_4OH + 1/2O_2 = 2Cu(NH_3)_4CO_3 + 3H_2O$$

The equations show that to leach metallic copper oxygen is required. The copper scrap and cement were leached in an agitated tank fitted with air injection. When leaching copper cement the final solution was filtered through a vacuum drum filter coated with a precoat of diatomaceous earth. The tank was charged with a weighed amount of scrap or cement followed by additions of leach liquor. The agitator and air blower were started and the leach progressed until cuprous copper could no longer be detected in the pregnant solution. The off-gases from the leach vessel were drawn through a bubble cap tower fitted with three plates to recover ammonia before passing to atmosphere.

2.5.2.9 Services

The following services supported the operation.

- Steam was provided by two Cleaver Brooks, 200HP, package boilers. The boilers were fired with Bunker C fuel oil until Stage 2 when the burners were converted to natural gas.
- Boiler feed water was supplied from four water softeners fed with water from the River Murray Water tank. The feed water was preheated with steam and tannin dosed for oxygen control. The feed water had a TDS level of 450 ppm and boilers were bled to maintain a TDS level of 2,500 ppm.
- General plant water was supplied from the Bon Accord Shaft sunk during the operations in the 19th century. This supply changed to the mine when dewatering of the open pit commenced.
- Power was supplied from an up-graded ETSA grid.
- Natural gas was available late in Stage 1 operations when the installation of a spur line from the main line feeding Adelaide was complete.
- Anhydrous ammonia was purchased from ICI Australia and road freighted from their plant in Newcastle, NSW.

3.0 Stage 1 Process Statistics

The operational statistics for Stage 1 to October 1973 are:

Process Manning (excl. Mining)		42
Waste Mined	m^3	318,187
Ore Mined	wmt	336,257
Ore Crushed	dmt	360,068
Crushed Grade	%Cu	1.15
Ore Leached	dmt	196,239
Leached Grade	%Cu	1.35
Slimes Ore Rejected	dmt	163,829
Slimes Grade	%Cu	0.90
Process Cu Recovery	%	60.0
Overall Cu Recovery	%	38.3 (Budget 42%)

Production Cupric Oxide

Ex. Ore	dmt	2,060
Ex. Copper Scrap	dmt	131
Ex. Copper Cement	dmt	305
TOTAL	dmt	2,496
CuO Grade	%Cu	77.0

Utilities

Oil	GJ/t CuO	53.6
NH_3	kg/t CuO	116.0
Power	kwh/t CuO	1554

4.0 Product Quality

The cupric oxide product quality specification in 1972 was:

		Min. %	*Max. %*
Cupric Oxide	CuO	96.0	
Total Copper	Cu	77.0	
Free Copper	Cu		0.01
Cuprous Oxide	Cu_2O		0.85
Iron	Fe		0.01
Lead	Pb		0.01
Arsenic	As		0.02
Cadmium	Cd		0.02
Zinc	Zn		0.02
Calcium Oxide	CaO		0.20
Magnesium Oxide	MgO		0.35
Chlorine	Cl		0.01
Sulphur as sulphate	SO_4		0.12
Carbonate	CO_3		1.00

Typical Sizing BSS	*% Wt. Retained*
100	2.9
150	5.9
200	8.6
300	10.8
350	6.9
-350	64.9

5.0 Capital Cost

The actual capital cost for Stage 1 operation was as follows at the currency value at the time.

Item	*Year of Expenditure*		
Initial Plant	1969/70 A$	1970/71 A$	1971/72 A$
General	157,476	223,038	9
Siteworks		78,716	
Services		100,328	
O/C Development		43,686	
Crushing		273,696	
Ore Preparation		197,972	16
Process Plant		618,783	380
Tailings Disposal		33,786	
Product Handling		82,175	350
Workshop & Store		65,841	264
Mine Office		125,484	1,003
Change Room		25,232	
Transport		943	4
Commissioning		32,824	
Housing General		831	5,622
Vehicles		10,561	
Housing		33,218	
Sub-Total	157,476	1,947,114	7,648
Modification & Additions			
Dryer			14,757
Dust Extractor			1,258
Alterations to Scrubber			2,525
Magnetic Separator			4,564
Pipes for Fines Dam			1,824
General			4,473
Sub-Total			29,401
Grand Total	A$2,141,639		

6.0 Marketing

The original economics for the business was based on the sale of the product as a concentrate to copper smelters. The added value from the sale of the cupric oxide as a chemical was recognised at the early stages of the Stage 1 operation and additional capital was committed for the installation of a dryer to lift product quality. Only one shipment was made to the ER&S Smelter at Port Kembla and all other sales were directed at the chemical markets. The principal markets during Stage 1 operations were trace elements for the fertiliser industry and as raw material for plant fungicide manufacture. To strengthen the latter Poseidon Ltd formed a Joint Venture with Vermont Chemical and Seed Pty Ltd in 1972 to produce copper oxychloride for the SE Asian and Australian markets. In the same year a technical and economic feasibility study was undertaken with I.C.I. (Australia) Ltd to expand copper chemical production to copper sulphate, cuprous oxide and copper powder.

STAGE 2

Synopsis

In parallel with Stage 1 operation the laboratory and pilot plant development of a process to treat the whole of the run-of-mine ore continued. The flotation option was revisited along with reduction roasting of the ore. Encouraging reduction roasting results had been show in laboratory tests, which were confirmed with pilot testwork in Europe and America at the engineering company laboratories of Lurgi and Envirotech, respectively. Natural gas would be available in the area at the time of a roaster installation and there was great interest in the cleaner fuel.

1.0 Introduction

The Stage 1 operation (discussed earlier) was essentially a pilot plant vat ammoniacal percolation leach of deslimed ore to gain greater appreciation of the problems and chemistry. Laboratory test work for Stage 2 was undertaken during this 2 year period to define the additional requirements for Stage 2. Stage 2 was the whole of mined ore treatment plant, which was not possible using the Stage 1 plant due to:

- The presence of chrysocolla, which was insoluble in ammoniacal liquors, in the mined ores.
- The presence of clay in the gangue minerals, which inhibited solid/liquid separation of leached slurries.

1.1 Stage 2 Operations

Plate 2 shows the operations nearing the end of Stage 2 construction, identifies the layout of the plant and northern end of the mine in the background. The identified items in the photograph are:

(1) Crushing station.
(2) Crushed ore storage.
(3) Roaster.
(4) Roasted and ground ore storage bins (Old Stage 1 leach tanks).
(5) Agitation leach tanks behind (Behind Stage 1 leach tanks).
(6) Filter station (Behind Stage 1 Leach Tanks).
(7) Evaporator section with the ammonia recovery and tank farm to the left.
(8) Cupric oxide storage, handling and packaging.
(9) Workshop and Store
(10) Administration office and Laboratory

Stage 2 operations were designed and developed by Samin engineers using metallurgical test data reported by:

- AMDEL's laboratory reduction roasting and leaching testwork.
- Lurgi roasting test results. A 30 ton crushed ore sample was despatched to Lurgichemie Research Centre in Frankfurt, West Germany, for roasting trials in a 0.5m diameter by 9m long rotary kiln using Bunker C fuel oil as the reductant.

PLATE 2 Burra Processing Plant nearing the end of Stage 2 construction, Oct 1973. PIRSA Photo 048500

- Envirotech roasting test results. A 20 ton sample of crushed ore was despatched to Brisbane, California laboratory of Bartlett-Snow-Pacific Division of Envirotech Systems Inc. The roaster was a 6 hearth with a total effective hearth area of 25 ft^2 and the reductant was natural gas.
- The knowledge gained during two years ammoniacal percolation leaching by Samin during Stage 1 operations.

Santos' willingness to enter into a 10 year contract to supply natural gas for the project was a major reason for the roaster supplier selection. It also proved to be of economic benefit as the world fuel oil crisis and price escalation followed the signing of the contract.

There were two major deviations from the original design process flow sheet prior to construction. These changes were:

- Two proposed 7 plate stills for distillation of pregnant liquors were abandoned in favour of the distillation vessels used during Stage 1 operation.
- Four 14ft diameter by 30 ft long vacuum drum filters for solid/liquid separation and washing of the leach slurry were abandoned in favour of 5 vertically mounted cylindrical pressure filters.

Again the technology was developed with no similar operation in the world (except perhaps the Nicaro ammonia leaching nickel plant in Cuba, access to which was not available at the time). This report discusses and describes the operation and the numerous problems encountered. Stage 2 commenced the commissioning phase in October, 1973, after a 14 day plant closure to “cut-in” new equipment above that used in Stage 1. The commissioning phase extended for almost 3 years before achieving economic viability. Stage 2 ore roasting ceased due to exhaustion of ore reserves on February 25, 1983. A 1976 flowsheet and equipment description is presented in the Appendix.

2.0 Mining & Mineralogy

2.1 Mining Contractor and Equipment

The mineralogy of the Burra reserves has been discussed in the report on the Stage 1 operations.

A 10 year mining contract with Roche Bros. was established at the start of the Stage 1 operations. The contract expired on 17th August, 1980, but was rewritten and extended by a further 6 months to cover the accelerated mining program for final ore extraction. The equipment used during the mining period has been stated in the Stage 1 operational report. During the latter stages of mining on narrow benches the Marion 111M face shovel was restricted in it’s movements and was replaced with a Kato backhoe excavator.

2.2 Mining Reserves

During 1971 the mining reserves were up-dated along with the pit design. In 1970 a re-evaluation of the reserves was undertaken by Samin staff and an independent geologist

Dr. E K Sturnfels, Consulting Geologists of Victoria, acting for Davis Contractors Pty Ltd of Sydney, who were open-pit engineering consultants to Samin Ltd. The diluted ore reserve was reported as 4.363 million long tons @ 1.53% Cu. In 1971 Samin staff revised the orebody model and arrived at a "more conservative" diluted ore reserve of 3.50 million long tons @ 1.52% Cu. The total ore mined during the Stage 1 & 2 operations was 1,895,104 metric tonnes @ 1.71% Cu. Therefore, mined ore data compared to reserve estimates were:

	%
Diluted ore tonnage	53.2
Ore Grade	112.5
Copper metal tonnage	60.1

The reason for the shortfall in the ore and metal recovery was that the expected extrapolated mineralisation did not extend into the area beneath the mine pool, an area not accessible for exploration rotary drilling. Also for economic reasons and the loss of ore beneath the old mine pool, the bottom two benches were deleted from the mining program during a pit design review in the mid 70's. This proved to be a fortunate decision as ore below the 430mRL also pinched out rapidly to a narrow sulphide mineralisation, which was unsuitable for the Stage 2 process. The oxide ore predicted in the 430mRL to 410mRL was not there.

2.3 Rock Mechanics

The original pit wall design angles were 47^o for the dolomitic western wall and 37^o for the eastern wall siltstones. Initial mining wall angles were 70^o on the western wall and 60^o in the silstone areas. Mine rock mechanic evaluations in the mid 70's supported the western wall angle of 70^o, but the eastern wall siltstone angle was reduced to 45^o. Kingston Fault walls were reduced to 60^o at the southern end of the mine and Graves Shaft areas. The manual pit wall angle reviews resulted in Kingston Fault being located in the middle of the western pit haul road leaving unsatisfactory road stability below the 480mRL level. The haul road was diverted across the pit at the 480mRL to the eastern side and the western haul road and ore was mined to competent rock and base of the pit, respectively. The western haul road was then backfilled with waste to gain access and mine all the remaining ore on the eastern side of the mine.

The mine benches were regularly inspected for cracking and any significant cracks were infra red theodolite monitored for movement. Two electronically activated alarm monitors were also established across cracks. In the last three months of mining an increased rate of wall movement was detected in the SE corner of the mine. Mining operations ceased, a trench was excavated along the surface crack and nearby pit dewatering water was pumped into the trench at a rate of 700 gallons per minute. The wall collapsed within 8 hours and mining activities resumed within 48 hrs.

2.4 Mine Dewatering

Mine dewatering appeared to be a straight forward exercise as we knew the major mine dewatering in the 19th century was achieved at Morphett Shaft at a pumping rate of 2,500 gallons per minute. The shaft was blocked at water level with fallen timber due likely to a fire pre the Samin mining period. Samin cleared about 10 ft of the blockage and attempted unsuccessfully to drill to ascertain the depth of the blockage material before abandoning Morphett Shaft as a site for dewatering pump installation.

Two attempts were made to sink a 24 inch diameter bore hole (one on the 512mRL bench level and one alongside Morphett Shaft) to install a 200kw 1000 gallon per minute submersible multi stage water pump. Both were unsuccessful, but a 200 gal/min pump was installed at the 512mRL bore hole and operated until Graves Shaft pumps were operational. Water inflow to the 512mRL borehole was below the installed pump capacity, which caused several pump and motor failures.

Examination of the 19th century mine records indicated mine access drives entered Morphett Shaft on the western side at the 40, 50 (approximately 430mRL) and 60 fathom levels. Also a drive entered Graves Shaft on the northern side at 50 fathoms. Two bore holes were installed at Graves Shaft, which intersected the identified drive. Two 200 gallon per minute multiple stage pumps were installed in each hole and pumping commenced with Graves shaft water pumped to the process plant. The next four pumps were all installed at Morphett Shaft with the bores intersecting each identified drive to the 60 fathom level. The original water table level, prior to pumping, was 491mRL. Over time the following bore pumps were installed at these locations.

	Design Pump Rate Gal/min			*Pumping Depth*
	200	*300*	*500*	*Approx. mRL*
Morphett Shaft	1	2	1	410
Graves Shaft	2			430

There was a steep cone of draw down between the southern siltstone ore and the limonitic waste beneath the "old mine pool". Water seeped through the northern faces of the southern pit. Four 200 gal/min pumps were installed across the mine in this area to create a cone of draw down to prevent this water seepage. These pumps were drawing water from approximately the 390 mRL level. About this time the Graves Shaft pumps ran dry and a 500 gal/min pump and borehole was installed on the then floor of the pit at the 440mRL bench level.

Final installed operating pumping capacity was 2600 gal/min from 9 boreholes and ore box cutting reached the 430mRL level, or the top of the 50 fathom old drive network. This drive network had ground water in it at the completion of mining. Pumps were progressively stopped and the last of the ore was box cut from the 480mRL bench. Over time the ground water level returned to the 485mRL or six meters below the original water table. In total 14 bores were drilled 12 of which were operational and 2 were abandoned prior to casing.

All mining activities ceased in February, 1981, with two years ore supply stockpiled at the crusher.

There was a wide cone of ground water draw down around the mine. First to lose water were two wells at a private residence in Burra North (about one km from the mine). The original depth of water in each well was approximately 1.0 meters. Samin paid for the connection of the residence to the town water supply. The water level dropped below Bon Accord shaft pump intake (approximately 10m fall in the water level) within three years of the end of mining. The Company owned and operated pump was shut down. Bon Accord Shaft was approximately 800m from the mine. About 4 months prior to ceasing mining and ground water pumping, a remote bore at a farm house about 5 km from the mine stopped producing. It was the sole water supply for the farm house and Samin immediately drilled a new deeper bore to reinstall supply. Four months after ceasing mine dewatering the old bore was active again, which confirmed it had been affected by mine water pumping.

3.0 Crushing

A description of the crushing operation and the changes made to the original design have been discussed in the Stage 1 report. A schematic of the section is shown in the Appendix. The major changes to the installed equipment during Stage 2 were:

- Engineering input to assist the manual operation to remove wood. The removal of wood prior to crushing had a chemical requirement as well as the obvious physical need. Any wood entering the roaster was reduced to carbon, which would absorb soluble copper in leaching which reported to the tailings solids.
- The primary and secondary screen decks were changed to 1/2 inch rubber to overcome screen blinding problems. The lower screen on the primary was removed due to frequent blinding and difficulty to clean.

4.0 Reduction Roasting

The roaster selected was the Envirotech 25 ft diameter, 12 hearth of Herreschoff design with air cooled rabble arms and central shaft. The refractories were designed for temperatures of 1000^{o}C and the unit was natural gas fired. The top four hearths were the drying/ore preheating hearths operating under oxidising conditions. The lower eight hearths operated with reducing gaseous conditions. Calcine discharging from the roaster was cooled in an inert atmosphere in a rotary cooler fitted with cooling tubes supplied with water from the steam water accumulator, prior to lifting to the roasted ore storage bin. Off-gases were dedusted in a cyclone bank and cooled in a steam raising economiser. Cooled gases passed through a baghouse and induced draft fan followed by discharge to the stack and atmosphere. The roaster had a total design steam raising capability of 27,000 lbs/hr, which exceeded the total requirement for the distillation and filtration sections.

Ore was delivered to the roaster through one of four variable speed extractor belts located in the concrete tunnel beneath the crushed ore stockpile. The ore before entering the roaster was relieved of accretions and foreign material by passing it over a 4ft x 2 ft vibrating screen fitted with a one inch deck. The ore passed over a belt weightometer located on the inclined belt to the roaster and entered the roaster through double dump valves. The principal roaster chemical reactions were:

$$CuCO_3.Cu(OH)_2 + CO + H_2 = 2Cu + 2CO_2 + 2H_2O$$
$$(2CuCO_3.Cu(OH)_2) + CO + 2H_2 = 3Cu + 3CO_2 + 3H_2O$$
$$CuSiO_3.2H_2O + H_2 = Cu + SiO_2 + 3H_2O$$
$$Cu_2O + CO = 2Cu + CO_2$$

There was no equivalent roasting operation to assist with the development of our personnel and operational problems that arose. The following discussions highlight the problems experienced and actions taken to optimise this unit operation.

4.1 Roaster Burners

The roaster was originally designed to treat ore at a rate of 45 dry tonnes/hr at a free moisture level of 4 percent. Burners were originally installed on Hearths 1, 2, 3, 4, 5, 8, 10 and 12. Two 20 million Btu/hr Urghurdt toroidal gas reformers adjusted to 0.4 gas stoichometry were installed on Hearth 12 to provide the reducing gases. Carbon monoxide and hydrogen levels in the 10 to 12 percent range were normal at Hearth 12. Two North American 2.6 million Btu/hr burners were installed on each of Hearths 5, 8 and 10 and were adjusted for 0.8 gas stoichometry. Three North American 4.8 million Btu/hr burners adjusted to 1.2 gas stoichometry were installed on each of Hearths 1, 2, 3 and 4. A proportion of the shaft cooling air exiting at the top of the roaster was returned to Hearths 2, 3 and 4 to burn excess reducing gases and recover full heating values.

Changes were required for the burner configuration largely as a result of a free moisture increase in the ore feed, which rose to an average of 11 percent. These changes were:

- The burners on Hearth 1 were removed as they were effectively only increasing the off-gas temperature and raising steam production levels above plant requirements. Solid/gas heat transfer and chemical reactions in hearth roasters takes place largely at the drop holes between hearths.
- Two North American 4.8 million Btu/hr burners adjusted for 0.8 gas stoichometry were fitted to Hearth 6. This became necessary as a result of the cooling effect on the roaster of the cyclone dust returning to Hearth 7.
- One gas reformer was removed from Hearth 12 as the reducing gas requirement was lower than design.

The overall affect of the higher ore feed moisture was a reduction in the roaster feed rate to 35 dtph at free moisture levels of 11 percent.

The refractory burner blocks supplied by North American did not penetrate fully the refractory wall of the roaster. A round burner block changed to a square section in the wall and dust laden roaster gases eddied in this area resulting in excessive burner slag requiring monthly shutdowns for cleaning. Full roaster refractory shell length burner blocks were cast at site and all burner blocks were replaced. Burner slag generation was markedly reduced.

4.2 Roaster Temperatures

Roaster temperatures had to be juggled to satisfy the following operational criteria.

- Temperatures needed to be as high as possible to satisfy the reduction rate reactions.
- Temperatures needed to be high enough to destroy the clay laminates for improved filtration during solid/liquid separation.

- Temperatures needed to be maintained as low as possible to avoid calcination of dolomites to minimise leach liquor carbon dioxide losses.

With the high moistures Hearths 2, 3 and 4 were normally controlled in the range 680^{o}C for drying and preheating the ore. Temperatures were decreased in Hearths 5 and 6 to 640^{o}C and Hearths 8 and 10 were controlled at 620^{o}C. The low stoichometry rectiformer flame normally resulted in lower temperatures again in Hearths 11 and 12. These seemed to be the best temperature conditions to satisfy the above process criteria.

All burners (except Hearth 12) were fitted with automatic temperature control and all burners had flame failure relays fitted.

The original roaster design ore residence time of 30 minutes was increased to 1 hour by decreasing the shaft rpm.

4.3 Roaster Internals

Roaster refractories were designed to operate at 1,000^{o}C.

- Wall insulation was 9 inch thick castable.
- The wall heat resistant refractory was 41/2 inch fire brick.
- The hearths were suspended arch off a strengthened shell ring using 9 inch fire brick.
- The shaft protective refractories were dense high temperature castable.

The plant was shut down for two weeks annually for maintenance. Furnace refractory repairs were minor and restricted generally to burner ports and spot repairs on the shaft. During the life of the plant the wall refractories were replaced in Hearth 1 as a result of damp ore drag on the wall fire brick and early burner damage to the insulation refractories.

The central bolted air cooled shaft was cast from cast iron with four rabble arm outlets per hearth. It was supported at the base of the roaster by on oil bath submerged thrust bearing and a grease lubricated bearing at the top. Sand seals were fitted to the shaft top and bottom to prevent furnace gases escaping. There were no operational or significant maintenance problems with the shaft during the 10 years operating life. The individual shaft castings per hearth were however not correctly installed, which resulted in ore from an upper hearth falling onto lower hearth rabble arms (discussed later). The bull ring and pinion drive gears were replaced once. The shear pin shaft protection at the drive coupling would periodically fail due to pin fatigue and if not replaced quickly often resulted in a bogged roaster. The shear pin protection was finally replaced with a torque limiter, which eliminated premature failures.

Rabble arms, weighing approximately 500 kg, were supplied cast from high nickel/chrome HH heat resistant stainless steel. The arms were socket located in the shaft at two machined faces and held in position with a locking pin. The machine faces formed seals for the inner and outer shafts. Shaft cooling air was blown into the central shaft and would pass through the central mild steel tube in the rabble arm to the outer end and return back along the rabble arm to the outer shaft. Air in the outer shaft could discharge to atmosphere or be returned back as combustion air to Hearth 1 to 3. Several rabble arm failures were experienced during the life of the roaster with the causes being:

- The initial failure in 1977 was at Hearth 12 and was a result of the incorrect installation of the shaft. Ore from Hearth 11 fell directly onto the rabble arms in Hearth 12 and impact wear reduced the casting thickness in it's high stressed area. On inspection the remaining arms showed similar wear, but remained operational for the roaster life after rock boxes were welded on the rabble arms at the wear areas to prevent further loss of metal.
- A further four single rabble arm failures occurred on the upper hearths the cause of which were initially more difficult to define. No rabble arm failure resulted from mechanical overload. The cause of the failure was finally attributed to sigma phase embrittlement of the HH alloy, which caused cracks in the casting after 5 years operation. At annual shutdowns all rabble arms were die checked for cracks near the shaft and when detected the arms were removed for heat treatment at 1,200^oC for 12 hours followed by grinding the crack out and weld repairing with 10 gauge 253 MA Sandvic sigma phase resistant welding rods. Rabble arms on hearths with burners fitted were more susceptible to sigma phase embrittlement. Four replacement rabble arms were cast from the higher grade HK heat resistant alloy to supplement the spares holding. The problem became manageable with the annual plant maintenance shutdown philosophy.
- The original rabble arms were supplied with square plugs welded in the square casting holes. Sigma phase cracking emanated from the square casting corners on several rabble arms. These arms were removed and the square casting holes and plugs machined out to round holes. Round plugs were welded into the holes. The rabble arm pattern was altered to include round casting holes on all new arms.

Tynes (rabble teeth), which were cast from the same HH alloy, weighed approximately 20 kg and were attached to the rabble arms by a dove tail connection. Like rabble arms they could be changed at temperature, but annual maintenance down time enabled tyne upgrades while the roaster was cold. The normal tyne blade depth was 8 inches and tynes with 4 inches or less at annual shutdowns were replaced. Wear understandably was greatest at the outer circumference and where the ore was less fluid in the moist upper hearths. Total tyne life probably averaged 2 to 3 years. To reduce the cost of the tynes cast from HH alloy some trials with white cast iron were tested in upper hearths, but these warped and twisted at the dove tail plate making removal difficult.

4.4 Roasted Ore Cooling and Storage

The object of the cooler was to cool the ore from 600^oC to less than 180^oC in an inert atmosphere to prevent the formation of copper ferrites, which were insoluble in the ammoniacal leach solutions. The inert atmosphere was developed with the introduction of water at a rate of one litre per minute to the feed end of the 8ft diameter, 75 ft long rotary cooler where it immediately generated steam when it contacted the ore at 600^oC. The majority of the cooling was effected by the ore cascading over 41/2 inch steam tubes stacked three deep for 2/3rd of the cooler length in the rotary cooler. Water for cooling was drawn from the base, and returned to, the steam accumulator. Ore discharged from the cooler at 120^oC and was lifted with a bucket elevator to the roasted ore storage bin.

Ore originally exited the roaster through a fixed grizzly located in the drophole at the base of the roaster. The grizzly blocked frequently (once or twice per shift) with accretions and burner slag, which required roaster shutdowns and opening of Hearth 12 doors for cleaning. The reducing atmosphere was lost and copper ferrites formed. Samin designed a vibrating grizzly with on-the-run dropping of the plus 2 inch material to a quench pit. This oversize material was returned to the crusher.

The original cooler atmospheric end seals were fitted on the 8 ft diameter. The cooler shell was not perfectly round and the shell would rise and fall on the trunnion rollers depending on the temperature, which resulted in the seal opening and closing each revolution. The cooler internals had to be maintained under positive pressure to prevent inward oxidising air leaks with the result dust emissions from the seal areas were excessive. The discharge end was addressed first by closing the 8 ft diameter cooler end and fitting a central round 24 inch diameter discharge tube and a cooler internal conical lifter. A floating dust seal on the smaller diameter was more effective. The discharging ore fell into a screw conveyer and bucket elevator as it passed to the roasted ore storage bin. The cooler feed end was closed and an additional 18 inch stainless steel screw conveyer located on the cooler central axis was fitted. Spring loaded seals between the 18 inch diameter screw conveyer and the cooler feed end plate were more effective. The original roaster discharge screw conveyer delivered ore to this new screw conveyer.

As the roaster throughput rate increased the cooling tube feed water rate decreased. On two occasions the feed water pump disintegrated. On investigation the pressure drop across the cooling water tubes caused the failures. Water entered and left the cooler through a 4 inch carbon sealed rotary Johnson Joint. The carbon seal in the joint failed on one occasion. Once inside the cooler, water and steam travelled in series through the entire length of pipe, which created high back pressures causing steam to flow backwards through the pump. The cooler tube bundle was split into four parallel units, which reduced the overall pressure drop and overcame the cooler feed water problems.

Apart from these major modifications to the supplied capital equipment the cooler was mechanically reliable and required only minor routine maintenance. Near the end of its operating life some leaks developed in the tube bundles as a result of ore wear. Several of the outer tubes were replaced. Screw conveyer flights required regular service.

Water as steam forming the inert atmosphere in the cooler was never a totally satisfactory option. It caused secondary problems with condensation at the end of the cooler and in the bucket elevator. It was reduced to low levels, but never totally removed. The most effective inert atmosphere was supplied through the roaster discharge screw conveyer, which “pumped” a small quantity of reducing gases from Hearth 12 into the cooler. This was a concern due to the creation of an explosive condition in the cooler and storage bin. Occasional minor explosions were noted in the cooler and ore bin when continuous operation was interrupted so the doors on top of the bin were left open at all times to allow the expanded gases to escape. We could not envisage a mechanical solution to prevent roaster gases entering the cooler, but prepared strict safety operating procedures to be followed during interruptions to normal operations.

The single chain bucket elevator fitted with cast iron buckets could not achieve 35 tph let alone the tender specification of 50 tph. The unit was replaced with a twin chain elevator fitted with continuous fabricated buckets. The replacement elevator was operationally reliable, but required major maintenance on the rollers, chains and buckets during routine and annual shutdowns.

No.1 Leach Tank from Stage 1 was used for storage of roasted ore prior to grinding.

4.5 Roaster Off-Gases

The roaster off-gases left the roaster at 600^{o}C and passed through 4 Buell Cyclones installed in parallel. The dust collected was returned to Hearth 7 of the roaster through counterweighted double dump valves. Gases leaving the cyclones passed to a steam raising economiser supplied by Senior Economiser where the temperature was reduced to 190^{o}C. The gases then passed through temperature controlled water sprays to reduce the temperature further to 160^{o}C prior to entry into a Buell-Norblo Norfelt dust collector fitted with Nomax bags. The baghouse and roaster were drafted with a 150HP centrifugal fan and the cleaned gases discharged to atmosphere via the stack. A summary of the equipment performance in this area follows.

The performance of the cyclones was satisfactory requiring only minor internal refractory repairs over the mine life. During plant shutdowns the dust underflow hopper would sweat and restrict the solids return to the roaster on start ups. It became an operator function to monitor the double dump valve operation until the hopper and ducts reheated.

The steam raising economiser was the finned tube variety. No mechanical problems arose directly attributed to this unit. Spalling of refractory on the walls of the duct directly above the economiser blocked sections of the fins. Larger pieces of refractory were removed by hand and the fins and tubes cleaned by water washing.

The temperature controlled water sprays directly beneath the economiser were a failure. The droplet size would not vaporise prior to the “drop out box” before the baghouse, which resulted in solid accretions in the ductwork. Throughput ore rates were initially slow which led to natural safe operating temperatures at the baghouse and commissioning of this unit continued.

The baghouse proved to be an engineering disaster with the following problems identified:

- Reverse air for bag cleaning was initially through poppet valves drawing in atmospheric air. This resulted in local cooling and condensation on a bag material, which was hydrolysed by steam.
- The suppliers attempted to correct the failure with the installation of a 200 cfm air compressor and a crude attempt to develop the Micropulse pulse bag cleaning system. This system also failed as the pulse air was cold and the pulses blew holes in the bags.
- The bags were supported on spiral spring cages hung from the base plate on the roof. With small pressure drops across the baghouse the bags and springs would suck upwards reducing the filter area and presenting bags closer to the pulse air. A complete set of replacement bag cages were fabricated and installed.

- Access to the bags was through the roof, the structure of which was light gauge and sealing poor. Rain water would leak through the seals. The roof was replaced and sealing improved.
- During the commissioning period the plant maintenance shutdown frequency was every 28 days for 32 hours. This time was necessary for baghouse bag replacement and repairs.

Nomax bag replacement costs were high and stack dust emission levels were not achieved. Finally trial high pressure water sprays were installed in the duct beneath the economiser and the baghouse bypassed. Surprisingly dust wetting was very effective in the moisture laden off-gases and required stack dust emission limits were immediately achieved. The baghouse no longer dictated the plant shutdown frequency, which was extended to 32 hours every 35 days. The dust slurry was discarded to the tails dam.

The induced draft (ID) fan maintained the controlled draft on the roaster. The automatic positioning of the discharge damper was controlled by the pressure in Hearth 1, which during operations was controlled slightly positive at +0.1 inches WG. This resulted in a pressure neutral condition at Hearth 12 and hence minimised inward air leakage into the roaster. With the change to wet rather than dry dust collection the wet gases caused impeller and damper corrosion and minor fan balance problems. A replacement impeller and damper were fabricated from stainless steel. On installation the problems disappeared. On the positive side the lower gas temperatures enabled a decrease in the installed fan motor size from 150 to 100 HP.

4.6 Roaster Statistics

The roaster operating statistics were:

Total Ore Roasted	wmt	1,724,760
Ore Moisture	%	11
Total Ore Roasted	dmt	1,535,036
Ore Grade	%Cu	1.85
Gas Usage *	GJ/t Ore	2.7 (2.4 to 2.9 dependent on ore moisture)

*Note: * Assumes all gas was consumed at the roaster. Actual roaster consumption would have been approximately 95% of the total.*

5.0 Grinding

The mill was a single compartment, grate discharge, air swept, bull and pinion gear drive, 9 ft diameter by 17 ft long Vickers Ruwold dry grinding mill fitted with Nihard shell liners. The drive was a 1000 HP, 11kV motor. The bearings were roller with water cooled recirculating oil lubrication. The sweep air was provided with a centrifugal fan through a Micropul Baghouse. The make-up ball charge was 21/2 inch forged steel. The mill was fed with ore from the roasted ore bin by a variable speed drive screw conveyer and discharged through a 1/4 inch trommel screen, for chat removal, to a single chain bucket elevator fitted with cast buckets identical to the unit that failed in the roaster circuit. Ground ore was stored in a modified Stage 1 No.2 Leach Tank. A schematic of the circuit is presented in the Appendix.

The only problem with the circuit was with the bucket elevator which would not originally convey the ground ore at the 50 tph rate. Drilling two 3/8 inch holes in the bottom of each bucket broke the suction restricting the ore discharge and design ore throughput rates were achieved. One set of mill shell liners were replaced during the life of the mine. The air sweep baghouse performance was excellent with only one replacement set of bags required.

The ball mill was significantly over designed and to prevent overgrinding operated for the entire time with a ball charge of 40%. Grinding media consumption was 0.12kg per tonne of ore roasted and the ground ore sizing was in the range of 60 percent minus 45 microns. Ground ore temperatures were approximately 100^{o}C.

6.0 Agitation Leaching

The original design called for three 15ft diameter by 18ft deep mechanically agitated leach tanks. The ore feed was metered from the ground ore bin with a variable speed drive screw conveyer and conveyed to the first of three inground agitated leach tanks. Leach solution was metered and rate controlled by the density probe installed in this tank. The leach slurry passed through overflow launders from the first to the second and finally third leach tank. Pumps were located in the third leach tank to deliver the slurry to filtration. During the early commissioning it was realised that ore degradation was occurring during the long leach time and agitation leaching was reduced to two vessels and 6 feet of the vessel depth was sealed with a cement base. Air was injected into all three tanks and the isolated third tank was used to leach scrap copper, copper cement and cupric oxide spillage. A flow schematic is shown in the Appendix.

The principal leach reactions are:

$$2Cu + (NH_4)_2CO_3 + 2NH_4OH + 1/2O_2 = Cu_2(NH_3)_4CO_3 + 3H_2O$$
$$Cu_2(NH_3)_4CO_3 + (NH_4)_2CO_3 + 2NH_4OH + 1/2O_2 = 2Cu(NH_3)_4CO_3 + 3H_2O$$

The pulp density used in leaching varied between 45 and 65 percent solids dependant on the head grade of the ore. At 45% solids the coarse ore chips accumulated on the floor of the leach tanks, but this build-up was floated forward when higher solid densities (lower ore head grades) were leached. This presented some problems for filtration, but were operator manageable. The original Stage 2 design called for a leach liquor of 100g/l NH_3 and a resultant pregnant liquor containing 60g/l Cu. For numerous reasons, associated with the leaching and filtration, the leach liquor target became 60g/l NH_3 and a pregnant liquor grade of 35g/l Cu. The leach reaction required oxygen which was supplied through air injection to the agitators. This required the maintenance of a draft on the leach surface to control ammonia losses and ammonia recovery from this draft air was effected by passing through a bubble cap tower supplied at the top with cold barren water. The optimum leach temperature to maintain acceptable copper leach rates and oxidation of the cuprous ion was 60^{o}C. Temperatures below 50^{o}C were avoided to minimise magnesium solubility which would report to the cupric oxide product. This temperature also kept ammonia vaporisation to acceptable levels. Operating at high solids density would drive the pulp temperature towards 70^{o}C and a simple recirculating design pulp cooler was introduced to the section along with cooling of leach liquor in a plate heat exchanger. Heat removal from the ore through the originally installed water cooled screw conveyers proved to be ineffective.

7.0 Filtration

This section took 3 years to commission and finalise appropriate equipment. Commissioning came within two months of forcing the closer of the operation. The original drum filtration operation was bypassed for the pressure filter Chemap option, which was finally abandoned and a modified vacuum drum filtering circuit was installed using available second hand equipment at the time. Site manning (excluding mining) during this period was 85 compared to a budget and final manning of 65.

7.1 Funda Filters

The 316 stainless steel Funda Filters manufactured by Chemap in Switzerland were installed by their Australian agents Alfa Laval. All process pipelines were manufactured from stainless steel. The filters (5) were approximately 5 meters high by 1.2 meters diameter and fitted with a nest of 70 by 1.0m diameter stainless steel filter trays located on a central shaft. Each filter tray was fitted with an 83 micron stainless steel filter cloth, which was retained at the circumference by a SS clip and at the central shaft by a SS plate spacing ring. The filters were operated and controlled to an automatic program. The basic filtration sequence was as follows.

1. The pulp flowing in the filter circuit ring main was stopped and the inlet and overflow valves on a filter opened.
2. The filter vessel would fill with pulp for a set time.
3. The filter overflow valve was then shut and the central shaft filtrate line opened. The slurry would filter for a set time.
4. The filtration cycle would then stop by opening the slurry recycle line, closing the filter feed and filtrate valves and opening the pulp drain line on the filter, which returned the excess pulp to the agitators. A second filter could then enter a filtration stage.
5. The filter would then draw wash water from the low ammonia wash tank and a similar program repeated to wash the retained solids on the trays. At the end of the cycle another filter could enter the wash stage.
6. Steam was then introduced to the filter and the cake steamed to recover ammonia. Vapours leaving the central shaft were cooled in a condenser for use as wash liquor.
7. At the completion of steaming, recycled plant water was pumped through the central shaft and under the filter cloth. At this time the central shaft would start to spin and a dump valve on the filter would open and the filter cake would slurry and discharge to the tailings pump pit. The filter was then ready for the next filtration cycle.
8. The tailings pump was controlled by pit level probes.

This was the basic program which was normally expanded to 14 steps in a 20 minute time frame. A summary of the principal operational and mechanical problems are as follows.

- All shafts were initially bent by the Chemap technicians as a result of uneven tightening of the filter nests. All nests were dismantled and shafts sent away for straightening.
- 316 SS is not a good wearing steel and several pipework holes were experienced.
- The original installed control valves were SS knife gate, designed to seal in one direction only. The filter sequencing required several valves to seal in both directions. Consequently copper and ammonia were lost through leaks.
- All gate valves were replaced with diaphragm valves along with the necessary pipework modifications and lost operational time.
- The nest of screen trays were tensioned using a torque wrench to avoid bending the shaft.
- The SS screens blinded with a calcium/magnesium scale, which could only be cleaned with an acid (HCl or HNO_3) wash followed immediately with a caustic wash. This lead to the installation of bulk acid and caustic washing equipment to wash screens insitu. This resulted in further complications of copper ion corrosion.
- Backwash water would lift the screens free of their retaining clips after minimal blinding.
- The centrifugal solids discharge caused wear at the tray circumference and holes in the screens.
- Holed screens resulted in solids contamination of the pregnant and leach liquor storage tanks. Stage 1 Leach Tanks 5 and 4 were introduced to settle solids prior to entry into these tanks, respectively.
- Nylon monofilament screen cloths were tried. They performed well at resisting scale blinding, but suffered high wear failures.
- The low upward pulp velocity in the filter lead to size segregation and reduced cake thickness from the base to the top of the filter. This resulted in a lowering of the washing and steaming efficiencies.
- The filter trays acted as distillation trays with the steam heat on the underneath side. Copper oxide scale formed on the trays.
- The filter nest was supported on a roller bearing in the top shell and the bottom journal ran in a Teflon sleeve. Shaft whipping grabbed the Teflon seal on numerous occasions during rotation. The Teflon was finally replaced with Rulon with some success.
- The maximum cycles on one nest was 1,800 or equivalent to 25 days. Only three nests exceeded 1,000 cycles and many nests failed with less than 100 cycles. A nest repair would take 24 hours for three tradesmen.
- During commissioning we learnt that the filters were the tallest ever made by Chemap and the first ever installed in a metallurgical operation. The majority of their installations were in breweries.
- All efforts to improve the filter performance were aimed at mechanical solutions in the hope the metallurgical performance would lift with greater mechanical

reliability. This did not occur and copper recoveries across the Funda Filters during commissioning were 50 percent.

7.2 Vacuum Drum Filtration

In 1975 it was agreed that it was unlikely the Funda Filter performance would ever be acceptable. An 8ft diameter batch pressure filter filtering 12 inch cakes was designed, fabricated, installed and tested with some encouraging results. However, again it was something new and we preferred to go with the well tested and accepted operation of vacuum drum filters. Time restrictions prevented the fabrication of new filters so the hunt was on for available units “on the ground”. A new 14ft diameter by 18 ft long Eimco vacuum drum filter was located in WA and two 12ft diameter by 14ft long Feinc string discharge vacuum filters were found at a disused operation in NT. The Feinc Filters were freighted to Melbourne for overhaul and the Eimco taken direct to Burra, where it was immediately installed for single stage filtration trials. Within eight hours of testing, the single stage filtration out-performed the Funda Filters and the latter were shut down. Design work for the vacuum drum filter circuit commenced immediately and the Funda Filter station was dismantled in readiness for the replacement filters. The single drum filter station operated until it was required to be repositioned to the secondary filter location in the new station.

The two Feinc Filters operated as primary filters. The cake thickness aim was for 4 to 6mm cakes with a two displacement wash using low copper and ammonia wash liquor. The cake discharge was repulped with low ammonia strength liquor to 60 percent solids and refiltered on the Eimco filter and washed with hot Barrens water. The solids from the Eimco were repulped with recycle water and pumped to the tailings dam. The vacuum vapours, process and hood vent air were scrubbed in one of three twelve tray stainless steel vacuum operated bubble caps operating in parallel.

The Feinc Filter internal pipework failed (holed) during early commissioning and the rotary vacuum valve for string discharge was unsuitable for air discharge. Samin replaced the internal pipework and fabricated a new rotary valve for air blow solids discharge and the plant became economically viable. Two months later Poseidon entered Receivership taking Samin with it. Samin was initially instructed to shut down based on the previous 5 years economics, but the decision was reversed when the last two months operating performance became more widely known.

During Receivership operation (approximately 2 years) Burra operated with a positive cash flow of $1.0 million per annum. For the last 4 years of operation Burra generated a positive cash flow, including administration, of $2.0 million per annum.

The principal changes made to the vacuum drum filter station to improve the performance were.

- The twelve tray vacuum bubble caps for gas cleaning were converted to 6 trays and the cap submersion reduced to reduce the overall pressure drop.
- The vacuum filter station was originally supplied with four 200HP Nash vacuum pumps. These were reduced to two (one operational and one spare) by the end of commissioning.

- Cold Barren water was applied to the top two trays of the bubble cap columns and recirculated cooling was applied to the lower sections.
- The bubble cap columns were converted to operate under atmospheric conditions.
- Air from the CO_2 absorption section was directed to the atmospheric bubble caps for ammonia recovery.

8.0 Distillation

There was little change to the evaporator section to that described in the report on Stage 1. Two additional evaporators were installed and the system was piped for two parallel lines of three in series. An additional condenser bundle was also installed along with the 5th gas scrubber. Evaporators were steamed two in line while the completed evaporator was vented to atmosphere and discharged to the thickener. Normal steaming rates for three in line evaporators was 10,000 lbs/hour, but was reduced to 6,000 lbs/hour for two evaporator in line steaming. An additional cooling tower was installed for scrubber and leach liquor cooling. A more detailed description of the operation and changes were presented in the Stage 1 report. A flow schematic of the area is presented in the Appendix. Typical operating conditions were.

Temperatures	^{o}C
Condenser Vapours In	100
Condenser Vapours Out	55
Condenser Water In	40
Scrubber PHE cooling water In	18
Leach Liquor PHE water In	35
Leach Liquor PHE water Out	38
Evaporator 1st in steam line	100 to 120^{o}C
Evaporator 2nd in steam line	120 to 135^{o}C
Evaporator 3rd in steam line	135 to 140^{o}C

The roaster supplied the total steam requirement for the evaporators under normal operating conditions, but could be automatically supplemented from the Stage 1 boilers during periods of shortfall. Roaster off-gases were drawn at a rate of 600 cfm to a scrubber irrigated with cooled leach liquor. Carbon dioxide was absorbed at a rate sufficient to replace that lost to the ore during leaching and the scrubber and bubble cap exhausts.

9.0 Product Drying and Packaging

Cupric oxide thickening, filtering and drying sections have been discussed in the Stage 1 report. The market demands for tighter product chemical and physical size control dictated changes in the leaching, filtering and distillation areas along with an expansion of the product handling area. Changes in the wet side were associated with temperature control and solids entrainment. In the packaging area the following occurred.

- Initially a water cooled screw conveyer and 15 tonne packaging bin was installed. This enabled packaging under more hygienic conditions and the use of plastic lined bags to replace the 50 kg drums. The 200 litre drum packaging option was always maintained.

- The market requirement for fine CuO increased along with a course rillable product. A Sturdivan air classifier was installed to provide both of these products along with two additional storage bins for each size fraction.
- The market demands for fine product increased further which could not be supplied through the classifier. A 15 inch diameter by 18 inch long vibrating ball mill and associated equipment was installed to supply the additional requirements.

The new equipment generally performed according to design and expectations except the discharge from the bucket elevator buckets, which recycled solids similar to that experienced with the initial roaster elevator. The problem was again overcome by drilling holes in the base of each bucket to break the suction. The chemical quality of the cupric oxide rose from 77 to 78.5% Cu during this period.

10.0 Stage 2 Process Statistics

The overall operational statistics during Stage 2 were:

Manning - Site (excl. Mining)		60
Manning - Mining (Contract)		12
Waste Mined	m^3	1,549,379
Ore Mined	wmt	1,607,598
Purchased Ore/Stage 1 Slimes	wmt	117,162
Ore Crushed	wmt	1,724,760
Ore Moisture	% H_2O	11
Roasted Ore	dmt	1,535,036
Roaster Head Grade	%Cu	1.85
Tails Grade	%Cu	0.39*
Ore Cu Recovery	%	79
Overall Cu Recovery	%	82

Production Cupric Oxide

Ex. Ore	dmt	28,394
Ex. Copper Scrap	dmt	3,246
Ex. Copper Cement	dmt	1,305
TOTAL	dmt	32,945
CuO Grade	%Cu	78

Utilities

Natural Gas	GJ/t CuO	115
	GJ/t Roasted	2.7
NH3	kg/t CuO	150
	kg/t Roasted	3.8
Power	kwh/t CuO	2600
Murray Water	m^3/t CuO	30
Grinding Media	kg/t Ore	0.12

Note * During operations with oxide siltstone ores, tails grades of 0.2% Cu were common (0.1% Cuinsol and 0.1% Cusol). The overall 0.39% Cu was a back calculated figure and includes:

- A higher than actual recovery credited to purchased materials.
- Losses associated with some minor pockets of sulphides.
- Higher losses in the high clay Tin Line Fault ore.
- Poor roasting of the core of course ore, which was shown to absorb copper ions.

11.0 Capital Costs

Records for the actual capital cost of Stage 2 equipment are no longer available. From the original feasibility study and the writer's memory the following capital cost estimate is presented.

	Feasibility *$*	*Revised Cost* *$*
Pre 1973 Expenses	20,378	
Feed Handling	64,074	
Roasting	1,169,550	1,500,000
Calcine Handling	48,330	
Grinding	248,294	
Agitation Leaching	27,405	
Filtering	533,615	750,000
Product Recovery	151,250	
Product Handling	23,910	
Residue Disposal	3,900	
Reagents NH3 & CO2	38,000	
Sampling	10,000	
Piping	69,000	
Electrical	75,127	
Steam Generation	5,070	
Engineering Facilities	29,000	
Spares & Stocks	72,750	
Mine Pumping	30,000	140,000
Site Establishment	30,000	
Eng. & Met. Research	250,000	
Commissioning	10,000	
Contingency	172,000	
Replacement Filter Station		500,000
TOTAL	**3,081,653**	

The cost revision is based on the following:

- The writer remembers the cost of the roaster to be $1,500,000.
- The writer remembers the cost of the Funda Filters to be $750,000.
- The cost of mine dewatering was approximately $10,000 per borehole or $140,000.
- The cost of the Eimco Drum filter was $120,000 and the writer believes the replacement cost of the filter station would have approximated $500,000.
- Therefore the estimated cost of Stage 2 was $4,238,000.

12.0 Marketing

Overseas marketing of cupric oxide was expanded during Stage 2 operations. Agents were appointed in Europe and North America to develop and grow our market share and bi-annual visits by our Marketing Manager to all customers were introduced. By the early 80's Burra was the largest single cupric oxide producer in the world with an estimated share of the world markets of 35 percent. Local sales to fertilisers for trace element supply was reduced in favour of export. The largest market was for the copper component in copper chrome asenate (CCA wood preserving) manufacture in North America and the UK. Other markets were:

- As a catalyst for phenol manufacture in America, Canada and Europe.
- As a colourant in ceramic Frits and glazes in USA, UK, Europe and Japan.
- As a brake lining component in Europe.
- As a growth promoter in animal feed (mainly pigs and poultry) in Europe, USA and UK.

13.0 Public Relations

Shortly after mine dewatering commenced the District Council of Burra Burra dammed the Burra Creek at a recreational area near the town centre. The previously dry creek helped boost tourism for the town and was the centre point for numerous community functions. At the completion of mining the Company drilled two bore holes close to the river and equipped them with 200 gal/min pumps and provided two spare pumps to the Council for replacement. The pumps are still operating today.

During mining the Golf Club Committee approached the Company to take water from our pipe line at the top of Bon Accord Shaft. The Golf Club installed the water line from the shaft to the golf course and the Company paid the pumping costs. When the water in Bon Accord Shaft dropped below the pump intake the Company back fed mine water from the process storage tank to the shaft and on to the golf course by gravity. The golf club used the water to develop the fairways the success of which is still evident today. The cricket club at Burra North also used the water for grass development around the oval wicket.

During the mining period the Company offered to use the waste being dumped to build a sporting complex for the future. The offer was declined due to recent Council expenditure on existing complexes. They requested we look at the layout for an airstrip. The Company surveyed the area for an airstrip, but nothing developed. In hindsight the Council and the Company should have started working together at an earlier stage to maximise the benefits from mining of waste.

Timber reclaimed from the old mining activities peaked at 6000 tonnes per annum. Fire wood was made available for collection at designated times free of charge for employees and "locals".

Stage 1 tailings was a much sort after commodity for vehicle driveways and yards. A nominal fee, administered by the employee social club, for non employees was introduced to prevent farm stockpiling and extend the life of the resource.

Process personnel were drawn from the local community, but tradesmen were employed from Adelaide and the mining sector. A large proportion of the workforce were young school leavers and local farm hands, the latter coming from a decreasing workforce requirement in the rural area. Samin employed and trained trades cadets in the electrical, fitting and boilermaker fields. Process personnel gained plant operating experience, which the area could not provide. Several young operators left Samin during the hard financial times, for the mining area of Kalgoorlie, where they were highly successful in the gold industry.

A unique chemical business was developed during Stage 1 & 2 operating periods and at the end of the ore resource many opportunities were explored to maintain an on-going business. Local copper scrap and cement production were the initial materials used and these were expanded to treat a range of raw materials. Cupric oxide is still being produced at the plant in Burra today.

Plate 3 was taken at the completion of mining prior to the cessation of mine dewatering. The two year ore stockpile behind the crusher is evident. The edge of the mined waste and tailings dams can be seen at the lower right of the photograph.

14.0 Acknowledgments

The writer thanks Mr. Neil Draper (Chief Metallurgist) and Mr. Jeff Schwier (Production Forman during Stage1 and Production Superintendent during Stage 2) for their comments and additions to this presentation prior to publication. The present owners of the cupric oxide processing business, AdChem, received a draft copy of this report and had no objection to the publication.

15.0 Bibliography

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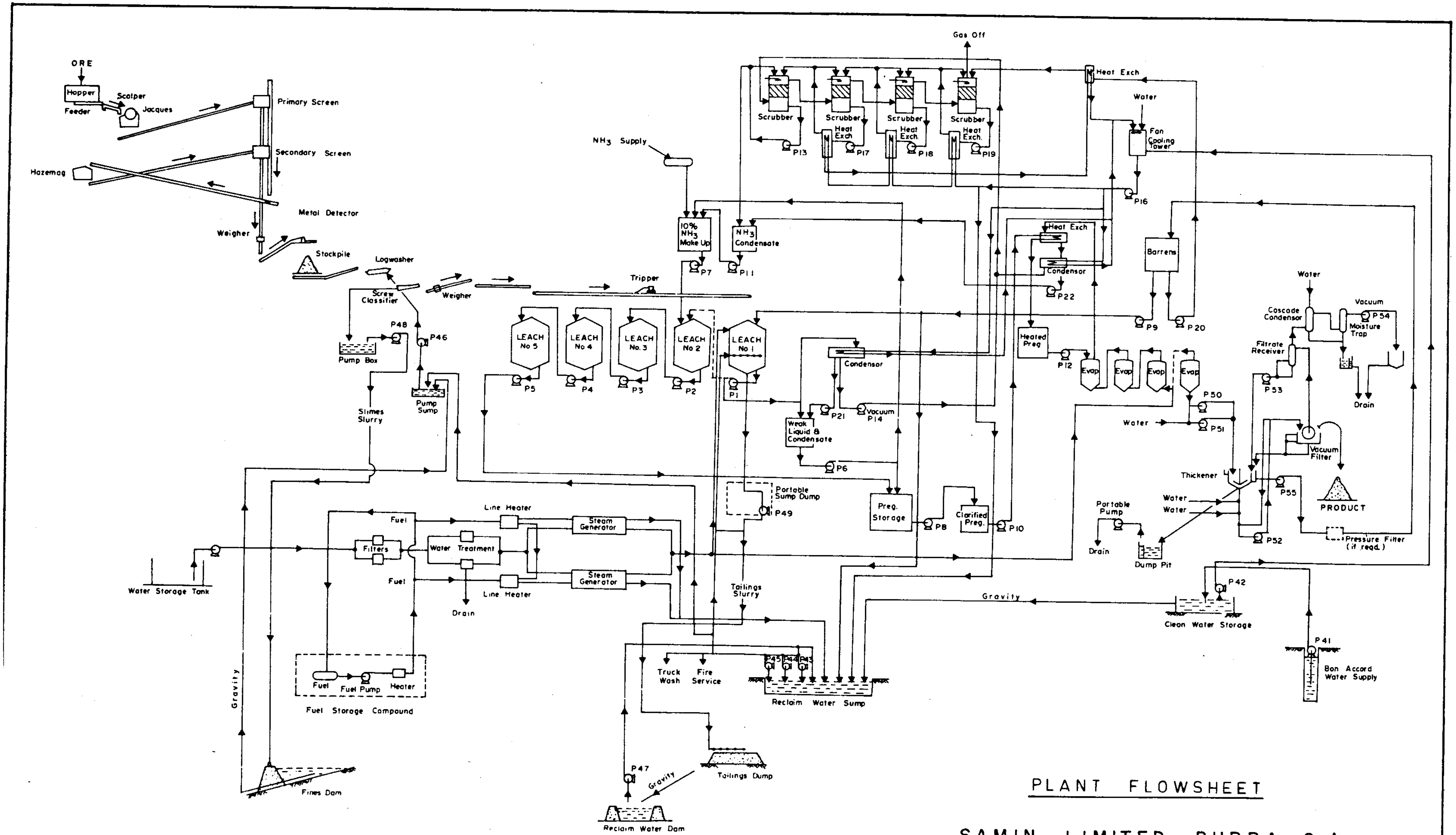
(2) Drexel, J F and McCallum, W S, 1986. Origin and age of the Burra copper ore body, Quart Geol Notes, 98, (Geol Surv S Aust).

PLATE 3 Aerial view of Burra Mine and Processing Plant, Jan 1981. PIRSA Photo 043488

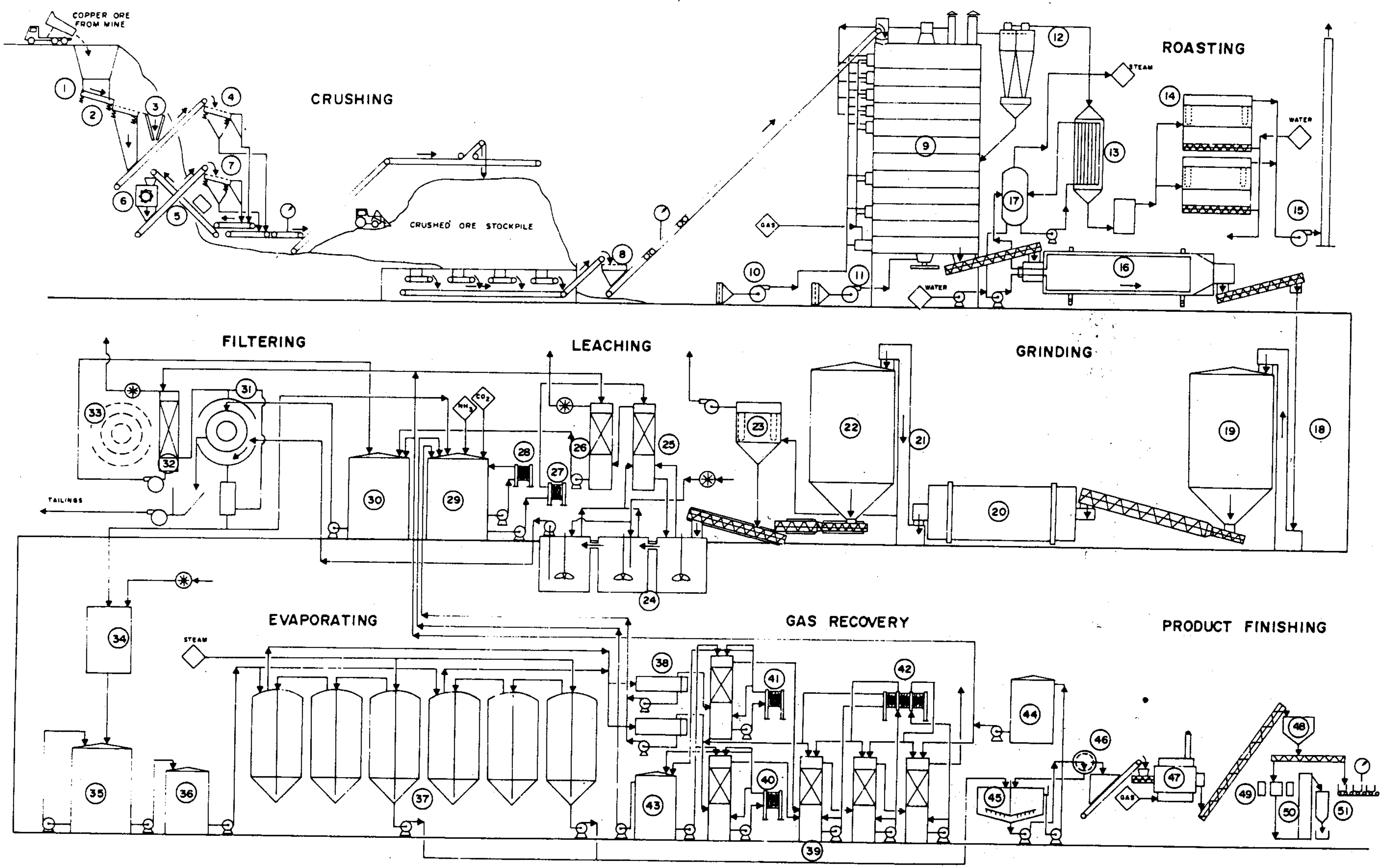
APPENDIX



PLANT FLOWSHEET

SAMIN LIMITED—BURRA S.A.

Stage I



SAMIN LIMITED

BURRA

PROCESS FLOWSHEET

STAGE 2

EQUIPMENT LISTS

Stage 2 Process Flowsheet

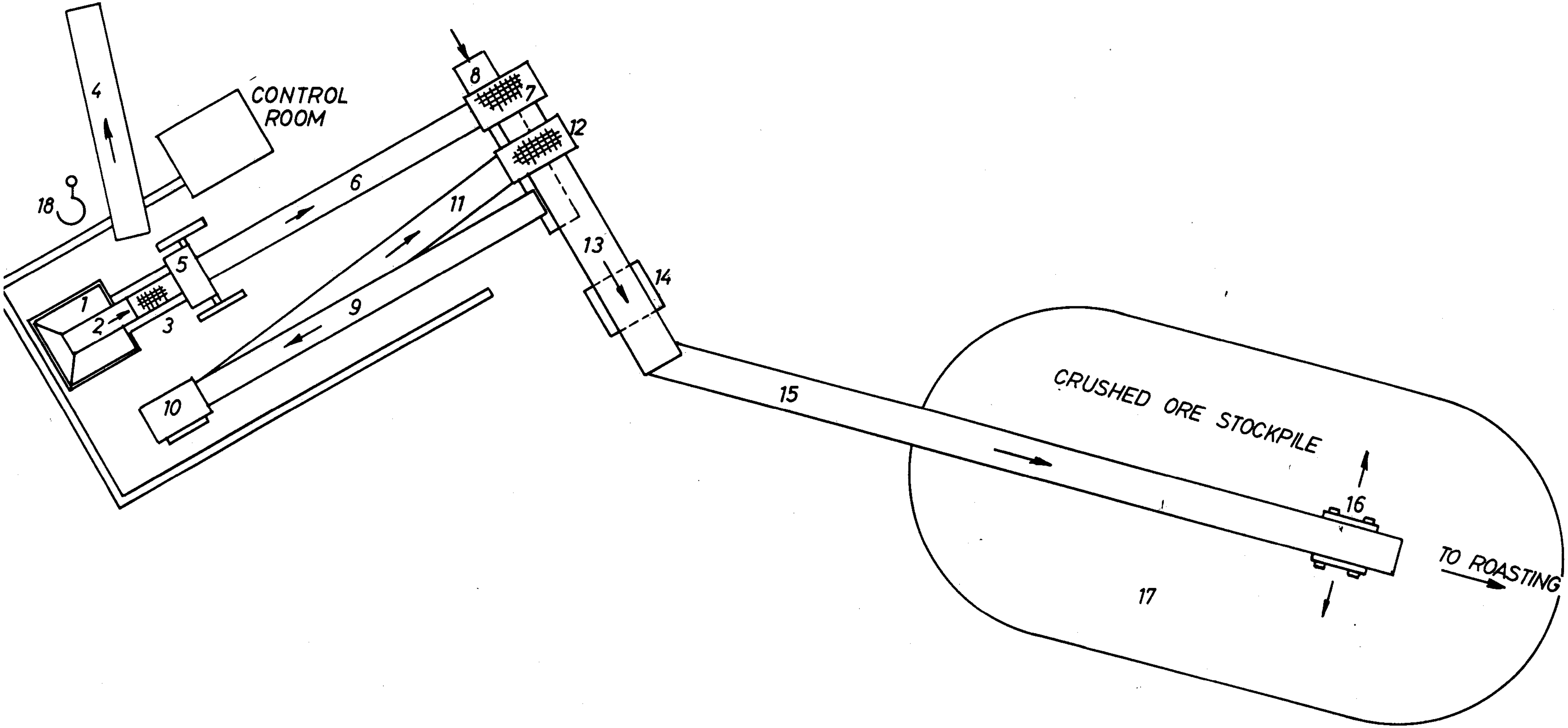
Item No.	*Description*
1	12ft Linkbelt Vibrating Feeder
2	8’x 4’ Linkbelt Scalping Screen
3	42”x30” Jaques Jaw Crusher 51/2”C.S.
4	10’x 5’ Primary Screen Aperture 1”x1/2”
5	Electro Magnet
6	AP4BR Hazemag Impact Crusher
7	10’x 5’ Secondary Screen Aperture 2”x 3/8”
8	4’x 2’ Vibrating Screen Aperture 1”x 1”
9	Envirotech 25’ 9” Diam 12 Hearth Roaster
10	31” Powermax Combustion Air Fan 100 hp
11	241/2” Powermax Shaft Cooling Fan 20 hp
12	4 x 48” Buell Cyclones
13	Economizer Rated 24,000 lbs/hr steam
14	Roaster Baghouse - 2 banks - Buell Norblo
15	55” Powermax ID Fan - 35,000 cfm - 150 hp
16	8’x 70’ Ore Cooler - 96 Tubes 50’ long
17	Steam Drum Accumulator
18	Bucket Elevator - 160 tph
19	No.1 Ore Bin - 250 t
20	Ball Mill - 9’x 17’ Vickers Ruwold
21	Bucket Elevator - 50 tph
22	No.2 Ore Bin - 250 t
23	Micropul Baghouse
24	Agitation Leach Tanks - 3 x 9,000 gal
25	Gas Scrubber
26	Gas Scrubber
27	Plate Cooler - Alfa Laval
28	Plate Cooler - Alfa Laval
29	Leach Liquor Storage Tank - 40,000 gal
30	Filter Wash Liquor Tank - 8,000 gal
31	14’x 18’ Eimco Vacuum Drum Filter
32	Gas Scrubber
33	Secondary Filtration - under construction
34	Pregnant Liquor Aeration Tank - 30,000 gal
35	Pregnant Liquor Storage Tank - 40,000 gal
36	Pregnant Liquor Measuring Tank - 8,000 gal
37	Evaporators - 6 x 16,000 gal
38	Steam Vapour Condensers - 2 x Multipass Tube
39	Gas Scrubbers (5) - 2 Parallel - 3 Series
40	Plate Cooler - Alfa Laval
41	Plate Cooler - Alfa Laval
42	Plate Cooler - Triple Unit - Alfa Laval
43	Condensate ex Evaporator Tank - 8,000 gal
44	Barren Liquor Tank - 40,000 gal
45	Thickener - 20’ Diameter Dorr Oliver
46	Vacuum Filter - Samin Design - 6’ Diameter
47	4’x 21’ Rotary Dryer
48	Product Storage Bin - 15 t
49	Vibrating Ball Mill - 15”x 18” Allis Chalmers
50	Bucket Elevator
51	Fine Product Bin - 10 t

Ore Crushing

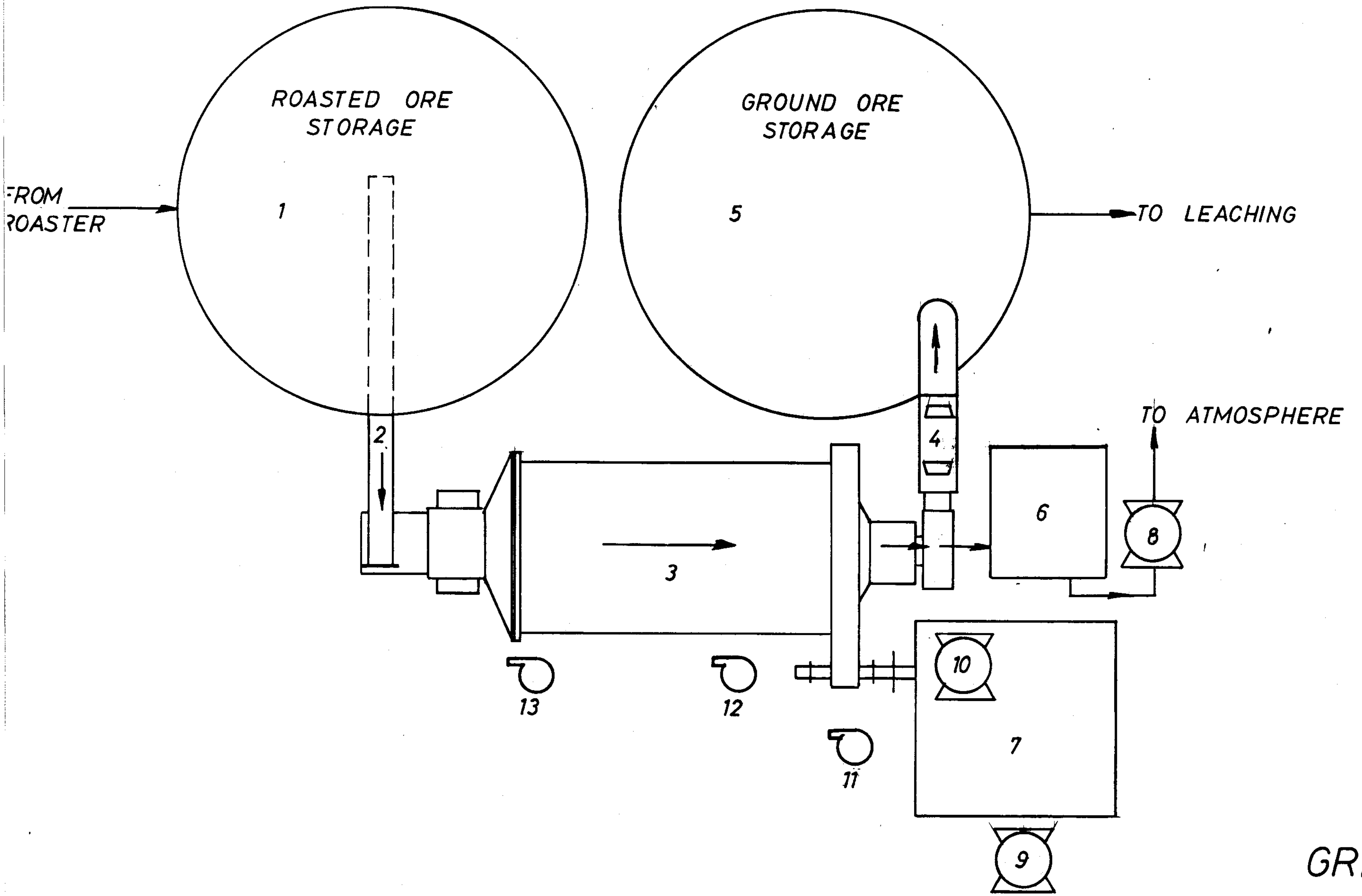
Item No.	*Desription*
1	Receival Hopper - 50t
2	Linkbelt Vibrating Feeder - 12 ft
3	Linkbelt 8'x 4'Vibrating Scalping Screen - 21/2" Aperture
4	Mine Timber Disposal Conveyer
5	42" x 30" Jaques Single Toggle Jaw Crusher
6	Primary Screen Feed Belt
7	Linkbelt 10' x 5' Primary Screen - 1/2" Aperture
8	Primary Screen O/S Belt
9	Secondary Crusher Feed Belt
10	AP4BR Hazemag Impact Crusher
11	Secondary Screen Feed Belt
12	Linkbelt 10' x 5' Secondary Screen - 1/2" Aperture
13	Crushed Ore Conveyer
14	Crushed Ore Weightometer
15	Crushed Ore Conveyer
16	Crushed Ore Conveyer Tripper Car
17	Crushed Ore Stockpile - 4,500 t
18	Mine Timber Crane

Ore Grinding

Item No.	*Desription*
1	Roasted Ore Storage Bin - 250 t
2	Mill Feed Screw Conveyer
3	Vickers Ruwolt 9' x 17' Ball Mill
4	Ground Ore Bucket Elevator
5	Ground Ore Storage Bin- 250 t
6	Micropul Baghouse
7	Ball Mill Motor House
8	Ball Mill Air Sweep Fan
9	Ball Mill Motor House Pressurising Fan and Filter
10	Ball Mill Motor - 11kV 1000hp
11	Ball Mill Bull-Ring Gear and Pinion Lubrication Pump
12	Ball Mill Bearing drive end Lubrication Pump
13	Ball Mill Bearing non-drive end Lubrication Pump



ORE CRUSHING


ROASTED ORE STORAGE
GROUND ORE STORAGE
FROM ROASTER
1
5
TO LEACHING
2
4
TO ATMOSPHERE
3
6
8
10
7
13
12
11
9

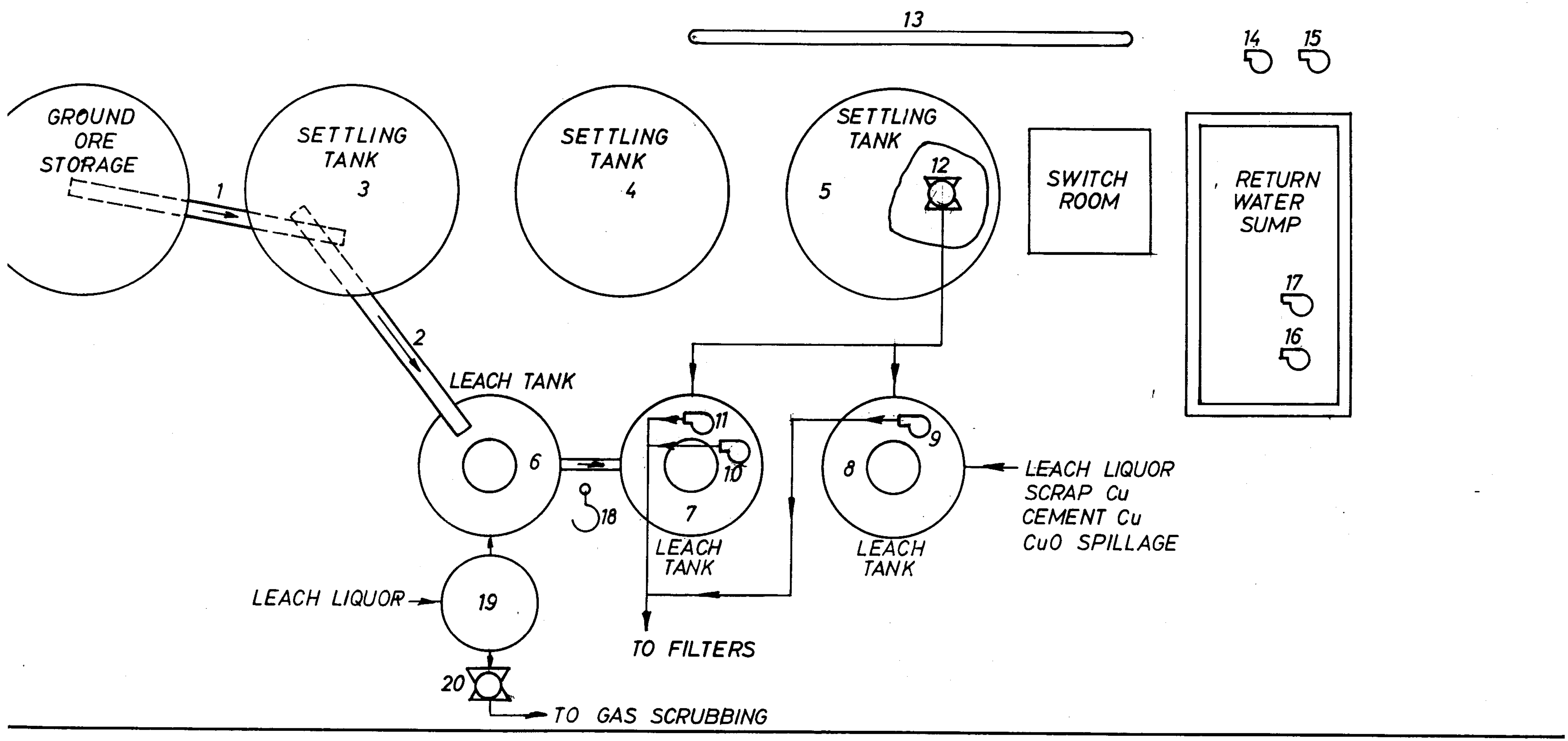
GRINDING

Ore Leaching and Service Water

Item No.	*Description*
1	Ground Ore Feed Variable Speed Screw Conveyer
2	Ground Ore Feed Screw Conveyer
3	Settling Tank No.3)
4	Settling Tank No.4) Solutions ex Filters
5	Settling Tank No.5)
6	Leach Tank No.1
7	Leach Tank No.2
8	Leach Tank No.3
9	Purchased Material Leach Pump
10	Leach Slurry Pump
11	Leach Slurry Pump
12	Leach Air Blower
13	Service Water Header Main
14 & 15	Service Water Pumps
16 & 17	Tails Water Pumps
18	Leach Service Crane
19	Gas Scrubber
20	Leach Section Draft Fan

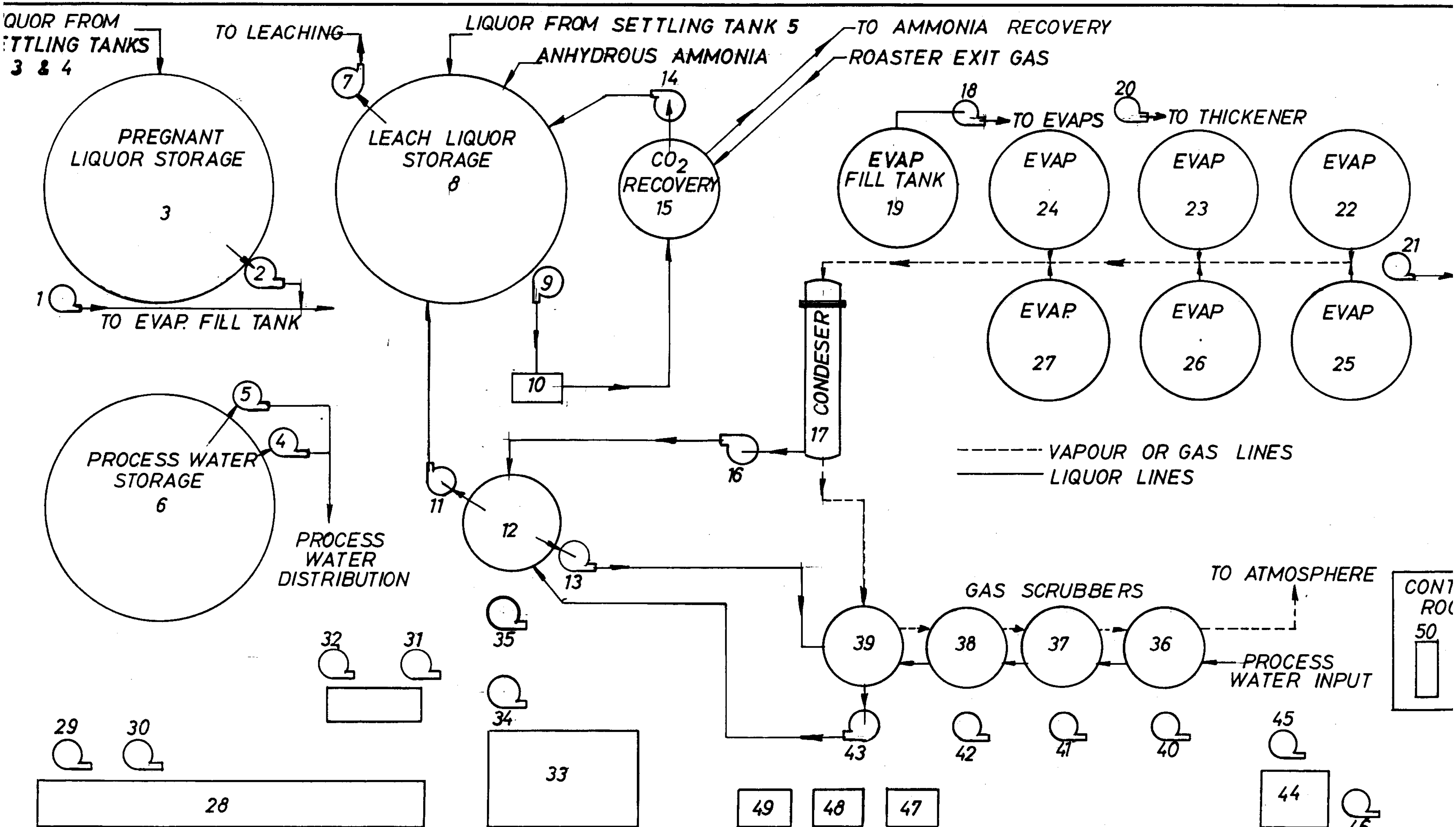
ORE LEACHING

SERVICE WATER RECLAIM & RETICULATION



Distillation and Tank Farm

Item No.	*Description*
1 & 2	Evaporator Fill Tank Pump
3	Pregnant Storage Tank - 40,000 gal
4 & 5	Process (Barrens) Water Distribution
6	Process (Barrens) Water Storage
7	Leach Liquor Pump
8	Leach Liquor Storage - 40,000 gal
9	Leach Liquor Pump to CO2 Recovery
10	Leach Liquor Cooling PHE
11	Condensate ex Evaporator Pump
12	Condensate ex Evaporator Storage Tank
13	Scrubber 4 & 5 Feed Liquor Pump
14	Leach Liquor CO2 Recovery Return Pump
15	CO2 Recovery Scrubber
16	Condenser Condensate Pump
17	Condensers 1 & 2
18	Evaporator Fill Pump
19	Evaporator Fill Measuring Tank
20 & 21	Evaporator Discharge Pumps
22, 23 & 24	Evaporator 1, 2 & 3 - Line 1
27, 26 & 25	Evaporatos 4, 5 & 6 - Line 2
28	Condenser Cooling Tower - CT 1
29 & 30	CT 1 Chemical Dosage Pumps
31 & 32	CT 1 Water Pumps
33	PHE Cooling Tower - CT 2
34	CT 2 Chemical Dosage Pumps
35	CT 2 Water Pump
36, 37 & 38	Scrubbers 1, 2 & 3
39	Scrubbers 4 & 5 in parallel
40, 41 & 42	Scrubbers 1, 2 & 3 Pumps respectively
43	Scrubbers 4 & 5 Pumps
44	PHE Cooling Tower - CT 3
45	CT 3 Chemical Dosage Pumps
46	CT 3 Water Pump
47	PHE for Scrubbers 1, 2 & 3
48	PHE for Scrubber 4
49	PHE for Scrubber 5
50	Distillation Section Control Room



EVAPORATION