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EURALBA MINING LIMITED

MT LYELL PROJECT

ANALYSIS OF TEST RESULTS '91/92
PRELIMINARY FINANCIAL FEASIBILITY

AUGUST 1992

AMG REFERENCE POINTS ADDED

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EXECUTIVE SUMMARY

The following report comprises a detailed analysis of the results of two test programmes aimed at the assessment of viability of extraction of copper from waste streams generated by the Mount Lyell copper mine in Tasmania.

The investigation has been based on the use of the Euralba Mining Limited Electrowin (EMEW) cell, a technology which has been under development for approximately three years. The principal characteristic of the new cell, proven in a number of test programmes, is its capability to electrowin metals from low to very low grade solutions at higher operating and cost efficiency than conventional systems.

Through extensive mining activity over the past 100 years substantial quantities of low grade copper bearing 'waste ore' have been generated at Mount Lyell. This waste now represents a continuing source of low grade copper bearing solutions.

The waste liquors are the result of high local rainfall and degeneration of minerals in large surface dumps (in excess of 40 million tonnes) and broken underground ore (mine dewatering). Site personnel estimate that some 800 tonnes of contained copper currently leaves the site per annum.

The study undertaken here envisages the possibility of installation of a treatment plant sized to a minimum daily production rate of 1.2 tonnes per day of copper.

Copper concentration in the Mount Lyell waste solutions (between 100 and 150 ppm Cu - 0.1 to 0.15 g/l) is approximately 400 times lower than that generally fed to conventional electrowin circuits. Although mechanisms for upgrading the solution (for example solvent extraction) are available, the capital cost entailed in installation of a conventional plant (SX/EW) are particularly prohibitive in the light of limited initial copper output.

The two investigations conducted by Euralba have examined performance of the new cell on these streams with a view to two potential treatment options:

a) Direct electrowinning, on a 'one pass' basis through a series of cells linked in hydraulic series (DIRECT ELECTROWIN), and

b) Liquor upgrading through a simplified solvent extraction circuit, followed by electrowinning through the new cell (SX/EMEW).

The latter recognises that there is considerable potential for simplification of solvent extraction (and therefore reduction in cost) through use of the cell.

The report attached hereto examines all of the physical and chemical parameters that would be expected to affect copper recovery performance in treating the Mt Lyell liquor; and analyses in that context the results of approximately 50 individual test runs completed to date.

In direct application of the technology to the waste stream, it has been shown to be capable of stripping copper to less than 1 ppm.

Although achieving significantly higher performance levels than would be possible using a conventional cell, overall efficiency (copper recovered versus theoretical yield for a specified current input) in direct electrowinning has remained lower than would be achievable on a primary leach liquor (of, say 1-3 g/l Cu).

The testwork has shown that a variety of chemical characteristics of the target liquor inhibit efficient electrowinning - including the low copper grade and the presence of competing metals which 'waste' power through cyclic oxidation and reduction reactions in the cell. The most important factor has been found to be the 'unfavourable' initial valency state of the high iron levels in solution (iron in the form of ferric sulphate being responsible for redissolution of electrowon copper).

The operating parameters demonstrated by the testwork to result in improved process performance have, therefore, been largely those that minimise dissolution of copper by ferric sulphate (or promote reduction to ferrous sulphate). Optimum initial performance has been achieved at low solution flow rate and high (comparative) current density (amps applied per square metre of cathode).

Due to the progressive changes in chemistry brought about by electrolysis, the effect of competing metals in solution changes continuously through treatment. It has been found, however, that the treatment cycle can be broken into a number of stages during which specific operating conditions result in optimum performance. Detailed analysis of the results of each test run have allowed modelling of an optimum circuit for the direct treatment route.

The model generated entails division of the circuit into two stages, operating under differing flow and current density conditions. Projected performance of the circuit would result in:

Copper recovery	:	90%
Overall EW power consumption	:	18.4 mWh/tonne copper

These figures are based on measured current efficiency levels between 13 and 16%. It should be noted, however, that overall current efficiency in excess of 30% has been achieved in liquor that has been 'pre-treated' to achieve limited iron reduction prior to electrowinning.

Due to high resistivity in the target solution direct electrowinning is only feasible at relatively low current density (between 50 and 70 A/m^2).

Financial analysis of the direct electrowin model demonstrates that, albeit low in efficiency, the treatment route offers significant potential for commercial viability from an operating cost standpoint.

Due to the relatively low current efficiency projected for this circuit, coupled with low operating current density, a large number of individual cells is required for treatment of the Mt Lyell liquor on a single pass basis. A requirement for approximately 8,600 cells, calculated for the above model, has obvious impact on projected capital cost.

Studies undertaken in conjunction with the direct electrowin testwork have clearly demonstrated that the application of solvent extraction, prior to electrowinning, has a number of significant advantages over the direct treatment route. Inter alia, these include:

* The effect of high iron concentration in the liquor is almost entirely obviated.

* Current efficiency within the range of liquor tenors expected in the electrowin circuit (50-10 g/l Cu) has been shown to be maintainable throughout at close to theoretical values.

* Resistivity of the resultant liquor will be far lower than the original waste, with the result that low voltage is maintained even at very high current density.

* Power consumption in electrowinning is reduced by a factor of at least six, with minor additional power cost in solvent extraction.

* The number of cells required to achieve projected minimum copper production (1.2 tonne/day) is reduced by a factor of 40 (to approximately 225 units).

A preliminary financial analysis of both potential treatment systems has been undertaken. Although both demonstrably offer potential commercial viability, the process advantages outlined above and their obvious financial impact lead to a preference for the SX/EMEW system for a future site facility.

Capital cost for the direct treatment route has been estimated on the basis of well established unit cost for the cells, allowing certain reductions due to the number of units required. Ancillary equipment has been based on the power requirements and projected liquor flow to the facility.

A projection of capital cost for the solvent extraction stage of the SX/EMEW system has been formulated, on behalf of Euralba Mining Limited, by Minproc Engineering (a firm well experienced in the field of conventional SX/EW technology). The required electrowin circuit has been sized and costed on the basis of results obtained during the current programme.

Operating costs for both electrowin circuits have been estimated on the basis of information recorded from the current programme, coupled with modelling and analysis thereof. Reagent and operating costs of the possible solvent extraction circuit have been calculated (utilising 'shake-out' information generated by Henkel Corporation) by Minproc Engineers.

Analysis of projected operating return has assumed conservative figures for product value from the two circuits, pending establishment of quality through continuous production.

The results of the studies undertaken (based on initial production of 1.2 tonnes per day of copper) may be summarised as follows:

	Treatment Options	
	SX/EMEW	DIRECT ELECTROWIN
PROJECTED CAPEX	$ 1,584,000	$ 2,587,000
ANNUAL SALES REVENUE	1,456,350	1,379,700
TREATMENT COSTS	555,530	750,075
PROJECTED SURPLUS	900,820	629,625
- AS % OF CAPITAL	56.9%	24.3%

It is suggested that production from a site facility could be readily expanded, through promotion of copper leaching and sourcing of liquor from subsidiary sources. Such expansion could be achieved at relatively low cost in the SX/EMEW option (the solvent extraction circuit being sized to liquor throughput rather than copper tenor), but would entail progressive purchase of larger numbers of cell for the direct electrowin system.

From a point of view of versatility for future expansion, therefore, the SX/EMEW process must again be preferred.

No formal analysis of the cost and result of any future expansion has been carried out here. However, any conclusion based on the results of this study must factor in the substantial size of the remnant resource in the Mount Lyell orebodies (and their low grade halo) as a long term source of leachable copper.

The report appended here strongly recommends immediate implementation of continued and detailed costing of the SX/EMEW treatment process as a facility for long term treatment of the Mt Lyell waste streams.

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1.0 INTRODUCTION

The following report has been compiled by Euralba Mining Ltd for presentation to Renison Golfields Consolidated Limited, as a comprehensive review of testwork conducted on the feasibility of viable extraction of copper from waste liquor streams at Mt Lyell, in Tasmania (Figure 1).

Mt Lyell mine has been a source of primary copper ore for more than a century. Its development commenced in the 1880's with selective mining of a series of high grade orebodies, by a variety of open cut and underground methods. The establishment of the Mount Lyell Mining and Railway Company in 1893, to develop the extensive Iron Blow deposit, marked the commencement of large scale mining in the area and bulk mining has been continuous since that time.

Mining and treatment methods employed on the Mount Lyell lease have varied over the years depending on the nature of the orebodies and beneficiation technologies available. Bulk open cut and underground mining have been prevalent through most of the mine's history, as has treatment by on-site smelting. It was only in 1969 that the current ore treatment method, grinding and floatation to produce a copper concentrate, was established.

The mine currently produces approximately 1.5 million tonnes of ore per annum, at an average grade of approximately 1.6% copper. The ore, all of which is produced by bulk (but selective) underground mining, is treated to produce a 25%+ concentrate through the site plant. Tailings from the latter are discharged directly into the Queen River.

Large quantities of waste material have been built up over the years of development of this mine. The largest of these comprise:

a) In excess of 40 million tonnes of low grade waste rock from the old West Lyell open cut, and

b) large quantities of broken low grade waste from selective mining, which have been allowed to cave into the workings beneath the old open cut (to a present depth of around 200 metres below sea level).

Due to the high rainfall in this area, both of these now represent a long term source of copper bearing waste liquor. Site personnel have estimated a total of approximately 800 tonnes copper contained in solutions leaving the site each year.

The current study has concentrated on extraction of copper from these two principal sources only, but has been cogniscant of the probability that in time copper leaching from both can be enhanced and output increased.

The programme conducted by Euralba, between December 1991 and May 1992, has addressed the possibility of viable extraction of this copper, based on the use of a novel electrowin technology which the company has under development.

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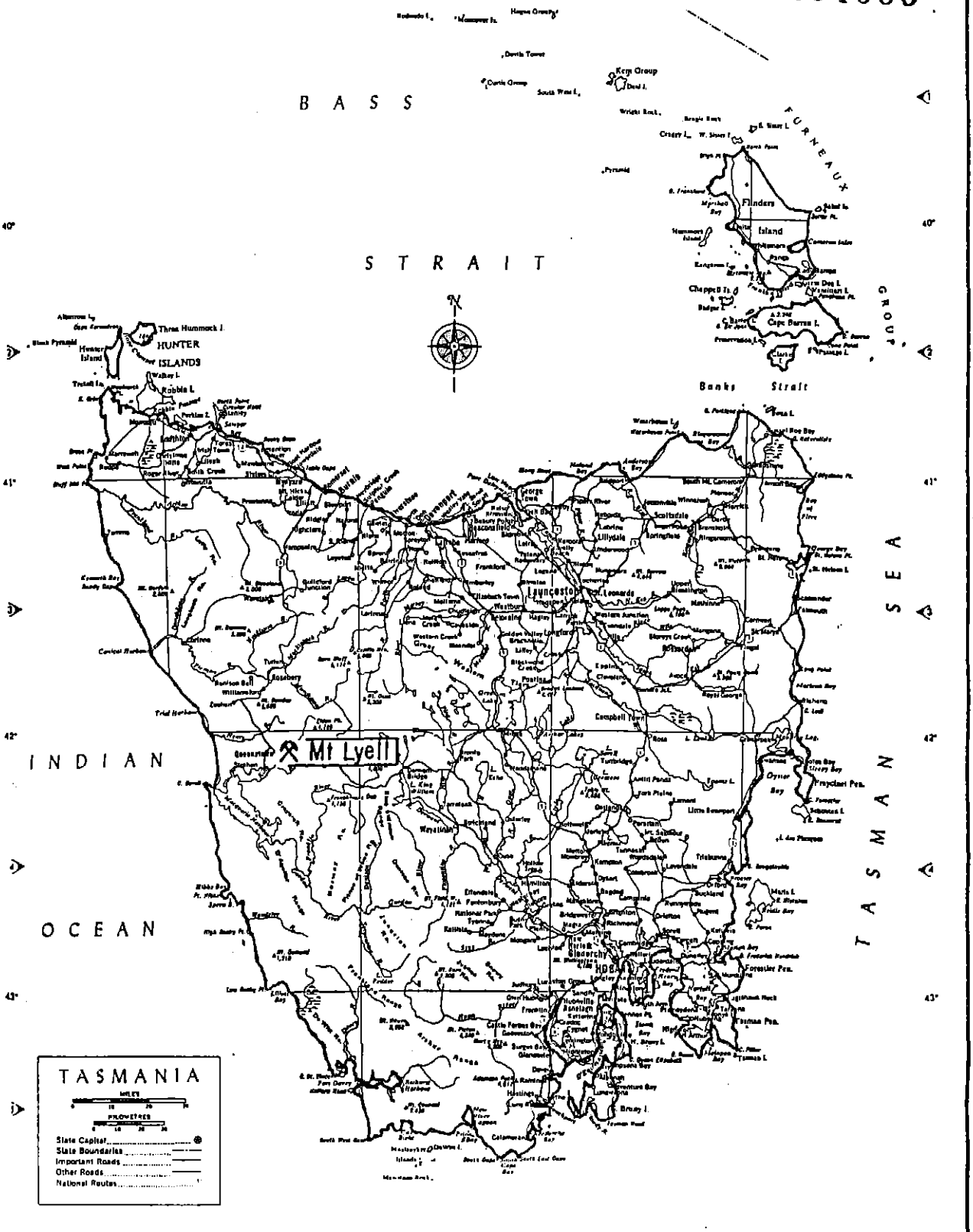


CONSULTANCY: Australian Geoscientists Pty Ltd		
CLIENT: Euralba Mining Limited		
PROJECT: Mt. Lyell		
TITLE: GENERAL LOCATION		
Author: T.Treasure	Date: August 1992	Dwg No:
Drawn: Down Under	Scale:	FIGURE: 1

Physical and operating aspects of the technology have been addressed in full in previous reports. Its potential viability in the current context arises through its proven capability to achieve copper recovery from low grade solutions, at far greater efficiency than conventionally available systems. The improvement possible is a result of high mass transfer capabilities of the novel cell design upon which the technology is based.

The study of the Mount Lyell waste streams has followed a methodical staged programme of:

a) Laboratory treatment of samples of liquor provided, under a wide variety of electrowin operating conditions.

b) Methodical operation of production scale cells on site.

Results from the initial study have been reported in full under separate cover ('Euralba Mining Limited, Mount Lyell Project' - Preliminary Test Report, Jan 1992). In addressing the results of the experimental work, this report concluded that the technology offered a potentially viable route for extraction of copper from the waste streams.

The subsequent site programme has investigated performance of the system on larger scale liquor samples, and has been aimed at the closer definition of potentially optimum operating parameters. The study has examined use of the new cell as an integral component of two possible treatment routes for these liquors:

a) Direct electrowinning through series of cells linked in hydraulic series, with no preliminary liquor treatment.

b) Initial liquor treatment by solvent extraction, followed by copper electrowinning utilising the new cell.

The following report concentrates on the results of the site programme, integrating those of the preliminary investigation where appropriate.

Detailed discussion is provided for chemical and physical parameters which have been shown to affect the performance of this technology at Mount Lyell; and of the rationale behind selection of settings regarded as appropriate for modelling of potential treatment circuits. A summary of the general effect that can be expected under differing operating conditions is followed by detailed analysis of individual parameters.

A preliminary financial analysis of both of the above treatment systems is provided in this report, along with recommendations for further development of the project.

2.0 OPERATING HARDWARE

Four separate test circuits were operated during the Mt Lyell site programme:

A: 4" Circuit - 5 x 100mm cells in hydraulic series
B: 2" Circuit - 10 x 50mm cells in hydraulic series
C: 'Drum' Circuit - single 50mm production cell
D: Laboratory circuit - single 35mm short cell

A number of photographs of the site facility are provided in Appendix 1 herewith.

Circuits A and B were utilised for the main productive tests; Circuit C for additional data collection. Circuit D was used to perform quantitative high grade tests, on liquor fabricated to duplicate the result of solvent extraction upgrading of the waste stream.

Tests aimed at defining progressive changes in liquor chemistry more closely were also performed in the latter.

To allow direct comparison of performance between test runs, the number of cells installed in the two major circuits was tailored to achieve approximately equivalent total cathode area.

Similarly to the initial laboratory work, all tests were performed on liquor circulated continuously through the circuits from a storage tank. The erection of a circuit 'on-line' with the waste stream is not be feasible, unless sufficient cells are installed to achieve depletion to target copper levels in a single pass. Progressive changes in liquor chemistry, throughout treatment, continually affect process performance- to the degree that installation of a partial circuit on line with the stream would examine only a minor portion of the treatment cycle.

Detailed analysis of performance at various stages in the test runs allows valid modelling of projected performance in a full scale treatment system.

The investigation been based totally around the use of the novel electrowin cell being developed by Euralba Mining Limited. Full details of cell design and performance to previous testwork are not provided here; the reader being referred to previous texts for this information.

The following hardware configurations were used in the Mount Lyell site programme:

4" CIRCUIT

Cells	:	Number in series	:	5 cells
		Cathode diameter	:	100 mm (4 inch)
		Cathode length	:	750 mm
		Cathode material	:	Copper
		Anode diameter	:	25 mm
Liquor storage			:	3,500 litres
Pumping			:	Onga KNA43 chemical pump
Flow measurement			:	Gemu rotameter type 807-50
Flow control			:	Valve at end of circuit
Rectifier			:	50A/60v, variable voltage

2" CIRCUIT

Cells	:	Number in series	:	10 cells
		Cathode diameter	:	54 mm (2 inch)
		Cathode length	:	850 mm
		Cathode material	:	Copper
		Anode diameter	:	25 mm
Liquor storage			:	2,650 litres
Pumping			:	Grundfoss CRN 2-70, 3-phase
Flow measurement			:	Gemu rotameter type 807-50
Flow control			:	Valve at circuit outlet
Rectifier			:	50A/60v, variable voltage

'DRUM' CIRCUIT

Cells	:	Number in series	:	1 cell
		Cathode diameter	:	50 mm
		Cathode length	:	850 mm
		Cathode material	:	Copper, stainless steel
		Anode diameter	:	25 mm
Liquor storage			:	210 litres
Pumping			:	Iwaki, model no. MD70R/10
Flow measurement			:	Manual
Flow control			:	Valve at circuit outlet
Rectifier			:	50A/60v, variable voltage

LABORATORY CIRCUIT

Cells	:	Number in series	:	1
		Cathode diameter	:	32.5 mm
		Cathode length	:	200 mm
		Cathode material	:	Copper
		Anode diameter	:	9 mm
Liquor storage			:	10 litres
Pumping			:	Iwaki, model no. MD30/RZ
Flow measurement			:	Gemu rotameter, type 807
Flow control			:	Valve at circuit outlet
Rectifier			:	15A/12v, variable voltage

3.0 OPERATING VARIABLES (General Introduction)

The testwork was undertaken over a relatively wide range of operating conditions; and was aimed at generating a comprehensive data base from which a future treatment operation could be designed (and costed).

Differing hardware configurations were operated under a variety of current and flow conditions. Coupled with variations in liquor chemistry (both imposed and due to climatic conditions), this work has produced valid data which would be representative of most potential operating conditions at the site.

The following notes provide a summary of variations in operating conditions during the testwork; in addition to their expected effect and the rationale behind changes imposed.

3.1 CELL DIMENSIONS

The choice of cell dimension will impact on the capital and operating costs of a treatment plant for Mt Lyell (controlling as it does the number of units required to achieve the target treatment level).

Diameter :

a) The productive capacity of the cell is a function of the surface area of cathode, vs applied current coupled with operating efficiency.

b) Operating voltage of the cell (and therefore 'power cost') is a function both of liquor chemistry and electrode spacing (in turn, governed by respective diameters of cathode and anode).

Length : Being the second factor that governs cathode surface area in the cell, unit length has direct influence on productive capacity of individual units.

In the current exercise, respective diameters of the two electrodes is the more significant feature. The Mt Lyell waste being a relatively 'poor' electrolyte, the marked effect of electrode spacing on cell voltage impacts significantly on projected operating costs. Electrode spacing has further effect in governing to a degree the harvest cycle of cathodes (electrode spacing vs. 'thickness' of plate achievable prior to replacement).

During the site programme only a single anode diameter (25mm) was available for testing on the production sized cells. Study of the effects of electrode spacing was achieved through varying cathode rather than anode diameter. In a production plant, it is envisaged that this factor will be controlled through the latter.

Since a large proportion of the capital cost of installing the new process is in fabrication of individual cells, any opportunity for reduction in the number required in a treatment facility should be taken (within process limits).

The geometry of the new cell is such that considerable dimensional variation is possible, without altering its basic design. Electrode spacing can be tailored using a variety of electrode diameters, and cathode length can be reasonably adapted to suit cell productivity requirements.

The following dimensions were examined during the Mt Lyell site programme:

'4 inch' cell

Cathode diameter (I.D. mm)	:	99.40
Anode diameter (O.D. mm)	:	25.00
Cathode length (mm)	:	750.00
Cathode area (m^2)	:	0.23

'2 inch' cell

Cathode diameter (I.D. mm)	:	52.00
Anode diameter (O.D. mm)	:	25.00
Cathode length (mm)	:	850.00
Cathode area (m^2)	:	0.14

'Laboratory' cell

Cathode diameter (I.D. mm)	:	33.00
Anode diameter (O.D. mm)	:	9.00
Cathode length (mm)	:	200.00
Cathode area (m^2)	:	0.02

3.2 CURRENT DENSITY

Current density is a measure of current applied in an electrical circuit, expressed as amps per square metre of electrode area. In conventional technology this value is equivalent for both anode and cathode (electrodes generally being the same size). In the case of the current cell, where respective electrode area varies considerably, the measurement is referred to cathode.

Under ideal conditions, the productive capacity of an electrowin cell is directly related to applied current - yield being 1.18 grams of copper per amp hour applied to the cell (Faraday's Constant for a cupric solution). In practice, competing reactions and inadequate 'supply' of metal ions to the cathode reduce theoretical yield. The resulting performance factor (actual vs theoretical yield) is expressed as 'current efficiency'.

Manipulating applied current in a cell has several well established effects, the more significant being:

a) Increasing current results in increased voltage for a given electrolyte, and thus

b) In a given electrolyte, power efficiency is generally greater at lower current density.

c) The 'quality' or 'nature' of plating is affected.

Development work over the past two years has continuously shown that the new cell is significantly less affected by b) and c) than conventional systems. This is a feature of its high mass transfer capabilities, which avoids the voltage increase normally suffered via the steep concentration gradient at the cathode boundary layer when operating at high current density.

The process does not however deny established electrochemical principles. Low conductivity of the target solution contributes to relatively high cell voltage, which is exacerbated by rise in current density. Given a specified electrode gap, current density limits for 'viable' power consumption vs. copper production are readily defined.

In both laboratory and site testing, priority has been given to the assessment of varying current settings on:

a) cell voltage, at various electrode gap spacings
b) current efficiency
c) product nature and quality

The following settings were utilised in the Mt Lyell programme:

Cell diameter (mm)	Electrode spacing (mm)	Cell current (A)	Equivalent current density (A/m^2)	
100	37.25	10.0	42	
		15.0	64	
50	13.50	5.0	35	
		7.5	52	
		10.0	69	
		15.0	104	
35	12.00	1.0	60	
		2.0	121	
		4.0	242	
		6.5	394	)
		8.5	515	) High grade
		10.2	618	) liquor only
		15.0	909	)

3.3 LIQUOR FLOW

Characteristically, improved performance in the new cell is a feature of more effective 'supply' of copper ions to the cathode surface. Variation in flow rate through the system can therefore govern process performance.

The various cell sizes utilised in this programme were operated at variable flow rate, as follows:

Circuit	Flow (l/hour)
4 Inch	2400
	3800
	4200
2 Inch	1400
	1800
	2400
	3400
Laboratory	240
	480
	720
	910

Operation of each circuit at all of the indicated flow settings, in conjunction with each of the other variables imposed, would have been impractical. However sufficient individual tests were undertaken, where the only variable was flow, to provide sound predictive data.

Initial planning of flow settings for the site programme was based on scale-up calculations from the laboratory work. Prediction of flow requirement in a full sized production cell related to calculation of cathode cross-sectional area. Data collected during the field programme now suggest that this comparison is more valid on a cathode circumference basis (performance of the larger cells matching that of the smaller units at lower than indicated optimum flow rate).

The initial laboratory work suggested that there were limitations to practical flow settings, due to rising pumping cost with increasing flow (due to resultant increase in 'pressure drop' across the cells). The calculations performed suggested 1800 litres/hour to be an upper limit for viable operation of the 2" cells.

The Mt Lyell programme therefore included pressure testing of the two main cell designs. Inlet and outlet pressure was measured, both on individual cells and groups in hydraulic series; and on specially modified cells with enlarged inlet and outlet spigots.

3.4 LIQUOR CHEMISTRY

The range of chemical conditions assessed, through the 50 or so tests completed to date, is regarded to be representative of varying conditions that may be encountered in a treatment facility for Mt Lyell.

The liquor contains a number of contaminants including copper, iron, aluminium and manganese. Certain of these have been shown to have marked effect on process performance. The programme has therefore concentrated on the assessment of changing concentrations of the following:

- Copper - as the target metal.
- Iron - as the principal contaminant; and that most likely to affect the efficiency of copper electrowinning.
- pH - as an indicator of conductivity of the solution and a gauge of changing chemistry under progressive electrolysis.

Variations in liquor chemistry during this programme have been brought about:

a) Through chemical treatment of the liquor (for example pH treatment to remove iron in the laboratory).

b) By climatic variations on site (dilution of the source material).

c) By sourcing solution from alternate sites.

d) By fabrication to suit likely future treatment options (examination of possible solvent extraction route).

Differing chemical conditions have resulted in the following ranges for significant (deemed for this programme) constituents prior to commencement of electrowinning:

Cu	:	90 - 170 ppm
Fe	:	230 - 2370 ppm
Al	:	400 - 850 ppm
pH	:	2.2 - 2.7

'High' grade tests were performed on a fabricated liquor intended to be representative of the result of a solvent extraction 'upgrade' of the waste stream. Sufficient copper sulphate was added (to a sample containing fresh water mixed with waste stream, in a ratio of 80:20) to achieve an initial liquor tenor of 40 g/l Cu. The resulting contaminants were of the order that would be expected with this treatment route.

All test runs were operated continuously beyond the point of effectively complete depletion of copper, chemical and physical conditions being monitored throughout the treatment cycle.

The detailed cross section of data collected, coupled with the results of previous programmes, allows valid predictions to be made for a future treatment circuit.

3.5 OTHER

As mentioned above, the method of liquor circulation for these tests differed from the route that would be chosen for a permanent facility. Due to the effect that progressive changes in chemistry have on process performance, 'batch' sample treatment was utilised. It is envisaged that a production unit would treat the liquor, to 'completion', in one pass through a bank of cells in hydraulic series.

Progressively larger samples have been used in programme to date, to allow close definition of changes in process performance due to chemical and operating variations. Although not strictly a process variable, therefore, note should be made of the differing sample volumes used for the various circuits:

4 Inch Circuit	:	3,500 litres
2 Inch Circuit	:	2,650 litres
Drum Circuit	:	210 litres
Laboratory Circuit	:	5 litres
		10 litres
		30 litres

4. DATA RECORDING

The following summarises determinations and measurements recorded throughout the test programme:

4.1 LIQUOR CHEMISTRY

Due to the extremely low grade of the target solution, copper recovery has been assessed on a continual basis through liquor assaying, rather than product weighing. Potential inaccuracies in recovery and weighing (of only 300 grams of copper in the main test circuits) were thereby avoided.

Periodic samples were assayed as follows:

Cu	:	AAS
Total Fe	:	AAS
Fe^{++}(as ppm Fe)	:	Titration
Al	:	AAS
pH	:	Portable meter

The Mt Lyell waste stream does contain other contaminants. For the purpose of this programme, however, the target metal has been copper and the above elements have the greatest potential effect on process performance.

Iron has been assayed continuously, being the element in this context which would be most expected to affect copper electrowin efficiency. Much of the following discussions of process performance covers the observed effect of high iron concentration in the waste stream. All samples were assayed for total iron, by AAS, selected ones for ferric:ferrous iron ratio.

A standard titration method was utilised for ferric iron determination:

1. 10 ml aliquot, diluted by 10 ml mixed H_2SO_4/H_3PO_4 and approximately 100 ml distilled water.

2. Indicator - 6 drops sodium diphenylamine solution (0.2% in water).

3. Titration with 0.1N $K_2Cr_2O_7$ solution

The results of ferric iron determinations differed considerably between initial laboratory tests on the Gold Coast and the work on site. Comparative analysis of the data suggests that this has arisen from slight differences in indicator/solution preparation at the two sites. Both sets of assays, however, allow qualitative assessment of the change in iron valency over the period of testing and are valid for predictive purposes.

Iron levels in the liquor have been further used to provide an indication of likely contamination of resulting copper product (any drop in concentration indicating product contamination).

4.2 OPERATING PARAMETERS

Detailed records have been kept of operating parameters throughout individual tests - including:

Amps : As an analogue measurement from the DC rectifiers and checked periodically on individual cells with current tongs (the latter being accurate to 1 amp).

Volts : Measured and recorded for each individual cell at each specified sample interval.

Flow : Rotameter reading at specified sample intervals.

4.3 OTHER

Profiles of volts vs. amps have been recorded for each of the cell sizes used, at various stages in the test treatment cycle. These data are of particular use in modelling optimum cell dimensions for a possible treatment plant - providing as they do quantitative plots of operating voltage for differing hardware configuration.

A variety of pressure tests have been performed during the programme, from which detailed readings of cell pressure drop (at varying flow rate) have been recorded. Data recorded on 'standard' and modified cells result in later prediction of pumping costs in an overall treatment circuit.

The nature of copper product has been assessed through physical examination of cathodes at the end of each test run. All powder product has been collected for future assay and size determinations. Selected material was assayed for copper on site at Mt Lyell

4.4 DATA PRESENTATION

Where appropriate, all of the data collected have been recorded in computer spreadsheet form - to allow automatic calculation of process performance variables, such as current efficiency and power consumption vs. copper recovery (both periodic and cumulative). Tables and graphs presented below are generated from this data base.

Information tables relating to the initial laboratory testing are provided in the initial report for this project. Those relating to the site investigation are presented here as Appendix 1.

5.0 DISCUSSION OF RESULTS

The purpose in this programme has been the generation of a treatment circuit capable of technically and commercially extracting copper from the Mt Lyell waste streams, utilising Euralba's novel cell design as an integral component.

As will be shown, through detailed discussion of results to date, the study has demonstrated that two separate treatment routes are technically feasible:

a) By direct electrowinning through a series of banks of cells, or

b) Through an initial upgrading step (utilising a solvent extraction circuit specially modified for this process) following by treatment through a reduced number of cells.

Chemical and process characteristics which support these observations are individually analysed below, both treatment routes being examined in the context of:

SOLUTION CHEMISTRY	-	Copper concentration Iron concentration and valency Acidity
CURRENT EFFICIENCY	-	Effects of changing chemistry Effects of changing operating conditions
VOLTAGE	-	Effects chemistry and physical conditions
POWER CONSUMPTION	-	Analysis of data and generation of predicted performance requirements
PUMPING	-	Process performance and predicted treatment costs
PRODUCT	-	Expected nature and quality

Each of the relevant sections deals initially with the result of direct treatment of the waste solution; followed by analysis of performance on 'high grade' (solvent extraction style) liquor.

A subsequent section of this report provides a preliminary analysis of project economics, upon which the eventual choice of a treatment route must be based.

Individual test results are presented as a series of tables in Appendix 1 herewith. Graphic representations of these data are utilised in support of the text below.

5.1 SOLUTION CHEMISTRY

The Mount Lyell liquor is characterised by chemical constituents and concentrations which inherently inhibit high copper electrowin efficiency. The effect of chemical composition has been well defined in the test programme, as has the effect of the chemical changes that occur during progressive electrolysis.

Relatively detailed analysis of these changes, particularly with respect to iron chemistry, is required to place observations on related performance characteristics into context. It has considerable further relevance to the choice of an optimum treatment route for this site.

5.1.1 Direct Electrowin

Critical characteristics from the point of view of direct electrowinning have been defined to be:

A: LOW COPPER CONCENTRATION

Concentration of the target metal in the Mount Lyell waste stream lies in a range far lower than would be normally contemplated in an electrowin circuit.

The effect of low tenor is progressively exacerbated as copper is depleted to the very low grades that would be the target of a treatment programme.

The tests undertaken have successfully achieved their principal aim - in depleting copper in the waste solution to levels around a few parts per million.

Liquors tested have assayed between 90 and 170 ppm copper initial tenor (or approximately 300 times less than that in a conventional SX/EW tankhouse). These grades are in concordance with the mean value of 130 ppm reported by the operators for the main waste dump stream (one of the principal targets for an initial treatment circuit).

Simplistically, efficient copper electrowinning relies on effective supply of metal ions to the cathode surface - a feature which is largely governed by concentration of the metal in the target liquor. If 'supply' of ions is not maintained at a level required by applied current, power is 'wasted' in plating other metals or in hydrolysing water.

Extremely 'inefficient' recovery would generally be expected for solutions of these tenors.

However, previous testing of the new cell on liquors of similar grade has in fact demonstrated high achievable current efficiency - up to 75% on commencement of liquor treatment (down to 20% between 170 and 18.8 ppm Cu).

The Mt Lyell series of tests, although resulting in a potentially economic route for direct electrowinning, has shown significantly lower efficiencies than may be expected for such an ideal solution.

The results suggest therefore that, although reduced copper concentration is a large contributor to low current efficiency, the effect is exacerbated by other electrochemical characteristics of the solution.

The most significant of these is considered to be the high content and unfavorable valency state of iron in the solution.

B: IRON CONCENTRATION AND VALENCY

Concentrations of up to 2.4 g/l iron have been encountered in the Mt Lyell liquor - with initial Fe:Cu ratio of up to 14:1.

High iron concentration has a known deleterious effect on the electrowinning of copper. This is generally accepted to be the result of a cyclical redox reaction in the cell (oxidation at the anode followed by reduction at the cathode) which results in 'wastage' of power.

The current programme has demonstrated that the valency state of the iron has an equally significant effect on process performance. Nearly all of the iron initially present is in the form of ferric sulphate, an effective oxidising agent capable of dissolving metallic copper.

The 'iron effect' in the Mt Lyell liquor is therefore twofold -

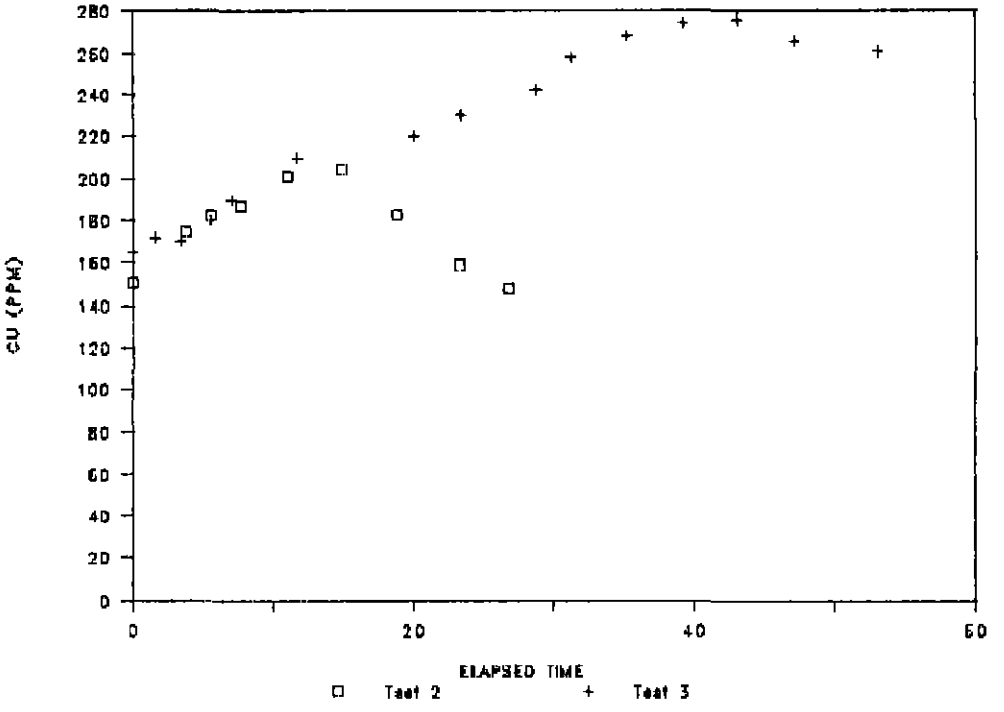
a) Power is 'wasted' through the cyclic redox reaction, and

b) The capacity of the solution to 'redissolve' copper that has already been electrowon has an obvious and significant effect on apparent process efficiency.

Given certain operating conditions (coupled with higher than normal iron concentration) the net effect on commencement of treatment of this liquor can in fact be an increase in copper concentration.

Potentially 'negative' recovery is illustrated by Figure 2, a graph showing copper recovery from two liquors of significantly different iron concentration (Test 2: 1400ppm, Test 3: 2400ppm) treated under the same operating conditions. The effect is attributed to copper leaching of the initial cathode surface by ferric sulphate.

FIGURE 2 : COPPER 'RECOVERY' PROFILES
TESTS ML 2, ML 3



A number of formal tests (ML8,9,11F,13F,12) were run without electrowin power, in an attempt to quantify the ferric iron effect. These demonstrated that the untreated waste solution readily dissolves copper:

$$Cu + Fe_2(SO_4)_3 = CuSO_4 + 2FeSO_4$$

Ferric sulphate — Ferrous sulphate

The reaction results in the production of ferrous sulphate, which does not have the oxidising power of ferric sulphate - itself a progressive beneficial effect.

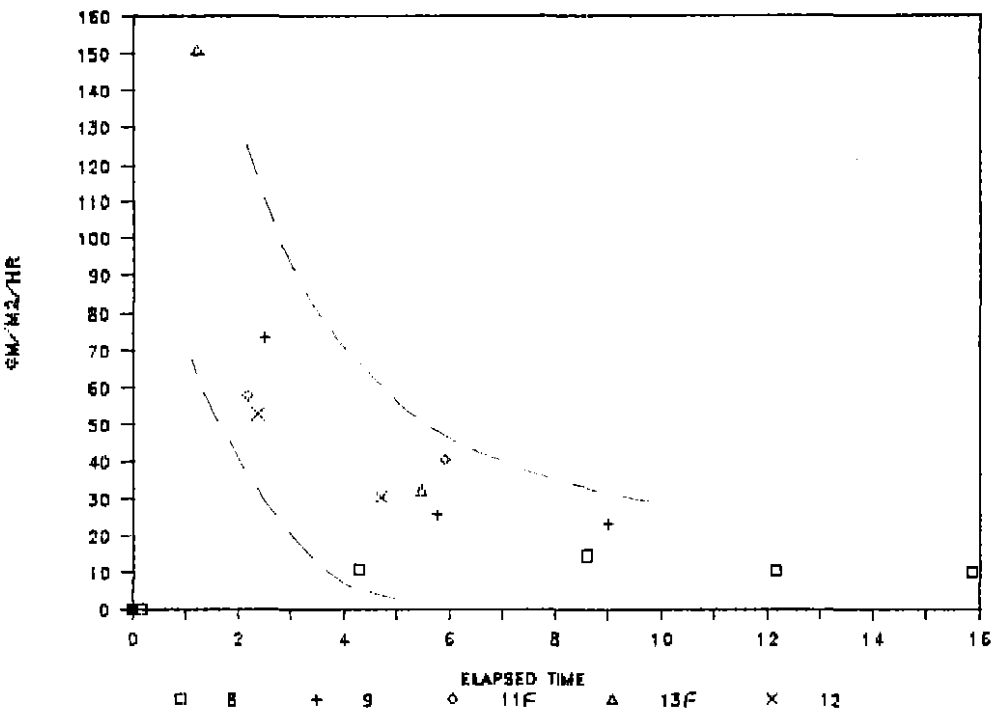
Re-oxidation to ferric state does occur in air, but rate of the reaction in a slightly acid solution is low. As the intention is to eventually treat the solution on a one-pass basis, the potential benefit achieved can be regarded as permanent.

Figure 3 traces the copper dissolution effect in this test series. As it results from contact between the solution and the copper cathodes, its progressive influence is assessed in terms of:

Copper (gms) dissolved,per square metre of cathode per hour,on a periodic basis through the treatment cycle.

This mode of interpretation allows a direct comparison of the effect for the different circuits used during the programme.

FIGURE 3 : COPPER DISSOLUTION
TESTS ML 8,9,11F,13F,12



This graph clearly shows a significant drop in 'leaching ability' of the solution with time, which occurs as progressive reduction of ferric iron is achieved (through the dissolution of copper).

Plots of all of these tests lie within a relatively well defined 'corridor', with dissolution commencing at high rate (between 60 and 150 gms/m^2/hour), falling rapidly and 'settling' at a rate between 25 and 10 gm/m^2/hour.

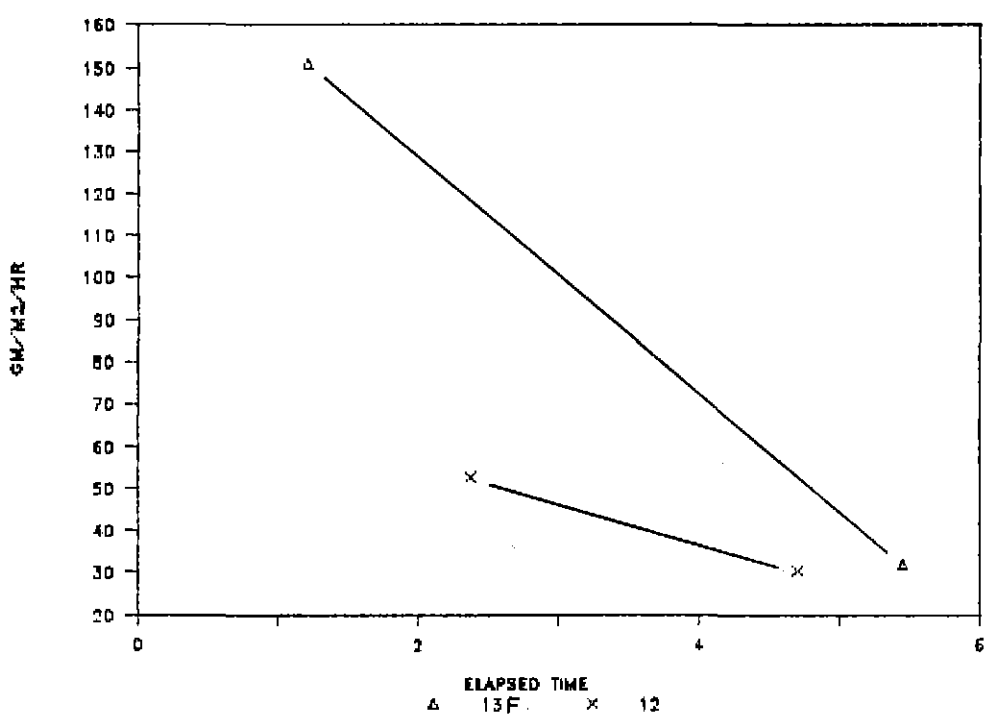
Insufficient data are available to exactly quantify the relationship between rate of dissolution and progressively changing ferrous iron concentration.

Calculation of theoretical values (stochiometrically against the above equation) for the individual tests suggests, however, that dissolution rates of below 40 gm/m²/hour are reached after conversion of only 20% of the contained ferric iron to ferrous sulphate (200-300 ppm as Fe++ as opposed to total Fe content of 1500-1700 ppm).

This suggests that the beneficial effect arising from the valency change can become significant, from a process point of view, even after only limited conversion.

Figure 4 illustrates the effect of differing initial ferrous iron concentration on copper dissolution rate (no other operating parameters being changed). Test 13F utilised 'fresh' liquor from the waste stream, Test 12 'previously electrowon' liquor (i.e in which a significant initial ferrous iron concentration was present).

FIGURE 4 : COMPARISON COPPER DISSOLUTION RATES
DIFFERING INITIAL FE++ CONC. ML12,13F



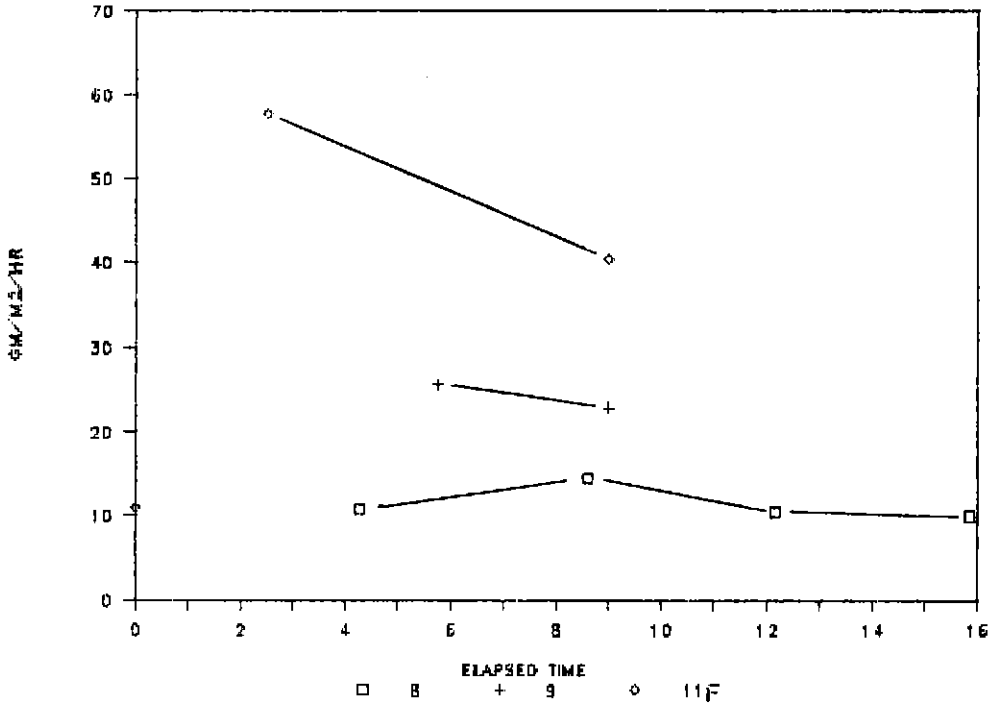
The plot clearly demonstrates reduction in copper leaching which with progressive increase in ferrous iron concentration. As will be discussed in more detail below, subsequent electrowinning from the solution from Test 12 achieved initial current efficiencies more than double that of most of the tests on fresh solution.

Iron valency has been found to have significant effect in testing of liquors from elsewhere, investigations at even higher concentration resulting in current efficiency in the region of 50% (at copper tenor similar to that at Mt Lyell).

Certain operating parameters integral to the new process have been established by this programme to affect the physical rate of dissolution of copper (in addition to their effect on the rate of chemical conversion of iron).

Figure 5 illustrates the effect of flow rate on this factor. Tests 8 and 11F, both on the same circuit, utilised liquor of similar chemical composition. Flow rate for the former was 1430 l/hour, the latter 1800 l/hour. Test 9 operated on a smaller sample volume (Drum Circuit) at even greater flow rate (3130 l/hour) (Tests 8, 9 and 11F).

FIGURE 5 : COMPARISON COPPER DISSOLUTION RATES
DIFFERING FLOW REGIME - ML8,9,11F

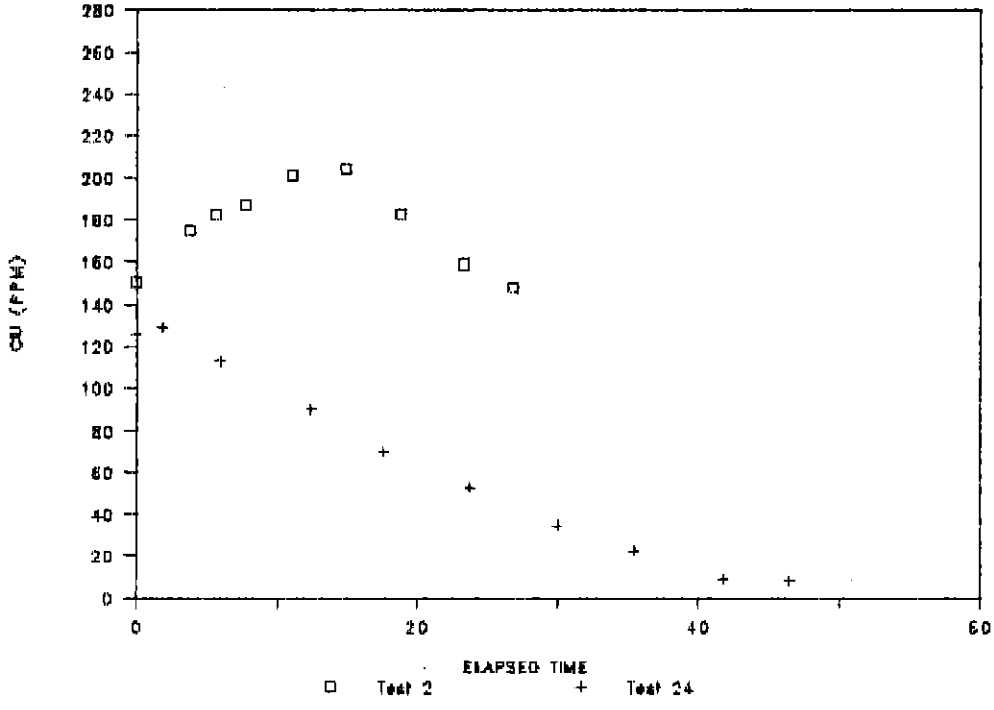


The plots for Tests 8 and 11FA illustrate a marked difference; reduction in flow by 20% resulting in a threefold reduction in copper dissolution rate.

The profile exhibited for Test 9 is a reflection of combining high flow rate (causing high initial dissolution) and limited sample volume. Faster circulation of the sample as a whole results in more rapid achievement of the levels of ferrous iron required to minimise copper dissolution - thus the more rapid drop-off than Test 11.

Electrowin tests on samples of similar chemistry (Fig 6) suggest that, under operating conditions, increasing current density does not result in an increased copper dissolution rate. Significant difference between the copper profiles for tests 2 and 24 (as all other conditions remained constant) reflects a marked decrease in the ratio of copper dissolved to that electrowon (i.e an increase in the latter, with constant or even decrease in the former).

FIGURE 6 : COPPER RECOVERY PROFILES
DIFFERING CURRENT DENSITY - ML2,24



Test 24, at similar grade to 2 and 3, resulted in positive copper recovery from commencement of treatment, in response to a 50% increase in current density. In effect, the higher current in Test 24 simply results in a 'net' balance between electrowon copper and that leached by ferric sulphate.

There are obvious potential operating benefits therefore in the initial treatment of this liquor at low flow. Relatively high current density at the 'front end' of a circuit will allow benefit to be taken of a further iron reduction mechanism, without a penalty in the form of increased copper dissolution.

In addition to the conversion achieved through oxidation and solution of copper, electrowinning of a Cu/Fe sulphate solution itself has a net reducing effect on the iron.

Whether this 'electro-reduction' is controlled by mass transfer of ions to the electrode surfaces, or is influenced by electrode current density, is not known at this stage. In either case, it is possible that the morphology of the new cell (with differing electrode area) will promote the effect.

Insufficient data have been collected here to quantify the respective proportions of iron reduction achieved (copper dissolution and/or electrolysis) under actual treatment conditions. Some general observations are, however, possible.

Being an electrochemical phenomenon, it can be reasonably assumed that lifting applied current will increase the rate of electrolytic iron reduction.

Figure 7 is a plot over time for two laboratory tests (Note: prefixed MDL) performed on the same liquor. It clearly demonstrates increase in ferrous iron formation in response to higher initial current density (Test 21: 70 A/m^2, Test 15: 140 A/m^2). As it is unlikely that copper dissolution rate has lifted (as suggested above) the increase is attributed directly to electrolytic reduction.

In the Mt Lyell context, current density can therefore be used to enhance process performance, through increasing the rate of ferric iron reduction (consequently reducing the 'leaching ability' of the circulating solution).

There is some evidence to suggest that size and configuration of the cells may have an effect on iron chemistry. Figures 8 and 9 plot the production of ferrous iron over time for selected tests on differing circuits. They examine production in terms of 'grams per amp hour' (in response to the electrolytic reduction effect) and grams per square metre of cathode per hour (the chosen method for copper leaching analysis).

FIGURE 7 : FERROUS IRON CONCENTRATION
DIFFERING CURRENT DENSITY - MDL 15,21

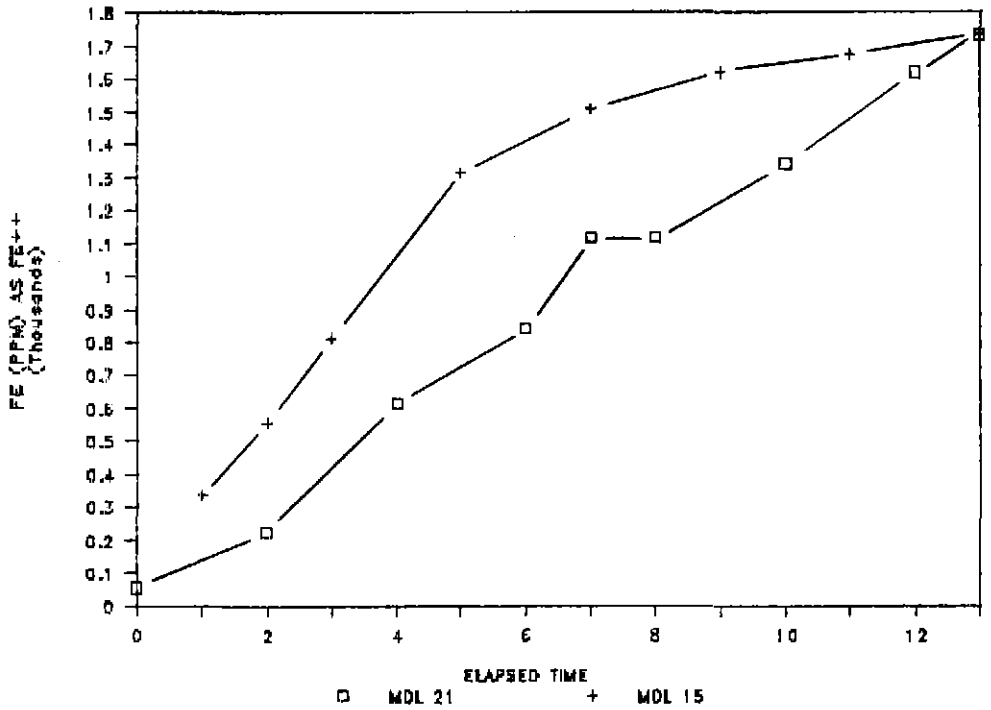


FIGURE 8 : FERROUS IRON CONVERSION (gm/A/hr)
DIFFERING CELL SIZES - MDL19,ML19,20

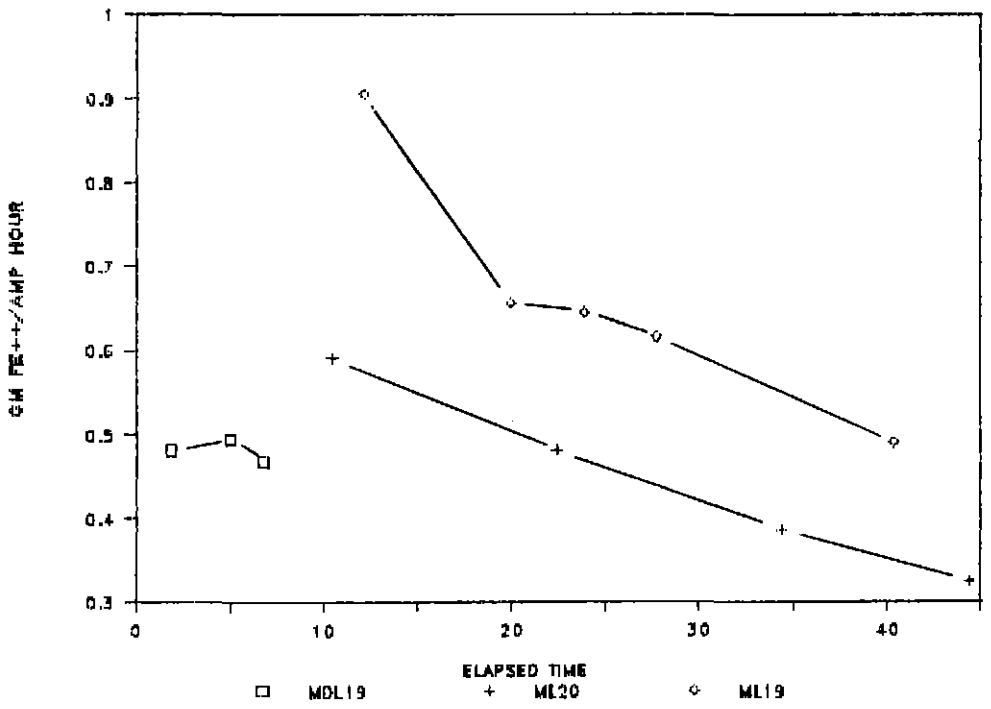
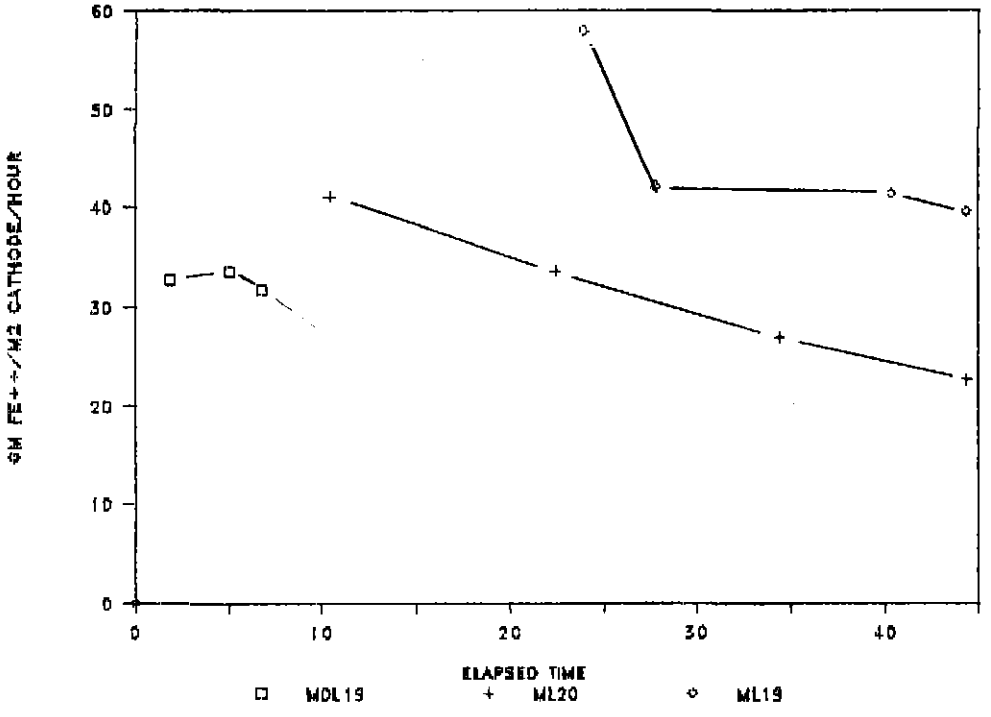


FIGURE 9 : FERROUS IRON CONVERSION (gm/m2/hr)
DIFFERING CURRENT DENSITY - MDL19,ML19,20



Both of the latter graphs show a higher conversion rate for the larger diameter cells, suggesting that this unit should be chosen for the initial section of a model treatment circuit. The tests in question were, however, conducted at differing flow rate, which (in affecting copper leaching rate) may be the cause of the variations recorded.

The results obtained clearly show process improvement when appreciable 'conversion' of ferric to ferrous sulphate has occurred (e.g. Test ML13). The above discussion has shown potential for enhancement of this through operating parameter changes, as a result of:

a) The effect that flow has on copper dissolution rates and consequent apparent electrowin efficiency.

b) The effect that the ratio of ferrous to ferric iron has on the ability of the circulating solution to dissolve copper.

c) The effect that operating current density has on overall reduction of iron.

d) The possibility that sizing of cells can be utilised to increase initial iron conversion.

There are several recognised mechanisms, outside of those operating in the circuit, through which iron valency conversion can also be promoted. Most entail contacting the solution with a reducing agent, such as:

Iron : $Fe_2(SO_4)_3 + Fe = 3FeSO_4$

This route is not feasible as a further reaction will result in 'cementation' of copper from the liquor:

$CuSO_4 + Fe = Cu + FeSO_4$

Sulphide : Contact with sulphide minerals e.g:

$Fe_2(SO_4)_3 + FeS_2 + 3O_2 + 2H_2O = 2H_2SO_4 + 3FeSO_4$

Although sulphide concentrates are available to achieve this reaction in the short term, the addition of iron and acid to the solution may not be acceptable from an environmental point of view.

It is worth noting that similar reactions to this result in the original dissolution of copper from the waste ore. The ferrous iron so produced is, however, oxidised during its travel through the heaps.

Hydrogen : In the presence of a catalyst, gaseous hydrogen will reduce ferric iron:

$2Fe^{3+} + H_2 = 2Fe^{2+} + 2H^+$

It is conceivable that hydrogen produced during electrolysis could be utilised to perform this reduction. The results of limited testing on site, using a platinum basket, were inconclusive. The route may however deserve future consideration.

Other methods such as agitation with carbon could be considered.

All however would result in costly additions to a treatment circuit at Mt Lyell, or even addition of further contaminants to the drainage.

It is likely that the benefits of iron reduction can only be considered in context of the reactions inherent to the electrowin process. Maximum advantage should be taken of those reactions, through manipulation of operating parameters where possible.

The influence of iron in the Mt lyell waste stream can therefore be summarised as follows:

a) Cyclic redox reactions in the liquor result in power 'wastage' and low apparent copper electrowinning efficiency.

b) Ferric sulphate in the waste stream redissolves previously electrowon copper. This effect is progressively reduced during treatment as ferrous sulphate is produced by reduction of the ferric iron.

c) Both a) and b) above result in progressive conversion of ferric sulphate to ferrous sulphate.

d) A 'net' reducing effect is suggested in the cyclic iron redox reaction during electrolysis, whose rate is related to applied current density.

Although the adverse effects of iron in the liquor cannot be obviated without its complete removal, the circuit can be designed to make maximum benefit of the features described above. As the changes in chemistry are progressive, 'tailoring' in hardware and operating parameters will be more effective at the 'front end' of the treatment system.

The initial stages of a model circuit for Mt Lyell would therefore include:

* Large diameter cells
* Low flow conditions
* Higher than 'normal' initial current density

Finally, the use of stainless steel rather than copper cathodes on the commencement of treatment will reduce the 'availability' of copper for immediate dissolution, and prevent any likelihood of damage to the cathode 'wall'.

C: 'ACIDITY'

pH of the Mt Lyell waste stream is relatively high, but reduces progressively - as ferric iron is converted to ferrous and 'free' acid is liberated through the electrowinning of copper.

'Acidity' of the solution governs its resistance and is a contributor, through its effect on cell voltage, to overall power cost of copper production. The work conducted has shown that relatively low current densities must be maintained in the process, to bring cell voltage into a viable range.

'Manipulation' of pH in this context is not possible, beyond the selection of operating parameters that may maximise conversion of ferric to ferrous iron. The effect of high resistivity may however, as will be discussed below, be somewhat obviated through changing electrode gap in the cells.

5.1.2 Liquor Upgrade Option

The adverse effect of the liquor's adverse chemical features would be obviated, if a liquor 'upgrading' step were incorporated into the treatment circuit.

Samples of the waste liquor were submitted to Henkel Australia, in Melbourne, to determine the technical feasibility of achieving this upgrade through a simplified solvent extraction circuit.

Appendix 2 herewith, provides the isotherm point calculations derived from this study. In effect, the work concludes that the upgrading step is feasible, utilising established methods and reagents. It would result in a liquor of the following characteristics:

Copper grade	:	50 g/l
Fe concentration	:	less than 100 ppm
Acid concentration	:	180 g/l (in spent electrolyte)

Expected recovery, for a 2 extraction/1 strip circuit, is projected at greater than 90%.

A series of tests (Tests ML4,5,10,15,22) were conducted during the site programme, on a liquor approaching this composition. In fabricating the solution, samples were purposely contaminated with other metals in the Mt Lyell waste (through addition of a proportion of untreated waste solution).

Although details of process enhancement achieved will be discussed further below, the potential advantage in this treatment step can be generalised as follows:

a) Current efficiency in copper electrowinning will approach 100% within the range of copper values to be treated (50-1 g/l Cu)

b) Cell voltage will be significantly less, which in conjunction with a) above will result in significant reduction in copper production cost.

c) The combination of a) above, and a consequent reduction in liquor volume, will result in a smaller, simplified and significantly cheaper electrowin circuit.

It has further become apparent, through testing of the technology on 'high' grade liquors and through discussions with proponents of the process, that performance of solvent extraction itself could be significantly improved, when operated in conjunction with the new cell.

The advantages that arise are in general the consequence of:

a) The wide range of copper values over which the cell maintains high efficiency.

b) The ability for the cell to operate at higher than conventional current densities.

c) The fact that the electrowin circuit is essential a 'closed' system.

5.2 CURRENT EFFICIENCY

Cost of copper production in any electrowin system is a function of:

a) Applied current

b) Cell voltage

c) Percentage copper recovery against theoretical yield for the applied current

As already noted, achievable efficiency is largely governed by solution chemistry. The relationship between 'supply' (liquor tenor) and 'demand' (current density) for copper; and the concentration of competing metals (such as iron) are the most important factors.

The cell under investigation favourably promotes the former, allowing high efficiency down to very low grades; but remains significantly affected by the latter (although some improvements over conventional cells are mooted).

Direct electrowinning of copper from the Mt Lyell liquor has been shown to result in significantly lower current efficiencies than would be possible on a normal 'tankhouse' liquor. However, it has also been demonstrated that:

a) Achievable efficiencies are significantly greater than would be possible in a conventional cell.

b) Efficiency, although low, could support a viable direct treatment route (power cost being the single largest cost to such an operation).

The programme has further demonstrated that an alternative treatment route (through solvent extraction followed by electrowinning in the new cell) would result far higher current efficiency (95%+). The potential viability of such an approach is supported by one of the major characteristics of the new system - its significantly lower capital cost than conventional systems.

The following discussion provides an analysis of achievable current efficiency, over a relatively wide range of chemical and operating conditions.

The data have been compiled to allow rapid visual comparison between tests. Simple plots of copper depletion over time serve, in most cases, to illustrate changes in current efficiency under differing conditions - copper recovery and progressive depletion being directly related to that factor.

5.2.1 Direct Electrowin

Initial laboratory testing, on small samples, suggested that cumulative efficiency, in winning copper down to a level of around 10ppm (original approximately 150ppm), would lie in the region of 7-9%.

Short intervals during which efficiency exceeded 20% suggested that, given optimum operating conditions, greater efficiency is possible.

The site programme has provided clearer definition of the capabilities of the process; and the degree to which hardware and operating conditions will govern current efficiency.

Detailed records of current efficiency achieved throughout the test runs, both for specified sample periods and on a cumulative basis, are provided in the individual test records supplied as Appendix 3 here.

Measurements of periodic and cumulative current efficiency are both important. Cumulative current efficiency to a target grade (10-20ppm Cu) is obviously an indicator of likely operating costs in a commercial treatment plant.

Periodic measurements, however, provide a clearer definition of the progressive effects of changing chemical conditions and operating parameters. The latter are used more specifically for modelling an idealised circuit for treatment of the liquor.

Table 1 below provides a summary of the current efficiencies achieved during testwork on the various circuits - under differing operating and chemical conditions:

TABLE 1 : Summary of Current Efficiency Measurements
Mount Lyell Field Programme

Test	Start Cu(ppm)	End Cu(ppm)	Current effic.		Flow l/hour	Current A/m^2
			Highest periodic	Cumul.		
ML6	91	16	15	10	1400	35
ML11	90	14	17	7	1400	52
ML18	130	18	17	8	1800	52
ML20	149	14	32	10	1800	69
ML25	135	15	12	9	2400	69
ML29	92	8	7	7	2400	104
ML33	119	9	13	8	3400	104
4" Circuit						
ML1	94	18	19	13	3800	42
ML2	150	-	23	-	3800	42
ML3	150	-	19	-	3800	42
ML7	155	-	24	-	4200	42
ML13	232	18	48	37	4380	42
ML19	126	15	16	12	4280	63
ML24	125	9	12	8	3800	63
'Drum' Circuit						
ML17	144	18	13	6	1700	52
ML21	138	16	12	7	1700	52
ML26	130	20	9	5	800	70
ML28	88	12	6	4	800	70

Note : Where no result indicated = test where initial dissolution effect resulted in initial negative recovery.

Although these results are covered in more detail below, several 'trends' are immediately apparent:

* In general, the 4 inch circuit was the best performer, current efficiency being approximately 30% improved over that of the 2 inch circuit under similar conditions.

  This is attributed to differing flow dynamics and possibly the higher flow rate used for the larger cells.

* At higher iron concentration (parallelling copper tenor in Table 1) low current density results in initially severe copper dissolution effects.

* Following reduction in copper redissolution, higher efficiency is generally a feature of lower current density.

* Discounting its effect on copper dissolution at the commencement of treatment, increased flow results in raised overall efficiency.

* Notably high current efficiency (up to 48%) can be achieved once a reasonable amount of contained iron is reduced (Test 13).

A complex relationship exists between the effects of separate operating conditions (both those imposed through liquor chemistry and those controllable in the process) on current density. It is only through detailed analysis of these effects that a model circuit for treatment of the Mt Lyell liquor can be generated.

The progressive effect of continually changing chemistry during treatment is an obvious feature of all of the testwork. The results demonstrate that copper recovery can be generalised into three stages with respect to recorded current efficiency levels:

Stage 1

Characterised by : High Fe^{+++}, potentially high copper solution rate.

Examination of copper concentration profiles shows 'negative' recovery for the initial stages of a number of tests (i.e negative current efficiency), characteristically a response to high iron concentration and rapid flow, in conjunction with low current density.

This phenomenon results from an imbalance between copper dissolution and electrowinning. It can be immediately overcome through increasing current density (an operating parameter that has been shown to have little effect on the rate of copper leaching), but it naturally has a significant effect on overall current efficiency.

Stage 2

Characterised by : Moderate Fe conversion has slowed copper dissolution rate.

An apparent second stage of recovery occurs when a certain proportion of iron has been reduced to ferrous sulphate. General analysis of the testwork suggests that, almost regardless of initial grade, this 'trigger' point occurs at around 25% conversion.

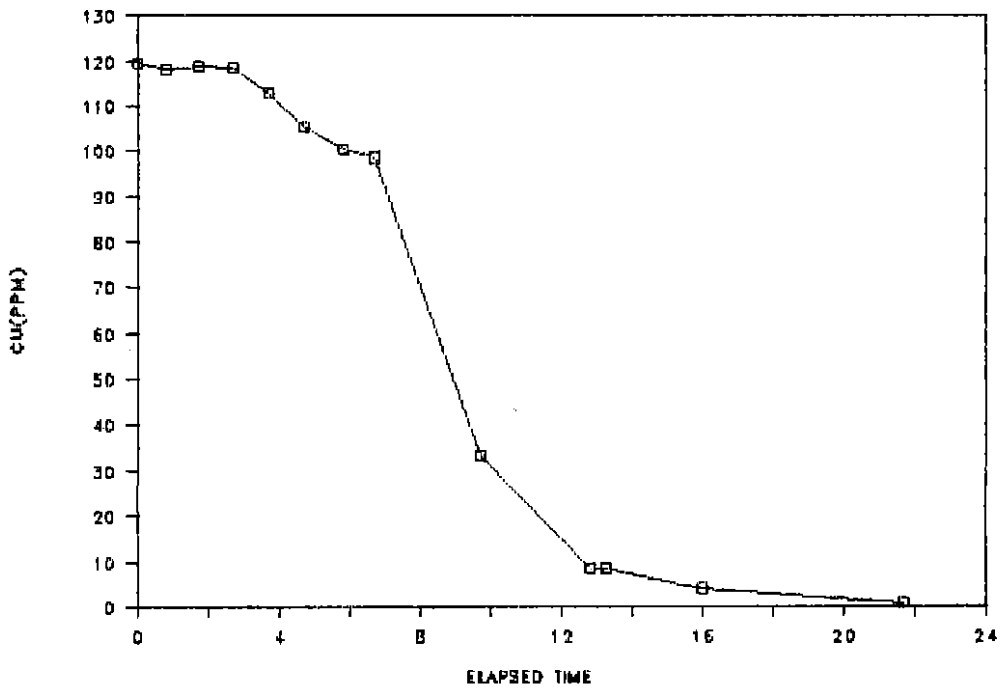
Copper recovery becomes a balance between:

a) reduction in current efficiency as copper tenor falls off, and

b) the continued lessening of the leaching power of the solution.

The effect in general is that, whereas progressively lower performance can be expected as copper tenor drops, relatively consistent current efficiency is maintained to very low copper grades - i.e straight line recovery (indicative of consistent current efficiency).

FIGURE 10 : MT LYELL ELECTROWIN CYCLE
Cu Profile - Staged Process Performance



Stage 3

Characterised by : Minimal Cu tenor

Stage 3 occurs at copper grades between 10 and 20 ppm, where the apparent benefit of the above can no longer be maintained. Current efficiency drops off rapidly to very low levels.

Copper recovery below 10 to 20 ppm would most likely not be contemplated in treatment of this liquor.

Although, changes in chemistry and the effects thereof are continuous this division is useful in determining optimum operating parameters for a practical treatment plant.

The impact of individual parameters on current efficiency has been examined in some detail. The detailed compilation of data undertaken allows the following observations to be made.

a) Effect of iron concentration and valency

The effects of the high iron concentration in this liquor have been discussed in some detail above. Its direct impact on calculated current efficiency lies in:

a) The 'loss' of copper to leaching, which obviously leads to a higher apparent power consumption per unit actually recovered.

b) The 'wastage' of power that occurs through the cyclic redox reaction.

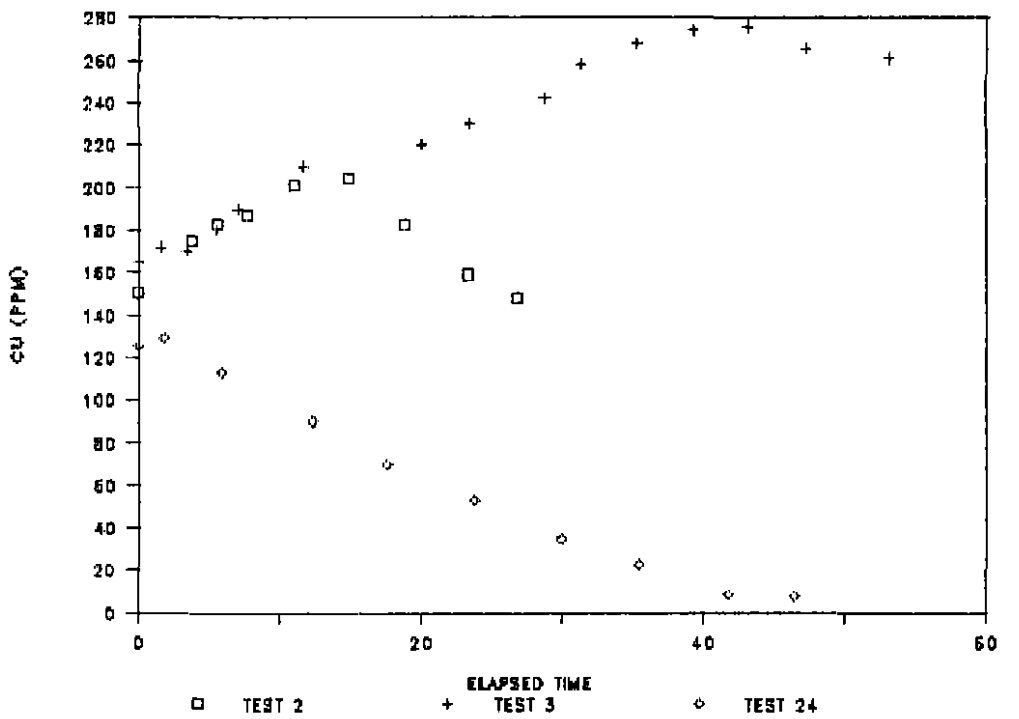
Its specific impact on current efficiency has been examined over a range of liquor concentrations, as illustrated by the following examples.

Figure 11 demonstrates the effect of increasing iron concentration in the initial stages of treatment. It is a plot of progressive copper concentration in liquor, for principally two tests operated under identical conditions (4" circuit). Solution for the two runs was of similar copper grade (90 ppm) but differing iron concentration (Test 2: 1410 ppm, Test 3: 2370 ppm).

Following a period of net dissolution of copper, Tests 2 and 3 achieved similar current efficiency levels to periodic electrowinning (of 22 and 19% respectively), indicating the performance levels possible once a proportion of the contained iron is reduced.

The period of negative recovery is prolonged by the increased iron concentration.

FIGURE 11 : FE VS CU PROFILE
VARYING FE CONCENTRATION - ML2,3,24



The graph included in Figure 11 for Test 24 illustrates the immediate effect of increasing current density on a similar liquor to Test 2. Test 24 achieved net copper recovery after only a very short interval, at current efficiency of 10-12% (as opposed to Test 2 which maintained negative copper balance for a considerable period of time).

Test 19, a similar run on the same circuit, again shows that negative recovery in stage 1 can be simply obviated - by lifting current density (changing the respective proportions of copper dissolved and electrowon). The result in this case was relatively consistent current density (8-16%) in depletion of copper from 126 ppm to 15ppm.

The ability of current density to obviate the negative recovery effect in stage 1 electrowinning will obviously be related to initial iron concentration. A comparison between tests 1 and 7 (Figure 13 : again same physical operating conditions but significantly different iron concentration (1430 as opposed to 730)) shows that for a given iron concentration a certain current density is required to maintain positive copper recovery from the commencement of treatment.

FIGURE 12 : FE VS CU PROFILE
VARYING CURRENT DENSITY - ML2,19

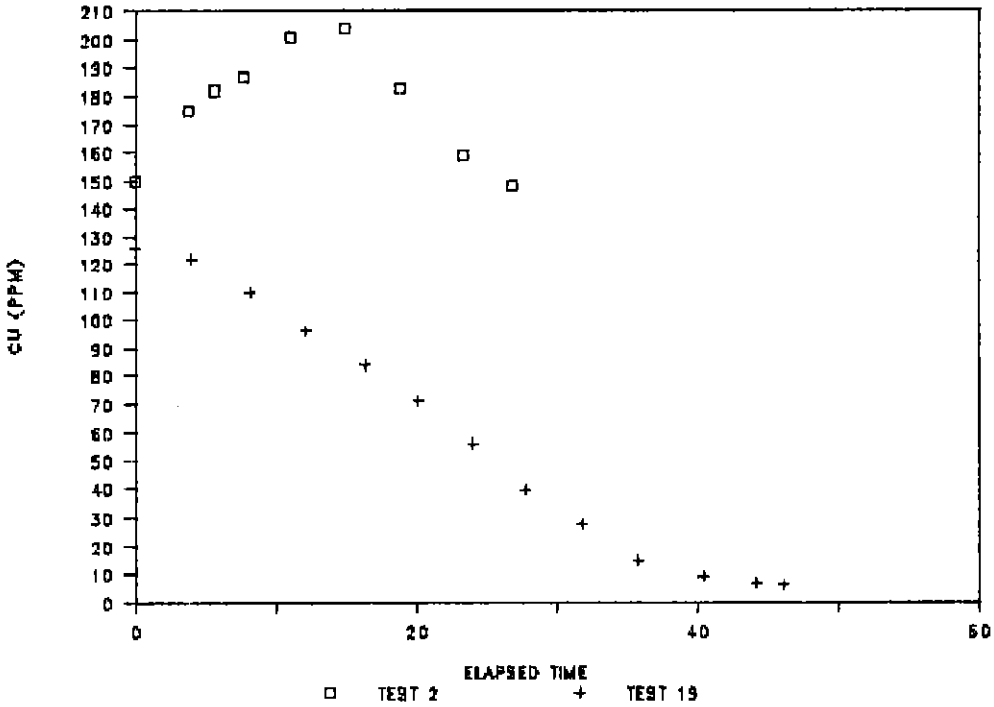
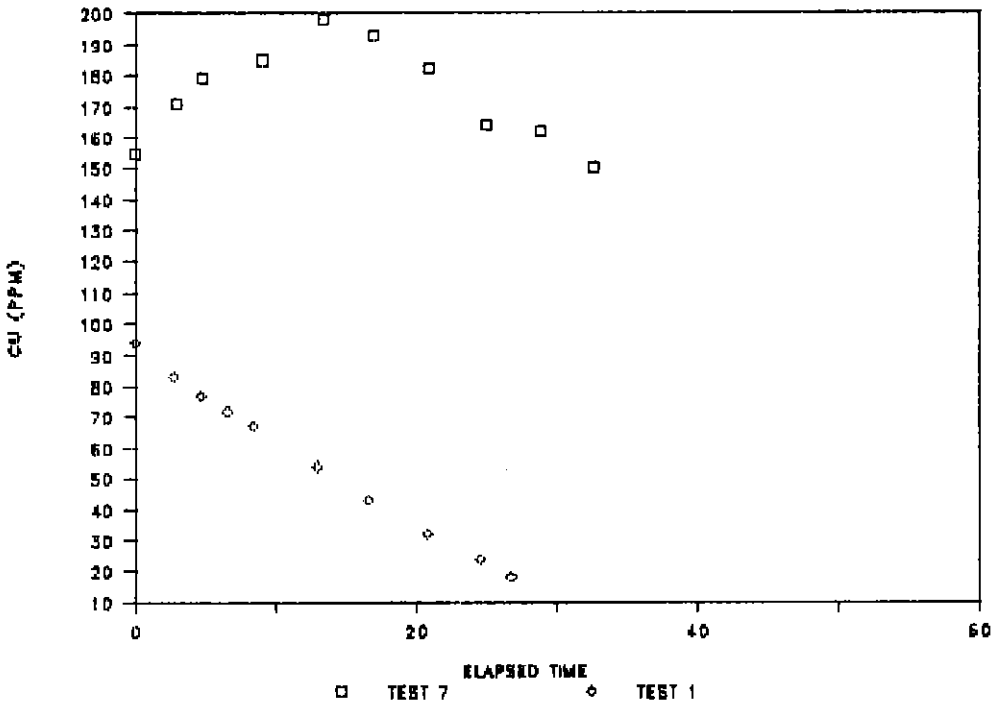


FIGURE 13 : FE VS CU PROFILE
VARYING CURRENT DENSITY - ML1,7



Although further data should be collected, it becomes possible to make predictions of the current density (and flow settings) required to maintain positive copper recovery for a given iron concentration.

Analysis of the results of the examples provided above, coupled with established copper leaching rates (section 5.1 above), can also provide an indication of the efficiency that could be achieved in Stage 1 electrowinning, without the ferric sulphate dissolution effect:

e.g. : Tests 2 and 24 after 5-6 hours treatment of same volume of similar chemistry liquor. Copper dissolution rates established in Test 13F, operated without electrowin on equivalent liquor volume.

Copper dissolved

Test 13F, 5-6 hrs = 378 grams

Copper recovered

Test 2, 5-6 hours = -78 grams
Test 24, 5-6 hours = +42 grams

Indicated 'Electrowon' copper

Test 2 = 300 grams
Test 24 = 420 grams

With five cells operating at the current density settings selected, a calculation on this basis would indicate achievable current efficiencies in excess of 90% for both tests on commencement of liquor treatment.

The calculation is obviously academic, but it does serve to substantiate the observation that changing current density does not increase copper dissolution rate; and to explain the marked difference in current efficiency achieved in Test 13 (a liquor which had undergone some iron conversion prior to electrowinning).

The effect of ferric to ferrous iron ratio is further emphasised by the results of a short series of tests run sequentially on a common liquor sample. Figures 14,15 and 16 here provide plots of two electrowin tests performed under the same operating conditions - on the same sample, but subjected to various 'treatment campaigns':

1. Preliminary electrowin test to assess operating performance under given conditions (Test 7)

2. Same liquor circulated through circuit without electrowin power - to assess copper dissolution rate given the iron reduction that occurred during first run (flush test 12).

3. Same liquor, retreated under same operating conditions as Test 7 - to assess current efficiencies achievable with some conversion of iron before treatment (Test 13).

Each of the steps preceding Test 13 resulted in progressive iron reduction, either due to copper dissolution (7 and 12) and/or electrolytic reduction (7).

As a result of the 'flush' in test 12 copper tenor in the liquor commenced at 230ppm as opposed to the original 156 ppm concentration for Test 7. As will discussed later initially higher copper tenor generally will result in an overall improvement in current efficiency, but Test 13 provides an ideal example of the efficiency that can be achieved on limited conversion of ferric to ferrous iron.

Figure 14 profiles copper concentration in the liquor over the period of both electrowin tests. Figures 15 and 16 detail the periodic current efficiency achieved by both over time and against progressively lower copper grade.

FIGURE 14 : COPPER RECOVERY PROFILE
TESTS 13,7

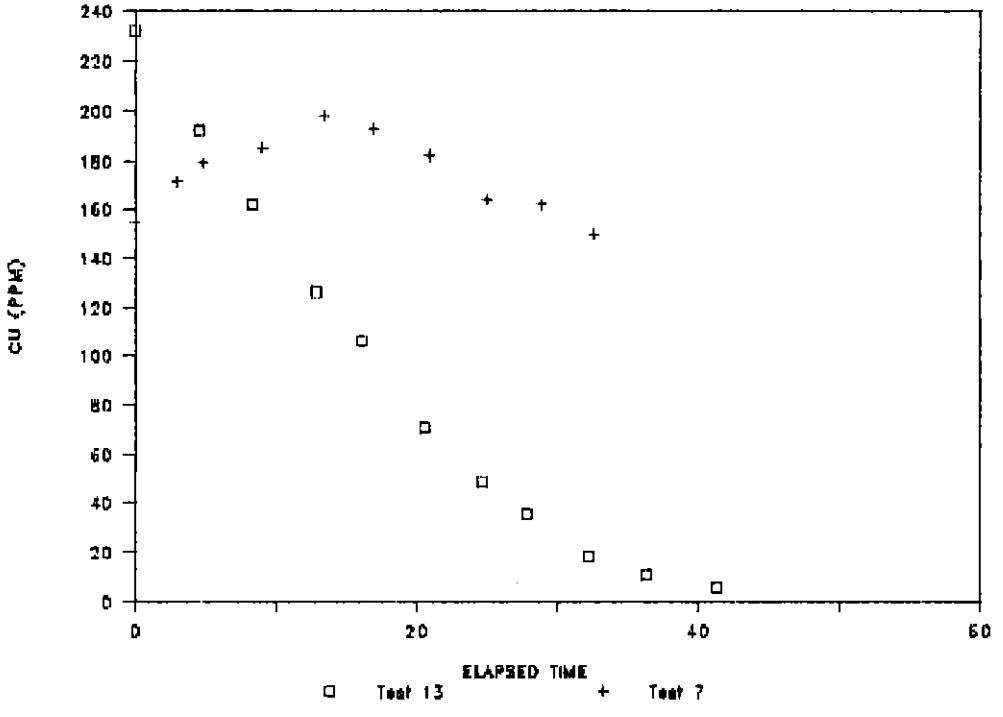


FIGURE 15 : COPPER PROFILE TEST 13
CU DEPLETION VS PERIODIC EFFICIENCY

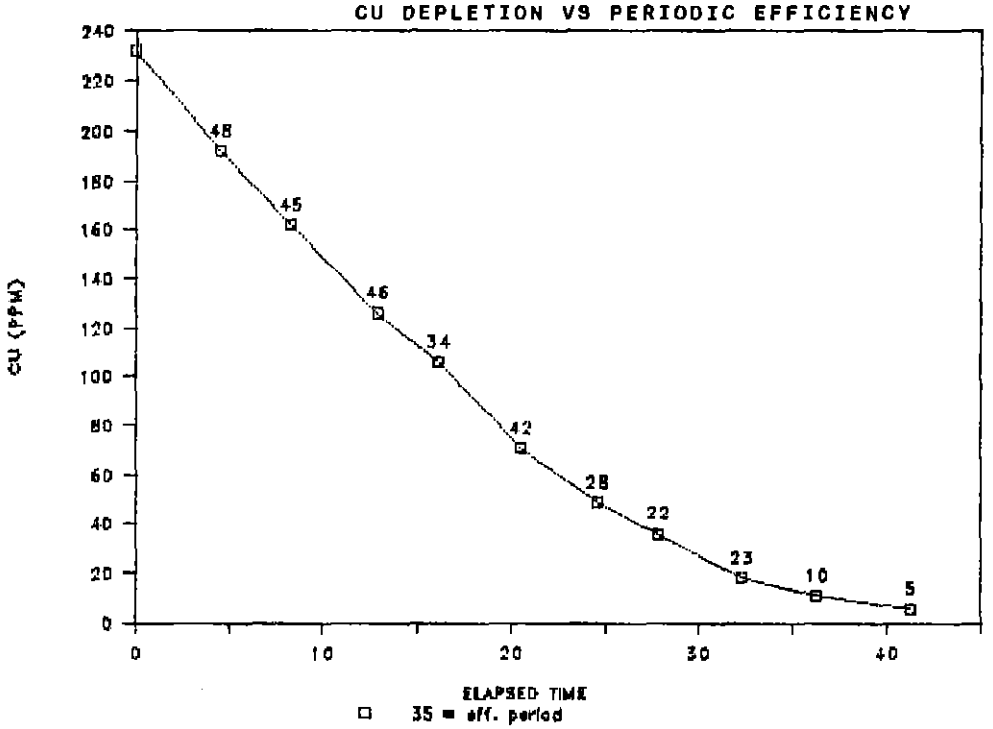
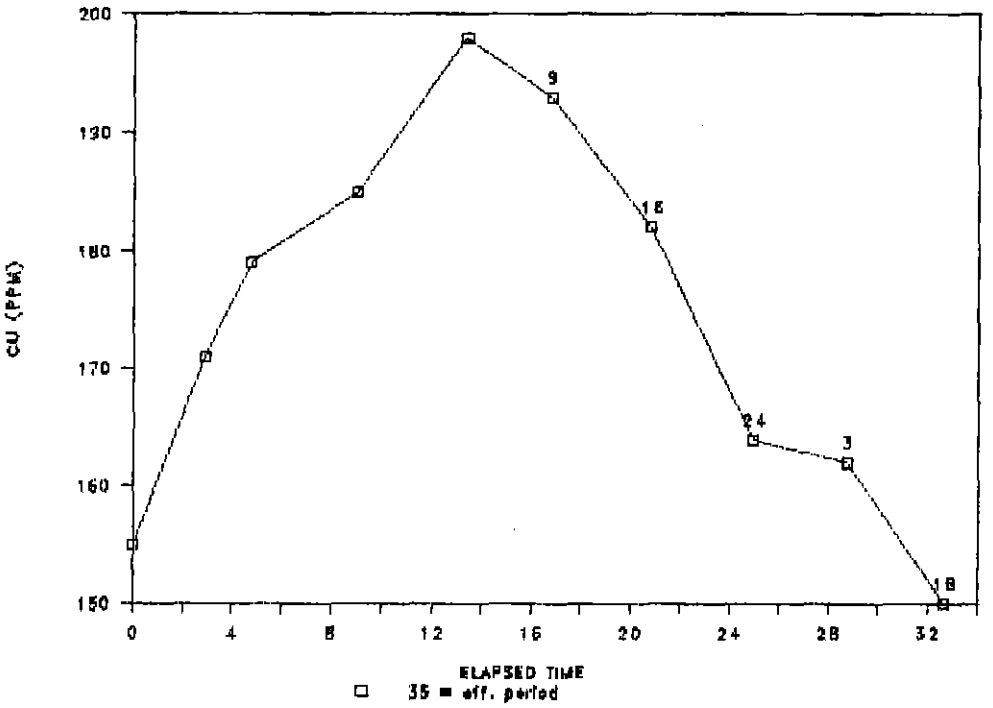


FIGURE 16 : COPPER PROFILE TEST 7
CU DEPLETION VS PERIODIC EFFICIENCY



Test 13 recording efficiency of 40+% in the copper range 240 to 70 ppm. Cumulative current efficiency was calculated at 37% down to a copper concentration of 18ppm.

On reaching of the iron 'trigger point' test 7 achieved current efficiency between 9 and 24% (dismissing one sample period at 3% efficiency as being the result of a rectifier trip) over a similar range of grade.

Assays conducted on site indicate an initial ferrous iron concentration for test 13 equivalent to approximately 15% of total iron.

b) Effect of copper grade on current efficiency

The true effect that copper grade has on current efficiency, within the relatively narrow range that will be encountered at Mt Lyell (90-160ppm) is somewhat masked, especially at the onset of electrowinning, by the effects outlined above.

Generally, it must be expected that progressively reducing grade will result in progressively lower current efficiency. It follows, therefore, that higher grade at commencement of treatment will result in a rise in overall cumulative current efficiency (as a result of 'averaging').

Grade related reduction in efficiency would be expected to be progressive, resulting in a gradually flattening slope for the recovery profiles plotted. As mentioned above, however, the lessening of other effects over time has the tendency to straighten these profiles during Stage 2 electrolysis (indicating consistent current efficiency).

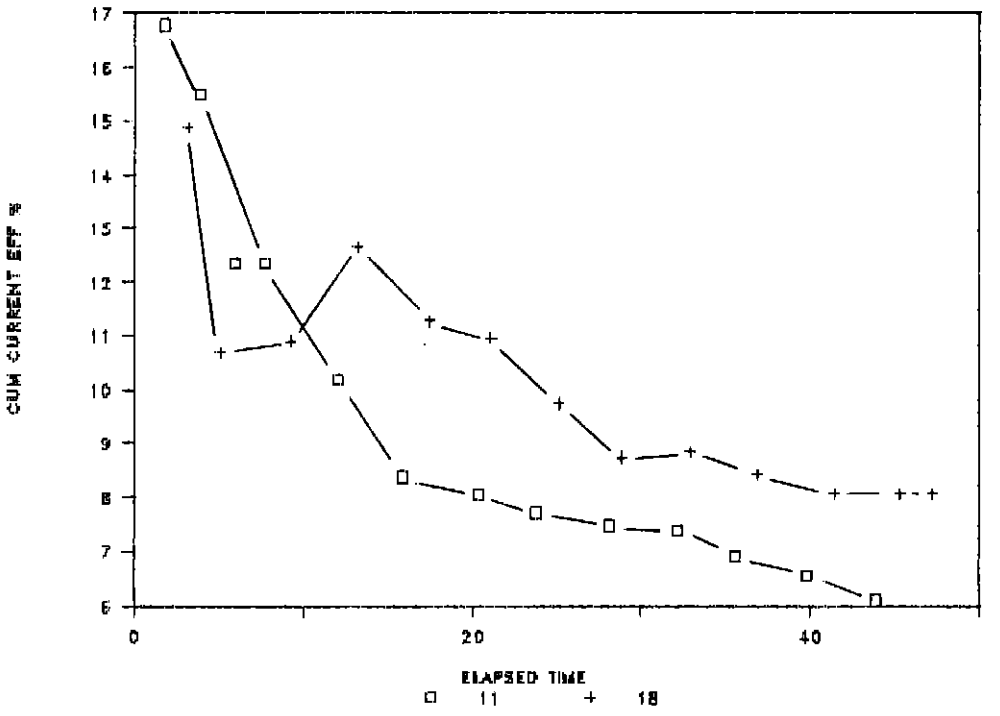
In other words, once copper dissolution by iron is reduced consistent efficiency can be expected in treating the liquor down to target grade.

Figure 17 plots cumulative current efficiency for two tests undertaken under similar operating conditions, but at different initial copper grade (Test 11: 90 ppm, Test 18: 130 ppm). On Stage 1 copper solution effects being overcome, the plots parallel each other, indicating similar current efficiency profiles. An increase of 1.5% cumulative efficiency is indicated for the higher grade sample.

During the period of time plotted, both tests achieved copper depletion down to less than 20 ppm.

The 'cross over' between plots for these tests should be noted. It is attributable to increased copper dissolution at the commencement of Test 18, the result of higher iron concentration (with no increase in applied current density).

FIGURE 17 : TESTS 11,18 (2 INCH CIRCUIT)
CUMULATIVE CURRENT EFFICIENCY VS TIME



Copper tenor has been shown to effect current efficiency on a periodic basis - lower grade generally resulting in lower efficiency. Higher initial copper grade in this case, however, is generally accompanied by higher iron concentration - which in turn can result in lower efficiency in stage 1 electrowinning. The net result is that, given a certain set of operating conditions, the effect of initial copper grade (within the range that may be expected at Mt Lyell) becomes a relatively minor factor.

c) Effect of changes in current density

Changing current density in treating this liquor has two opposing effects:

a) In general it has been shown that raising current density reduces current efficiency on a periodic basis.

a) High initial current density can however result in a net benefit to the process - through increasing copper recovery and minimising the overall effect of copper dissolution with high iron concentration.

It is likely, therefore, that a treatment system for this waste stream would entail progressive changes in current density through the circuit, matched to the perceived recovery stages outlined above.

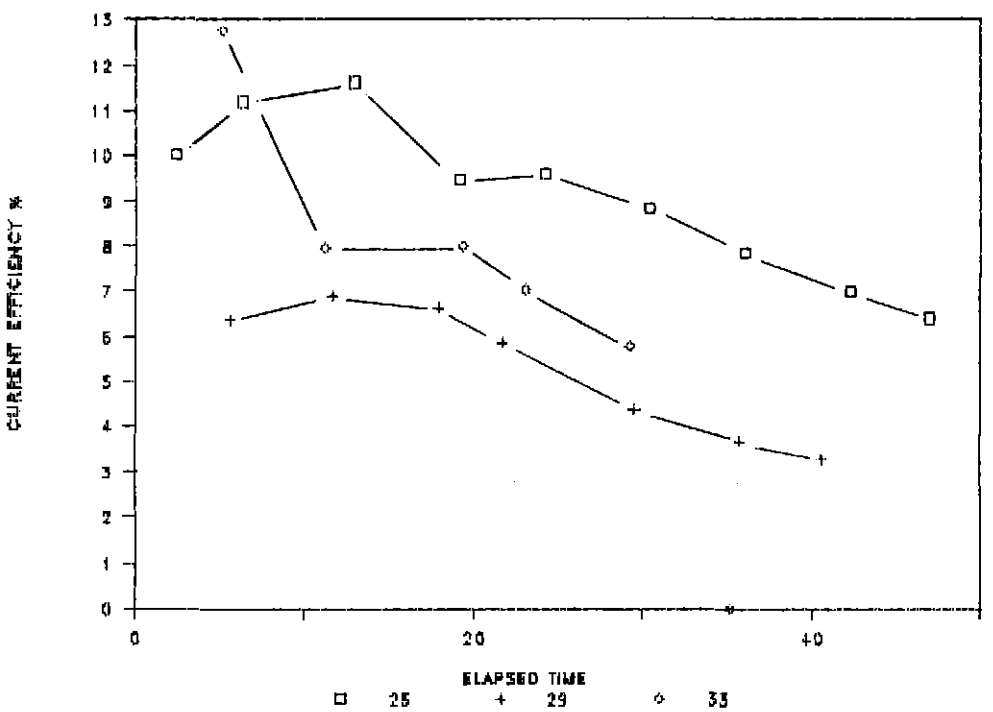
The potential benefits of high initial current density have been discussed above. Analysis of individual test results suggests that initial current density settings be governed (along with flow) in relation to iron concentration in the total waste stream.

If sufficient data were available, the required current setting for given iron concentration and flow could be mathematically calculated. For practical purposes, however, the present test programme suggests that the following settings should be used:

	Fe (ppm)	Current density (A/m^2)
2" Circuit	<1,000	>30
	>1,000	>70
4" Circuit	<1,000	>40
	>1,000	>60

Figure 18 compares performance between tests 25, 29 and 33; in terms of cumulative current efficiency over time.

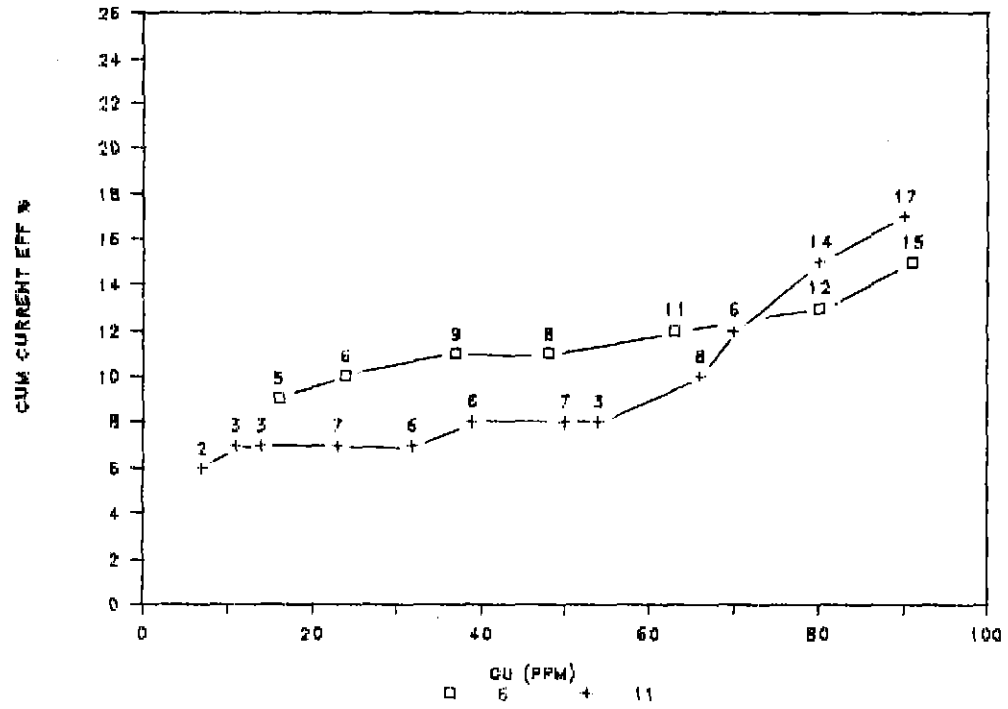
FIGURE 18 : CUM CURRENT EFFICIENCY VS CURRENT DENSITY TESTS ML 25,29,33



Although operating conditions differed slightly in these tests the plots provide a general view of the effect of increased current density (tests 25 vs 33) and copper grade (Test 29 vs 25).

The change in current efficiency with varying current is more precisely defined when analysed against copper grade rather than time. Figures 19 and 20 examine both cumulative and periodic efficiency at varying copper concentration, in a series of tests conducted with the 2" circuit.

FIGURE 19 : EFFICIENCY VS CU GRADE
2 INCH CIRCUIT LOW CU



Tests 6 and 11 were conducted at relatively low copper tenor (90ppm). Their plots demonstrate both of the effects summarised above. Increased current density (from 35 A/m^2 in Test 6 to 52 A/m^2 in Test 11):

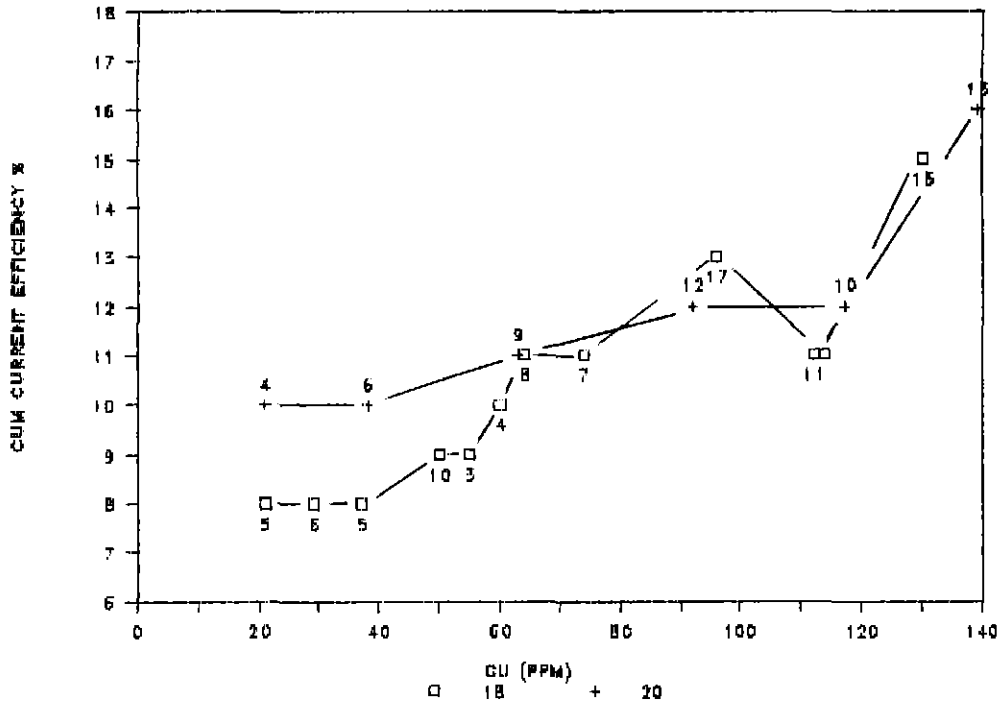
a) initially resulted in higher efficiency (approximately 15%) but,

b) yielded lower efficiency at lower grade.

Thus, again, the cross over between the plots.

On a cumulative basis, the higher current test resulted in lower overall current efficiency. This would not have been the case if the tests had been operated at higher iron concentration.

FIGURE 20 : EFFICIENCY VS CU GRADE
2 INCH CIRCUIT HIGH CU



At higher copper grade (accompanied by higher iron concentration), the initial dissolution effects complicate the graphic representation for tests 18 and 19.

Figure 20 demonstrates that at higher grade, the lower current density test achieved a lower cumulative current efficiency. Examination of the values given for periodic current efficiency show, however, that the lower current does generally result in higher efficiency at specific copper grade; and that the lower overall reading is a result of the significantly higher apparent efficiency achieved through high initial current density on the higher iron content sample.

The data collected show that it is possible, given specific copper/iron concentration, to optimise current efficiency through changes in current density during progressive liquor treatment. Although actual levels will be determined by iron content in the liquor, the range of solutions tested suggest the following base:

Stage 1 : 60-80 A/m²
Stage 2 : 40-60 A/m²

The establishment of operating guidelines would, however, have to take into closer account the relationship between initial and progressively changing chemical conditions. If current is reduced too early in the treatment cycle (i.e prior to the perceived iron reduction 'trigger point') the effect can be commencement at positive recovery and reversion to net loss if the change is too early (see previous report on Mt Lyell lab work : MDL test 14).

d) Flow effects on current efficiency

The rate of solution flow through the circuit has significant demonstrated impact on apparent current efficiency. Again, the test work has established two opposing influences:

a) Increased flow in the first stages of treatment result in an increase in copper leaching and, therefore, reduction in apparent current efficiency

b) In later stages of treatment increasing flow through the cells should enhance supply of copper ions to the cathode surface; bringing about an improvement in current efficiency at lower grades.

The former, as has been demonstrated in previous discussion, has a marked impact on overall performance in terms of cumulative current efficiency.

The marked initial difference in current efficiency recorded for tests 20 and 25 (Figure 21) is directly attributable to this effect. Both were operated in the 2 inch circuit, under similar chemical conditions but at varying flow rate (Test 20: 1800 l/hr; Test 25: 2400 l/hr).

A different presentation of the data from these two tests (Fig 22- cumulative efficiency vs copper grade) suggests that improvements possible through increased flow in later stages of treatment may be minimal. Following a period of high apparent efficiency in response to low flow (test 20) cumulative current efficiency curves against copper concentration are virtually identical

FIGURE 21 : CUMULATIVE EFFICIENCY VS TIME
DIFFERING FLOW RATE (ML 20,25)

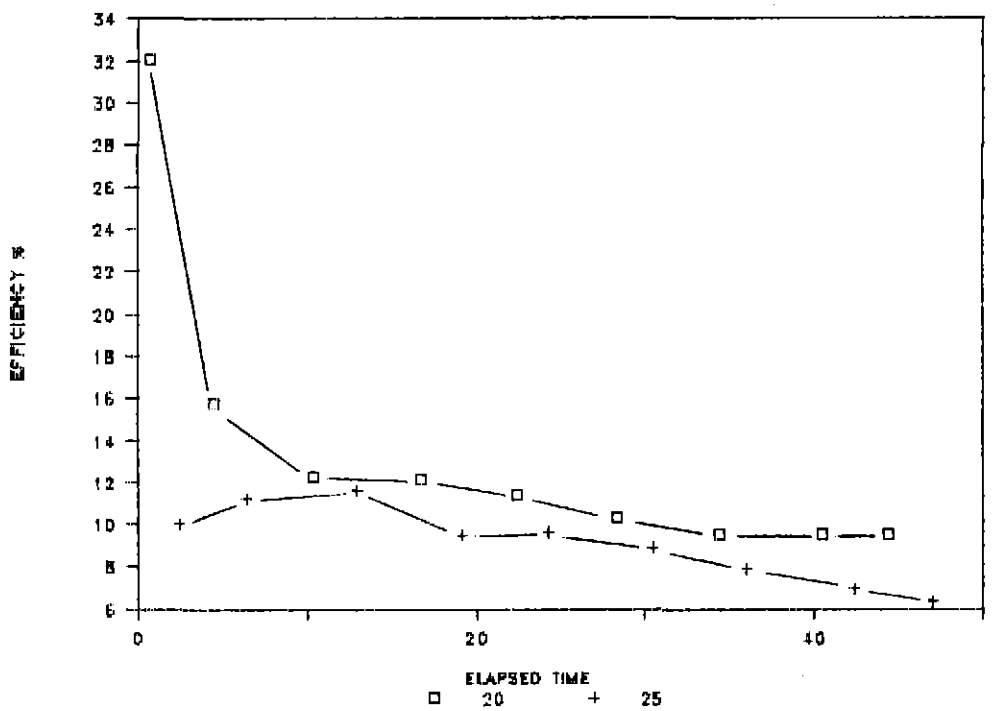
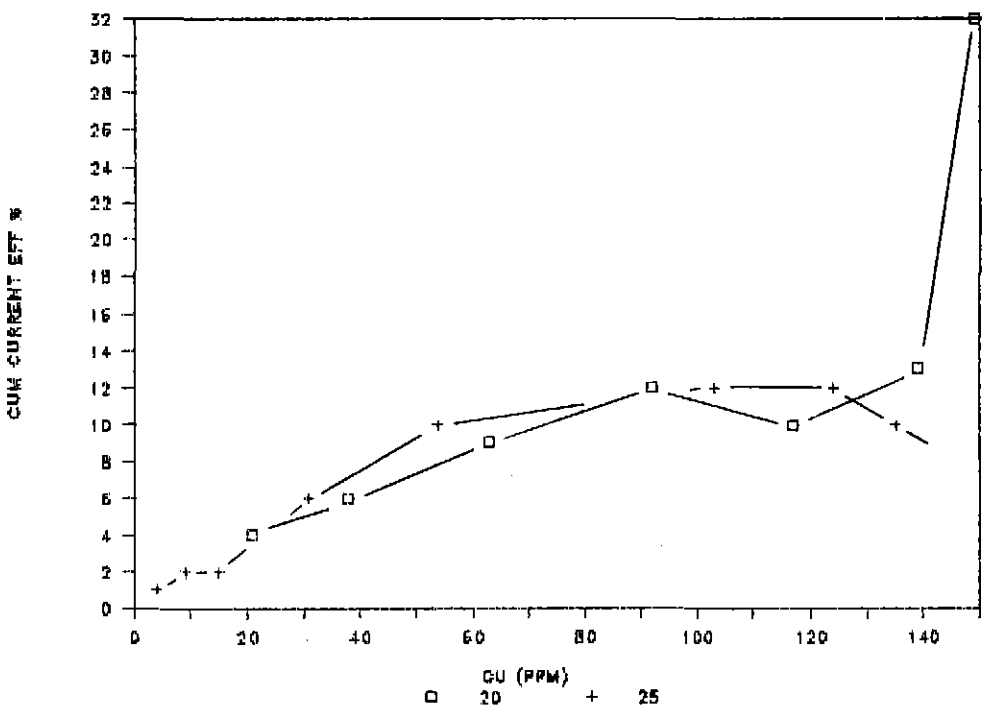


FIGURE 22 : CUMULATIVE EFFICIENCY VS COPPER GRADE
DIFFERING FLOW RATES (ML 20,25)



e) Effect of cell size on current efficiency

An examination of the comparative results presented in Table 1 suggests that cell size may have a significant effect on process performance. If this is the case, the governing factor must relate to differing flow patterns in the cells in question.

In designing the field test programme cross-sectional area calculation was utilised as a basis of 'scale up' from laboratory to production sized cells. Allowance was made for differing anode diameter; and pumps were selected accordingly.

The following factors were utilised to project 'scale up' operating parameters, from the small cell on which most of the laboratory testing was based:

Cell diameter	X-sectional area (net of anode) mm^2	Scale up factor (vs lab cell
32mm	766	
54mm	1800	x 2.4
100mm	7366	x 9.6

On the basis of these calculations, a flow rate of 7,200 litres per hour for the 4 inch cells would be equivalent to 1,800 litres per hour for the 2 inch circuit.

Results from the field programme, however, clearly show that performance in the 4 inch circuit has matched (or exceeded) that of the 2 inch circuit at significantly lower than this theoretical rate.

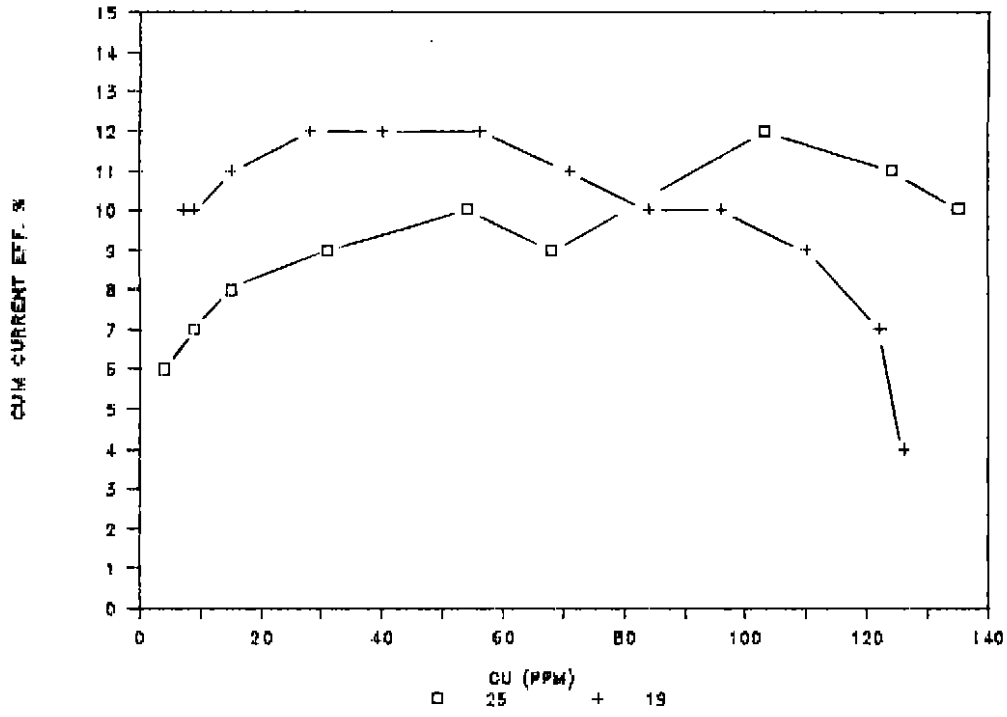
Clearly relative circumferences of the cells would appear to yield a more valid comparison. Even so, the larger cell retains the possibility of inherent process improvement. Operating at approximately double the flow of the 2" circuit, the 4 inch cells have achieved between 25 and 35% improvement in overall performance, for example:

Test 25 (2 inch circuit) : At flow rate 2400 l/hour, cumulative current efficiency to 15 ppm Cu of 9%

Test 19 (4 inch circuit) : At flow rate 4280 l/hour, cumulative current efficiency to 15 ppm Cu of 12%

Figure 23 examines the progressive relationship between copper grade and current efficiency in these two tests. The graph shows a clear separation of performance capabilities in the two cells during the electrowin cycle.

FIGURE 23 : CUMULATIVE EFFICIENCY VS CU GRADE
VARIED CELL SIZES



The smaller cell displays higher cumulative current efficiency at the commencement of treatment. This is attributed in this analysis to the effect of lower overall flow rate on copper dissolution.

Following apparent poor initial performance, due to exacerbated copper dissolution at high flow, the large cell displays progressively higher periodic current efficiency and therefore improved overall performance (at progressively reducing copper grade).

A comparison between Tests 1 (4 inch circuit) and 6 (2 inch circuit), operated under similar chemical and operating settings, supports the enhanced performance capability of the larger cell - the former achieving cumulative efficiency of 13% to 18ppm Cu, the latter 10% to a similar level.

Comparison between the latter tests shows, however, improvement in the larger cell throughout the test. This difference is attributed to the initial relatively low iron content in these tests. It is suggested that to maintain the enhanced performance at higher levels, the 4" cell requires operation at current density settings governed by iron concentration in the liquor.

The data suggest that the larger cell be used in a direct electrowin treatment route for Mt Lyell - flow and current density settings being tailored to minimise copper dissolution at the front end of the circuit.

f) Summary of Current efficiency Factors

The data collected demonstrate that current efficiency in direct treatment of the Mt Lyell liquor will be generally low. This is exacerbated by the fact that apparent recovery reflects a net relationship between copper electrowon and that dissolved (or redissolved) by ferric sulphate.

The study has shown that copper dissolution can be minimised and current efficiency can be enhanced through:

a) Changing flow rate through the electrowin cycle:

   i) Low initial flow to minimise the rate of copper dissolution, followed by

   ii) Higher flow rate in later treatment, to promote more rapid supply of copper ions to the cathode.

b) Changing current density, to:

   i) Maintain initial positive copper recovery by increasing the ratio of electrowon to dissolved copper - i.e increased initial current; and

   ii) Lower current density in later stages, to enhance the supply and demand requirements of this low grade liquor.

c) Selection of cell size

The data collected suggest that the following operating parameters (matched to the electrowin stages outlined above) will result in optimum current efficiency in treating the Mount Lyell liquor:

Stage 1

Cu range	:	150+ to 100 ppm
Cell size	:	4 inch
Flow rate	:	1800 l/hour
Current density	:	70 A/m^2
Projected efficiency: (cumulative)		16%
Test examples	:	Test 20 (extrapolated to 4" cell)

Stage 2

Cu range	:	100 to 15 ppm
Cell size	:	4 inch
Flow rate	:	4000 l/hour
Current density	:	50 A/m^2
Projected efficiency: (cumulative)		13%
Test examples	:	Tests 19,1

These settings will be utilised later in erecting a model circuit and calculating projected operating performance for costing and feasibility purposes.

5.2.2 Liquor Upgrade Option

The tests conducted at Mt Lyell on 'high' grade liquor, as would be expected, achieved much higher current efficiency. It is particularly notable that the new cell maintained efficiency in excess of 90% in stripping copper down to very low tenor (from 40 g/l down to less than 1 g/l).

Significant enhancement of performance, over a conventional cell, is the major reason that the use of solvent extraction in this context is possible. Significant capital cost savings are possible in both solvent extraction and electrowin circuits; and the proven versatility of the new cell allows maintenance of high performance over a wide range of chemical conditions.

Tests 4,5,10,15 and 22 were performed utilising the laboratory circuit on liquor made up as follows:

Test 4/5/10

Water	-	9.5 litres
Mt Lyell creek liquor	-	0.5 litres
Laboratory grade $CuSO_4$	-	1.6 kg

Test 15

As for 4/5/10

Test 22

'Stripped' liquor from Run 15	-	10 litres
Laboratory grade $CuSO_4$	-	1.6 kg

Sufficient copper sulphate was added to achieve an initial liquor grade of 40 g/l, a tenor that would be achieved through solvent extraction on the Mt Lyell waste stream (as tested by Henkel Corporation in Melbourne).

In the initial two runs, approximately 5% of the sample comprised liquor from the Mt Lyell waste stream, intended to introduce 'contaminants' to the target solution, equal to or greater than that which would result from solvent extraction.

The final test utilised 'spent' liquor from run 15 as a base solution, made up to approximately 40 g/l Cu through adding copper sulphate. The purpose was to determine cell voltage levels, at free acid concentrations that would more closely approach those of a solvent extraction eluate (liquor 'pre-treatment' having resulted in generation of 'free' sulphuric acid).

Table 2 below summarises efficiency levels achieved at differing current density settings.

TABLE 2 - Current Efficiency Measurements, High Grade Liquor

Test No	Current density (A/m^2)	Cu range (ppm)		Current Effic. (cumulative %)
		From	To	
4	515	37,920	24,360	98
5	394	24,200	1,000	100
10	121	980	158	81
15	618	40,800	524	100
22	900	39,800	9,900	98

Note : Current efficiency calculations are subject to limits of accuracy in copper assays and liquor volume measurements - efficiency of 98-100% therefore represents a performance envelope within which current inefficiency is not quantifiable.

Detailed records of periodic efficiency against depleting copper grade and time are presented in Appendix 3 herewith.

Two main facts are immediately apparent from the above table:

* Close to theoretical current efficiency is achieved in treating high grade liquor with minimal iron contamination.

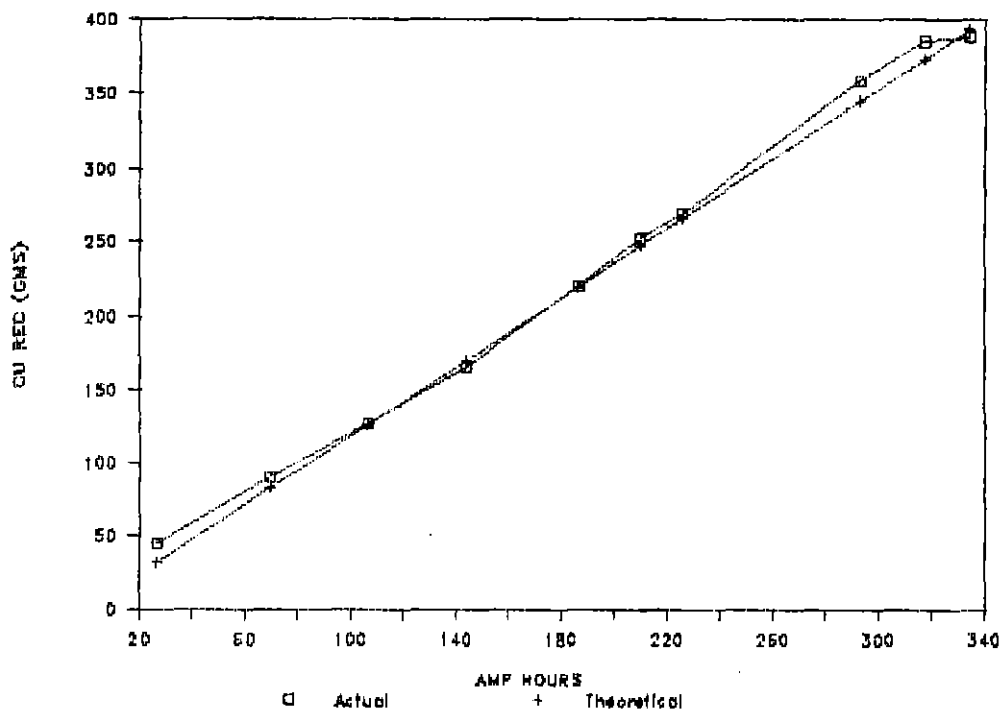
* The high efficiency achieved was at much higher current density than is appropriate for direct electrowinning.

Both observations have significant economic impact in the current context. High efficiency results in a reduction in power cost in copper electrowinning, by a factor of at least 8 (in comparison with the direct electrowin above).

Maintenance of high efficiencies at very high current density and down to very low copper grades greatly reduces the size of electrowin plant required for a specified production level.

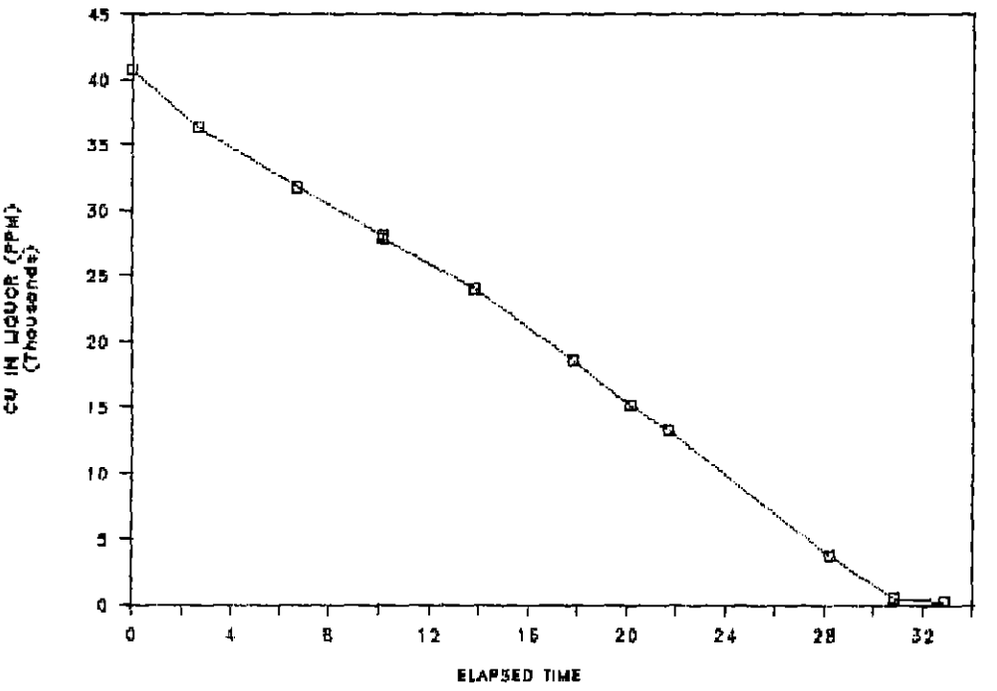
Figures 24 and 25 here provide an example of the result of testing in the middle range of operating parameters used (Test 15, current density 600+ A/m^2). The graph of copper depletion in liquor over time (Figure 25) shows almost straight line recovery between 40g/l and 0.24 g/l Cu. Figure 24 examines the relationship between copper recovered, in grams per amp hour of applied current, against theoretical yield.

FIGURE 24 : HIGH GRADE TEST 15
ACTUAL VS THEORETICAL GM CU/AMP HR



Plots of the other tests show the same close relationship between actual and theoretical recovery, even at a current density in excess of 900 A/m^2 (approximately 4 times that at which a conventional EW tankhouse operates).

FIGURE 25 : HIGH GRADE TEST 15
COPPER RECOVERY PROFILE



Current efficiency of 76% recorded for the last sample interval of Test 10 (460 to 158 ppm Cu) demonstrates the performance that can be expected from this process, on a relatively 'clean' liquor of similar grade to the main waste stream.

The data presented ably demonstrate the unique performance capabilities of the new cell, on a liquor of a composition that would be expected from solvent extraction treatment of the waste stream.

5.3 VOLTAGE

The cost of metal recovery in electrowinning is a function of:

a) Applied current
b) Current efficiency
c) Cell or circuit voltage

Although nature of electrodes and liquor circulation play their part, cell voltage is in turn governed principally by:

a) Applied current
b) Liquor chemistry, and
c) Spacing between the cell electrodes

Recorded voltage measurements, in testing both the untreated waste stream and the high grade fabricated liquors at Mt Lyell, obviously reflect the differing liquor chemistries. Both sample series are discussed below.

5.3.1 Direct Electrowin

The test programme has shown that changing initial copper and iron grade in the liquor, over the range tested, causes only limited variation in voltage for a given circuit configuration (Figure 26). Progressive chemical changes during electrowinning do, however, have a more marked effect.

For predictive purposes, therefore, it is assumed (within reasonable limits) that chemical composition of the untreated waste stream will have little impact on operating voltages.

In treating the Mount Lyell waste liquor by direct electrowinning, changes in current density and electrode spacing are the major contributors to variation in cell voltage.

Figure 27 illustrates the direct effects of current density, electrode spacing and progressive chemical change on cell voltage - being graphic representations of recorded measurements at the beginning and end of two separate test runs (Test 24 : 4 inch cell; Test 25 : 2 inch cell).

FIGURE 26 : CURRENT DENSITY VS CELL VOLTAGE
VARYING GRADE LIQUOR - ML 11,25

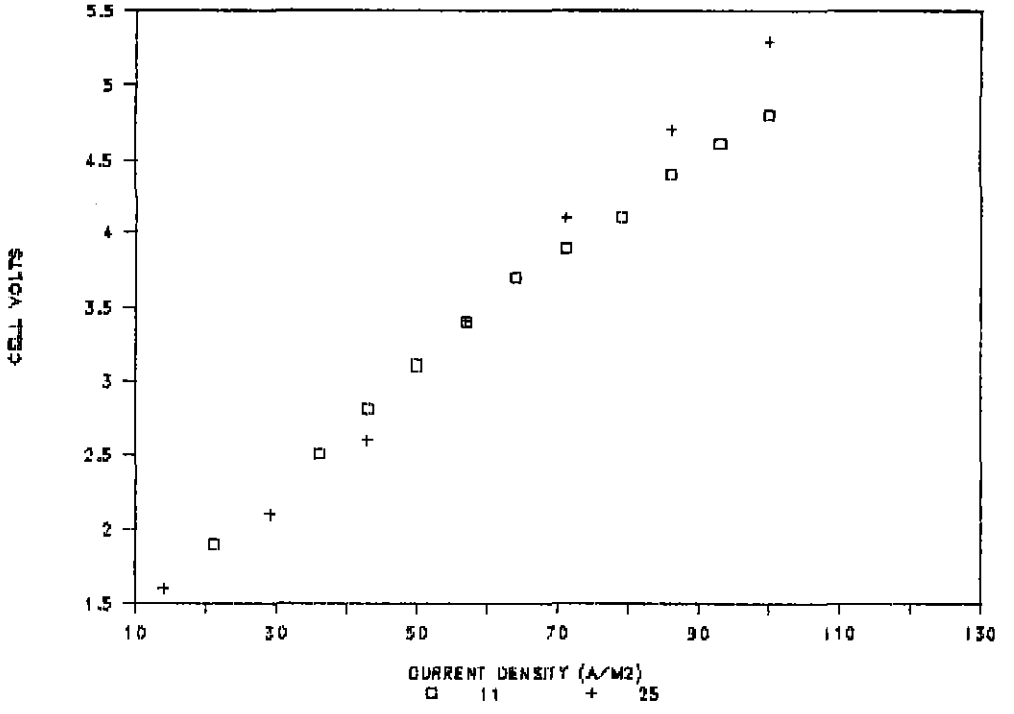
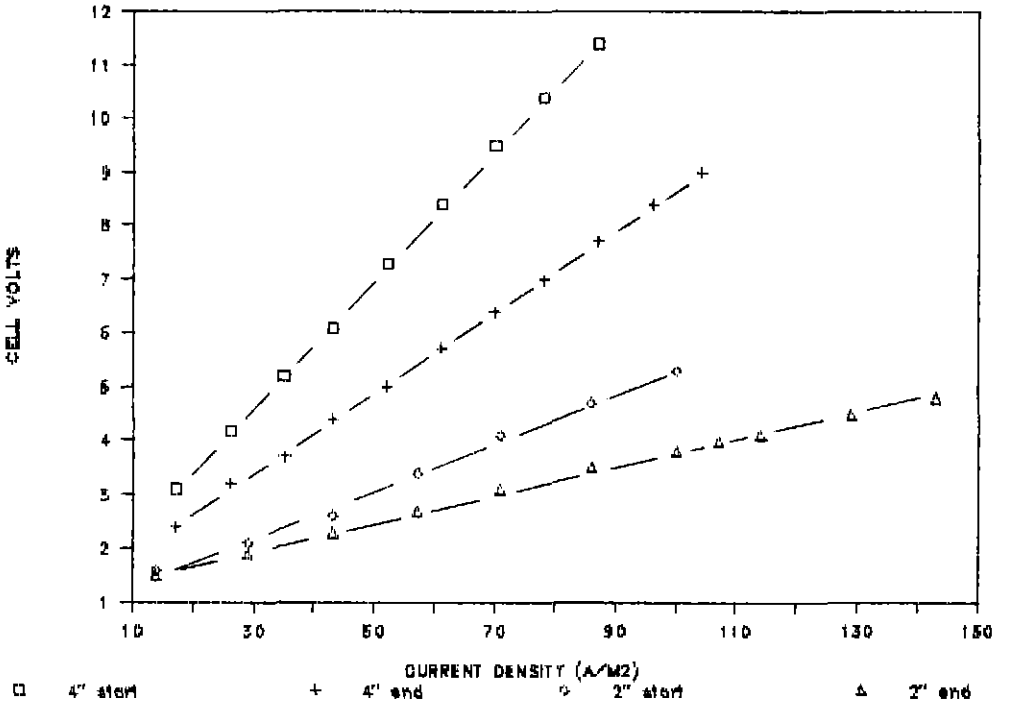


FIGURE 27 : CURRENT DENSITY VS CELL VOLTAGE
VARIATION ELECTRODE GAP (ML24,25)



The graphs supplied in Figure 27 clearly demonstrate:

a) A straight line relationship between current density and operating voltage for a given cell.

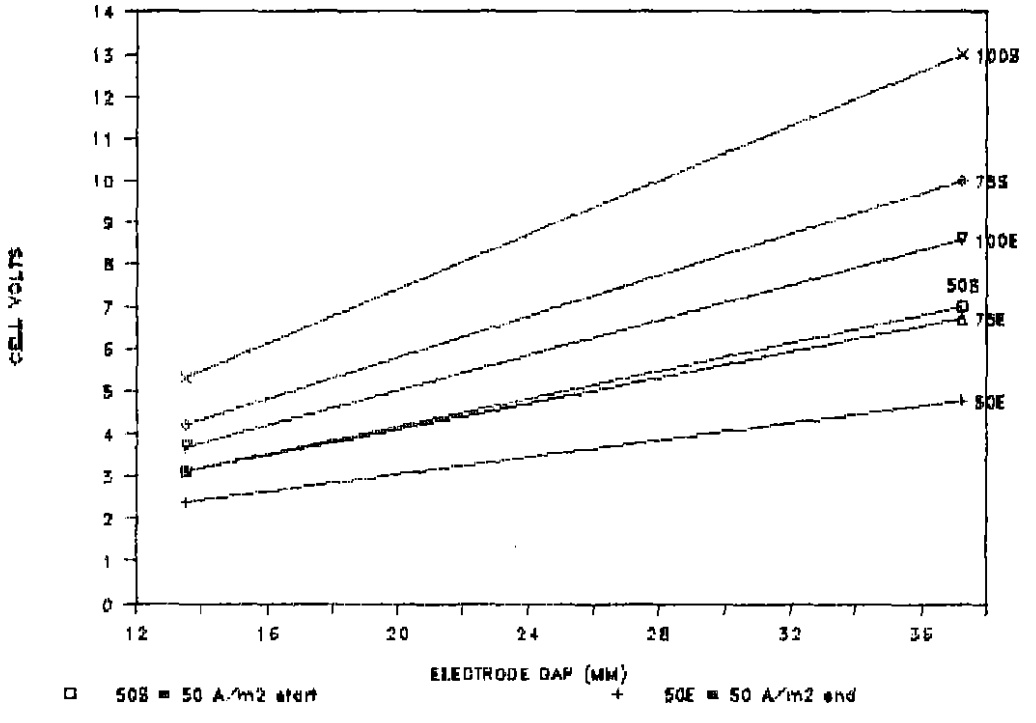
b) Comparison between cell voltage profiles in the 2 inch (electrode gap 13.5mm) and 4 inch (electrode gap 37.25mm)

c) The reduction in voltage that occurs between start and completion of liquor treatment, in response to the pH change occurring through reduction of iron and plating of copper (consequent release of free acid).

Even at low current density (30 A/m²), operating voltage for the larger cell exceeds twice that of the narrower unit. Within the range of settings selected as appropriate to the project (50-75 A/m² - see Current Efficiency above), the narrower electrode gap results in a threefold decrease in operating voltage.

Although the range of cells tested is not large enough to perform quantitative calculation, the relationship between voltage and electrode gap would appear to be linear (Fig 28).

FIGURE 28 : CELL VOLTAGE VS ELECTRODE GAP
VARYING CURRENT DENSITY (ML24,25)



The lack of an intermediate measurement between electrode gaps of 13.5 and 37.25mm may be the cause of the apparent linearity of this relationship. If it were to hold true, the effect of increasing electrode gap could be quantified at varying current densities:

	A/m^2	Increase per 10mm electrode gap
Commence treatment	100	+ 3.2 volts
	75	+ 2.5 volts
	50	+ 1.6 volts
End treatment	100	+ 2.1 volts
	75	+ 1.5 volts
	50	+ 1.0 volts

The data available demonstrate that the increase in voltage with widening electrode gap is exacerbated at high current density; but also that progressive liquor treatment and lowering current density reduce the effect.

Electrode gap in the cell under development is governed by the respective diameters of anode and cathode.

The advantage in use of a larger cell at Mt Lyell is clear - in reducing the number of cells required for a given production rate. Use of the larger cell from the field testing at high current density, would be prohibited due to high power cost. There are, however, no major constraints against altering anode diameter.

The narrower electrode gap used (13.5 mm) resulted in 'acceptable' voltages in the 2 inch circuit, over the range of current densities selected above. Even further narrowing of the gap (through further reducing diameter of the cathode due to plating) would have a continued beneficial effect on voltage; but could result in flow constriction and consequent increase in pumping costs.

A factor which will affect practical variation in electrode gap will be the nature of copper product chosen. Progressive plating of copper on the cathode automatically narrows the electrode gap. There is a resulting benefit, in reducing voltage, but the available space in the cell for plating before harvesting is restricted.

As will be detailed below, the conditions under which the Mt Lyell waste stream would be treated (with respect to both chemistry and projected operating mode) are such that formation of copper powder rather than plate is promoted. As this product will automatically 'flush' from the cell the constraints with respect to electrode gap are removed.

Sizing of the ideal cell for direct treatment of the Mt Lyell stream, therefore, becomes a balance between a number of factors:

a) The minimum practical electrode gap - to reduce operating voltage as low as possible without adversely affecting product harvesting conditions.

b) Preferred product nature and harvesting method.

c) Optimum current density settings and the balance between this factor and optimum cell production rate (a function of the area of the cathode and current efficiency)

The choice of cell dimensions for a model treatment circuit will be discussed in further detail below. However, in the context of operating voltage, use of a large diameter cell (4 inch or 100mm) is recommended; fitted with a larger anode sized to achieve an electrode spacing of between 12 and 15mm.

5.3.3 Liquor Upgrade Option

Voltage in the high grade liquor tests was significantly lower than for direct treatment of the waste, due to higher sulphate concentration and significant free sulphuric acid tenor following electrowinning.

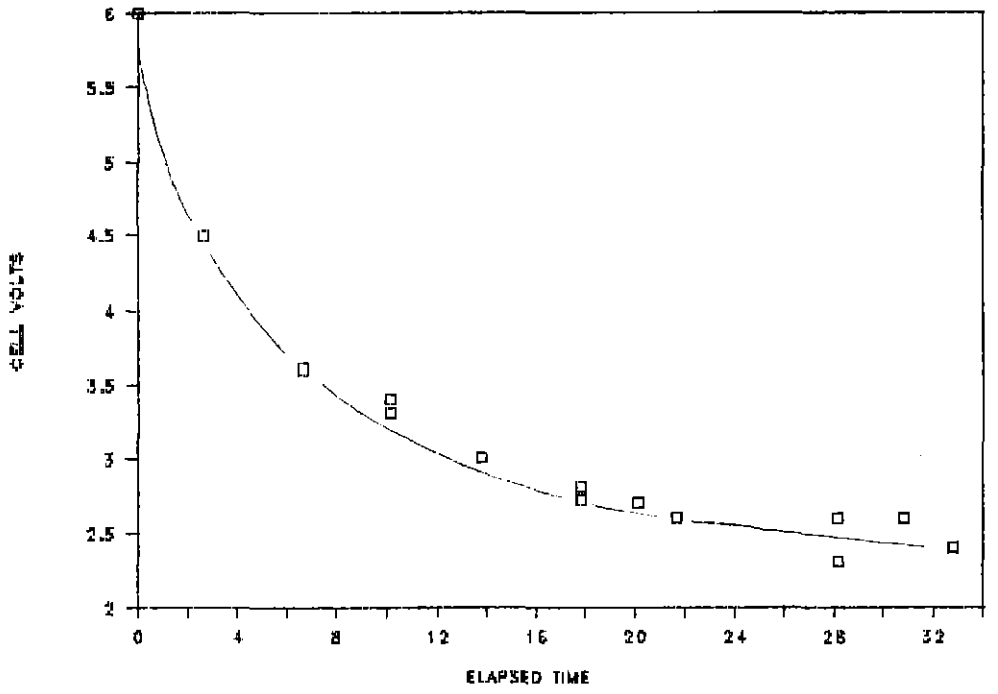
The following table details operating voltage and relevant chemical conditions at the commencement of each of Tests 4,5,10,15 and 22:

Test No.	Cu (g/l)	H_2SO_4 (g/l)	pH	Current density (A/m^2)	Cell volts
4	40	0	2.4	515	5.2
5	24	27	0.8	394	2.7
10	1	58	0.4	121	2.3
15	40	0	2.7	618	6.0
22	40	60	0.6	909	3.0

Voltage in all dropped rapidly during copper depletion, due to the progressive liberation of free acid; and to 'closure' of the electrode gap through copper plating. As recorded in the tabular results provided here (Appendix 3), even at the extremely high current densities tested, the resulting calculation of power cost for copper production are acceptable.

Figure 29 here plots the progressive lowering of cell voltage over time in Test 15 (which commenced at a pH of 2.6 and 0 free acid concentration).

FIGURE 29 : CELL VOLTAGE VS TIME
TEST 15



With a reasonable initial free acid concentration of 6% (60 g/l), operation as high as 900 A/m² current density recorded a relatively low voltage (Test 22 : start voltage 3.0).

All of the high grade testwork was conducted in a laboratory cell, with an initial electrode gap of 12mm. As no tests were conducted on the larger cell, a profile of projected voltage against electrode gap cannot be formulated.

It is notable that operation on this liquor at 900 A/m² recorded a cell voltage similar to operation of the larger cells directly on the waste stream - but at a current density 18 times higher (direct electrowin in 2" tube at 50 A/m² = 3 volts).

Utilising solvent extraction for 'upgrading' the Mount Lyell liquor would result in a treatment stream containing higher concentration of acid than used here. This will result in lower resistivity in the liquor and consequent lower operating voltage at a given electrode gap and current density. A lesser increase in voltage with cell size can therefore be expected.

Previous testwork supports the projection of cell potential of less than 3 volts, for a 100mm cell with 37 mm electrode spacing, used for later financial analysis of this treatment route. This assessment is based on operation at current density between 400 and 600 A/m^2.

5.4 E/W POWER CONSUMPTION

Being the dominant operating cost component, electrowin power consumption per unit of copper product will obviously be the final arbiter in economic analysis of the direct electrowinning route.

Through the detailed information recorded, this is obviously calculated on the basis of the following formula:

$$\frac{wh}{gm\ Cu} = \text{Power consumption (wh/gm Cu)}$$

Note: Being a simple metric conversion the calculation can be expressed also as kwh/kg or mwh/tonne Cu

Watt hours = circuit volts x amps x elapsed time
Gm copper = calculated recovery (in this case based on liquor assays)

Electrowin costs differ considerably between direct electrowinning and treatment of fabricated high grade liquor. The results of both test series are outlined below.

5.4.1 Direct Electrowin

The site testwork was aimed at assessing performance of the cell under a wide variety of conditions in a short period of time. Although early results indicated that changing conditions during treatment were required to achieve optimum performance, methodical approach dictated that settings be kept constant through individual test runs.

The detailed data recorded and discussed above, however, allows confident evaluation of performance during specific stages in the electrowin cycle.

Calculations of both periodic and cumulative performance have been recorded for all tests. Although the cumulative data are useful for analysis and interpretation of trends, detailed analysis of periodic performance data is required for potential treatment plant modelling.

The testwork entailed continual recycling of solution through a small number of cells. The analysis of periodic data undertaken, however, allows valid performance projections for a complete treatment cycle (based on continuous, 'single pass' liquor treatment).

Power consumption recorded through the test runs reflects the interrelationship of the performance factors discussed above.

Power consumption in the 4 inch circuit tests is obviously weighted by high operating voltages, due to wide electrode spacing (a factor that can be reduced through mechanical change to the cell).

On a cumulative basis direct 'cost' of copper production ranged from 10 mwh/tonne (Test 13, 4" circuit) to more than 60 mwh/tonne (for tests at high current density and wide electrode gap). The most promising runs on the two inch circuit recorded overall power cost of 25.8 and 29.0 mwh/tonne in depleting copper to less than 20 ppm (Tests 6 and 18).

Power consumption for short test intervals ranged from as low as 7.2 to in excess of 60 mwh/tonne.

Modelling of a potential on-line treatment system requires correlation of performance from different stages of specific tests, during which optimum conditions (of flow, current density and current efficiency as outlined above) were operating.

Table 3 summarises pertinent data from the tests utilised for this modelling.

Test 13 is included in the table as an example of the performance achievable from the process on partial reduction of ferric sulphate. In winning copper at current efficiencies in excess of 40%, the test recorded power consumption of around 7.5 mwh/tonne copper. This result is not used for modelling at this stage.

Detailed power consumption calculations appear in the tabulated results appended herewith (Appendix 3). It should be noted that the values recorded do not reflect 'optimised' operation for entire test runs. For example Test 19, which recorded relatively high current efficiency of 13% in stage 2 electrowinning (between 110 and 15 ppm Cu), resulted in relatively high cumulative power cost (approx. 40 mwh/t Cu). This was entirely due to high voltage (as a result of wide electrode gap in the 4 inch cells) and reduced initial recovery (the result of high flow rate).

Test 20 resulted in an overall power consumption of 39 mwh/tonne, higher than would be expected under ideal conditions due to low efficiency levels during stage 2 electrowinning (low flow and relatively higher current density).

TABLE 3 : Summary of Operating Data
Tests Selected for Modelling Purposes

TEST NO : ML 20
CIRCUIT : 2 Inch
CURRENT : 69 A/m2
FLOW : 1800 l/hour

Elapsed Time	Cu rec period (gms)	Cell Volts	Cell Amps	Curr eff. period
0.70	26.5	3.5	10.0	32
4.55	58.3	3.5	10.0	13
10.42	66.3	3.5	10.0	10
16.00	76.9	3.0	9.7	12
22.05	66.3	3.0	10.0	9
28.12	45.1	2.8	10.0	6
32.03	18.6	2.8	10.0	4

TEST NO : ML 19
CIRCUIT : 4 Inch
CURRENT : 64 A/m2
FLOW : 4280 l/hour

Elapsed Time	Cu rec period (gms)	Cell Volts	Cell Amps	Curr eff. period
2.03	0.0	7.3	15.0	8
3.90	13.6	7.4	16.0	10
8.17	40.8	6.9	16.0	14
12.12	47.6	6.5	14.0	11
16.35	40.8	6.6	15.8	13
20.03	44.2	6.5	15.5	14
23.95	51.0	6.0	15.2	16
27.73	54.4	5.8	15.5	11
31.77	40.8	5.8	15.5	12
35.78	44.2	5.6	15.5	5
40.37	20.4	5.8	15.5	2

TEST NO : ML 13
CIRCUIT : 4 Inch
CURRENT : 42 A/m2
FLOW : 4380 l/hour

Elapsed Time	Cu rec period (gms)	Cell Volts	Cell Amps	Curr eff. period
4.52	132.0	4.2	10.5	48
8.27	99.0	3.9	9.2	45
12.85	118.8	4.2	10.0	46
16.10	66.0	4.2	10.0	34
20.50	115.5	3.9	11.0	42
24.60	72.6	3.8	10.8	28
27.85	42.9	3.9	10.0	22
32.28	59.4	3.7	9.4	23
36.30	23.1	4.2	11.0	10
41.30	16.5	3.8	9.5	5

TEST NO : ML 1
CIRCUIT : 4 Inch
CURRENT : 42 A/m2
FLOW : 3800 l/hour

Elapsed Time	Cu rec period (gms)	Cell Volts	Cell Amps	Curr eff. period
2.67	29.2	6.8	10.0	19
4.62	15.9	6.4	10.0	14
6.50	13.3	5.9	10.0	12
8.33	13.3	5.9	10.0	12
12.92	34.5	5.7	10.0	13
16.55	29.2	5.5	10.0	14
20.77	29.2	5.3	10.0	12
24.50	21.2	5.3	10.0	10
26.73	15.9	5.1	10.0	12

Detailed analysis of the data recorded allows the design of an on line treatment system that would approach optimum performance for this liquor; and for which projected power consumption levels can be calculated. Different operating conditions would be applied to separate stages of the circuit, in accordance with the variations in the electrowin cycle defined above (see CURRENT EFFICIENCY - Direct Electrowin):

		Stage 1	Stage 2
Cu concentration			
Range (ppm)	- from	150	100
	- to	100	15
Applied operating settings			
Flow (l/hour)		1800-2000	3600-4000
Current density (A/m^2)		70	50
Cell diameter (mm)		100	100
Electrode gap (mm)		15	15
Resulting performance factors			
Projected voltage		3.7	2.7
Projected current efficiency (%)		16	13
Cu production/projected power consumption			
Gm Cu per m^2 cathode/per hour		13.22	7.67
Watts per m^2 cathode per hour		259	135
kwh/kg (mwh/tonne) Cu		19.6	17.6
Circuit power consumption			
% of copper produced in circuit		42.5	57.5
Overall power use (kwh/kg or mwh/t)		18.4	

NOTES:

1. Cu concentration range is that expected from the test programme and through analysis of company records for the waste stream.

   If copper grade proves lower, higher efficiency can be expected due to accompanying lower iron grade. Higher grade would be expected also to result in higher overall efficiency.

2. Applied operating settings are as defined in preceding sections of this report.

3. Voltage assumptions are based on the profiles established for the respective current densities proposed, extrapolated to an electrode gap of 15mm.

4. Current efficiencies assumed match those from the relevant stages of the tests outlined.

5. Gm Cu/hour = current x 1.18 x efficiency

6. Watts = current x voltage

7. Overall power usage is calculated from the copper ranges specified for Stage 1 and 2 electrowinning. Percentage production is utilised to calculate 'weighted' combined power consumption.

5.4.2 Liquor Upgrade Option

Similarly to the direct electrowin tests, variation in power 'cost' in the high grade tests is governed by solution chemistry and current density. All of the tests were conducted with constant electrode gap in the laboratory cell.

The tables provided (Appendix 3) detail calculated power usage on both periodic and cumulative bases. In stripping copper between 40 g/l and 0.5 g/l, the latter ranged from 1.9 - 3.6 kwh/kg.

Periodic power consumption reflected differing initial solution chemistry (low vs moderate acid concentration) and the progressive lowering of pH through formation of free acid by electrowinning. Measurements as low as 1.33 kwh/kg were recorded, for appreciable periods even at relatively high current density (e.g. Test 5: 17.6 g/l to 5.8 g/l Cu, at current density +/- 400 A/m^2).

The test operated at highest current density (Test 22, 900+ A/m^2) achieved overall copper recovery, between 39.8 g/l to 9.9 g/l, at 2.35 kwh/kg. As acid concentration remained below that which would be utilised for organic stripping in a solvent extraction step, this value can be regarded as a maximum for modelling purposes.

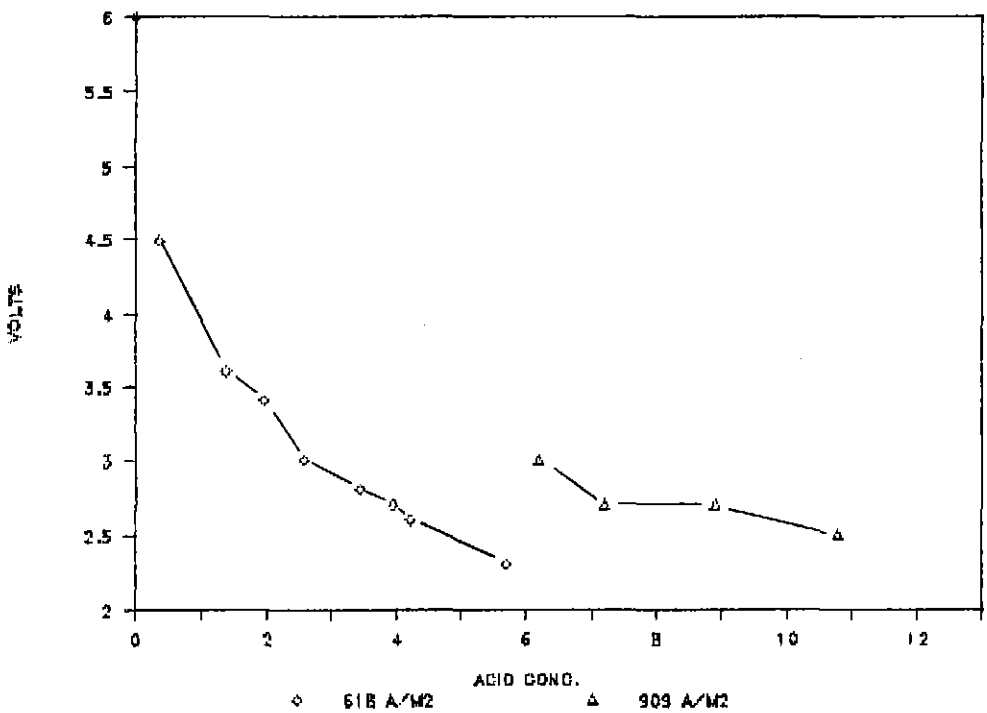
In providing a series of plots of voltage vs calculated acid concentration, at varying current density, Figure 30 can be used to predict operating voltage in this treatment route.

Projected operating voltage for the laboratory cell at varying current density and 12% free acid are either recorded or can be projected at:

A/m^2	Voltage
400	<1.7
600	1.7
900	2.4

Naturally, the size of the selected production cell will affect power consumption, through the effect of electrode gap on operating voltage.

FIGURE 30 : ACID CONCENTRATION VS VOLTAGE
HIGH GRADE TESTS VARYING CURRENT DENSITY



The results gained through the Mt Lyell programme do not allow a precise projection, but the data collected from the direct electrowinning work (and from previous studies) can be utilised to provide an indicative calculation of performance of a given cell.

Sizing of the cell that would be required for treating a solvent extraction stream at Mt Lyell is, again, somewhat dependant on the nature of product required. The testwork conducted yielded a dense cathode plate between 40 and 7 g/l, even for the higher current density runs.

In preliminary analysis, if plate is the preferred product, it is suggested that a two metre long 80mm diameter cell would be close to an optimum for this project. Commencing at 27.5 mm, electrode gap would decrease to approximately 16.5mm on progressive cathode loading to 50kg of copper.

On the basis of the data collected, projected operating factors for this cell would be as follows:

	Treatment cycle	
Cu concentration	START	END
Range (g/l)	40	10
Applied operating settings	START	END
Current density (A/m^2)	400	400
Cell diameter (mm)	80	80
Electrode gap (mm)	27.5	17.5
Resulting performance factors		
Projected voltage	3.0	1.7
Projected current efficiency (%)	95	95
Cu production/projected power consumption		
Gm Cu per m^2 cathode/per hour	448	448
Watts per m^2 cathode per hour	1200	680
kwh/kg (mwh/tonne) Cu	2.7	1.5

NOTES:

1. Copper range is estimated at this stage as 'delta copper' for electrowin circuit. Start and end factors reflect change in liquor chemistry through electrowinning.

2. If cell current is maintained constant actual current density will increase as the surface area of cathode decreases through progressive plating. Similarly operating voltage will decrease as electrode gap narrows through the same effect.

3. Current efficiency is as measured.

4. Estimation of cell numbers required in a treatment plant examine the possibility of operating at current density 600 A/m^2 as well as 400 A/m^2.

On the basis of averaging initial and final voltage and maintaining constant current density through the treatment cycle, projected power consumption for this model will be:

2.1 kwh/kg copper

It should be noted that the system can be operated in a mode that would promote powder, rather than plate, production - principally through increasing current density further.

If such an operating mode were chosen for a future treatment plant, narrower cells could be utilised. This would result in a drop in operating voltage; which would to some degree 'balance' out the higher voltage that would result from operating at higher current density.

The above calculation is, therefore, used to make reasonable projections of capital and operating costs for this treatment route.

5.5 PUMPING

5.5.1 General

As the effective use of this technology relies on high volume liquor transfer, the test programme has examined pressure drop through the various cells and circuits operated. These data provide a reasonable basis for calculation of pumping cost in a treatment circuit.

The relevant factor here is the pressure loss that occurs between inlet and outlet of individual cells - expressed in psi. Increased flow rate results in higher loss and an approximate 'cubed' increase in pump power consumption. Any treatment circuit using this cell should, therefore, be designed to minimise pressure drop at the flow rates envisaged.

Testing of a full sized cell during the initial laboratory programme suggested that flow rate above 1600 litres per hour for a 2 inch cell would result in unacceptable pumping cost (in a direct electrowin circuit at Mt Lyell), in accordance with the following table:

Flow (l/hr)	Pressure Drop (psi)	Pumping power (watts)
3200	6.2	75
2400	3.5	32
1600	1.5	9.4
800	0.4	1.2

In the context of direct electrowinning cells operating between 89 and 170 watts (2 metre long 100mm diameter cell, under the modelled conditions above), the higher flow rates would result in proportional pumping power up to 90% of electrowin power consumption.

Pressure tests on site, however, have demonstrated that minor alterations to the hardware will significantly reduce pumping costs.

Pressure drop across the cell is largely governed by the sizing of inlet and outlet spigots. Additional pressure losses in reticulation and fittings have been found to be minor in comparison with cell losses.

Although formal investigation of potential change in cell hydraulics have yet to be undertaken, it is expected that performance will not be

significantly affected by changing diameter of the liquor ports. The effect on pumping cost is, however, demonstrated to be marked.

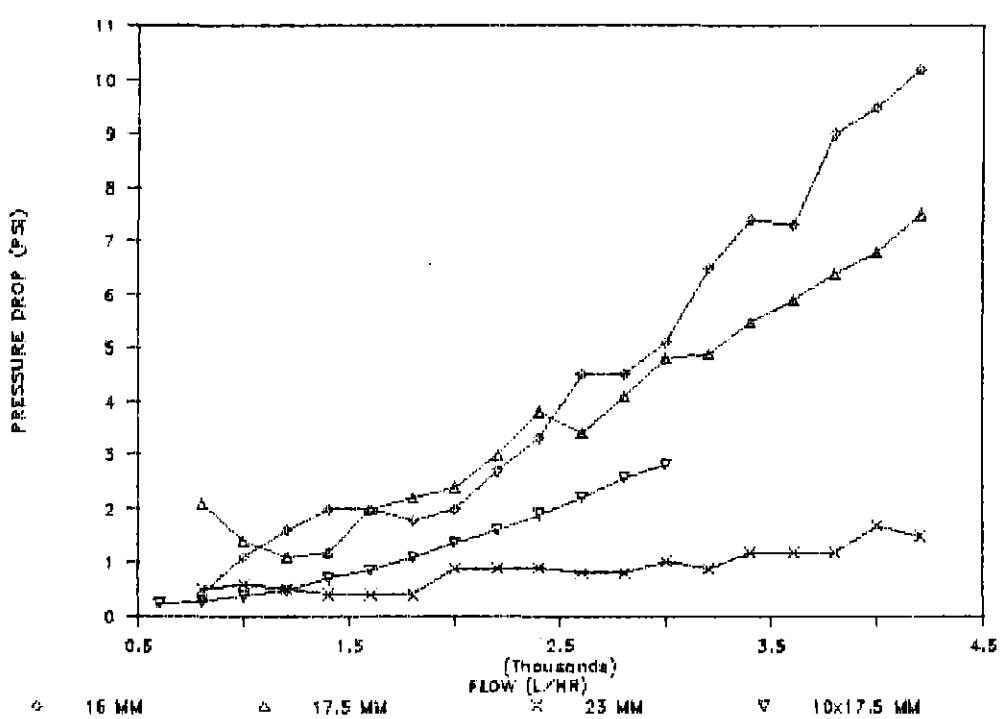
A variety of flow/pressure profiles have now been generated for the cells used at Mt Lyell - on both 2 inch and 4 inch units. Size of liquor ports in the cells has been varied, within limits which would not be expected to significantly alter hydrodynamics. These include:

Cell diameter	liquor ports (I.D. mm)
2 inch	16.0
	17.5
	23.0
4 inch	17.5
	23.0

The tests have shown little difference in pressure drop across individual 2 inch and 4 inch cells with the same liquor port diameter. Minor increase in diameter on both units has, however, been proven to significantly reduce pressure drop.

Figure 31 illustrates the effect of increased port diameter on testing of single cells. Also included is a plot of pressure drop across 10 cells connected in hydraulic series (total pressure loss being divided by 10 to achieve a value for comparison with the single cell tests).

FIGURE 31 : PRESSURE DROP INLET/OUTLET
VARIOUS DIAMETERS



The graphs clearly show that a significant decrease in pressure drop is achieved at high flow, even through a very small increase in port diameter (16 to 17.5 mm). Markedly lower pressure drop is indicated for a cell fitted with 23mm spigots.

At the higher flow rates modelled for a possible treatment circuit (3600-4000 l/hour), pressure loss across the cell is decreased by a factor of more than four through an increase in port diameter of only 5.5mm.

The graph for pressure loss over 10 cells in hydraulic series (2 inch circuit) suggests that losses in a circuit will in fact be less than that measured for individual cells. The data recorded does not take into account pressure drop through fittings required to mount pressure gauges on the cells. Measurement of a number of cells in hydraulic series will 'spread' such a loss - yielding the overall 'improved' plot in question.

For the sake of conservative projection, measurements on single cells are used in modelling potential treatment routes for Mount Lyell.

The optimum operating flow rates cited above would result in the following indicated pressure drop for a cell fitted with 23mm inside diameter ports:

1800-2000 l/hour	:	1.0 psi
3600-4000 l/hour	:	1.5 psi

Pumping power, in watts per cell, can be calculated in accordance with the following formula:

$$\text{Pump power (watts)} = \frac{Q \times \text{delta } P}{3.6 \times 145 \times E}$$

The formula being a combination of conversion factors and:

Q	= Flow in litres/hr
delta P	= Pressure drop in psi
E	= Pump efficiency (e.g. 50%)

The above pressure losses convert to an indicated power consumption to pumping per cell of:

Flow rate	Watts per cell
1800-2000 l/hr	7.3
3600-4000 l/hr	21.9

5.5.2 Direct Electrowin

Projected electrowin and pumping power consumption, for the model circuit outlined in section 5.4.1 above, can therefore be presented as follows:

	Electrowin stage	
	Stage 1	Stage 2
Operating parameters		
Flow (l/hr)	1800-2000	3600-4000
Cell pressure drop (psi)	1.0	1.5
Current density	70	50
Cathode area per cell (m^2)	0.63	0.63
Amps per cell	44	31
Projected voltage	3.7	2.7
% copper output	42.5	57.5
Power consumption		
Electrowin (watts per cell)	163	85
Pumping - watts per cell	7.3	21.9
- % x electrowin	4.4	25

At the higher flow rate for stage 2 electrowinning, pumping cost remains high as a proportion of electrowin power. However, averaging of the calculations (on a weighted basis against proportion of the plant operating at specific flow rate) indicates the following projected pump cost as a proportion of operating electrowin power:

Pumping power cost = 16% x electrowin power

On the basis of previous calculations of projected electrowin power consumption (5.4.1 above) pumping cost in the circuit will approximate to:

2.9 kwh/kg copper

It should be noted that testing of a circuit has recorded lower pressure drop factors than single cells, making this calculation a conservative estimate.

5.5.3 Liquor Upgrade Option

Pumping power cost in the use of this technology is presented above as a comparative factor against power usage in individual cells.

Proportionately, therefore, pumping power usage, although remaining constant and related to dimensional factors in the cell, will decrease as applied electrowin power is increased.

In other words operation of a cell at higher current will reduce the cost of pumping per tonne of product.

It has been suggested that cells would be operated at a current density of 400 A/m^2 (up to 600 A/m^2) on high grade liquor from a solvent extraction plant. At 200 amps per cell (projected cell dimensions = 0.5m^2) electrowin power will equal approximately 740 watts per cell.

On the basis of the above calculations, pumping power consumption becomes a minor cost in comparison with electrowin power.

5.6 PRODUCT NATURE AND QUALITY

Selected cells were opened, cleaned and examined at the end of each test run. Again, the nature of product is related to chemical and operating conditions applied - the most significant difference being between direct electrowinning from the waste stream; and from high grade fabricated liquor.

5.6.1 Direct electrowin

Both copper powder and plate were produced during direct electrowinning. The method of operation chosen (recycling of a chosen sample volume) coupled with chemical conditions in the liquor (redissolution of copper by ferric sulphate) masks the conditions under which each product is formed. The following general comments are supported by the observations made:

a) The formation of plate occurs at higher grade and under low current density conditions.

b) Low grade, low flow and high current density promote powder formation.

c) Chemical conditions are such that plate formed in initial treatment of the solution is largely redissolved by subsequent 'attack' by ferric sulphate

The requirements for plate formation through most of the treatment cycle are outside of those proposed for the model treatment circuit- the latter entailing:

Stage 1 : Higher copper grade in solution.

High current density, to alleviate the effect of ferric iron dissolution of copper on overall power consumption. Low flow regime designed for further minimisation of the iron effect.

Stage 2 : Generally low copper grade.

Lower current density and higher intended flow would promote plate production only if copper grade were maintained.

Operation of an on-line circuit for the waste stream would result in any plate formed, in initial treatment, being subject to continual attack by 'fresh' ferric sulphate solution.

Therefore, although the technology allows tailoring of product through changing operating parameters, the product expected from continuous direct electrowinning from the Mount Lyell waste stream comprises fine copper powder.

Powder forms as a fine layer on the cathode surface - which, on reaching a certain thickness, detaches under the influence of flow related shear. All of the test runs have demonstrated that the material detaches readily during operation; allowing harvesting thereof on a continuous basis.

Product is fine grained, but has been demonstrated to settle rapidly due to the high specific gravity of copper. It is envisaged that harvesting would entail simple gravity settling at the end of each bank of cells.

A number of batches of powder product were retained for assay and reference purposes. Indicative testing of one representative sample, by the site laboratory, returned an assayed grade of 99.3% copper.

Further assays were not undertaken for two reasons:

a) Electrowinning was generally continued beyond almost total depletion of copper - to a point that it would be generally expected that other metals in solution may commence plating.

b) The volume of product from individual test runs was limited (maximum contained copper in samples being less than 300 grams) resulting in high possibility of sample contamination.

The one assay performed, however, indicates that grade achievable in the product is extremely high. Analysis of iron and aluminium grades through testing indicates that little or none of these metals will plate, given the right conditions. The 0.7% residual in the sample is suggested to reflect oxidation or contamination during collection.

5.6.2 Liquor Upgrade Option

In a conventional plant (due to limitations in flow and efficiency) it would be expected that most of the product from operating at the high current densities chosen for the Mount Lyell programme would be in the form of powder.

All of the tests recorded here recovered copper as dense cathode plate down to very low liquor tenors.

The grade at which powder formation commences varies in relation to current density and flow through the cell. Generally, however, coherent plate is maintained down to a grade of approximately 5 g/l at current densities between 400 and 600 A/m^2. The apparent 'boundary' between product types at 900+ A/m^2 is slightly higher, at approximately 7 g/l Cu.

Operation on a liquor of this chemistry allows extreme versatility in control of product, the system (through lowering of current density and increasing flow velocity) being capable of dense cathode production down to far lower copper grades.

It should be noted that progression from plate to powder is not accompanied by reduction in current efficiency in the tests, until copper grade in the liquor reaches levels below 500ppm.

In the Mount Lyell tests, nature of the plate changed progressively with increasing current density - in general terms as follows:

A/m^2	Nature of plating
400	Dense plate, slight roughness on surface
500	Dense plate, minor roughness on surface
600	Dense plate, slightly nodular
900	Dense nodular plate

The surface characteristics observed were maintained throughout the test runs at given current density - i.e the nodularity at higher settings did not increase with progressively thicker plating. Required loading on the cathode can therefore be expected even at these high current densities without interference with the electrical circuit across the electrolyte.

It has been suggested that a cell diameter of 80mm and current density setting of $400A/m^2$ would be appropriate to the Mt Lyell project (5.2.2 above). A 2 metre cell would reach a harvest weight, of 50 kg dense copper product, on plating of the cathode to an approximate thickness of 10mm.

Similarly to the results of testing of liquors from other primary mining projects, a product purity in excess of 99.9% copper can be expected.

6.0 THE MOUNT LYELL RESOURCE

Although examination of the Mt Lyell orebody and its environs suggests that copper leaching could be promoted at a number of localities, the scope of the current exercise has been limited to existing and easily containable waste streams.

The major targets for this programme therefore comprise:

1. The West Lyell waste rock underflow (Fig 32, locality 5), and

2. The conveyor tunnel outflow (Fig 32, locality 8a).

The former arises through leaching of the substantial dumps of waste ore from the old West Lyell open cut; and is naturally confined to a creek which flows past and through the current treatment plant and office areas (see Figure 32).

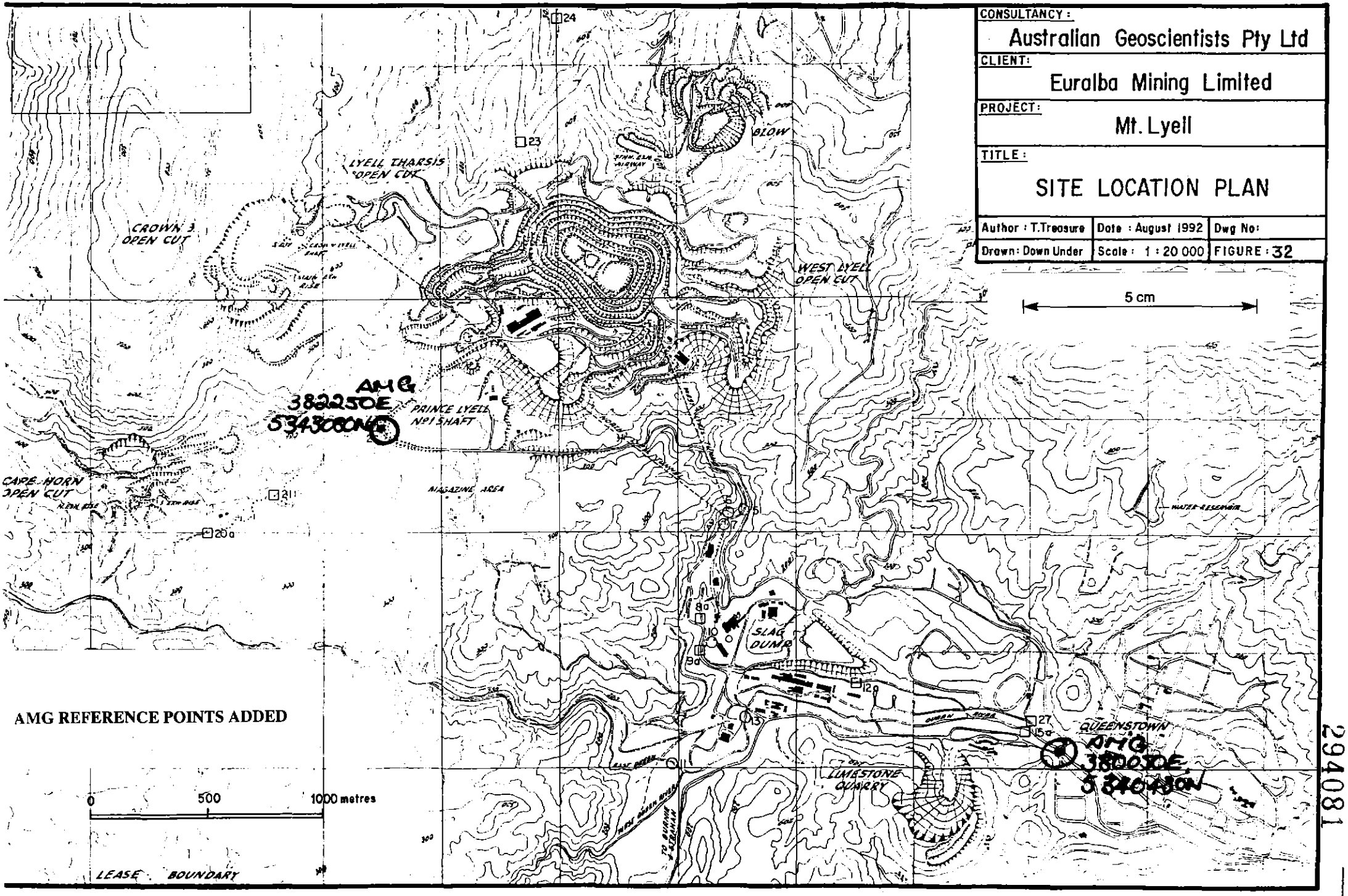
The 47 million tonnes of waste rock produced over the life of the West Lyell open cut contains approximately 10% pyrite and appreciable (albeit subeconomic at the time of operation of the open cut) copper concentrations. Pyrite is present in a finely divided form, amenable to oxidation and acid formation. These factors combine with consistent annual rainfall to produce high flows of acidic solution which are high in soluble copper concentration.

The current mining operation is concentrated on the Prince Lyell orebody, which extends below the old West Lyell open cut. Sub-level open stoping produces approximately 1.5 million tonnes of ore annually, at average grade 1.6% copper, for treatment by floatation. As the mine moves deeper, lower grade material flanking the mined orebody is allowed to cave into previous mining levels.

As mining is limited to a cut-off grade of around 0.7% copper, the broken ore underground represents a substantial potential resource of slow leaching ore.

As a consequence, rain falling on the open cut and its catchment percolates through 350 metres of fractured and caving pyritic rock before it reaches the current mining level. Mine dewatering therefore results in a substantial stream of copper bearing waste liquor, of generally higher grade than the waste dump effluent.

The operating company (Mount Lyell Mining and Railway Co.) has maintained a methodical sampling programme on the waste streams over a number of years. The data collected over the past two years from the localities of particular interest have been examined during the course of this programme. In general, the interpretation conducted supports results reported by the local environmental manager (I.Woods, Renison Consolidated Limited), following detailed collation in 1991.


CONSULTANCY :
Australian Geoscientists Pty Ltd
CLIENT:
Euralba Mining Limited
PROJECT:
Mt. Lyell
TITLE :
SITE LOCATION PLAN
Author : T.Treasure
Date : August 1992
Dwg No:
Drawn: Down Under
Scale : 1 : 20 000
FIGURE : 32
5 cm
CROWN 3 OPEN CUT
LYELL THARSIS OPEN CUT
GLOW
WEST LYELL OPEN CUT
AMG
382250E
5343080N
PRINCE LYELL NO1 SHAFT
CAPE HORN OPEN CUT
MAGAZINE AREA
SLAG DUMP
LIMESTONE QUARRY
QUEENSTOWN
AMG
380050E
5340450N
AMG REFERENCE POINTS ADDED
0
500
1000 metres
LEASE BOUNDARY

2940081

Typical characteristics of the initial target streams have been reported by Woods as follows:

Factor	Waste rock	Mine water	Combined
Cu (ppm)	130	170	150
Fe (ppm)	1160	520	854
Al (ppm)	300	200	252
Zn (ppm)	13	7	10
Mn (ppm)	140	40	92
SO_4 (ppm)	8200	3800	6097
pH	2.5	2.9	
Flow (m^3/day)	4700	4300	9000

The resulting combined flow contains approximately 1.35 tonnes of copper per day - 90% extraction of which would require a 1.2 tonne per day production facility.

Records of flow from the mine dewatering show a significant increase in copper concentration in that source, for the period Jan-Dec 1991. Formal records for this period (Locality 8a, Appendix 4) suggest an average copper content in excess of 300 ppm, for a flow in the region of 5,500 m^3/day.

Such an increase would yield an additional 0.9 tonnes of copper daily- requiring a production capacity of approximately 2 t/day.

Woods' comprehensive study in 1991 suggested that a total of approximately 700 tonnes of copper leaves the site annually in the form of acid mine drainage.

Several pertinent observations impact on the current study:

* There are a number of waste streams currently being generated by the site. The two major sources will yield a minimum of 1.2 tonnes of copper per day, based on recovery of 90% of contained metal.

* There are several factors that could significantly increase this yield:

  a) Copper concentration in mine dewatering appears to have increased in recent years.

  b) Subsidiary AMD streams could be captured and added to the treatment system.

  c) More rapid leaching of copper from the dumps and underground ore could be promoted.

Note this source will only contribute [illegible] pumped

The former would have immediate impact on the project; the latter being a future possibility only.

* Typical characteristics generated for the liquor from records over a reasonable period of time differ, possibly significantly, from that tested in the current programme:

  Specifically: Iron concentration in the dump waste stream is lower than that tested (850 ppm for the combined stream as opposed to approximately 1500 ppm in the current programme).

Given a decision to proceed to the next stage of this project, detailed analysis of these factors would be undertaken. For the purpose of this study, however, minimum facility requirements are confidently projected as follows:

Waste volume	:	9,000m3/day
Copper tenor	:	150 ppm
Projected recovery	:	90%
Production capacity	:	1.2 tpd copper

Naturally, a project based on treatment of the waste streams would entail provision for significant short term variations in flow - either by establishment of a storage dam and/or through isolation of the liquor sources.

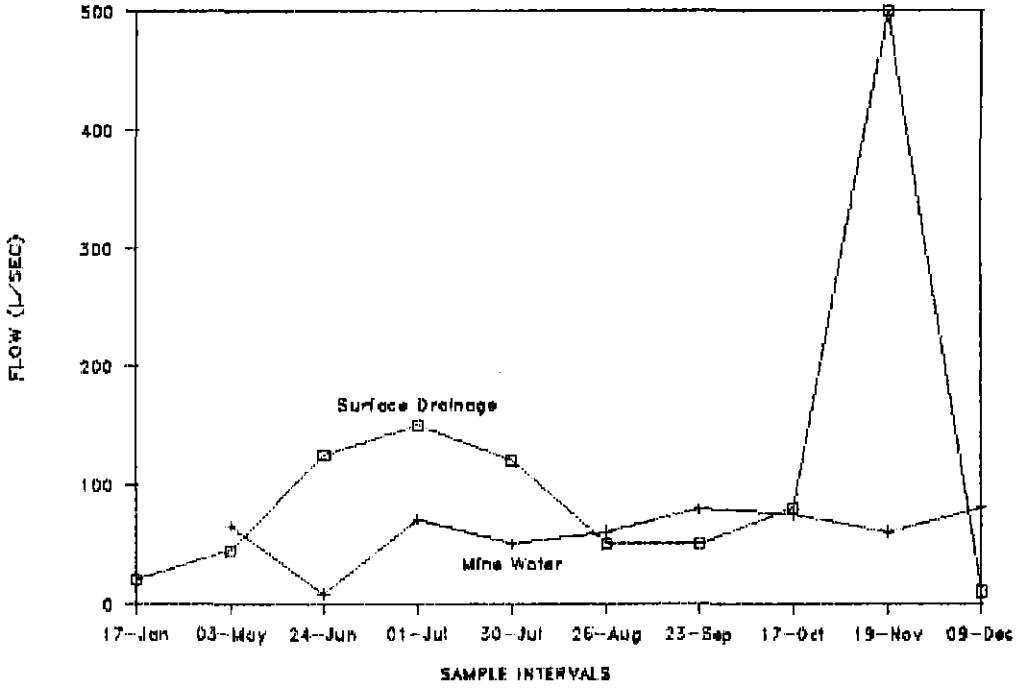
The effect of torrential rainfall on the surface drainage system has been observed to be of short duration, reflecting rapid surface run-off from the total catchment area. The result (at, for example, the test site chosen) is obviously a significant short term increase in volume, accompanied by decrease in contained copper grade.

A liquor management scheme, which would isolate the major copper-bearing solutions from fresh run-off could be readily imposed.

Large volume buffers exist at the point of source of both major waste streams. Percolation through ore in the waste dumps and underground should be relatively evenly distributed and slow, as a result of which the effect of occasional torrential rain should be 'smoothed'.

Comparison of recorded flows from both sources during 1991 (Fig 32) supports this observation - mine pumping rates being relatively constant through the year, with an increase by approximately 30% following a dramatic increase in surface flow (probably the result of a short duration storm - on or close to the 19th Nov).

FIGURE 33 : FLOW MEASUREMENTS, 1991
PRINCIPAL WASTE SOURCES



The mounting of a treatment operation on these liquors should contemplate the possible closure of the Mount Lyell mine in four years. Once closed, the underground workings could represent and ideal reservoir for the operation.

It is considered, therefore, that the liquor storage facilities required for a treatment operation can be minimised (to possibly equivalent to 7 days plant throughput). In addition, both treatment routes under review will be able to handle reasonable variations in throughput without significantly affecting copper recovery (up to 15% variation in flow).

7.0 TREATMENT CIRCUIT OPTIONS

Two treatment routes are considered to offer potential viability in extraction of the copper from the Mount Lyell waste streams:

OPTION A : Direct Electrowinning, on a single pass basis, through banks of cells mounted in hydraulic series.

OPTION B : 'Upgrading' of the liquor through solvent extraction, in conjunction with copper recovery from a recycle stream.

Solvent extraction, in this context, is a well established technology. Its basic principles will not be elaborated here, aside from references to possible process improvements that may be achieved in using the new cell for subsequent copper electrowinning.

The choice of a treatment circuit will naturally rely on study of capital and operating costs - a preliminary analysis of which is presented later in this report.

It is stressed, however, that the viability of the solvent extraction route relies heavily on the use of the new cell:

a) By virtue of the reductions in capital cost that are possible in the electrowin circuit, and equally

b) By virtue of simplification of the solvent extraction circuit, possible through use of the new technology.

Model circuits for both treatment options are outlined below. Although combination of the major waste streams and sourcing of other liquors may increase performance or expected copper yields (Section 6.0 above), modelling will assume worst case conditions, similar to those encountered in the current test programme.

7.1 DIRECT ELECTROWIN ROUTE

Modelling of performance factors detailed above, would require installation of the following:

7.1.1 Electrowin cells

The number of cells required in a direct electrowin circuit is calculated on the basis of the following criteria:

		STAGE 1	STAGE 2
Cu grade in liquor	ppm	150	
Liquor volume	m^3/hr	375	
Target copper recovery	%	90	
Cell dimensions:			
Length	mm	2000	
Diameter	mm	100	
Cathode area	m^2	0.63	
Current density	A/m^2	70	50
Current efficiency	%	16	13
Amps per cell		44	32
Production per cell	gm Cu/hr	8.3	4.9
Proportion of plant	% of total	30	70
Proportion of production	% of total	42.48	57.52
NO. OF CELLS REQUIRED	per stage	2580	6020
	TOTAL	8600	

Although a large number of cells are therefore required for the direct electrowin route, it should be noted that plant operation (designed to produce powder) would essentially be automatic.

7.1.2 Pumping/liquor reticulation

A direct electrowin facility would be intended to achieve the desired copper depletion in one pass through banks of cells linked in hydraulic series. At the designed flow rate (section 5.2.1 above), this will require establishment of approximately 100 banks of cells (375,000 l/hr at 3,600 l/hr through individual banks).

On the basis of the criteria outlined above each bank of cells would comprise two sections:

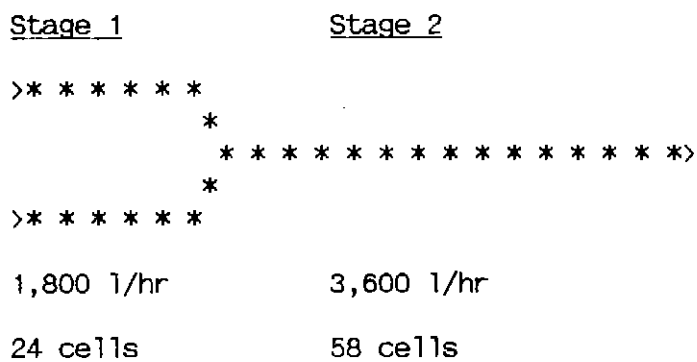
Section 1 : Reflecting optimum chosen parameters for stage 1 electrowinning.

Two parallel sets of 12 cells operating at 1800 l/hour and 70 A/m².

Section 2 : Under chosen parameters for stage 2 electrowinning.

One line of 58 cells in hydraulic series operating at 3,600 l/hour and 50 A/m².

Flow through the two sets in Section 1 would join to run directly through Section 2, thus achieving the differing rates required- i.e:



A relatively complex reticulation system would be required, probably entailing supply of liquor via manifold to a number of banks of cells. Connection between cells is simple and requires only flexible hoses.

Situation of the facility could be such that advantage could be taken of natural 'head' (from points of source of the solution) for operation at flow required in Stage 1. For modelling purposes, however, pumping from a static reservoir is assumed.

These cell configurations and flow parameters require pumping of a total of approximately 6,000 litres per minute at a head of 80-100 psi. This requirement could be met by a single pump (with a stand-by unit) or through a series of units. In order to maintain close control of flow; and to allow possible variation under changing conditions, a 'booster' pump system could be installed within the circuit.

Exact pump specification can await further study as it is likely that direct electrowinning would require further piloting prior to final design.

7.1.3 Power supply/Rectifier

Analysis of power consumption levels (section 5.4 above) suggests the following AC-DC conversion requirement for the plant modelled:

Production rate	:	50 kg Cu per hour
Operating power requirement	:	920 kW
Projected rectifier efficiency	:	90%
Rectifier specification	:	1,000 kW

Calculation of power requirement in pumping (section 5.5 above- equivalent to approximately 16% of electrowin power), dictates supply of approximately 150 kW for this purpose (pump efficiency factor already built in).

Total plant power requirement is therefore indicated to be between 1.1 and 1.2 MW.

7.1.4 Product Harvesting

As copper production is relatively low (1.2 tonnes Cu as powder) it is suggested that product harvesting and storage facilities would not be elaborate. A simple system comprising settling tanks, product washing and filtering ('wet' bagging at around 20% w/w moisture) is currently envisaged.

Although, as will be discussed further below, it is likely that the direct electrowin route will yield significant operating return, it requires large numbers of cells and relatively large pump and power facilities. It operation would, however, be simple and automated.

Being based on modelling of experimental data, prudence would require further piloting before finalising plans for a production plant of this nature. This could be readily effected through operating a single bank of cells, in the pattern outlined above.

7.2 LIQUOR UPGRADE ROUTE

The attractions in using solvent extraction on the Mount Lyell waste are obvious, from the discussions of operating performance outlined above. In general, they arise through alleviation of the adverse effects of high iron levels in the liquor; and in the established performance potential of the new cell.

7.2.1 Solvent Extraction Circuit

Minproc Engineers Pty.Ltd., a consulting group well experienced in the relevant field, were commissioned by Euralba to prepare a preliminary study of process requirements and capital/operating costs for an SX/EMEW (solvent extraction/Euralba Mining ElectroWin process) facility at Mount Lyell.

In completing that study, Minproc have relied on 'shake-out' tests conducted by Henkel Corporation (supplier of solvent extraction chemicals) on the waste liquor and on a careful analysis of the EMEW cell capabilities.

The isotherm calculations provided by Henkel are supplied here as Appendix 2. The Minproc study, which is relied upon for the following details, is available under separate cover if required.

The facility was initially sized for 330m³ per hour throughput, approximately 10% less than the projected total volume in the two major waste streams. Adaptation to slightly higher volume throughput can be readily achieved, the only effect being a slight rise in capital cost.

The designed circuit contemplates 90% copper recovery from two extraction stages and one strip stage. It has been conservatively designed, using the traditional mixer-settler systems developed in the 1960-1970's, and has been formulated from first principles using the design criteria outlined in Table 4.

Minproc have provided a preliminary flow chart for the operation, presented here as Figure 34.

7.2.2 Electrowin Circuit

The electrowin circuit required to match this plant would be small, modular and comprised of the following (target production = 1.2 tonnes Cu per day):

Pumping

Provided for in Minproc SX plant design - no further provision required.

TABLE 4 : DESIGN CRITERIA, MT LYELL SOLVENT EXTRACTION PLANT

FACTOR	UNIT	CRITERION
Solution feed rate	m^3/hr	330
Plant availability	% overall	95
days/year		346
hours/day		24
days/week		7
PLS - typical liquor analysis		
Cu	g/l	0.150
Total Fe	g/l	2.0
H_2SO_4	g/l	0.1
Si		* as SiO_2
TSS	ppm	*
pH		2.55
Temperature	°C	*
Viscosity	cP	*
Extractant		
Type		Salicylaldoxime and modifier
SG	kg/l	0.91 - 0.97
Viscosity	cP	200 - 45 (15-30oC)
Diluent		
Type		High flashpoint kerosene
SG	kg/l	0.8
Viscosity	cP	1.5 (15°C)
Flashpoint	°C	80 (100 kPa)
Aromatic content	%	20 max
Organic		
Volume % extractant	%	2
Loading	g/Cu/Vol %	0.50
Cu:Fe selectivity		500:1 (design)
Org.entrainment	ppm organic cts	70
	ppm aqueous cts	100
Aqu.entrainment	ppm organic cts	300
	ppm aqueous cts	250
Org. entrainment in raffinate recovery		baffles in pond
Aqu. entrainment in L.O. recovery		baffles in tank
Diluent evap. open	mm/day	2
closed	mm/day	0.1
Crud removal - suction pump into coffer dam		

TABLE 4 : Continued

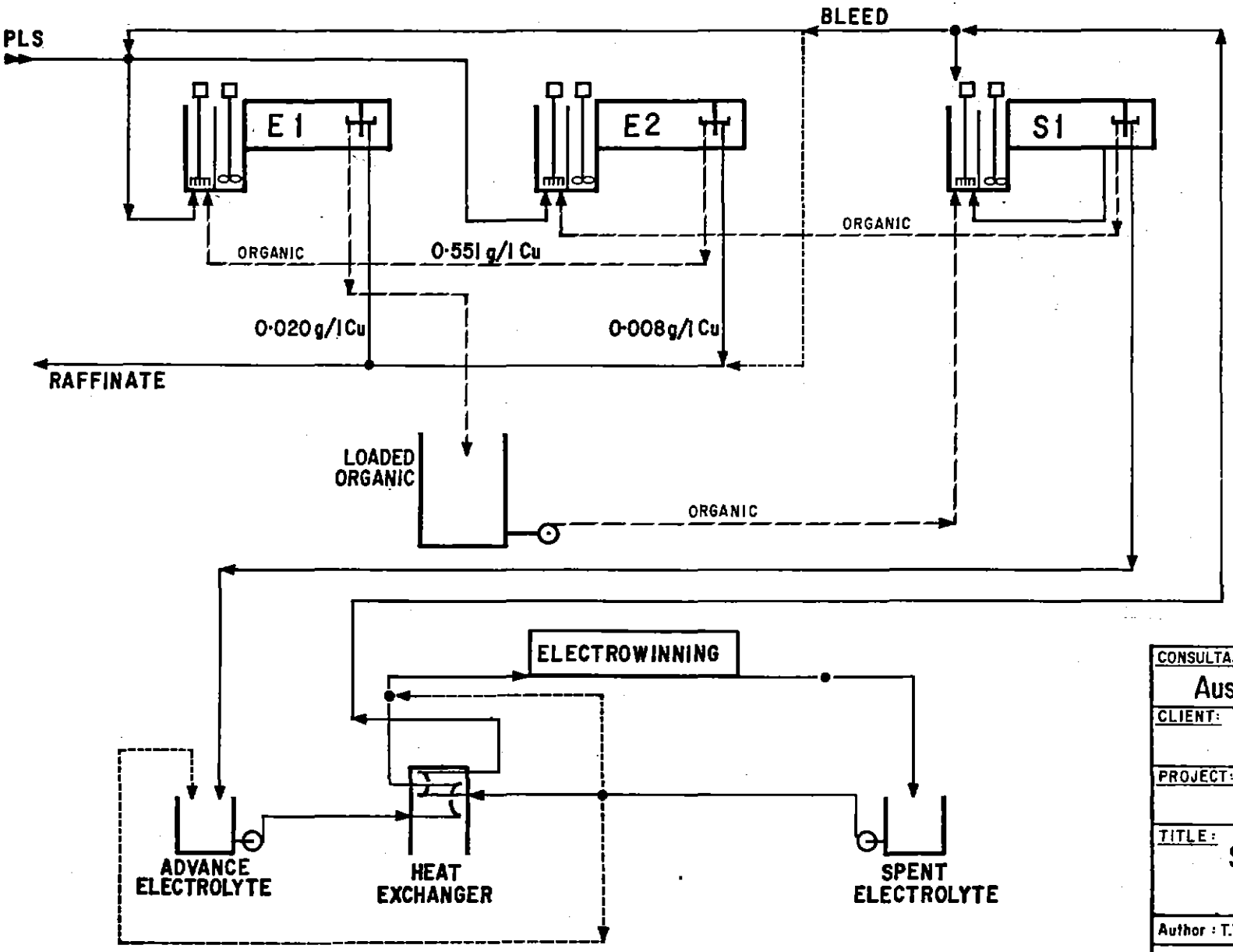
Electrolyte

Copper delta		g/l	20
Spent electrolyte:			
	Cu	g/l	30
	H_2SO_4	g/l	180
	Total Fe	g/l	0.5 (design)
	Co	g/l	*
	Cl	ppm	*
	Advance Cu	g/l	50

SX configuration

Continuous	E_1		aqueous
	E_2		aqueous
	S_1		organic
O/A ratio	E		1:1
	S		1:1
O/A advance ratio	E		1:1
	S		99:1
Stage efficiency	E_1	%	87
	E_2	%	95
	S_1	%	40
Overall Cu extraction		%	91
Stages -	extract		2 in parallel
	strip		1
Addition point spent electrolyte			Strip first mixer
Mixer retention time	E	sec	180
	S	sec	120
Settler rating		$m^3/h\ m^2$	5
Fire protection			water deluge

* = not available at time of study



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CONSULTANCY: Australian Geoscientists Pty Ltd		
CLIENT: Euralba Mining Limited		
PROJECT: Mt. Lyell		
TITLE: SOLVENT EXTRACTION FLOW CHART		
Author : T.Treasure	Date : August 1992	Dwg No:
Drawn: Down Under	Scale :	FIGURE : 34

Electrowin cells

Cell dimensions - diameter	:	80mm		
length	:	2000mm		
Current efficiency	:	95%		
Current density	:	400 A/m²	or	600 A/m²
Production per cell (g/hr Cu)	:	225		336
Number cells required	:	225		148

The cells can be hydraulically connected either to achieve target depletion of the liquor in a single pass, or as a recycle circuit.

Rectifier

AC-DC Transformer @ 85% projected efficiency : 150 kilowatt

Neither of the two process steps requires substantial civil works, beyond concrete footings. Cover is limited for the solvent extraction plant; and the electrowin circuit would occupy an area of only approximately 5 x 5 metres.

In addition to the inherent simplicity of the EMEW system, the study undertaken has suggested that several features of the combined SX/EMEW process are of particular note:

* The capacity of the cell to maintain high efficiency, in winning copper down to low grades even at high current density, allows a wider delta copper range (the difference between grade of feed to the tankhouse and liquor returned to the organic strip stage) than in a conventional tankhouse. As a result 'free' acid concentration in the 'spent electrolyte' is higher, offering greater efficiency in stripping and concomitant circuit simplification.

* As well as the cell being modular and requiring limited civil works (in comparison with a conventional tankhouse) its proven capability to operate at high efficiency, at significantly greater current density than is usual, has a direct impact in reducing the cathode area required to achieve a specified production rate.

* There is no requirement in the circuit for elaborate organic scavenging from advanced electrolyte. In a conventional tankhouse, any residual organic phase collected concentrates at the electrode 'water-line', and burns onto the cathode surface. In the EMEW cell, the closed turbulent system simply returns any organic content back to the stripping mixer settler feed.

* Expansion of a facility at Mt Lyell could be rapidly achieved, on a modular basis, in both solvent extraction and electrowin steps.

* Liquor reticulation requirements for the electrowin circuit are simplified in comparison with conventional tankhouses.

Other pertinent features, not necessarily specific to use of the process in conjunction with solvent extraction, include:

- No lead contamination of cathode
- No acid mist emission
- Reduced likelihood of 'hot spots' and short circuits

All of these factors contribute to the view that combining solvent extraction with the new electrowin technology is an appropriate method for treatment of the Mount Lyell waste streams. Process characteristics of the new cell lead to a significantly lower capital cost than could be contemplated using a conventional SX/EW system.

Unlike the direct electrowin route, it can be confidently argued that such a treatment system would not require further piloting on site. Solvent extraction is a well established process for upgrading of low grade solutions, whose performance on this liquor is confidently predicted by both Henkel and Minproc. The attached electrowin facility would be operating under conditions that are well tested. It is a comparatively small facility, which is well within the current engineering and mechanical status of the technology.

If required, piloting would comprise operation of a 'mini' solvent extraction circuit in conjunction with a few cells.

8.0 PRELIMINARY FINANCIAL ANALYSIS

Both possible treatment circuits are subjected to preliminary financial analysis below. The data provided on the solvent extraction circuit is extracted from the study undertaken on the company's behalf by Minproc; and is based on that company's experience in planning, fabrication and installation of such facilities.

Capital costing of the direct electrowin route is based on modular costs that have been established over two years development of the technology. Operating costs are based on the results of the recent site investigation.

8.2 CAPITAL COST PROJECTIONS

8.2.1 Direct Electrowin

The principal cost in the direct electrowin route lies in the fabrication of cell end caps and purchase of anodes. It is possible to make a reasonable estimate of these costs on the basis of individual cell manufacture, given the knowledge built up through development of the technology.

a) CELL FABRICATION

The cost in an individual cell assembly is built up of the following components:

End Caps

Current manufacturing cost of end caps is relatively high, as they are fabricated through complex machining of commercially available PVC components. The number of units required in this facility would, however, justify (and require) their manufacture through injection moulding.

Preliminary estimates of manufacture by this method indicate a unit cost of between $5.00 and $7.00.

A final cost projection of $15 per unit would provide for manufacture of dies; and for any ancillary fittings to the caps (bringing cost per cell to $30.00).

A plant of this size would justify engineering and fabrication of common end caps for multiples of cells (probably 20). The result would be simplification of reticulation and electrical wiring, but probably not significant decrease in the above cost.

Anodes

Again, the anodes currently in use are expensive items (equivalent to approximately $90 per meter, being titanium with iridium/yttrium oxide coating) - designed to operate at very high current density. The low settings proposed for this route would allow significantly cheaper materials to be used.

The cell proposed for the direct electrowin route is 100 mm diameter, with an electrode gap of 15mm. The anode required would therefore be 70 mm in diameter. The ratio of cathode:anode surface area being approximately 1.3:1, maximum operating anode current density would be 93 A/m^2 (at cathode current density 70 A/m^2).

An investigation of available materials and coatings for 'lighter' duty anodes has yet to be conducted, but it is considered probable that their cost would be less than 50% of the ones currently in use.

A preliminary estimate of $90 per cell is therefore conservatively placed on the anode cost per cell.

Cathodes

The cost of cathodes is dependant on the type of material used and the quantity entailed in a single purchase. The former is governed partially by the nature of product from electrowinning.

In all projects examined to date, where plate is the desired product, it has been proposed that the cathode be commercially available copper pipe. This mode of operation envisages harvesting and sale of the entire cathode, on its reaching desired weight. The cost of the electrode therefore becomes an operating rather than capital item.

Preliminary experimentation has shown that stainless steel cathodes could be used as a starter 'tube', from which copper plate could be mechanically stripped for sale. However, as the technique has not been developed on a production scale cell, use of a copper cathode would remain the current preferred choice.

Direct electrowinning on the Mt Lyell liquor has been modelled on the basis of powder, rather than plate, production. In this case, it would be proposed that 316 stainless steel cathodes be used.

'Off-the-shelf' cost of stainless steel tube is relatively high- approximately $40 per metre (as opposed to copper at approximately $10 per metre). However, as has been found to be the case with purchases of copper tube, bulk 'trade' ordering can result in an approximate halving of this price.

A projection of $40 per 2 metre cell is therefore provided as a capital item in this study.

Inter-cell connections

The cells are readily connected in series utilising flexible hose, commercially available at approximately $3 per metre.

Electrical connection between cells entails simple wiring, of relatively low current rating (maximum 50 Amps). Being closely juxtaposed wiring of the cells does not require excessive cable lengths (for the purpose of this study assumed at 0.6m per cell, at a cost of around $3 per metre).

Assembly of the cells into multiples with common end caps would reduce costs in connections. For this study, however, a combined cost of $5 per cell is assumed.

Mounting

The cells would be mounted (in multiples) on racks, unless the system is modularised into units of 20, in which case it is expected that the modules would be free-standing.

A provision of $10 per cell would more than adequately cover the cost involved.

Total cell cost

Total cell costs in the model presented are therefore as follows:

	Per unit $	per cell $	Plant $ (8,600 cells)
End caps	30	60	516,000
Anode	90	90	774,000
Cathode	40	40	344,000
Connections	5	5	43,000
Mounting	10	10	86,000
Totals		205	1,763,000

Any savings possible in fabrication costs would impact significantly on this projection - for example, a saving of $50 per cell (25% of its projected cost) would reduce capital expenditure by close to $0.5 million.

b) PUMPING

Costs of pumps in this circuit would depend on the final reticulation pattern chosen. Preliminary estimates are however possible on the basis of the data generated.

A single pump capable of delivering 25% of the flow envisaged (6,000 l/min) at the pressure required (up to 80 psi) has been costed at approximately $5,000.

If a single pump were utilised, its unit value has been estimated at approximately 3 times this figure. Provision of a back-up unit would double projected expenditure.

On this basis a capital projection for pump purchase is placed at $40,000.

c) RECTIFIER

The price of rectifiers is largely governed by the degree of internal circuit complexity (controllability) and the level of control over grid current variations desired.

On the basis of a unit currently being installed by GEC in New South Wales ($1.4 million, 4 mW capacity), price of a new rectifier for this project (1 mW) would be in the region of $350,000.

Discussions with manufacturers have, however, indicated that second hand rectifiers are readily available - at a price of equal to or less than $200,000 for the size required.

d) OTHER

Further provisions are made in the estimate for the following:

a)	Liquor storage and management	$100,000
b)	Rectifier shed, working area and cover to cells	$ 50,000
c)	Plant reticulation	$ 25,000
d)	Product harvesting equipment	$ 75,000

An estimate of total capital costs in installing a direct electrowin plant of the size required at Mount Lyell is presented in Table 5 below:

Table 5 : SUMMARY OF PROJECTED CAPITAL COSTS
MT LYELL DIRECT ELECTROWIN PLANT

Item	Projected $
Electrowin cells	1,763,000
Pumps	40,000
Rectifier	200,000
Liquor storage/management	100,000
Product harvesting	75,000
Building	50,000
Plant reticulation	25,000
	2,253,000
Contingency	225,000
Design, project management etc	100,000
	2,578,000

8.2.2 Solvent Extraction/EMEW System

a) Solvent Extraction Circuit

Minproc have provided a budget price for the solvent extraction circuit of A$1.23 million.

The price includes all necessary electrical, piping and instrumentation costs, plus all equipment purchased as proprietary items or fabricated to order; and its installation.

Indirect cost of engineering design, project management and procurement is included, plus a 10% contingency allowance. The costs do not allow for the civil works required to form the plant site, as these may already be available in the form of a flat site with concrete pads.

The accuracy of the estimate given is stated to be +/- 15-20%.

The capital costs were derived from data compiled during a 1992 tender for a conventional SX-EW plant. Inter alia, they are based on the following equipment schedule and itemisation of capital costs:

TABLE 6 : PRELIMINARY SOLVENT EXTRACTION EQUIPMENT LIST

Item	No.	No. of drives	kW	Total kW
Extraction mixer settler	2	2	22	44
		2	11	22
Loaded organic tank 4m ø x 6m	1	-		
Loaded organic pump	1	1	7	7
Stripping mixer settler	1	1	22	22
		1	11	11
Advance electrolyte tank 2m ø x 3m	1	-		
Advance electrolyte pump	1	1	1.1	1.1
Spent electrolyte tank 1.5m ø x 2m	1	-		
Spent electrolyte pump	1	1	1.1	1.1
Heat exchanger	1	-		
				107.2

TABLE 7 : CAPITAL COST PROJECTIONS SX EQUIPMENT

Item	Cost (A$)
Diluent tank	Supplied by diluent supplier
Mixer Settler E_1	190,000
Mixer Settler E_2	190,000
Mixer Settler S_1	220,000
Loaded organic tank	40,000
Loaded organic pump	30,000
Heat exchanger	20,000
Spent electrolyte tank	30,000
Spent electrolyte pump	20,000
Piping and valves	115,000
Electrical	125,000
Total direct cost	1,030,000
10% contingency	103,000
Design, project management etc	97,000
TOTAL SX COSTS	1,230,000

b) Electrowin Circuit

Capital cost projections for the electrowin portion of the SX/EMEW facility, are based on the following general items:

Pumping - Provided for in SX circuit
EW cells - Maximum number 225 cells
Rectifier - 150 kW unit

Cost of fabrication of individual cells will be similar to that calculated for the direct electrowin system, aside from a requirement for more expensive, heavy duty anodes (significantly higher operating current density) - as follows:

Item	Unit	$ per cell
Caps	$30 ea.	60
Anode	$90/m	180
Cathode	$15/m	30
Connections		10
Mounting		10
Total		290
Total for 225 units		$65,250

The model provides for use of copper cathodes, the first set of which are included in the capital estimate.

Cost of the rectifier required can only be estimated at this stage, but a projected figure of $60,000 is in accordance with discussions with potential suppliers.

Civil works required for this small plant are limited - to workshop/rectifier shed and cover for the cells. Reticulation outside of that provided in the SX circuit is minimal. A provision of $15,000 is considered sufficient to cover these requirements.

Projected costs in the electrowin circuit are summarised as follows:

Item	$
Cell fabrication	65,250
Rectifier	60,000
Civils etc	15,000
	140,250
Contingency	14,750
Total	155,000

c) Combined SX/EMEW Capital Projections

Although the use of solvent extraction will simplify liquor management on site, a provision of $100,000 is made for this facility (similar to that for the direct electrowin route).

The 'first fill' of reagents to the SX facility should be included also as a capital provision:

-	extractant	2,207 kg	$35,312
-	diluent	118,836 litres	$61,795
-	acid (98%)	+/-16 tonnes	$ 2,044
			$99,151

Thus - total projected capital costs:

Item	$
SX direct costs	1,030,000
EW direct costs	140,250
Liquor management	100,000
First SX fill	99,500
Design/management	97,000
Contingency	117,750
TOTAL	1,584,500

8.3 OPERATING COST PROJECTIONS

Operating cost projections for each of the possible treatment routes are based on calculations provided by Minproc, and on the measurements taken, recorded and analysed above.

It should be noted that the projections are based on the minimum copper production rate calculated for the two main waste streams at Mt Lyell.

8.3.1 Direct Electrowin option

A preliminary estimate of operating costs in a direct electrowin facility is based on the following established data and assumptions:

Daily copper production		:	1.2 tonnes
Grid power cost		:	5 cents/kWh
Electrowin power consumption (see section ? above)		:	18 mWh/tonne Cu
Pump power consumption (see section ? above)		:	2.9 mWh/tonne Cu
Personnel requirement		:	2 labour 1 supervisory
Personnel cost (incl.on costs)	Labour	:	$45,000 p.a
	Supervisory	:	$75,000 p.a

Personnel costs are recalculated to derive a daily projection by dividing the projected annual cost by 320.

Maintenance requirements on the plant should be low, as few moving parts are incorporated. An annual provision of $100,000 is provided.

Projected direct daily operating costs are summarised as follows:

Item	$/day
Electrowin power	1,080
Pump power	145
Labour	280
Supervision	240
Maintenance	275
Total	2,020

= $1,683 per tonne Cu

It is noted that the above table does not make provision for administration, product transport or government royalties.

8.3.2 SX/EMEW Option

Detailed mass balance calculations were performed by Minproc to achieve an estimate of consumables (reagent cost) in the solvent extraction circuit. This cost has been projected at $0.36/kg Cu produced, on the basis of the following table:

	$/year	$/kg Cu	Recover 50% of organic
Diluent	148,039	0.40	0.20
Extractant	84,594	0.23	0.11
98% acid	18,343	0.05	0.05
	250,976	0.68	0.36 $/kg Cu

Projected operating costs are based upon this figure, coupled with the following assumptions and calculations:

Daily copper production		: 1.2	tonnes
Grid power cost		: 5	cents/kWh
Electrowin power consumption		: 2.1	mWh/tonne Cu
SX power consumption (at 70% of total installed power)		: 1.8	mW/day
SX consumables		: 0.36	$/kg copper
Personnel requirement		: 4	labour
		1	supervisory
Personnel cost (incl.on costs)	Labour	: $45,000 p.a	
	Supervisory	: $75,000 p.a	
Plant maintenance		: $50,000 p.a.	

Due to the higher projected labour component in this route, resultant daily costs are calculated on a 346 day annual factor.

Projected direct daily operating costs in the combined SX/EMEW system are summarised as follows:

Item	$/day
Electrowin power	126
SX power	90
SX consumable	432
Labour	520
Supervision	217
Maintenance	137
Total	1,522

= $1,268 per tonne Cu

The schedule does not include the potential operating cost entailed in continuing supply of copper tubing as cathode material. Such a cost would reflect the difference between purchase price and value of the contained copper (which would be recovered on sale of product) - i.e. a premium for fabrication.

This cost is not provided as this stage, because:

a) It is possible that a favourable arrangement can be made with a manufacturer, entailing 'toll' production of tubing from the project's own product, and

b) Depending on the timing of the project, plans for use of stainless steel tubes (as 'starters' from which the copper plate can be stripped) may be further progressed.

At a cost of copper tubing of around $10 per metre and a target cathode loading of 50kg per cell, the resultant manufacturing premium (the initial cell being approximately 5 kg in weight) would be equivalent to approximately $100 per tonne of product.

Again, it should be noted that the above table does not make provision for administration, product transport or government royalties.

8.4 PROJECTED OPERATING SURPLUS VS CAPITAL COST

Projections of potential operating surplus in the two treatment operations assume 365 days/year operation - a mode that will suit the nature of the source (as a continually running waste stream); and would minimise requirement for liquor storage dams.

In reflecting lowest potential copper production from the two main waste streams, the calculation represents minimum projected return.

Increased production rates (although possibly entailing additional capital cost) would allow significant economies of scale, especially with regard to personnel costs.

The projections provided below assume a discounted value for product, against current copper value around A$3,500, as follows:

Copper powder	:	90%
Cathode	:	95%

The discounts are applied as, although quality of product is projected to be extremely high (especially for cathode), final specifications can not be determined at this stage. Dependant on early operation, full LME price could be expected for cathode. Dependant on purity, powder produced could achieve a significant premium.

No formal sensitivity analyses are presented here, but observations on the potentially adverse effects of changes in major factors (copper price and power cost) are presented in a later comparison of the two treatment systems.

The analysis therefore indicates the following potential annual surplus for each of the treatment options (as a calculated amount and as a percentage of projected capital cost):

Direct Electrowin Option	$
Projected annual sales	1,379,700
Treatment costs	750,075
Surplus	629,625
Capital expenditure	2,587,000
Annual surplus as % of capital	24.3 %

SX/EMEW Option	
Projected annual sales	1,456,350
Treatment costs	555,530
	900,820
Capital expenditure	1,584,000
Annual surplus as % of capital	56.9 %

Both treatment routes offer, therefore, significant potential return on capital invested.

The SX/EMEW option however, in addition to presenting an opportunity for more than doubling potential rate of return, has a number of process and financial advantages over direct electrowinning. Not least of these is a significant difference in sensitivity to the cost of grid power at Mt Lyell.

9. COMPARISON OF TREATMENT OPTIONS (Process and Financial)

Observations on respective performance of each treatment route for Mount Lyell have appeared continuously in the above discussion. The following is a summary of the key characteristics, both process and financial, which lead naturally to a preference for the SX/EMEW option.

9.1 PROCESS

The chemical 'upgrading' possible through a simplified solvent extraction step has been demonstrated to have the following results:

* The effect of iron on electrowin efficiency is almost entirely obviated.

* Due both to the above and to the higher concentration of copper in the electrowin liquor, high current efficiency can be maintained at high to very high current density.

* The direct consequence of resulting:

  a) increase in current efficiency by a factor of around 8
  b) increase in current density between 6 and 10 times

  - has a multiplier effect on the number of cells required to achieve the desired production rate (by a factor of approximately 40).

* The resultant electrowin circuit is sized within current engineering status of the technology.

* At the expense of limited additional power requirement in the solvent extraction circuit, power consumption in the plant is reduced by a factor of at least 6.

* As increased copper production capability would be largely through increase in the number of electrowin cells, the SX/EMEW route (with 40 times less units) will lend itself more readily to expansion.

* The solvent extraction circuit is sized to match liquor throughput rather than liquor tenor. As expansion is more likely to arise through increase in copper grade, rather than liquor volume, the circuit has significant advantage with respect to the capital works required.

* It is conceivable that significant variations in liquor availability would be more easily handled in the solvent extraction circuit rather than direct electrowin.

* Conditions in the electrowin circuit will remain constant, as opposed to the possible requirement for change in operating parameters (to reflect changing liquor composition) in the direct electrowin plant.

The EMEW system has a significant further advantage in its potential for significant improvement in the operation of a solvent extraction circuit. Such advantages have been summarised in section 7.2 above and do not require elaboration here - aside from noting that they lead to a more compact plant than would be necessary with conventional electrowinning at Mt Lyell. It is noted that studies into the latter have been conducted in the past.

9.2 FINANCIAL

There are a number of obvious financial reasons for the preference for the SX/EMEW route:

a) Reduced capital cost
b) Reduced operating cost
c) Reduced cost of expansion

These obviously have direct effect on projected operating returns with respect to capital investment, but they impact significantly also in analysing project sensitivity to various factors:

a) CHANGE IN COPPER PRICE

Reduction copper value to $3,000 per tonne has the following effect:

Direct electrowin surplus reduced to	:	$ 432,525
% vs capital reduced to	:	16.7%
SX/EMEW surplus reduced to	:	$ 690,945
% vs capital reduced to	:	43.6%

Although the reduction in return (as a percentage of the expected scenario above) is not significantly different, the reduction in return increases significantly the 'risk' level in the direct electrowin plant.

Given that projected capital and operating costs remain constant, break even value for copper (on a 'cash' basis without provision for capital) in the two scenarios is equivalent to the calculated 'cost' of production indicated:

Direct EW	:	$ 1,680
SX/EW	:	$ 1,286

b) INCREASED POWER COST

Power cost in the direct EW option exceeds that of SX/EMEW by an amount approximately equivalent to $830 per tonne of copper produced.

A 50% increase in projected power cost (5 to 7.5 cents/kWh) would increase daily operating costs in the two options as follows:

Direct electrowin	:	+ $615
SX/EMEW	:	+ $175

The result for the former would be a decrease in apparent return on capital of nearly 9%. Coupled with a fall in copper price, such an increase in power cost would make the operation marginal.

The SX/EMEW system is obviously far more robust under adverse power cost conditions, adding therefore to the overall strength of the projections.

c) INCREASED CAPITAL COSTS

Both systems are relatively resistant to unforeseen increase in capital cost. Under the model provided in section 8.2 above, increase in capital cost by $0.5 million would result in reduction in apparent return (%) per annum of about 20%.

Naturally, however, increases in capital cost will result in greater sensitivity to other factors.

A significant increase in copper grade in the liquor to be treated would probably have a beneficial effect on direct electrowinning- through a concomitant increase in current efficiency. Expansion of the plant to suit a higher grade would require a lesser number of cells to achieve the same result.

Even so, the capital cost of expansion would significantly exceed that in the SX/EMEW plant.

The solvent extraction facility would not require significant expansion, as it is sized to the volume of liquor throughput. Production from the electrowin section could be expanded through increasing current density to the cells, or increasing the number of units installed. In the latter case, the incremental number that may be required is would be relatively low.

From both process and financial angles therefore the SX EMEW route offers greater flexibility and increased greater resistance to adverse changes in operating conditions.

10.0 CONCLUSION AND RECOMMENDATIONS

Detailed analysis of the operating performance of the new technology suggests that both of the treatment routes examined offer opportunity for commercial extraction of copper from the Mount Lyell waste streams.

It is clear however, from both technical and financial standpoints, that the SX/EMEW route offers greater flexibility and greater resistance to adverse changes in operating costs and conditions. Among the more obvious advantages in this option is its intrinsically lower sensitivity to increased power costs.

Solvent extraction is a well developed technology of proven capability to perform the liquor treatment required. In applying the new technology, the number of cells required is dramatically reduced, to a degree well within the limits of its current development status.

As its design is currently modelled on the basis of the selection of operating parameters from individual test runs, use of the direct electrowin route at Mount Lyell would require further piloting (continual treatment of, say, 5% of the waste stream).

It is suggested, due to established 'state of the art' in both sections of the SX/EMEW system, that no further piloting is required for this option. Requirement for further technical study is considered to be limited to further 'shake-out' solvent extraction tests on the liquor; and to further operation of a single production electrowin cell on solution of the indicated chemical composition.

It is considered that the establishment of a conventional SX/EW treatment plant at Mount Lyell would be prohibited by high capital cost - especially at the low levels of copper production contemplated in the initial operation.

The current study, coupled with work on other projects, has shown that capital cost of an electrowin circuit is markedly reduced using the new cell. Coupled with simplification achievable in the solvent extraction circuit, this results in the significantly lower projected capital cost, which is shown to be ably supported by operating returns even at low copper output.

It is recommended, therefore, that development of the project proceed with immediate implementation of a detailed feasibility study on the SX/EMEW treatment option.

Early implementation of this plan is recommended in the light of the long term potential of a copper leach/extraction programme at Mount Lyell. Based initially on limited production from the two major current sources, it is envisaged that significant expansion could be achieved through a number of routes (on which the proposed closure of the mine in 1996 would have little impact).

294111

APPENDIX 1

Site Photographs

294112

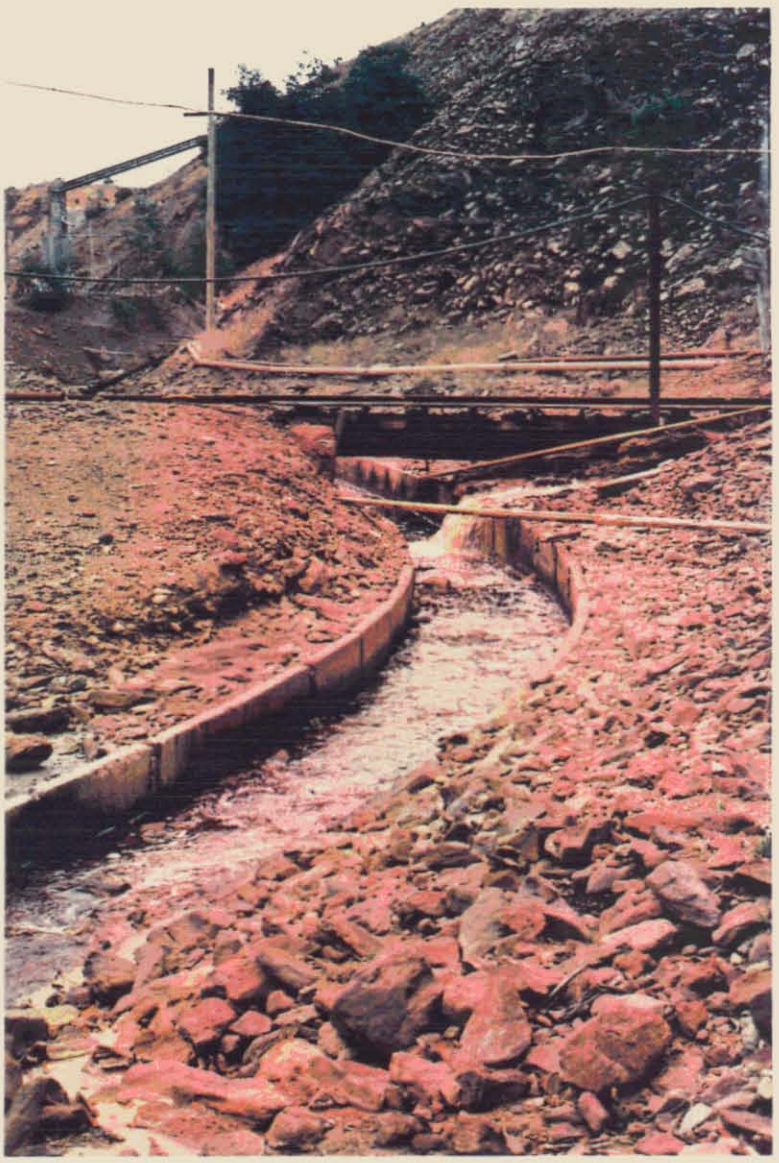
PHOTOGRAPH 1 : West Lyell Open Cut

294113

PHOTOGRAPH 2 : Dumps from West Lyell Open Cut

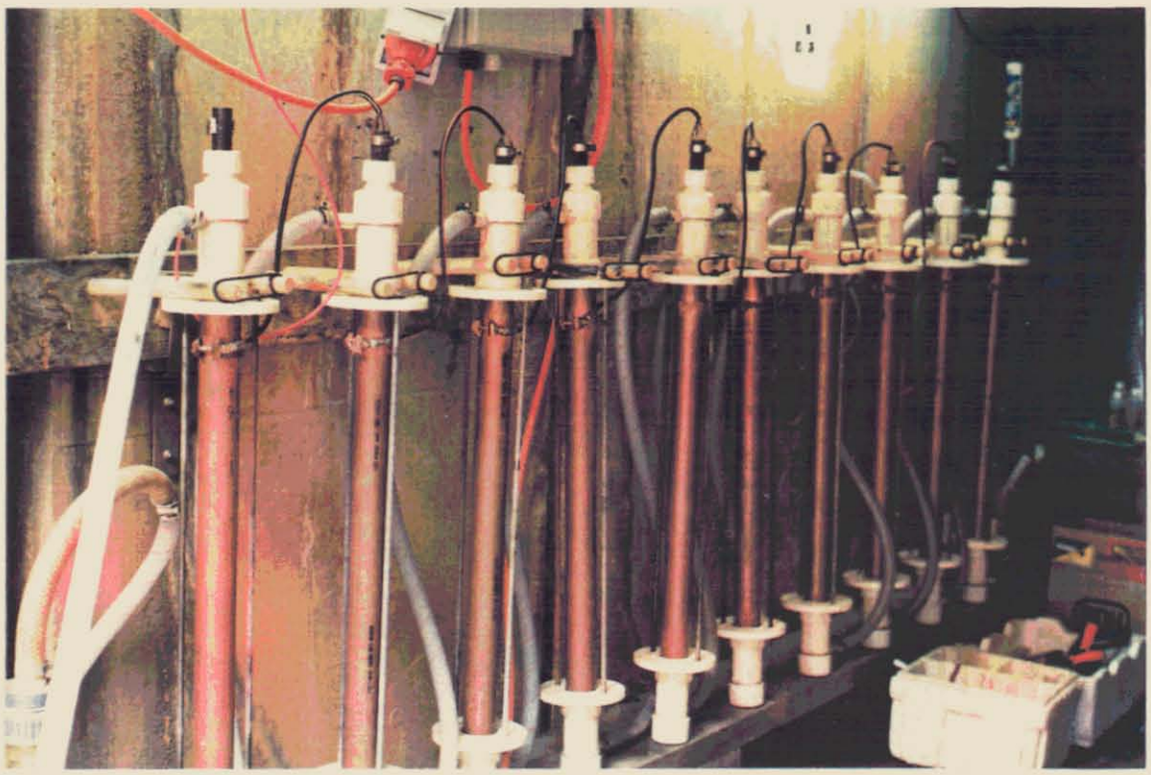
294114

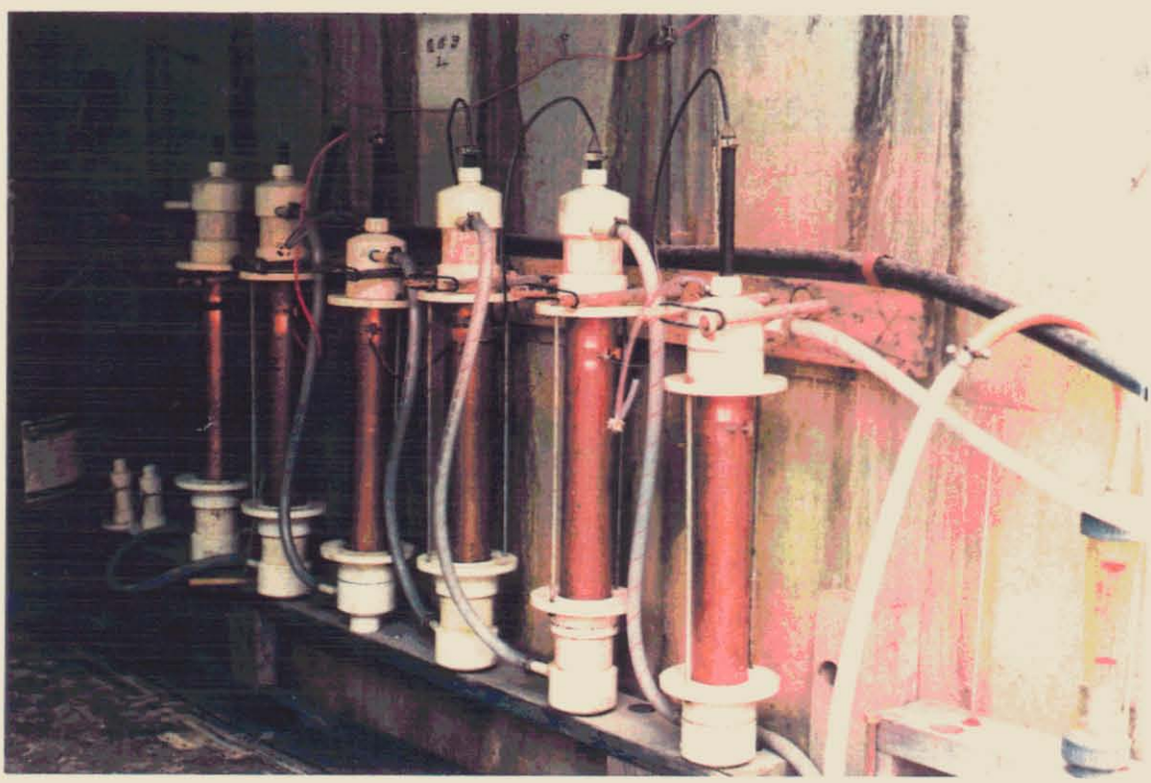
PHOTOGRAPHS 3 AND 4 : Waste Liquor Stream - Locality 5



PHOTOGRAPHS 5 AND 6 : 2 Inch and 4 Inch Trial Circuits

294115





APPENDIX 2

Henkel Isotherm Calculations

ISOCALC 1 PROGRAM - HENKEL CORP. - Cu/H_2SO_4
ISOTHERM POINT CALCULATIONS AND GRAPHICS/McCABE-THIELES

Reagent selected : LIX 984
Volume pct. of reagent selected : 2.00
Copper content of the spent electrolyte, g/l Cu: 30
Copper content of pregnant electrolyte, g/l Cu: 50
Number of strip stages: 1
H_2SO_4 in spent electrolyte, g/l: 180.0
Copper content of stripped organic, g/l Cu: .408
Copper content of aqueous solution tested, g/l Cu: 0.150
pH of solution tested: 2.55
Total sulphate content of solution tested, g/l: 10
Assumed activity of th SO_4= ion, %: 28
Other parameters of circuit: 1E X 1E X 1S

The MAX LOAD for 2.00 pct. LIX 984 is 0.994 g/l Cu.

O/A = 10.0 : ORG.= 0.422 g/l Cu; AQ.= 0.002 g/l Cu
O/A = 5.0 : ORG.= 0.437 g/l Cu; AQ.= 0.003 g/l Cu
O/A = 2.0 : ORG.= 0.481 g/l Cu; AQ.= 0.003 g/l Cu
O/A = 1.5 : ORG.= 0.505 g/l Cu; AQ.= 0.004 g/l Cu
O/A = 1.0 : ORG.= 0.553 g/l Cu; AQ.= 0.005 g/l Cu
O/A = 0.5 : ORG.= 0.687 g/l Cu; AQ.= 0.010 g/l Cu
O/A = 0.2 : ORG.= 0.898 g/l Cu; AQ.= 0.052 g/l Cu
O/A = 0.1 : ORG.= 0.956 g/l Cu; AQ.= 0.095 g/l Cu

O/A ratio in extraction: 1.0
Mixer efficiency for EP-1 extraction = 98%, EP-2 extraction = 93%

EP-1 and EP-2 extraction stage profiles as follows (g/l Cu)

Aq. Feed	---> EP-1 Aq.(Raff.)	Aq.Feed	---> EP-2 Aq.(Raff.)
0.150	0.020	0.150	0.008

(L.O.) EP-1 Org.	<---------- EP-2 Org.	<---Stp.Org.
0.681	0.551	0.408

The expected recovery: EP-1 = 86.4%, EP-2 = 94.9%, OVERALL = 90.65%

The loaded organic is 68.51% of max load.
The net transfer is 0.137 g/l per 1 vol.% of LIX 984
The strip O/A = 110/1 using 30 g/l Cu in the spent electrolyte.

Materials balance across the strip circuit as follows (g/l Cu)

Load.Org.	------> S-1 Org.
0.681	0.408

S-1 Aq.(P.E.)	<-----Spent Elec.
50.000	30.00

APPENDIX 3

Individual Test Data Tables

HIGH GRADE LIQUOR TESTS

TEST NO : ML4

Sample Volume (l) : 10
Head grade (Cu ppm) : 37920
Liquor flow (l/hour) : 350
Cathode diametre (mm) : 35
Cathode length (mm) : 200
Masking (mm) : 50
Anode diameter (mm) : 9
Number cells : 1
Amps - per cell : 8.5
- circuit : 8.5
Current density (A/m2): 515

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe (ppm) total	pH	Cu rec period	Current eff. period	Current eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
01-Mar	0.51	0.00	5.2	8.5	37920	200	2.4					
01-Mar	0.59	2.03	4.4	8.9	35540	200	1.1	23.8	114	114	3.56	3.56
01-Mar	0.75	5.78	3.7	8.5	31940	200	0.79	36.0	94	101	3.68	3.64
02-Mar	0.08	13.87	3.2	8.0	24360	200	0.58	75.8	96	98	3.04	3.30

TEST NO : ML 5

Sample Volume (l) : 10 (= resid liquor #4)
Head grade Cu ppm) : 24200
Liquor flow (l/hour) : 350
Cathode diametre (mm) : 35
Cathode length (mm) : 200
Masking (mm) : 50
Anode diameter (mm) : 9
Number cells : 1
Amps - per cell : 6.5
- circuit : 6.5
Current density (A/m2): 394

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe (ppm) total	pH	Cu rec period	Current eff. period	Current eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
04-Mar	0.52	0.00	2.7	6.5	24200	200						
04-Mar	0.75	5.45	2.5	6.8	23500	200	0.78	7.0	15	16	13.44	13.44
05-Mar	0.08	13.48	2.3	6.3	16300	200	0.80	72.0	116	76	1.75	2.79
Off for inspection												
05-Mar	0.47	13.48	2.2	6.5	17600	200	0.76					
05-Mar	0.94	24.85	2.3	5.8	5820	200		117.8	143	105	1.33	1.91
Off for inspection												
06-Mar	0.50	24.85	2.2	6.8	5720	200	0.90					
06-Mar	0.67	29.00	2.1	6.3	2560	200		31.6	99	104	1.82	1.90
06-Mar	0.75	30.87	2.5	7.2	1000	200		15.6	105	104	1.85	1.90

294120

TEST NO : ML 10

Sample Volume (l) : 10
Head grade (Cu ppm) :
Liquor flow (l/hour) :
Cathode diametre (mm) : 35
Cathode length (mm) : 200
Masking (mm) : 50
Anode diameter (mm) : 9
Number cells : 1
Amps - per cell : 2
- circuit : 2
Current density (A/m2): 121

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe (ppm) total	pH	Cu rec period	Current eff. period	Current eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
08-Mar	0.50	0.00	2.3	2.0	980	200	0.38					
08-Mar	0.61	2.63	2.2	2.0	460	200	0.44	5.2	84	84	2.28	2.28
Off for inspection												
08-Mar	0.63	2.63	2.2	1.5	460	200	0.44					
08-Mar	0.73	5.22	1.7	1.1	158	200	0.59	3.0	76	81	2.16	2.23

TEST NO : ML 15

Sample Volume (l) : 10
Head grade (Cu ppm) :
Liquor flow (l/hour) : 350
Cathode diametre (mm) : 35
Cathode length (mm) : 200
Masking (mm) : 50
Anode diameter (mm) : 9
Number cells : 1
Amps - per cell : 10.2
- circuit : 10.2
Current density (A/m2): 618

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe (ppm) total	pH	Cu rec period	Current eff. period	Current eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
10-Mar	0.50	0.00	6.0	10.0	40800	200	2.67					
10-Mar	0.60	2.62	4.5	10.5	36400	200	1.62	44.0	139	139	3.20	3.20
10-Mar	0.77	6.68	3.6	10.5	31800	200	1.40	46.0	91	110	3.76	3.49
10-Mar	0.92	10.15	3.4	10.8	28100	320	1.27	37.0	85	101	3.49	3.49
Off for inspection												
11-Mar	0.46	10.15	3.3	10.0	27800	320	1.24					
11-Mar	0.62	13.80	3.0	10.4	24000	200	0.69	38.0	86	97	3.09	3.40
11-Mar	0.79	17.88	2.8	10.4	18500	320	0.69	55.0	110	100	2.24	3.11
Off for inspection												
12-Mar	0.51	17.88	2.7	10.2	18500	320	0.69					
12-Mar	0.60	20.16	2.7	10.3	15100	320	0.59	32.3	117	102	1.96	2.96
12-Mar	0.67	21.71	2.6	10.3	13300	320	0.52	17.1	91	101	2.46	2.93
12-Mar	0.94	28.18	2.6	10.5	3800	320	0.40	90.3	114	104	1.94	2.68
Off for inspection												
13-Mar	0.47	28.18	2.3	10.0	3800	320	0.40					
13-Mar	0.58	30.80	2.6	8.2	524	330		26.2	93	103	2.23	2.65
13-Mar	0.67	32.83	2.4	9.0	246	330	0.12	2.2	11	98	19.66	2.75

294121

TEST NO : ML 22

Sample Volume (l)	:	10
Head grade (Cu ppm)	:	39800
Liquor flow (l/hour)	:	350/500
Cathode diametre (mm)	:	35
Cathode length (mm)	:	200
Masking (mm)	:	50
Anode diameter (mm)	:	9
Number cells	:	1
Amps - per cell	:	15
- circuit	:	15
Current density (A/m2)	:	909

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe (ppm) total	pH	Cu rec period	Current eff. period	Current eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
16-Mar	0.56	0.00	3.0	15.0	39800							
16-Mar	0.75	4.43	2.9	15.0	33300	380	0.63	65.0	83	83	2.99	2.99
Off for inspection												
16-Mar	0.76	4.43	2.7	15.0	33300	380	0.63					
16-Mar	0.99	9.95	2.7	15.0	22300	380	0.64	110.0	113	99	2.03	2.39
Off for inspection												
17-Mar	0.44	9.95	2.7	15.0	22300	380	0.64					
17-Mar	0.75	17.48	2.5	14.0	9900	380	0.64	124.0	96	98	2.29	2.35
Off for inspection												
18-Mar	0.59	17.48	2.5	14.5	9900	380	0.64					
19-Mar	0.02	27.91			1000	380	0.60	89.0				

2 INCH CELL TESTS

TEST NO : ML 6

Sample Volume (l) :	1802
Head grade (Cu ppm) :	91
Liquor flow (l/hour):	1400
Cathode diametre (mm:	54
Cathode length (mm) :	850
Masking (mm) :	0
Anode diameter (mm) :	20
Number cells :	10
Amps - per cell :	5
- circuit :	5
Current density (A/m:	35

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe(ppm) total	Fe(ppm) as Fe++	pH	Cu rec (gm) period	Cu rec (gm) Cumul.	Curr. eff. period	Curr. eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
04-Mar	0.65	0.00	3.05	5.0	91	640		2.69						
04-Mar	0.75	2.30	3.03	5.0	80	660		2.70	19.8	19.8	15	15	17.63	17.63
04-Mar	0.92	6.30	2.95	5.5	63	680		2.72	30.6	50.5	12	13	20.50	19.38
05-Mar	0.08	10.33	2.81	5.0	48	680		2.66	27.0	77.5	11	12	22.58	20.49
05-Mar	0.25	14.30	2.76	5.2	37	680		2.60	19.8	97.3	8	11	28.26	22.08
05-Mar	0.43	18.55	2.69	5.0	24	680		2.51	23.4	120.7	9	11	25.08	22.66
05-Mar	0.60	22.67	2.66	4.9	16	680		2.46	14.4	135.2	6	10	37.80	24.27
05-Mar	0.74	26.15	2.55	4.8	11	710		2.05	9.0	144.2	5	9	48.58	25.79

294123

TEST NO : ML 11

Sample Volume (l) : 2650
Head grade (Cu ppm) : 90
Liquor flow (l/hour): 1400
Cathode diametre (mm: 54
Cathode length (mm) : 850
Masking (mm) : 0
Anode diameter (mm) : 20
Number cells : 10
Amps - per cell : 7.5
- circuit : 15 = 2 banks in parallel
Current density (A/m: 52

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe (ppm) total	Fe (ppm) as Fe++	pH	Cu rec (gm) period	Cu rec (gm) Cumul.	Current eff. period	Current eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
10-Mar	0.43	0.00	4.1	7.5	90	640	20	2.56						
10-Mar	0.50	1.78	3.7	7.5	80	650		2.47	26.5	26.5	17	17	19.68	19.68
10-Mar	0.59	3.87	3.5	7.5	70	640		2.44	26.5	53.0	14	15	21.15	20.42
10-Mar	0.75	5.67	3.6	8.8	66	630		2.31	10.6	63.6	6	12	49.44	25.25
10-Mar	0.93	10.02	3.3	7.8	54	650		2.25	31.8	95.4	8	10	39.16	29.89
11-Mar	0.09	13.77	3.1	7.1	50	660		2.29	10.6	106.0	3	8	83.40	35.24
11-Mar	0.28	18.34	3.3	8.3	39	660		2.23	29.2	135.2	7	8	38.18	35.87
11-Mar	0.42	21.70	3.2	7.5	32	650		2.17	18.6	153.7	6	8	46.83	37.17
11-Mar	0.60	26.00	2.9	7.5	23	790		2.25	23.9	177.6	6	7	41.34	37.73
11-Mar	0.77	30.07	2.9	7.5	14	790	280	2.17	23.9	201.4	7	7	36.89	37.63
11-Mar	0.91	33.54	2.7	6.8	11	800		2.21	8.0	209.4	3	7	86.68	39.50
12-Mar	0.09	37.79	2.6	6.0	7	810		2.18	10.6	220.0	3	7	67.73	40.86
12-Mar	0.25	41.79	3.2	8.8	5	800		2.16	5.3	225.3	2	6	161.42	43.69
12-Mar	0.46	46.84	2.8	6.5	7	800		2.15						

294124

TEST NO : ML 18

Sample Volume (l) : 2650
Head grade (Cu ppm) : 130
Liquor flow (l/hour): 1800
Cathode diametre (mm: 54
Cathode length (mm) : 850
Masking (mm) : 0
Anode diameter (mm) : 20
Number cells : 10
Amps - per cell : 7.5
- circuit : 7.5
Current density (A/m: 52

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe (ppm) total	Fe (ppm) as Fe++	pH	Cu rec (gm) period	Cu rec (gm) Cumul.	Current eff. period	Current eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
12-Mar	0.54	0.00	3.5	7.5	130	1300	5	2.50						
12-Mar	0.67	3.22	3.0	7.5	114	1340		2.47	42.4	42.4	15	15	18.22	18.22
12-Mar	0.75	5.05	3.1	7.5	112	1350		2.46	5.3	47.7	3	11	77.83	24.84
12-Mar	0.93	9.25	2.8	7.8	96	1330		2.39	42.4	90.1	11	11	22.28	23.64
13-Mar	0.09	13.22	2.7	7.0	74	1340	235	2.31	58.3	148.4	17	13	13.80	19.77
13-Mar	0.27	17.52	2.9	8.0	64	1340		2.28	26.5	174.9	7	11	33.53	21.86
13-Mar	0.59	19.00	2.9	8.0	60	1360	330	2.24	10.6	185.5	8	11	32.13	22.44
13-Mar	0.74	22.67	2.7	8.0	55	1350	330	2.09	13.3	198.8	4	10	61.99	25.08
13-Mar	0.91	26.75	2.8	8.2	50	1340		2.06	13.3	212.0	3	9	68.65	27.80
14-Mar	0.08	30.73	2.5	7.0	37	1340		2.02	34.5	246.5	10	9	22.98	27.13
14-Mar	0.27	35.30	2.6	7.5	29	1340		1.98	21.2	267.7	5	8	39.04	28.07
14-Mar	0.43	39.13	2.6	7.7	21	1360		1.98	21.2	288.9	6	8	35.59	28.62
14-Mar	0.51	41.00	2.7	7.5	18	1340			8.0	296.8	5	8	47.11	29.12

294125

TEST NO : ML20

Sample Volume (l) : 2650
Head grade (Cu ppm) : 149
Liquor flow (l/hour): 1800
Cathode diametre (mm: 54
Cathode length (mm) : 850
Masking (mm) : 0
Anode diameter (mm) : 20
Number cells : 10
Amps - per cell : 10
- circuit : 10
Current density (A/m: 69

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe (ppm) total	Fe (ppm) as Fe++	pH	Cu rec (gm) period	Cu rec (gm) Cumul.	Current eff. period	Current eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
15-Mar	0.56	0.00	3.9	10.0	149	1670	5	2.62						
15-Mar	0.59	0.70	3.5	10.0	139	1670	70	2.58	26.5	26.5	32	32	9.79	9.79
15-Mar	0.75	4.55	3.5	10.0	117	1650		2.53	58.3	84.8	13	16	23.15	18.97
15-Mar	1.00	10.42	3.5	10.0	92	1680	232	2.43	66.3	151.1	10	12	30.77	24.15
16-Mar	0.74	16.00	3.0	9.7	63	1670		2.27	76.9	227.9	12	12	22.90	23.73
16-Mar	1.00	22.05	3.0	10.0	38	1690	503	2.29	66.3	294.2	9	11	26.54	24.36
17-Mar	0.25	28.12	2.8	10.0	21	1690		2.21	45.1	339.2	6	10	38.99	26.30
17-Mar	0.41	32.03	2.8	10.0	14	1690	544	2.38	18.6	357.8	4	10	59.86	28.04

294126

TEST NO : ML 25 Poss power cut 0855/19th

Sample Volume (l) : 2650
Head grade (Cu ppm) : 135
Liquor flow (l/hour): 2400
Cathode diametre (mm: 54
Cathode length (mm) : 850
Masking (mm) : 0
Anode diameter (mm) : 20
Number cells : 10
Amps - per cell : 10
- circuit : 10
Current density (A/m: 69

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe (ppm) total	Fe (ppm) as Fe++	pH	Cu rec (gm) period	Cu rec (gm) Cumul.	Current eff. period	Current eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
18-Mar	0.49	0.00	3.8	10.0	135	1410	5	2.91						
18-Mar	0.59	2.47	3.5	10.0	124	1410	109	2.87	29.2	29.2	10	10	30.97	30.97
18-Mar	0.76	6.42	3.4	10.0	103	1410		2.78	55.7	84.8	12	11	24.38	26.65
19-Mar	0.03	12.95	3.2	10.0	68	1410		2.58	92.8	177.6	12	12	22.96	24.72
19-Mar	0.29	19.17	3.2	10.0	54	1410		2.46	37.1	214.7	5	9	53.03	29.62
19-Mar	0.50	24.32	3.0	10.0	31	1470	430	2.04	61.0	275.6	10	10	26.02	28.82
19-Mar	0.76	30.42	3.1	10.0	15	1450	458	1.99	42.4	318.0	6	9	44.17	30.87
19-Mar	0.99	36.07	3.0	9.5	9	1480	469	1.98	15.9	333.9	2	8	106.36	34.46
20-Mar	0.26	42.38	3.3	10.0	4	1430	480	1.98	13.3	347.2	2	7	145.72	38.71
20-Mar	0.45	47.00	3.3	10.0	2	1450	480		5.3	352.5	1	6	284.84	42.41

294127

TEST NO : ML 29

Sample Volume (l) : 2650
Head grade (Cu ppm) : 92
Liquor flow (l/hour): 2400
Cathode diametre (mm: 54
Cathode length (mm) : 850
Masking (mm) : 0
Anode diameter (mm) : 20
Number cells : 10
Amps - per cell : 15
- circuit : 15
Current density (A/m: 104

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe (ppm) total	Fe (ppm) as Fe++	pH	Cu rec (gm) period	Cu rec (gm) Cumul.	Current eff. period	Current eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
20-Mar	0.76	0.00	5.5	14.0	92	1030	31	2.36						
20-Mar	1.00	5.63	5.3	16.0	68	1050	209	2.70	63.6	63.6	6	6	71.68	71.68
21-Mar	0.25	11.73	5.0	17.0	35	1050	282	1.89	87.5	151.1	7	7	59.50	64.63
21-Mar	0.51	17.93	4.3	15.0	8	1030	335	1.79	71.6	222.6	6	7	64.47	64.58
21-Mar	0.67	21.73	4.3	15.5	3	1050	251	1.76	13.3	235.9	2	6	188.06	71.52
21-Mar	1.00	29.55	4.1	14.0	3	1040	363	1.75						
22-Mar	0.25	35.73	4.2	14.0	3	1040	341	1.73						
22-Mar	0.46	40.58	4.2	14.0	3	1050	349							

294128

TEST NO : ML 33

Sample Volume (l)	:	2650
Head grade (Cu ppm)	:	119
Liquor flow (l/hour)	:	3400
Cathode diametre (mm	:	54
Cathode length (mm)	:	850
Masking (mm)	:	0
Anode diameter (mm)	:	20
Number cells	:	10
Amps - per cell	:	15
- circuit	:	15
Current density (A/m	:	104

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe (ppm) total	Fe (ppm) as Fe++	pH	Cu rec (gm) period	Cu rec (gm) Cumul.	Current eff. period	Current eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
22-Mar	0.78	0.00	5.53	14.0	119	1170	25	2.58						
22-Mar	1.00	5.17	5.24	18.0	72	1170	215	2.34	124.6	124.6	13	13	35.74	35.74
23-Mar	0.25	11.25	4.2	15.0	54	1170	341	2.12	47.7	172.3	4	8	98.90	53.23
23-Mar	0.59	19.42	4.1	15.8	9	1180	396	2.04	119.3	291.5	8	8	43.77	49.36
23-Mar	0.75	23.20	3.9	15.0	4	1170	396	2.00	13.3	304.8	2	7	177.43	54.93
24-Mar	0.01	29.37	4.0	14.5	1	1180	377	2.00	8.0	312.7	1	6	450.79	64.99
24-Mar	0.25	35.22	4.0	14.5	2	1160	391	2.00						
24-Mar	0.46	40.22	4.0	15.5			413							

4 INCH CELL TESTS

TEST NO : ML 1

Sample Volume (l)	:	2650
Head grade (Cu ppm)	:	83
Liquor flow (l/hour)	:	3800
Cathode diametre (mm)	:	100
Cathode length (mm)	:	750
Masking (mm)	:	0
Anode diameter (mm)	:	20
Number cells	:	5
Amps - per cell	:	10
- circuit	:	10.00
Current density (A/m2)	:	42.42

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe(ppm) total	Fe(ppm) as Fe++	pH	Cu rec (gm) period	Cu rec (gm) Cumul.	Curr. eff. period	Curr. eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
28-Feb	0.40	0.00	7.6	10.0	94	730								
28-Feb	0.51	2.67	6.8	10.0	83	730		2.67	29.2	29.2	19	19	32.93	32.93
28-Feb	0.59	4.62	6.4	10.0	77	710		2.66	15.9	45.1	14	17	40.48	35.60
28-Feb	0.67	6.50	5.9	10.0	72	720		2.38	13.3	58.3	12	15	43.80	37.46
28-Feb	0.74	8.33	5.9	10.0	67	730		2.41	13.3	71.6	12	15	41.09	38.13
28-Feb	0.93	12.92	5.7	10.0	54	730		2.46	34.5	106.0	13	14	38.64	38.30
29-Feb	0.09	16.55	5.5	10.0	43	730		2.45	29.2	135.2	14	14	34.89	37.56
29-Feb	0.26	20.77	5.3	10.0	32	740		2.39	29.2	164.3	12	13	39.27	37.87
29-Feb	0.42	24.50	5.3	10.0	24	730		2.35	21.2	185.5	10	13	46.84	38.89
29-Feb	0.51	26.73	5.1	10.0	18	740		2.29	15.9	201.4	12	13	36.66	38.72

294130

TEST NO : ML 2

Sample Volume (l) : 2465
Head grade (Cu ppm) : 150
Liquor flow (l/hour) : 3800
Cathode diametre (mm) : 100
Cathode length (mm) : 750
Masking (mm) : 0
Anode diameter (mm) : 20
Number cells : 5
Amps - per cell : 10
- circuit : 10 = 2 banks in parallel
Current density (A/m2): 42

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe(ppm) total	Fe(ppm) as Fe++	pH	Cu rec (gm) period	Cu rec (gm) Cumul.	Curr. eff. period	Curr. eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
03-Mar	0.44	0.00	5.3	10.0	150	1410								
03-Mar	0.51	1.63	5.1	10.0	150	1410			0.0	0.0				
03-Mar	0.59	3.75	4.9	10.0	175	1420			-61.6	-61.6				
03-Mar	0.67	5.57	4.7	10.0	182	1410			-17.3	-78.9				
03-Mar	0.76	7.68	4.5	10.0	187	1440			-12.3	-91.2				
Off for inspection														
03-Mar	0.78	7.68	4.5	10.0	187	1440				-91.2				
03-Mar	0.92	10.97	4.3	10.0	201	1460			-34.5	-125.7				
Off for inspection														
03-Mar	0.93	10.97	4.3	10.0	201	1460				-125.7				
04-Mar	0.09	14.87	4.2	10.0	204	1440			-7.4	-133.1				
04-Mar	0.25	18.80	4.3	10.0	183	1450			51.8	-81.3	22		16.19	
04-Mar	0.44	23.23	4.2	10.0	159	1410			59.2	-22.2	23		15.93	
04-Mar	0.58	26.75	4.0	10.0	148	1440			27.1	4.9	13		26.59	

294131

TEST NO : ML 3

Sample Volume (l) : 3500 Main dump stream
Head grade (Cu ppm) : 165
Liquor flow (l/hour) : 3800
Cathode diametre (mm) : 100
Cathode length (mm) : 750
Masking (mm) : 0
Anode diameter (mm) : 20
Number cells : 5
Amps - per cell : 10
- circuit : 10.00
Current density (A/m2): 42.42

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe(ppm) total	Fe(ppm) as Fe++	pH	Cu rec (gm) period	Cu rec (gm) Cumul.	Curr. eff. period	Curr. eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
04-Mar	0.45	0.00	4.7	10.0	165	2370	25.0	2.61						
04-Mar	0.52	1.57	4.5	10.0	172	2350			-24.5					
04-Mar	0.59	3.37	4.4	10.0	170	2350			7.0					
04-Mar	0.68	5.50	4.3	10.0	180	2370			-35.0					
04-Mar	0.74	7.03	4.3	10.0	190	2390			-35.0					
04-Mar	0.94	11.63	4.0	10.0	210	2360			-70.0					
05-Mar	0.08	15.17	4.0	10.0	210	2370			0.0					
05-Mar	0.28	20.00	4.0	10.0	220	2430			-35.0					
05-Mar	0.43	23.40	4.0	10.0	230	2390			-35.0					
05-Mar	0.65	28.67	3.7	10.2	242	2460		2.40	-42.0					
05-Mar	0.75	31.17	3.6	10.0	258	2460		2.31	-56.0					
05-Mar	0.92	35.17	3.6	10.0	268	2500		2.49	-35.0					
06-Mar	0.08	39.20	3.5	10.0	274	2480		2.47	-21.0					
06-Mar	0.24	43.00	3.8	10.3	276	2480	495	2.43	-7.0					
06-Mar	0.42	47.17	3.7	10.0	266	2460	525	2.40	35.0	35.0	14	14	22.38	22.38
06-Mar	0.60	51.50	3.5	10.0	266	2520	551	2.35	0.0	35.0	0	7	ERR	44.52
06-Mar	0.66	53.05	3.5	9.9	261	2400		2.10	17.5	52.5	19	9	15.42	34.82

294132

TEST NO : ML 7

Sample Volume (l) : 3500
Head grade (Cu ppm) :
Liquor flow (l/hour) : 4200
Cathode diametre (mm) : 100
Cathode length (mm) : 750
Masking (mm) : 0
Anode diameter (mm) : 20
Number cells : 5
Amps - per cell : 10
- circuit : 10.00
Current density (A/m2): 42.42

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe(ppm) total	Fe(ppm) as Fe++	pH	Cu rec (gm) period	Cu rec (gm) Cumul.	Curr. eff. period	Curr. eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
06-Mar	0.55	0.00	5.2	10.0	155	1430		2.48						
06-Mar	0.67	2.88	4.8	9.8	171	1460		2.48	-56.0	-56.0		-33	-12.77	-12.77
06-Mar	0.75	4.72	4.9	10.0	179	1480		2.49	-28.0	-84.0		-30	-15.72	-13.75
06-Mar	0.93	9.00	4.6	11.0	185	1460		2.54	-21.0	-105.0		-19	-50.76	-21.15
07-Mar	0.11	13.40	4.2	9.0	198	1460		2.54	-45.5	-150.5		-19	-21.27	-21.19
07-Mar	0.25	16.87	4.6	10.0	193	1460		2.45	17.5	-133.0	9	-13	41.40	-29.43
07-Mar	0.42	20.87	4.5	10.6	182	1460		2.35	38.5	-94.5	16	-8	24.45	-51.38
07-Mar	0.59	24.98	4.5	11.0	164	1540		2.30	63.0	-31.5	24	-2	15.98	-186.10
07-Mar	0.75	28.77	4.0	10.0	162	1540		2.31	7.0	-24.5	3	-1	121.45	-273.97
07-Mar	0.91	32.60	4.1	10.2	150	1270		2.27	42.0	17.5	18	1	18.71	428.48

294133

TEST NO : ML 13

Sample Volume (l) : 3300 Liquor from #7 flushed
Head grade (Cu ppm) :
Liquor flow (l/hour) : 4380
Cathode diametre (mm) : 100
Cathode length (mm) : 750
Masking (mm) : 0
Anode diameter (mm) : 20
Number cells : 5
Amps - per cell : 10
- circuit : 10.00
Current density (A/m2): 42.42

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe(ppm) total	Fe(ppm) as Fe++	pH	Cu rec (gm) period	Cu rec (gm) Cumul.	Curr. eff. period	Curr. eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
09-Mar	0.75	0.00	4.0	10.0	232	1320		2.27						
09-Mar	0.93	4.52	4.2	10.5	192	1310		2.20	132.0	132.0	48	48	7.19	7.19
10-Mar	0.09	8.27	3.9	9.2	162	1310		2.15	99.0	231.0	45	47	7.52	7.33
10-Mar	0.28	12.85	4.2	10.0	126	1300		2.08	118.8	349.8	46	47	7.44	7.37
10-Mar	0.42	16.10	4.2	10.0	106	1310		2.02	66.0	415.8	34	44	10.27	7.83
10-Mar	0.60	20.50	3.9	11.0	71	1620		2.10	115.5	531.3	42	44	8.10	7.89
10-Mar	0.77	24.60	3.8	10.8	49	1650		2.08	72.6	603.9	28	41	11.82	8.38
10-Mar	0.91	27.85	3.9	10.0	36	1630		2.06	42.9	646.8	22	39	15.05	8.80
11-Mar	0.09	32.28	3.7	9.4	18	1630		2.04	59.4	706.2	23	37	13.68	9.21
11-Mar	0.26	36.30	4.2	11.0	11	1640		2.01	23.1	729.3	10	34	34.94	10.03
11-Mar	0.47	41.30	3.8	9.5	6	1650		2.01	16.5	745.8	5	30	62.28	11.19

294134

TEST NO : ML 19

Sample Volume (l) : 3400
Head grade (Cu ppm) :
Liquor flow (l/hour) : 4280
Cathode diametre (mm) : 100
Cathode length (mm) : 750
Masking (mm) : 0
Anode diameter (mm) : 20
Number cells : 5
Amps - per cell : 15
- circuit : 15.00
Current density (A/m2): 63.64

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe(ppm) total	Fe(ppm) as Fe++	pH	Cu rec (gm) period	Cu rec (gm) Cumul.	Curr. eff. period	Curr. eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
12-Mar	0.59	0.00	8.2	15.0	126	1310	50.0	2.53						
12-Mar	0.67	2.03	7.3	15.0	126	1330		2.51	0.0	0.0	0	0	ERR	ERR
12-Mar	0.75	3.90	7.4	16.0	122	1320		2.49	13.6	13.6	8	4	78.08	164.87
12-Mar	0.93	8.17	6.9	16.0	110	1360		2.44	40.8	54.4	10	7	59.90	86.14
13-Mar	0.09	12.12	6.5	14.0	96	1350	242	2.37	47.6	102.0	14	9	41.82	65.46
13-Mar	0.27	16.35	6.6	15.8	84	1370		2.28	40.8	142.8	11	10	50.63	61.22
13-Mar	0.42	20.03	6.5	15.5	71	1370	290	2.20	44.2	187.0	13	10	42.58	56.82
13-Mar	0.59	23.95	6.0	15.2	56	1330	341	2.23	51.0	238.0	14	11	36.72	52.51
13-Mar	0.74	27.73	5.8	15.5	40	1390	377	2.10	54.4	292.4	16	12	31.44	48.59
13-Mar	0.91	31.77	5.8	15.5	28			2.05	40.8	333.2	11	12	44.21	48.05
14-Mar	0.08	35.78	5.6	15.5	15	1370		2.01	44.2	377.4	12	12	39.93	47.10
14-Mar	0.27	40.37	5.8	15.5	9	1380	438	1.96	20.4	397.8	5	11	99.25	49.78
14-Mar	0.43	44.17	5.9	16.2	7	1360			6.8	404.6	2	10	259.96	53.31
14-Mar	0.51	46.07	5.4	14.5	6	1370	461		3.4	408.0	2	10	243.18	54.89

TEST NO : ML 24

Sample Volume (l) : 2650
Head grade (Cu ppm) : 83
Liquor flow (l/hour) : 3800
Cathode diametre (mm) : 100
Cathode length (mm) : 750
Masking (mm) : 0
Anode diameter (mm) : 20
Number cells : 5
Amps - per cell : 15
- circuit : 15.00
Current density (A/m2): 63.64

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe(ppm) total	Fe(ppm) as Fe++	pH	Cu rec (gm) period	Cu rec (gm) Cumul.	Curr. eff. period	Curr. eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
18-Mar	0.51	0.00	8.7	15.0	125	1330	25.0	2.86						
18-Mar	0.59	1.83	8.4	15.0	129	1360	67	2.89	-10.6	-10.6	-7	-7	-110.91	-110.91
18-Mar	0.76	5.87	7.6	15.0	113	1380		2.83	42.4	31.8	12	6	57.08	113.07
19-Mar	0.03	12.37	7.3	15.5	90	1380		2.73	61.0	92.8	10	8	60.58	78.58
19-Mar	0.25	17.58	6.5	15.0	70	1390		2.52	53.0	145.8	11	9	51.79	68.83
19-Mar	0.50	23.75	6.3	15.0	53	1420	317	2.21	45.1	190.8	8	9	65.70	68.10
19-Mar	0.76	29.92	6.0	15.0	35	1430	363	2.15	47.7	238.5	9	9	59.63	66.40
19-Mar	0.99	35.42	6.0	15.0	23	1400	402	2.10	31.8	270.3	7	9	77.83	67.75
20-Mar	0.25	41.75	6.0	15.0	9	1410	460	2.00	37.1	307.4	7	8	76.82	68.84
20-Mar	0.45	46.42	6.1	15.0	8	1410	450		2.7	310.1	1	8	799.06	75.08

DRUM CIRCUIT

TEST NO : ML 17

Sample Volume (l)	:	210
Head grade (Cu ppm)	:	144
Liquor flow (l/hour)	:	1700
Cathode diametre (mm)	:	50
Cathode length (mm)	:	900
Masking (mm)	:	0
Anode diameter (mm)	:	20
Number cells	:	1
Amps - per cell	:	7.5
- circuit	:	7.50
Current density (A/m2)	:	53.03

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe(ppm) total	Fe(ppm) as Fe++	pH	Cu rec (gm) period	Cu rec (gm) Cumul.	Curr. eff. period	Curr. eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
12-Mar	0.57	0.00	3.5	7.5	144	1490	17	2.24						
12-Mar	0.67	2.28	3.2	7.5	136	1500	25	2.29	1.7	1.7	8	8	34.15	34.15
12-Mar	0.75	4.25	3.0	7.5	130	1520	28	2.29	1.3	2.9	7	8	36.29	35.07
12-Mar	0.93	8.67	2.9	7.5	106	1530	56	2.26	5.0	8.0	13	10	19.39	25.16
13-Mar	0.09	12.42	2.7	7.0	90	1520	140	2.22	3.4	11.3	10	10	22.66	24.42
13-Mar	0.27	16.67	2.7	7.5	76	1520	209	2.15	2.9	14.3	8	10	28.30	25.22
13-Mar	0.42	20.33	2.6	7.0	64	1540	302	2.2	2.5	16.8	8	10	27.95	25.63
13-Mar	0.59	24.33	2.7	8.0	65	1550	346	2.23	-0.2	16.6	-1	8	-378.57	30.75
13-Mar	0.74	27.98	2.7	7.5	57	1530	392	2.12	1.7	18.3	5	7	45.46	32.10
13-Mar	0.91	32.08	2.7	8.5	49	1530	429	2.06	1.7	20.0	4	7	52.71	33.84
14-Mar	0.08	36.12	2.5	7.0	39		460	2.02	2.1	22.1	6	7	38.70	34.30
14-Mar	0.26	40.58	2.5	7.5	29	1530	489	1.95	2.1	24.2	5	7	38.55	34.67
14-Mar	0.43	44.48	2.6	7.3	21	1520	495		1.7	25.8	5	7	43.66	35.25
14-Mar	0.50	46.35	2.7	7.5	18	1520	525		0.6	26.5	4	6	57.91	35.79

294137

TEST NO : ML 21

Sample Volume (l)	:	210
Head grade (Cu ppm)	:	138
Liquor flow (l/hour)	:	1700
Cathode diametre (mm)	:	54
Cathode length (mm)	:	850
Masking (mm)	:	0
Anode diameter (mm)	:	20
Number cells	:	1
Amps - per cell	:	7.5
- circuit	:	7.50
Current density (A/m2)	:	51.99

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe(ppm) total	Fe(ppm) as Fe++	pH	Cu rec (gm) period	Cu rec (gm) Cumul.	Curr. eff. period	Curr. eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
15-Mar	0.61	0.00	3.4	7.5	138	1480	5	2.62						
15-Mar	0.67	1.42	3.2	7.5	132	1500	25	2.58	1.3	1.3	10	10	27.83	27.83
15-Mar	0.75	3.37	2.9	7.5	123	1490	56	2.55	1.9	3.2	11	11	23.80	25.29
15-Mar	1.00	9.25	2.9	7.0	95	1490	161	2.46	5.9	9.0	12	11	21.04	22.52
16-Mar	0.26	15.65	2.8	7.8	68	1490	268	2.3	5.7	14.7	10	11	23.81	23.02
16-Mar	0.50	21.37	2.6	7.3	49	1490	286		4.0	18.7	8	10	29.11	24.32
16-Mar	0.75	27.27	2.6	7.0	39	1530	394	2.32	2.1	20.8	4	9	51.55	27.07
16-Mar	0.99	33.20	2.6	7.2	31	1530	441	2.27	1.7	22.5	3	8	64.57	29.87
17-Mar	0.25	39.40	2.4	7.3	22	1520	469	2.22	1.9	24.4	4	7	59.46	32.17
17-Mar	0.41	43.30	2.2	6.0	16	1530	490	2.26	1.3	25.6	4	7	47.34	32.91

294138

TEST NO : ML 26

Sample Volume (l)	:	187
Head grade (Cu ppm)	:	
Liquor flow (l/hour)	:	800
Cathode diametre (mm)	:	50
Cathode length (mm)	:	900
Masking (mm)	:	0
Anode diameter (mm)	:	20
Number cells	:	1
Amps - per cell	:	10
- circuit	:	10.00
Current density (A/m2)	:	70.71

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe(ppm) total	Fe(ppm) as Fe++	pH	Cu rec (gm) period	Cu rec (gm) Cumul.	Curr. eff. period	Curr. eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
18-Mar	0.49	0.00	4.4	10.0	130	1340	42	2.92						
18-Mar	0.59	2.50	4.0	10.0	124	1360	81	2.87	1.1	1.1	4	4	93.75	93.75
18-Mar	0.76	6.42	3.8	10.0	117	1370	170	2.77	1.3	2.4	3	3	116.90	106.22
19-Mar	0.03	12.95	3.3	10.0	95	1370	296	2.56	4.1	6.5	5	4	56.48	74.95
19-Mar	0.25	18.22	3.2	10.0	76	1360	367	2.45	3.5	10.1	6	5	48.28	65.56
19-Mar	0.50	24.35	3.2	10.0	43	1420	430	2.03	6.2	16.2	9	6	31.86	52.78
19-Mar	0.76	30.53	3.2	9.8	41	1420	461	2	0.4	16.6	1	5	524.70	63.38
19-Mar	0.99	36.05	3.1	9.5	20	1420	463	1.99	3.9	20.5	6	5	42.78	59.45
20-Mar	0.26	42.42	3.2	10.0	13	1430	491	2	1.3	21.8	2	4	149.65	64.85
20-Mar	0.45	47.00	3.1	10.0	7	1390	500		1.1	23.0	2	4	128.91	67.97

294139

TEST NO : ML 28

Sample Volume (l)	:	175
Head grade (Cu ppm)	:	88
Liquor flow (l/hour)	:	800
Cathode diametre (mm)	:	50
Cathode length (mm)	:	900
Masking (mm)	:	0
Anode diameter (mm)	:	20
Number cells	:	1
Amps - per cell	:	10
- circuit	:	7.50
Current density (A/m2)	:	70.71

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe(ppm) total	Fe(ppm) as Fe++	pH	Cu rec (gm) period	Cu rec (gm) Cumul.	Curr. eff. period	Curr. eff. Cumul.	wh/gm Cu period	wh/gm Cu Cumul.
20-Mar	0.78	0.00	5.3	10.0	88	930	20							
20-Mar	1.00	5.17	3.9	10.0	87	950	109		3.7	3.7	6	6	64.67	64.67
21-Mar	0.25	11.27	4.0	10.0	61	930	221		1.1	4.7	1	4	229.48	101.29
21-Mar	0.51	17.42	3.3	10.0	44	940	265		3.0	7.7	4	4	75.45	91.31
21-Mar	0.75	23.27	3.6	10.0	27	950	304	2	3.0	10.7	4	4	87.84	84.77
21-Mar	1.00	29.08	2.9	8.5	12	950	307	1.96	2.8	13.3	4	4	66.61	81.19
22-Mar	0.25	35.23	3.3	8.0	6	940	307	1.9	1.1	14.4	2	4	149.80	86.21
22-Mar	0.45	40.10	3.6	12.0	9	950	304		-0.5	13.8	-1	3	-319.81	101.63

FLUSH TESTS

Drum Circuit

TEST NO : ML 9

Sample Volume (l) : 210 Fresh creek liquor
Head grade (Cu ppm) :
Liquor flow (l/hour) : 3130
Cathode diametre (mm) : 54
Cathode length (mm) : 750
Masking (mm) : 0
Anode diameter (mm) : 20
Number cells : 1
Amps - per cell : 0
- circuit : 0.00
Current density (A/m2): 0.00

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe(ppm) total	Fe(ppm) as Fe++	pH	liquor vol(l)	number cells	Cu rec (gm) period	Cu rec (gm) Cumul.	Cu diss /hr
07-Mar	0.48	0.00			138	1760			210.0	1			
07-Mar	0.58	2.50			266	1410		2.58	210.0	1	-26.9	-26.9	10.8
07-Mar	0.72	5.73			324	1440		2.6	210.0	1	-12.2	-39.1	3.8
07-Mar	0.85	8.98			376	1420		2.63	210.0	1	-10.9	-50.0	3.4

TEST NO : ML 14

Sample Volume (l) : 210 Fresh creek liquor
Head grade (Cu ppm) :
Liquor flow (l/hour) : 3130
Cathode diametre (mm) : 54
Cathode length (mm) : 750
Masking (mm) : 0
Anode diameter (mm) : 20
Number cells : 1
Amps - per cell : 0
- circuit : 0.00
Current density (A/m2): 0.00

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe(ppm) total	Fe(ppm) as Fe++	pH	liquor vol(l)	number cells	Cu rec (gm) period	Cu rec (gm) Cumul.	Cu diss /hr
10-Mar	0.47	0.00			376	1760		2.69	210.0	1			
10-Mar	0.60	3.30			378	1410		2.70	210.0	1		PUMP OFF IN ERROR	
10-Mar	0.77	7.37			372							LIQUOR LEFT IN ERROR	
10-Mar	0.92	10.98			380								

TEST NO : ML 23

Sample Volume (l) : 210
Head grade (Cu ppm) :
Liquor flow (l/hour) : 1800
Cathode diametre (mm) : 54
Cathode length (mm) : 750
Masking (mm) : 0
Anode diameter (mm) : 20
Number cells : 1
Amps - per cell : 0
- circuit : 0.00
Current density (A/m2): 0.00

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe(ppm) total	Fe(ppm) as Fe++	pH	liquor vol(l)	number cells	Cu rec (gm) period	Cu rec (gm) Cumul.	Cu diss /hr
17-Mar	0.50	0.00			16	1530	490		210.0	1			
17-Mar	0.59	1.95			31	1540	503	2.38	210.0	1	-3.2	-3.2	1.62
17-Mar	0.71	4.90			39	1520	489	2.37	210.0	1	-1.7	-4.8	0.57
17-Mar	0.75	5.95			42	1520	483	2.37	210.0	1	-0.6	-5.5	0.60

2 Inch Circuit

TEST NO : ML 8 (post #6)

Sample Volume (l) : 2650 Fresh creek liquor
Head grade (Cu ppm) :
Liquor flow (l/hour) : 1400
Cathode diametre (mm) : 54
Cathode length (mm) : 850
Masking (mm) : 0
Anode diameter (mm) : 20
Number cells : 10
Amps - per cell : 0
- circuit : 0.00
Current density (A/m2): 0.00

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe(ppm) total	Fe(ppm) as Fe++	pH	liquor vol(l)	number cells	Cu rec (gm) period	Cu rec (gm) Cumul.	Cu diss /hr	Cu diss /cell/hr
07-Mar	0.75	0.00			150	1750		2.54	2650.0	10				
07-Mar	0.76	0.17			242	1620		2.61	2650.0	10	-243.8	-243.8	1462.8	146.3
07-Mar	0.93	4.25			266	1610		2.63	2650.0	10	-63.6	-307.4	15.6	1.6
08-Mar	0.11	8.58			300	1610		2.65	2650.0	10	-90.1	-397.5	20.8	2.1
08-Mar	0.26	12.13			320	1600		2.65	2650.0	10	-53.0	-450.5	14.9	1.5
08-Mar	0.41	15.83			340	1600		2.62	2650.0	10	-53.0	-503.5	14.3	1.4

TEST NO : ML 11F

Sample Volume (l) : 2650 Fresh creek liquor
Head grade (Cu ppm) :
Liquor flow (l/hour) : 1800 (varied)
Cathode diametre (mm) : 54
Cathode length (mm) : 850
Masking (mm) : 0
Anode diameter (mm) : 20
Number cells : 10
Amps - per cell : 0
- circuit : 0.00
Current density (A/m2): 0.00

Date	Time	Elapsed Time	Avg cell Volts	Amps	Cu (ppm)	Fe(ppm) total	Fe(ppm) as Fe++	pH	liquor vol(l)	number cells	Cu rec (gm) period	Cu rec (gm) Cumul.	Cu diss /hr	Cu diss /cell/hr
11-Mar	0.64	0.00			144	1520		2.13	2650.0	10				
11-Mar	0.73	2.17			212	1470	5	2.51	2650.0	10	-180.2	-180.2	83.2	8.3
11-Mar	0.89	5.90			294	1460	64		2650.0	10	-217.3	-397.5	58.2	5.8

4 Inch Circuit

TEST NO : ML 13F

Sample Volume (l) : 3500 Fresh creek liquor
Head grade (Cu ppm) :
Liquor flow (l/hour) : 4000 plus
Cathode diametre (mm) : 100
Cathode length (mm) : 750
Masking (mm) : 0
Anode diameter (mm) : 20
Number cells : 5
Amps - per cell : 0
- circuit : 0.00
Current density (A/m2): 0.00

Date	Time	Elapsed Time	Avg cell Volts	Amps Cu (ppm)	Fe(ppm) total	Fe(ppm) as Fe++	pH	liquor vol(l)	number cells	Cu rec (gm) period	Cu rec (gm) Cumul.	Cu diss /hr	Cu diss /cell/hr
11-Mar	0.68	0.00		144	1520		2.13	3500.0	5				
11-Mar	0.73	1.22		206	1480		2.16	3500.0	5	-217.0	-217.0	178.4	35.7
11-Mar	0.91	5.45		252	1480			3500.0	5	-161.0	-378.0	38.0	7.6

TEST NO : ML 12

Sample Volume (l) : 3500 Liquor from test 7
Head grade (Cu ppm) :
Liquor flow (l/hour) : 4000 plus
Cathode diametre (mm) : 100
Cathode length (mm) : 750
Masking (mm) : 0
Anode diameter (mm) : 20
Number cells : 5
Amps - per cell : 0
- circuit : 0.00
Current density (A/m2): 0.00

Date	Time	Elapsed Time	Avg cell Volts	Amps Cu (ppm)	Fe(ppm) total	Fe(ppm) as Fe++	pH	liquor vol(l)	number cells	Cu rec (gm) period	Cu rec (gm) Cumul.	Cu diss /hr	Cu diss /cell/hr
09-Mar	0.49	0.00		154	1310	390	2.26	3500.0	5				
09-Mar	0.59	2.37		196	1290		2.24	3500.0	5	-147.0	-147.0	62.1	12.4
09-Mar	0.69	4.70		220	1300		2.15	3500.0	5	-84.0	-231.0	36.0	7.2

APPENDIX 4

Waste Liquor Measurements 1991

WATER QUALITY MEASUREMENTS 1991

LOCATION	DATE	Cu ppm	Fe ppm	Flow l/sec	pH	SO4 ppm	Al ppm
05	17-Jan	150	1450	20	2.5	10980	0
05	03-May	65	644	45	2.7	1667	283
05	03-Jun	73.2	359	0	2.7	2340	173
05	24-Jun	0	0	125	0	0	0
05	01-Jul	116.8	679	150	2.6	4309	275
05	30-Jul	170.5	1007	120	2.5	5845	283
05	26-Aug	137.5	620	50	2.5	5110	345
05	23-Sep	186.6	940	50	2.6	6707	562
05	17-Oct	55.2	520	80	2.7	4322	331
05	19-Nov	34	85	500	3	1108	99
05	09-Dec	126.3	870	10	2.5	8094	99
06	17-Jan	25	570	2	2.7	4780	0
06	03-May	25.1	538	2	3	1409	231
06	03-Jun	79.1	614	0	2.8	4085	247
06	24-Jun	0	0	2	0	0	0
06	01-Jul	83.9	805	4	2.8	4725	292
06	30-Jul	47.8	488	4	2.7	2916	125
06	26-Aug	25.7	499	1.5	2.9	3871	214
06	23-Sep	18.1	485	3.5	2.9	2965	191
06	17-Oct	23.8	550	4	3	4549	245
06	19-Nov	35.8	427	4	2.9	3288	189
06	09-Dec	34.9	459	3	2.8	3263	189
07	03-May	65	419	50	2.9	1020	59.4
07	03-Jun	108.3	327	0	2.7	1661	58
07	24-Jun	0	0	45	0	0	0
07	01-Jul	141.8	461	120	2.5	1925	69.2
07	30-Jul	128.7	402	150	2.8	1873	47.2
07	26-Aug	123.6	365	40	2.6	1649	59.6
07	23-Sep	64.9	353	35	2.7	1637	71
07	17-Oct	64.2	334	50	2.7	1780	72
07	19-Nov	65.5	290	75	2.7	1708	73
07	09-Dec	129.3	358	45	2.7	1692	73
08a	03-May	65	252	65	3	2226	252
08a	03-Jun	150.3	209	0	3	2955	140
08a	24-Jun	0	0	8	0	0	0
08a	01-Jul	532	607	70	2.7	7414	386
08a	30-Jul	320	745	50	2.6	6783	215
08a	26-Aug	405	476	60	2.9	6685	325
08a	23-Sep	274	384	80	2.9	4300	238
08a	17-Oct	298	355	75	2.8	5815	263
08a	19-Nov	261	244	60	2.9	3925	197
08a	09-Dec	290	306	80	2.8	5290	197
09a	03-May	93.5	308	200	3	1536	182
09a	03-Jun	70	444	0	2.8	3266	195
09a	24-Jun	0	0	300	0	0	0
09a	01-Jul	85.5	505	300	2.6	3538	196
09a	30-Jul	0	0	300	0	0	0
09a	26-Aug	75.3	447	170	2.7	3680	209
09a	23-Sep	337.9	510	180	2.8	3876	263
09a	17-Oct	89.6	372	350	2.8	3915	221
09a	19-Nov	25	1956	600	3	1121	93
09a	09-Dec	89.4	336	160	2.8	3879	307

WET	DRY	
3"	6"	9"

294147

97-4023(A)
Vol 2 of 3
(OF)

NOTE: Regarded as Appendix 5 of 97-4023

EURALBA MINING LIMITED

Mt Lyell Waste Dumps Effluent Treatment

The Mt Lyell SX/EW Copper Project

Feasibility Study

Gunn Metallurgy
1993

294B

EURALBA MINING LTD

MT LYELL WASTE DUMPS EFFLUENT TREATMENT

THE MT LYELL SX/EW COPPER PROJECT

FEASIBILITY STUDY

GUNN METALLURGY
1993

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1.0 EXECUTIVE SUMMARY

1.1 Background to the Study

Euralba Mining Ltd have developed an electrowinning technology (the EMEW cell) which is capable of plating metals at a broad range of concentrations from aqueous solutions at high current densities and efficiencies. This technology is compact and relatively low in capital cost. An application to which it is well suited is the extraction of metals from waste streams, where the concentration of the target metal is often very low relative to that required for conventional electrowinning plant.

This technology has been pilot tested extensively across Australia, including two campaigns at the Mt Lyell Mining and Railway Company mine at Queenstown in Tasmania. The tests were carried out on acid mine drainage from the waste ore dumps formed during open pit mining operations between 1932 and 1970. Bacterial activity is now at levels such that the copper sulphide minerals in the dumps are being slowly broken down. The drainage from the dumps carries sufficient copper in solution to justify investigation of methods for its recovery by a commercial venture. The environmental impact of this copper in solution provides added incentive for the extraction of the copper.

Euralba Mining commissioned Australian Geoscientists to produce a prefeasibility study in 1992. This was based upon pilot scale site trials of direct electrowinning from a combination of waste dump drainage and surface runoff. The trials were successful, but the size of plant required (the number of EMEW cells) was larger than anticipated due to the very high iron levels in the drainage stream, and was sensitive to power cost. A conceptual design and cost estimate for a solvent extraction (SX) process was included in the study. This demonstrated some cost savings over the direct electrowin option when SX is used in conjunction with the EMEW cell, and a more manageable operation. In effect the impact of the high iron content on current efficiency was eliminated, and the size of the EMEW plant was reduced.

In November and December of 1993 a pilot scale SX plant was run on a variety of drainage streams at the Mt Lyell site, with copper from the upgraded solution recovered by EMEW cells. Sufficient data was collected to enable both the design and cost estimation of a full scale process plant, and the commercial analysis of a full scale operation. In conjunction with this trial, information on the streams available for processing was gathered, and a more definitive assessment of the available resource was undertaken. At the time of writing this report, the main issue which had not been determined was the future of the Mt Lyell site. The future could bring either a change of owner and some form of ongoing operation, a care and maintenance phase with several possible formats, or total cessation of operations and site remediation. This study assumes that the major infrastructure elements such as power supply and access to the site will be available to the SX/EMEW project, and otherwise that the plant site will be below the conveyor tunnel elevation on flat ground.

1.2 Technical Overview

The efficiency of the EMEW cell derives from the high flow and thus mass transport rates achieved in the cell and the use of an advanced dimensionally stable anode. The technology has been fully demonstrated on a small scale (to 1 tonne per day copper), and is currently undergoing a staged trial at the Young Australian mine near Cloncurry which will establish the engineering parameters for plants capable of producing from 2 to 10 tonnes of copper per day. The pilot plant operation for the SX/EMEW process at Mt Lyell used conventional open SX settlers in a single stage each of extraction and stripping. The continuous capacity of the plant ranged between 2500 and 3600 litres per hour of copper bearing solution, upgrading the feed from typically 150ppm (parts per million) copper and 2500ppm iron to an electrowin feed solution of concentration 20000ppm (20 g/l) copper and 5500ppm iron. The recovery of copper from the waste streams, with only one stage of extraction, varied between 60 and 88%, with the plant operating continuously for most of the 28 days of operation. The full scale flowsheet will use two stages of extraction, and thus it is anticipated that recoveries will be in the 80 to 90% range.

In technical terms the plant operated very well, particularly when account is taken of the metal concentrations in the feed solution. The plant achieved design performance with far higher iron to copper ratios than any operation in existence, and with copper tenors at the very lowest limit of current world practice. A further technical challenge which was overcome was the volume of crud (amorphous gels) which formed in the SX process, initially causing exceptionally high losses of the organic phase which carries the copper selective extractant. The plant was successfully modified on site to allow operations to continue at acceptable performance levels. Well tried techniques are available for the removal and processing of crud from either phase, and these have been used in the full scale plant design.

The technical approach proposed for the full scale plant is sound and has been based upon significantly more testing at both laboratory and pilot scale than is customary. The quality of copper produced has been determined from the pilot program, and is of a purity which is superior to most electrowon copper. The resource assumptions are considered to be conservative, based on reliable data from the streams which have been monitored. No attempt has been made to rationalise stream collection across the site or to enhance the leaching processes which have been occurring naturally over the last 60 years - both of which would considerably increase the resource base. The copper tenor and flowrate selected for commercial analysis of the project have been conservatively selected.

1.3 Commercial Highlights

The attached table provides a summary of the financial analyses detailed in section 10. of this report.

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EURALBA MINING LTD
MT LYELL SX/EW COPPER PROJECT
FEASIBILITY STUDY MAY, 1994

FINANCIAL ANALYSIS

SUMMARY - BASE CASE

BASE CONDITIONS			ASSUMPTIONS		
FLOW COPPER TENOR	0.120	g/l	COPPER PRICE	2.68	A$/KG
RECOVERY	80	%	EXCHANGE RATE	0.85	US$/LB
CAPITAL COST A$	1041889		DAYS/YEAR	0.7	US$/A$
PREPRODUCTION COST	100000		HRS/DAY	364	
BASE CASE ANNUAL PRODUCTION			INTEREST RATE	24	
FLOWRATE M3/HR	500		DISCOUNT RATE	10	%
COPPER g/l	0.12		TERM	12	%
RECOVERY %	80		NPV ON REVENUE	10	YEARS
COPPER TONNES	419.3		NPV ON CFBIT	3241679	
OPERATING COST FACTOR	1		NPV	2112660	A$ REVENUE - CAPITAL
				1980367	A$ AFTER DEBT SERVICE AND TAX

SENSITIVITIES - A$ x 1000

COPPER PRICE	0.75	0.85	0.95	US cents/pound Cu
NPV - REVENUE	2510	3242	3973	
NPV - CFBIT	1381	2113	2844	
NPV - DEBT FINANCED	1487	1980	2471	

CAPITAL COST	-20%	base	+20%
NPV - REVENUE	3242	3242	3242
NPV - CFBIT	2321	2113	1904
NPV - DEBT FINANCED	1996	1980	1962

OPERATING COST	-20%	base	+20%
NPV - REVENUE	3837	3242	2646
NPV - CFBIT	2708	2113	1517
NPV - DEBT FINANCED	2379	1980	1579

PRODUCTION	0.070	0.120	0.170	0.220	g/litre Cu
NPV - REVENUE	1056	3242	5427	7613	
NPV - CFBIT	-73	2113	4298	6484	
NPV - DEBT FINANCED	388	1980	3445	4909	

DISCOUNT RATE	9%	12%	15%
NPV - REVENUE	3693	3242	2871
NPV - CFBIT	2551	2113	1753
NPV - DEBT FINANCED	2239	1980	1765

2.0 INTRODUCTION

2.1 General Background

This study has been commissioned by Euralba Mining Ltd to examine the viability of producing copper metal from acid mine drainage effluents at the Mt Lyell Mining and Railway Company site in Queenstown, Tasmania. The Mt Lyell mining field has been producing copper metal or mineral from a range of orebodies over the last 100 years. The current owner, Renison Goldfields intends to cease mining and processing operations in December 1994, and at the time of writing of this report there is no indication of the extent of remediation, mothballing or ongoing liquidation of assets which may take place. Several parties have expressed interest in taking over the operation as a going concern.

The Tasmanian Government is involved with regard to both remediation of the minesite and in seeking ways by which some form of profitable activity can continue in the lease. They are supportive of Euralba's interest in extracting copper from the waste streams, possibly in part because this will contribute to the remediation of the site and significantly reduce the impact of the long term contamination of the Queen and King Rivers and Macquarie Harbour by the copper content of the acid mine drainage.

2.2 Corporate Details

The corporate background of Euralba Mining Limited is provided under separate cover in the form of an annual report, a company presentation document, a recent quarterly report and several analyses carried out by stockbrokers.

2.3 Objectives of the Study

The primary objective of this study is to assemble a definitive analysis of and proposal for the design, construction and operation of a solvent extraction and electrowinning (SX and EW) operation located at Mt Lyell in Tasmania. This operation proposes to process a variety of established copper bearing effluent streams to recover the copper as LME grade A cathode at a nominal rate of one tonne per day using commercially proven and available solvent extraction technology and the proprietary Euralba Mining Ltd electrowinning cell.

This study also projects the costs and income associated with this proposed operation, and presents a financial analysis of the project. The financial viability of the project is established within the context of the stated assumptions regarding the long term fate of the site and the associated tenure and infrastructure issues. The project is dependent on the data supplied by the current owner regarding the stream flows and their metal content, and the more comprehensive and specific data collected by Euralba during the two trial programs at the mine site, for estimation of the long term resource of copper in solution available to the project.

2.4 Study Organisation

The study is structured in the form of a conventional feasibility study, with the background to the project, resource definition and technical design issues preceding the presentation of operating and capital cost estimates. The study concludes with a financial analysis presenting pre-tax and pre-debt service cash flows, and calculation of the Net Present Value of the cash flow. A risk analysis is undertaken with the illustration of sensitivity of the project to copper price, exchange rate, copper levels in process feed and extraction levels achieved by the process.

2.5 Project History

The Mt Lyell Mining & Railway Company have been seeking to process their effluents for many years in such a way that the cost of treatment is offset by revenue from recovery of the contained copper. The origin of the copper is from the bacterial leaching of copper sulphide minerals in the large volumes of disturbed and broken ground at the Mt Lyell copper mine site. This endeavour was extra to their legal obligations with regard to environmental remediation, due in large part to the long mining history in the area and the commercial impracticability of nullifying the effects of past mining and management practices in the context of current day standards.

The Euralba Mining Ltd technology for electrowinning base metals from solutions carrying very low concentrations of the target metal was developed over the 1980's and early 1990's. The EMEW cell is now engineered to a point where commercial application can be undertaken. A pilot scale trial was undertaken in 1991 and 1992 at the mine site, with copper being electrowon directly from the various effluent streams. This trial was successful, but indicated that the purification and concentration of the effluent streams would reduce the capital and operating costs and improve the flexibility of the plant. The iron levels in solution reduced the overall power efficiency of the direct electrowinning process, and the addition of a solvent extraction process would reduce this effect and the size of plant required.

A pilot plant trial of the combined SX and EW was undertaken in late 1993 at the mine site, and provided the design parameters for a full scale plant.

2.6 Information Sources

This study is largely based on the pilot plant trials conducted in November 1993, and the information gathered at that time on stream flows, copper concentrations and infrastructure issues. The study also both directly and indirectly draws upon data gathered over the entire duration of Euralba Mining Ltd's involvement with the Mt Lyell site, and the large database of performance data accumulated during the process and engineering development phase of the EMEW cell. The capital cost estimates for the full scale plant have been assembled by obtaining quotations from suppliers, these quotations based upon designs and specifications supplied by Euralba Mining Ltd for the electrowinning process. The experience of the author and the significant body of information in the public domain has determined the process and engineering design of the solvent extraction plant. The Mt Lyell Railway and Mining Company provided invaluable assistance in the establishment of the pilot plants, by contributing financially and by providing access to materials, services and data.

2.7 Technology Background

Solvent extraction (SX) and Electrowinning (EW) are now thoroughly established and commercially viable technologies, but remain unfamiliar to many people even in the main stream of the mining industry.

SX is used to separate a relatively low concentration of a specific metal cation from a solution containing a wide range of ionic species (such as occurs with the leaching of metalliferrous ores), and form a relatively pure solution at high concentration of that cation. In the extraction of copper the specific extractant molecule - known as an aldoxime - has the capacity to carry a proton - a hydrogen ion - in a low pH environment. If the pH is raised and copper is present in intimate contact with the extractant molecule as a sulphate, then the copper will swap places with two of the protons. This contacting takes place in the mixer compartment of the "mixer/settler". The two protons associate themselves in the aqueous phase with the sulphate to form sulphuric acid. The copper cation remains attached to the extractant molecule until contacted with a low pH (high proton concentration) aqueous phase, and under these conditions the copper exhanges, or is stripped, into the aqueous phase. The volume of the low pH strip solution is much lower than the solution from which the cation was originally extracted, enabling the increased concentration. The ability to move the organic from one aqueous environment to another, from the extraction to the stripping stages, is facilitated by the natural separation of the oily organic from the aqueous phase in the settler compartment of the "mixer settler". The organic phase is composed of the extractant dissolved in kerosene.

Once the copper has been selectively transferred to a low pH aqueous solution (electrolyte), the EW process is used to convert the ionic copper to metallic copper by adding electrons. This is done by pumping electrons through the solution, which causes the copper to plate onto the stainless steel surface supplying the electrons. To preserve the electrochemical balance water breaks down so that hydrogen replaces the copper removed from the solution and the excess oxygen is released as a gas. The EMEW cell is designed to perform this task at very high efficiencies, and is not as vulnerable to impurities and trace organic phase in the electrolyte as the conventional EW processes.

The EMEW process has now been demonstrated at commercial scale, producing a readily harvested, high purity metal product at low capital and operating cost relative to conventional technology. Assay data obtained for the copper produced during the pilot program at Mt Lyell is attached in appendix 6.

The quality of the copper produced in the EMEW cell is significantly higher than that produced by conventionally electrowon copper, and is thus readily marketable. ?

3.0 PROJECT ENVIRONMENT

3.1 Location and Access

The Mt Lyell low tenor copper SX/EW project is based on the Mt Lyell Mining and Railway Co site at Queenstown on the west coast of Tasmania. Queenstown is accessed from the north by the Zeehan and Murchison highways which link the town to the regional centre of Wynyard/Burnie, and from the south by the Lyell highway which links to the state capital of Hobart. The rail link (Emu Bay Railway) is for freighting of Mt Lyell concentrates, and runs from a siding at Melba flats (40 kilometres north on the Zeehan highway) to Burnie. An airstrip at Queenstown is used by a regular light aircraft service from Hobart and the northern cities.

The proposed solution transport system and process plant will be located downslope to the west of the waste dumps and close to the outfall from the conveyor tunnel. The plant site is subject to change depending on the status of the site when the project commences. The location will not, within limits, have an impact on the project performance. There will be cost advantages in locating the plant as close as possible to the existing power infrastructure, and minimising the pumping requirements for aquisition of feed solutions. There are a number of suitable sites at lower elevations than the sources of the various streams of interest.

3.2 Topography

The topography of the area of interest is shown on figure 1. It is very steep and rugged, with roads and drainage ways confined either to the topographic features or, where this has hindered the effective movement of men and materials in the past, to railway tunnels driven through the mountains. The bulk of the extensive tunnel systems are now unused, with some used as drainage ways. The topography also justified the use of aerial ropeways in past years, whereas currently ore is moved by a sophisticated overland conveyor system and by truck.

The topography can be used to advantage in this project, enabling the pipeline transport of solutions to and from the process plant. The location of the waste dumps along the western and southern side of the open pit confines any drainage to either the valleys on the west which converge to haulage creek or to the open pit which drains into the underground. Minimisation of the dilution effect of surface runoff can be achieved by development of the drainage systems around the critical dump catchment area and provision of containment structures in the gullies at the toes of the shaft and haulage dumps.

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Fig 1.

Topo map of site

3.3 Geotechnical Features

Geotechnical factors do not bear significantly on the design of the project. The waste dumps are stable on their western battered exposure. The effects of block caving, with the pit walls now visibly failing and spalling into the open cut, may break back into the waste dumps but should not significantly effect leaching (and would remain inside the leaching zone). At the point where the dumps adjoin the top edge of the open pit the depth of waste is at its lowest and some exposure may assist air ingress.

A total mine closure may create a need for occasional maintenance of cuttings and dump slopes, largely due to the high rainfall. The plant foundation requirements are not onerous geotechnically. There are many sources for the quantities of fill which will be required for pond walls and foundation pads for the mixer/settlers.

3.4 Hydrology & Meteorology

The Mt Lyell site by Australian standards is wet and cold. The weather pattern is dominated primarily by the southern ocean westerlies, with less frequent fine weather when high pressure zones form in the east of the state. The rainfall and temperature data for the area is shown in appendix 1. The temperatures affect the performance of the solvent extraction and electrowin plants, particularly at night, and the rainfall creates significant variations in the flow and copper tenor of the drainage from the dumps and underground.

The ground water system is not relevant to this study and has not been investigated. The effluent from the plant is released into the Queen River at much lower copper concentrations, but with a small decrease in pH which is stoichiometrically related to the amount of copper recovered. This increase in acid concentration in the river would be undetectable due to the buffering effect of the high iron concentration, which exists as a balance between hydroxides and sulphates of iron.

3.5 Infrastructure & Services

The existence of comprehensive infrastructure and services at Mt Lyell and Queenstown is of major benefit to the project. Every facility required by the proposed project is currently available, although a complete shutdown of existing operations may require some intervention and renegotiation to ensure ongoing supply of power, water and communications.

Power supply is provided by both the state grid and the local Lake Margaret hydroelectric facility. Substations and the power distribution across the site are established.

Water supply is from rainfall catchment dams on the Mt Lyell site.

3.6 Demography & Social Issues

Queenstown has a population of ..... people, and a workforce directly dependent on the existing Mt Lyell operation of approximately 300. Tourism has contributed to the local economy, but when the mine closes there will be very little to compensate for loss of jobs in the local area. Some families have lived in Queenstown for 5 generations or more, and it is likely that many will welcome any opportunity to stay in the district. The last major contraction was the exodus of the hydroelectric scheme team after completion of the King River Dam project, contributing to the steady contraction in government services over recent years.

The only near term reprieve for the working population is the Henty Gold Project located to the east of Roseberry, which would be manned out of Queenstown. Otherwise the area will be reliant purely on tourism for income, with a declining contribution from the significant number of retirees who have remained in the town.

3.7 Political & Regulatory Factors

These issues have not been determined at this stage of the project development.

3.8 Tenure & Associated Interests

These issues have not been determined at this stage of the project development.

4.0 THE RESOURCE

4.1 Existing Stream Flows

The Mt Lyell Mining & Railway Co have monitored the acid mine drainage from the site for many years. Much of this data is of limited use in assessing the availability of metal to a process plant, as it is comprised only of assays with no stream flowrate data. The absence of flow data prevents the calculation of actual copper tonnage emerging from the various areas of the Mt Lyell site in the earlier years. Since 1991, and the installation of V notch weir flow measurement systems, a more quantitative data set has been compiled.

The three sites of immediate interest are Haulage Creek, which carries all the acid drainage from the West Lyell open cut waste dumps and large volumes of surface runoff, the North Lyell Tunnel, which carries drainage from the Tharsis open pit area, and the Conveyor Tunnel which carries the water pumped from the underground areas currently in production. These underground workings, accessed from the Prince Lyell shaft, are directly below the West Lyell open pit which ceased production in 1974. Rainfall drains from the open pit and its catchment through the caved areas above the production stopes, and then into the underground sumps. The water is continuously pumped from the sumps, with an occasional clearance of settled solids (see figure 1).

The sampling points have been selected such that addition of the data from each source yields the total flow from that area of the minesite. The North Lyell Tunnel and Conveyor Tunnel flows discharge into haulage creek which in turn flows into the Queen River (which flows into Macqarie Harbour). The data does not enable calculation of the clean water dilution from surface runoff which flows into haulage creek. The pilot program carried out in 1993 took solution flows from the toe of the waste dumps and the outflow of the two tunnels, thus avoiding the dilution. A back calculation of the flowrate which the plant will obtain at full scale from the waste dumps has been based on the change in metal concentration between the haulage creek data and the waste dump outflow samples (see figure 2 and table 1).

Over the last 3 years a simple arithmetic average of 1.56 tonnes of copper per day has been transported to the Queen River in solution as copper sulphate in these three streams. The pattern of flow appears to follow the rainfall curve, shown on figure 3. Appendix 1 contains the meteorological data for the Queenstown area (see section 3.4).

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TABLE 1.

MT LYELL SX/EW COPPER PROJECT
PROJECTION OF LEACHING RATES
PROJECTION OF STREAM FLOWS AND COPPER TENORS
TO THE PROPOSED SX/EW PROCESS PLANT

HAULAGE CREEK	N. LYELL TUNNEL	CONVEYOR TUNNEL		TOTAL
0.084	0.059	0.125	g/l	
107	61	68	l/sec	
8.98	3.54	8.47	g/sec	
0.776	0.306	0.731	T/day	1.814 T/day
474	302	758	kg/day	1534 kg/day
282	111	266	T/yr	660 T/yr

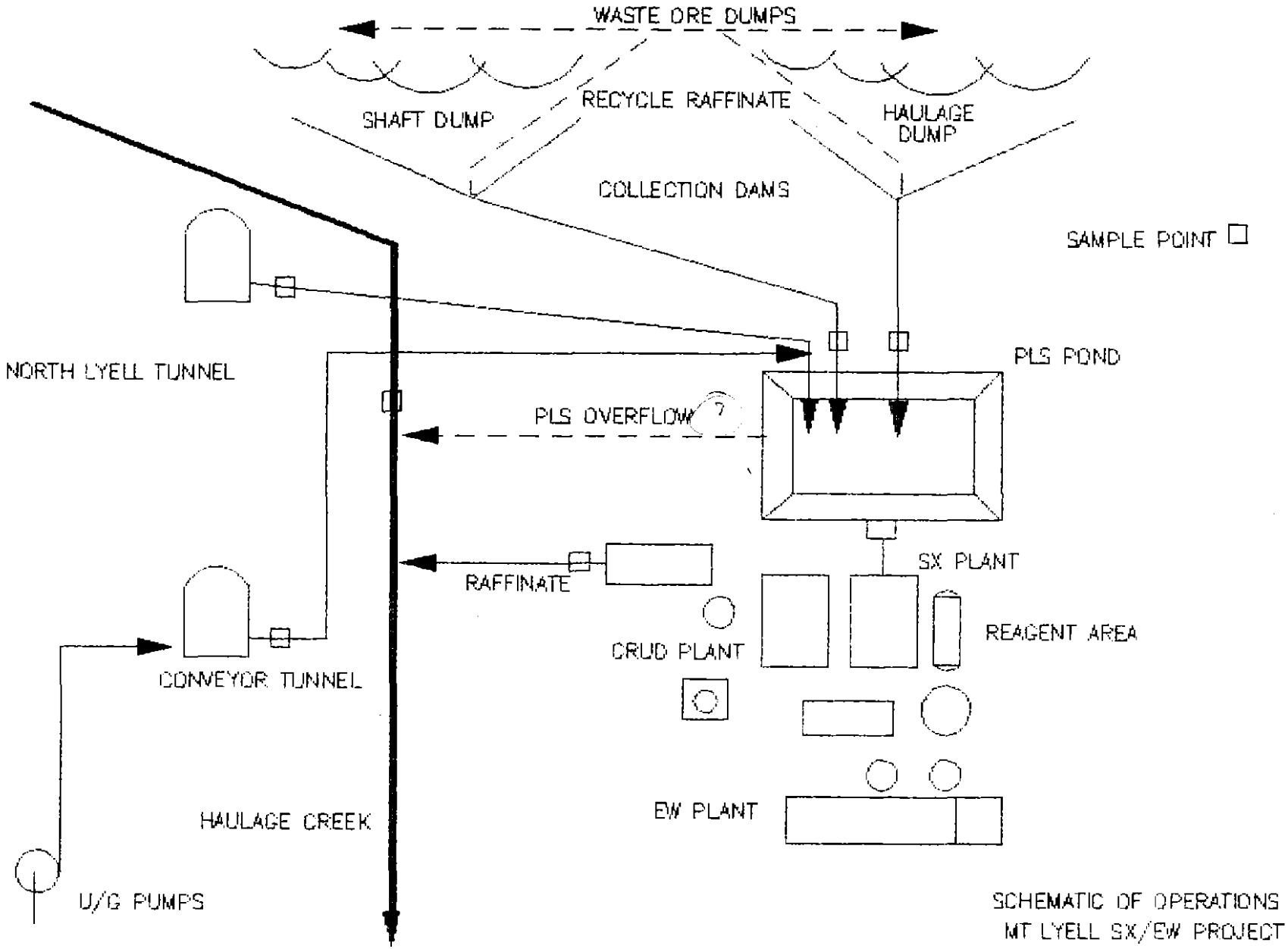
FOR HAULAGE CREEK - DUMPS
18076 T Cu extracted in 64 years
81784 T Cu in dumps
22.1 % extraction to date
0.35 % extraction /year

1000000 T ore/year
0.17 %Cu av
5.87 T Cu/yr

assume linear dump construction over 64 years & leaching over 60

YEARS	60	59	58	57	56	55	54	53	TOTALS
T Cu	352.2	346.4	340.5	334.6	328.8	322.9	317.0	311.2	2653.5
	52	51	50	49	48	47	46	45	
	305.3	299.4	293.5	287.7	281.8	275.9	270.1	264.2	2277.8
	44	43	42	41	40	39	38	37	
	258.3	252.4	246.6	240.7	234.8	229.0	223.1	217.2	1902.1
	36	35	34	33	32	31	30	29	
	211.3	205.5	199.6	193.7	187.9	182.0	176.1	170.3	1526.3
	28	27	26	25	24	23	22	21	
	164.4	158.5	152.6	146.9	140.9	135.0	129.2	123.3	1150.6
	20	19	18	17	16	15	14	13	
	117.4	111.5	105.7	99.8	93.9	88.1	82.2	76.3	774.94
	12	11	10	9	8	7	6	5	
	70.4	64.6	58.7	52.8	47.0	41.1	35.2	29.4	399.21
	4	3	2	1					
	23.5	17.6	11.7	5.9					58.707

10685 T Cu tonnes copper in 60 years
13.1 % extraction



SCHEMATIC OF OPERATIONS
MT LYELL SX/EW PROJECT

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PROJECTION OF ACTUAL FLOWS

	HC	NLT	CT		
COPPER	0.084	0.059 ✓	0.125	g/l ✓	sampled
	107.0	60.5	67.6	l/sec	sampled
IRON	0.792	0.142	0.744	g/l	sampled
copper	8.980	3.544	8.465	g/sec	sampled
	0.170	0.059 ✓	0.125	g/l ✓	projected
	50.0	30.0	60.0	l/sec	projected
copper	8.500	1.757	7.516	g/sec	projected
iron	39.6	4.3	44.6	g/sec	

?

COMBINED FLOW PROJECTION

0.127 g/l copper
0.632 g/l iron
5.0 iron/copper ratio
140.00 l/sec
504.00 m3/hr
1.54 T/day in feed
0.80 extraction
1.23 T/day extracted

LINE SIZING

HC	NLT	CT	
180.0	108.0	216.0	M3/HR
235	235	297	ID
0.043	0.043	0.069	M2
1.15	0.69	0.87	M/SEC

OD line size	ID nominal
500	470
450	423
400	377
355	334
315	297
280	263
250	235

MT LYELL STREAM FLOW DATA
HAULAGE CREEK

PERIOD	Cu	Fe	l/sec	Cu kg/day	AV RAIN	1992 RAIN
1	58.1	679	150	753	235	0
2	85.6	1007	120	888	260	0
3	68.7	620	50	297	263	0
4	92.8	940	50	401	248	0
5	55.2	520	80	382	226	0
6	16.7	85	500	721	199	0
7	109	870	10	94	184	0
8	142	934	20	245	151	173
9	167	1164	20	289	123	80
10	94	727	40	325	163	99
11	109	1400	25	235	224	173
12	72.1	700	65	405	244	195
13	149	1480	30	386	235	202
14	75	743	60	389	260	403
15	47.5	505	60	246	263	336
16	72.9	690	40	252	248	245
17	63	578	25	136	226	202
18	84.5	840	30	219	199	210
19	138	1540	25	298	184	114
20	87.3	955	25	189	151	0
21	83.2	754	20	144	121	0
22	78.3	510	20	135	163	0
23	82.4	711	20	142	224	0
24	38.6	401	360	1201	244	0
25	26	225	720	1617	235	0
26	87.7	880	216	1637	260	0
27	82.5	919	108	770	263	0
AVERAGE	84	792	107	474		
	27	27	27	27		
				776		
AV. Fe/Cu	9.43					

NORTH LYELL TUNNEL

	Cu	Fe	l/sec	Cu kg/day
	68.3	461	120	708
	66.3	402	150	859
	61.6	365	40	213
	64.9	353	35	196
	64.2	334	50	277
	65.5	290	75	424
	63.8	358	45	248
	58.8	286	45	229
	51.9	228	25	112
	76	205	35	230
	73.1	449	35	221
	55.1	335	50	238
	56.7	284	35	171
	47.6	284	45	185
	42.5	294	45	165
	62.4	418	30	162
	49.1	319	30	127
	39.2	276	35	119
	63	449	25	136
	53.3	350	20	92
	58.7	388	30	152
	65.2	455	15	84
	61.6	437	25	133
	48.4	312	180	753
	58.4	388	54	272
	47.7	285	180	742
	58	387	180	902
AVERAGE	59	348	61	302
	27	27	27	27

AV. Fe/Cu 5.94

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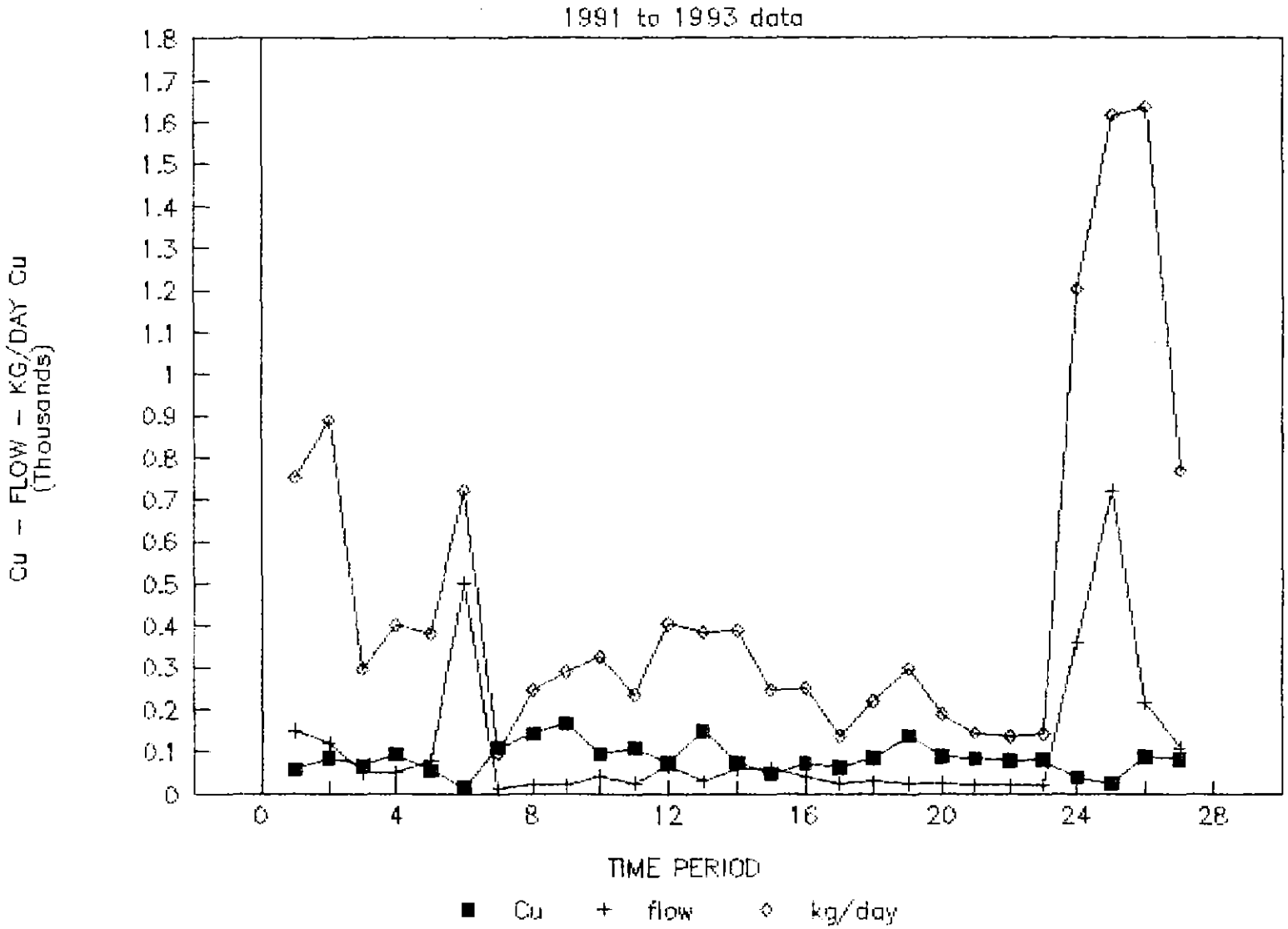
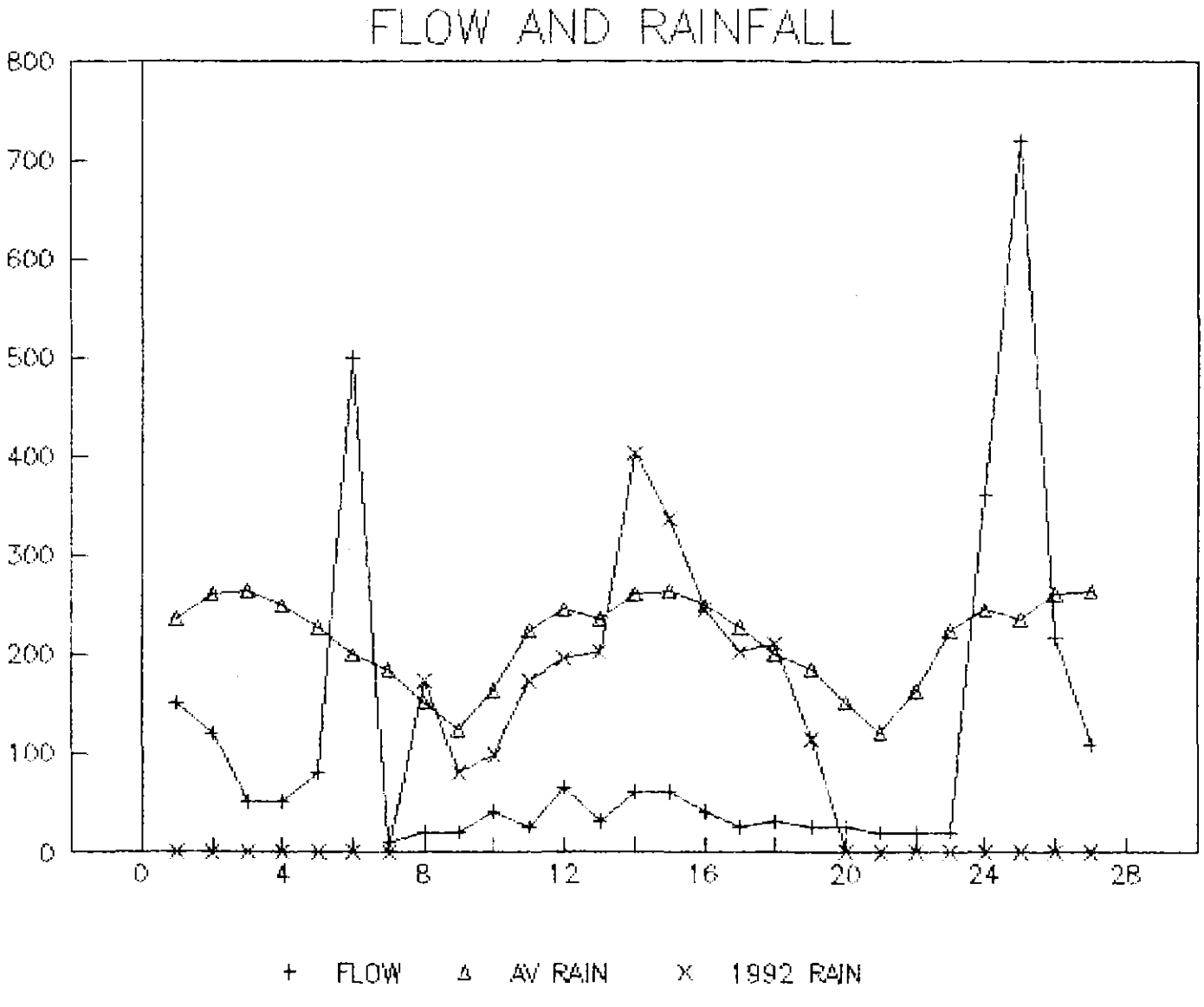
CONVEYOR TUNNEL

	Cu	Fe	l/sec	Cu kg/day
	237	607	70	1433
	114	745	50	492
	200	476	60	1037
	134	384	80	926
	147	355	75	953
	121	244	60	627
	129	306	80	892
				0
	83.2	109	80	575
	98.5	155	80	681
	33.3	44.7	40	115
	98.52	373	70	596
	138	224	20	238
	227	453	60	1177
	161	405	50	696
	182	507	60	943
	91.1	131	20	157
	44.8	68	100	387
	102	206	40	353
	91.1	131	30	236
	152	201	45	591
	98.1	246	40	339
	51.2	115	25	111
	30.2	146	108	282
	173	496	126	1883
	144	308	144	1792
	176	486	144	2190
AVERAGE	125	305	68	758
	26	26	26	26

AV. Fe/Cu 2.43

HAULAGE CREEK – MT LYELL




FLOW AND RAINFALL
800
700
600
500
400
300
200
100
0
0
4
8
12
16
20
24
28
+ FLOW
Δ AV RAIN
X 1992 RAIN

4.2 Bacterial Leaching Process

The copper and iron in solution in the various streams originates from bacterial leaching of the sulphide minerals contained in the waste rock from the mining areas. The bacteria are thermophyllic, acidophyllic and chemolithotropic (thrive in low pH, elevated temperatures and do not require organic input other than natural sources to construct their basic protein structure). The dominant variety are Thiobacillus Ferro-Oxidans (TFO), a unicellular rod shaped bacillus (class Achizomycetes, order Pseudomonodales) which are essentially ubiquitous (create the appropriate conditions and they appear), and are about 0.4 microns diameter and 1 micron long. TFO coexist with other species which may or may not play a role in the overall leaching process.

TFO thrive in a mixed aqueous and solid environment in the temperature range 30 to 37 degrees centigrade and in solutions with pH of 1.2 - 2.2. Their metabolism is based on enzyme action using complex amino acids to catalyse the oxidation of iron (from ferrous to ferric - loss of an electron) from which they derive their energy. They adapt to a specific environment over time, and their population growth is governed by supply of air, availability of mineral surface and the presence of any inhibiting materials such as surfactants or acid consumers. Other bacteria assist in the fixing of nitrogen and possibly in the oxidation of sulphur to sulphate, but the essential ingredients are ferrous iron, air and water. TFO attach themselves from the aqueous phase to a mineral surface - possibly hydrophobically by lipids in their outer membrane. The mineral surface is made up of iron and sulphur (pyrite) or iron, sulphur and copper (chalcopyrite). The internal reactions generate an electron :

$$2Fe2+ + 0.5O2 + 2H+ \rightarrow 2Fe3+ + H2O + e-$$

and possibly:

$$2S + 2H2O + 3O2 \rightarrow 2H2SO4$$

The overall reaction starting with pyrite may be:

$$2FeS2 + 7O2 + 2H2O \rightarrow 2FeSO4 + 2H2SO4$$
$$4FeSO4 + O2 + 2H2SO4 \rightarrow 2Fe2(SO4)3 + H2O$$
$$2Fe2(SO4)3 + (CuFe)S2 \rightarrow CuSO4 + 5FeSO4 + 2S$$

Curiously, a better understanding exists of the internal enzyme processes than the bulk mineral leaching due to the complex ionic environment. Various galvanic effects have also been proposed which participate in the leaching process (between different minerals). An investigation by ANSTO at Mt Lyell has suggested that the leaching rate is controlled by air supply, which in waste dumps derives from the thermal convection established in the dump due to the temperature profile set up by the exothermic oxidation reactions. The upward movement of warm air draws in fresh air from the bottom and sides of the dump structure. The presence of bacteria, temperature profiles and obviously high levels of ferric iron have been established in the West Lyell waste dumps.

4.3 Waste Dump Leaching

The West Lyell dumps are formed from waste rock deriving from the operation of the West Lyell open pit between 1930 and 1972. An estimate of the tonnage and grade has been compiled by difference. The tonnage and grade mined are documented (58 million tonnes at 0.72% copper) and the original resource was well defined by drilling and a kriged block model using a cut off grade of 0.4% copper. The resulting estimate for the waste dumps is:

tonnes	%Cu	range %Cu
16,251,650	0.0	0.0
985,565	0.08	0.0 - 0.1
6,751,461	0.1	0.1 - 0.2
9,833,789	0.26	0.2 - 0.3
11,286,020	0.35	0.3 - 0.4
48,108,505	0.17	

Previous external estimates had derived a figure of 59 million tonnes at 0.20% Cu. There is evidence to support a higher grade as ore was sometimes dumped to waste when the stockpile areas were full. It is reasonable to assume a copper metal resource of 81,784 tonnes.

The leaching of chalcopyrite, the dominant copper mineral in the waste dumps, is a very slow process and in laboratory simulations usually displays a flat and linear extraction of copper as a function of time. At present the stream flow monitoring suggests that the dumps are releasing an average of 20 grams of copper per hour. If this process has been underway since 1930, a total of 18,000 tonne of copper has been lost from the dumps from a total resource of 82000 tonnes copper. This equates to an extraction of 22%, or 0.35% copper extraction per year on average. This estimate is most likely excessive, as many years would have been required to establish the leaching process at the low ambient temperatures, and waste was dumped up to 1974 thus staggering the start of the leaching process over 44 years. Recalculation on this basis (linear dump construction at 1 million tonnes per year at 0.17% Cu, 0.35% extracted per year since) the tonnage extracted to date is close to 11,000 tonnes copper, equating to 13% extraction of the total copper metal in the dump.

A feature of these dumps indicated by drilling is that there is very little water retained inside. There is also some evidence that the total copper flow is directly related to the rainfall pattern. The peak rainfall period is May through September, which coincides with the peak copper flows. This leads to the supposition that regular irrigation may enhance the rate at which copper is extracted. At present any rainfall appears to be flushing soluble species - iron, aluminium, copper and sulphate - from the dumps. This is supported by observation; on wet days the iron precipitation from the leach liquors is visibly greater. A full scale extraction plant should incorporate the facility to recycle leach solution or raffinate back onto the heaps between rainfall events, gaining greater use of the contained acid and ferric sulphate, which normally is effective on a "once through" (flushed) basis only before flowing down the creek. Figure 4. shows a projection of the leach curve implied by the stream sampling data.

The stream flow data on haulage creek is useful only to a limited extent. It indicates the total copper exiting the West Lyell dumps from the action of rainfall, but does not give a realistic flowrate estimate due to the large volumes of surface runoff also carried by the creek. A full scale plant would tap feed from the toe of each of the two main dumps - known respectively as the shaft and haulage dumps. The shaft dump is higher in grade than the haulage, especially in iron. This is possibly due to the surface runoff which percolates through the Prince Lyell shaft ore stockpile, which is located above the waste dump area, and joins the drainage emerging from the toe of the waste dump. The haulage dump stream emerges south of the shaft dump drain point and probably contains contaminants due to its use as a garbage tip. It is intended to instal a collection system at the two dump toes to minimise the degree of dilution by surface runoff, and possibly allow the recycling of solution back onto the dumps during periods of lower rainfall.

4.4 Underground Potential

The stream flow data on the conveyor tunnel reveals a very similar copper metal flow (0.72 tonne per day Cu) to the waste dumps (0.78 tonne per day Cu). This must result from bacterial leaching in the caved stopes above the 60 series production areas which extend to the base of the West Lyell open pit. This stream is pumped from the underground sumps, and contains a relatively high level of suspended solids. The situation would change should the mine be shut down and allowed to flood.

This study makes no attempt to quantify the resource in the underground as it is virtually impossible to characterise the way in which solutions and air move through this area (and thus determine how much material is exposed to the leaching process). Should the mine close, further study of the mine hydrology would be recommended to determine how best to manage the underground for the purpose of sustaining, or perhaps enhancing, the leaching process. Very large tonnages of broken rock are contained in the caving areas, and with low grades and slow leaching rates it is unlikely that any expenditure underground to create more leaching zones could be commercially justified. Investigations into insitu leaching are underway at the Tasmanian Department of Mines. These studies may develop a methodology which will increase the rate of leaching in the underground, but would require a major technical breakthrough to increase the tenors much beyond their current values.

5.0 PROJECT DESCRIPTION

5.1 Waste Dumps and Mine

The waste dumps are formed on the western and southern rim of the West Lyell open pit, which is now a caving area due to the underground mining activity since closure of the open cut operation in 1974. The rain falling on the dumps drains through and emerges in two primary areas - adjacent to the Prince Lyell shaft collecting the effluent from the shaft dump, and to the east in a gully running northerly which collects the catchment of the haulage dump. Both emerge in the same major valley and currently flow down haulage creek to the existing plant area and into the Queen River. The two flow streams will be isolated at the dump toe and piped down the valley in a general westerly direction to the proposed process plant area near the existing mine office complex.

For the purposes of this study it has been assumed that the operator will be permitted to establish the plant in the vicinity of the existing mobile underground equipment workshop near the mine office, or in a lower area near the existing crushing plant. The solution stream from the conveyor tunnel (at the discharge end of the overland conveyor) will be piped across to the process plant area. A small pump may be required for this transfer depending on the final location of the plant feed pond. An extra settling pond may be required to remove suspended solids if the mine water continues to be pumped from the underground sumps .

5.2 Process Plant

The process plant is fed from an on-ground plastic lined pond which receives the various copper bearing streams, and which is sufficiently elevated to enable gravity flow to the plant. The pond includes a spillway and diversion system such that any excess flow during storms , or the entire flow, can be diverted to haulage creek. A pipe running through the dam wall delivers the Pregnant Leach Solution (PLS) to the first extraction mixer settler at a rate regulated by a flowmeter and control valve. The mixer settler is a conventional design with two mix stages overflowing to a settler, passing through a picket fence and separating the organic and aqueous phases by adjustable launders. The mixer is fed through a draft tube, with the first stage mixing PLS and organic, which is gravity fed from the organic weir of the second extract stage. The aqueous from the first settler is gravity fed to the second mixer/settler unit and the aqueous phase from this stage is gravitated to the raffinate after-settler, where any residual organic phase is recovered by use of a packed bed coalescence zone.

The loaded organic from the first stage of extraction is gravity fed to the loaded organic tank, where a coalescence zone enables the separation of any entrained aqueous phase. This aqueous phase is pumped from the bottom of the loaded organic tank to the crud decant system. The clean organic phase is pumped from a separate compartment to the strip circuit, where the copper is extracted into electrolyte and the organic returned as feed to the second stage extract mixer/settler. The organic from the second mixer/settler is fed to the first stage of extraction.

The extract and strip circuits are run on a 1:1 organic to aqueous ratio in the mixers, although the O/A ratio in the system is adjustable. This depends on the level of organic recycle required in the extraction stages, determined by the loading on the organic phase required to achieve the target copper transfer. This arises from the very low PLS grades relative to conventional practice, and the need to run a higher fraction of extractant in the organic phase than required by the copper transfer rate. This high extractant level is determined primarily by mixing efficiency limitations below a 4% by volume extractant level. The mass balance and flowsheets illustrate the flexibility of the system, with the mass balance showing the particular configuration which was successfully employed in the pilot plant tests. The system has been designed to permit a wide range of operating practice as it is probable that a more efficient system will be evolved in time on the full scale plant.

Crud formation was shown by the pilot plant work to be a problem which can be simply dealt with. The bulk of the pilot plant campaign was devoted to development of methods to manage the collection and disposal of crud with low levels of organic loss. This was achieved by use of an air curtain in the mixer settler to force the crud out of the aqueous phase to the dispersion band. A floating boom was required to control the increased aqueous entrainment in the organic phase. A pump suction manifold mounted behind the organic weir, the height of which could be adjusted to optimise the crud transfer, enabled a diaphragm pump to transfer the crud accumulation to the crud decant tanks. Any organic phase withdrawn from the mixer settler is rapidly recovered and recycled from the top of the decant tanks, and clean aqueous phase despatched to the raffinate after settler. The crud band is transferred to an agitated tank and mixed with diatomaceous earth. This batch process gives complete separation of the crud to the diatomaceous earth, allowing rejection of the aqueous residue and recycling of the recovered organic. The crud is readily and rapidly removed from the diatomaceous earth (DE) by water washing. A filtration step was also successfully employed to separate the DE and organic, and has been included in the flowsheet to provide the option of more rapid processing.

Two portable diaphragm pumps are employed to transfer the various solutions from tank to tank. Maximum use of gravity flow has been made in the plant layout, with air entrainment in launders eliminated by careful selection of relative levels and the use of goosenecks in the piping layout. Regulating flow valves are used on recycle streams and the organic advance to strip.

The advanced electrolyte from the advanced electrolyte tank (AET) is pumped at a controlled rate, regulated by a flowmeter and control valve, to the electrowin (EW) feed. The electrowin is comprised of 20 lines each containing 20 cells in packs of 5, a total of 400 cells. Each line is fed in parallel off a manifold from the advance electrolyte pump, with the flowrate to each line ranging from 2500 to 3600 litres per hour. The spent electrolyte is returned through a back pressure valve to the spent electrolyte tank.

The EMEW cells are mounted in packs of 5, enabling simple electrical and hydraulic connections, and are removed as packs for washing with hot water in a booth and stripping of the cathode in a proprietary jig. This is achieved by removal of the end caps and anodes, knocking out of the inner stainless steel sleeve which carries the cathode deposit and reloading a new sleeve. With the anodes and end caps refitted the pack is ready to be reinstalled in the winning circuit after a very short turnaround time. The copper cathode is removed from the inner sleeve very rapidly as it can be opened down its length. The electrical connections can be manipulated to suit the current and voltage characteristics of the rectifier, but for the purpose of this study each flow line is in parallel, with every cell in the line in series, and the electrical connections follow the hydraulic connections. The flexibility of this system is effectively unlimited, but the concentration change between advance and spent has been minimised for this study to maintain simplicity.

With 20 cells per line and a design voltage drop of 2.8 volts per cell, the overall voltage requirement is 56 volts. The current per cell is 140 amps, and with 10 lines per rectifier the overall current requirement is 1400 amps. The rectifiers will have extra capacity to cope with higher current density operation when higher tenor solutions are treated.

From the spent electrolyte tank a proportion of the flow is pumped to the feed of the strip mixer/settler for contacting with the loaded organic. The remainder is recirculated to the advance electrolyte tank to maintain the EW feed flow requirement. A bleed stream is pumped from this recirculation stream to the first extract stage to maintain the iron level in advanced electrolyte below 6 g/l. The loss of acid to raffinate is made up by fresh acid addition to the spent electrolyte (and the copper is extracted in the normal manner as the bleed stream is effectively a component of the PLS feed).

The acid lost in the electrolyte bleed is a significant cost to the operation. A dialysis technique, developed for recycling of acid pickling solutions, will be trialled on the bleed stream to recycle the acid component. An electrolysis process utilising membrane technology is also under development for the separation of acid from concentrated solutions, and may prove to be more effective than the dialysis route.

The recycling of raffinate onto the waste dumps has not been included in the scope of this study. The practice would be intermittent, particularly productive during the drier months, and would be relatively inexpensive. The power requirement would be approximately 0.19 kW/cubic metre per hour based on a 50 metre lift. This cost would be justifiable if the solution tenors and flows could be stabilised during the drier low flow periods, and if the recycle stimulated the leaching rate. This could be trialled at full scale once the plant was commissioned and the project cash positive. This recycle would be sourced from the effluent containment dams at the toe of the waste dumps.

5.3 Process Services & Facilities

Process services and facilities are primarily systems for the supply of power, water, compressed air and reagents to the SX-EW process. This study assumes that power is available and can be taken from a 6.6kV board at the boundary of the process plant area. Water is not required in large amounts, but the quality is important, and the study assumes that a potable quality supply is available at the plant boundary (primarily for domestic use, washing of cathodes and cleanup purposes). The solution balance is shown in the process mass balance calculations in Table 2.

Compressed air is provided by a dedicated compressor, and is utilised for driving diaphragm pumps, the settler air curtain, and vacuum cleanup devices used for removing crud accumulations. A diesel driven pump is provided for failsafe firewater supply under power outage and will cut in on a pressure switch if the external water supply system fails. This pump is connected to a water main, which is fitted with sprays at each corner of the SX plant arranged to be capable of laying a curtain of water over the mixer settlers and process tankage. The system is also backed up by connection to the PLS pond.

Reagents used in the process are high flashpoint kerosene stored in a tank supplied by the vendor and resupplied by bulk tanker, the specific copper extractant which is stored in 200 litre drums and the concentrated acid for makeup to the electrowinning circuit. The acid is stored in a mild steel tank which conforms to well established occupational health and safety regulations, with specifically configured filling, venting and overflow pipework and all contained in a bund designed to hold the entire tank contents. The acid is mixed with water in a stainless steel tank, with the water injection used to achieve adequate mixing prior to transfer by dosing pump into the electrowin circuit. The dosing pump can also be used in recirculation mode to improve mixing, and is designed to withstand the temperatures experienced in this operation. The concentrated acid is delivered to the mixing tank by a solenoid diaphragm pump. Diatomaceous earth for crud processing is stored in 25 kg bags on wood pallets, and can be manually added to the crud circuit as required..

TABLE 2.

EW MASS BALANCE

	FLOW M3/HR	g/l Cu	KG/HR Cu	g/l acid	KG/HR acid	g/l Fe	KG/HR Fe
ADVANCE EX STRIP	76.2	25.70	1958	179	13633	6.01	458
ADDITION	0.107			198	21.309		
ADVANCE ELECTROLYTE	76.3	25.66	1958	179	13654	6.00	458
EW FEED	72.0	25.71	1851	179.0	12886	6.00	432
NO EW LINES	20						
Cu WON/ACID MADE			50.998		78.719		0.000
RECIRCULATED SPENT	-4.3	25.00	-107	180	-768	6	*******
SPENT EX EW	72.0	25.00	1800	180	12957	6	432
BLEED	0.093	25.00	2.325	180	16.740	6	0.558
SPENT TO STRIP	76.2	25.00	1904	180	13712	6	457

WHERE RECIRC.D SPENT IS NEGATIVE - THIS EQUATES TO THE STRIP AQUEOUS RECYCLE FLOW

MASS BALANCE OVERALL

	FLOW M3/HR	Cu g/l	Cu kg/hr	acid g/l	acid kg/hr	pH	Factors
PLS	500	0.120	60.0	0.28	138.2	2.55	0.85 recovery
iron bleed	0.093	25.0	2.3	180	16.7		
SX feed	500	0.125	62.3	0.03	154.9	3.47	
raffinate	500	0.018	9.0	0.47	233.7	2.32	1.54
loaded organic	500	1.50	750.0				1 org/aq
ext. org. recycle	424	1.50	635.7				
organic advance	76	1.50	114.3				
stripped organic	76	0.800	60.9				
delta organic		0.700	53.3				
advance electrolyte	76	25.7	1957.7	179	13633		
copper transferred			53.3				
copper production			51.0				98.076
spent electrolyte	76	25.0	1904.4	180	13712		delta

A compact laboratory for routine monitoring of the process is incorporated into a site office and store building. Also included in this building is a workshop for maintenance of proprietary cell stripping equipment, pump and agitator maintenance, and an ablutions facility. The laboratory will have a small Atomic Absorption Spectrometer (AAS) unit, and glassware appropriate for the routine testing of phase disengagement characteristics, entrainment levels and the wet chemistry methods required to back up the AAS. The wet chemistry methods will also enable monitoring of the leaching processes occurring in the waste dumps and underground.

5.4 Infrastructure

For this study, the infrastructure is assumed to be essentially available as it exists at present. The future of Mt Lyell and the site facilities have not been clearly determined at this time, thus the existing access, power, water and communications capability has been adopted. The infrastructure at the Mt Lyell site is comprehensive and in excess, in every aspect, of the needs of this project.

5.5 Expansion Capability

The plant is expandable in a modular form in the EW area. The solvent extraction plant can cope with an increase in tenor up to 1 g/l copper without any changes in reagent levels or consumption.

6.0 ENGINEERING

6.1 Pilot Plant and Process Design

The design of the pilot plant SX circuit utilised well established parameters which had been obtained from the literature and the various SX operations in Australia. Shakeout tests, a laboratory simulation of the extraction and stripping capability of the organic phase, were difficult to interpret for several reasons. This was primarily due to the very low levels of extractant in the organic phase compared with standard operating practice. These tests were conducted by the suppliers of the copper specific extractants on solutions taken from the Mt Lyell mine site.

Conflicting data was obtained from the laboratory scale SX tests, and the pilot scale plant was designed to accommodate several operating strategies. The copper levels in the feed solution were more than 10 times lower than most operations, however the extractant level in the organic phase could not be reduced to less than 4% by volume - about seven times more than was required. This was due to the realities of mixing, where below 4% the volume of extractant is too low to enable good contact with the aqueous phase within the limits of the mixing regime. This limit is the production of stable emulsions, which would impair the disengagement of the two phases in the settler. Consequently, the copper loading of the organic phase was very low and to minimise the volume of organic to be stripped the facility to recycle organic in the extraction stage was incorporated.

An advantage of this excess of extractant in the circuit is that during periods of high copper tenor in plant feed, there will be adequate reagent in the system. Otherwise an increase in the extractant inventory would be required, a practice which is not suitable for recovering short term increases in the copper in feed solution.

The pilot strip mixer settler was designed to accomodate organic at flows equivalent to the PLS (feed - pregnant leach solution) to the extract stage. This was in case the organic loadings did not increase with recycling and the full flow had to be stripped to achieve the target copper transfer. This scenario was taken seriously because of the high iron levels in the PLS, and the tendency for iron to load onto the organic in the absence of copper. This scenario did not eventuate and the target loadings were acieved by recycling. The selectivity of copper extraction over iron achieved was acceptable, despite the high levels of extractant.

The problems encountered during the pilot scale trials were not those anticipated from the testwork. The effect of temperature was more severe than expected on the phase disengagement, and the volumes of amorphous silica and iron hydroxide based "crud" phase were high. The process of copper extraction and stripping worked well, achieving 70 to 90% extraction in one stage under steady conditions. The focal effort of the pilot trial was to deal with the crud formation, which caused very large losses of organic phase to the raffinate (effluent from the process - PLS after copper extraction).

The solution developed has been incorporated into the design proposed in this study for the full scale plant. This revolved around running the extract stage aqueous continuous so the crud formed on the walls of the aqueous dispersion. It was observed that a shock would cause the amorphous crud "cloud" to consolidate at the phase boundary in the settler, so an air curtain was installed at the picket fence with a boom floating on the organic phase to retain the aqueous phase carried to the surface by the air bubbles. It is anticipated that this will be better performed by vibration of the picket fence in the full scale plant. The accumulation of crud at the phase boundary is then pumped from the dispersion band to a separation process developed to reclaim the organic content.

The design area of the settlers has also been modified for the full scale plant due to the slower phase disengagement times experienced at the low temperatures prevailing during the pilot testing. A standard factor is 5 m3/hour of settler feed per square metre of settler plan area. The proposed design is based on 3 m3/hour/m2. The extraction levels obtained from the single stage of extraction in the pilot tests will be higher in the full scale plant as two stages of extraction will be employed.

The separation process entails the decanting of the organic which readily disengages from the crud, and a diatomaceous earth medium to absorb the remaining crud and enable either filtration or organic continuous medium mixing separation of the remaining organic phase. These processes were conducted at pilot scale and were successful. A further precaution was taken to minimise organic losses by use of an after settler on the raffinate to trap any entrained organic, and at full scale this final separation will be enhanced by the use of coalescing media - a practice which was developed in the Chilean SX/EW operations and is now being introduced to the Australian industry. This technique will also be used on the organic phase to minimise aqueous entrainment (which carries iron into the electrolyte).

The remainder of the plant design is very simple. A major advantage of the EMEW process is its tolerance for high iron levels and entrained organic. There is no opportunity for the organic to burn onto the cathode and cause poor deposition morphology. Thus there is no requirement for the filtration or flotation scavenging of trace organic from advance electrolyte. The EMEW process current efficiency is very high at high copper concentrations, thus a heating and heat exchange system has been provided for in the plant layout but not included in the cost estimate. The bulk of the spent electrolyte is recirculated with a small volume diverted to the strip mixer, thus most of the heat generated by the EW is conserved.

6.2 Civil Works

Although the study assumes a flat site for the process plant, there is a requirement for considerable civil works. These take two forms. The procurement, transport, placement and compaction of fill for the PLS pond walls, the dump toe dams and the base for the mixer settlers will be obtained locally. Concrete will be used for the mixer settlers, the raffinate flume and coalescer box, and as a base base for the electrowin plant, pumps and tanks. Excavation has been avoided in the design as it is likely that this would be expensive. The exception is the area in which the extract stage mixers stand, which is excavated to 1 metre depth and is concrete lined with polyethylene protection.

The concrete construction of the mixer settlers will be achieved by use of prefabricated reinforced panels, protected from the acidic solutions by a 3mm layer of polyethylene in the extract settlers and 6mm polyethylene in the strip settler and all mixers. The panels will be 2.5 metres high, placed on a concrete strip foundation at ground level, and supported by 1.5 metres of fill on each side of the panel. This leaves a metre of concrete wall exposed and a sand lined base which are polyethylene lined to form the settler. This approach was taken due to the large size of the extract settlers and the difficulties associated with transport and assembly of fibreglass structures on-site. The fill platform is extended around the outside of the settler to form access ways. The mixers are constructed of prefabricated bolt fastened panels in a hexagonal geometry using a system developed for water tanks, with the pipe feed base section isolated from the mixer zone by a polyethylene false bottom.

The raffinate flume has been costed in concrete with a 1mm polyethylene liner. It is possible that this will be excavated if the ground conditions are suitable, and lined with polyethylene. Concrete sections may be used where culverts are required for traffic areas.

6.3 Structural

The only structural components of the plant are wooden walkways over the extract stage mixer boxes and the building which incorporates the office, laboratory, store, workshop and ablutions. This is comprised of a series of four transportable units internally configured to each requirement. A basic enclosure is provided for the power board and the rectifiers. The electrowinning modules are supplied with their structural framework, which carries the DC busbars. All pipework is on ground.

6.4 Mechanical

The mechanical equipment in the plant is detailed in the equipment list in appendix 5. By setting the tankage and mixer settler's operating levels at a common elevation, most flows do not require pumping. The pump mixers in the first stage mixer boxes are capable of lifting 200mm above operating levels to ensure against backmixing from the feed distributor in the settler, and to compensate for any friction head losses. All pipework has been sized to achieve a fluid velocity of 1 metre per second or less, thus minimising losses.

All impellors are designed for the vessel in which they are fitted and the specific duty. The pump mixers have variable speed drives, as do the dosing pumps for acid mixing and addition to the circuit and the iron bleed pump. All equipment is selected for resistance to acidic attack. A compressor is provided to drive the diaphragm pumps, the output of which can be regulated by a combination of air flow and discharge choking, to provide an air curtain in the settlers, and to drive tools. A small mobile hoist with manual hydraulic drives is employed for carrying EW packs and reagent drums.

6.5 Tankage

Most of the tankage is constructed from high density polyethylene, with fittings and baffles fusion welded into off the shelf units. This material is ideal for acidic solutions, is simple to instal and modify. The acid mixing tank is constructed in 316L stainless steel due to the high temperatures which can occur in the mixing process.

The concentrated acid tank is of mild steel construction and enclosed in a concrete bund to meet safety requirements. The diluent (high flash point kerosene) is also stored in a steel tank which is provided by the diluent supplier.

As mentioned above, the mixer boxes and the raffinate after settler/coalescer are constructed in polyethylene lined concrete. This strategy has been employed due to any or all of size, geometry and rigidity requirements which cannot be provided in polyethylene alone.

6.6 Electrical

The electrical system is simple due to the very few drives in the plant, and will be internally distributed on cable racks from a power board located in the rectifier enclosure. No allowance has been made for transformers or motor control centres, as the power required is either 415v 3 phase or domestic. The demand will not exceed **246** kW, based on a 70% draw factor from an installed poer of **351** kW.

The electrowinning circuit is driven as two independent sections, with regard to both electrolyte pumping and rectifiers. Each rectifier is designed for a base load of 100kW, are air cooled and will be current variable. It is not anticipated that capacitance filtering will be required. The type and capacity of rectifier specified are readily available both new and second hand.

6.7 Piping & Valves

Most pipework is in high density polyethylene class 3, with flexible segments to prevent undue stresses on fittings from pipe movement when temperature changes. The exceptions are diluent lines and the concentrated acid lines, both in 316 stainless steel, and the interconnecting pipework within the EW circuit. These are a flexible, wire reinforced plastic.

Valves used for isolation are plastic ball type, with the major flow regulating valves and isolation valves in the SX circuit being viton coated butterfly type where a large pipe bore is utilised. The electrowinning regulating valves are ball valves, with fast release couplings to enable removal and installation of the EW cell packs.

6.8 Controls

The control system proposed for the study design is confined to the flow regulation of PLS, organic recycle in extract, organic advance to strip, the spent electrolyte flow to strip and the flowrate to electrowin feed. These will be manually regulated on the basis of turbine flowmeter signals. The use of more sophisticated flow measuring devices, such as ultrasonics, will be tested after plant commissioning. The installation of closed digital control loops has been considered, however this will be delayed until a thorough knowledge of the flow dynamics and optimum flow regime has been obtained, and the higher cost techniques can be justified.

A proprietary control and alarm system will be supplied with the electrowin system. This will focus on detecting short circuits and abnormal plating phenomena by monitoring voltage and current at various points in the rectifier/cell system This information will also be used for calculation of process performance. The rectifier and solution temperatures will be monitored, alarmed and logged.

The electrical conductivity of the contents of the mixer boxes will be monitored continuously and alarmed to alert the operators of phase flipping. This is essential in the organic continuous regime used in the extract stages. A sensor will also be installed in the loaded organic tank to prevent the loss of organic phase in the aqueous stream coalesced from the loaded organic, and linked to an alarm and isolation solenoid valve on the aqueous bleed line.

7.0 ENVIRONMENTAL IMPACT

7.1 Effluent Definition & Control

The project has a net environmental benefit, with negligible environmental risk associated with the process. The removal of the copper from the effluent streams creates a stoichiometric addition of sulphuric acid to the existing effluent. This acid addition theoretically drops the pH of the plant effluent by approximately 0.15 pH units, however the acid is rapidly consumed by the extraordinarily high iron hydroxide content of the effluent. This is in turn massively diluted by the local drainage stream into which the plant effluent (raffinate) is directed.

7.2 Regional & Local Drainage

The regional and local drainage will not be impacted in any significant way by the proposed plant. Some local drainage will be diverted around the facility, but the proposed process simply diverts copper bearing liquors, removes the copper, then adds the same flow back into the drainage system.

7.3 Noise, Dust & Land Use

The technology in use is effectively silent. Being a hydrometallurgical plant, there will be no dust emmissions other than from road traffic on the plant access road. A relatively small amount of space is required for the plant (40 by 40 metres). The feed (PLS) pond requires a separate area of 60 by 60 metres. It is anticipated that existing level sites will be utilised within the Mt Lyell lease.

7.4 Flora & Fauna

This issue is not addressed by this study. A preliminary impression is that flora and fauna will not be a significant issue in relation to the plant area.

TABLE 4.

MEASURED pH - PROJECTED CHANGE IN ACID CONCENTRATION

FEED TYPE	DATE		pH PLS	pH RAFF	ACID g/l PLS	ACID g/l RAFF	ESTIMATED g/l Cu EXTRACTED
	NOV 1993						
SHAFT		12	2.4	2.19	0.39	0.63	0.159
SHAFT		13	2.36	2.17	0.43	0.66	0.153
SHAFT		14	2.47	2.23	0.33	0.58	0.160
		16					
		17					
HAUL		18	2.46	2.54	0.34	0.28	-0.037
HAUL		19	2.45	2.2	0.35	0.62	0.177
HAUL		20	2.69	2.63	0.20	0.23	0.019
MIX		21	2.23	2.26	0.58	0.54	-0.025
		22					
		23					
		24					
MIX		25	2.14	2.09	0.71	0.80	0.057
MIX		26	2.21	2.15	0.60	0.69	0.059
MIX		27	2.16	2.09	0.68	0.80	0.078
MIX		28	2.32	2.18	0.47	0.65	0.117
MIX		29	2.35	2.23	0.44	0.58	0.091
MIX		30	2.36	2.24	0.43	0.56	0.089
	DEC 1993						
MIX		1	2.34	2.23	0.45	0.58	0.084
MIX		2	2.28	2.2	0.51	0.62	0.068
MIX		3	2.2	2.1	0.62	0.78	0.105
NLT		4	2.3	2.21	0.49	0.60	0.074
		5					
NLT		6	2.13	2.01	0.73	0.96	0.151
NLT		7	2.13	2.09	0.73	0.80	0.046
NLT		8	1.99	1.9	1.00	1.23	0.151
HAUL		9	2.36	2.24	0.43	0.56	0.089

8.0 OPERATING COSTS

8.1 Fixed Operating Costs

The fixed operating costs are shown on tables at the back of this section, and reflect the labour and overhead component of the operation. This includes maintenance and cost of services such as purchasing, the laboratory and communications. The labour rates and cost burden are generous. The materials for the mechanical and civil aspects of the operation have been carefully chosen and the maintenance allowances are considered to be very conservative.

8.2 Variable Operating Costs

The variable costs are essentially consumables and the EW power. Strictly speaking the pumping and auxiliary power usage is a fixed cost as the project does not anticipate increases in flowrate beyond 500 m3/hour. Any expansion in production will come from improved extraction or increases in copper concentration in process feed, which will increase the EW power on a variable basis. The SX reagent consumption costs are also fixed up to the point of feed copper tenors of 1 g/l (1000ppm) due to the excess of extractant in the organic phase. These costs have been calculated on the basis of the process design calculations and mass balance - which are based on the performance levels achieved at the pilot scale.

8.3 Accuracy

The accuracy of these costs is a function of the quantities determined in the mass balance and the unit cost for each item. Typical values have been used for consumables, usually with some contingency to account for the fluctuations experienced with items such as petrochemical based reagents and acid. The cost of power is yet to be finalised, however the values used reflect the current industrial tariffs in the Tasmanian mining industry.

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EURALBA MINING LIMITED
MT LYELL SX/EW COPPER PROJECT
OPERATING COST SUMMARY

COST CATEGORY	COST A$/YEAR
FIXED COSTS	
LABOUR	310128
MAINTENANCE & SERVICES	44575
VARIABLE COSTS	
CONSUMABLES	175513
TOTAL	530216

~ 74 people?

A$ PER KG OF COPPER PRODUCED	1.26
US$/lb OF COPPER PRODUCED	0.40
EXCHANGE RATE - (US$/A$)	0.7

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EURALBA MINING LTD
MT LYELL SX/EW COPPER PROJECT
OPERATING COSTS

FILE: OPC01

HOURS/YEAR	8387
PRODUCTION	419
INST kW	197

CONSUMABLES - VARIABLE COSTS

AREA/ITEM	CONSUMPTION	COST/UNIT	$ COST	UNITS	
ACID	179	180	32154	TONNES	
DILUENT	60803	0.34	20673	LITRES	
EXTRACTANT	2516	15	37740	LITRES	
POTABLE WATER	9691	0.25	2423	MLITRES	
EW POWER	1651823	0.04	66073	kWH	3.9
GEN POWER	411270	0.04	16451	kWH	197
TOTAL			175513		
			A$/KG		0.4

LABOUR - FIXED COSTS

CATEGORY	NUMBER	OVERHEAD	RATE	COST
SUPERVISOR	1	1.35	16	44928
DAYWORK	1	1.25	12	31200
SHIFTWORK	6	1.25	15	234000
TOTAL				310128
			A$/KG	0.74

Average rate looks low for shiftwork - $25000 p.a.

MAINTENANCE AND SERVICES - FIXED COSTS

ITEM	RATE	COMMENT	COST
MECHANICAL	4	ANNUAL % VALUE	11983
ELECTRICAL	6	ANNUAL % VALUE	4067
CIVILS	5	ANNUAL % VALUE	1325
LABORATORY		AAS - GLASSWARE - CHEMICALS	15000
OFFICE		STATIONARY/SOFTWARE OFFICE EQUIPMENT	5000
PURCHASING			6000
COMMUNICATIONS			1200
TOTAL			44575

9.0 CAPITAL COSTS

9.1 Direct Costs

Direct costs have been estimated either by quotes from manufacturers, the application of current cost rates to quantities of materials derived from a specific design, or by the application of factors based on common practice to a known cost. An example of the latter is the estimate for electrical distribution costs, where the estimate is derived as 12% of the equipment cost. The piping costs were estimated by sizing the pipelines, calculating the mass of HDPE required and costing this at $3.00 per kg (compared with current rates of $2.40/kg).

9.2 Indirect Costs

The indirect costs are engineering and project management services, a contingency factor plus insurance and fees. These are estimates based on experience of similar projects, and are subject to review if the scope of the design and construction phase changes.

The amount of contingency allowed is usually directly related to the accuracy of the estimate, and is often different across either disciplines or areas of the cost estimate. The contingency used here is slightly lower than the accuracy estimate, primarily because much of the estimate already has a significant contingency built in. An example of this is the EW cell costs, which allow for the use of seamless tube where the lower cost welded tube is now demonstrated to be adequate. The rates used for concrete and HDPE are conservative, and much of the tankage estimate is based on list prices.

9.3 Accuracy

The overall accuracy of the cost estimates in this study ranges between 12 and 16%. There are several areas in the project concept which cannot be definitively scoped at this time, and thus the accuracy of the cost estimates in the electrical and civil works areas fall into the 20 to 25% range.

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CAPITAL COST SUMMARY

AREA	DESCRIPTION		TOTAL
1	SOLVENT EXTRACTION		226611
2	ELECTROWINNING		199640
3	SERVICES AND FACILITIES		172913
4	CIVIL WORKS		66241
5	ELECTRICAL & INSTRUMENTS	@ 12% of areas 1 - 3	61334
6	PIPING & VALVES		120324
	DIRECT CAPITAL COST		847064
	CONTINGENCY @ 12%		101648
	ENGINEERING, D & C, PROJECT MANAGEMENT		84706
	INSURANCE, FEES		8471
	INDIRECT CAPITAL COST		194825
	TOTAL CAPITAL COST		1041889

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MOUNT LYELL COPPER PROJECT
CAPITAL COSTS
FILE: MLFCAPCO

AREA 1. - SOLVENT EXTRACTION CAPITAL COST

EQUIP NO.	DESCRIPTION	NO.	COST	FREIGHT INSURE INSTAL	TOTAL
PP01	PLS PRIME PUMP	1	0		0
					0
MS01	EXTRACT M/S 1	1	42456	0	42455.9
					0
AG01	M/S 1 PUMP MIXER	1	19070	0	19070
					0
AG02	M/S 1 AGITATOR	1	4610		4610
					0
MS02	EXTRACT M/S 2	1	42456	0	42455.9
					0
AG03	M/S 2 PUMP MIXER	1	19070		19070
					0
AG04	M/S 2 AGITATOR	1	4610		4610
					0
PP02/03	CRUD REMOVAL PUMPS	2	5602		5602
					0
					0
					0
TK01	LOADED ORGANIC TANK	2	8200	0	8200
					0
PP04	LOADED ORGANIC PUMP	1	0		0
					0
PP05	ORGANIC RECYCLE PUMP	1	7480		7480
					0
PP06	CRUD/AQUEOUS EX LOT PUMP	1	871		871
					0
MS03	STRIP M/S	1	17881	0	17880.8
					0
AG05	STRIP M/S PUMP MIXER	1	13100		13100
					0
TK02A	ADVANCED ELECTROLYTE TANK	1	4100	0	4100
					0
TK02B	SPENT ELECTROLYTE TANK	1	4100	0	4100
					0
PP07	IRON BLEED PUMP	1	2243		2243
					0
PP08ab	ADVANCE ELECTROLYTE PUMP	3	30762		30762
					0
					0
TOTAL COST - SOLVENT EXTRACTION EQUIPMENT					226611

AREA 2. - ELECTROWINNING CAPITAL COST

EQUIP NO.	DESCRIPTION	NO.	COST	FREIGHT INSURE INSTAL	TOTAL
EP01/90	EW PACK (5 CELLS)	90	148500		148500
					0
FM01	FLOWMETERS	4	24000		24000
					0
WH01	WATER HEATER	2	2140		2140
					0
TK10	WASH BAY ASSEMBLY	1	2000		2000
					0
XX01	CELL STRIP - ASSEMBLE JIG	1	3000		3000
					0
RT01/02	RECTIFIER	2	20000		20000
					0
MH02	MOBILE HOIST	0	0		0
					0
					0
					0
					0
					0
					0
					0
					0
					0
					0
TOTAL COST - ELECTROWINNING					199640

AREA 3. - SERVICES AND FACILITIES CAPITAL COST

EQUIP NO.	DESCRIPTION	NO.	COST	FREIGHT INSURE INSTAL	TOTAL
					0
					0
TK03/04	CRUD DECANT TANKS	2	11398		11398
					0
TK05	CRUD EXTRACTOR	1	1030		1030
					0
AG07	CRUD AGITATOR	1	2900		2900
					0
PF01	CRUD/DE FILTER	1	0		0
					0
PP09	G.P. DIAPHRAGM PUMP	2	5602		5602
					0
TK06	RAFFINATE COALESCER	1	9800		9800.44
					0
AC01	AIR COMPRESSOR	1	14000		14000
					0
PP10	FIRE DELUGE PUMP/SPRAY	1	12000		12000
					0
PP11	EXTRACTANT PUMP	2	1360		1360
					0
TK07	DILUENT TANK	1	0		0
					0
TK08	ACID STORAGE TANK	1	5699		5699
					0
TK09	ACID MIXING TANK	1	1500		1500
					0
PP12	STRONG ACID PUMP	1	3473		3473
					0
PP13	ACID DOSING PUMP	1	4651		4651
					0
MH01	MOBILE DRUM HOIST	1	13500		13500
					0
	OFFICE/LAB W'SHOP/STORE	2	40000	20000	60000
					0
	OFFICE/LAB. EQUIPMENT	LOT	26000		26000
					0
					0
TOTAL COST - SERVICES AND FACILITIES					172913

AREA 4. - CIVIL WORKS

EQUIP NO.	DESCRIPTION	UNIT	RATE	QUANTITY	TOTAL COST	COMMENT
	DUMP TOE DAMS	LOT				NOT QUANTIFIED
	SITE CLEAR	LOT				NOT REQUIRED
	ACCESS ROAD	LOT				NOT REQUIRED
	PLS POND EXCAVATION/FILL	M3	5.75	3911	22488	EXCAVATE,PLACE COMPACT
	PLS POND LINER	M2	6.00	4000	24000	1 mm HDPE
	SITE DRAINAGE	M	5	500	2500	PERIMETER DRAIN
	SX CONCRETE (STAIRS/WALLS)	LOT	120	50	6000	MINOR WORKS
	EW CONCRETE PAD	M2	120	60	7200	CELL AREA
	TANKFARM CONCRETE PAD	M2	120	34	4053	TANK BASES
TOTAL COST - CIVIL WORKS					66241	

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LINE AND VALVE LIST COST/KG $3.00

LINE NO.	DESCRIPTION	LENGTH	BORE	KG/M	COMMENTS	COST
	PLS (FEED) SYSTEM					
315/3/HDP1	SHAFT DUMP LINE	1000	297	8.661	DRAIN/COUPLING	25983
315/3/HDP2	HAUL DUMP LINE	600	297	8.661	DRAIN/COUPLING	15590
315/3/HDP3	N. LYELL TUNNEL LINE	300	297	8.661	DRAIN/COUPLING	7795
315/3/HDP4	CONVEYOR TUNNEL LINE	100	297	8.661	DRAIN/COUPLING	2598
450/3/HDP5	PLS TO E1	100	423	17.260	FV/FM	5178
	AREA		0.14			
	PRIME LINE				OPTION	
	RAFF CIRCUIT					
450/3/HDP6	RAFF E1/E2/TK01	6	423	17.260		311
CL/400/1		17	600	$90.00	400X400	918
					CONCRETE LAUNDER	
CL/400/2	TK01 TO HAULAGE Ck	100	600	$90.00	400X400	5400
					CONCRETE LAUNDER	
	ORGANIC CIRCUIT					
400/3/HDP7						
400/3/HDP8	E2 - E1	4	377	13.735	LARGE BORE	165
400/3/HDP9	E1 - TK01	14	377	13.735	TO MAINTAIN	577
400/3/HDP10	E1 - RECYCLE	4	377	13.735	VELOCITY AT	165
400/3/HDP11	S1 - E2	12	377	13.735	1 M/SEC	494
	ELECTROLYTE CIRCUIT					
400/3/HDP12	TK02	10	103	1.140		34
400/3/HDP13	TK02 - S1	6	103	1.140		21
	EW CIRCUIT					
110/3/HDP14	TK02a - EW - TK02b	20	103	1.14	MANIFOLD	68
25/3/HDP15	S1/EW/TK03/TK04	40	21.6	0.12		14
50/3/HDP16	IRON BLEED LINE	10	46.7	0.61	CLASS 9	18
16/3/HDP17	OXYGEN BLEED LINE	210	12.7	0.08	CLASS 9	50
16/3/HDP18	CONC ACID LINE	10	12.7	0.08	CLASS 9	2
25/3/HDP19	MIXED ACID LINE	10	21.6	0.16	CLASS 9	5
25/3/HDP20	DILUENT LINE	10	21.6	0.16	CLASS 9	5
	CRUD CIRCUIT					
50/3/HDP21	FEED	40	46.7	0.61		73
50/3/HDP22	DECANT LINES	20	46.7	0.61	MANIFOLD	37
50/3/HDP23	G.P. PUMPING	40	46.7	0.61	FLEXIBLE	73
	TOTAL LINE COST					65574

VALVE LIST

LINE NO.	DESCRIPTION	NO.	BORE	UNIT CO	FITTINGS	COST
	PLS (FEED) SYSTEM					
315/3/HDP1	SHAFT DUMP LINE	5	297	10	COUPLINGS,1 BLEED	50
315/3/HDP2	HAUL DUMP LINE	3	297	10	COUPLINGS,1 BLEED	30
315/3/HDP3	N. LYELL TUNNEL LINE	1	297	10	BLEED	10
315/3/HDP4	CONVEYOR TUNNEL LINE	1	297	10	BLEED	10
450/3/HDP5	PLS TO E1	1	423	10000	BV,FV,IVs	10000
	PRIME LINE					
	RAFF CIRCUIT					
450/3/HDP6	RAFF E1/E2/TKO1	1	423	3000	IV	3000
CL/400/1	LAUNDER	0	600			0
CL/400/2	TKO1 TO HAULAGE Ck	1	600	3000	IV	3000
	ORGANIC CIRCUIT					
400/3/HDP8	E2 - E1	1	377	2500	IV	2500
400/3/HDP9	E1 - TKO1	1	377	2500	IV	2500
400/3/HDP10	E1 - RECYCLE	3	377	2500	IV	7500
400/3/HDP11	S1 - E2	1	377	2500	IV	2500
		1		4000	RV	4000
	ELECTROLYTE CIRCUIT					
400/3/HDP12	TKO2a - TKO2b	1	377	4000	RV	4000
400/3/HDP13	TKO2b - S1	3	377	2500	IV	7500
		1		4000	RV	4000
	EW CIRCUIT					
110/3/HDP14	TKO2a - EW - TKO2b	4	103	40	IV	160
		2		1000	RV	2000
25/3/HDP15	S1/EW/TKO3/TKO4	60	21.6	20	IV	1200
				600	RV	0
50/3/HDP16	IRON BLEED LINE	1	46.7	30	IV	30
16/3/HDP17	OXYGEN BLEED LINE	1	12.7	20	IV	20
16/3/HDP18	CONC ACID LINE	4	12.7	20	IV	80
25/3/HDP19	MIXED ACID LINE	3	21.6	25	IV	75
25/3/HDP20	DILUENT LINE	3	21.6	25	IV	75
	CRUD CIRCUIT					
50/3/HDP21	FEED	2	46.7	30	IV	60
50/3/HDP22	DECANT LINES	12	46.7	30	IV	360
50/3/HDP23	G.P. PUMPING	3	46.7	30	IV	90
	TOTAL VALVE COST					54750

10.0 COMMERCIAL ANALYSIS

10.1 Basis

The basis on which the financial analyses have been constructed are nominated at the top of the spreadsheet calculations presented on the attached tables. The two critical parameters are copper price and exchange rate. The copper price used for the base case is 85 cents US per pound of copper metal produced - A$2.68 per kilogram at an exchange rate of US$0.70 to the Australian dollar. This is a very conservative projection considering that the copper price on an annually averaged basis fits a 3.5% compound growth rate, and that the locus of the low points on the price plot fitted to a 3.5% simple growth (linear) rate projects to 85 cents for 1994.

10.2 Method of Analysis

The analysis method used is standard Lotus spreadsheet cash flow projections over a term of 10 years on an unescalated basis. A net present value calculation is provided on revenue (before capital costs have been deducted), on cash flow after capital has been deducted which essentially represents an equity financing approach, and on cash flow after debt service and taxation has been accounted for.

10.3 Outcomes

The outcome of the financial analysis is favourable for revenue, equity financing, and for full debt financing after debt service and tax. For the latter scenario the project is cash positive in year 1. of operations.

This implies a very rapid return on investment.

10.4 Sensitivities

The sensitivity analysis is attached to the financial analysis, and examines the sensitivity of the project viability to changes from the base case presented on the following pages. The areas tested were:

1. copper price
2. discount rate
3. capital cost
4. operating cost
5. production rate - whether due to feed grade/rate, process performance or running time gains or deficits.

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EURALBA MINING LTD
MT LYELL SX/EW COPPER PROJECT
FEASIBILITY STUDY MAY, 1994

FINANCIAL ANALYSIS

BASE CONDITIONS		ASSUMPTIONS		
FLOW COPPER TENOR	0.120 g/l	COPPER PRICE	0.85 US$/LB	2.68 A$/KG
RECOVERY	80 %	EXCHANGE RATE	0.7 US$/A$	
CAPITAL COST $	1041889	DAYS/YEAR	364	
PREPRODUCTION COST	100000	HRS/DAY	24	
BASE CASE ANNUAL PRODUCTION		INTEREST RATE	10 %	
FLOWRATE M3/HR	500	DISCOUNT RATE	12 %	
COPPER g/l	0.12	TERM	10 YEARS	
RECOVERY %	80	NPV ON REVENUE	3241679	
COPPER TONNES	419.3 -	NPV ON CFBIT	2112660 A$ - REVENUE LESS CAPITAL OUTLAY	
OPERATING COST FACTOR	1	NPV	1980367 A$ - AFTER DEBT SERVICE AND TAX	

ALL IN A$
PRODUCTION

	YEAR 0	YEAR 1	YEAR 2	YEAR 3	YEAR 4	YEAR 5	YEAR 6	YEAR 7	YEAR 8	YEAR 9	YEAR 10	TOTAL
FLOWRATE M3/HR		500	500	500	500	500	500	500	500	500	500	
COPPER g/l		0.12	0.12	0.12	0.12	0.12	0.12	0.12	0.12	0.12	0.12	
RECOVERY %		72	78	80	80	80	80	80	80	80	80	
COPPER TONNES		377.4	408.8	419.3	419.3	419.3	419.3	419.3	419.3	419.3	419.3	4140.9
INCOME		1010017	1094185	1122242	1122242	1122242	1122242	1122242	1122242	1122242	1122242	11082135
FIXED COSTS		354703	354703	354703	354703	354703	354703	354703	354703	354703	354703	
VARIABLE COSTS		157962	171125	175513	175513	175513	175513	175513	175513	175513	175513	
REVENUE		497353	568357	592026	592026	592026	592026	592026	592026	592026	592026	5801914
NPV		444065	897156	1318548	1694791	2030722	2330661	2598463	2837572	3051062	3241679	

EQUITY FINANCED BEFORE TAX

	YEAR 0	YEAR 1	YEAR 2	YEAR 3	YEAR 4	YEAR 5	YEAR 6	YEAR 7	YEAR 8	YEAR 9	YEAR 10	
CAPITAL COST	1041889		30000		40000		30000		40000		20000	1201889
CUM CASH FLOW	-1041889	-544536	-6179	585847	1137872	1729898	2291923	2883949	3435974	4028000	4600025	
CASH FLOW	-1041889	497353	538357	592026	552026	592026	562026	592026	552026	592026	572026	4600025
DISCOUNT RATE	12											
NET PRESENT VALUE		-597824	-148649	252743	603565	939497	1224236	1492038	1714992	1928483	2112660	

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DEBT FINANCED

	YEAR 0	YEAR 1	YEAR 2	YEAR 3	YEAR 4	YEAR 5	YEAR 6	YEAR 7	YEAR 8	YEAR 9	YEAR 10	
REVENUE		497353	568357	592026	592026	592026	592026	592026	592026	592026	592026	5801914
CAPITAL $A	1201889											
DEPRECIATION												
DECLINING BALANCE		1201889	949492	750099	592578	468137	369828	292164	230810	182340	144048	
- RATE	21.00	252397	199393	157521	124441	98309	77664	61354	48470	38291	30250	
ANNUAL DEPRECIATION		252397	199393	157521	124441	98309	77664	61354	48470	38291	30250	1088091
									ADD TO FINAL YEAR DEP.N			113798
CAPITAL TERM YRS	10											
PRINCIPLE REPAID $A		64858	71343	78478	86326	94958	104454	114899	126389	139028	152931	168224
PRINCIPLE OUTSTANDING	1201889	1137031	1065688	987210	900884	805926	701472	586573	460184	321155	168224	
INTEREST RATE %	10											
INTEREST EFFECT	1322078	1250734	1172257	1085931	990973	886519	771619	645230	506202	353271	185047	
INTEREST PAYMENT	120189	113703	106569	98721	90088	80593	70147	58657	46018	32116	16822	
TOTAL REPAYMENTS	185047	185047	185047	185047	185047	185047	185047	185047	185047	185047	185047	2035513
DEPRECIATION $A		252397	199393	157521	124441	98309	77664	61354	48470	38291	144048	
TAXABLE INCOME $A	-120189	131253	262395	335784	377496	413124	444214	472014	497537	521619	431155	
CARRY LOSS FORWARD $A		11064	262395	335784	377496	413124	444214	472014	497537	521619	431155	
								7TH YEAR				
TAX RATE %	33											
TAX PAID $A		3651	86590	110809	124574	136331	146591	155765	164187	172134	142281	
YEAR		1	2	3	4	5	6	7	8	9	10	
OPERATING PROFIT	-120189	127602	175805	224975	252922	276793	297624	316249	333350	349485	288974	
ADD BACK DEPRECIATION		252397	199393	157521	124441	98309	77664	61354	48470	38291	144048	
- CAPITAL PAYMENT	0	64858	71343	78478	86326	94958	104454	114899	126389	139028	152931	
- PREPRODUCTION COST	100000											
ADD BACK INTEREST	120189	113703	106569	98721	90088	80593	70147	58657	46018	32116	16822	
CASH FLOW $A MILLIONS	-100000	428844	410423	402739	381126	360736	340981	321362	301449	280863	296813	3425337
CUM CASH FLOW	-100000	328844	739267	1142006	1523133	1883869	2224850	2546212	2847660	3128524	3425337	
NPV	-100000	282896	610083	896745	1138958	1343649	1516401	1661768	1783519	1884801	1980367	
DISCOUNT RATE	12											

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EURALBA MINING LTD
MT LYELL SX/EW COPPER PROJECT
FEASIBILITY STUDY MAY, 1994

FINANCIAL ANALYSIS

PRODUCTION RATE SENSITIVITY

BASE CONDITIONS			ASSUMPTIONS			
FLOW COPPER TENOR	0.170	g/l	COPPER PRICE	0.85	US$/LB	2.68 A$/K
RECOVERY	80	%	EXCHANGE RATE	0.7	US$/A$	
CAPITAL COST A$	1041889		DAYS/YEAR	364		
PREPRODUCTION COST	100000		HRS/DAY	24		
BASE CASE ANNUAL PRODUCTION			INTEREST RATE	10	%	
FLOWRATE M3/HR	500		DISCOUNT RATE	12	%	
COPPER g/l	0.12		TERM	10	YEARS	
RECOVERY %	80		NPV ON REVENUE	5427441		
COPPER TONNES	419.3		NPV ON CFBIT	4298422	A$ REVENUE - CAPITAL	
OPERATING COST FACTOR	1		NPV	3444827	A$ AFTER DEBT SERVICE AND TAX	

FLOW COPPER TENOR	0.220	g/l	COPPER PRICE	0.85	US$/LB	2.68 A$/K
RECOVERY	80	%	EXCHANGE RATE	0.7	US$/A$	
CAPITAL COST A$	1041889		DAYS/YEAR	364		
PREPRODUCTION COST	100000		HRS/DAY	24		
BASE CASE ANNUAL PRODUCTION			INTEREST RATE	10	%	
FLOWRATE M3/HR	500		DISCOUNT RATE	12	%	
COPPER g/l	0.12		TERM	10	YEARS	
RECOVERY %	80		NPV ON REVENUE	7613204		
COPPER TONNES	419.3		NPV ON CFBIT	6484184	A$ REVENUE - CAPITAL	
OPERATING COST FACTOR	1		NPV	4909288	A$ AFTER DEBT SERVICE AND TAX	

FLOW COPPER TENOR	0.070	g/l	COPPER PRICE	0.85	US$/LB	2.68 A$/K
RECOVERY	80	%	EXCHANGE RATE	0.7	US$/A$	
CAPITAL COST A$	1041889		DAYS/YEAR	364		
PREPRODUCTION COST	100000		HRS/DAY	24		
BASE CASE ANNUAL PRODUCTION			INTEREST RATE	10	%	
FLOWRATE M3/HR	500		DISCOUNT RATE	12	%	
COPPER g/l	0.12		TERM	10	YEARS	
RECOVERY %	80		NPV ON REVENUE	1055916		
COPPER TONNES	419.3		NPV ON CFBIT	-73103	A$ REVENUE - CAPITAL	
OPERATING COST FACTOR	1		NPV	388450	A$ AFTER DEBT SERVICE AND TAX	

FLOW COPPER TENOR	0.120	g/l	COPPER PRICE	0.85	US$/LB	2.68 A$/K
RECOVERY	80	%	EXCHANGE RATE	0.7	US$/A$	
CAPITAL COST A$	1041889		DAYS/YEAR	364		
PREPRODUCTION COST	100000		HRS/DAY	24		
BASE CASE ANNUAL PRODUCTION			INTEREST RATE	10	%	
FLOWRATE M3/HR	500		DISCOUNT RATE	12	%	
COPPER g/l	0.12		TERM	10	YEARS	
RECOVERY %	80		NPV ON REVENUE	3241679		
COPPER TONNES	419.3	*	NPV ON CFBIT	2112660	A$ REVENUE - CAPITAL	
OPERATING COST FACTOR	1		NPV	1980367	A$ AFTER DEBT SERVICE AND TAX	

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EURALBA MINING LTD
MT LYELL SX/EW COPPER PROJECT
FEASIBILITY STUDY MAY, 1994

FINANCIAL ANALYSIS

COPPER PRICE SENSITIVITY +/-20%

BASE CONDITIONS			ASSUMPTIONS			
FLOW COPPER TENOR	0.120	g/l	COPPER PRICE	0.95	US$/LB	2.99 A$/KG
RECOVERY	80	%	EXCHANGE RATE	0.7	US$/A$	
CAPITAL COST A$	1041889		DAYS/YEAR	364		
PREPRODUCTION COST	100000		HRS/DAY	24		
BASE CASE ANNUAL PRODUCTION			INTEREST RATE	10	%	
FLOWRATE M3/HR	500		DISCOUNT RATE	12	%	
COPPER g/l	0.12		TERM	10	YEARS	
RECOVERY %	80		NPV ON REVENUE	3973249		
COPPER TONNES	419.3		NPV ON CFBIT	2844230	A$ REVENUE - CAPITAL	
OPERATING COST FACTOR	1		NPV	2470519	A$ AFTER DEBT SERVICE AND TAX	

BASE CONDITIONS			ASSUMPTIONS			
FLOW COPPER TENOR	0.120	g/l	COPPER PRICE	0.75	US$/LB	2.36 A$/KG
RECOVERY	80	%	EXCHANGE RATE	0.7	US$/A$	
CAPITAL COST A$	1041889		DAYS/YEAR	364		
PREPRODUCTION COST	100000		HRS/DAY	24		
BASE CASE ANNUAL PRODUCTION			INTEREST RATE	10	%	
FLOWRATE M3/HR	500		DISCOUNT RATE	12	%	
COPPER g/l	0.12		TERM	10	YEARS	
RECOVERY %	80		NPV ON REVENUE	2510108		
COPPER TONNES	419.3		NPV ON CFBIT	1381089	A$ REVENUE - CAPITAL	
OPERATING COST FACTOR	1		NPV	1486813	A$ AFTER DEBT SERVICE AND TAX	

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EURALBA MINING LTD
MT LYELL SX/EW COPPER PROJECT
FEASIBILITY STUDY MAY, 1994

FINANCIAL ANALYSIS

OPERATING COST SENSITIVITY +/-20%

BASE CONDITIONS			ASSUMPTIONS			
FLOW COPPER TENOR	0.120	g/l	COPPER PRICE	0.85	US$/LB	2.68 A$/KG
RECOVERY	80	%	EXCHANGE RATE	0.7	US$/A$	
CAPITAL COST A$	1041889		DAYS/YEAR	364		
PREPRODUCTION COST	100000		HRS/DAY	24		
BASE CASE ANNUAL PRODUCTION			INTEREST RATE	10	%	
FLOWRATE M3/HR	500		DISCOUNT RATE	12	%	
COPPER g/l	0.12		TERM	10	YEARS	
RECOVERY %	80		NPV ON REVENUE	2646345		
COPPER TONNES	419.3		NPV ON CFBIT	1517326	A$ REVENUE - CAPITAL	
OPERATING COST FACTOR	1.2		NPV	1578605	A$ AFTER DEBT SERVICE AND TAX	

BASE CONDITIONS			ASSUMPTIONS			
FLOW COPPER TENOR	0.120	g/l	COPPER PRICE	0.85	US$/LB	2.68 A$/KG
RECOVERY	80	%	EXCHANGE RATE	0.7	US$/A$	
CAPITAL COST A$	1041889		DAYS/YEAR	364		
PREPRODUCTION COST	100000		HRS/DAY	24		
BASE CASE ANNUAL PRODUCTION			INTEREST RATE	10	%	
FLOWRATE M3/HR	500		DISCOUNT RATE	12	%	
COPPER g/l	0.12		TERM	10	YEARS	
RECOVERY %	80		NPV ON REVENUE	3837013		
COPPER TONNES	419.3		NPV ON CFBIT	2707994	A$ REVENUE - CAPITAL	
OPERATING COST FACTOR	0.8		NPV	2379240	A$ AFTER DEBT SERVICE AND TAX	

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EURALBA MINING LTD
MT LYELL SX/EW COPPER PROJECT
FEASIBILITY STUDY MAY, 1994

FINANCIAL ANALYSIS

CAPITAL SENSITIVITY +/-20%

BASE CONDITIONS			ASSUMPTIONS			
FLOW COPPER TENOR	0.120	g/l	COPPER PRICE	0.85	US$/LB	2.68 A$/KG
RECOVERY	80	%	EXCHANGE RATE	0.7	US$/A$	
CAPITAL COST A$	1250267		DAYS/YEAR	364		
PREPRODUCTION COST	100000		HRS/DAY	24		
BASE CASE ANNUAL PRODUCTION			INTEREST RATE	10	%	
FLOWRATE M3/HR	500		DISCOUNT RATE	12	%	
COPPER g/l	0.12		TERM	10	YEARS	
RECOVERY %	80		NPV ON REVENUE	3241679		
COPPER TONNES	419.3		NPV ON CFBIT	1904282	A$ - REVENUE LESS CAPITA	
OPERATING COST FACTOR	1		NPV	1962180	A$ - AFTER DEBT SERVICE AND TAX	

BASE CONDITIONS			ASSUMPTIONS			
FLOW COPPER TENOR	0.120	g/l	COPPER PRICE	0.85	US$/LB	2.68 A$/KG
RECOVERY	80	%	EXCHANGE RATE	0.7	US$/A$	
CAPITAL COST A$	833511		DAYS/YEAR	364		
PREPRODUCTION COST	100000		HRS/DAY	24		
BASE CASE ANNUAL PRODUCTION			INTEREST RATE	10	%	
FLOWRATE M3/HR	500		DISCOUNT RATE	12	%	
COPPER g/l	0.12		TERM	10	YEARS	
RECOVERY %	80		NPV ON REVENUE	3241679		
COPPER TONNES	419.3		NPV ON CFBIT	2321037	A$ REVENUE - CAPITAL	
OPERATING COST FACTOR	1		NPV	1996241	A$ AFTER DEBT SERVICE AND TAX	

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EURALBA MINING LTD
MT LYELL SX/EW COPPER PROJECT
FEASIBILITY STUDY MAY, 1994

FINANCIAL ANALYSIS

DISCOUNT RATE SENSITIVITY

BASE CONDITIONS			ASSUMPTIONS			
FLOW COPPER TENOR	0.120	g/l	COPPER PRICE	0.85	US$/LB	2.68 A$/KG
RECOVERY	80	%	EXCHANGE RATE	0.7	US$/A$	
CAPITAL COST A$	1041889		DAYS/YEAR	364		
PREPRODUCTION COST	100000		HRS/DAY	24		
BASE CASE ANNUAL PRODUCTION			INTEREST RATE	10	%	
FLOWRATE M3/HR	500		DISCOUNT RATE	15	%	
COPPER g/l	0.12		TERM	10	YEARS	
RECOVERY %	80		NPV ON REVENUE	2871018		
COPPER TONNES	419.3		NPV ON CFBIT	1752585	A$ REVENUE - CAPITAL	
OPERATING COST FACTOR	1		NPV	1765292	A$ AFTER DEBT SERVICE AND TAX	

BASE CONDITIONS			ASSUMPTIONS			
FLOW COPPER TENOR	0.120	g/l	COPPER PRICE	0.85	US$/LB	2.68 A$/KG
RECOVERY	80	%	EXCHANGE RATE	0.7	US$/A$	
CAPITAL COST A$	1041889		DAYS/YEAR	364		
PREPRODUCTION COST	100000		HRS/DAY	24		
BASE CASE ANNUAL PRODUCTION			INTEREST RATE	10	%	
FLOWRATE M3/HR	500		DISCOUNT RATE	9	%	
COPPER g/l	0.12		TERM	10	YEARS	
RECOVERY %	80		NPV ON REVENUE	3692640		
COPPER TONNES	419.3		NPV ON CFBIT	2550753	A$ REVENUE - CAPITAL	
OPERATING COST FACTOR	1		NPV	2239415	A$ AFTER DEBT SERVICE AND TAX	

11.0 CONCLUSIONS AND RECOMMENDATIONS

The recovery of copper cathode of LME grade "A" quality from the effluent streams at Mt Lyell is technically feasible and commercially viable.

The process route and techniques have been pilot tested at the Mt Lyell site, and thus a high level of confidence in the process is justified.

The financial analysis is based upon cost estimates derived by well established methods and current pricing data. These cost estimates are considered to be conservative.

Several areas require more detailed definition. The plant site and the availability of power, water and access have not been determined. It is recommended that these issues be resolved rapidly, and then a final cost estimate for the entire project implementation can be undertaken. This final estimate should also take into account the potential for use of an existing building at the site, and the potential for aquisition of other elements of the existing facilities.

It is worthy of note that, to the author's knowledge, there is no other process route and technology commercially available which is capable of achieving the viable extraction of copper from the Mt Lyell effluent streams.

APPENDIX 1.

METEOROLOGICAL DATA

AVERAGE AND EXTREME MAXIMUM AND MINIMUM TEMPERATURES - SELECTED TASMANIAN STATIONS

(Temperatures measured in degrees Celcius)

STATION NAME & NUMBER		PERIOD	JAN	FEB	MAR	APR	MAY	JUNE	JULY	AUG	SEPT	OCT	NOV	DEC	ANNUAL
QUEENSTOWN (7xs)	MEAN MAX	1965-1986	21.1	22.0	19.7	16.6	14.5	12.3	11.6	12.5	13.6	15.9	17.6	19.3	16.4
97034	MEAN MIN	1965-1986	8.3	8.5	7.7	6.4	4.4	2.6	2.3	3.1	4.2	5.0	6.4	7.8	5.6
	EXT. MAX	1965-1987	37.3	36.3	35.6	29.5	25.0	19.5	19.5	21.0	26.8	27.8	33.3	35.3	37.3
			1981	1982	1966	1985	1985	1974	1975	1977	1987	1965/77	1966	1980	1/1981
	EXT. MIN	1965-1987	0.0	0.0	-1.1	-2.6	-6.0	-6.2	-6.7	-5.5	-3.9	-3.3	-1.5	-0.6	-6.7
			1965	1966	1966	1965	1974	1973	1978	1974	1983	1965/68	1974	1965/66	7/1978
ROSEBERY	MEAN MAX	1979-1986	21.1	22.1	20.1	17.0	14.4	11.3	10.8	12.6	13.4	16.3	19.1	20.5	16.6
97073	MEAN MIN	1979-1986	10.1	9.2	9.4	7.8	6.2	4.2	2.9	4.8	5.3	6.5	7.7	9.4	7.0
	EXT. MAX	1979-1987	38.0	36.8	31.3	28.0	24.5	17.7	15.6	19.1	22.5	26.2	33.0	35.6	38.0
			1981	1982	1983	1985	1985	1981	1981	1980	1985	1981	1987	1980	1/1981
	EXT. MIN	1979-1987	4.0	1.8	1.9	0.5	0.0	-3.0	-3.0	-2.0	-1.5	-1.0	-0.5	2.8	-3.0
			1986	1980	1987	1984	1981	1983/85	1982/83	1986	1983	1984	1980	1979	various
SAVAGE RIVER *	MEAN MAX	1966-1987	18.9	20.2	17.5	14.7	12.1	10.2	9.3	9.9	11.1	13.5	15.4	17.3	14.2
97047	MEAN MIN	1966-1987	9.2	10.2	8.5	7.3	5.8	4.2	3.5	3.6	4.2	5.3	6.8	8.0	6.4
	EXT. MAX	1966-1987	34.1	34.4	29.3	25.2	21.3	18.3	18.6	26.1	23.9	26.0	30.8	32.4	34.4
			1981	1975	1974	1987	1985	1968	1986	1971	1987	1977	1987	1980	2/1975
	EXT. MIN	1966-1987	0.8	2.3	-0.1	-2.9	-3.9	-3.3	-3.5	-5.0	-3.8	-1.1	-4.0	-1.2	-5.0
			1967	1982	1967	1967	1970	1982	1986	1978	1972	1966	1980	1968	8/1978
STRAHAN	MEAN MAX	1971-1989	20.7	21.4	19.5	17.0	14.8	12.4	12.1	12.8	14.1	16.0	17.8	19.1	16.5
97067	MEAN MIN	1971-1989	10.6	10.6	9.8	8.8	7.4	5.4	4.7	5.4	6.4	7.2	8.2	9.7	7.9
	EXT. MAX	1971-1989	36.2	37.2	34.2	29.3	23.1	19.4	19.1	20.2	26.9	31.2	33.5	36.2	37.2
			1973	1982	1989	1985	1988	1974	1975	1977/80	1987	1987	1977	1976	2/1982
	EXT. MIN	1971-1989	4.3	2.4	2.4	0.3	-0.5	-3.0	-2.0	-2.0	-1.2	0.2	1.5	2.6	-3.0
			1978	1980	1987	1988	1976	1983	1978/82	1974	1976	1975	1974	1975	6/1983

*Denotes Closed Station

2042210

294211

THE MOUNT LYELL MINING AND RAILWAY COMPANY LIMITED

RAINFALL RECORD

JANUARY, 1992

	QUEENSTOWN (SUBSTATION) (mm)	LAKE MARGARET DAM (mm)	LAKE MARGARET POWER STATION (mm)
THIS MONTH			
RAINFALL	172.8	204.0	209.8
AVERAGE	151.1	232.1	-
RECORD LOW	28.0 (1939)	53.0 (1939)	-
RECORD HIGH	433.0 (1923)	610.0 (1923)	-
WET DAYS	21	16	21
YEAR TO DATE			
RAINFALL	172.8	204.0	209.8
AVERAGE	151.1	232.1	-
WET DAYS	21	16	21

294212

THE MOUNT LYELL MINING AND RAILWAY COMPANY LIMITED

RAINFALL RECORD

FEBRUARY, 1992

	QUEENSTOWN (SUBSTATION) (mm)	LAKE MARGARET DAM (mm)	LAKE MARGARET POWER STATION (mm)
THIS MONTH			
RAINFALL	80.2	113.2	101.0
AVERAGE	120.7	185.1	-
RECORD LOW	15.0 (1912)	46.0 (1975)	-
RECORD HIGH	322.0 (1940)	426.0 (1940)	-
WET DAYS	12	11	12
YEAR TO DATE			
RAINFALL	253.0	317.2	310.8
AVERAGE	271.8	417.2	-
WET DAYS	33	27	33

294213

THE MOUNT LYELL MINING AND RAILWAY COMPANY LIMITED

RAINFALL RECORD

MARCH, 1992

	QUEENSTOWN (SUBSTATION) (mm)	LAKE MARGARET DAM (mm)	LAKE MARGARET POWER STATION (mm)
THIS MONTH			
RAINFALL	97.8	114.0	105.8
AVERAGE	163.2	237.7	-
RECORD LOW	49.0 (1910)	106.0 (1974)	-
RECORD HIGH	420.0 (1945)	576.0 (1945)	-
WET DAYS	16	16	16
YEAR TO DATE			
RAINFALL	350.8	431.2	416.6
AVERAGE	435.0	654.9	-
WET DAYS	49	43	49

294214

THE MOUNT LYELL MINING AND RAILWAY COMPANY LIMITED

RAINFALL RECORD

APRIL, 1992

	QUEENSTOWN (SUBSTATION) (mm)	LAKE MARGARET DAM (mm)	LAKE MARGARET POWER STATION (mm)
THIS MONTH			
RAINFALL	173.2	260.8	231.8
AVERAGE	223.6	313.4	-
RECORD LOW	17.0 (1923)	25.0 (1923)	-
RECORD HIGH	454.0 (1914)	649.0 (1914)	-
WET DAYS	22	20	23
YEAR TO DATE			
RAINFALL	524.0	692.0	646.4
AVERAGE	658.6	968.3	-
WET DAYS	71	63	72

294215

THE MOUNT LYELL MINING AND RAILWAY COMPANY LIMITED

RAINFALL RECORD

MAY, 1992

	QUEENSTOWN (SUBSTATION) (mm)	LAKE MARGARET DAM (mm)	LAKE MARGARET POWER STATION (mm)
THIS MONTH			
RAINFALL	194.8	259.0	252.8
AVERAGE	244.1	340.9	-
RECORD LOW	46.0 (1974)	87.0 (1974)	-
RECORD HIGH	557.0 (1948)	694.0 (1948)	-
WET DAYS	18	19	19
YEAR TO DATE			
RAINFALL	718.8	951.0	901.2
AVERAGE	902.7	1309.2	-
WET DAYS	89	82	91

294216

THE MOUNT LYELL MINING AND RAILWAY COMPANY LIMITED

RAINFALL RECORD

JUNE, 1992

	QUEENSTOWN (SUBSTATION) (mm)	LAKE MARGARET DAM (mm)	LAKE MARGARET POWER STATION (mm)
THIS MONTH			
RAINFALL	202.4	213.0	223.6
AVERAGE	234.5	319.2	-
RECORD LOW	55.7 (1978)	50.0 (1937)	-
RECORD HIGH	575.0 (1962)	781.0 (1952)	-
WET DAYS	15	15	15
YEAR TO DATE			
RAINFALL	921.2	1164.0	1124.8
AVERAGE	1137.2	1628.4	-
WET DAYS	104	97	106

294217

THE MOUNT LYELL MINING AND RAILWAY COMPANY LIMITED

RAINFALL RECORD

JULY, 1992

	QUEENSTOWN (SUBSTATION) (mm)	LAKE MARGARET DAM (mm)	LAKE MARGARET POWER STATION (mm)
THIS MONTH			
RAINFALL	403.0	470.0	454.2
AVERAGE	260.0	358.0	-
RECORD LOW	66.0 (1910)	107.0 (1932)	-
RECORD HIGH	525.0 (1970)	769.0 (1917)	-
WET DAYS	26	27	27
YEAR TO DATE			
RAINFALL	1324.2	1634.0	1579.0
AVERAGE	1397.2	1986.4	-
WET DAYS	130	124	133

294218

THE MOUNT LYELL MINING AND RAILWAY COMPANY LIMITED

RAINFALL RECORD

AUGUST, 1992

	QUEENSTOWN (SUBSTATION) (mm)	LAKE MARGARET DAM (mm)	LAKE MARGARET POWER STATION (mm)
THIS MONTH			
RAINFALL	336.0	379.2	380.4
AVERAGE	262.9	359.4	-
RECORD LOW	91.0 (1965)	123.0 (1961)	-
RECORD HIGH	480.0 (1936)	679.0 (1946)	-
WET DAYS	26	21	26
YEAR TO DATE			
RAINFALL	1660.2	2013.2	1959.4
AVERAGE	1660.1	2345.8	-
WET DAYS	156	145	159

294219

THE MOUNT LYELL MINING AND RAILWAY COMPANY LIMITED

RAINFALL RECORD

SEPTEMBER, 1992

	QUEENSTOWN (SUBSTATION) (mm)	LAKE MARGARET DAM (mm)	LAKE MARGARET POWER STATION (mm)
THIS MONTH			
RAINFALL	245.4	323.2	304.0
AVERAGE	247.7	352.1	-
RECORD LOW	85.0 (1951)	130.4 (1977)	-
RECORD HIGH	586.0 (1980)	829.0 (1980)	-
WET DAYS	27	27	27
YEAR TO DATE			
RAINFALL	1905.6	2336.4	2263.4
AVERAGE	1907.8	2697.9	-
WET DAYS	183	172	186

294220

THE MOUNT LYELL MINING AND RAILWAY COMPANY LIMITED

RAINFALL RECORD

OCTOBER, 1992

	QUEENSTOWN (SUBSTATION) (mm)	LAKE MARGARET DAM (mm)	LAKE MARGARET POWER STATION (mm)
THIS MONTH			
RAINFALL	202.0	260.4	221.6
AVERAGE	226.2	325.4	-
RECORD LOW	66.0 (1914)	111.0 (1914)	-
RECORD HIGH	568.0 (1988)	784.4 (1988)	-
WET DAYS	20	16	17
YEAR TO DATE			
RAINFALL	2107.6	2596.8	2485.0
AVERAGE	2134.0	3023.3	-
WET DAYS	203	188	203

294221

THE MOUNT LYELL MINING AND RAILWAY COMPANY LIMITED

RAINFALL RECORD

NOVEMBER, 1992

	QUEENSTOWN (SUBSTATION) (mm)	LAKE MARGARET DAM (mm)	LAKE MARGARET POWER STATION (mm)
THIS MONTH			
RAINFALL	210.4	296.8	264.2
AVERAGE	199.3	298.1	-
RECORD LOW	66.0 (1989)	99.6 (1989)	-
RECORD HIGH	475.0 (1915)	641.0 (1915)	-
WET DAYS	21	21	21
YEAR TO DATE			
RAINFALL	2318.0	2893.6	2749.2
AVERAGE	2333.3	3321.4	-
WET DAYS	224	209	224

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294222

THE MOUNT LYELL MINING AND RAILWAY COMPANY LIMITED

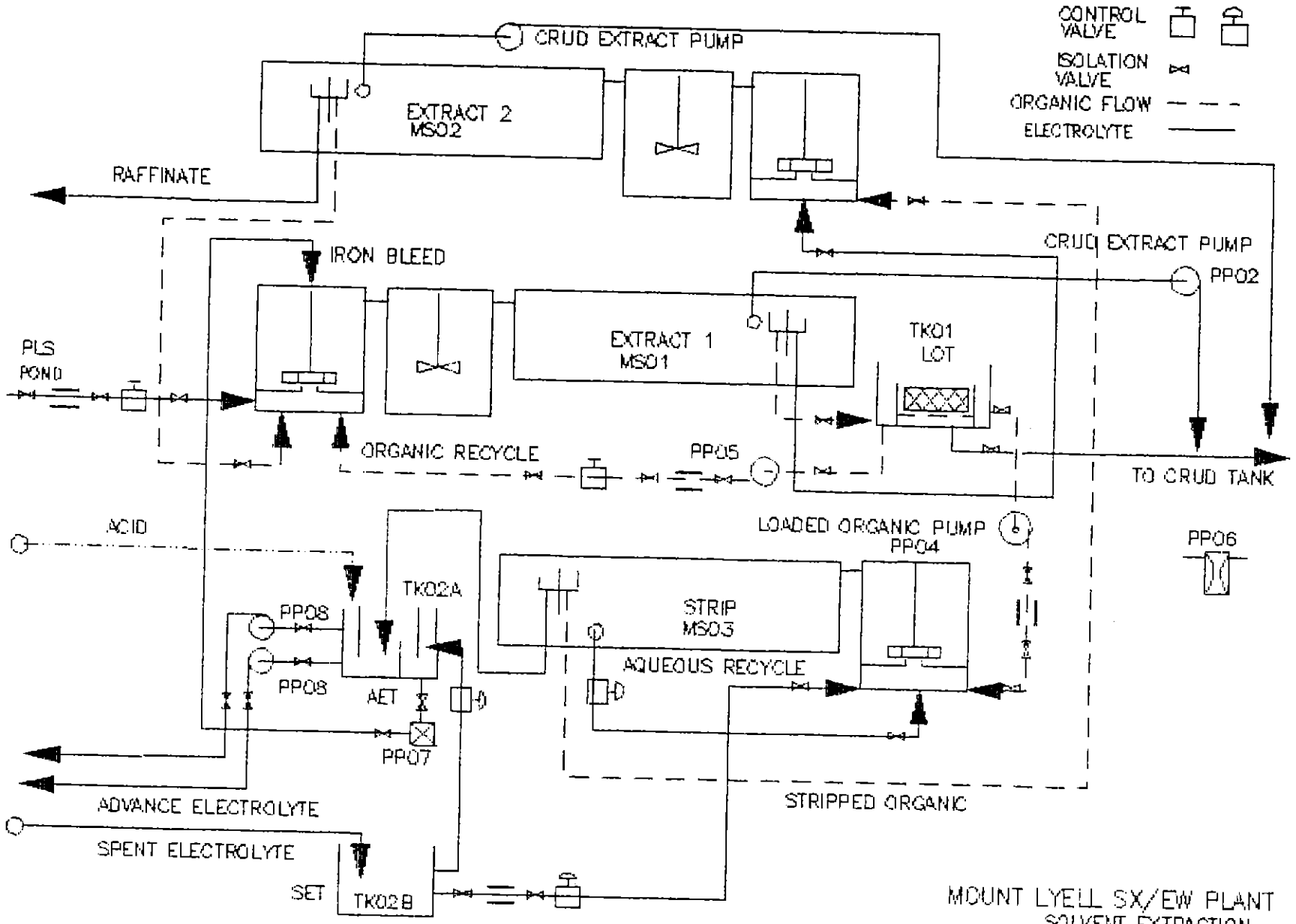
RAINFALL RECORD

DECEMBER, 1992

	QUEENSTOWN (SUBSTATION) (mm)	LAKE MARGARET DAM (mm)	LAKE MARGARET POWER STATION (mm)
THIS MONTH			
RAINFALL	113.6	164.8	159.0
AVERAGE	184.0	256.8	-
RECORD LOW	9.0 (1960)	62.0 (1960)	-
RECORD HIGH	433.0 (1976)	611.0 (1976)	-
WET DAYS	16	12	13
YEAR TO DATE			
RAINFALL	2431.6	3058.4	2908.2
AVERAGE	2517.3	3578.2	-
WET DAYS	240	221	237

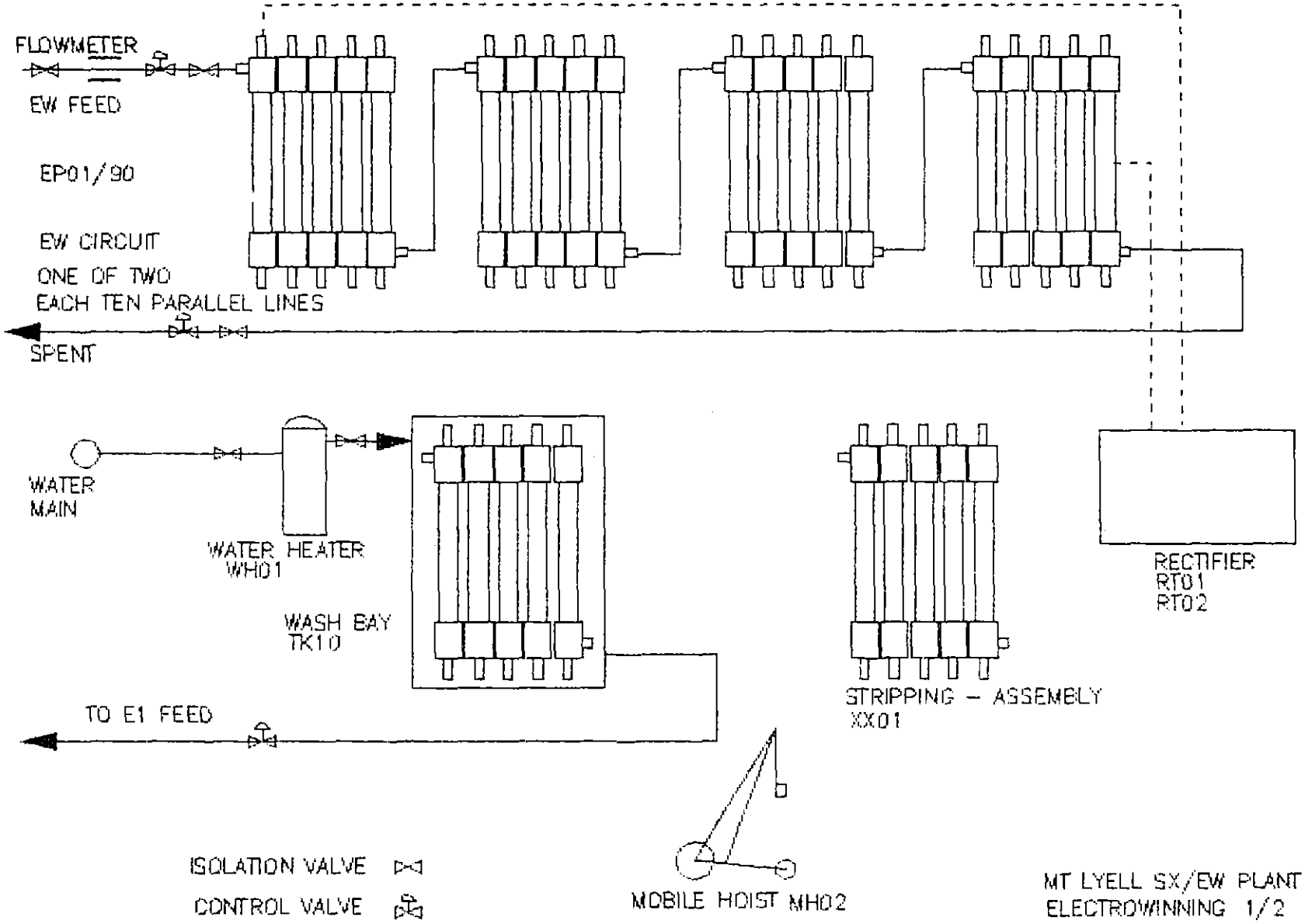
APPENDIX 2.

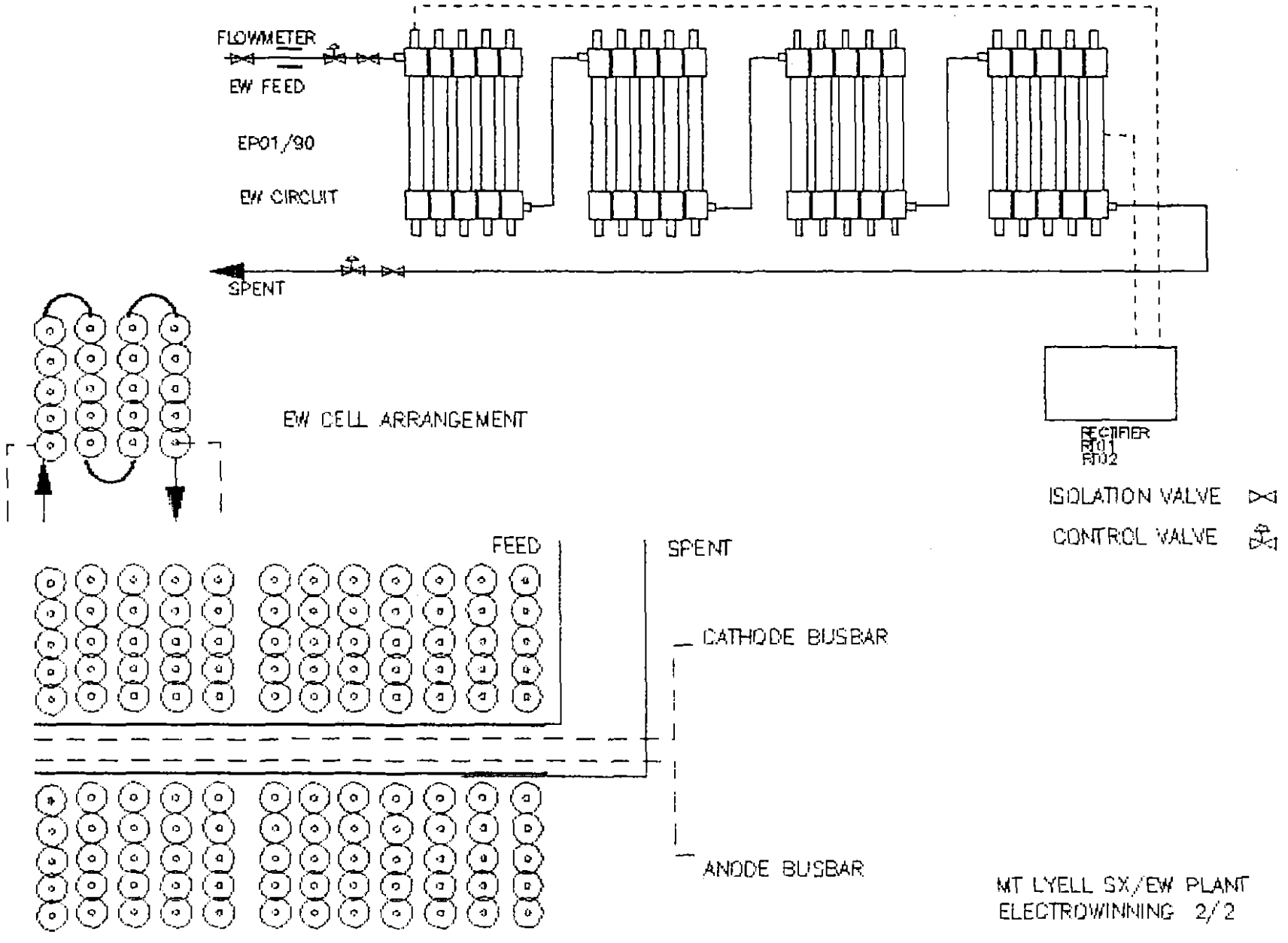
FLOWSHEETS

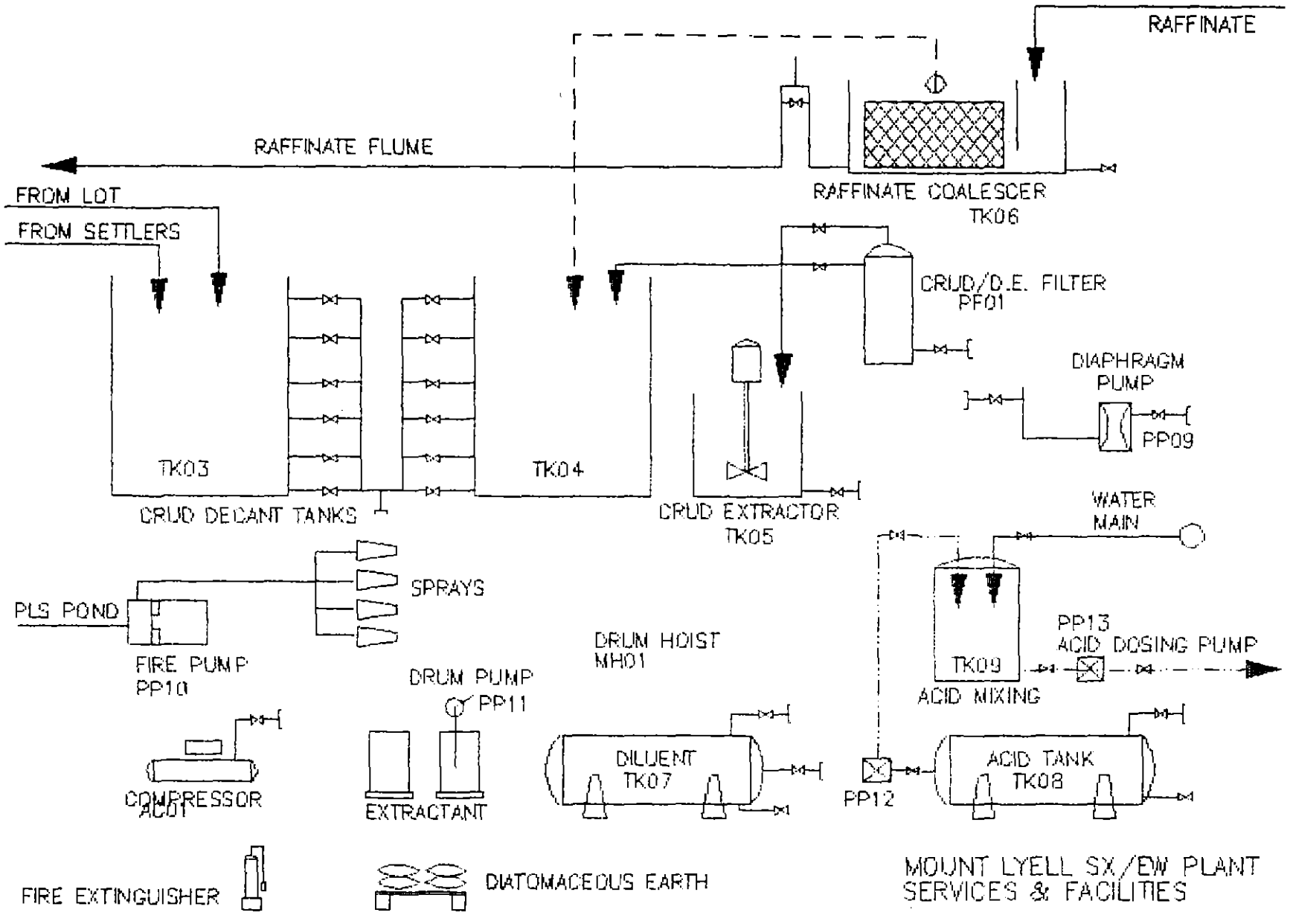


MOUNT LYELL SX/EW PLANT
SOLVENT EXTRACTION

2042224


FLOWMETER
EW FEED
EP01/90
EW CIRCUIT
ONE OF TWO
EACH TEN PARALLEL LINES
SPENT
WATER
MAIN
WATER HEATER
WH01
WASH BAY
TK10
TO E1 FEED
STRIPPING – ASSEMBLY
XX01
RECTIFIER
RT01
RT02
ISOLATION VALVE
CONTROL VALVE
MOBILE HOIST MH02
MT LYELL SX/EW PLANT
ELECTROWINNING 1/2


FLOWMETER
EW FEED
EP01/90
EW CIRCUIT
SPENT
EW CELL ARRANGEMENT
RECTIFIER
R01
R02
ISOLATION VALVE
CONTROL VALVE
FEED
SPENT
CATHODE BUSBAR
ANODE BUSBAR
MT LYELL SX/EW PLANT
ELECTROWINNING 2/2


RAFFINATE
RAFFINATE FLUME
RAFFINATE COALESCER
TK06
FROM LOT
FROM SETTLERS
CRUD/D.E. FILTER
PF01
DIAPHRAGM
PUMP
PP09
TK03
TK04
CRUD DECANT TANKS
CRUD EXTRACTOR
TK05
WATER
MAIN
SPRAYS
PLS POND
FIRE PUMP
PP10
DRUM HOIST
MH01
PP13
ACID DOSING PUMP
TK09
ACID MIXING
DRUM PUMP
PP11
DILUENT
TK07
ACID TANK
TK08
PP12
COMPRESSOR
AC01
EXTRACTANT
FIRE EXTINGUISHER
DIATOMACEOUS EARTH
MOUNT LYELL SX/EW PLANT
SERVICES & FACILITIES

APPENDIX 3.

DESIGN CRITERIA & PROCESS CALCULATIONS

294229

FILE: MLPILOT
MT. LYELL SX/EW COPPER PROJECT
EURALBA MINING LTD

DESIGN CRITERIA AND PROCESS DESIGN CALCULATIONS
SCALEUP FROM DEMONSTRATION PLANT PERFORMANCE

WASTE DUMP EFFLUENT DAMS

294230

SHAFT DUMP

CAPACITY	12 HRS	
FLOW AV.	35 M3/HR	

HAULAGE DUMP

CAPACITY	12 HRS	
FLOW AV.	35 M3/HR	

PLANT FEED

COPPER PRODUCTION	1224	KG/DAY	
FEED COPPER RECOVERED	0.102	G/L	
FEED FLOW	12000	M3/DAY	
	500	M3/HR	
DESIGN PLS FLOW	500	M3/HR	
	8333	L/MIN	0.579655
	139	L/SEC	

PLS FEED LINE

	450 OD		500 OD
LINE DIAM.	423	mm	470
VELOCITY	0.99	M/SEC	0.80
MAX HEAD	20	M	20
POWER	42.79	KW	42.79

PLS POND

FLOW IN	500	M3/HR
CAPACITY	12	HRS
	6000	M3
DEPTH	3	M
SLOPE OFF VERTICAL	30	deg
CUTBACK	5.2	M
TOP AREA	2500	M2
BOTTOM AREA	1569	M2
VOLUME	6049	M3
WIDTH	50	M
LENGTH	50	M
WALL AREA	18.6	M2
FILL VOLUME	3911	M3
COMPACT DENSITY	1.75	T/M3
TONNES FILL	6844	T
SURFACE AREA	3388	M2
DESIGN AREA FOR LINER	4000	M2

SOLVENT EXTRACTION

SX EXTRACTION	85	%		
PLS TENOR	0.12	G/L		
SALICYLALDOXIME SG	0.94	KG/L		
VOL %	4	%		
DILUENT SG	0.8	KG/L		
EVAP	0.1	MM/DAY		
ORGANIC LOADING				
Cu/Fe SELECTIVITY	40	CHECK		
ORGANIC ENTRAIN RAFF	15	PPM		
DESIGN Cu DELTA	1	G/L		
EXTRACT CONTINUITY E1		ORGANIC		
EXTRACT CONTINUITY E2		ORGANIC		
STRIP CONTINUITY		AQUEOUS		
ORG/AQU. RATIO STRIP	1			
ADVANCE ELECT TENOR	25.70	G/L		
PLS + BLEED FLOW	500	M3/HR		
ORGANIC FLOWRATE EXT.	500	M3/HR		
Cu ADVANCE FROM PLS	0.12	G/L		
COPPER IN FEED	62	G/HR		
	EXTRACT		STRIP	
TOTAL FLOW TO M/S	1000	M3/HR	152	M3/HR
EXTRACT STAGES	2			
STRIP STAGES			1	
MIXERS/STAGE EXT.	2			
MIXERS/STAGE STRIP			1	
RESIDENCE TIME EACH	2	MINS	1.5	MINS
SETTLER RATING	3.2	M3/M2HR	3	M3/M2HR
ORGANIC SPACE VEL. DESIGN	0.05	M/SEC MAX	0.05	M/SEC
	5	cm/sec		cm/sec
ORGANIC DEPTH	200	mm MAX	200	mm
AQUEOUS DEPTH	400	mm	400	mm
FREEBOARD	150	mm	50	mm
SETTLER WIDTH	12000	mm	6000	mm
LENGTH	26047	mm	8464	mm
DESIGN LENGTH	26000	mm	12000	mm

Note: each extract M/S in four units of 6x14m

ORGANIC SPACE VEL ACTUAL			
EXTRACT	0.006	cm/sec	0.002 cm/sec

ELECTROWINNING

COPPER RECOVERY	50998	G/HR		50998	
MAX PROD RATE	1223.96	KG/DAY	DESIGN	1224	
CELL DIAM	0.101	M		0.101	
CELL LENGTH	1.000	M		1.000	
CATHODE AREA	0.319	M2		0.319	
CURRENT DENSITY	440	A/M2		440	
AMPS/CELL	140	A		140	
CURRENT EFFICIENCY	75	%		80	********
THEORETICAL CAPACITY	1.185	G/AHR		1.185	
NO OF CELLS	409.29			383.71	
FLOWRATE	3600	L/HR/CELL			
EW FEED FLOW	72000	L/HR			
NO OF LINES	20.0				
CELLS/LINE CALC	20.5				
CELLS/LINE DESIGN	20				
TOTAL CELLS	409.3				
AV WT DEPOSIT	44.1	KG	half gap		
NO CELLS HARVESTED/DAY	27.7		ie 6 packs		

NO OF PUMPS	2		ANODE DIAM	0.025	M
LINES/PUMP	10		CELL X-SECT	0.008	M2
EW FEED FLOW/PUMP	10	L/SEC	CELL VOLUME	0.008	M3
	600	L/MIN	RT	7.58	SEC
	36000	L/HR		107.44	G/PASS
PRESSURE DROP/LINE	32	kPa			
PRESSURE HEAD/LINE	2.74	M			
TOTAL HEAD	5	M			
SG	1.19	KG/L			
POWER	0.8	kW			

VOLTS/CELL	2.8	V
RECTIFIER VOLTS	57.3	V
DESIGN	60	V
AMPS/CELL	140	A
NO OF RECTIFIERS	2	
RECTIFIER AMPS	1402	A
DESIGN	1800	A
POWER FACTOR	0.85	
RECTIFIER EFFICIENCY	0.96	

ATOMIC WEIGHTS		ACID R
Fe	55.847	1.756182
Cu	63.54	1.543555
SO4	96.0616	
Fe	55.847	
H	1.00797	
O	15.9994	

CRUD SYSTEM & PROCESS TANKAGE

RAFFINATE COALESCER				500 OD	450
RESIDENCE TIME	5 MIN		LINE D	470 mm	4
FLOW	500 M3/HR		NO	1	
VOLUME	41.7 M3		VEL	0.80 M/SEC	0.
DEPTH	1.2 M		GO TO LAUNDER?		
AREA	34.72 M2				
WIDTH	4.0 M	65.16 m2 concrete			
LENGTH	8.7 M				
LOADED ORGANIC TANK					
RESIDENCE TIME	8 MIN		LINE D	100 mm	
FLOW	76.2 M3/HR		NO	2	
VOLUME	10.2 M3	5000	VEL	1.35 M/SEC	
NO OF TANKS	2				
DEPTH	2.5 M	2.61			
DIAMETER	1.6 M	1.86			
AREA	2.0 M2				
ADVANCE & SPENT ELECTROLYTE TANKS					
RESIDENCE TIME	3 MIN	3.9	LINE D	150 mm	
FLOW	76.2 M3/HR		NO	1	
VOLUME	3.8 M3	5000	VEL	1.20 M/SEC	
DEPTH	1.6 M	2610			
DIAMETER	1.7 M	1.86			
AREA	2.4 M2				
CRUD DECANT TANKS					
RESIDENCE TIME	720 MINS	675	LINE D	50 mm	
NO OF TANKS	2	5	NO	1	
FLOW	4.0 M3/HR		VEL	0.57 M/SEC	
VOLUME EACH	24.0 M3	9000			
DEPTH	4.5 M	2.075			
DIAMETER	2.6 M	2.7			
AREA	5.3 M2				
CRUD SEPARATOR/MIXER					
VOLUME	1.0 M3	1.05			
DEPTH	1.2 M	1.38			
DIAMETER	1.0 M	1.115			
AREA	0.8 M2				

294234

REAGENT STORAGE

ACID STORAGE TANK

DAILY CONSUMPTION	278.5	LITRES	
STORAGE CAPACITY	20	DAYS	
TANK VOLUME	5.6	M3	9000
DEPTH	2	M	2.075
DIAMETER	1.9	M	2.7
AREA	2.8	M2	

NOTE: EXTEND TO A SINGLE TANKER LOAD + 25%

BUND DEPTH	0.17	M
AREA	36	M2

ACID MIXING TANK

DAILY ADDITION + 15%	2.97	M3 WATER + ACID
DEPTH	1.5	M
DIAMETER	1.6	M
AREA	2.0	M2

EXTRACTANT IN DRUMS

DILUENT TANK

DAILY ADDITION	174	LITRES/DAY
STORAGE CAPACITY	30	DAYS
TANK VOLUME	5.2	M3

294235

PROCESS DESIGN CALCULATIONS

FEED - target average	120	ppm Cu	
for design	2500	ppm Fe	
IRON IN PLS	2.5	g/l	
IRON BLEED	6	g/l	
FEED FLOW	500	M3/HR	
Cu TENOR	0.12	g/l	
Cu RECOVERY	85	%	
AVAILABILITY	0.96		
Cu FLOW IN PLS	51.000	KG/HR	1224 kg/day
RAFF. TENOR	0.018	g/l	
Cu/Fe REJECTION	40		CHECK
AQUEOUS ENTRAINMENT			CONSERVATIVE
EXTRACT ORGANIC CTS	100		DESIGN ESTIMATE
STRIP AQ CTS	300		DESIGN ESTIMATE
SPENT Cu TENOR	25	g/l	
SPENT ACID TENOR	180	g/l	
PLS pH	2.4		
PLS ACID TENOR	0.39	g/l	
IRON BLEED FLOW	0.093	M3/HR	ITERATE ****
Cu TRANSFER BY SX	53.33	KG/HR	
Fe TRANSFER BY SX	1.333	KG/HR	
Fe TRANSFER BY ENTRAIN.			
EXTRACTION TO STRIP	0.125	KG/HR	
STRIP TO EXTRACTION	0.900	KG/HR	
BALANCE OF Fe TRANSFER	0.000		TO ZERO ****
TOTAL Fe TRANSFER	0.5580	KG/HR	
FOR PLS + IRON BLEED			
COPPER	0.12	g/l	
ACID	0.42	g/l	
IRON	2.50	g/l	
TOTAL ACID LOSS	21.309	KG/HR	
Cu TRANSFERRED	53.323	KG/HR	
DELTA TENOR	0.70		FROM ASSAY
SPENT FLOW TO STRIP	76.176	M3/HR	
SPENT EX T'HOUSE	76.269	M3/HR	not relevant
ORGANIC RECYCLE EXTRACT	423.917	M3/HR	
ORGANIC ADVANCE TO STRIP	76.176	M3/HR	
SYSTEM O/A	0.15		STRIP ORGANIC/PLS
AQUEOUS RECYCLE ON STRIP	0.000	M3/HR	
STRIP O/A RATIO	1.00		
LOADED ORGANIC TENOR	1.5		
DELTA organic	0.7		
Cu WON (TRANSFER-BLEED)	50.998	KG/HR	1224.0 KG/DAY
ACID GENERATED	78.719	KG/HR	
WATER LOST BY REACTION	14.459	KG/HR	
OXYGEN GENERATED	12.841	KG/HR	

424

EW MASS BALANCE

	FLOW M3/HR	g/l Cu	KG/HR Cu	g/l acid	KG/HR acid	g/l Fe	KG/HR Fe
ADVANCE EX STRIP	76.2	25.70	1958	179	13633	6.01	458
ADDITION	0.107			198	21.309		
ADVANCE ELECTROLYTE	76.3	25.66	1958	179	13654	6.00	458
EW FEED	72.0	25.71	1851	179.0	12886	6.00	432
NO EW LINES	20						
Cu WON/ACID MADE			50.998		78.719		0.000
RECIRCULATED SPENT	-4.3	25.00	-107	180	-768	6	*******
SPENT EX EW	72.0	25.00	1800	180	12957	6	432
BLEED	0.093	25.00	2.325	180	16.740	6	0.558
SPENT TO STRIP	76.2	25.00	1904	180	13712	6	457

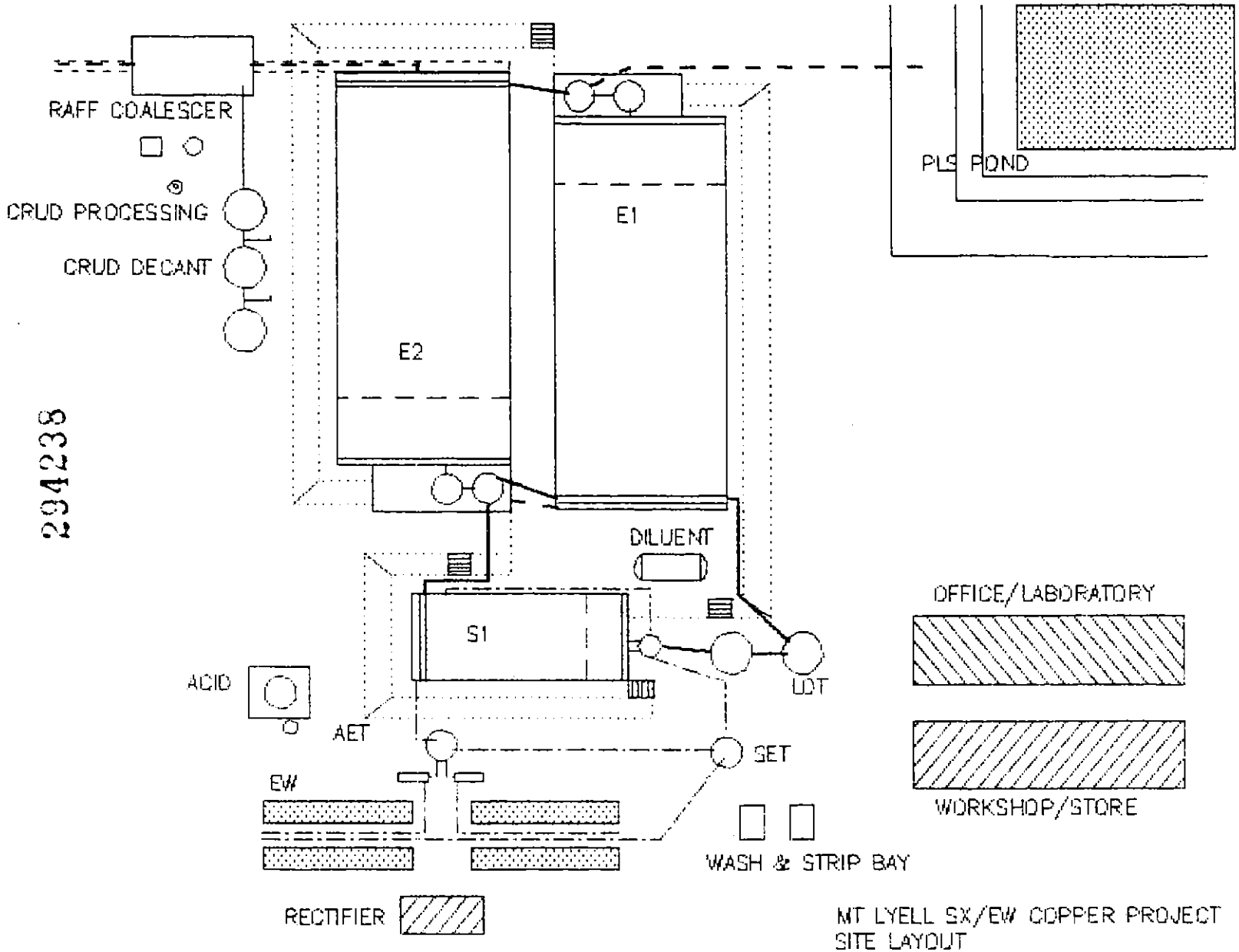
WHERE RECIRC.D SPENT IS NEGATIVE - THIS EQUATES TO THE STRIP AQUEOUS RECYCLE FLOW

MASS BALANCE OVERALL

	FLOW M3/HR	Cu g/l	Cu kg/hr	acid g/l	acid kg/hr	pH	factors
PLS	500	0.120	60.0	0.28	138.2	2.55	0.85 recovery
iron bleed	0.093	25.0	2.3	180	16.7		
SX feed	500	0.125	62.3	0.03	154.9	3.47	
raffinate	500	0.018	9.0	0.47	233.7	2.32	1.54
loaded organic	500	1.50	750.0				1 org/aq
ext. org. recycle	424	1.50	635.7				
organic advance	76	1.50	114.3				
stripped organic	76	0.800	60.9				
delta organic		0.700	53.3				
advance electrolyte	76	25.7	1957.7	179	13633		
copper transferred			53.3				
copper production			51.0			98.076	
spent electrolyte	76	25.0	1904.4	180	13712		delta

APPENDIX 4.

DRAWINGS



MT LYELL SX/EW COPPER PROJECT
SITE LAYOUT

APPENDIX 5.

EQUIPMENT LIST

MOUNT LYELL COPPER PROJECT

FILE: MLFEQUIP

TOTAL INSTALLED POWER - MT LYELL SX/EW PROCESS PLANT - kW 351.39

ESTIMATED POWER DRAW 246 kW

AREA 1. - SOLVENT EXTRACTION

EQUIP NO.	DESCRIPTION	NO.	INSTALLED POWER kW EACH	DESCRIPTION	TOTAL INSTALLED kW
PP01	PLS PRIME PUMP	1	0	OPTION DEP. ON POND ELEVATION	0
					0
MS01	EXTRACT M/S 1	1	0	CONCRETE FREFAB + MDPE LINER	0
					0
AG01	M/S 1 PUMP MIXER	1	11	TOP SHROUDED PUMP/MIXER - SS316L	11
				1.5 m D, 0 - 58 RPM type SX-6	0
AG02	M/S 1 AGITATOR	1	0.55	HYDROFOIL TURBINE MIXER	0.55
				1.1 m D, 0 - 53 RPM type HA-700	0
MS02	EXTRACT M/S 2	1	0	CONCRETE FREFAB + MDPE LINER	0
					0
AG03	M/S 2 PUMP MIXER	1	11	TOP SHROUDED PUMP/MIXER - SS316L	11
				1.5 m D, 0 - 58 RPM type SX-6	0
AG04	M/S 2 AGITATOR	1	0.55	HYDROFOIL TURBINE MIXER	0.55
				1.1 m D, 0 - 53 RPM type HA-700	0
PP02/03	CRUD REMOVAL PUMPS	2	0	WILDEN M2	0
					0
					0
					0
TK01	LOADED ORG. TANK	2	0	HDPE ROTOMOULD + FUSED FITTINGS	0
				5000 LITRE	0
PP04	LOADED ORG. PUMP	1	0	NOT REQUIRED	0
					0
PP05	ORG. RECYCLE PUMP	1	22	MODEL 200-315	22
					0
PP06	CRUD/AQ.LOT PUMP	1	0	WILDEN M1	0
					0
MS03	STRIP M/S	1	0	CONCRETE FREFAB + MDPE LINER	0
					0
AG05	STRIP M/S PUMP MIX	1	4	TOP SHROUDED PUMP/MIXER - 904L SS	4
				1.0 m D, 0 - 74 RPM type SX-6	0
TK02A	ADV. ELEC. TANK	1	0	HDPE ROTOMOULD + FUSED FITTINGS	0
				5000 LITRE	0
TK02B	SPENT ELECTROLYTE	1	0	HDPE ROTOMOULD + FUSED FITTINGS	0
				5000 LITRE	0
PP07	IRON BLEED PUMP	1	0	WILDEN M2	0
					0
PP08ab	ADV. ELEC. PUMP	2	2.2	IWAKI MDF L505 CFVW	4.4
					0
					0
TOTAL POWER - SOLVENT EXTRACTION EQUIPMENT					53.5

AREA 2. - ELECTROWINNING

EQUIP NO.	DESCRIPTION	NO.	INSTALLED POWER kW	DESCRIPTION	TOTAL INSTALLED kW
EP01/90	EW PACK (5 CELLS)	90	0	PROPRIETARY PACKAGE - PACKS OF 5	0
					0
FM01	FLOWMETERS	4	0.96	ULTRASONIC	3.84
					0
WH01	WATER HEATER	2	15	ELECTRIC RHEEM MODEL 603/315	30
				- 1000 L/HR @ 82 DEG C	0
TK10	WASH BAY ASSEMBLY	1	0	PROPRIETARY ASSEMBLY	0
					0
XX01	CELL STRIP JIG	1	2.4	PROPRIETARY ASSEMBLY	2.4
					0
RT01/02	RECTIFIER	2	100	CUSTOM BUILT	200
					0
MH02	MOBILE HOIST	0	0	ABBEY CR40P 2 TONNE	0
				utilise MH01	0
					0
					0
					0
					0
					0
					0
					0
					0
					0
					0
TOTAL POWER - ELECTROWINNING					236.24

AREA 3. - SERVICES AND FACILITIES

EQUIP NO.	DESCRIPTION	NO.	INSTALLED POWER kW	DESCRIPTION	TOTAL INSTALLED kW
					0
					0
TK03/04	CRUD DECANT TANKS	2	0	9000 L ROTOMOULD/FUSED FITTINGS	0
					0
TK05	CRUD EXT. TANK	1	0	1100 L ROTOMOULD/FUSED FITTINGS	0
					0
AG07	CRUD AGITATOR	1	0.55	TURBINE MIXER - 0.37D, 0 - 318 RPM type HA-750	0.55
PF01	CRUD/DE FILTER	1	0	SAND FILTER	0
					0
PP09	GP DIAPHRAGM PUMP	2	0	WILDEN M2	0
					0
TK06	RAFF. COALESCER	1	0	CONCRETE FREFAB + MDPE LINER	0
					0
AC01	AIR COMPRESSOR	1	30	TBA	30
					0
PP10	FIRE PUMP/SPRAY	1	7.5	MODEL 200-260	7.5
					0
PP11	EXTRACTANT PUMP	2	0	WILDEN M.025	0
					0
TK07	DILUENT TANK	1	0	PROVIDED BY DILUENT SUPPLIER	0
					0
TK08	ACID STORAGE TANK	1	0	5000 L ROTOMOULD/FUSED FITTINGS	0
					0
TK09	ACID MIXING TANK	1	0	316L SS - CUSTOM	0
					0
PP12	STRONG ACID PUMP	1	2.2	ALDOS KM 253-100/2 PVC	2.2
					0
PP13	ACID DOSING PUMP	1	2.2	ALDOS KM 255-403 PVC	2.2
					0
MH01	MOBILE DRUM HOIST	1	0	ABBEY CR40P 2 TONNE	0
					0
	OFFICE/LAB W'SHOP STORE	3	4.8	12x3m, ablutions, transportable	14.4
					0
	OFFICE/LAB. EQUIP.	1	4.8	furniture, AAS, glassware, misc. computer, FAX/phone, heating	4.8
					0
					0
TOTAL POWER - SERVICES AND FACILITIES					61.65

APPENDIX 6.

COPPER CATHODE ANALYSES

CRL QUALITY

COPPER REFINERIES PTY. LTD.

A.C.N. 009 676 975
A MEMBER OF THE M.I.M. GROUP OF COMPANIES

294244

HEAD OFFICE: HUNTER STREET, STUART, TOWNSVILLE, QUEENSLAND
P.O. BOX 5484 M.C.
TOWNSVILLE QLD. 4810
AUSTRALIA

TELEPHONE: (077) 81 8211
TELEGRAMS: COPPERFINE
TELEX : AA 47088 COPTOWN
FACSIMILE : (077) 78 2308 (General)
(077) 78 2755 (Supply)

7 February 1994

ASSAY REPORT

EURALBA COPPER

REF: NONAS 803

SAMPLE MTL2

	RESULTS	
SILVER	0.4	ppm
ARSENIC	<0.1	ppm
BISMUTH	<0.1	ppm
CADMIUM	<0.1	ppm
COBALT	<0.1	ppm
CHROMIUM	<0.1	ppm
IRON	6.5	ppm
MANGANESE	<0.1	ppm
NICKEL	0.3	ppm
PHOSPHOROUS	<0.1	ppm
LEAD	<0.1	ppm
SULPHUR	11.2	ppm
ANTIMONY	<0.1	ppm
SELENIUM	0.3	ppm
TIN	<0.1	ppm
TELLURIUM	<0.1	ppm
ZINC	0.2	ppm

Method of Analysis - CMS01

Sample was taken as supplied, melted in a graphite crucible under a nitrogen atmosphere, cast as an ingot and drawn to wire prior to analysis.

N. Gerrard
CHIEF CHEMIST

Ag Above = 189 ppm

CRL QUALITY

COPPER REFINERIES PTY. LTD.

A.C.N. 009 676 975
A MEMBER OF THE M.I.M. GROUP OF COMPANIES

HEAD OFFICE: HUNTER STREET, STUART, TOWNSVILLE, QUEENSLAND
P.O. BOX 5484 M.C.
TOWNSVILLE QLD. 4810
AUSTRALIA

TELEPHONE: (077) 81 9211
TELEGRAMS: COPPERFINE
TELEX : AA 47068 COPTOWN
FACSIMILE : (077) 78 2308 (General)
(077) 78 2755 (Supply)

7 February 1994

ASSAY REPORT

EURALBA COPPER

REF: NONAS 093

SAMPLE MTL1

RESULTS

SILVER	<0.1	ppm
ARSENIC	<0.1	ppm
BISMUTH	<0.1	ppm
CADMIUM	<0.1	ppm
COBALT	<0.1	ppm
CHROMIUM	<0.1	ppm
IRON	4.8	ppm
MANGANESE	<0.1	ppm
NICKEL	0.3	ppm
PHOSPHOROUS	<0.1	ppm
LEAD	0.1	ppm
SULPHUR	3.7	ppm
ANTIMONY	<0.1	ppm
SELENIUM	0.2	ppm
TIN	<0.1	ppm
TELLURIUM	<0.1	ppm
ZINC	0.2	ppm

Method of Analysis - CMS01

Sample was taken as supplied, melted in a graphite crucible under a nitrogen atmosphere, cast as an ingot and drawn to wire prior to analysis.

Total Above = 93 ppm

N. Garrard
CHIEF CHEMIST

EURALBA MINING LIMITED

Mt Lyell Project

Process description and design criteria

Project Costing

97-4023

294247

Uralba
MINING LIMITED
A.C.N. 000 751 093

GOLD COAST OFFICE
23 RIVERSDALE ROAD
OXENFORD, QLD., 4210

TEL: 075-731800
FAX: 075-736097

MOUNT LYELL PROJECT

1. Process description and design criteria
2. Project Costing

GPO BOX 202, SYDNEY NSW 2001
TELEPHONE: (02) 247 9855 FACSIMILE: (02) 251 5961
REGISTERED OFFICE: LEVEL 2, 28 THE ESPLANADE, PERTH W.A 6000

PROCESS DESCRIPTION

Waste dump drainage (PLS) is pumped or gravity fed from the drainage channel and fed to the extraction mixer settler. Flowrate is controlled by throttling valve or bypass line from a flowmeter reading (rotameter). The PLS is contacted with organic phase from the strip mixer settler in a single mixer box of a volume yielding a residence time of 2 minutes. The impellor design enables organic continuous operation by organic flooding at startup, and is a pump mixer turbine. The organic aqueous mixture is in a 1:1 ratio, and overflows the mixer into a lateral distributor at the head of the settler. Flow down the settler is moderated by a picket fence, and at the end of the settler the dispersion band between the two disengaged phases rides against the wall of the organic weir. Organic phase overflows the organic weir and into the loaded organic tank (LOT). The aqueous phase passes under the organic weir and into the aqueous weirbox. The aqueous phase flows over the adjustable weir and out of the settler into a baffled settling pond prior to discharge into the drainage channel as raffinate. The baffles trap any entrained organic, and the settler also acts as emergency containment.

Loaded organic is pumped via a throttle flow control valve into the strip mixer. The aqueous recycle and spent electrolyte is contacted with the loaded organic, and the mixture proceeds in similar fashion to the extraction settler. The aqueous recycle, taken from near the end of the settler, constitutes the bulk of the aqueous flow, and the small advance electrolyte stream flows over the aqueous weir into the advanced electrolyte tank (AET). This circuit is operated in aqueous continuous mode. The stripped organic flows to the extraction mixer box.

The advance electrolyte is pumped into the electrowinning circuit and back to the AET. A small bleed flow equal to the flowrate of the advance electrolyte from the strip settler is tapped from the spent electrolyte into the strip mixer, and depending on the iron levels in the electrolyte circuit, a bleed may also be taken from the spent electrolyte stream to the extraction mixer.

Acid and water are periodically added to the advance electrolyte to compensate for the iron bleed, evaporation and the breakdown of water in the electrowinning cells. Organic phase, a mixture of extractant and high flashpoint kerosene, is also periodically added to the extraction mixer to compensate for entrainment losses into raffinate, evaporation of kerosene and degradation of extractant.

A crud phase may form in the dispersion bands, depending on silica and suspended solids levels in PLS. This will need to be removed by suction and would also constitute a loss of organic from the system. The system is small and simple, thus the reagent additions and iron bleed can be conducted manually.

The electrowinning feed flow is controlled by a throttling valve to the desired flow rate. Electrical regulation is achieved at the rectifier. Cathode harvesting is a manual process, and can be carried out by shutting down the three pumps. The tankage is designed to accommodate the system draindown.

5 cum/hr (= 5kg/hr)

is this going to be enough? 2?

impellor – high min?

Aqueous or organic continuous??

Ok.

SIZE?

Why a throttle control valve?

SIZE? 5m³/hr

Recycle.

Ok.

Ok.

Feed inchin wall

294249

DESIGN CRITERIA AND PROCESS DESIGN CALCULATIONS
DEMONSTRATION PLANT FOR EURALBA MINING
MT. LYELL WASTE DUMPS

PROCESS CALCULATIONS

IRON IN PLS	2 g/l	ESTIMATED
IRON BLEED	0.5 g/l	NOMINATED
FEED FLOW	5.04 M3/HR	SELECTED FOR DESIGN
Cu TENOR	0.12 g/l	ESTIMATED
Cu RECOVERY	90 %	PROVISIONAL ESTIMATE
AVAILABILITY	100 %	✓
Cu FLOW IN PLS	0.54 KG/HR	✓
RAFF. TENOR	0.012 g/l.	depends on recovery.
Cu/Fe REJECTION	500	DESIGN ✓
aqueous entrainment		
EXTRACT AQUEOUS CONT.	500	DESIGN ESTIMATE
STRIP ORGANIC CONT.	600	DESIGN ESTIMATE
SPENT Cu TENOR	7 g/l	SELECTED FOR DESIGN
SPENT ACID TENOR	180 g/l	SELECTED FOR DESIGN
PLS pH	2.55	ESTIMATED
PLS ACID TENOR	0.28 g/l	
IRON BLEED FLOW	0.009513 M3/HR	ITERATE ****
Cu TRANSFER BY SX	0.61 KG/HR	
Fe TRANSFER BY SX	0.001 KG/HR	
Fe TRANSFER BY ENTRAIN.		
EXTRACTION TO STRIP	0.005 KG/HR	
STRIP TO EXTRACTION	0.002 KG/HR	
BALANCE OF Fe TRANSFER	-0.000000	TO ZERO ****
TOTAL Fe TRANSFER	0.0048 KG/HR	
FOR PLS + IRON BLEED		
COPPER	0.13 g/l	
ACID	0.61 g/l	
IRON	1.9972	
TOTAL ACID LOSS	1.823 KG/HR	
Cu TRANSFERRED	0.611 KG/HR	
DELTA TENOR	5	SELECTED FOR DESIGN
SPENT FLOW TO STRIP	0.122 M3/HR	
SPENT EX T'HOUSE	0.132 M3/HR	
RECYCLE ON STRIP	4.927 M3/HR	
STRIP O/A	41.34	
Cu WON (TRANSFER-BLEED)	0.544 KG/HR	
ACID GENERATED	0.840 KG/HR	
WATER LOST BY REACTION	0.154 KG/HR	18.01534
OXYGEN GENERATED	0.137 KG/HR	

— ? how AE ?
✓ 10kg day max. feed.
OK.
check with ICI [but AE may be less]

?
Delta Cu g/l ?
✓
✓
?
CHECK
predict iron build up.

Fe = 0.0048 KG HR
Cu = 0.5 KG HR
What is max Fe.
Fe in RE ?
? 2 g/l. ?

294250

EW MASS BALANCE

	FLOW M3/HR	g/l Cu	KG/HR Cu	g/l acid	KG/HR acid	g/l Fe	KG/HR Fe
ADVANCE EX STRIP	0.122	12.55	1.533	173	21.149	0.54	0.066
ADDITION	0.010				1.823		
ADVANCE ELECTROLYTE	0.132	11.63	1.533	174	22.972	0.50	0.066
EW FEED	3.600	7.15	25.744	179.8	647.160	0.50	1.800
Cu WON/ACID MADE			0.544		0.840		0.000
RECIRCULATED SPENT	3.468	7	24.278	180	624.299	0.5	1.734
SPENT EX EW	0.132	7	0.922	180	23.701	0.5	0.066
BLEED	0.010	7	0.067	180	1.712	0.5	0.005
SPENT TO STRIP	0.122	7	0.855	180	21.989	0.5	0.061

MASS BALANCE OVERALL

	FLOW l/hr	Cu g/l	Cu kg/hr	acid g/l	acid kg/hr	pH	factors
PLS	5040	0.12	0.605	0.28	1.393	2.55	
iron bleed	9.513	7	0.067	180	1.712		
SX feed	5050	0.13	0.671	0.34	3.105	2.46	
raffinate	5050	0.12	0.606	0.78	3.945	2.10	1.54
loaded organic	5040	0.516	2.601				1
stripped organic	5040	0.395	1.990				
delta organic		0.121	0.611				
advance electrolyte	122	12.55	1.533	173	21.149		
copper transferred			0.611				
copper production			0.544				98.078
spent electrolyte	122	7.00	0.855	180	21.989		

ATOMIC WEIGHTS		ACID R
Fe	55.847	1.756183
Cu	63.54	1.543556
SO4	96.0616	
Fe	55.847	
H	1.00797	
O	15.9994	

SPENT ELECTROLYTE FROM CELLS			
Cu	7 g/l	2.315 KG/HR	CuSO4
Fe	0.5 g/l	0.179 KG/HR	FeSO4
H2SO4	180 g/l	23.701 KG/HR	H2SO4
ION FLOW		26.195 KG/HR	
SG	1.175 KG/L		
MASS FLOW	154.715 KG/HR		
WATER FLOW	128.520		

SPENT ELECTROLYTE TO STRIP			
Cu	7 g/l	2.148 KG/HR	CuSO4
Fe	0.5 g/l	0.166 KG/HR	FeSO4
H2SO4	180 g/l	21.989 KG/HR	H2SO4
ION FLOW		24.303 KG/HR	
SG	1.175 KG/L		
MASS FLOW	143.537 KG/HR		
WATER FLOW	119.235 KG/HR		

STRONG ELECTROLYTE - EW FEED			DEPOSITION CHECK
FLOW	0.122159369		ERR
Cu	g/l	2.859 KG/HR	CuSO4
Fe	g/l	0.179 KG/HR	FeSO4
H2SO4	187 g/l	22.861 KG/HR	H2SO4
ION FLOW		25.899 KG/HR	
SG	1.188 KG/L		
MASS FLOW	145.125 KG/HR		
WATER FLOW	119.226		

EW CELL MASS BALANCE	RATIO		KG/HR
CuSO4		2	1.37
e		2	
H2O		2	0.15
	BAL.		1.521249
H2SO4		2	0.84
Cu		2	0.54
O2		1	0.14
	BAL.		1.521249

ATOMIC WEIGHTS	
Cu	63.54
SO4	96.0616
Fe	55.847
H	1.00797
O	15.9994

294252

DESIGN CRITERIA

COPPER PRODUCTION	10	KG/DAY	
FEED COPPER RECOVERED	0.11	G/L	
FEED FLOW	90909	L/DAY	
	3788	L/HR	
DESIGN PLS FLOW	5040	L/HR	
	84	L/MIN	0.579655
	1.4	L/SEC	
LINE DIAM.	25.4	MM	
VELOCITY	2.76	M/SEC	
MAX HEAD	20	M	
POWER	0.43	KW	
PLS FLOWRATE	5.04	M3/HR	
SX			
SX EXTRACTION	90	%	
PLS TENOR	0.12	G/L	
SALICYLALDOXIME SG	0.94	KG/L	
VOL %	2	%	
DILUENT SG	0.8	KG/L	
EVAP	0.1	MM/DAY	
ORGANIC LOADING			
Cu/Fe SELECTIVITY	500		
ORGANIC ENTRAIN RAFF	70	PPM	
DESIGN Cu DELTA	5	G/L	
EXTRACT CONTINUITY		ORGANIC	
STRIP CONTINUITY		AQUEOUS	
ORG/AQU. RATIO	1		
ADVANCE ELECT TENOR	12	G/L	
DESIGN PLS FLOW	5040	L/HR	
Cu ADVANCE FROM PLS	0.11	G/L	
COPPER RECOVERY	554.4	G/HR	
MIXERS/STAGE	1		
RESIDENCE TIME	2	MINS	
SETTLER RATING	5	M3/M2HR	
ORGANIC SPACE VEL.	0.05	M/SEC	MAX
ORGANIC DEPTH	200	MM	MAX
AQUEOUS DEPTH	400	MM	

12 g/L → 7 g/L.

5

is this enough.

allow free board = 600 mm.

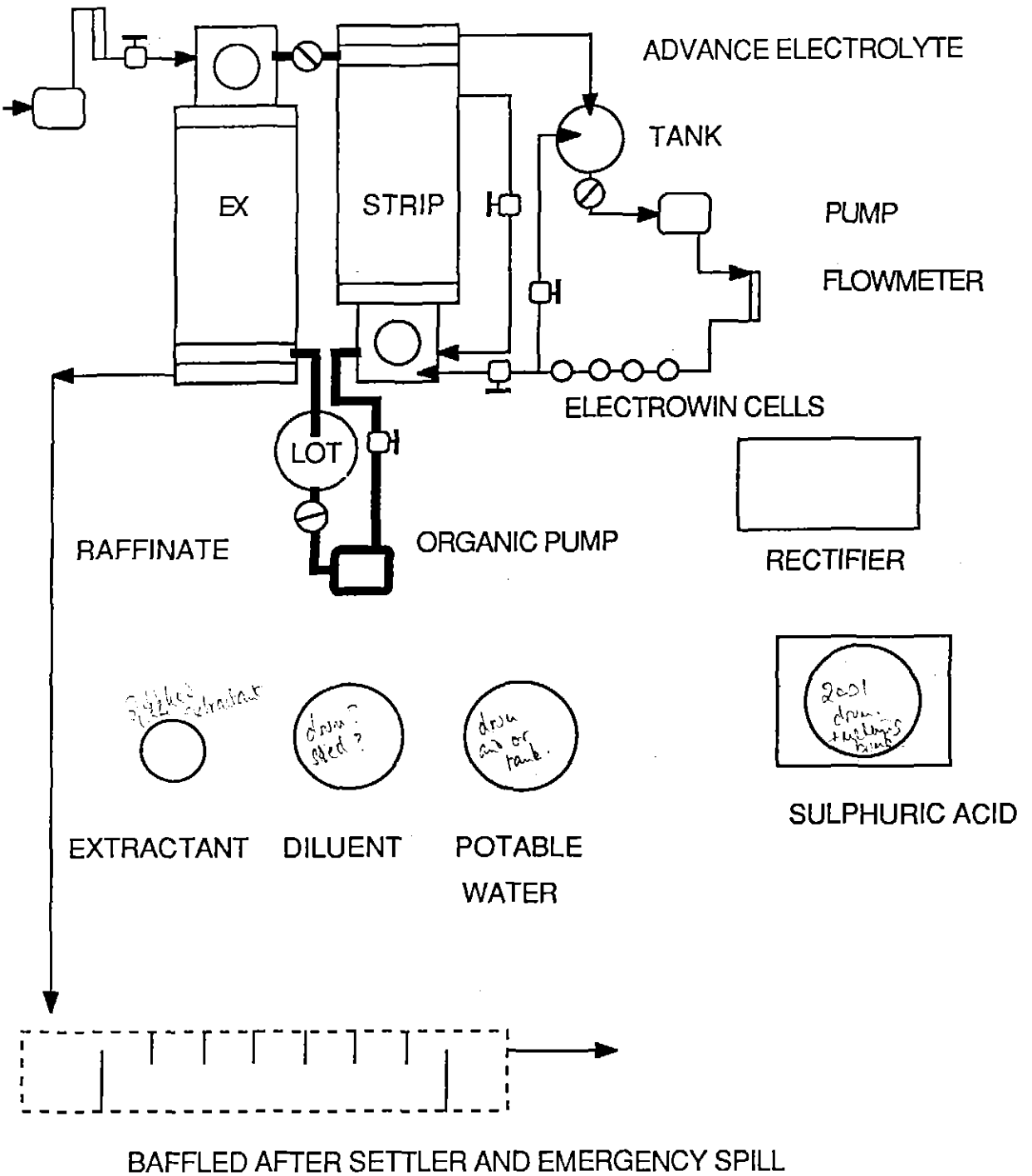
TOTAL LIQUOR/ORGANIC IN CIRCUIT.

294253

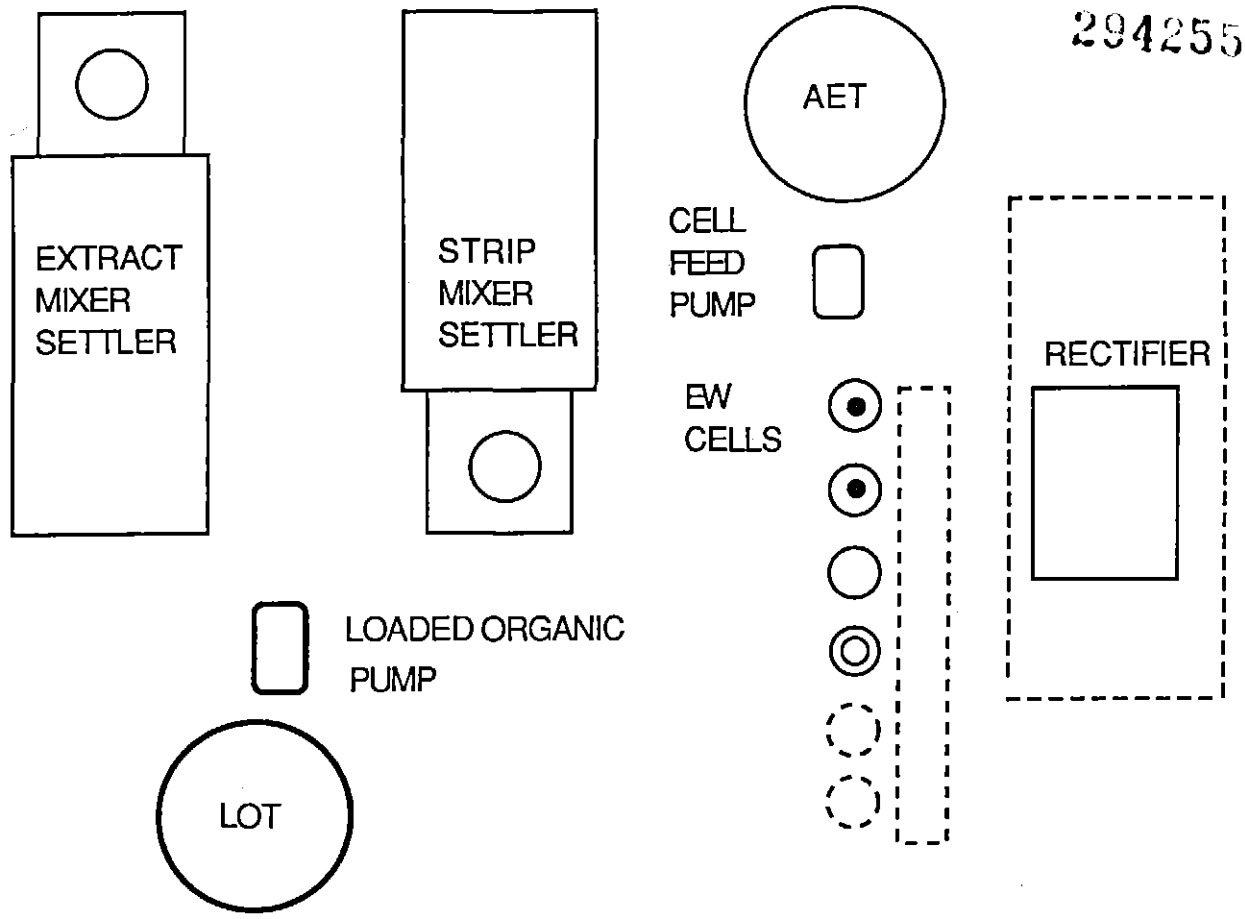
EW

COPPER RECOVERY	554.4 G/HR	
MAX PROD RATE	13.31 KG/DAY	DESIGN
CELL DIAM	0.100 M	
CELL LENGTH	0.850 M	
CATHODE AREA	0.267 M2	
CURRENT DENSITY	300 A/M2	
AMPS/CELL	80 A	
CURRENT EFFICIENCY	99 %	
THEORETICAL CAPACITY	1.185 G/AHR	
NO OF CELLS	~~6~~	
FLOWRATE	3600 L/HR	
ANODE DIAM	0.025 M	
CELL X-SECT	0.007 M2	
CELL VOLUME	0.01 M3	
RT	~~6.26 SEC~~	
	0.96 G/PASS	

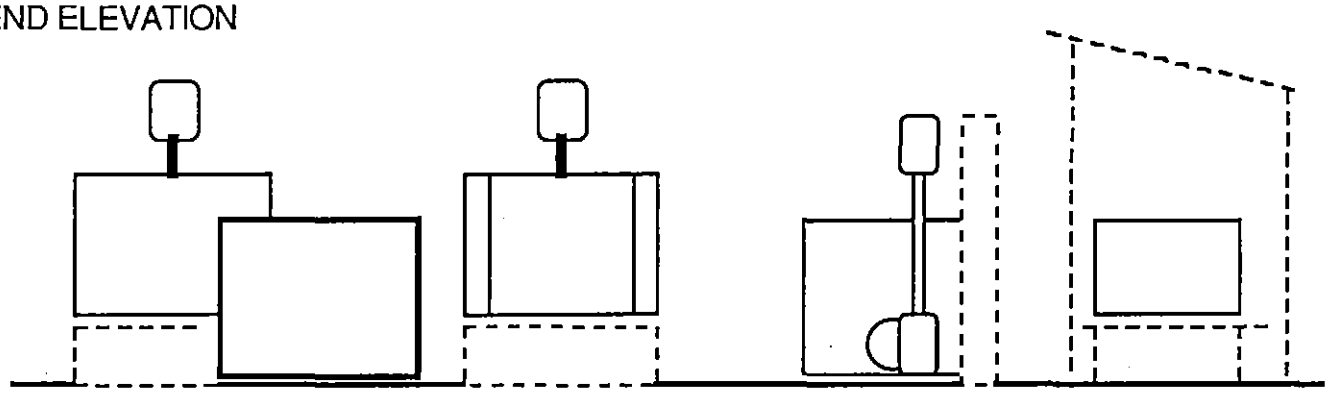
MT LYELL DEMONSTRATION PLANT FLOWSHEET



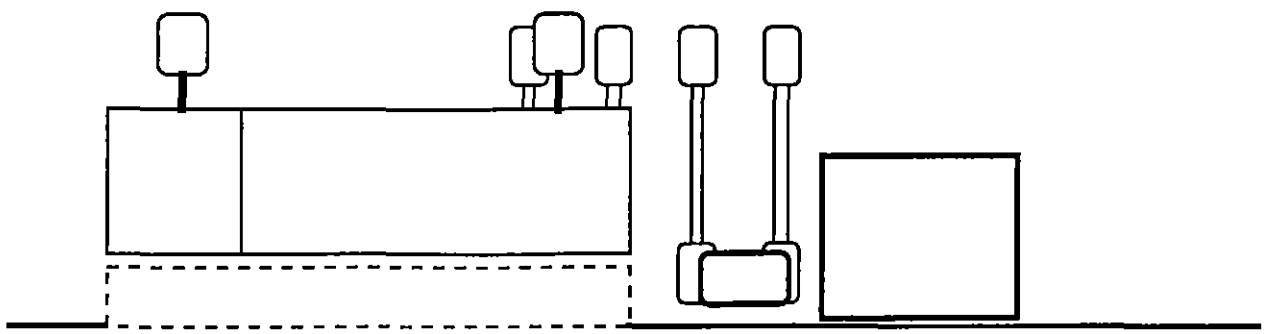
PLAN VIEW



END ELEVATION



SIDE ELEVATION



Mt Lyell demonstration plant conceptual layout

MEMORANDUM

To: Michael Wren
From: Tony Treasure
Date: 23.10.92

Re: COSTING STAGE 2 MT.LYELL PROJECT

The following comprises an assessment of projected costs in the next stage of development of the Mt Lyell project.

Although, as detailed in our recent report, the requirement for continued piloting at Mt Lyell may not be strictly necessary from a process point of view, the plan presented here includes operation of a complete model plant on site for a period of approximately 4 weeks.

The tasks required to take the project through to development of a commercial facility may be generalised as follows:

a) Design, fabrication and operation of demonstration plant.

The facility designed and costed here will be capable of production of approximately 10kg/day of cathode copper.

The system is based on the SX/EMEW treatment option and will comprise in general, the following:

- Mini scale SX plant, incorporating 1 extract and 1 strip stage. It is envisaged that the final production plant will contain 2 extraction stages to provide in excess of 90% copper recovery. Final recovery is not necessary for the demonstration plant, but during its operation the efficacy of two extraction stages will be demonstrated on a batch basis.

- Small scale electrowin plant comprising four production scaled cells

Operation of the plant will establish final operating parameters for the production facility and will allow potential process problems to be addressed prior to finalisation of the development plan.

The costings detailed below provide for fabrication of the plant from scratch, purchase of new equipment and some utilisation of equipment already owned by the company.

b) Detailed design and costing of full scale facility.

In detailed design and costing of the production facility, the company will be relying on the services of one or two consultants who are completely familiar with the installation of solvent extraction facilities.

The exercise will take advantage of existing detailed plant designs, with modification where required. Modified drawings will be produced, which will form the basis for detailed costing of supply and installation.

The provision made, of 40 man days for this work is based on an assessment made by M.Gunn, a process engineer who has been previously responsible for such tasks.

Much of the work entailed can be conducted in conjunction with operation of the demonstration plant on site, physical operation once set up requiring only one man. Direct personnel costs in operating the small plant can therefore be reduced.

c) Assessment of site facilities.

A full assessment of site facilities, such as working area, office facilities, power requirement etc will be undertaken. This will naturally concentrate on the establishment of the operation in such a manner as to avoid interference with the current mining operation.

d) Liquor storage and treatment

The study will examine the possible requirement for liquor storage and treatment facilities. The latter will address the question of entrained solids in the mine effluent (which would result in crud production in the SX plant).

A complete assessment of all previous stream measurements and rainfall statistics for the area will be undertaken, principally with a view to isolation of the target waste sources from external rainfall drainage.

The study will lead to a liquor management plan which will detail requirement for liquor storage, primarily to cater for severe volume variations.

PROJECTED COSTS

a) Demonstration Plant

SX

4 x 250 litre mixing tanks
4 x agitators
1 x 500 litre loaded organic tank
1 x 500 litre advance electrolyte tank
2 x 1m3 settling tanks
1 x 5,000 l/hour pump
1 x 3,000 l/hour pump
1 x 3,000 l/hour pump
Rotameters
Valves and pipework
Sundry fittings

EW	
4 x electrowin cells complete 4 x anodes Cu cathode tubes Rectifier (modified stock item) Rotameter Frames, wiring, piping	
Allowing for some equipment already owned by the company, a budget price for fabrication of the plant has been calculated at:	$25,000
ENGINEERING DESIGN AND PROCUREMENT (Estimate)	$10,000
Sub-total	$35,000
OPERATION	
Personnel	
- Consultant 30 days @ $400 (50% captive to Design and costing programme	$ 6,000
- Assistant 30 days @ $200	$ 6,000
Equipment transport	$ 2,000
Accommodation/meals 30 days @ $150	$ 4,500
Consumables	$ 2,000
Travel	$ 2,500
Supervision/reporting	$ 5,000
Sub-total	$28,000
b) Engineering and other studies	
Plant design/costing 40 days @ $400	$16,000
Site and liquor details 10 days @ $400	$ 4,000
Reporting/supervision 10 days @ $400	$ 4,000
Sub-total	$24,000
TOTAL	$87,000
Contingency	$10,000
TOTAL BUDGET PROVISION REQUIRED	$97,000

It should be noted that there are areas where this expenditure can be reduced (*for example through operating a smaller demonstration plant*) but prudence would dictate that a budget close to $100,000 should be provided for the next stage of the project.

The programme outlined would result in a well established bankable feasibility document.

Appendix a

67 ST. PAUL'S TERRACE
P.O. BOX 544
SPRING HILL QLD. 4004
TELEPHONE: (07) 839 0383
FACSIMILE: (07) 832 0101
TELEX: 44114

MINPROC ENGINEERS PTY. LTD.

A.C.N. 008 992 694

294259

14 August 1992

Mr Tony Treasure
Director – Euralba Mining
23 Riversdale Road
OXENFORD QLD

Dear Sir

On behalf of Minproc Engineers I have reviewed your reports documenting the work carried out by Euralba Mining at Mt Lyell in Tasmania. The reports specifically address the results of electrowinning testwork on low tenor copper bearing solutions emerging from the waste ore heaps.

As detailed in previous correspondence, the Euralba Mining electrowin cell is regarded as a commercially viable and technically sound device for electrowinning metals, with distinct and demonstrable advantages over conventional tankhouses. This opinion is based on detailed study of the technology, with regard to its demonstrated performance on a number of liquors, and a site visit during your Mt Lyell field investigation.

Your report is rigorous, comprehensive and consistent and I support your financial conclusions within the bounds of the stated assumptions. I do not believe that the preferred option, of solvent extraction and your low cost cellhouse, requires a pilot scale testing stage to enable detailed engineering design to proceed with confidence. Solvent extraction of copper is now so widely practiced that design criteria can be developed from laboratory scale tests. Minproc would be prepared to provide appropriate guarantees of performance.

Should you require any further comment or reference, please do not hesitate to contact us.

Yours sincerely

M. J. GUNN
PRINCIPAL METALLURGIST

GN410990.DN

Appendix 3

Euralba
MINING LIMITED
A.C.N. 000 751 093

294260

Ref: SB0141

6 June, 1994

Mr Brian McBride
Assistant General Manager
Mineral Mining Section
Tasmanian Development and Resources
30 Gordans Hill Road
Rosny Park Hobart Tasmania

Dear Brian

Please find the penaltment draft of this feasibility study on the Mt Lyell waste streams. I hope you and the rest of the members of the Task Force look favourably upon the conclusions of this study and also on our submission to treat the waste streams emanating from the remaining, underground and aboveground, low grade ore. The Company, in association with Tasmanian Government, is looking forward to embarking upon an environmental remediation program, initially aimed at eliminating the copper, the most toxic element contained in the Mt Lyell waste streams.

Euralba feels confident that given assistance further inroads could be made into the beneficiation of the waste streams and would look forward to further discussions with the Government on this question.

This "world first" clearly points the way for the mining industry in utilising technology such as ours, which not only will reduce the cost of metal production but also reduces the impact on the environment.

The final version of the report should be ready by Friday 17 June, 1994.

Look forward to your response.

Yours sincerely

Michael Wren
Managing Director

PO BOX R868, ROYAL EXCHANGE, SYDNEY NSW 2000
TELEPHONE: (02) 247 9855 FACSIMILE: (02) 247 9898
REGISTERED OFFICE: LEVEL 3, 37-49 PITT STREET, SYDNEY NSW 2000

Appendix 4

MINPROC ENGINEERS PTY LTD
A.C.N. 008 992 694

67 ST. PAUL'S TERRACE,
SPRING HILL, QLD. 4004
P.O. BOX 544,
TELEPHONE: (07) 839 0383
FACSIMILE: (07) 832 0101
TELEX: 44114

294261

9 June 1992

Mr T Treasure
Director
Euralba Mining
23 Riversdale Road
OXENFORD QLD 4210

Dear Sir,

With reference to our recent meetings regarding Mt Lyell, I am now in a position to present the capital and operating costs for a solvent extraction process feeding the Euralba Mining Direct Electrowinning cell. The costs presented are based on careful analysis of the EMDEW cell capabilities, and adaptation of the solvent extraction process to suit it.

The solvent extraction circuit design is based upon the isotherms calculated by Henckel from shakeout tests on Mt Lyell liquor. The parameters adopted by Henckel are acceptable, and this includes the parallel - series circuit configuration adopted for this analysis. Better than 90% stage extraction was predicted at these low tenors in a single extraction stage (on a stage basis). Unlike a "normal" leaching project, the raffinate will most likely <u>not</u> be recirculated back to the leaching site, and thus this "stage extraction" represents the net copper recovery.

Several features of the overall SX-EW process are worthy of note. In particular, there is no requirement for elaborate organic scavenging from advance electrolyte. In a conventional tankhouse, any residual organic phase collects and concentrates at the electrode "water-line", and burns onto the cathode surface. In the EMDEW cell, the closed turbulent system simply returns any organic content back to the stripping mixer settler feed. The entrainment of organic into advance electrolyte has been costed as a loss because the effect of the EW process on organic is not known. This may well constitute a saving on the costs tabulated and a projected operating cost assuming 50% recycle recovery has been included.

A further simplification is in the recirculation system. In a conventional tankhouse, the total flow cross-sectional area to cathode area ratio is quite high. A high level of spent electrolyte recirculation is required to maintain adequate specific flow rates through the tankhouse. Because the EMDEW cell current densities are so much higher, recirculation is not necessary to maintain specific flow (litres/min per m^2 of cathode). This further simplifies the circuit, and demands only a small tank/pump arrangement to return spent electrolyte to the SX system (with the option to recycle to advance electrolyte).

In a conventional tankhouse the anodes are a lead/calcium/tin alloy, usually cast and rolled to maintain dimensional stability. The EMDEW cell has a sinter coated titanium tube anode, which eliminates the lead contamination problem experienced in many tankhouses. As a result, the major copper cathode impurity in the proposed circuit is iron, and the electrolyte bleed has accordingly been designed to maintain 1.5 g/l iron in the electrolyte circuit. This bleed stream removes acid in the form of 180 g/l acid, $CuSO_4$ and $FeSO_4$ - and there is an operating cost for replacing this bleed with clean acid. The iron level in PLS is only 2 g/l, and at 500:1 Cu:Fe selectivity, very little iron will transfer across the organic barrier into electrolyte. The cost of the bleed calculated to maintain 1.5 g/l is realistic as a 1.5 g/l iron level can be tolerated in electrolyte and produce cathode quality to LME grade A specifications.

The circuit has been conservatively designed using the traditional mixer-settler developed in the 1960-70's. A Krebs double settler system may be worthy of consideration when the project is implemented, giving benefits such as simpler layout and smaller footprint. The agitators specified (but not sized) are derived from the Holmes and Narver system, and the three tanks in the circuit have been sized with allowance incorporated for settler and EW cell draindown, plus adequate surge volume. The pumps have been sized, but further discussions would be required with Mt Lyell to assess the degree of automation and plant downtime which would be required. This in turn will influence the pump materials, control systems and the desirability of a standby unit. This design has incorporated no standby equipment.

Capital cost for a plant capable of extracting and transferring to electrowin feed 45 kg/hr of copper from 330 m^3/hr of feed solution is A$1.23 million. This price includes necessary electrical, piping and instrumentation costs, plus all equipment purchased as proprietary items or fabricated to order, and its installation. The indirect cost of engineering design, project management and procurement is included, plus a 10% contingency. The costs do not allow for the civil works necessary to form the plant site, as these may already be available in the form of a flat site with concrete pads.

The capital costs were derived from data compiled during a 1992 tender for a conventional SX-EW plant. The accuracy of these costs are in the range ± 15 to ± 20%.

The process design was carried out from first principles using the attached design criteria. The design criteria are based upon the Henkel shake-tests and Isotherms, and upon well established practice in similar plants. To derive the operating costs, a detailed mass balance was undertaken. This also enabled calculation of the cost of a first fill of reagents, usually classified as capital and amounting to A$91,151.00.

The operating costs have been calculated for reagent usage only. Power draw by pumps will depend on plant layout which in turn is a function of terrain and aspects beyond the battery limits of this exercise. The equipment list contains a reasonable estimate of installed power.

Manning requirements will be minimal, but may have to conform to other needs and thus may deviate from these recommendations. Assuming continuous operation, only one shift worker will be required per shift, with an extra hand on day work to assist with reagents, copper cathode harvesting and housekeeping.

The operating costs are as follows for reagent consumption (annual basis):

	$/year	$/Kg Cu	Recover 50% of organic
diluent	148039	0.40	0.20
extractant	84594	0.23	0.11
98% acid	18343	0.05	0.05
	250976	0.68	$0.36/Kg Cu produced

The third column represents the anticipated cost where 50% of organic losses are recovered in the raffinate pond, and where 50% of the organic lost from the strip stage is returned to the strip stage in spent electrolyte.

A conservative estimate of power draw in SX is 70% of the total installed power of 107.2 kW, (75kW). At a power cost of 6c/kWhr, the cost will be approximately 10c per kg of Cu produced.

The overall consumable cost will be close to $0.46/Kg of copper produced.

I trust the information included in this letter is adequate for your immediate needs. The process calculations and flowsheet have not been put into a formal format (i.e. typed and drafted) as this would simply add to your costs. This information is available at short notice if required.

Looking forward to your further instructions.

Yours faithfully

M J GUNN
PRINCIPAL CONSULTING METALLURGIST

MT LYELL SOLVENT EXTRACTION DESIGN CRITERIA

Feed rate		m^3/hr	33.0	max.
Plant availability		%	95	annual overall
	days		346	
	hours/day		24	
	days/week		7	
PLS - dump drainage analysis				
	Cu	g/l	0.150	
	Total Fe	g/l	2.0	
	Fe^{3+}			
	Fe^{2+}			
	H_2SO_4	g/l	0.1	
*	Si			as SiO_2
*	TSS	ppm		
	pH		2.55	
*	temperature	°C		
*	viscosity	cP		
*	Cl			
Extractant				
	type		Salicylaldoxime and modifier	
	SG	kg/l	0.91 - 0.97	
	Viscosity	cP	200 - 45 (@15 - 30°C)	
Diluent				
	type		High flashpoint kerosene	
	SG	kg/l	0.8	
	Viscosity	cP	1.5	(15°C)
	flashpoint	°C	80	(100 kPa) - ASTM D-93
	aromatic content	%	20	max

Organic

	Volume % extractant	%	2
	loading	g/Cu/Vol % extractant	0.50
	Cu:Fe selectivity		500:1 (design)
	Organic entrainment in organic cts	ppm	70
	Organic entrainment in aqueous cts	ppm	100
	Aqueous entrainment in organic cts	ppm	300
	Aqueous entrainment in aqueous cts		250
	Organic entrainment in raffinate recovery		?
	Aqueous entrainment in L.O. recovery		baffles in tank
	Diluent evaporation open	mm/day	2
	closed	mm/day	0.1
	Crud removal	- suction pump into coffer dam.	

Electrolyte

	Copper delta	g/l	20
	Spent Electrolyte Analysis		
	Cu	g/l	30
	H_2SO_4	g/l	180
	Total Fe	g/l	0.5 (design Note)
*	Co	g/l	(100 ideal)
*	Cl	ppm	
	Advance Cu	g/l	50

SX Configuration

Continuous	E_1	aqueous (note)
	E_2	aqueous (note)
	S_1	organic

O/A (organic to aqueous ratio by volume)	-	E		1 : 1
	-	S		1 : 1
O/A advance ratio	-	E		1 : 1
	-	S		99 : 1
Stage efficiency	-	E_1	%	87
	-	E_2	%	95
	-	S_1	%	40
Overall extraction			%	91
Stages - extract				2 in parallel
- strip				1
Addition point of spent electrolyte				Strip first mixer
Mixer retention time	-	E	sec	180
	-	S	sec	120
Settler rating			$m^3/h\ m^2$	5
Fire protection				water deluge

MT LYELL SOLVENT EXTRACTION

PRELIMINARY EQUIPMENT LIST

Item	No	No. of Drives	kW	TOTAL kW
Extraction Mixer Settler	2	2	22	44
		2	11	22
Loaded Organic Tank 4 m ø x 6 m	1	-		
Loaded Organic Pump	1	1	7	7
Stripping Mixer Settler	1	1	22	22
		1	11	11
Advance Electrolyte Tank 2 m ø x 3 m	1	-		
Advance Electrolyte Pump	1	1	1.1	1.1
Spent Electroylte Tank 1.5 mø x 2 m	1	-		
Spent Electrolyte Pump	1	1	1.1	1.1
Heat Exchanger	1	-		
				107.2

@ 70% distribution
av. draw 74 kW

$\simeq$ 1.65 kW hrs/kg Cu

MT LYELL SX PLANT

CAPITAL COST

ITEM	COST (A$)	
Diluent Tank	Supplied by diluent supplier	
Mixer-Settler E_1	190 000	includes civil works, drives and fabrication
Mixer Settler E_2	190 000	
Mixer Settler S_1	220 000	
Loaded Organic Tank	40 000	
Loaded Organic Pump	30 000	
Heat Exchanger	20 000	
Spent Electrolyte Tank	30 000	
Spent Eletrolyte Pump	20 000	
Piping and Valves	115 000	
Electrical	125 000	
Total Direct Cost	1 030 000	
10% Contingency	103 000	
Design, Project Management, etc	97 000	
Total SX Costs	$1 230 000	

Exclusions

- ponds (raffinate/PLS)
- associated pumps
- buildings
- water supply, air supply.

First Fill

		Quantity	Cost (A$)
-	extractant	2207 kg	35 312
-	diluent	118 836 litres	61 795
-	acid (98%)	15.73 tonnes	2 044
			$99 151

Appendix 5

294271

11th July 1994

Mr Michael Wren
Managing Director
Euralba Mining Limited
PO Box R868
ROYAL EXCHANGE
SYDNEY NSW 2000

Dear Sir,

MT LYELL PROJECT

Thank you for your draft submission of 28 April 1994 outlining Euralba's continued interest in Mt Lyell.

As you are aware from our various meetings and previous correspondence, the Mt Lyell Task Force has been actively seeking a new operator to assess, and hopefully exploit, the total remaining resources at Mt Lyell. Following assessment of the proposals received and detailed negotiations, the government has signed an agreement with Gold Mines of Australia (GMA) to take over the mining leases and carry out an assessment of the Mt Lyell resource.

As stated at the meeting between officers from Mineral Resources Tasmania, the Department of Environment & Land Management and yourselves on 14 April 1993, any involvement of Euralba at Mt Lyell would be dependant on the scope of the work proposed by any new operator and could not in any way inhibit that operation.

The first phase of GMA's work programme will involve a complete assessment of the resources and mining and processing options. Until such time as a final feasibility study is completed (possibly January 1996) it would not be possible to define what impact a mining operation would have on potential water sources. While it is likely that some sources would be unaffected, at least for some years, these could only be defined in consultation with the GMA.

.../2

It is noted that your feasibility study is based on treatment of all mine waters at existing liquor tenors. If the Prince Lyell mine were flooded it is likely that the outflow from the cave area would change and the tenor would probably be lower than current conveyor tunnel values.

During our previous discussions it was stressed that it was unlikely that access could be guaranteed to all mine water. It was understood that the EMEW process could be tailored to various sized operations. We believe that your financial studies should be expanded to assessment of individual streams and various combinations thereof, to determine whether a smaller scale option would be economically viable.

The government would not consider any proposal which was not commercially viable and cannot accept any of the conditions which you have proposed regarding funding, compensation, royalties or non-repayment of the Environment Fund loan when the operation is profitable.

The government still considers that Euralba can play a part at Mt Lyell and will continue to pursue with GMA the best way of achieving a mutually satisfactory arrangement.

Yours sincerely,

Brian McBride
Assistant General Manager
MINING & MINERAL PROCESSING

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