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PROCESS EVALUATION OF COLUMN FLOTATION AT ERGO

Jonathan Charles Joshua Eves

A dissertation submitted to the Faculty of Engineering, University of the Witwatersrand, Johannesburg, in fulfilment of the requirements for the degree of Master of Science in Engineering.

JOHANNESBURG, 1990.

I

D E C L A R A T I O N

I declare that this dissertation is my own, unaided work, except where specifically stated. It is being submitted for the Degree of Master of Science in Engineering at the University of the Witwatersrand, Johannesburg. It has not been submitted before for any degree or examination in any other University.

Jonathan Rees

Jonathan Rees

(FOR SIGNATURE)

15th DAY OF AUGUST 1990.

A C H I E V E M E N T A T T A I N E D

Column flotation is an innovation in flotation technology and my research demonstrated the superior performance of column flotation over the technology currently employed on the Witwatersrand gold field.

A B S T R A C T

The static Rand gold price has put pressure on the South African gold mines to improve efficiency. Superior metallurgical performance and lower costs attributed to column flotation prompted the construction of a pilot plant (238 millimetres by 10 metres) at ERGO, an Anglo American tailings retreatment plant.

The column out-performed a conventional batch cell, showing a 13 percent improvement in gold recovery at a 29 percent sulphide sulphur grade. Because overall column performance was dependent on froth phase recovery, the column could easily be operated suboptimally. The USBM bubble generator, which worked well with potable water, tended to clog up with plant water.

The column's carrying capacity was a tenth that predicted by a d_{80} correlation. A correlation using the Sauter mean diameter, the $d_{(3,2)}$, as the correct measurement of particle size for carrying capacity is presented for Witwatersrand ore.

Recovery of coarse material by the column was superior to that of the existing plant.

IV

TO THE LORD JESUS CHRIST

A C K N O W L E D G E M E N T S

I would like to thank the Manager, Mr J.A. Turnbull, and the Head of the Metallurgical Laboratory, Mr C.G. Schmidt, of the Anglo American Research Laboratories for the opportunity to study for this degree. I am also grateful to Mr P.J. Mostert, Manager of the East Rand Gold and Uranium Company, for providing the financial and engineering support.

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CHAPTER 1

INTRODUCTION

1.1 CURRENT STATE OF THE SOUTH AFRICAN GOLD MINING INDUSTRY

40 percent of South Africa's gold will be produced at a loss in 1990 if current trends continue (Lunsche 1990). This situation has been brought about by a static Rand gold price and high domestic inflation since 1987. Indications are that for this, and the following reasons, several gold mines will be forced to close by the end of the year.

The political environment of the gold mining industry is uncertain, with the unions and the African National Congress (ANC) calling for nationalization of the industry to see it used for the economic upliftment of the country's poor.

Also significant in the current political environment is the effect of labour action on production. During January to March 1990, Vaal Reefs lost the equivalent of two days production due to labour unrest.

The mines have been caught in a profit squeeze for some time now, as the dollar gold price remained static but the average cost of gold production per kilogram rose by approximately 17 percent per annum. In an attempt to offset this the government devalued the Rand to allow an average annual increase in the Rand gold price of 9 percent over the last ten years. No further devaluations are being contemplated.

Furthermore, South Africa's gold mines are ageing and their profitability is declining. Costs naturally increase as a mine ages and the gold grade of the mined ore decreases.

Capital expenditure has been cut back by all the mining houses to help keep profitability up. Mines require capital expenditure to open up new mining areas and to maintain and improve mine efficiency. Therefore reducing capital expenditure can only be a temporary measure if the mine is to remain viable.

Two different strategies have been adopted by the mines to cope with the financial pressure.

Some mines, particularly those managed by Gencor, are mining more selectively to keep profits up. They mine fewer tons of ore, but of higher grade. These mines are kept profitable in the short term, but if the gold price does not rise the strategy will shorten the mines' lives.

Anglo American, on the other hand, has tended to keep its mills full at the expense of grade. The success of this approach depends on all operations running at optimum efficiency.

1.2 FLOTATION PLANT PRACTICE ON ANGLO AMERICAN GOLD MINES

Flotation in the Anglo American gold mines is used as a scavenging process to recover gold from gold plant tailings. Because a significant quantity of gold is locked in pyrite, the pyrite is floated along with the gold. The pyrite is then roasted to liberate the locked gold.

As the financial pressure on the gold mines increases, flotation as a gold concentrating process will play an even more important role. Work at the Anglo American Research Laboratories has shown that more gold can be recovered if the flotation process is moved from

the back to the front of the circuit (Tumilty et. al. 1990). The gold plant feed can then be separated by flotation into a high and a low grade fraction (Figure 1.1), and the concentrate can be subjected to intensive leaching.

Existing flotation plants are limited by the efficiency of conventional flotation cells and operate in the region of 45 percent gold recovery. If the efficiency of the existing flotation plant is to improve, as for survival it must, then a more effective flotation cell must be developed. The new cell must yield higher gold recovery and cost less to build and operate.

1.3 OVERALL AIMS OF THE STUDY

Given the financial constraints on the gold mining industry at present, efficient recovery of gold is vital. Because gold is such a valuable metal, small improvements in flotation plant performance mean large increases in mine revenue.

The main objective of this study was to quantify the improvement in gold recovery using column rather than conventional flotation, under plant conditions.

1.4 SPECIFIC AIMS OF THE STUDY

In 1986 Dobby and Finch wisely commented "The column differs dramatically from conventional mechanical flotation machines, both in design and operating philosophy, which has been a principal reason for its slow acceptance by the mineral industry" (Dobby and Finch 1986). The project was to overcome this barrier and to make column

flotation technology available to the Anglo American Corporation. The specific aims of the project were:

a) to evaluate the suitability of column flotation for the flotation of the Witwatersrand gold ores

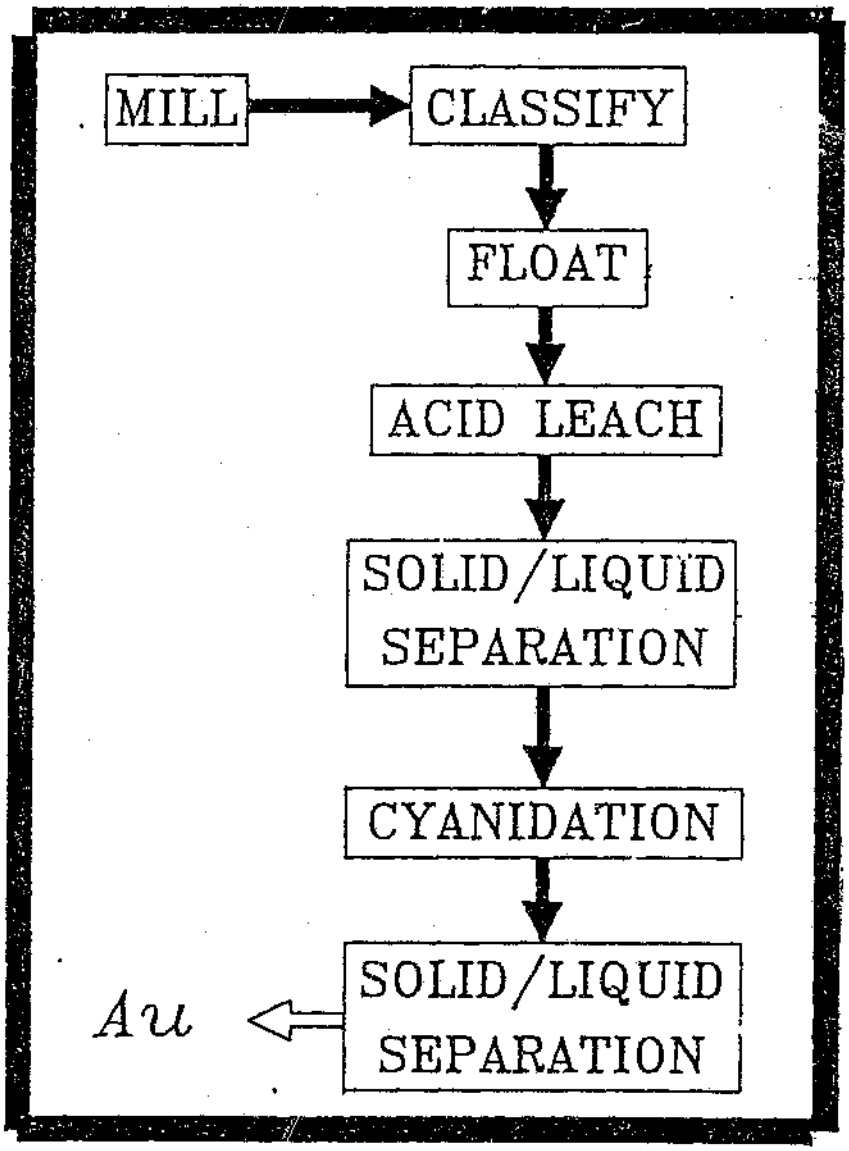
b) to learn an operating philosophy for column flotation of Witwatersrand gold ores

c) to identify the important design parameters

d) to quantify and develop suitable models for important parameters, e.g. flotation kinetics.

Figure 1.1
The FUG Circuit

(Tumilty et al 1990)



CHAPTER 2

COLUMN FLOTATION PLANT PRACTICE, DESIGN AND MODELLING

2.1 INTRODUCTION

The concept of column flotation was developed in Canada by Wheeler and Tremblay around 1960. An attempt was made to introduce a commercial scale machine in the mid-sixties. This failed and there was little further development in the non-communist countries until the patent expired in the 1980s. Since then, many pilot and commercial columns have been built and operated on various ores throughout the world. This situation has led to various designs (Figure 2.1) and operating philosophies, and conflicting views on the effect of the operating parameters on column performance. The variation in design and operating philosophies make it difficult to generalise and describe the typical flotation column. However, most columns fit into the description given below.

2.2 DESCRIPTION OF THE FLOTATION COLUMN

The basic flotation column is depicted in Figure 2.2. Ground ore is fed into the column approximately two thirds of the way up. The ore can be fed through the side of the column or through a vertical pipe positioned in the centre of the column.

As the feed descends, the particles meet a rising stream of bubbles, produced by a sparger situated near the bottom of the column. This section of descending material is called the collection or recovery zone and its height determines the recovery of valuable

Figure 2.1

Various Designs of flotation columns

(Tyurnikova and Naumov 1981)

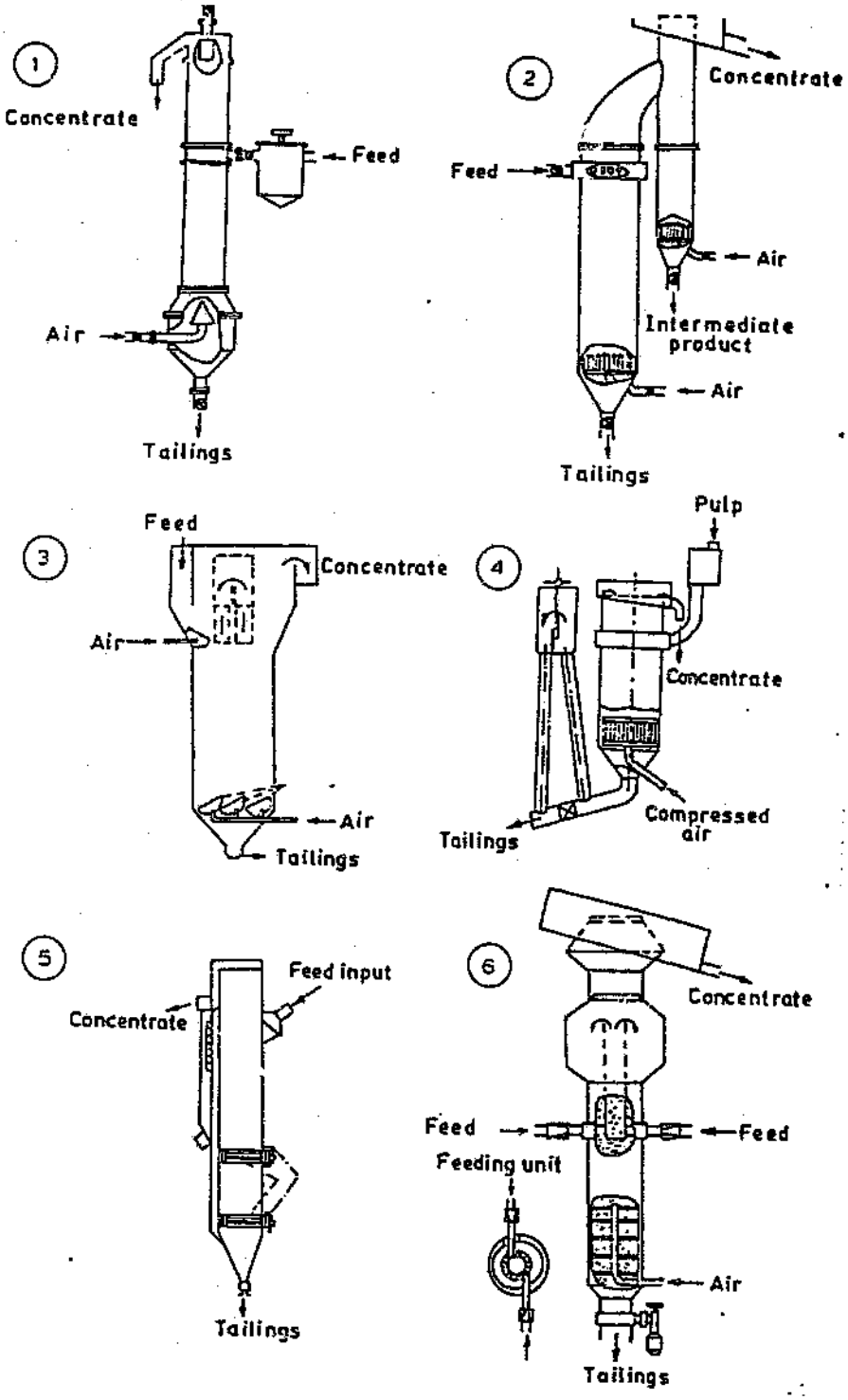
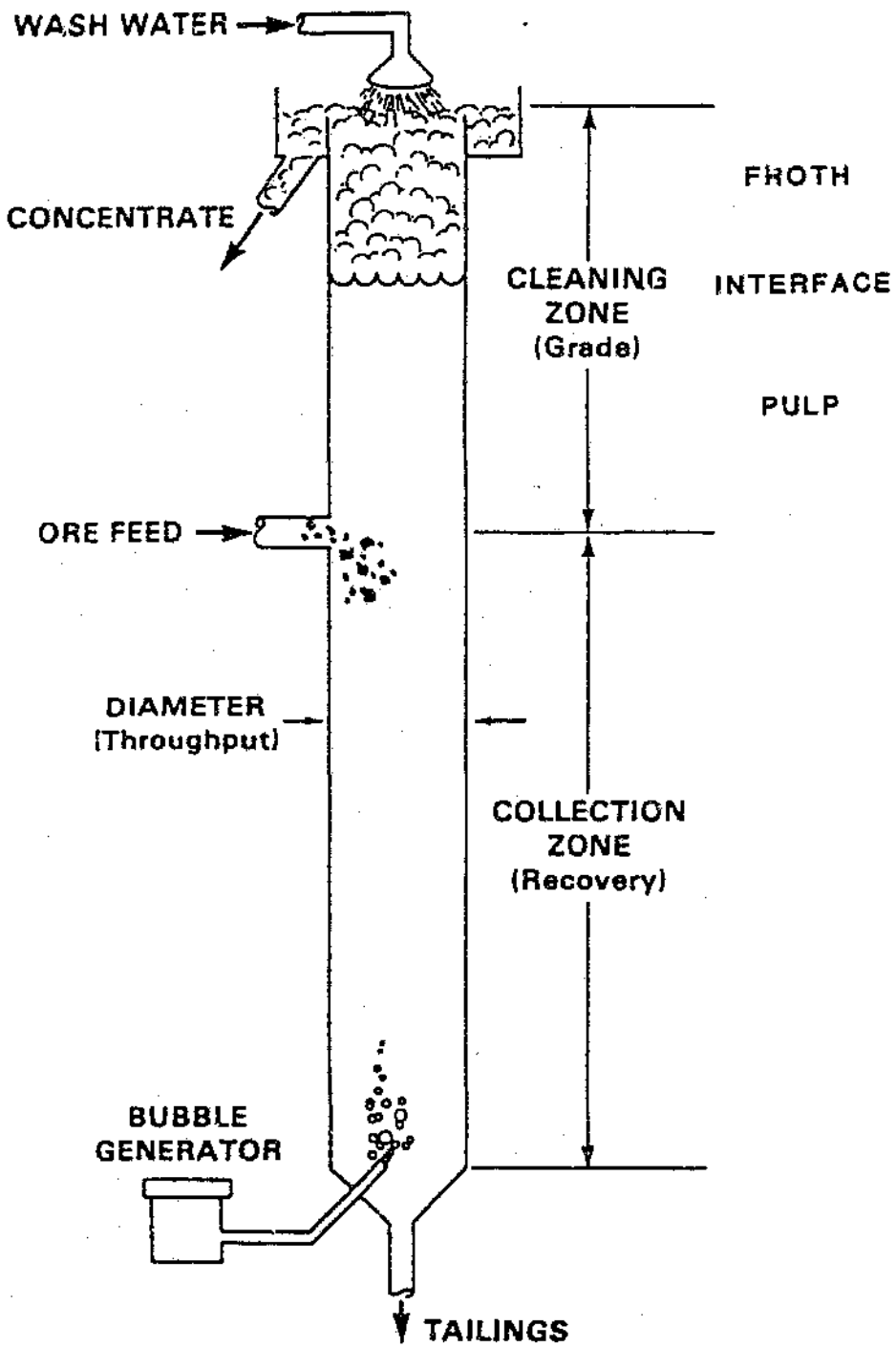


Figure 2.2

SCHEMATIC DIAGI'AM OF THE FLOTATION COLUMN

(Mc Kay et al. 1988)



minerals, which attach themselves to the rising bubble stream. The non-flotable gangue passes out of the bottom of the cell in the tailings.

The region above the feed entry but below the cell lip is called the cleaning zone. The concentrate grade is a function of the height of this zone, especially the froth height above the froth/pulp interface. The clean wash water added to the top of the froth displaces gangue out of the froth and therefore improves grade.

The wash water is also used to stabilise the froth in the deep froth bed. The wash water can be distributed above the froth or in the froth itself.

The critical parameter defining the capacity or throughput of the column is the diameter or cross-sectional area. The column cross-section can be either square or circular and the decision taken depends on structural rather than metallurgical considerations (McKay et al 1989).

2.3 COMPARISON BETWEEN THE COLUMN AND THE CONVENTIONAL CELL

The mechanism of particle-bubble attachment differs between conventional and column cells. In the conventional cell, the rotating impellor breaks down the stream of incoming air into small bubbles, which then attach to particles. The turbulence from the impellor can separate a weakly-attached mineral particle from a bubble.

The column is a counter current contacting device and relies on plug type flow for good performance. A good approximation of plug flow is achieved in laboratory columns because of their narrow

diameter and high length to diameter ratios (Dobby and Finch 1986). In the early commercial cells, however, poor gas dispersion and large diameters gave rise to axial mixing and poor performance. The problem of axial mixing has been reduced by the use of baffles in the recovery zone.

Unlike conventional flotation, descending mineral particles in the quiescent recovery zone of the column cell meet rising air bubbles continuously over two thirds of the length of the column. Providing plug flow or near plug flow conditions are maintained, every mineral particle spends its total residence time near an air bubble, thus effectively increasing its active residence time over that found in conventional cells.

Work done by Whelan and Brown (1956) has shown that attachment efficiency between a particle and bubble is dependant on the angle between the particle and bubble vectors on collision. An angle of zero degrees or a 'head on' collision is most efficient for attachment. Attachment efficiency is also dependant on the relative velocities between particle and bubble. A relative velocity in the range 7-9 cm/sec was found to be the most efficient for coal particles (Tyrunikova et al 1981). The conditions for efficient attachment are more likely to occur in the quiescent collection zone of the column, than in the agitated turbulent conditions of the conventional cell.

The flotation activity of air bubbles increases with air bubble conditioning time (Tyurnikova et al 1981). The substantial height of the column allows bubble residence times of about 40 seconds in the collection zone, which allows enough time for conditioning. The bubble residence times in conventional cells is considerably less, probably 1-5 seconds.

The combination of counter current flow and quiescent conditions of the column are associated with higher attachment efficiencies than the turbulent, co-current conditions in the conventional cell.

2.3.1 Advantages And Disadvantages Of Column Flotation Cells

Metallurgically, the following advantages are claimed for column cells:

a) Improved recovery

The recovery of both fine and coarse particles is enhanced in the column. The longer 'useful' residence time in the column achieves better recovery of fine particles due to the increased probability of collision. The quiescent conditions allow a large particle to be attached to several bubbles at once, and so be carried to the overflow lip. The absence of turbulence reduces the probability of detachment.

b) Improved concentrate grade

The displacement of process water and its entrained fine gangue by a blanket of clean wash water significantly improves the concentrate grade.

Other advantages given in the literature are:

c) Single unit operation

The inherent short circuiting in conventional cells is less likely to occur in the counter current regime of the flotation column. Thus

a whole bank of mechanical cells can be replaced by a single column cell. This has been demonstrated industrially at the Gaspe Copper Mines Limited (Coffin 1982) where sixteen cells were replaced by a single column. Therefore the requirement of a minimum number of conventional cells, usually six, does not apply to the column cell.

d) Large capacity unit operation

A single column can process the feed to a small plant. At Harbour Lights Mine, Western Australia, (Subramanian et al 1988) a single 2,5 metre column acts as a rougher for a plant processing 100 tons per hour. Often columns are able to process higher feed densities.

e) Low maintenance

There are no moving parts in the column and, because of the quiescent conditions, there is little wear by abrasion.

f) Simple scale-up

The relatively simple mechanism of descending particles and ascending bubbles and the lack of turbulence makes scale-up and modelling easier using conventional chemical engineering techniques (Dobby and Finch 1986).

g) Small floor space

The column is a vertical unit and requires less floor space than conventional cells.

h) Low capital cost

The cost of the column is low because there are no moving parts and electric motors. Often it is a homemade device and can be constructed from standard steel pipe.

i) Low operating cost

The absence of moving parts means that the energy consumption of the column is approximately 50 percent of that of a conventional cell. Table 2.1 shows data generated in China comparing the important cost factors of column and conventional cells of the same capacity.

The disadvantages of column cells noted in the literature are:

a) Problems of plugging and scaling of the sparger

"The sparger is the heart of the column" (Wheeler 1986) and any deterioration in its performance will significantly reduce the efficiency of the cell. The best sparger available to date is the USBM bubble generator. The device is, however, expensive to operate in South Africa, as it requires significant quantities of pure water (about five percent of the column's feed capacity).

b) More reagent conditioning is required

Efficient conditioning of the feed after reagent addition should take place before the pulp enters the column cell due to the absence of turbulence. The turbulent conditions in a conventional cell will, to some extent, compensate for inadequate reagent conditioning.

Table 2.1

COST OF COLUMN AND CONVENTIONAL CELLS OF THE SAME CAPACITY COMPARED.

(Hu & Liu 1988)

	Unit	Column	Conventional
Power consumption	kWh/t	0,477	0,889
Area Occupied:			
Flotation Section	m^2	216	360
Compressor Section	m^2	54	-
Total	m^2	270	360
Floor Area:			
Flotation Section	m^2	256	672
Compressor Section	m^2	540	-
Total	m^2	796	672
Weight of Equipment:			
Flotation Unit	T	18,14	45,387
Auxiliary (pump, compressor, conditioner)	T	10,53	2,72
Total	T	28,67	48,107
Installed Power:			
Flotation Unit	kW	30,2	225,1
Auxiliary (pump, compressor, conditioner)	kW	113,0	24,0
Total	kW	143,2	249,1

c) More difficult startup and shutdown

Care has to be taken during start up and shut down that the tailings line does not block with settled solids or that the contents of the column do not settle out.

d) Low open area

The open area of a column is relatively low compared to that of a conventional cell. The open area to pulp volume ratio of the column is approximately one fifth that of a conventional cell. The bubble surface area for carrying concentrate to the overflow lip is the carrying capacity limitation of the column. This disadvantage becomes important when the majority of the feed is recovered to the concentrate, for example, in coal flotation.

2.4 COMMERCIAL APPLICATIONS AND DESIGNS OF FLOTATION COLUMNS

2.4.1 Mines Gaspé

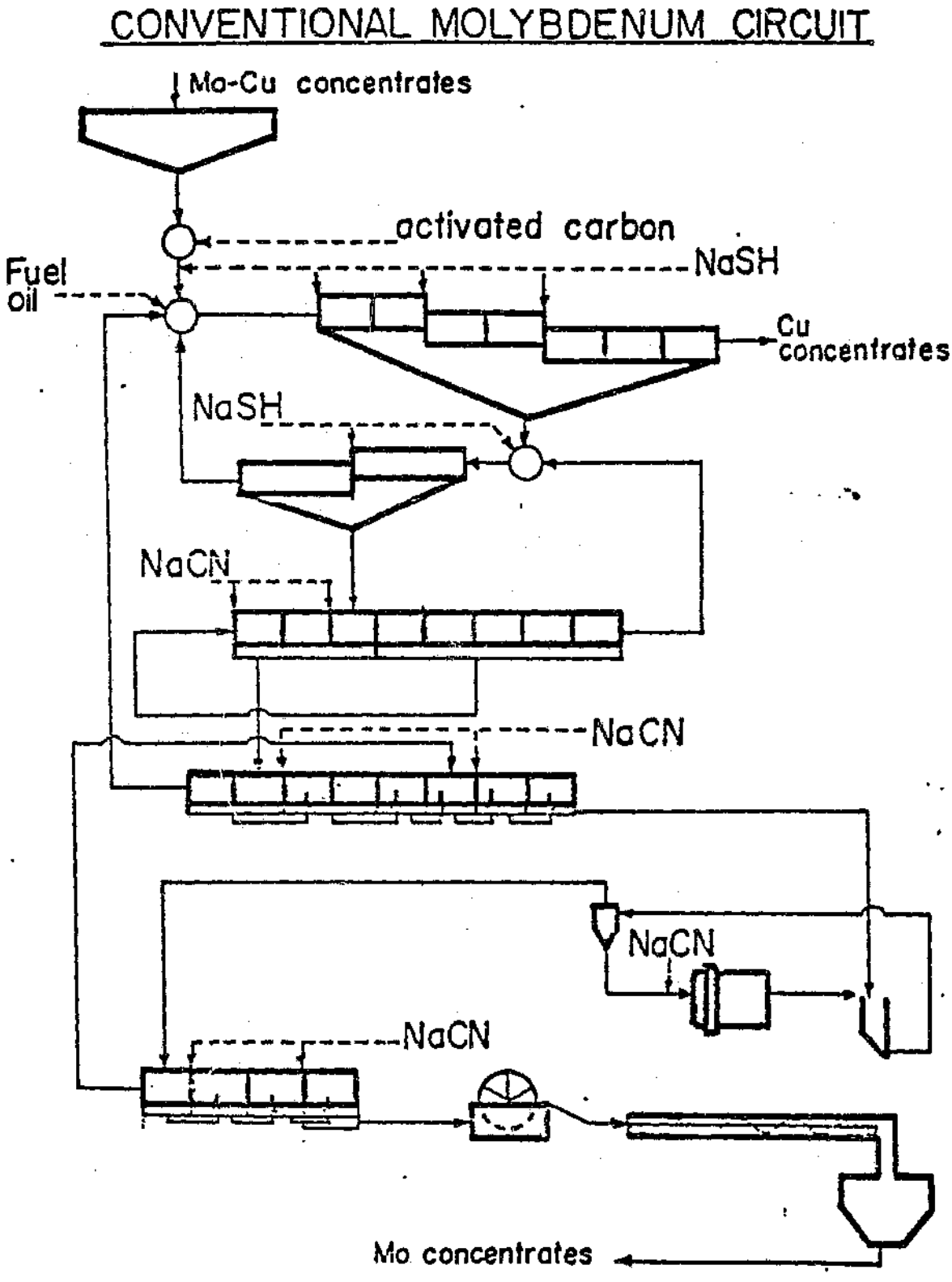
Comparison of the conventional and column circuits at Mines Gaspé, Quebec, Canada (Coffin 1982).

During the eighties, major process and equipment modifications were implemented to cope with changing feed rate and mineralisation. The major change was the replacement of conventional flotation with column flotation in the cleaning section. The bulk copper/ molybdenum concentrates from all flotation circuits were originally

Figure 2.3

Mines Gaspé

(Coffin 1982)



fed to the circuit shown in Figure 2.3. The circuit consisted initially of a bank of seven 8,4 m^3 Wemco cells, the tailings of which was the copper concentrate. All the rougher concentrate was then cleaned in a single stage in two banks of eight 1,4 m^3 Denver DR cells. The first cleaner concentrate was routed to two cleaning stages in a row of eight 1,1 m^3 Denver DR machines. The third cleaner concentrate was floated in six stages of cleaning in another bank of eight 1,1 m^3 Denver DR cells. The ninth cleaner concentrate was pumped into a 15cm cyclone in close circuit with a ball mill. The cyclone overflow was routed to the final four stages

of cleaning in a row of eight 0,65 m^3 cell-to-cell Denver floating machines.

As can be appreciated, the main problems with this circuit were the long retention times and the large circulating loads. The circuit was inherently unstable and difficult to control.

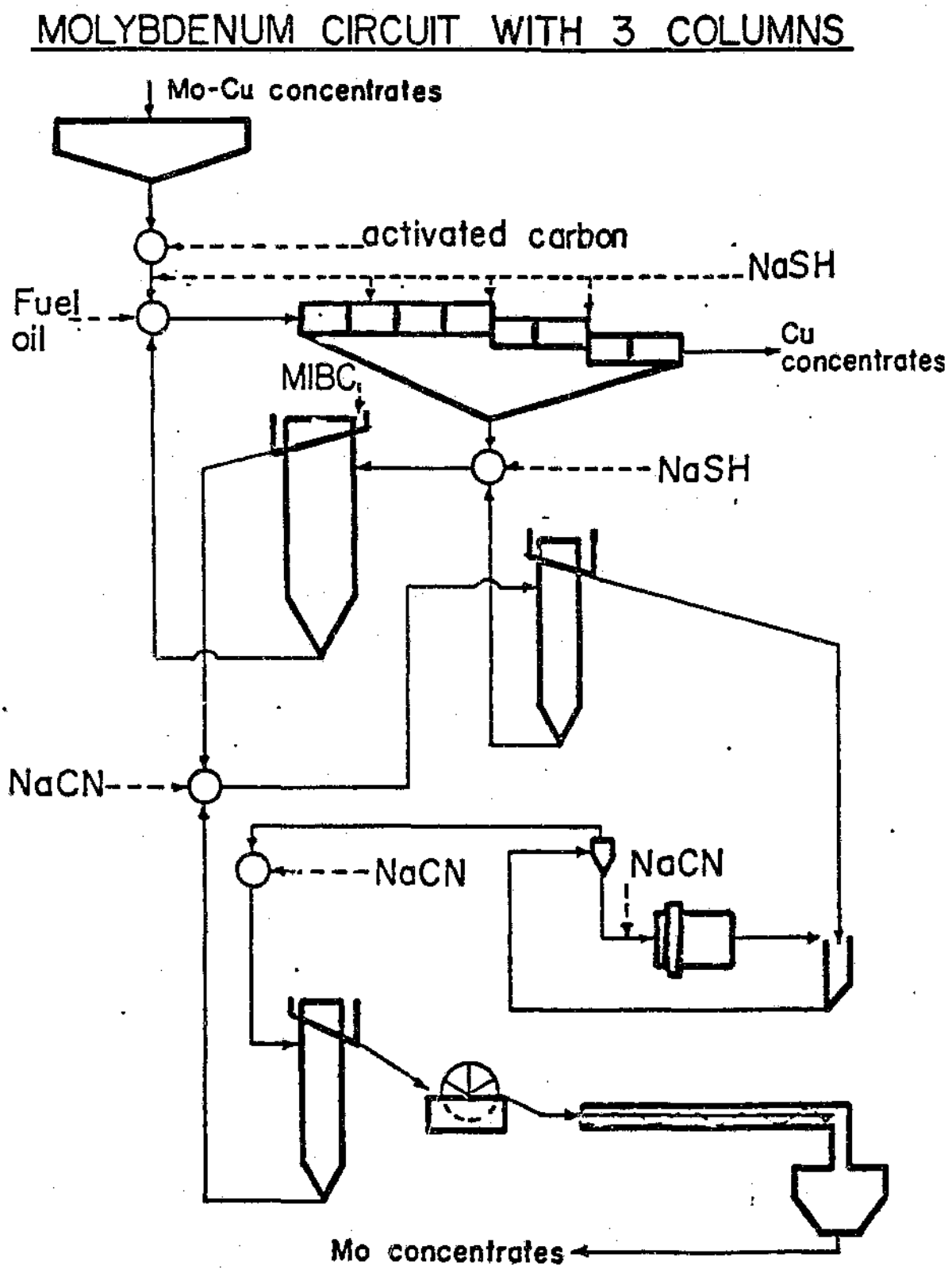
The decision was made to investigate the use of two-stage cleaning by column cells (0,9 m x 0,9 m x 12 m primary cleaner and 0,45 m x 0,45 m x 12 m recleaner). Provision was made for further cleaning of the column concentrates if necessary, and the circuit that evolved consisted of four conventional flotation banks and two columns performing a total of eight flotation steps.

The final step was to replace the first cleaner row of sixteen 1,4 m^3 Denver DR cells by a 1,8 m x 1,8 m x 12 m column. The resultant flowsheet, shown in Figure 2.4, consists of one conventional bank of cells and three columns for a total of four flotation steps. The overall results are summarised in the following table.

Figure 2.4

Mines Gaspé

(Coffin 1982)



Circuit	Recovery %	Upgrading	
		From % Mo	To % Mo
Conventional	64.5	0,69	50,3
3 Columns	80,3	1,00	52,4

Despite the reduction in the number of cleaning stages, the recovery as well as the grade of the molybdenum concentrates improved.

Column design at Mines Gaspé

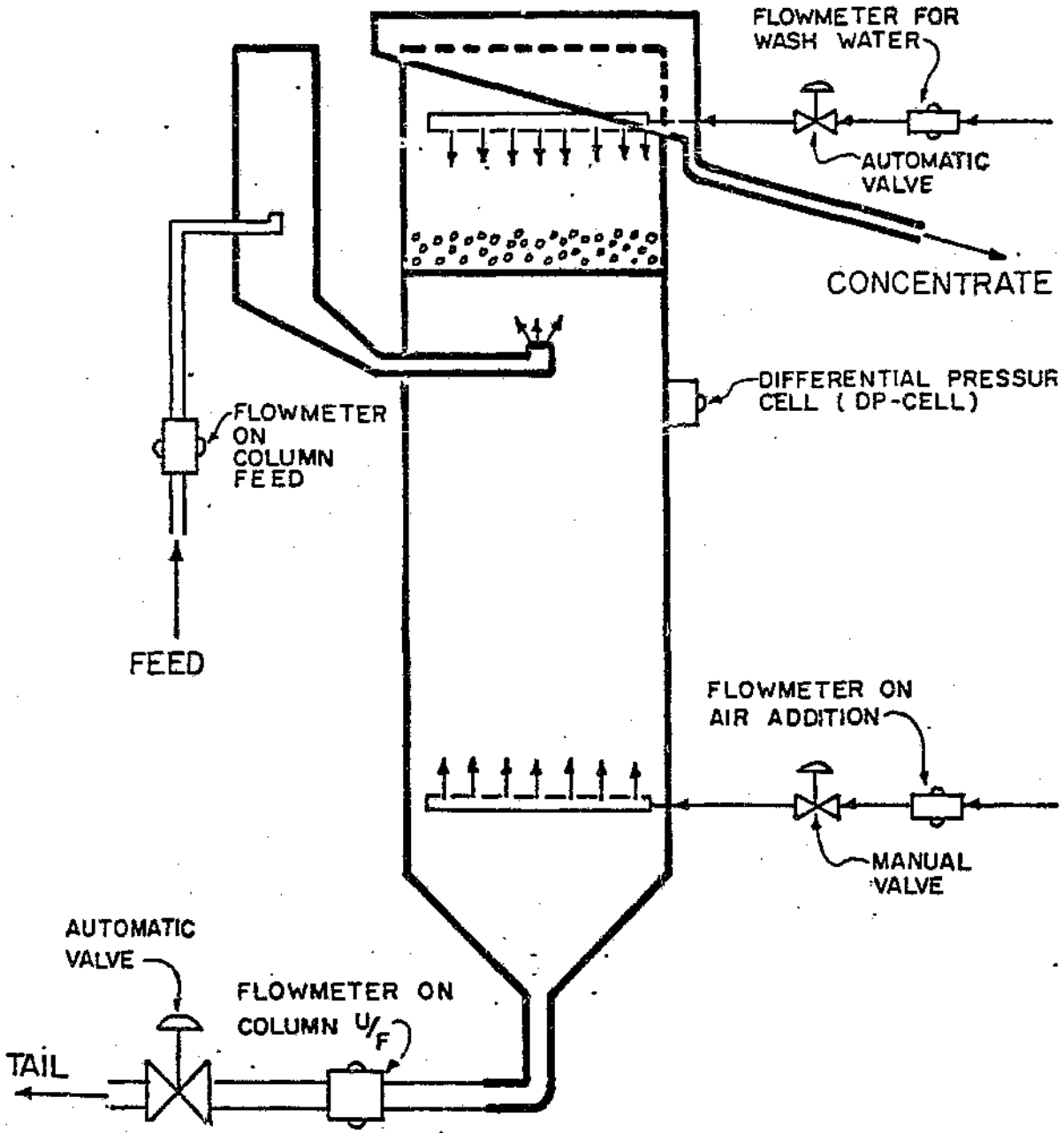
The feed port on the columns was fitted 3 m, a quarter of the column height, below the overflow lip. Wash water spray pipes were installed a few inches below the froth surface. Temporary disruptions of the wash water supply were found to result in a drop in concentrate grade. The columns were well instrumented (Figure 2.5). The tailings flowrate was maintained at a higher rate than the feed by a fixed amount. This control strategy ensured good washing of the froth and a net flow of liquid down the column (positive bias). The pulp/froth level was detected by a differential pressure cell and the froth depth was controlled at a constant value (between 1 m and 1,3 m) by manipulating the wash water flowrate. If the level dropped the wash water flowrate increased and vice versa.

Figure 2.5

Mines Gaspé

(Coffin 1982)

FLOTATION COLUMN AND REQUIRED INSTRUMENT CONTROL LOOPS



Column operation at Mines Gaspé

The following operating factors were found to be important:

a) Feed flow: Stable feed flowrate was critical for good performance. If the design feed rate was exceeded (see Figure 2.6), a dramatic drop in recovery was observed.

b) Air: The condition of the bubbles was important and an on-going inspection and maintenance was required. The bubble flowrate was set in proportion to the feed solids flowrate.

c) Wash water: Water pressure fluctuations caused disruption in the operation of the column and poorer performance.

2.4.2 Harbour Lights

Harbour Lights Mining is a gold mine located in Leonora, Western Australia. The deposit consists of an oxide ore body lying above a sulphide ore body. The oxide ore is treated by a conventional carbon-in-pulp plant and the sulphide was to be treated by flotation. (Subramanian 1988).

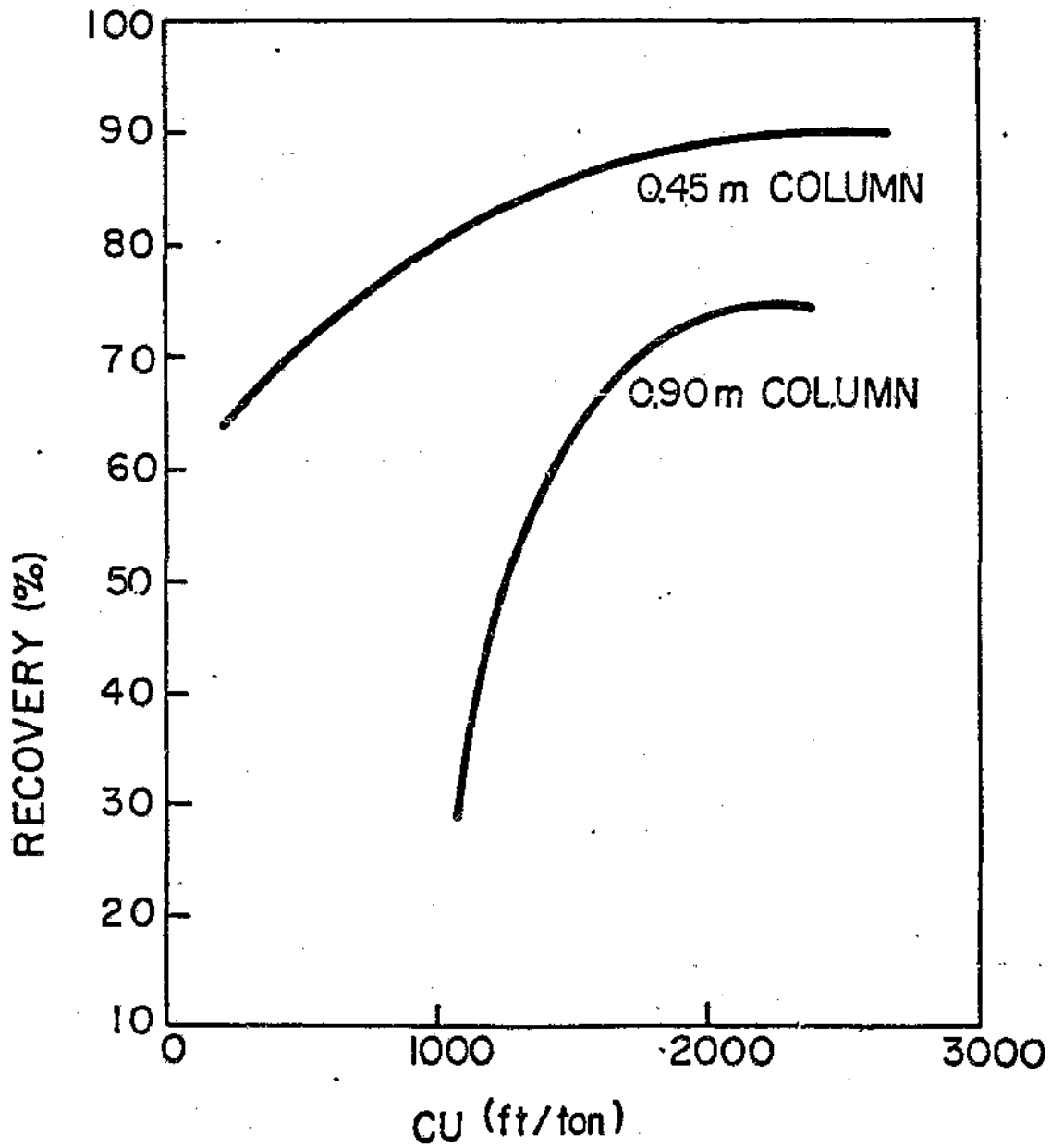
The gold present in the sulphide ore is free gold or is associated with pyrite and arsenopyrite. The sulphide ore contained significant amounts of talc minerals and initial laboratory testwork indicated that this would have an adverse effect on grades. Two pilot campaigns during 1986-1987 proved successful and a commercial plant was constructed and commissioned in August 1987. The final circuit consisted of a primary rougher column 2,5 m in diameter and 12 m high which recovered material at finished grade. The column tailings were treated by six Outokumpu OK17 cells which acted as scavengers. The scavenger concentrate was cleaned in a second

Figure 2.6

Mines Gaspé

(Coffin 1982)

COLUMN RECOVERY VS FLOWRATE PER TON OF FEED



Volume of air
Feed tonnage

column cell which was 1,2 m in diameter and 12 m high. This concentrate was also final grade, while the cleaner tailings were recirculated back to the conditioning tank of the first column. The flowsheet is shown in Figure 2.7. Absolute figures are not given but the following points are listed:

a) Stage 1 (rougher) column flotation produced concentrates equivalent to, or better than, conventional with three cleaning stages.

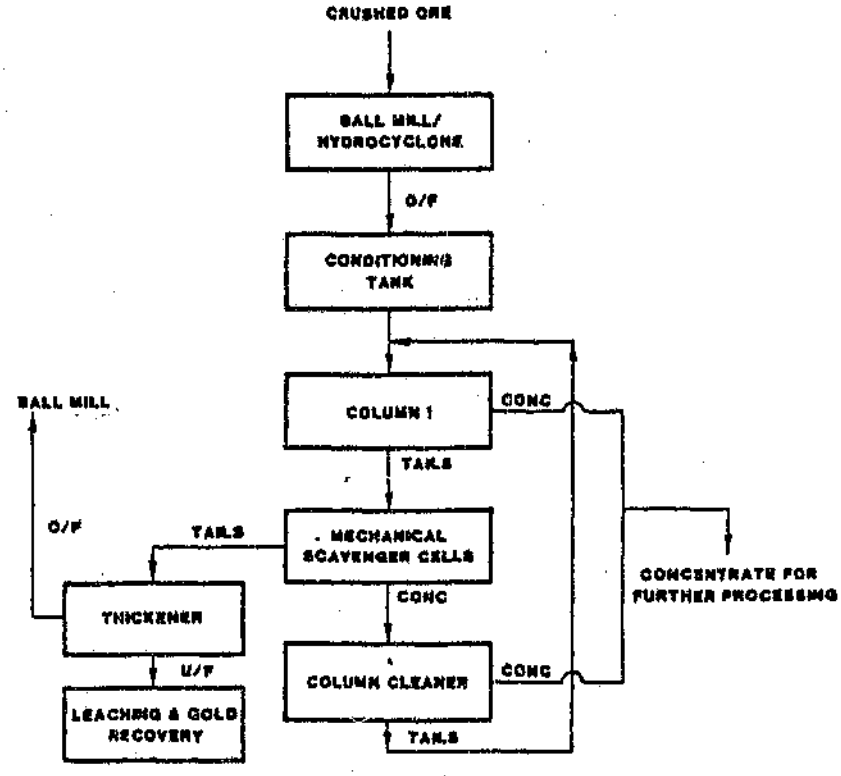
b) The table below compares the results from the column as a single stage rougher to multistage tests in conventional cells. It is noteworthy that the column concentrate grades are better than those achieved in open cycle laboratory tests with two cleaning stages.

Type of cell	Scale	Concentrate Assay		
		Gold g/t	Arsenic %	Sulphur %
Conventional	Laboratory	135	12	28
Column	Pilot	149	11	33

c) The results indicated that an all column circuit was feasible, but it was considered too great a risk at the design stage and therefore a hybrid design was used. Although not specifically stated, the authors indicate that in future they would be prepared to go for an all column circuit.

Figure 2.7

Harbour Lights Flowsheet
(Subramanian 1988)



Column Design at Harbour Lights

The superficial feed and gas rates were 1,6 and 1,5 cm/sec respectively for the larger roughing column. The slurry entered the column through a centre feed well and was distributed by a multiport system to minimise turbulence. Six internal vertical baffles were provided to a height of about eight metres from the bottom to minimise axial mixing. The area of each baffled compartment was 0,8 m^2. The total height of the column was 12 m and the minimum retention time was 11 minutes. The gas sparger consisted of a perforated pipe wrapped in filter cloth.

The wash water distributor resembled a spokes-and-wheel arrangement and acted as a shower, presumably above the froth. Apparently wash water was only used when the concentrate grades dropped.

The design of the 1,2 m diameter cleaner column was similar except that it had one baffle.

The control system involved simple level control. A differential pressure cell was used to measure the interface, which was controlled via a pinch valve on the tailings line. To prevent contamination of the product with feed, an emergency system was installed to direct the feed to the sump if the level in the column approached the overflow lip. The gas and wash water flowrates were set using rotameters. Sight glasses were used to check the froth/slurry interface. No bias controls or overflow measurement devices were installed.

2.4.3. Magma Copper Company

Comparison of Conventional Flotation and Column Flotation at Magma Copper Company, Arizona, USA (Clingan and McGregor 1987).

The mining circuit treats 57 000 tons per day of a porphyry copper

deposit with molybdenum as a by-product. The successful implementation of column flotation in other molybdenum circuits with the reduction in the number of cleaning stages, the increase in recovery and grades and the reduced operating costs cited by molybdenum plants using columns as cleaners, prompted a decision to replace the conventional cleaning section (Figure 2.8) with columns.

Two column cells 1,5 m and 1,8 m in diameter and 12 m high were installed. The height was dictated by the flotation plant building.

The columns were operated in parallel as a single stage (Figure 2.9), and replaced two stage conventional cell cleaning. The columns replaced 36 Agitair 1,13 m^2 cells, 28 on the first cleaner and eight on the final cleaner.

The column volume was 1,23 times the volume of the conventional cells.

The molybdenum recovery to the bulk copper/molybdenum concentrate improved by 5 percent after the installation of the columns. The reason for this improvement is thought to be the superior performance of the columns on fine material. The copper recovery and grades remained similar.

2.4.3 Column Design at Magma Copper Company

The column designs were based on the flotation column scale up method of Dobby and Finch (1985). No pilot plant work was done, and the design parameters were based on the performance of the existing two-stage copper cleaning circuit. Calculations indicated that two columns would be needed with diameters between 1,5 m and 1,8 m. It was decided to build one column of 1,5 m diameter and

Figure 2.8

San Manuel Flotation Circuit with Conventional Cleaning

(Clingan and Mc Gregor 1987)

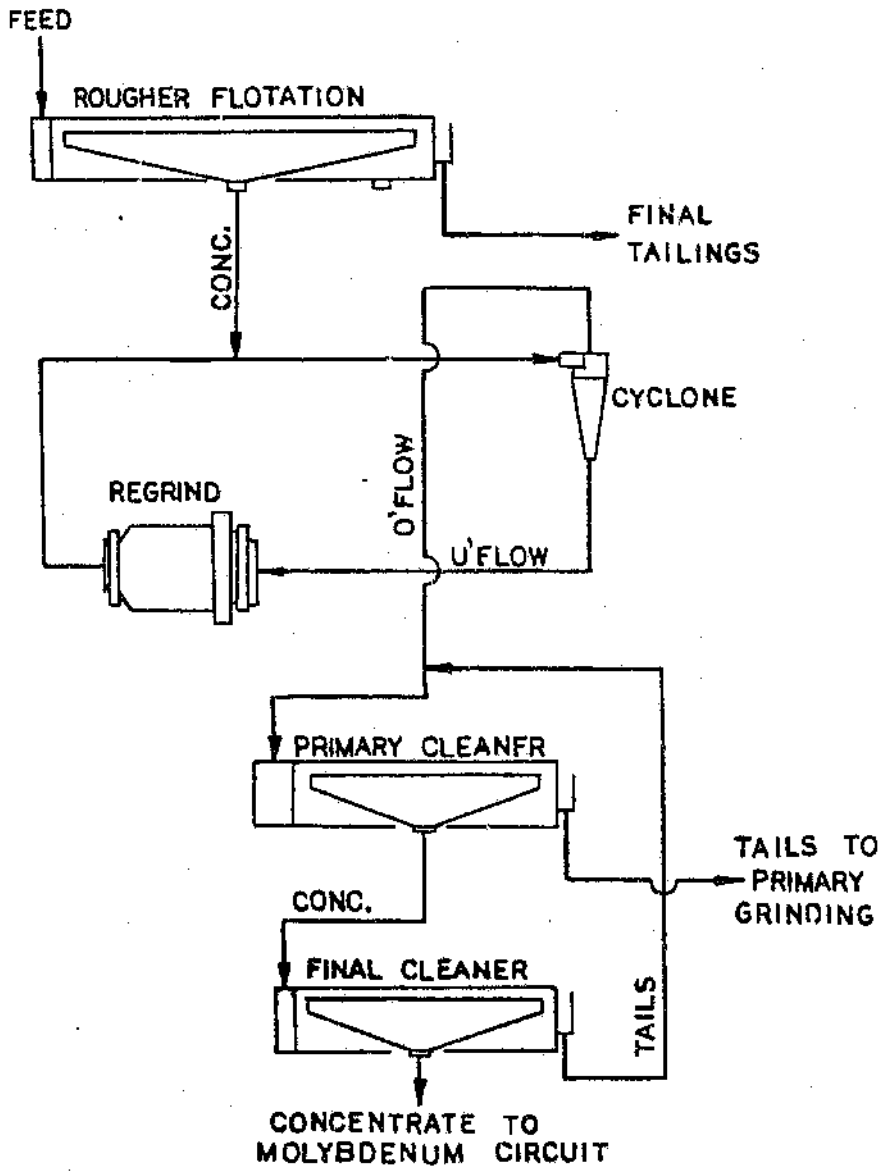
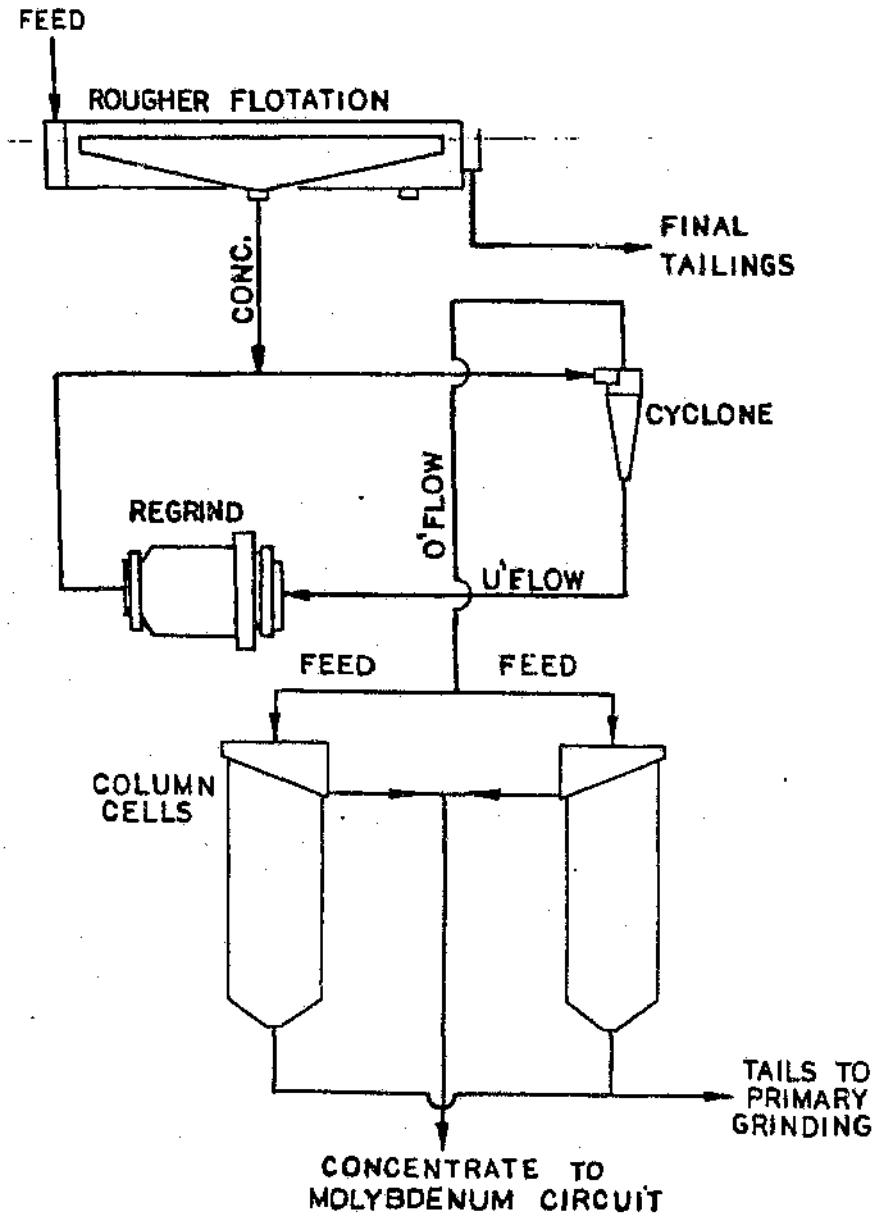


Figure 2.9

San Manuel Flotation Circuit with Column Cell Cleaning

(Clingan and Mc Gregor 1987)



one of 1,8 m to build up design data for future installations. They were designed to be operated in parallel.

An extensive column cell control system, which is described by Amelunxen and Redfearn (1985), was fitted to each column.

A bias ratio control loop adjusted the valve on the tailings line to maintain a constant ratio between the tailings and feed flowrate. The pulp level control loop adjusted the wash water flowrate to maintain a constant pulp level. The gas rate to the cell was controlled by correlating the wash water flowrate to the bias. Part of the air control loop involved measuring air holdup (percent air present in the slurry). Operation of the columns indicated that a simple level control manipulating the tailings flowrate would have been sufficient.

The spargers were fabric-covered filter disk sectors. The sectors were arranged to fit within the cross section of the cell with the narrow end of the sector towards the centre, having a "pie-like" appearance. The tightly woven polypropylene fabric covering the spargers required replacement every two months.

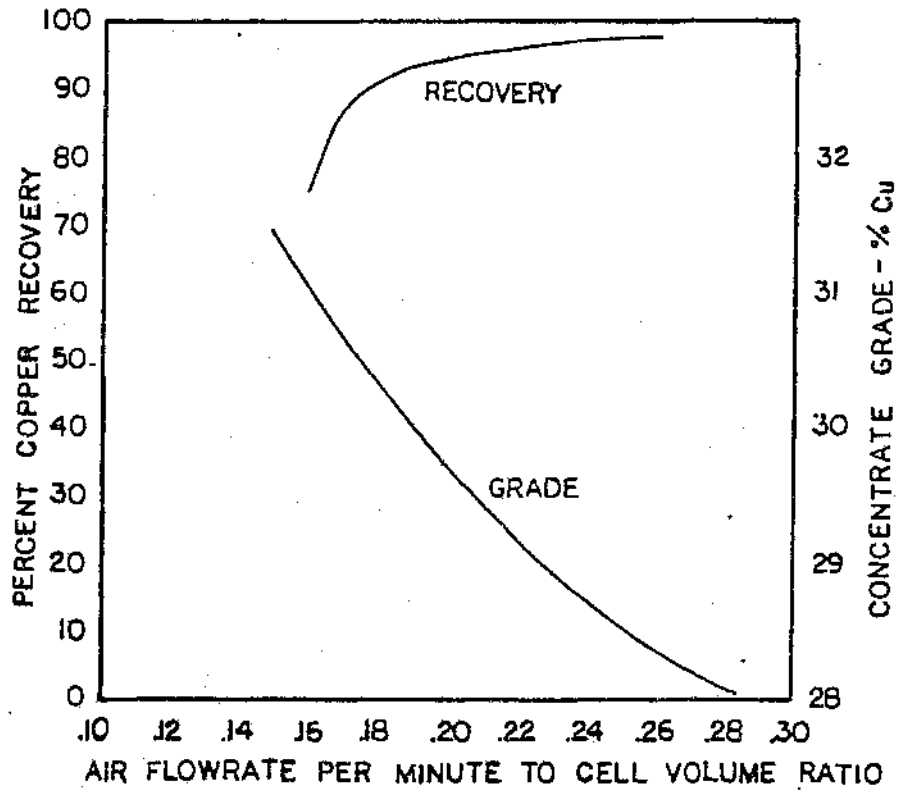
The metallurgical results for the two cells were similar and the additional retention time of 44 percent in the larger column was thought to be offset by a higher sparger surface area to air flow ratio and a smaller bubble size in the smaller column. The greater axial mixing in the larger column may also be responsible for its poorer performance. The froth height varied between 0,6 m to 1 m and no effect was observed on grade or recovery. No difference in performance was observed when the bias ratio was varied between 1 and 1,15. Grade decreased when the bias rate was less than 1 and molybdenum recovery decreased when the bias ratio was greater than 1,15. Gas rate proved to be the most important variable at San Manuel (Figure 2.10). The recovery was

Figure 2.10

Copper Concentrate Grade and Recovery
with Air Flowrate to Cell Volume Ratio

(Clingan and Mc Gregor 1987)

SAN MANUEL



found to improve with gas rate at the expense of grade until bubble coalescence became so extreme that the froth surface became disrupted.

2.4.4 Summary of Applications Worldwide

There are many other applications documented in the literature: Amelunxen and Redfearn (1985), at Gibraltar Mines, Espinosa et al (1989) at Mount Isa Mines, Feeley et al (1987) at Inco's Matte Separation Plant, Mauro and Grundy (1984) at Lorex Mining Company, Fickling et al (1989) at Black Mountain in the Cape Province. Lists of applications of column cells have been published by Hu and Liu (1988) relating to China, Table 2.2, relating to North America by Moon and Sirois (1988), Table 2.3, and worldwide by Yianatos (1989), Table 2.4.

Such lists give an overall view but are out of date as soon as they are published. Clingan and McGregor (1987) and Yianatos et al (1986) published column flotation surveys, Tables 2.5 and 2.6 respectively, which summarise the important design parameters of existing columns.

2.5 CARRYING CAPACITY

2.5.1 Carrying Capacity Models

The carrying capacity of a flotation column in a particular application is the maximum mass flowrate of the concentrate stream divided by the column cross section. It is limited by the capacity of the bubbles to carry mineral to the froth overflow lip. Carrying capacity (C_a) is an important scale-up parameter, especially for fine

Table 2.2

Use of Column Cells in China

(Hu and Liu 1988)

The following flow sheets have been used in practice:

a) Rougher-scavenger: used for simple Cu sulfide ore and phosphate ore.

b) Scavenger alone (conventional flotation cells for roughing): for Cu ore.

c) Two-stage roughing: for complex clayey Cu ore.

d) Rougher-scavenger-cleaner: for complex clayey Cu ore.

e) Rougher-cleaner: for coal and iron ore.

f) Rougher-scavenger (conventional flotation cell for cleaning): For Sn ore, W ore and mixed sulfide and oxide ore of Cu-Pb-Sn-Mo-Bi etc.

Table 2.3

Applications
of Column Flotation in North America

(Moon and Sirois 1988)

1. INCO, Thompson, Manitoba.
2. INCO, Copper Cliff, Ontario.
3. Hudson Bay Mining and Smelting Co, Ruttan Mine, Manitoba.
4. Falconbridge Nickel Mines Ltd, Falconbridge, Ontario.
5. Cominco Ltd, Kimberley, British Columbia.
6. Cominco Ltd, Polaris Project, Greenland.
7. Cominco Ltd, Red Dog Project, Alaska.
8. Utah Mines Ltd, Island Copper Mines, British Columbia.
9. Noranda Mines Ltd, Gaspé Copper Mine, Murdochville, Quebaec.
10. Noranda Mines Ltd, GECO, Manitouwadge, Ontario.
11. Noranda Mines Ltd, Brunswick Mining and Smelting Corp Ltd., Bathurst, New Brunswick.
12. Gibralter Mines Ltd., McLeese Lake, British Columbia.

Table 2.4

Summary of Pilot Plant and Plant Operations - Yianatos (1989)
(* indicates column testing reported)

AUSTRALIA	
Amdel, Adelaide	*several
BHP Central Laboratory	*coal
Blue Spec/Golden Spec W.A.	gold
Harbour Lights, W.A.	gold
Hellyer, Tasmania	lead, zinc, silver
Kambalda Nickel Operations (WA)	nickel
Mount Isa Mines, Queensland	copper, lead, zinc
Paddington, W.A.	gold
Renison Ltd Tasmania	*tin
Riverside Coal Preparation Plant	coal
Tefler, W.A.	copper
Woodcutters, N.T.	zinc
CANADA	
Geco Mines, Ontario	copper, lead
Gibraltar Mines, B.C.	copper
Inco, Ontario	copper
Inco, Thompson	copper
Lornex Mining Co., B.C.	copper, moly
Mines Gaspé, Quebec	moly
Cominco, Polaris	lead, zinc
Niobec, Quebec	*carbonatos, niobium
Noranda, New Brunswick	*copper, moly
Noranda, Mattabi, Ontario	*copper, lead
CHILE	
Cia, Minera del Pacifico	*phosphates
Codelco, Chuquicamata	copper, moly
Codelco, Andina	*copper, moly
Codelco, El Teniente	*copper
Exxon, Disputado, Les Bronces	copper
Exxon, Disputada, El Soldado	*copper
La Escondida	*copper
Mantos Biancos, Antofagasta	copper
Soquimich	*astrakanite
PAPUA, NEW GUINEA	
Bougainville Copper, Panguna	copper

Table 2.4 Continued

PERU	
Cuajone, Tacna	copper
U S A	
Cominco, Alaska (project Red Dog)	*lead, zinc
Cyprus Minerals, Sierrita, Arizona	copper, moly
Magma Copper Co., Pinto Valley	moly
Magma Copper Co., San Manuel, Arizona	copper, moly
Kennecott Copper Co, New Mexico	moly
US Burea of Mines, Salt Lake	*fluorite, chromite

Table 2.5

(Clingan and McGregor 1987)

Column Flotation Survey Summary

	Cell Dimensions Height x Diameter (m)	Height to Diameter Ratio	Type of Circuit	Application	Sparger Type	Air Flow Rate to Cell Volume Ratio	Mineral Recovery (%)	Pulp Level Control	Bias Control	Air Flow Control	Froth Depth (m)
Cyprus Minerals, Sierrita	15.0 x 0.9	16.5:1	Mo	Inter. Clnr.	Fabric	0.08:1	36.2	Auto	Man.	Man.	0.3
	15.0 x 0.9	16.5:1	Mo	Inter. Clnr.	Fabric	0.08:1	36.2	Auto	Man.	Man.	0.3
	15.0 x 0.9	16.5:1	Mo	Inter. Clnr.	Fabric	0.08:1	36.2	Auto	Man.	Man.	0.3
	15.1 x 2.0	7.5:1	Cu/Mo	Final Clnr.	Fabric	0.21:1	97.0	Auto	Man.	Man.	ND
Exxon Minerals, Los Bronces (Chile)	12.0 x 0.9	13.1:1	Cu	Final Clnr.	Fabric	0.20:1	86.0	Auto	Auto	Man.	0.9
	12.0 x 0.45	26.2:1	Cu	Final Clnr.	Fabric	0.20:1	86.0	Auto	Auto	Man.	0.9
Inco Limited, Copper Cliff	11.1 x 1.1	10.4:2	Cu	Final Clnr.	Perf. Rubber	0.14:1	50.0	ND	ND	ND	1.2
	11.1 x 1.8	6.1:1	Cu	Final Clnr.	Perf. Rubber	ND	ND	ND	ND	ND	1.2
Kennecott Mining Co., Chino	11.4 x 0.4	28.5:1	Mo	Final Clnr.	Fabric	0.095:1	85.0	Auto	Man.	Man.	1.8
	11.6 x 0.6	19.0:1	Mo	Inter. Clnr.	Fabric	0.01:1	34.0	Auto	Man.	Man.	1.8
	11.5 x 0.9	12.7:1	Mo	Scavenger	Fabric	0.07:1	75.0	Auto	Man.	Man.	1.8
Magma Copper Co., Pinto Valley	7.6 x 0.6	12.5:1	Mo	Inter. Clnr.	Fabric	0.20:1	70.0	Auto	Man.	Man.	0.3
	7.6 x 0.6	12.5:1	Mo	Scavenger	Fabric	0.20:1	70.0	Auto	Man.	Man.	0.15
Magma Copper Co., San Manuel	1.8 x 12.1	8.0:1	Cu/Mo	Final Clnr.	Fabric	0.22:1	96.0	Auto	Man.	Man.	0.8
	1.5 x 12.1	6.6:1	Cu/Mo	Final Clnr.	Fabric	0.22:1	96.0	Auto	Man.	Man.	0.8
	9.7 x 1.2 x 1.2	8.0:1	Mo	Inter. Clnr.	Fabric	0.19:1	50.0	Auto	Man.	Man.	0.9
	3.6 x 0.8	4.8:1	Mo	Final Clnr.	Fabric	ND	40.0	Auto	Man.	Man.	0.9
Noranda Mines, Gaspe	13.6 x 0.45	30:1	Mo	Final Clnr.	Perf. Rubber	0.30:1	72.09	Auto	Auto	Auto	ND
	13.6 x 0.9	15:1	Mo	Inter. Clnr.	Perf. Rubber	0.15:1	82.97	Auto	Auto	Auto	ND
	13.6 x 1.8	7.5:1	Mo	Inter. Clnr.	Perf. Rubber	0.08:1	35.79	Auto	Auto.	Auto	ND
Southern Peru Copper, Cujone	12.1 x 1.2	10.0:1	Cu/Mo	Final Clnr.	Fabric	ND	ND	ND	ND	ND	ND
US Bureau of Mines, Salt Lake	5.5 x 0.06	86.4:1	Fluorite	Rougher	USBM	0.32:1	95.8	Auto	Man.	Auto	0.9
	4.2 x 0.06	67.2:1	Fluorite	Final Clnr.	USBM	0.38:1	99.0	Auto	Man.	Auto	1.8
	4.2 x 0.06	67.2:1	Mica	Rougher	USBM	0.47:1	90.0	Auto	Man.	Auto	0.9

ND = No Data

Table 2.6

Survey by Yianatos (1987)

Column Dimensions (m)	Shape	Application	Gas Sparger	Operation
0,9 x 15,0 2,0 x 15,1	Circ. Circ.	Mo cleaner Cu/Mo clnr	Fil. cloth Fil. cloth	Cyprus Minerals Sierrite
0,9 x 12,0 0,45x 12,0	Circ. Circ.	Cu cleaner Cu cleaner	Fil. cloth Fil. cloth	Disputada Exxon Chile
0,9 x 13,6	Square	Mo cleaner	perforated rubber	Codelco Chuquic-amata
1,1 x 11,0 1,8 x 11,0	Circ. Circ.	Cu cleaner Cu cleaner	perforated rubber	INCO Copper Cliff
0,6 x 11,0 0,9 x 12,8	Circ. Circ.	Mo cleaner Mo cleaner	Fil. cloth Fil. cloth	Lornex Mining Co.
0,6 x 7,6 0,6 x 7,6	Circ. Circ.	Mo cleaner Mo scav.	Fil. cloth Fil. cloth	Magma Copper Co Pinto Vall.
1,8 x 12,1 1,5 x 12,1 1,2 x 1,2 x 9,7 0,8 x 3,6	Circ. Circ. Square Circ.	Cu/Mo clnr. Cu/Mo clnr. Mo cleaner Mo cleaner	Fil. cloth Fil. cloth Fil. cloth Fil. cloth	Magma Copper Co. San Manuel
0,45 x13,6 0,9 x 13,6	Square Square	Mo cleaner Mo cleaner	perforated rubber perforated rubber	Noranda Gaspe Mines
0,4 x 11,4 0,6 x 11,6 0,9 x 11,5	Circ. Circ. Circ.	Mo cleaner Mo cleaner Mo scav.	Fil. cloth Fil. cloth Fil. cloth	Kennecott Min. Co. Chino
2,1 x 13,0	Circ.	Cu cleaner	Fil. cloth	Gibraltar Mines
1,2 x 12,1	Circ.	Cu/Mo cleaner	Fil. cloth	Sthn Peru Copper Cuajone

particle applications (Yianatos 1989). The maximum carrying capacity can be determined from pilot plant experiments (Espinosa-Gomez et al 1988) as mass of solids per unit column cross-sectional area per unit time. Alternatively, C_a can be estimated from the following semi-theoretical relationship (Espinosa-Gomez et al 1988A).

$$C_a = \frac{60\ K_1\ \pi d_p \rho_p J_g}{d_b} \qquad \text{Equation 2.5.1}$$

where K_1 is a combined efficiency factor to account for unknowns such as a bubble coverage and solids drop-back from the froth. d_p is a measure of particle size and ρ_p is the particle density. J_g is the superficial gas velocity and d_b is the average bubble diameter.

A simpler model for carrying capacity has also been developed by Espinosa-Gomez et al (1988) based on Equation 2.5.1.

$$C_a = 0{,}0682\ (d_{80}\rho_p) \qquad \text{Equation 2.5.2}$$

Equation 2.5.2 is a "best-fit" equation for eleven column flotation applications of both laboratory and industrial scale. The d_{80} is defined as the aperture size through which 80 per cent of the particles would pass. The d_{80} was used as an indication of paricle size distribution because it "is relatively easy to obtain compared to, say, the d_{20}" (Espinosa-Gomez et al 1988).

The finer particles consume more bubble surface area per gram than

coarser particles. If the distribution is dominated by fines, the carrying capacity will be limited by the fine rather than the coarse part of the size distribution, and the d_{20}, although perhaps more difficult to measure, would probably be a better measure of par'' size for carrying capacity than the d_{80}. If the distribution is dominated by coarse particles the d_{80} would probably be the better measure However, neither are able to describe a whole particle size distribution using one point.

In any case, in flotation the critical size parameter is really the area occupied by the particle on the bubble surface, because once the bubble is covered it is fully loaded. Therefore the cross-sectional area of a particle, which determines bubble loading, should be used to calculate mean diameter (Moys and Finch 1988).

2.5.2 Derivation of a Suitable Mean Particle Size

Figure 2.11 represents a typical plot of mass distribution $p_c(l)$ for particle size (diameter l).

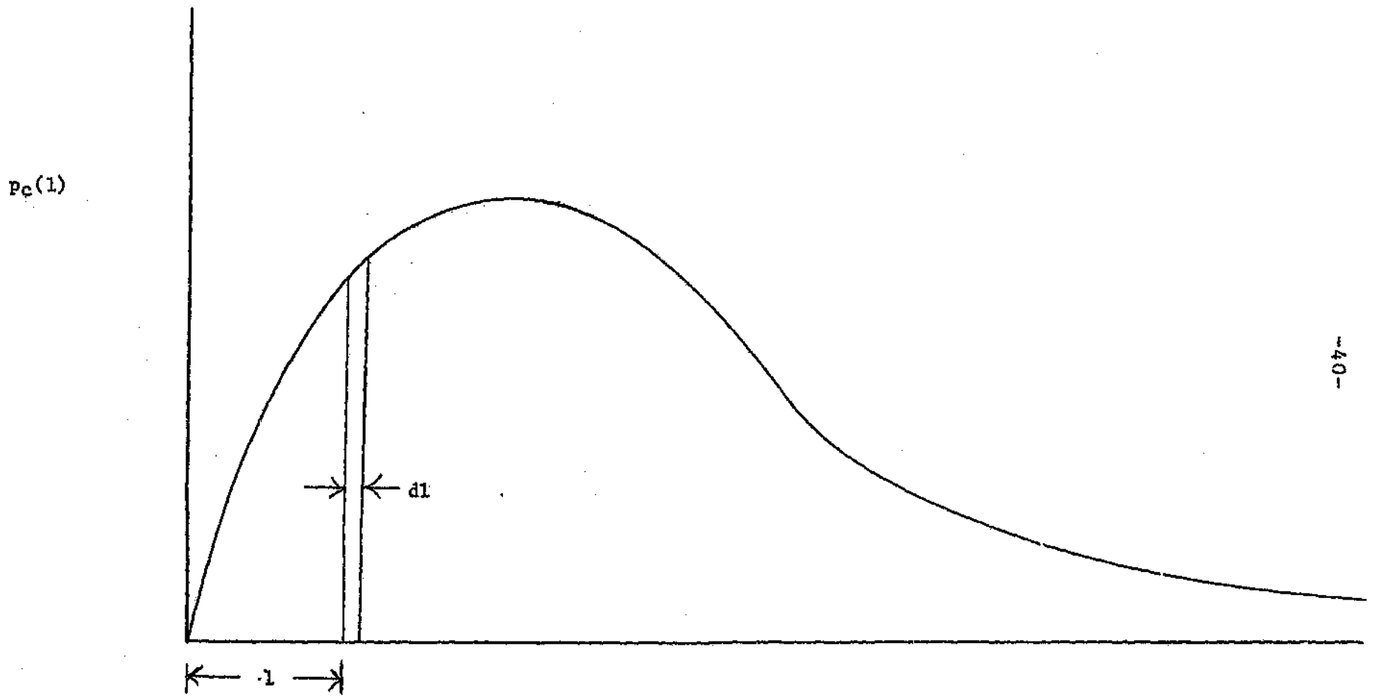
Spherical particles are assumed, and the particle size parameter used in this study is the particle diameter d_p that has the mean cross-sectional area of the particles in the sample.

The total cross-sectional area of a group or sample of particles is equal to the total number of particles multiplied by the mean particle cross-sectional area of that sample.

$$\text{The mean cross-sectional (CS) area} = \frac{\pi d_p^2}{4}.$$

Figure 2.11

PARTICLE SIZE DISTRIBUTION BY MASS



$$\text{The total CS area} = \frac{N \pi d_p^2}{4}$$

where N = the total number of particles

$$N = \frac{\text{Total Mass}}{\text{volume of mean particle} \cdot \text{particle density}}$$

$$\text{As the total mass} = \int_0^\infty p_c(l)\,dl = 1.$$

$$N = \frac{1}{\frac{\pi d_p^3 \rho_p}{6}}$$

ρ_p is the mean density of the particles and is assumed to be a constant.

The total CS area of the sample

= (CS area of the mean particle) . N

$$= \frac{\pi d_p^2}{4} \cdot \frac{1}{\frac{\pi d_p^3 \rho_p}{6}}$$

$$= \frac{6}{4 \rho_p d_p} \qquad \text{Equation 2.5.3}$$

The CS area of the particles of diameter l to l+dl (whose distribution is described by a number distribution n(l))

= (number of particles in the interval)

. (the CS area of a sphere of diameter l)

$$= \frac{n(l)\,dl\,\pi l^2}{4}$$

$$= \frac{p_c(l)\,dl}{\frac{\rho_p \pi l^3}{6}} \cdot \frac{\pi l^2}{4}$$

The cross-section of the whole sample from l = 0 to l = ∞

$$= \int_0^\infty \frac{\pi l^2 \, p_c(l)\,dl}{4 \rho_p \pi l^3/6}$$

$$= \frac{6}{4\rho_p}\int_0^{\infty}\frac{p_c(l)\,dl}{l} \qquad \text{Equation 2.5.4}$$

Equating Equation 2.5.3 and Equation 2.5.4

$$\frac{6}{4\ \rho_p\ d_p} = \frac{6}{4\ \rho_p}\int_0^{\infty}\frac{p_c(l)\,dl}{l}$$

$$d_p = \left(\int_0^{\infty}\frac{p_c(l)\,dl}{l}\right)^{-1} \qquad \text{Equation 2.5.5}$$

The formula is equivalent to the $d_{(3,2)}$ or the Sauter mean diameter, as shown in Appendix I.

2.6 FLOTATION KINETICS IN COLUMN FLOTATION

2.6.1 Introduction

In attempting to find a suitable model for the kinetics of the flotation column at ERGO, a search was made of the literature for an adequate mathematical model to describe the flotation kinetics and estimate important parameters.

From the literature it was clear that first order kinetics adequately describe the recovery of a single flotable species (Harris and

Chakravarti 1970). Two models specifically used for column flotation have been proposed by Dobby and Finch (1986), and Contini et al (1988). Dobby and Finch (1986) use the Kelsall (1961) model which assumes two flotable species; one fast floating and one slow floating.

$$r = (1 - x)(1 - \exp(-k_f t)) + x(1 - \exp(-k_s t))$$

This is a three parameter model and x is the fraction of valuable mineral which is slow floating, k_s is the rate constant for that fraction, and k_f is the rate constant for the fast floating fraction.

Contini et al (1988) propose using the first order rate equation.

$$r = 1 - \exp k_n t$$

for each narrowly spaced size fraction and individual species.

Dowling et al (1986) and (1986A) tested 13 different kinetic models on industrial conventional circuit and batch cell data, and in their list of preferences the Kelsall model fared badly. They also found that "flotation models with more than two descriptive parameters... cannot be statistically justified since these assumptions introduce model error measureable by a poorer fit and for wider confidence limits". The best model they found was that which assumed a uniform distribution of component flotabilities and a first order rate equation:

$$r = R\left(1 - \frac{1}{k_u t}(1 - \exp(-k_u t))\right) \qquad \text{Equation 2.6.1}$$

Seven different ores were used to generate the batch cell data for different reagent additions and different flotation stages e.g. roughing and cleaning (Dowling et al 1986A).

Although the 13 models were applied to batch cell data, their performance is relevant to the choice of a column flotation model because the reaction rate in a batch reactor is described by the same models as that in a plug flow reactor (Levenspiel 1972).

The two industrial rougher flotation circuits modelled (Dowling et al 1986) were Nchanga Consolidated and Tennessee Copper Company. The roughing circuits for Nchanga and Tennessee Copper consisted of 20 and 24 conventional cells in series. The abilities of the various flotation rate models to predict the performance of conventional flotation circuits may be used as criteria for choosing a column flotation model, as it may be demonstrated that the performance of n back mixed reactors in series approaches that of a plug flow reactor for a simple first order reaction as n approaches infinity (Levenspiel 1972). Ten back mixed reactors perform like a plug flow reactor within experimental error, and therefore plug flow flotation columns may be modelled in the same way as the rougher flotation circuits at Nchanga and Tennessee Copper.

Further, there are a number of practical considerations in choosing a kinetic model for column flotation. Firstly, recovery versus time data is limited to the number of different feed rates that the column can operate at without the pulp lines blocking. This is particularly important for pilot plant work because the narrow pipe sizes are prone to blocking. Indeed, experience at ERGO was such that each flowrate setting (or residence time) required its own pipe and essentially three sets of residence times were achieved.

Secondly, due to the continual variation of the feed to the column, and its fineness, recovery data for individual size fractions are difficult to determine with any confidence.

It was decided to use the model preferred by Dowling et al (1986) and (1986A) because it has many advantages for column flotation.

a) It can be used to predict recovery after infinite flotation time. This parameter is useful for evaluating the potential of column flotation for improving the performance of a conventional flotation circuit.

b) It is a two parameter model and therfore can be statistically justified. A two parameter model requires only three data points.

c) It was found to yield a good fit on conventional circuit data (Dowling et al 1986) and it was the preferred model out of the thirteen tested.

The disadvantage of the model is that it cannot be easily developed to account for axial mixing in a column.

2.6.2 Derivation of Rate Equation

The model was derived for conventional batch flotation cells by Huber-Panu et al (1976). The derivation given below is for the flotation column.

The following assumptions were made:

a) the flow of slurry down the column is assumed to be plug flow.

b) Steady state conditions prevailed over the sampling period. There was no important variation in either the quantity and or quality of the feed pulp and there was no variation in the operation of the column.

c) The non-flotable valuable mineral particles of the ore are denoted by (1-R).

d) The feed of the pilot plant ERGO scavenger concentrate was considered to be monosized material. While not realistic, this assumption was made necessary by the fineness of the material. Ideally the material would have been screened into several size fractions and the flotation rate constant of each considered individually. However, the feed material was in the region of 87 percent minus 45 micrometres (see BR2 head size analyses - Appendix II), and cyclosizing was required for sizing. The assay laboratory requirement for accurate gold analysis is 100g. As only a small quantity of material may be cyclosized at one time, and the majority of that reports to the ultrafine fraction, many cyclosizer runs would be required for each sample. The cost of such subdivision was prohibitive.

Flotability

The degree of hydrophobicity of the valuable minerals is determined by the physical (e.g. liberation) and chemical (e.g. reagents) conditions in the column. The hydrodynamics of the column determined the collision probabilities and collision efficiencies between the bubbles and particles. These factors are reflected in the flotability (k) of the valuable minerals. The flotability can also be described as the rate of flotation.

The kinetics of flotation of flotable particles of flotability k can be accurately represented by a rate equation of the form

$$\frac{d\,C(k)}{d\,t} = -\,k\,N\,C(k)$$

where C(k) is the concentration of potentially flotable minerals in the pulp of flotability k, and N is the number of bubbles per unit

volume of pulp and t denotes time. Using the following terms the equation may be rewritten; X(k) as the fraction of potentially flotable valuable mineral of flotability k in the feed which remains unfloated i.e. in the pulp phase and C_0 as the feed value of C.

$$C_0 \frac{d\,X(k)}{d\,t} = - k\,N\,C_0\,X(k)$$

The C_0 terms cancel on both sides.

$$\frac{d\,X(k)}{d\,t} = - k\,N\,X(k) \qquad \text{Equation 2.6.2}$$

As the flotability of valuable mineral particles differ from one to another, the distribution of the flotabilities is characterized by a density function for particle flotability f(k).

Let the fraction of valuable flotable mineral in the feed which is not recovered be X_T. The fraction of valuable mineral recovered r, can be written as the fraction of the valuable mineral which is recoverable in the feed R, times the fraction of the recoverable mineral which actually is floated $(1-X_T)$.

$$\text{then } r = R(1-X_T) \qquad \text{Equation 2.6.3}$$

The fraction of the flotable valuable mineral particles of flotability between k and k + dk in the feed is f(k) dk. The fraction of material of flotability k in the column feed which is lost to the column tails is X(k) by definition of X(k) i.e. the fraction of

valuable flotable mineral of flotability k in the feed which remains in the pulp.

$$X_T = \int_0^{k_m} X(k)\ f(k)\ dk$$

where k_m is the maximum flotability of all the particles in the feed.

Substituting Equation 2.6.3

$$r = R\left(1 - \int_0^{k_m} X(k)\ f(k)\ dk\right) \qquad \text{Equation 2.6.4}$$

Flotable valuable mineral balance

Balance over time t to $t+\delta t$ of the flotable valuable mineral of flotability k in the pulp phase as shown in Figure 2.12.

$$IN = F\ C_0\ X(k)$$

where F is the volumetric flowrate down the column.

$$OUT = F\ C_0\ (X(k) - \delta X(k))$$

Rate of removal of valuable mineral from from the pulp phase

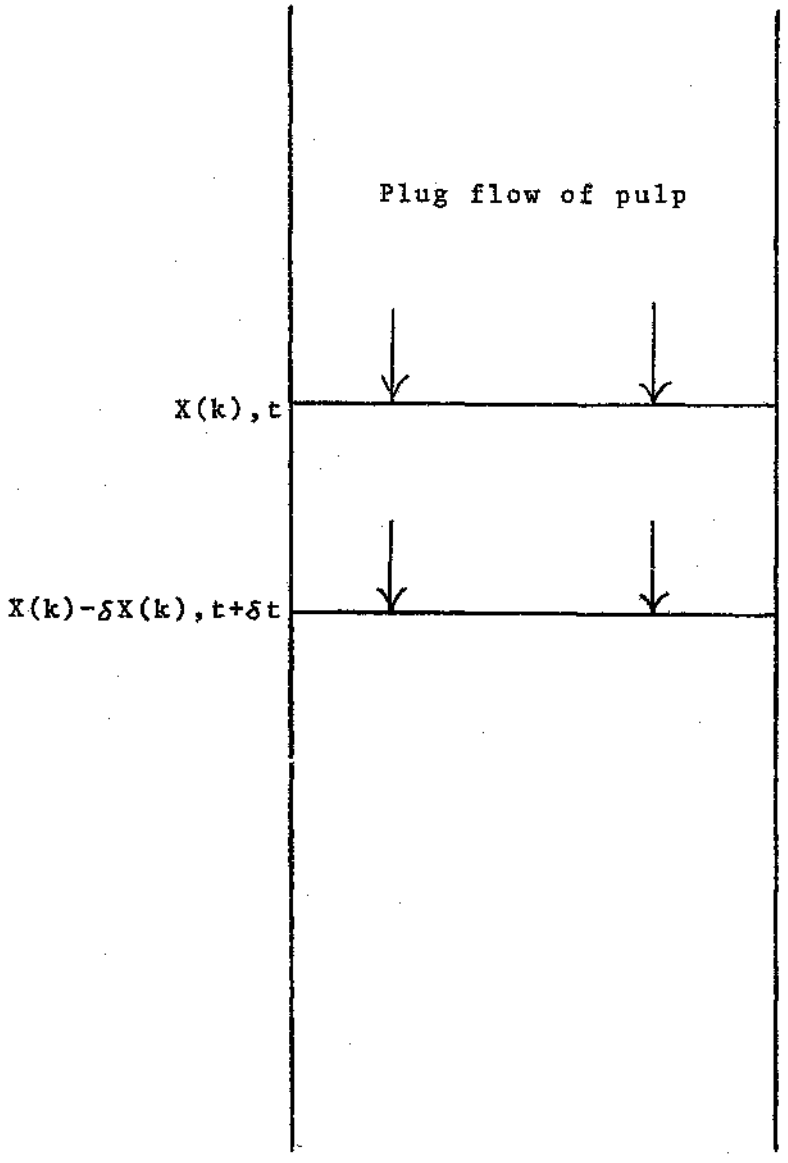
$$= F\ C_0\ \frac{d\ X(k)}{d\ t}\ \delta t$$

Substituting Equation 2.6.2

$$= -F\ C_0\ k\ N\ X(k)\ \delta t.$$

Figure 2.12

MASS BALANCE OF VALUABLE MINERAL IN THE PULP PHASE



Using a valuable mineral balance i.e.

IN - OUT = ACCUMULATION

the constants F, C_0, cancel.

Therefore

$$X(k) - (X(k) - \delta X(k)) = - k N X(k) \delta t$$

$$\Rightarrow \frac{\delta X(k)}{\delta t} = - k N X(k)$$

taking the limit as $\delta t \to 0$

$$\frac{d X(k)}{dt} = - k N X(k)$$

$$\Rightarrow \frac{d X(k)}{X(k)} = - k N dt$$

Integrating over time for a constant flotability k

$$\ln X(k) = - k N t$$

$$\Rightarrow X(k) = \exp (- k N t) \qquad \text{Equation 2.6.5}$$

Substituting equation 2.6.5 in Equation 2.6.4

$$r = R\left(1-\int_0^{k_m} \exp\ (-\ k\ N\ t)\ f(k)\ dk\right) \qquad \text{Equation 2.6.6}$$

Case 1

Assuming a uniform flotability (see Figure 2.13)

$$\text{i.e.} \quad f(k) = \frac{1}{k_m} \qquad 0 \leq k \leq k_m$$

$$r = R\left(1-\int_0^{k_m} \exp\left(\frac{-\ k\ N\ t}{k_m}\right)\ dk\right)$$

$$\Rightarrow \quad r = R\left[1 - \frac{-\exp(-\ k\ N\ t)}{k_m\ N\ t}\right]_0^{k_m}$$

$$\Rightarrow \quad r = R\left[1- \frac{(1-\exp(-k_m\ N\ t))}{k_m\ N\ t}\right]$$

Substituting equation 2.6.5 in Equation 2.6.4

$$r = R\left(1-\int_0^{k_m} \exp\ (-\ k\ N\ t)\ f(k)\ dk\right) \qquad \text{Equation 2.6.6}$$

Case 1

Assuming a uniform flotability (see Figure 2.13)

$$\text{i.e.} \quad f(k) = \frac{1}{k_m} \qquad 0 \leq k \leq k_m$$

$$r = R\left(1-\int_0^{k_m} \exp\frac{(-\ k\ N\ t)}{k_m}\ dk\right)$$

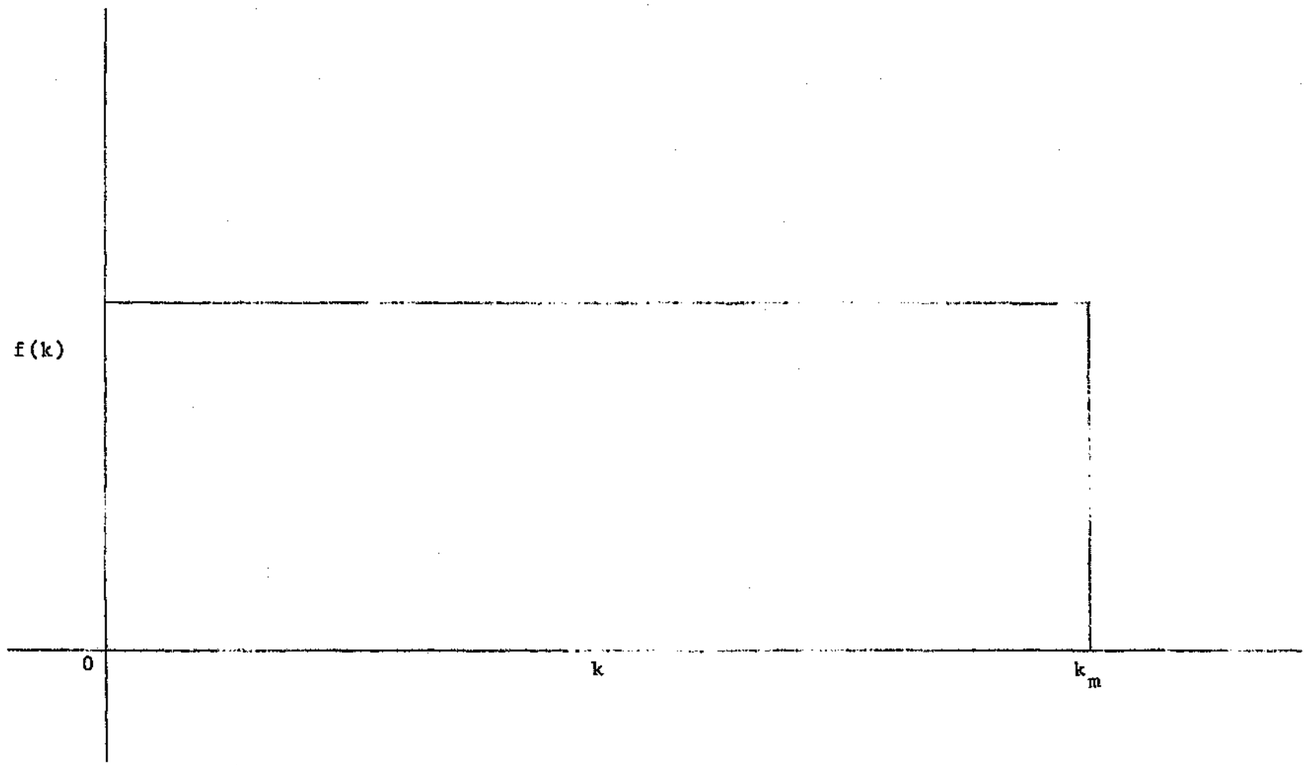
$$\Rightarrow \quad r = R\left[1 - \frac{-\exp(-\ k\ N\ t)}{k_m\ N\ t}\right]_0^{k_m}$$

$$\Rightarrow \quad r = R\left[1- \frac{(1-\exp(-k_m\ N\ t))}{k_m\ N\ t}\right]$$

Figure 2.13

FREQUENCY PLOT OF FLOTABILITIES

for a Sample of Particles of Uniform Flotability



let $k_m N = k_u$

then $$r = R\left[1 - \frac{1}{k_u t}(1-\exp(-k_u t))\right]$$ Equation 2.6.7

Case 2

Assume a single flotability for all the valuable mineral particles

Let $k = k_a$ and therefore $f(k)dk = 1$.

$\therefore r = R(1 - \exp(-k_a N t))$

let $k_a N = k_s$

then

$$r = R(1 - \exp(-k_s t))$$ Equation 2.6.8

2.7 AXIAL MIXING IN THE ERGO COLUMN

2.7.1 Calculation of vessel dispersion number

It was important to determine whether the ERGO pilot plant deviated from plug flow significantly. LaPlante et al (1988) present an empirical equation 2.6.1 to estimate the dispersion coefficient.

$$D = 0{,}000298 . d^{1{,}31} . J_g^{0{,}33} . \exp(-0{,}025\ S).$$

where D is the dispersion coefficient in m/sec, d is the diameter of the column in cm, J_g is the superficial gas rate in cm/sec, S is the mass percent solids.

Typical values for the ERGO pilot plant were:

$d = 23{,}8$ cm

$J_g = 1{,}2$ cm/sec

$S = 15$ per cent

These values yielded an estimate of D to be about 0,014 m^2/sec.

The vessel dispersion number N_d was

$$= \frac{D}{U_i L}$$

where U_i is the interstitial pulp velocity and L is the collection zone length. Typical values for the ERGO pilot plant were:

$U_i = 0{,}006$ m/sec

$L = 8$ m,

These values yielded an estimate of $N_d = 0,3$.

The dispersion number (0,3) indicated that a significant amount of mixing did take place. The mixing in the column affected the flotation rate constant but not the infinite time recovery values of equation 2.6.1.

2.7.2 Comparison of dispersion coefficients

La Plante et al (1988) cited a dispersion coefficient of 0,001 m^2/sec for a 5,1 cm diameter, 9,2 m high laboratory scale pilot column, 0,54 m^2/sec for a 0,516 m diameter commercial column. The dispersion coefficient for the ERGO pilot plant of 0,014 m^2/sec was between the two.

The relative variance of the residence time distribution in the recovery zone of the column was calculated to be 0,01 from a formula given by La Plante. Typical values for a bank of conventional cells vary between 0,1 and 0,3. The ERGO pilot column did not have plug flow, but had considerably less axial mixing than commercial flotation cells.

Although the use of the rate constant in scale-up was not explored, nor was scale-up of concern in this project, the infinite recovery values may be used for comparison purposes between conventional cell and column flotation tests.

CHAPTER 3

PILOT PLANT

3.1 HOST FLOTATION PLANT

The pilot plant was built at the East Rand Gold and Uranium Company Limited (ERGO). ERGO is situated near Springs on the eastern side of the Witwatersrand and treats approximately two million tons of mine dump material per month. The ERGO circuit is shown in Figure 3.1. Slime is reclaimed from several slime dams by monitoring with high pressure water, and is pumped to the flotation plant conditioning tanks. The different dam materials are kept separate in three streams: A, B and C. The pulp is diluted from about 50 to about 39 percent solids and the pH is reduced to 3,8. It is then pumped to a distribution box at the head of each stream, where the stream is split into two. The reagents are added just before the distribution box with a reagent conditioning time of about 45 seconds. The pulp residence time in the roughing circuit is about 17 minutes and 90 in the cleaning circuit.

The plant reagents added are:

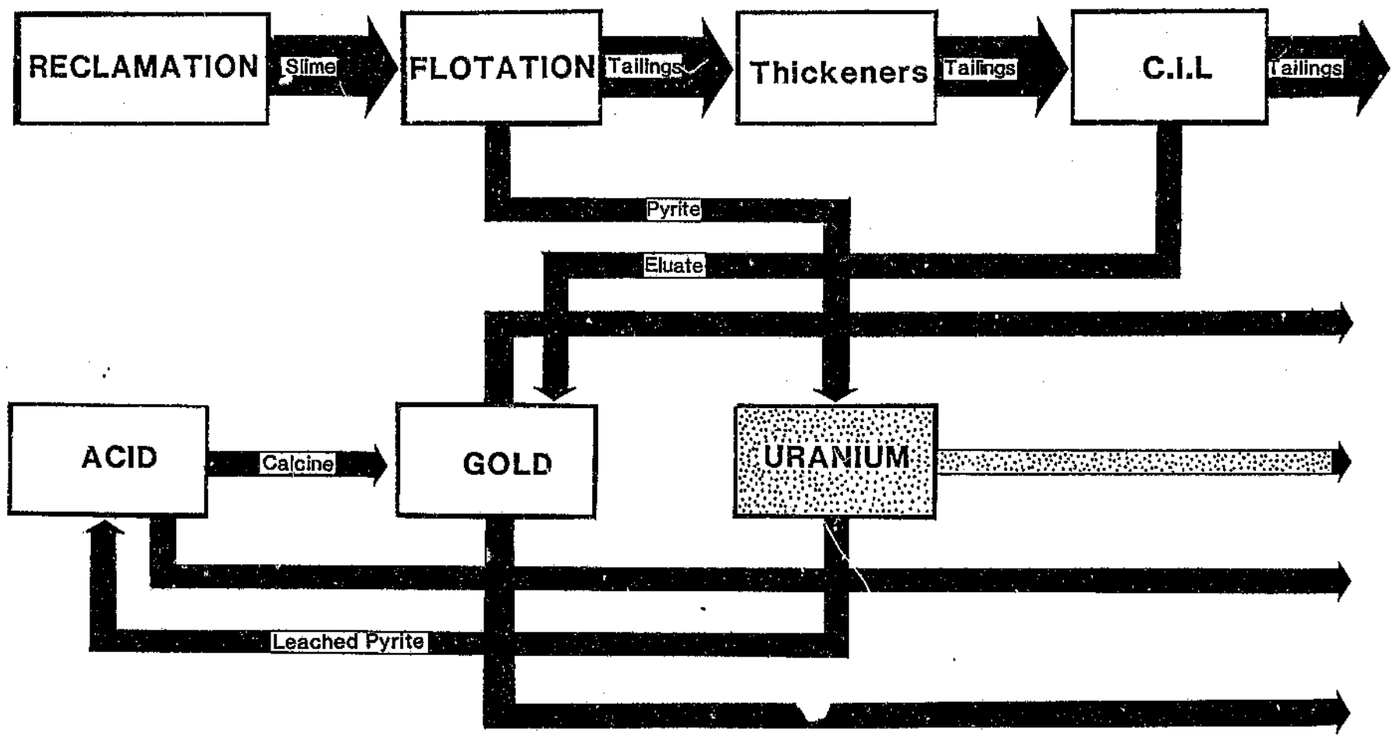
collectors: sodium mercapto benzo thiazole at 85 g/t
dithiophosphate at 12 g/t

frother: Polypropylene glycol methyl ethers (Dow 200) at 17 g/t.

The flotation plant tailings are treated in the Carbon-in-Leach (CIL) plant to recover any cyanidable gold.

Figure 3.1

ERGO FLOWSHEET



3.2 BASIS OF PILOT PLANT DESIGN

It was decided to build the pilot plant on an existing commercial plant so that the column's sensitivity to normal changes in operating conditions could be assessed. The feed to a flotation plant is continually changing due to the inhomogenous nature of ore bodies. Feed grade can change from hour to hour, and operating parameters also fluctuate, as no control system is perfect. Parameters such as percent solids, pH and reagent addition are not constant and may change rapidly. The pH control at ERGO was particularly poor and values between 1,5 and 5 were measured.

The pilot plant was designed to be large enough to experience normal industrial scale problems, and was an order of magnitude larger than the existing column flotation pilot plant at the University of the Witwatersrand.

The ERGO flotation column was designed around existing equipment and the largest slurry pump available was a Bredel peristaltic pump capable of pumping 9 600 ℓ/hr. Assuming a maximum superficial feed pulp velocity of 5 cm/sec, the column diameter was calculated to be in the region of 260 millimetres. The nearest size of pipe available was used, with an inside diameter of 250 millimetres. It was mild steel with a 6 millimetre rubber lining for corrosion protection, which reduced the inside diameter to 238 millimetres.

3.2.1 General Dimensions and Capacity

The column height was ten metres, similar to commercial operations around the world (Yianatos 1987). The other important parameters were designed to give maximum flexibility. Three feed entry ports were fitted; 900, 1300 and 1700 millimetres from the top. The 1700

millimetre port was found to be satisfactory and was used throughout the testwork. In general the instrumentation (e.g. rotameters) was designed to allow for an order of magnitude variation on set point. The designed flowrate ranges were specified by the project supervisor and are summarised below.

Feed rate

Superficial velocity J_f: 0,5 to 5 cm/sec
Volumetric flowrate Q_f: 13 to 130 l/min

Gas rate (STP)

Superficial velocity J_g: 5 cm/sec
Volumetric flowrate Q_g: 130 l/min

Wash water

Superficial velocity J_w: 0,1 to 0,5 cm/sec
Volumetric Flowrate Q_w: 3 to 13 l/min

Summary of Specifications

Column height (H):	10 metres
Column Inside Diameter (d_c):	238 mm
Feed Entry Point:	1700 mm from the top
Gas Injection Point:	800 mm from the bottom
Height for Collection Zone Hold Up Measurement:	7470 mm

3.2.2 Feed and Tailing Pipe Diameters

The minimum pipe velocity to keep slurries in suspension is 1,5 m/sec. The column was designed to operate with superficial velocities varying from 0,5 to 5 cm/sec. Such a wide variation in feed/tailings flowrates required different overall pipe diameters. Three hose diameters were chosen so that over the wide range of slurry feed rates the slurry velocity was kept in the range 1,5 to 3,5 m/sec. The hoses chosen had diameters of 30, 20 and 14 millimetres. (The closest available hose size to 14 millimetres was 12,5 millimetres).

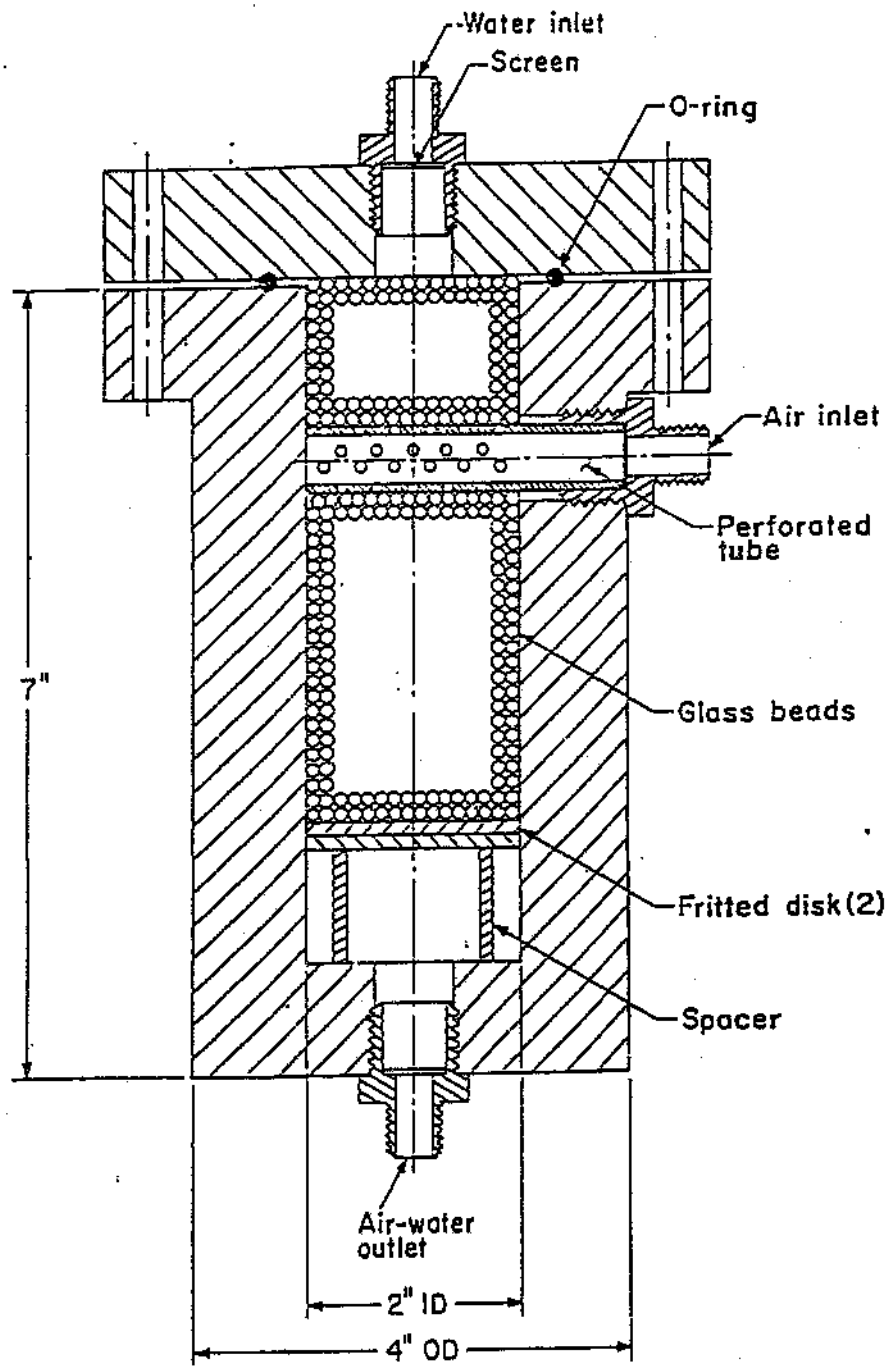
3.3 BUBBLE GENERATION SYSTEM

The bubble generator is the heart of the column, because small, uniform bubbles are essential for the column to perform satisfactorily.

Fine bubbles are required because the open area of the column is approximately a fifth of that of a conventional cell. Nonetheless, the bubble surface area in the column must be similar to that found in the conventional cell at the concentrate overflow lip if a similar recovery is to be achieved. This extra surface area is achieved through the use of finer bubbles. The bubbles must be uniform in size to avoid large bubbles travelling up the column at a high velocity and coalescing with the finer bubbles.

The USBM bubble generator described by McKay et al (1988) was fitted to the pilot plant. It uses an external contactor to intimately mix the air and frother phases, as shown in Figure 3.2. The size of contactor indicated in Figure 3.2 was used successfully (McKay et al 1988) to generate bubbles for columns of diameters between 64 and 300 millimeters, so it was decided to use the same size of contactor for the pilot plant. The two-phase mixture from the contactor is fed to the injection tube inside the column.

Figure 3.2
CROSS SECTION OF THE USBM
BUBBLE GENERATOR



3.3.1 Injector Design

McKay (1988) recommends 10 to 12 injection holes per square foot of column. From this recommendation the number of holes required for a column of 238 millimetres in diameter is five injection holes.However, the holes are positioned in pairs along the injection tube, so the number of holes used was six.

The cross-section of the column was divided into three segments of equal areas: A, B and C (Figure 3.3). The dimensions of the circular segments were calculated using Table 1-19a from Perry (1984). Pairs of injection holes were drilled in each segment.

3.4 WASH WATER DISTRIBUTOR

The design of the wash water distributor is as important as the wash water itself. The wash water must be distributed equally over the cross-section of the froth to stabilise the froth evenly; and the jet from the distributor must not mechanically destroy the froth. Preferably, the distributor should be able to be operated at various levels above and below the froth surface.

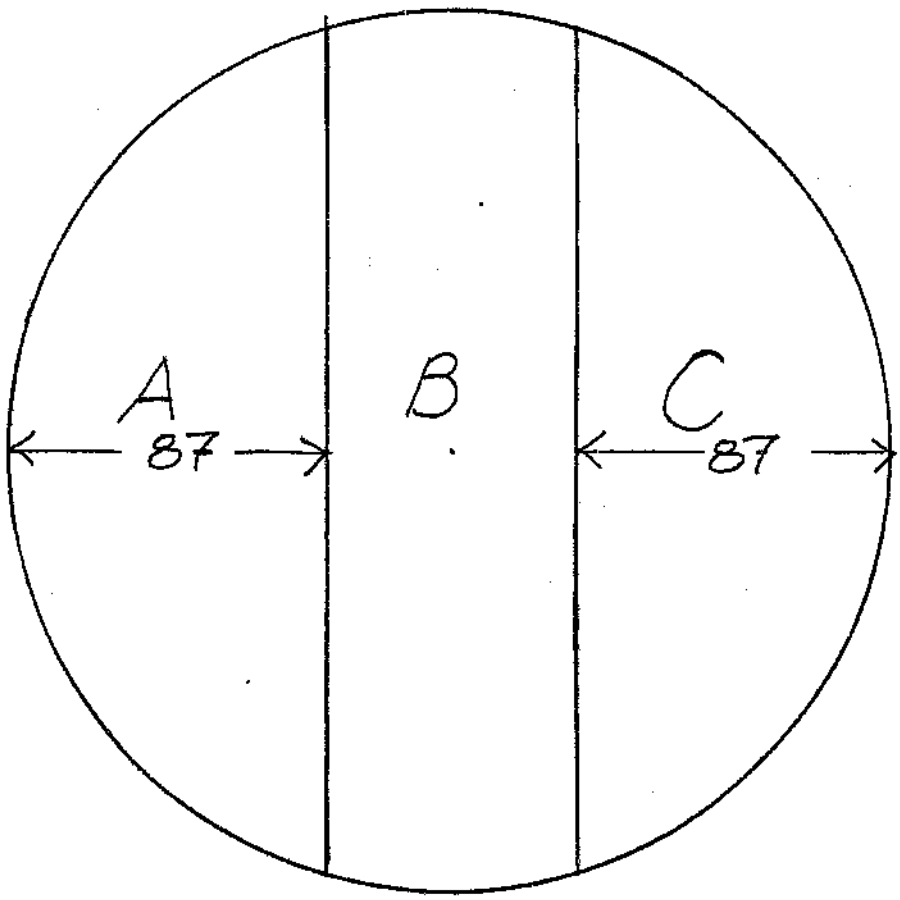
Two designs of wash water distributor design were attempted, the bucket distributor and the concentric coil design. The bucket design was found not to be suitable on the ERGO column, but may be valid in other applications.

3.4.1 Bucket distributor design

In order to distribute the wash water evenly, it was decided to let it drip onto the top of the froth from many points just above the froth. The more holes, the more evenly the wash water could be distributed. A flat-bottomed plastic bucket was mounted just above the surface of the froth, with pin holes punched in a regular pattern in the bottom.

Figure 3.3

COLUMN SEGMENTS OF EQUAL AREAS



Holes were punched rather than drilled to form a protruding lip around each hole on the bottom of the bucket to facilitate droplet formation at each hole and to prevent the drops from running together.

It was decided to use a 0,75 millimetre hole to provide a gentle flow of water without the hole blocking. The flowrate through a 0,5 millimetre hole was determined experimentally for 10 to 200 millimetres head of water (see Figure 3.4). The results were adjusted for a 0,75 millimetre hole, and it was calculated that 413 holes were required for the maximum wash water flowrate with a head of 160 millimetres.

The wash water used was Rand Water Board water. An air agitated steady head tank was fitted about 2 metres above the top of the wash water bucket, with a ballcock valve for level control. A pH controller was fitted to the tank to maintain a pH of 3,8 in the wash water, the same as the feed pulp. A frother dosing pump added a known amount of frother to the tank so that frother washed from the froth would be replaced to maintain the stability of the froth phase. The wash water was fed from the steady head tank through a rotameter to the top of the wash water distributor.

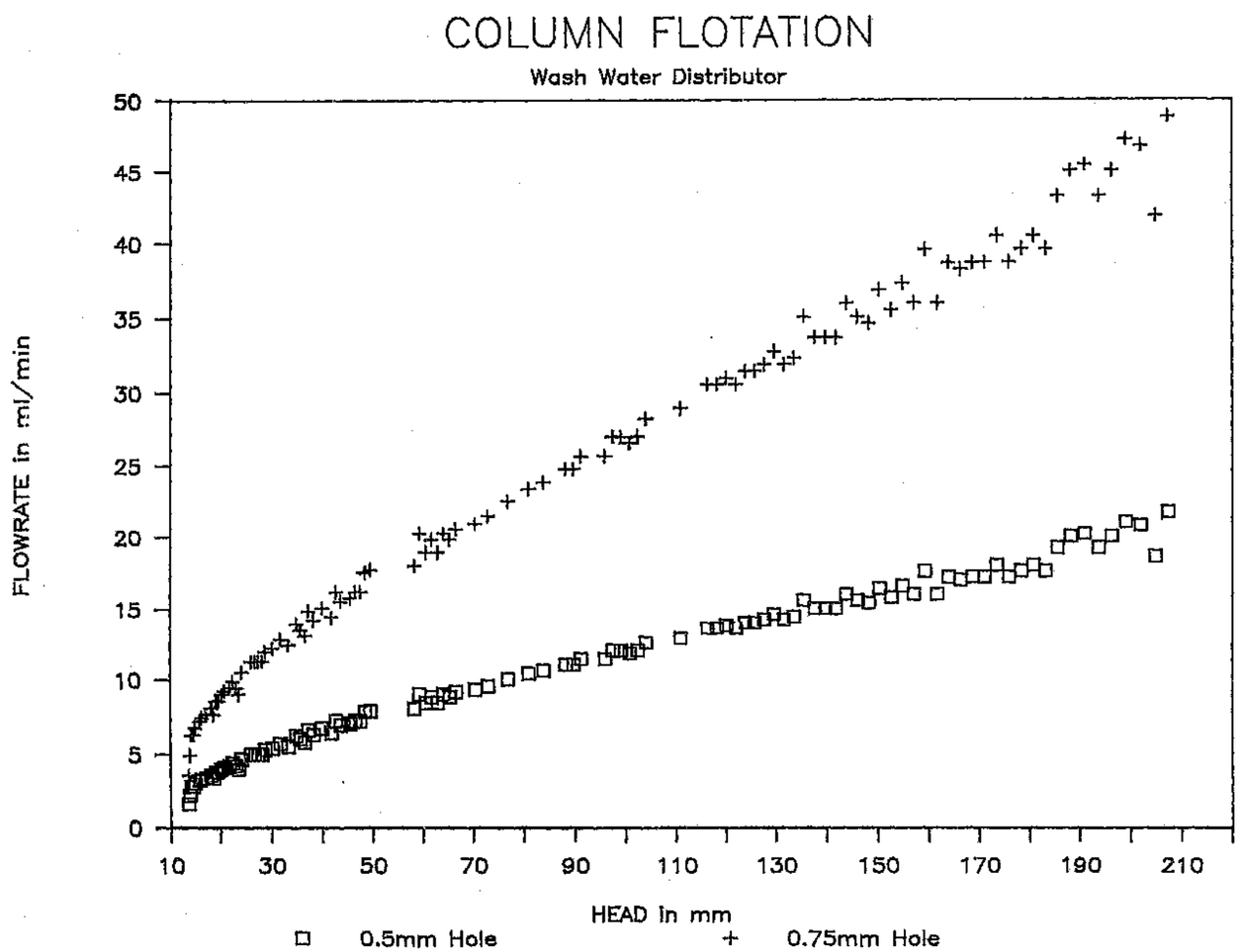
During commissioning, however, it became apparent that this arrangement was unsuitable. Firstly, it became clear that wash water addition above the froth did not stabilise the froth sufficiently and needed to be added below the froth surface. Secondly, pin holes in the plastic closed over so the wash water failed to reach its design capacity.

3.4.2 Concentric Coil Design

It was decided to replace the wash water bucket distributor with a coil-type distributor.

Stainless steel tubing was bent into three concentric coils, which were joined together radially by another length of tubing. The

Figure 3.4


COLUMN FLOTATION
Wash Water Distributor
FLOWRATE in ml/min
50
45
40
35
30
25
20
15
10
5
0
10
30
50
70
90
110
130
150
170
190
210
HEAD in mm
0.5mm Hole
0.75mm Hole

tubing in turn was fed by a vertical section of pipe connected to the wash water rotameter. Pairs of holes were drilled downwards at 45° in the tubing as exits for the wash water.

Because the coils were fed at only one point in their circumference, the distribution was found to be uneven, with most water exiting near the addition points. The coil was inclined at a slight angle which improved distribution. However, this was not really satisfactory, and it would have been better to have had four addition points.

Because it was decided to use three concentric coils to distribute the wash water, the cross-section of the column was divided into three equal areas: A, B and C. A concentric ring served each area with wash water (see Figure 3.5).

Area A = Area B = Area C

Assuming the diameter of the column to be 238 millimetres,

$a = 98$ mm
$b = 69$ mm

To find the radius of the coil r_a serving area A, the area A was divided into two equal areas (see Figure 3.5) and r_a was calculated to be 109 millimetres.

Similarly, for Area B and Area C:

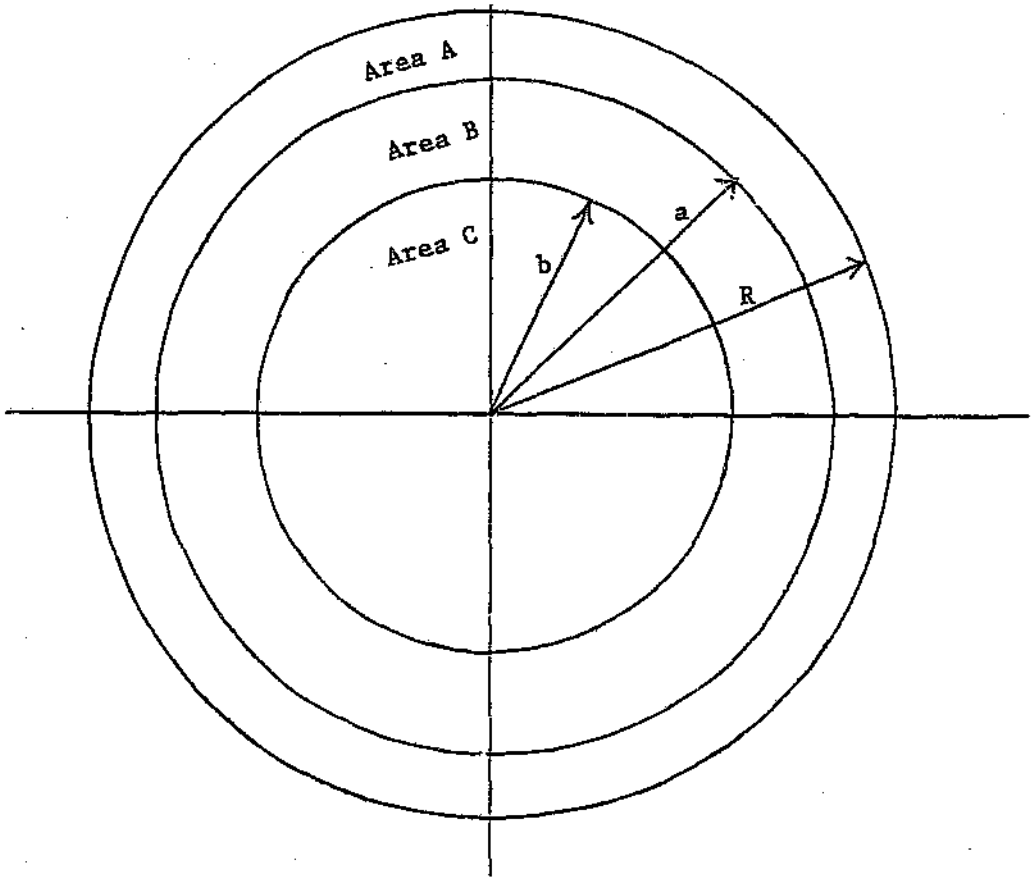
$r_b = 85$ mm

$r_c = 49$ mm

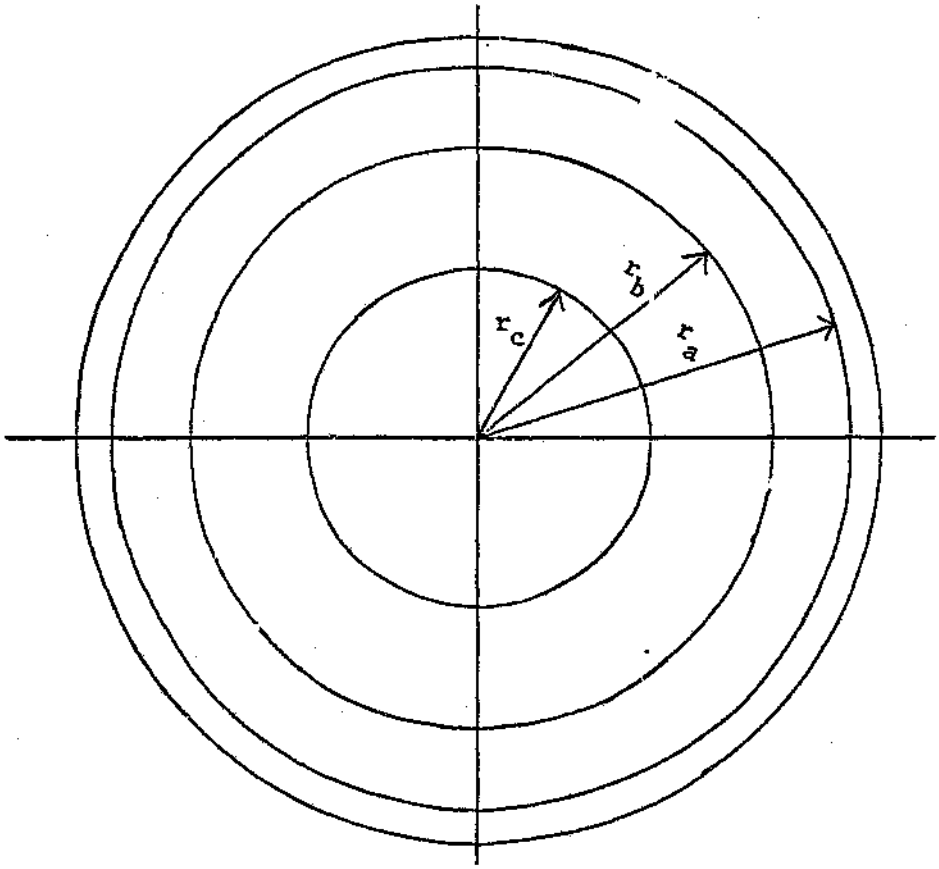
Figure 3.5

WASH WATER DISTRIBUTOR DESIGN

RADII OF EQUAL AREAS



Radii of Concentric Coils



Hole spacing

As the hole diameter was 0,75 millimetres, the total number of holes required for good distribution was approximately 400. Since each ring served the same surface area, the number of holes per ring were equal.

The holes were drilled downwards at 45° in pairs, and spaced ten millimetres apart in coil A, eight millimetres apart in coil B and five millimetres apart in coil C.

3.5 DESCRIPTION OF THE PILOT PLANT

A schematic diagram of the pilot plant is shown in Figure 3.6. The column itself was constructed of three sections. The main section was a rubber-lined ten-inch mild steel pipe six metres in length. The rubber lining reduced the inside diameter from 250 millimetres to 238 millimetres. This section was fixed to the plant steelwork at its top and bottom.

The top section was made of clear PVC and was two metres in length. At the top of the column the wash water coil was fitted just inside the column about 20 millimetres below the overflow lip. An overflow launder was fitted at the top to direct the concentrate overflow into the sampler situated below. The top section is shown in the photograph in Figure 3.7. The differential pressure transducer was connected to the side of the column via a flexible plastic tube filled with water. A section of the tube was wound into a coil to prevent solids from the column getting into the differential pressure cell. The water line was flushed each day to remove any solids that had accumulated in the line.

The bottom two-metre section was fabricated from two millimetre stainless steel plate. This facilitated the welding on of "after-thoughts", which would not be possible with rubber-lined mild steel. The USBM bubble generator injector was fitted to this section 700 millimetres from the bottom of the column.

Figure 3.6

SCHEMATIC DIAGRAM OF ERGO PILOT PLANT

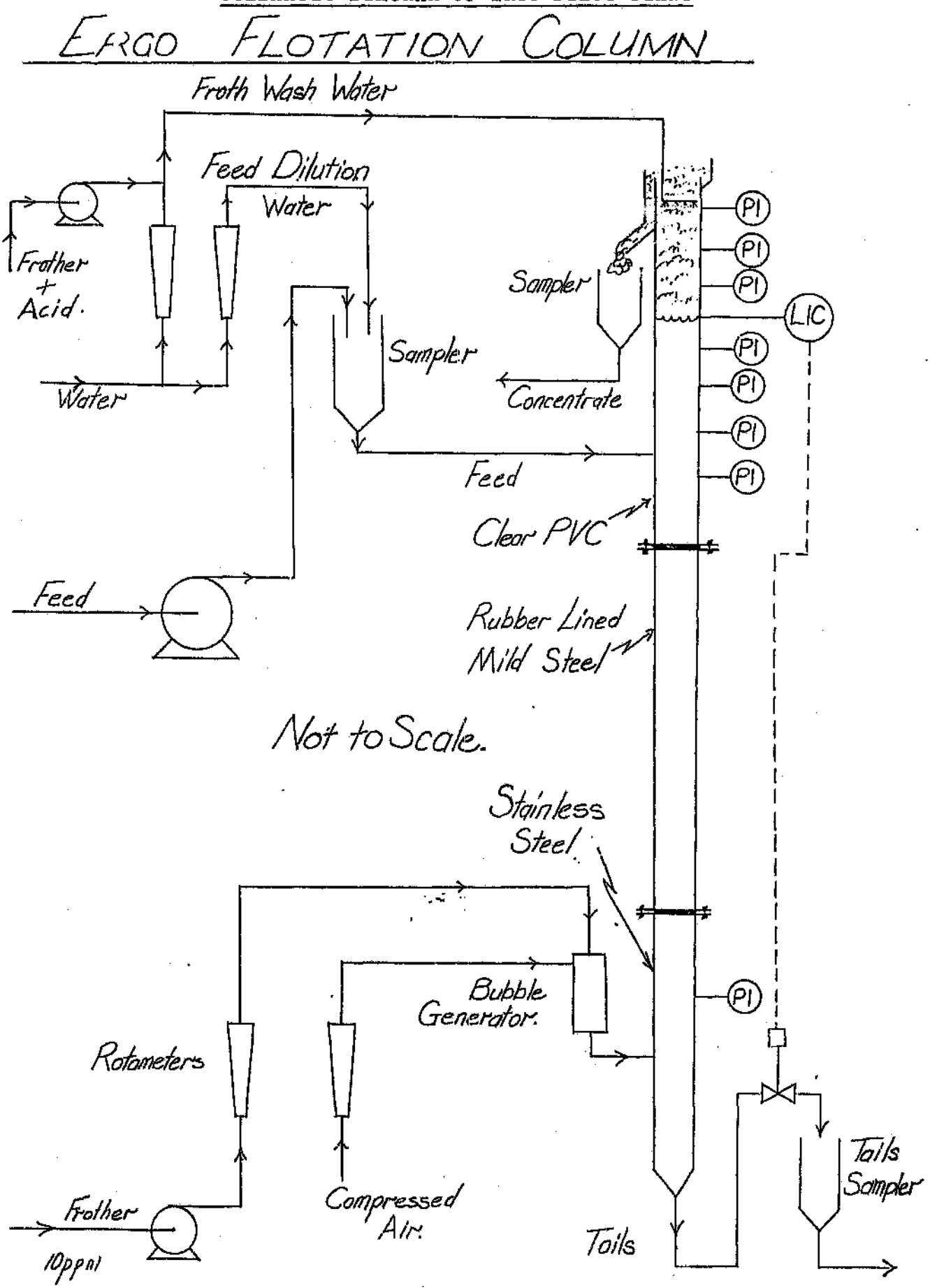
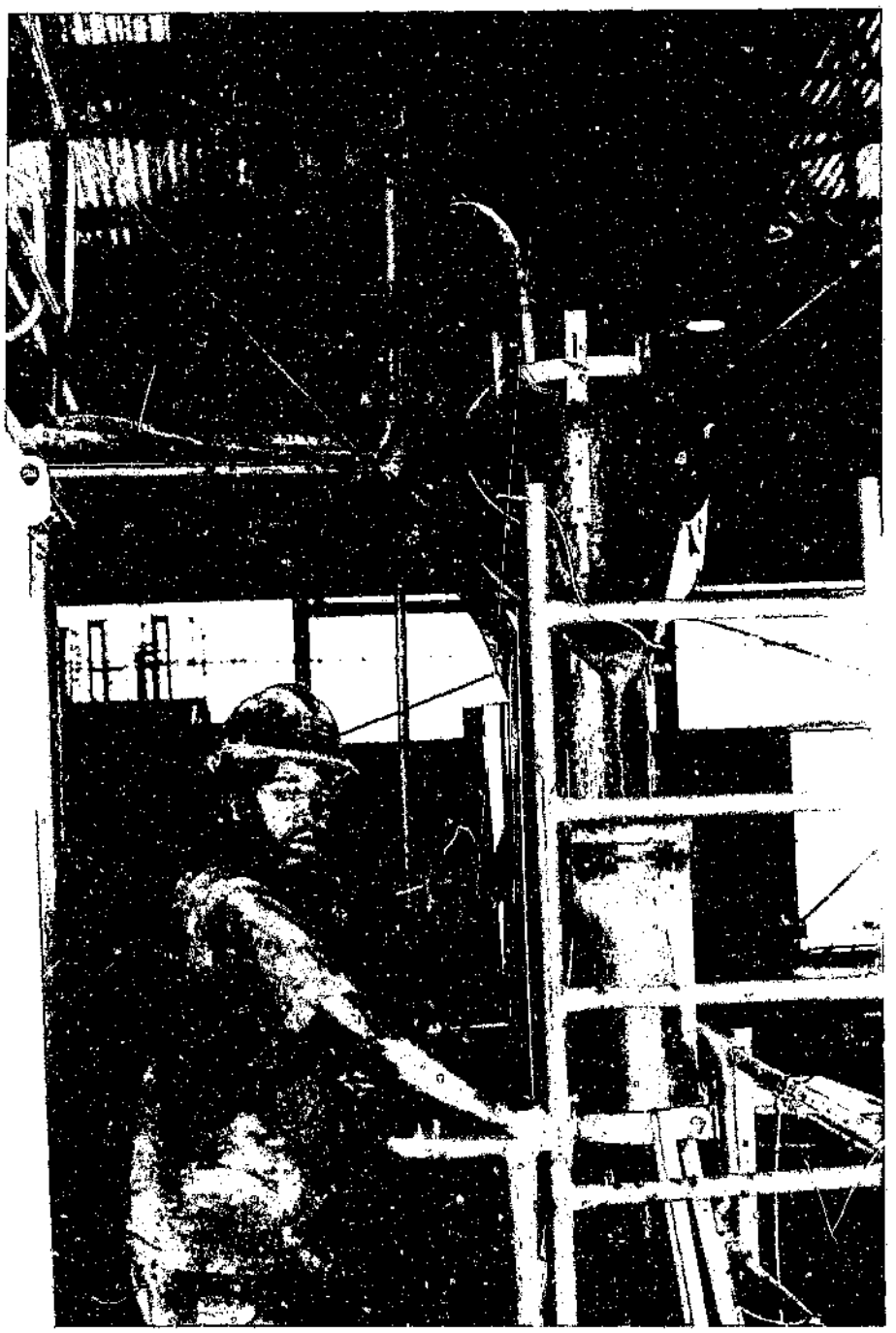


Figure 3.7

TOP SECTION OF THE PILOT PLANT



Six manometers were fitted at 200 millimetre intervals up the side of the top section to measure gas hold-up in the cleaning zone. To measure gas hold-up accurately in the collection zone a pressure gauge was fitted at the top of the column and connected via a water manometer to a point 500 millimetres above the injector. A U-tube was used at the point of connection to the column to prevent the manometer from draining when the column was emptied. The use of the manometer leg meant that a very sensitive pressure gauge could be used. The pressure difference between the lowest of the six manometers in the top section and the pressure gauge reading was used to calculate the relative density in the zone between the two measuring points. The relative density of the pulp in the zone was assumed to be that of the tailings, and the difference between actual and calculated relative density was used to calculate gas hold-up. The gas hold-ups in the top section were calculated similarly.

The USBM bubble generation system was controlled from the top of the column while the generator itself was situated by the injector. The 20 ppm Dow 200 frother solution was made up in a one cubic metre make-up tank. The pH of the solution was adjusted to 3,8 at the beginning of every day. The frother solution was pressurised using a variable speed Bredel SP15 peristaltic pump connected to an accumulator to dampen the pulses. Three in-line filters were used to ensure the water was clean enough so as not to block up the bubble generator or sparger holes, and a pressure gauge was fitted into the line to detect when the filters needed cleaning. The flowrate was measured using a GEC 14X5 rotameter. A ball valve and non-return valve were fitted to the line for ease of operation. A pressure gauge was fitted between the bubble generator and the injector to monitor the pressure drop across the bubble generator.

Initially the bubbler was operated with mine water, which had high dissolved and suspended solids content. This water proved unsuitable for the bubble generator, as the silica bed and fitted disks became inoperative due to suspended solids, and the sparger holes became sealed. The problem was overcome by the use of Rand Water Board water, which proved an unfortunate condition, as

RWB water is expensive and makes the USBM bubble generator unattractive for commercial applications.

The compressed air or gas was instrument air, and was available at seven bar. The air contained significant amounts of oil, and two in-line coalescing filters were fitted to ensure that the air had less than 1 ppm oil present. The air rate was controlled by an air regulator fitted before a GEC 14K air rotameter. A pressure gauge was fitted at the top of the rotameter to measure the pressure in the rotameter. The rotameter was calibrated for air at one atmosphere and 15°C. The reading was adjusted for pressure according to the formula below in kPa.

$$Q_{actual} = Q_{calibration} \sqrt{\frac{P_{gauge} + P_{atmosphere}}{P_{calibration}}}$$

where

Q_{actual} is the volumetric flowrate at 1 atmosphere and 15°C.

$Q_{calibration}$ is the flowrate obtained from the manufacturer's calibration curve.

P_{gauge} is the gauge pressure at the top of the rotameter.

$P_{atmosphere}$ is atmospheric pressure at the measurement location.

$P_{calibration}$ is the pressure at which the rotameter was calibrated by the manufacturer.

A non-return valve and gate valve were fitted to the air line for ease of operation. All the controls for the bubble generation system were situated at the top of the column for ease of control. The feed system pumped feed from the scavenger concentrate pump

discharge line up to a steady-head tank. The tank was cylindrical with a conical base and had a volume of about 200 litres. A slurry hose was fitted to the apex of the cone and the slurry was fed by gravity to the column. The pH control on the plant was poor and a pH meter was fitted so tests could be stopped when the plant pH went out of control. As mentioned previously, three different sizes of hose were used to allow a wide range of flowrates to be tested without the line blocking up. All the hose connections were two-inch Table D type flanges and each slurry hose was fitted to a polypropylene two-inch flange. Thus the hoses could be interchanged quickly and efficiently. A reagent pump was installed to allow reagent addition to the feed tank.

The major operating problem of the whole pilot plant was the blocking of the line from the feed steady-head tank to the column. Eventually an air line was fitted to the middle of the slurry line so that when the line blocked it could be immediately cleared by compressed air. This proved to be a satisfactory arrangement.

The feed dilution water was Rand Water Board water and connected via a gate valve to a GEC 65S rotameter. The line fed into the feed steady-head tank. The line was designed to dilute the maximum feed rate by 100 percent.

The wash water was fed from a stainless steel steady-head tank. The tank was air agitated and the level control was by a ballcock valve. A Hanna Instruments pH controller was fitted and frother was added at a pre-determined rate by a Watson Marlow 101 UR variable speed peristaltic pump.

The wash water flowrate was measured using a GEC 24S rotameter and controlled manually with a gate valve. The line was prone to air locking, and a ball valve was fitted below the rotameter to remove air locks and to allow the wash water to be sampled for pH measurement during a test.

All the controls were fitted at the top of the column for ease of

operation and are illustrated in the photograph in Figure 3.8. A data sheet was used to record the operating parameters (Figure 3.9).

3.6 CONTROL STRATEGY

The control strategies employed in column flotation worldwide are discussed in Section 2. The simple 'level' control strategy was chosen for ERGO because it gives adequate control and is the most flexible for pilot plant operation, which, unlike plant operation, has the undivided attention of the engineer. Because of this close attention, the differential pressure cell finally used gave adequate level control, whereas in normal plant operation its sensitivity to fluctuating conditions would make it unreliable.

Initially, a conductivity meter was used as the transducer for level measurement. The method was recommended by Moys and Finch (1988) to overcome the problems associated with differential pressure transducers in column flotation. Moys and Finch state: "The use of pneumatic pressure measurements for obtaining estimates of interface level is subject to large errors because the measurement is a strong function of all variables which define the state of the slurry and the froth above the point of measurement". The variables referred to are bubble size, gas hold up, particle size and gas rate. Deviations of 400 millimetres have been measured on an industrial plant (Moys and Finch 1988).

The conductivity probes consisted of stainless steel bolts with the centres drilled out fitted every 200 millimetres along the side of the top clear PVC section of the column. The two probes above and below the interface were connected to the conductivity meter via banana clips inserted into the bolts.

Figure 3.8

PILOT PLANT INSTRUMENTATION

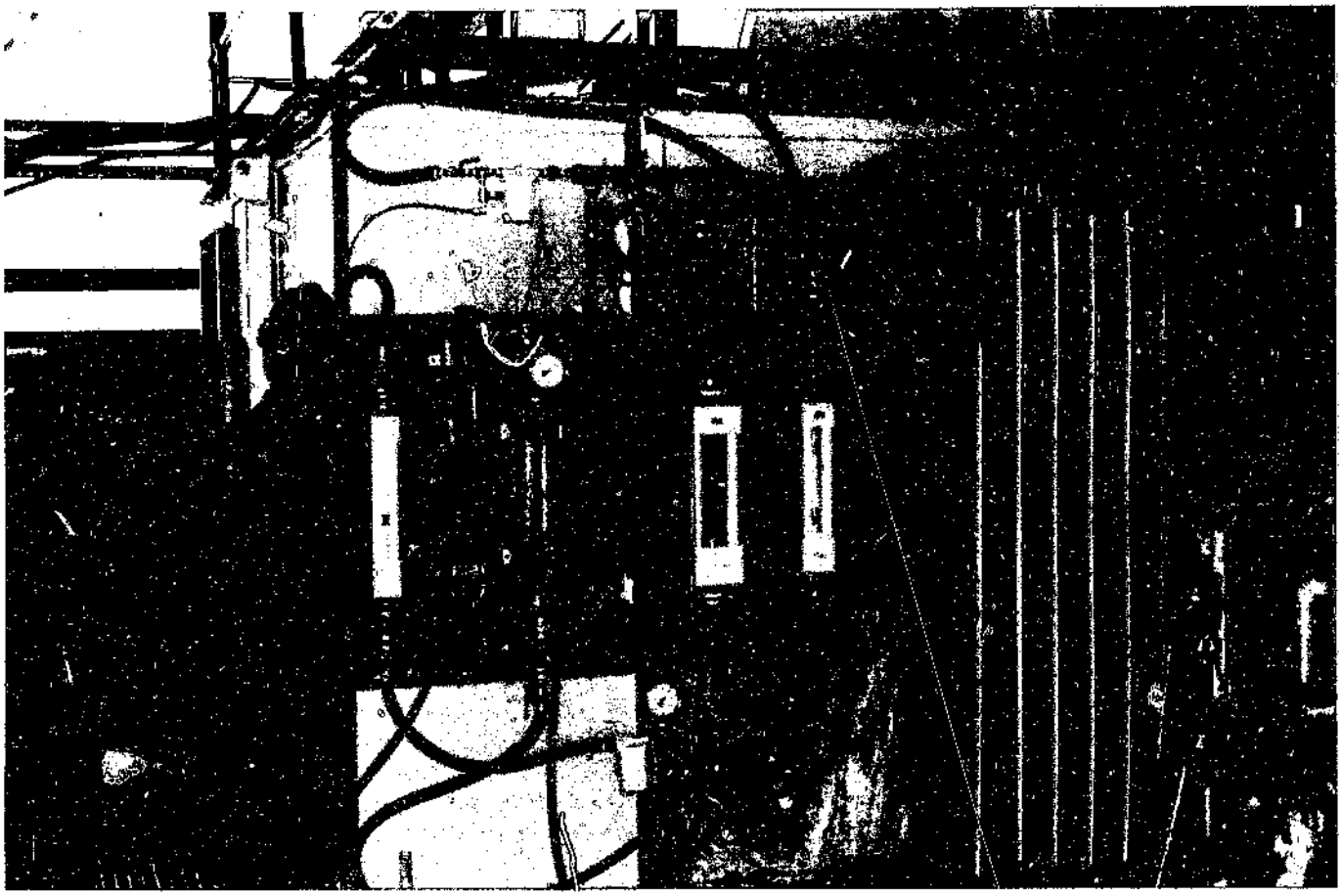


Figure 3.9

DATA SHEET

E R G O FLOTATION COLUMN.
XXXXXXXXXXXXXXXXXXXXXXXXXXXXX

Date Stream

	TEST NUMBER	UNITS		
	Time			
BUBBLE GENERATOR:	Water Pressure	kPa		
	Water Rotameter Reading			
	Water pH			
	Water Frother Concentration	ppm		
	Gas Pressure	kPa		
	Gas Rotameter Reading			
	Injector Pressure	kPa		
WASH WATER:	Rotameter Reading			
	pH			
	Frother Concentration	ppm		
FROTH PHASE:	Froth Height	mm		
	W/W Distributor Position	mm		
HOLDUP:	Collection Zone Dp.	kPa		
	Manometer Reading 1	mm		
	Manometer Reading 2	mm		
	Manometer Reading 3	mm		
	Manometer Reading 4	mm		
	Manometer Reading 5	mm		
	Manometer Reading 6	mm		
	Manometer Reading 7	mm		
CONTROL:	Set Point (Dp Cell)			
	Conductivity			
CONCENTRATE:	Sample Volume	l		
	Sample Mass	kg		
	pH			
TAILS:	Flowrate	l/min		
	Sample Volume	l		
	Sample Mass	kg		
	pH			
HEAD:	Sample Volume	l		
	Sample Mass	kg		
	Sample pH			
	pH Meter Reading			

The conductivity meter was connected to a Rex F9 temperature controller. Only proportional and integral control was used, and control set points were determined empirically using the quarter wave dampening technique. The controller output was connected to a current to pneumatic converter, i.e. an I to P converter. The pneumatic pressure controlled a Fischer pinch valve on the tailings line. All electrical signals were 4 to 20 milliamperes.

However, the conductivity meter proved to be sensitive to changes in the feed pH, which fluctuated widely because of poor plant control, so the conductivity meter was replaced by the differential pressure cell. One side of the DP cell was connected 1,5 metres from the top of the column and the other side to the atmosphere. The differential pressure cell gave good level indication once all the column variables, such as gas and wash water rate, were fixed. However, if these variables were changed, the level maintained by the control system also changed, but not usually significantly. The final control system worked satisfactorily.

Moys (1989) adapted the conductivity meter to measure conductivity across the pulp/froth interface relative to the conductivity of the pulp, which essentially eliminates the problem of pH variations in the feed.

CHAPTER 4

EXPERIMENTAL PROCEDURE

4.1 OPERATION OF THE USBM BUBBLE GENERATOR

4.1.1 Testing Of The Bubble Generator

Plug flow is the ideal flow regime for column flotation, and although it is never possible, it may be approximated. Bubble flux behaviour is important in promoting plug flow. Bubbles must have uniform size and be evenly distributed across the cross-section of the column.

McKay et al (1988), the designers of the USBM bubble generator, claimed that its bubble distribution is uniform 0,46 m above the sparger, and the system was tested at the University of the Witwatersrand before being installed at ERGO. It was to be tested on a two-metre section of clear PVC to observe bubble size and bubble size distribution, but the clear section failed under pressure testing, and was replaced by a stainless steel section.

The uniformity of bubble distribution was measured using conductivity. Conductivity, however, is dependent on a number of factors, including pH, temperature, gas hold-up and bubble size. So to minimise variation, a flow of fresh water with a superficial velocity of 0,7 cm/sec was used to keep the frother concentration in the column at 10 ppm of Dow 200. The bubble generator pressure was 515 kPa. The conductivity was measured at 16 points throughout the cross-section of the column as shown in Figure 4.1. The probe, depicted in Figure 4.2, was held one metre above the injector.

Figure 4.1

MEASURING POINTS FOR THE CONDUCTIVITY PROBE

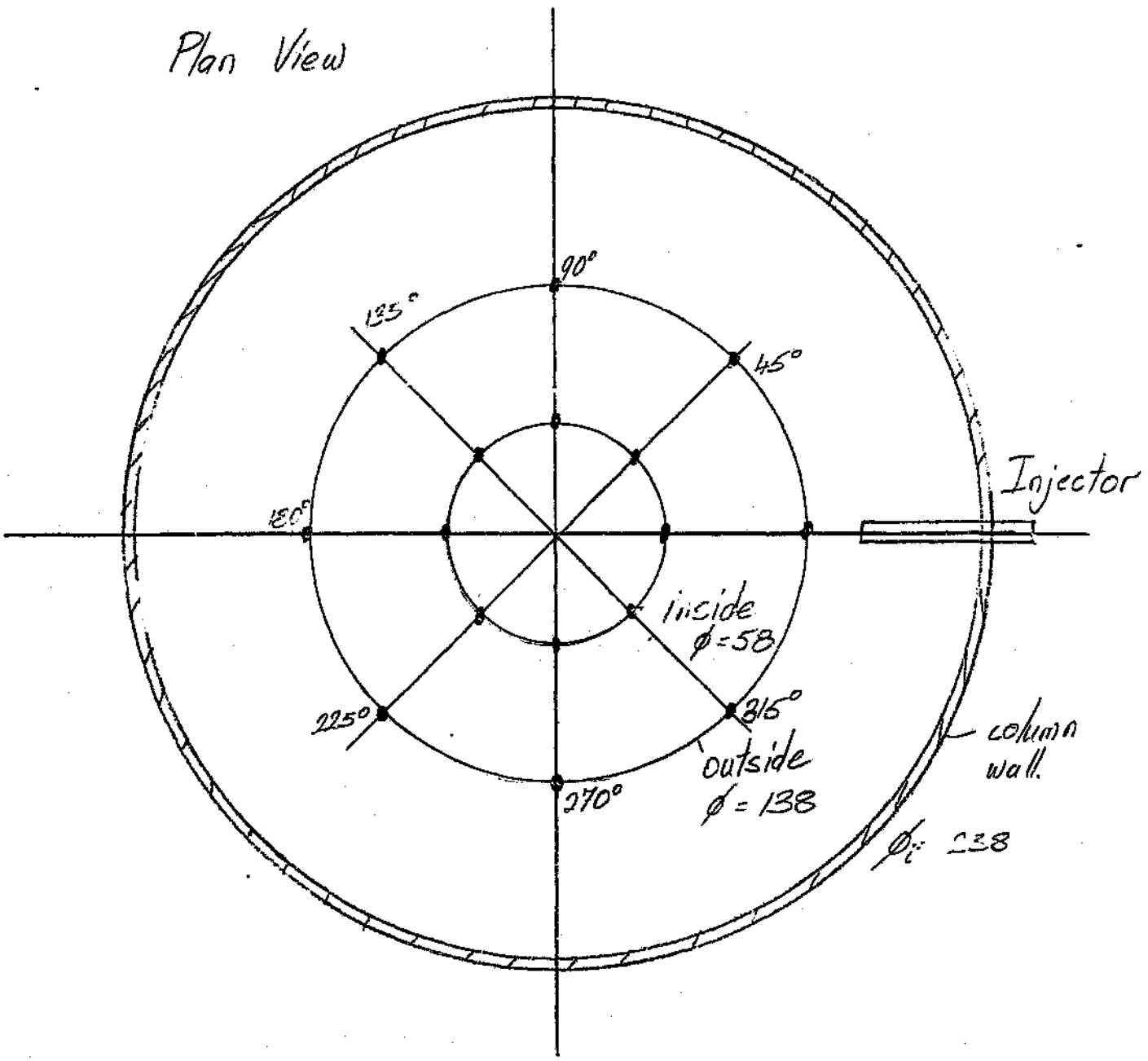
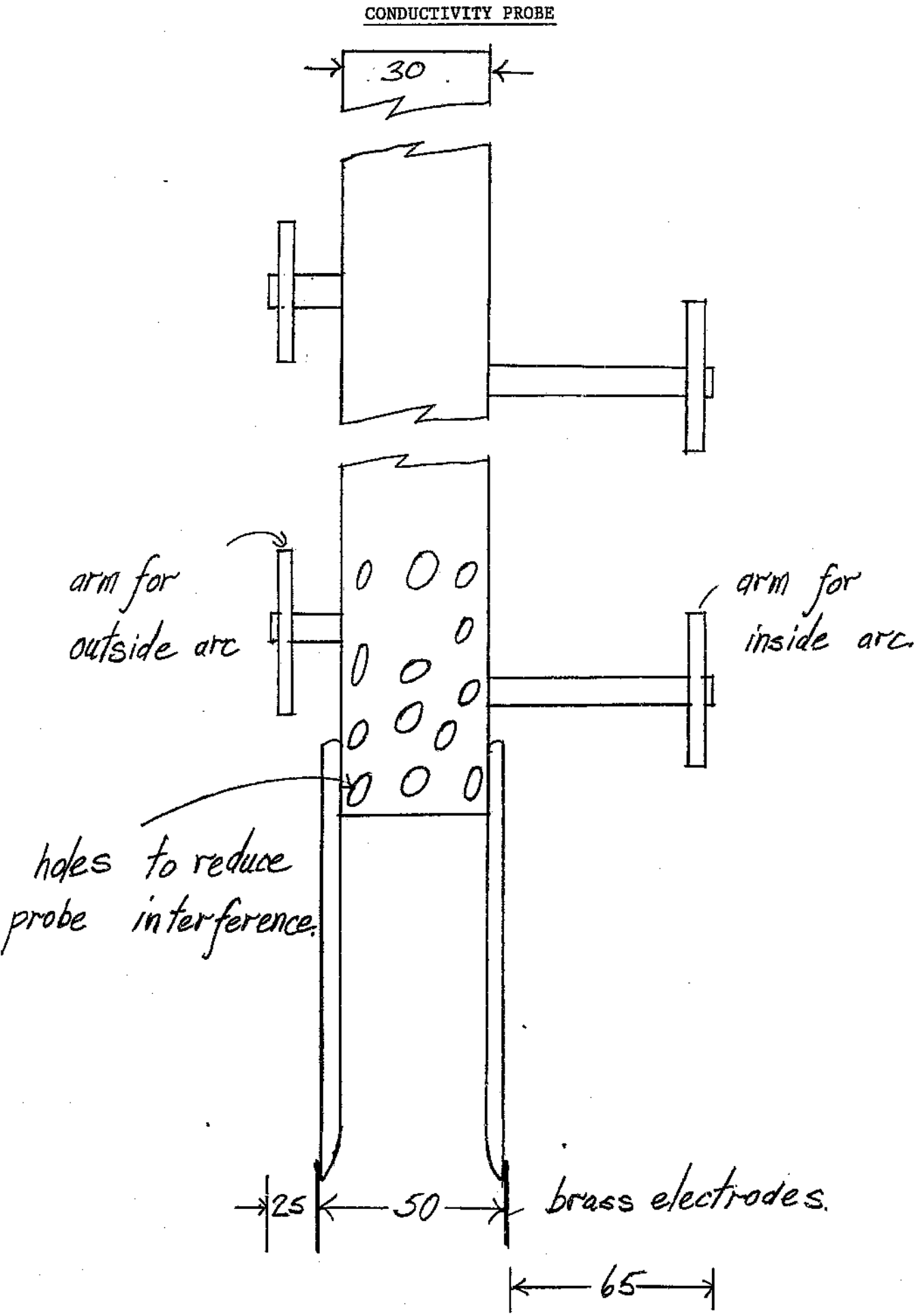


Figure 4.2


CONDUCTIVITY PROBE
30
arm for outside arc
arm for inside arc.
holes to reduce probe interference.
25
50
brass electrodes.
65

The measurements were begun on the outside arc above the injector fitting, and the probe was moved counter-clockwise from measuring point to measuring point, some distance from the column wall to prevent the stainless steel interfering with the conductivity reading. Eight readings were taken. The inside arc was measured similarly, and the readings were recorded by computer.

Five such measurements of conductivity were made at different gas rates, ranging between 2 and 2.8 cm/sec. The first two tests, BS/1 and BS/2, are plotted in Figure 4.3. The conductivity is represented as a percentage of the conductivity reading without bubbles present, and is plotted against probe position. Uniform bubble distribution is indicated by a straight horizontal line, so the steady rise in percent conductivity in BS/1 suggested that the electrode was probably greasy.

Gas hold-up is another important parameter for good flotation column performance, and it was measured at gas rates between 1,6 cm/sec and 2,8 cm/sec. The results are tabulated in Table 4.1 and show that hold-up increased with gas rate.

It was found, however, that a considerable increase in gas rate gave only a small increase in hold-up, presumably because of bubble size also increasing with gas rate.

During commissioning at ERGO, Sallies Mine water was used in the bubble generator. However, this water was found to scale up the injector holes. When the injector was operated outside the column it was discovered that the holes furthest from the inlet blocked first. This was because the holes nearest the inlet produced wet spray and the holes furthest from the inlet produced a very fine dry spray with associated evaporation and consequent scaling.

Figure 4.3

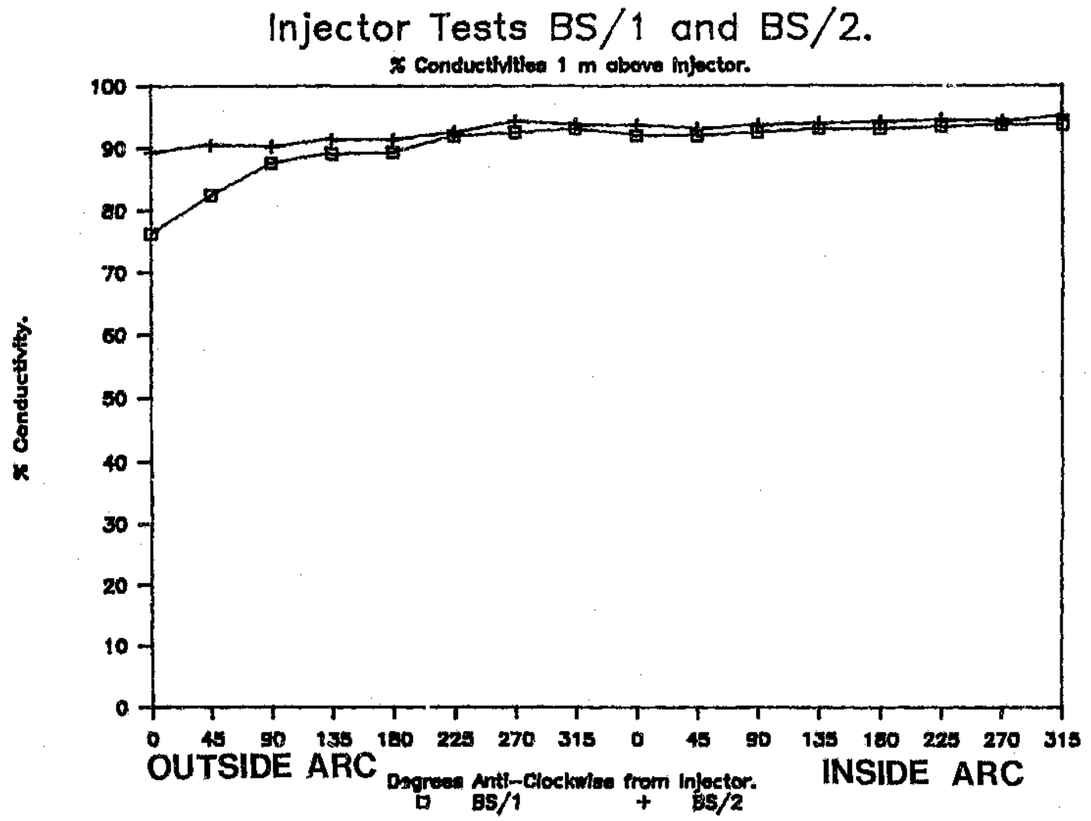

Injector Tests BS/1 and BS/2.
% Conductivities 1 m above injector.
% Conductivity.
100
90
80
70
60
50
40
30
20
10
0
0 45 90 135 180 225 270 315 0 45 90 135 180 225 270 315
OUTSIDE ARC
Degrees Anti-Clockwise from Injector.
INSIDE ARC
□ BS/1
+ BS/2

Table 4.1

Bubble Generator Test

Test Number	Air Flowrate cm/sec	Hold-Up %
1	1,6	14,8
2	1,8	16,3
3	2,4	18,5
4	2,8	18,7
5	2,8	20,1

As a result the decision was made to clean the injector regularly with a rifle cleaner and acid, and the injector holes were checked for scaling. The apertures of the holes were checked by connecting the injector to a water supply and ensuring, by measuring or visually, that there was a similar water flowrate from each hole.

4.2 EXPERIMENTAL PROCEDURE

Several tests were conducted on ERGO cleaner feed using conventional laboratory flotation machines, to give a reference for the column flotation results. The first test was a release analysis test described by Dell for coal flotation. The second set of tests were routine laboratory flotation tests with and without reagent addition. The experimental procedure for start-up and shut-down of the flotation column is also described in this section.

4.2.1 Release Analysis

The objective of the release analysis is to make a perfect separation of a sample into a number of fractions of diminishing flotability (Dell 1964). The release anlysis was conducted to establish the optimum grade/recovery curve for ERGO scavanger concentrate.

A conventional laboratory flotation machine which allowed air rate to be controlled separately to impeller speed was used. The testwork was conducted at the ERGO metallurgical laboratories. A sample of ERGO cleaner feed (scavenger concentrate) was taken at the plant and immediately placed in the flotation machine. The machine had a capacity of approximately two kilograms. After conditioning the sample was floated in the normal way, using every means possible

(such as keeping up the pulp level, scraping well, adding extra frother) to obtain a high recovery. The cell was then emptied and the froth returned to the cell. The tailings were filtered and dried and prepared for analysis.

This procedure was repeated with further additions of reagent as necessary.

Thus a separation had been effected between the floatable and the non-flotable material, and the final separation was then to separate the flotable material into fractions of progressively less flotable material.

This was done by returning the concentrate to the cell. The air was turned off and the impeller speed was reduced to the point where froth formation and flotation cease. Both were then increased cautiously until flotation was just discernable. This froth was scraped for as long as it appeared.

The concentrate sample basin was then changed; the impeller speed and air rate were slightly increased, and the froth was scraped for as long as it appeared. The concentrate sample basin was again changed and the procedure repeated until the impeller was operating at full speed with the air valve fully open. Reagent was added as needed. Eight fractions in all were collected. The tailings were filtered, dried and prepared for assay, as were the eight concentrate samples.

4.2.2 Routine Conventional Laboratory Cell Tests

The routine conventional laboratory tests were conducted at Anglo American Research Laboratories by Mr J.F. Muller of the Ore Processing section. The tests were done in one kilogram D12 Denver cells.The cleaner feed stream at ERGO was sampled in the morning, and the sample transported to the AARL where it was floated the same day. This was to minimise any oxidation or alteration of the pyrite surfaces and consequent deterioration of the flotation

response of the sample. The sample was floated "as received" i.e. no adjustments were made to the pH or RD of the sample.

Four tests in all were conducted. The first two tests, one without and one with reagent addition, were repeated, because the pH of the first sample was 1,56, whereas the correct pH for acidic pyrite flotation is 3,8. The poor pH control on the plant proved to be a headache throughout the testwork, and the pH readings recorded prompted plant management to upgrade their pH control. A further sample was taken with a pH of 3,96 which was considered to be within experimental limits, and tests with and without reagent addition were also conducted on this material.

The experimental procedure described below assumes reagent addition. The tests conducted without reagent addition had a similar procedure except for reagent addition and reagent conditioning.

The impellor speed was set at 1 350 rpm. A head sample was dipped for RD measurements. The sample was conditioned by agitating for five minutes and the reagents were then added. The dosages of the reagents are given in Tables 4.2 and 4.3.

The reagent was conditioned for one minute and concentrate was pulled for five minutes. During flotation the froth is pulled from the back of the cell to the cell overflow lip four or five times a minute, manually, with a scraper. Senkol 50 and more copper sulphate were added, and this time conditioned for half a minute. The sample was floated for a further five minutes. Again reagent was added and conditioned for half a minute, and the sample floated for five minutes. This procedure was repeated three more times for flotation times of five, ten and ten minutes respectively. The procedure is summarised in Tables 4.2 to 4.5. Each concentrate was collected in a separate tray and filtered, dried, and rubbed down through a sieve to break up agglomerates. The sample was then weighed and split for gold, sulphide sulphur and

Table 4.2

Conventional Laboratory Tests - Experimental Conditions

Test 1

Stage	Reagent Addition g/t		Time in Minutes	
	Copper Sulphate	Senkol 50	Conditioning	Flotation
C1	-	-	5	5
C2	-	-	-	5
C3	-	-	-	5
C4	-	-	-	5
C5	-	-	-	10
C6	-	-	-	10
T	-	-	-	-

Test 2

Stage	Reagent Addition g/t		Time in Minutes	
	Copper Sulphate	Senkol 50	Conditioning	Flotation
			5	-
C1	49	-	1	5
C2	24	49	0,5	5
C3	10	19	0,5	5
C4	10	19	0,5	5
C5	10	19	0,5	10
C6	5	10	0,5	10
T	-	-	-	-

Table 4.3

Conventional Laboratory Tests - Experimental Conditions

Test 3

Stage	Reagent Addition g/t		Time in Minutes	
	Copper Sulphate	Senkol 50	Conditioning	Flotation
C1	-	-	5	5
C2	-	-	-	5
C3	-	-	-	5
C4	-	-	-	5
C5	-	-	-	10
C6	-	-	-	10
T	-	-	-	-

Test 4

Stage	Reagent Addition g/t		Time in Minutes	
	Copper Sulphate	Senkol 50	Conditioning	Flotation
			5	
C1	38	-	1,0	5
C2	19	38	0,5	5
C3	8	15	0,5	5
C4	4	8	0,5	5
C5	4	8	0,5	10
C6	4	8	0,5	10

uranium analyses. The analyses were conducted by the Anglo American Research Laboratories Analytical Department according to the methods described by Lenahan and Murray-Smith (1988).

4.2.3 Sampling Procedure

The feed, concentrate and tailings streams were each fitted with a sampler, a short vertical pipe with a conical outflow supported on a tripod. The slurry hose was held in the pipe by a ring fixed above the pipe mouth. A sample was taken by lifting the hose out of its ring holder and placing it into a sample bucket at the same elevation. In this way the whole stream was sampled. The hose was returned to the sampler after the sampling period.

The feed sampler was considerably larger than the concentrate and tailings samplers because it was also used as a conditioning tank, and had a volume of approximately 200 litres.

Each stream was sampled four times over approximately a residence time to ensure representative sampling. The sampling time depended on the flowrate of the stream. Typically the feed and tailings streams were sampled four times for ten seconds each; a total sampling time of 40 seconds. The concentrate stream was typically sampled four times for 20 seconds each.

Samples were taken in 20 litre buckets lined with fresh plastic bags each time to avoid the possibility of contamination. The samples were transported to the Anglo American Research Laboratories for sample preparation and analysis. The sample volumes and wet and dry masses were recorded and used to calculate flowrates, with the exception of the tailings flowrate which was measured separately.

Midway through the sampling procedure, all the relevant process variables were recorded. The data sheet shown in Figure 3.9 was used. The operating parameters of the bubble generator were recorded: pressures and flowrates of the water and air, and the operating pressure of the injector. The observed bubble size at

the base and top of the froth were recorded. The pH of each slurry and water stream was recorded. The wash water flowrate and the wash distributor depth was measured. The reagent addition was recorded. The treatment of the samples is given in Section 4.3.

4.2.4 Flotation Column Operation

Reagent addition

Frother was added to the wash water and bubble generator water in excess (more than 70 g/t) so that its actual concentration did not influence the results. In some earlier tests copper sulphate was added at 10 g/t as a sulphide activator and Senkol 50 was added at 20 g/t as a sulphide collector. When the conventional batch tests were done and their results indicated that these reagents did not improve final recoveries, they were no longer added.

Start-up (see Figure 3.6)

Initially the bubble generator frother tank was made up with RWB (Rand Water Board) water and 20 ppm Dow 200 frother, and the pH was adjusted to 3.8. The USBM bubble generator was started up first to prevent the injector holes from being blocked. The water pump for the generator was switched on and the pressure allowed to build up. The water pump was a variable speed peristaltic Bredel pump connected to an accumulator to dampen its pulses. As the pressure in the water system built up, so the air regulator was turned up to the corresponding water pressure, until the whole bubble generation system reached the pressure of five bar.

Once the bubble generation system was operating satisfactorily, and the gas rate set correctly by adjusting the gas regulator, the column was ready for pulp. The feed dilution water was turned on to prevent the line from the feed steady-head tank to the column from blocking.

The feed pump was switched on. The feed was taken from a sampling point in the scavenger concentrate pump discharge line.

This valve had to have been opened before the feed pump was operated. Once the slurry was flowing into the column the feed dilution water was shut off and compressed air was added to thebottom of the column to agitate the contents of the column. The wash-water valve was adjusted to give the correct flowrate on the rotameter.

When the slurry level in the column rose to the correct level, the compressed air at the bottom of the column was closed off and the control system switched on. The timing for the change-over was important because, if effected too early, the tailings line would block up.

Once the control system was controlling the slurry/froth interface correctly and the froth was overflowing freely, the column was deemed to have been started-up.

Before samples were taken all the water lines were flushed out e.g. the manometer lines and the differential pressure cell water line.

Shut down

At the end of a day's testwork the column was shut down. First, the controller was switched off (the tailings valve was set to fail open) and the column began to drain. The feed dilution water was switched on and the feed line was flushed with process water. The feed pump was then switched off. When the feed to the column was free of solids the feed dilution water was switched off. The wash-water pH controller and frother pump were switched off. When the column had completely drained the wash-water was shut off and the bubble generator water pump shut off. The gas pressure was reduced at the pressure regulator as the water pressure dropped, until the bubble generator reached atmospheric pressure when all the valves in the bubble generation system were closed. Finally the column was washed down and the water and air supplies to the column were shut off.

4.3 SAMPLE PREPARATION AND ANALYSIS

4.3.1 Sample Preparation

The pilot plant samples were transported back to the Anglo American Research Laboratories for sample preparation and analysis.

Initially the samples were removed from the buckets in their sample bags and weighed. The sample bags were doubled, as it was found that single bags tended to burst. The average weight of the bags was known and subtracted from the total mass to yield the sample mass. The volumes of the samples were measured using a two-litre measuring cylinder: first the tailings samples, then head and concentrate samples to avoid contamination. The acute problem of gold cross-contamination between samples was greatly reduced because no large gold particles were present in the material. (ERGO material is mine dump material which has already been leached for free gold).

The samples were then filtered in a pressure filter and dried in a drying oven. The dried sample was rubbed through a 300 micrometre screen with a large rubber bung to break up the agglomerates formed during filtering and drying. The sample was then weighed to obtain its dry mass, and split into four duplicate samples as follows:-

2 x 150 gms for gold analysis (50 gms for concentrate gold analysis)
2 x 5 gms for sulphide sulphur
2 x 20 gms for uranium analyses
2 x 20 gms for Malvern size analyser

An additional hundred gram sample was screened at 75 micrometres and the plus and minus fractions were submitted for gold and sulphide analysis.

4.3.2 Sample Analysis

Analysis was done by the Anglo American Research Laboratories analytical department by procedures outlined by Lenahan and Murray-Smith (1986).

Gold analysis was done by fire assay; uranium is analysed as U_3O_8 by X-ray fluorescence spectrometry.

Sulphide sulphur is approximated as available sulphur at 850 °C in the South African gold mining industry. The method for determining available sulphur was originally developed because of the importance of combustible sulphur content in ensuring sufficient heat of reaction for roaster operation. Although not an exact determination of sulphide sulphur, it is considered a fair approximation, and the more correct X-ray diffraction method is not performed at AARL.

4.4 DATA ANALYSIS

The flotation column has many more operating variables than conventional cells. This, together with the thorough treatment of the samples and extensive instrumentation on the column, meant that a considerable amount of data was collected each run. A Lotus 1-2-3 spreadsheet was used to process the data. The spreadsheets for the testwork are contained in Appendix I.

The top section of the spreadsheet included all instrument readings and sample data, including bubble size at the base of the froth and pH readings of the three streams. The lower part of the spreadsheet contained the calculated results e g. percentage recovery and residence times. The screening results were also tabulated. The important data were summarised in table 5.1 and a regression was done to establish the relationship between the dependent and independent variables.

Table 5.1

Column Cell Results on Ergo Scavanger Concentrate

Test Number	Mass Recovery	Gold Recovery	Gold Grade	Gold in Tails	S/S Recovery	S/S Grade	S/S in Tails	Residence Time	Superficial Gas Velocity	Wash Water Rate	Bias Ratio	Carrying Capacity	Hold-up	Feed Solids	Feed Gold Grade	Feed S/S Grade	Froth Height
Units	%	%	g/t	g/t	%	%	%	min	cm/sec	ℓ/min	-	g/min/cm²	%	%	g/t	%	mm
TTL1	23	67	10,7	1,62	77	21,4	1,95	12,4	1,24	10,3	1,156	1,85	-	13	2,9	7,3	710
TTL2	15	48	11,7	2,17	62	20,6	2,20	13,6	1,13	8,5	1,185	1,06	-	13	4,7	5,6	870
TTL3	8	35	10,6	1,67	48	21,6	2,01	12,1	1,31	8,5	1,131	0,73	-	15	2,7	3,4	850
TTL4	8	37	13,2	2,05	57	19,3	1,34	12,6	1,50	8,5	1,078	0,77	-	14	3,8	4,0	820
BR1	8	47	14,8	1,57	66	27,1	1,27	14,3	0,94	6,1	1,20	0,79	-	16	3,7	4,6	410
BR2	16	55	13,6	2,15	80	26,3	1,28	16,4	0,94	4,8	1,15	1,36	-	16	3,5	4,7	390
BR3	10	42	9,94	1,59	75	32,6	1,26	13,8	0,94	3,7	1,08	1,05	-	17	5,3	4,8	410
CF1	30	42	8,97	5,33	50	24,9	10,96	21,5	1,43	9,0	0,855	2,42	22	17	7,8	20,8	150
CF2	49	57	9,24	6,66	69	27,9	12,22	33,5	1,24	0,0	0,47	3,86	19	N/A	N/A	N/A	270
RT1	18	55	12,0	2,17	71	24,0	2,10	14,2	1,125	6,7	1,14	1,16	15	12,4	6,4	4,2	315
RT2	16	68	14,5	1,29	62	19,3	2,22	14,0	1,125	6,7	1,14	0,86	15	12,6	4,6	3,8	290
JB1	18	67	13,5	1,48	86	29,2	1,09	20,2	1,05	6,5	1,33	0,71	15	14	3,4	4,9	500
JB2	12	58	12,7	1,28	82	24,1	0,74	23,4	1,20	4,8	1,11	0,48	16	11	3,3	5,6	490
JB3	31	73	7,45	1,23	90	21,1	1,10	42,8	1,20	4,0	1,61	0,37	14	17	3,1	5,0	420
JB4	26	71	9,01	1,27	87	22,8	1,74	42,6	1,20	4,0	1,83	0,34	14	15	3,6	5,2	420

CHAPTER 5

RESULTS AND DISCUSSION

5.1 PRESENTATION OF RESULTS

5.1.1 Introduction

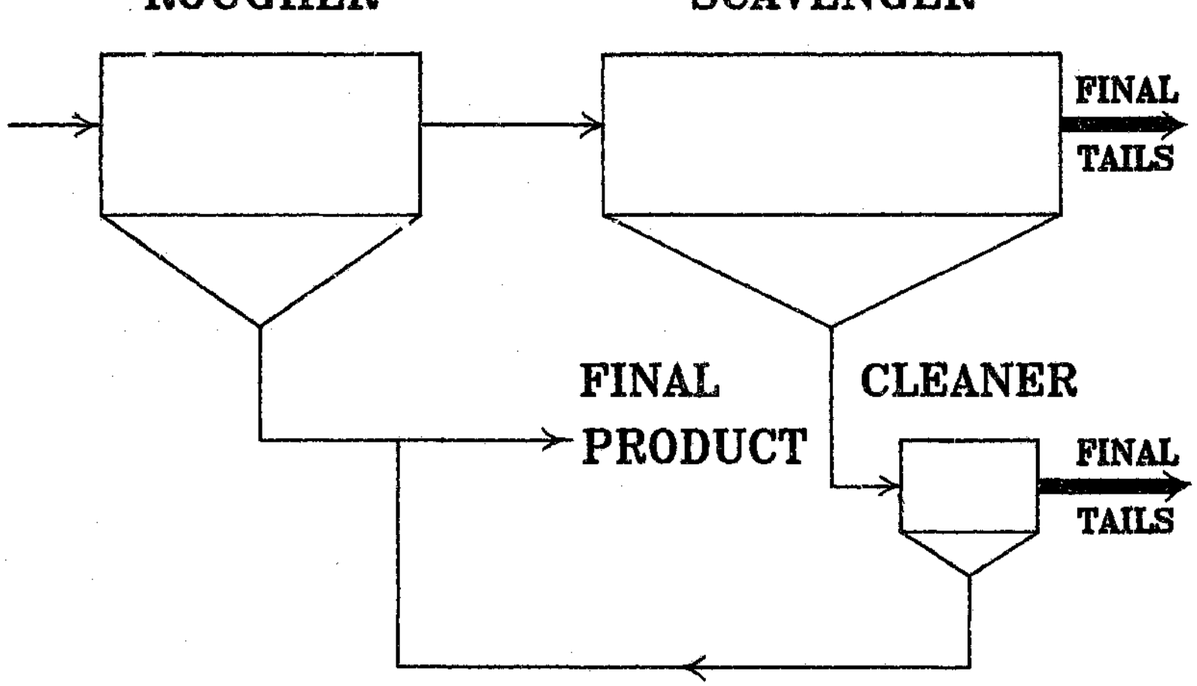
The objective of the project was to evaluate column flotation under plant conditions at ERGO. Once the column flotation pilot plant had been commissioned the best operating philosophy for ERGO had to be established. Previous experience in conventional flotation proved to be a handicap, and only after several months of frustration and heart-searching was a good operating philosophy learned.

The next stage was to operate the column as close to plant conditions as possible for comparison purposes. This proved difficult for three main reasons. Firstly, the residence times of column and plant were completely dissimilar. The maximum residence time in a column is limited by the settling rate of the solids, and 20 minutes appears to be the upper limit for industrial columns. ERGO's cleaner circuit, on the other hand, has a residence time of 90 minutes, due to a deviation in plant usage from original design. Originally, the cleaner circuit was intended to clean both rougher and scavenger concentrates, but after commissioning the concentrate from the 10 rougher cells was routed to the final product (see figure 5.1 page 95). The lower feed rate to the cleaner bank resulted in the increase in the residence time, and consequently the cleaner cells' performance was unusually good.

The second difficulty in comparison was that ERGO's cleaner circuit treats scavenger concentrate from A, B, C and D streams, whereas the column flotation pilot plant operated on A stream scavenger concentrate only. So although the feed materials were similar, they were not identical.

Figure 5.1

E R G O FLOTATION PLANT



Thirdly, although the column feed rate was measured, the cleaner circuit's performance could only be estimated because the feed rate was neither measured nor constant.

Conventional laboratory cell data have been used for design at the Anglo American Research Laboratories for many years. Experience has shown that the laboratory cell can predict the grade/recovery relationship for the plant. The flotation time for a full scale conventional flotation plant is assumed to be 2,5 times that of the laboratory cell. Therefore laboratory-scale conventional tests were carried out, and the results were used for comparison with the flotation column.

5.1.2 Comparison Of The Grade/Recovery Curves For Column And Conventional Flotation

The grade/recovery data for both the column cell and conventional flotation are plotted in Figure 5.2 for gold flotation, and 5.3 for sulphide sulphur flotation. Sulphide sulphur is the limiting grade at ERGO because a minimum sulphide sulphur grade of 29 percent is required for efficient operation of the pyrite roaster. Sulphide sulphur grade is thus the abscissa in both graphs.

Because of the large number of variables in column flotation, it is possible for the column to be operated suboptimally. This was the case with most of the data, as the column was operated to establish the effect of the variables on column performance, not to obtain optimal results. For instance, sulphide sulphur grades achieved by the column were unexpectedly low, only slightly better than conventional grades (see Figure 5.3). Because of the froth washing they should have been better, had the operating parameters been optimised.

Therefore the curves drawn on the graph were an attempt to represent optimum behaviour, and demonstrate the superior metallurgy of the column cell for ERGO scavenger concentrate. The result most significant for the South African gold mining industry is the 13 percent improvement in gold recovery achieved by the column at a sulphide sulphur concentrate grade of 29 percent (Figure 5.2) and it is a pity that this conclusion is supported by only one data point. This is

Figure 5.2

COMPARISON OF COLUMN AND BENCH CELL

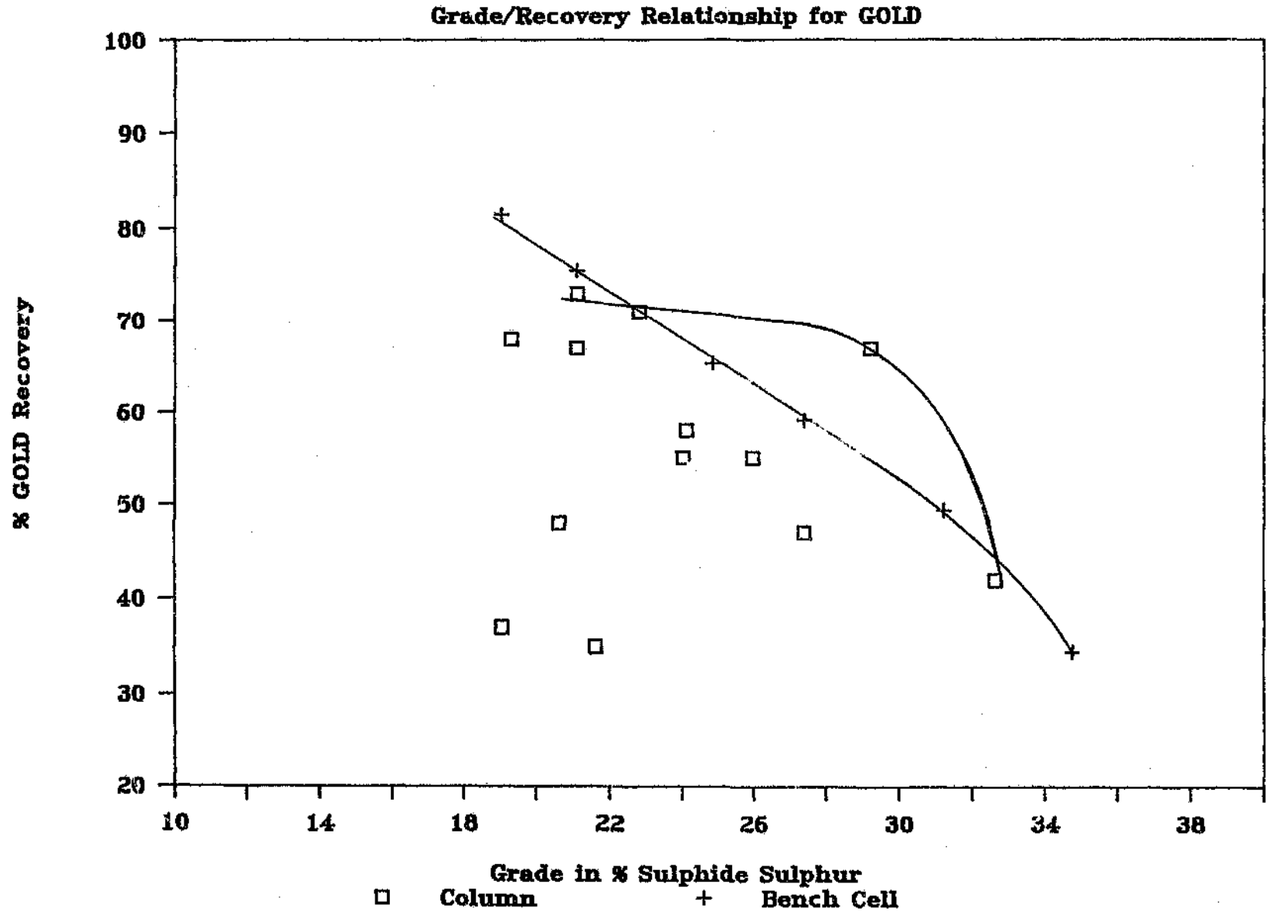
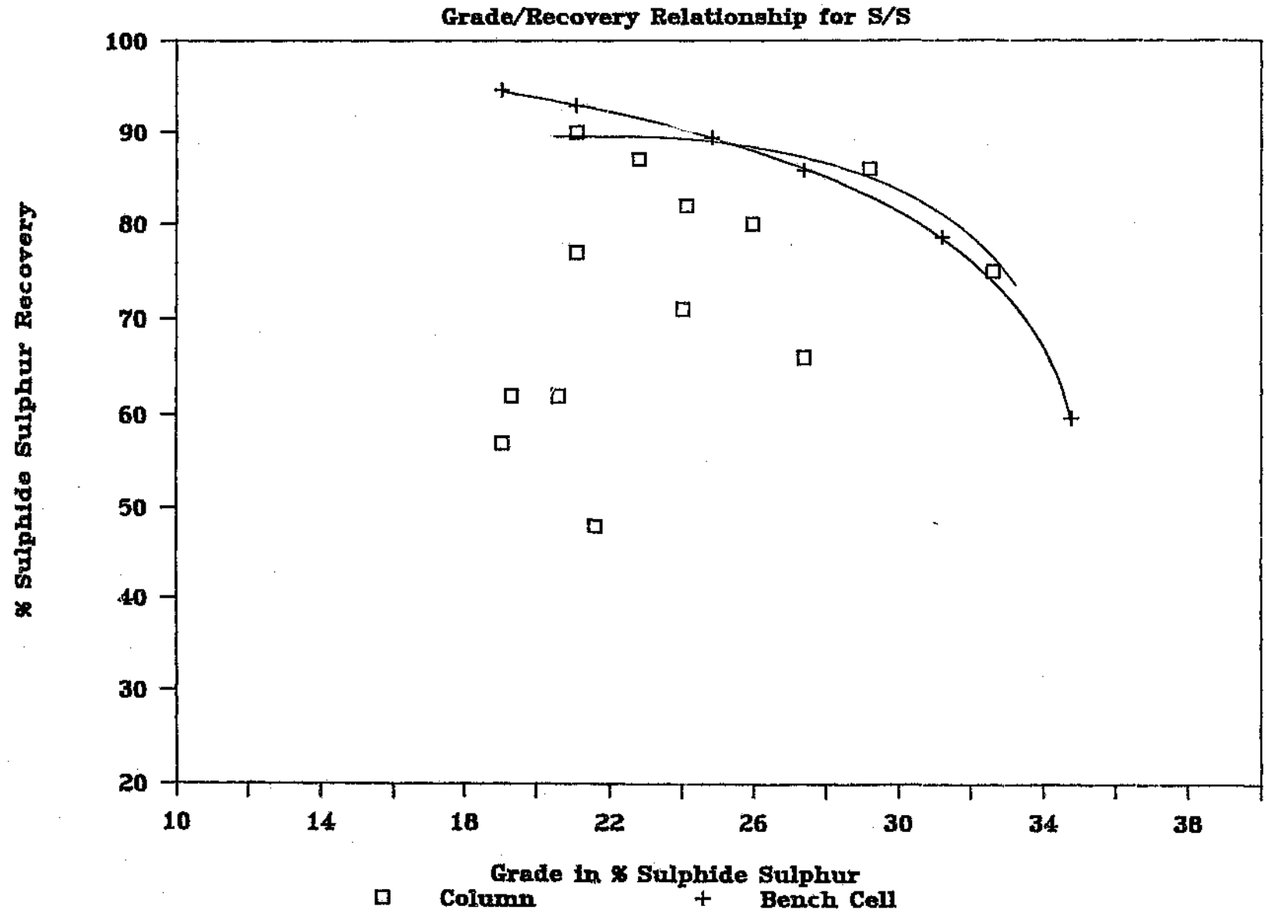


Figure 5.3

COMPARISON OF COLUMN AND BENCH CELL



obviously the best operating point for the column and the residence time for this test (JB1 in Table 5.1) was 20 minutes, which is in the upper range of residence times of flotation columns.

In order to appreciate the significance of the column flotation results for ERGO, it is important to understand the economics and mineralogy of the ERGO operation.

The sulphide sulphur has little economic value and is floated at ERGO not for itself, but because it contains gold. Gold is "locked" inside the pyrite particles. It is not known whether the gold is present in the pyrite in solid solution or as sub-micron-size particles, but it is liberated by roasting the pyrite (FeS_2) to form haematite (Fe_2O_3). Haematite is a porous mineral and a significant proportion of the "locked" gold is liberated on roasting (see ERGO flowsheet Figure 3.1). The concentrate grade should be at least 29 percent sulphide sulphur for efficient operation of the roaster and the objective of the ERGO flotation plant is to maximise gold recovery at a concentrate grade of 29 percent sulphide sulphur. If column flotation improved sulphide sulphur recovery alone, for ERGO it would have no economic advantages over conventional flotation. However, because higher gold recoveries were achieved at the same sulphide sulphur grades as conventional flotation, the column has the potential to significantly improve the profitability of the ERGO flotation plant, because higher gold recoveries at similar sulphide sulphur grades would increase the plant's revenue per ton of concentrate produced.

The reason for the improved gold recovery in column flotation may to some extent be explained by the gold deportment by size fraction. Table 5.2 summarises the gold and sulphide sulphur deportment in the plus and minus 75 micrometre fractions of the head, concentrate and tailings. The feed was fine, and 93 percent was minus 75 micrometres. The gold and sulphide sulphur minerals were significantly upgraded in the plus fraction of the feed. (The scavenger concentrate was the slower floating fraction of the plant feed and contained the finer and coarser material). The concentrate

Table 5.2

E R G O FLOTATION COLUMN

Test No. JB1

Screen Size um	HEAD Mass %	HEAD Gold g/t	HEAD S/S %	HEAD Gold %	HEAD S/S %	CONCENTRATE Mass %	CONCENTRATE Gold g/t	CONCENTRATE S/S %	CONCENTRATE Gold %	CONCENTRATE S/S %	TAILINGS Mass %	TAILINGS Gold g/t	TAILINGS S/S %	TAILINGS Gold %	TAILINGS S/S %
+75	7.40	9.82	21.70	22.72	30.24	22.00	18.40	27.30	31.28	22.13	6.30	1.45	0.26	5.61	1.39
-75	92.60	2.67	4.00	77.28	69.76	78.00	11.40	27.10	68.72	77.87	93.70	1.64	1.24	94.39	98.61
Overall	100.00	3.20	5.31	100.00	100.00	100.00	12.94	27.14	100.00	100.00	100.00	1.63	1.18	100.00	100.00

was much coarser than the feed at 78 percent minus 75 micrometres and the tailings was consequently finer than the feed. The plus fraction of the tailings had a lower gold grade compared to the finer fraction. Calculations indicate that while the gold recovery of the minus 75 micrometre fraction was 65 percent, the recovery of gold in the plus fraction was 95 percent. The ability of the flotation column to float the plus 75 micrometre particles was a significant result, as columns had earned the reputation of yielding superior results on fine cleaner feed materials. However a sampling campaign on the cleaner circuit conducted by plant personnel revealed that on average the plant cleaner concentrate was 88 percent minus 75 micrometres. The equivalent figure for the column testwork was 78 (section 5.1.4), indicating that the column was in this case better than the plant at floating coarse material. Thus the column would appear to be superior to conventional flotation for both the coarser and finer fractions which are the slow-floating fractions.

The superior performance of the column may also be due to its stable froth phase. Unpublished work done by Loveday and discussed in 1990 has shown that 90 percent of the collected material in conventional flotation is rejected in the froth phase. This high figure is due to the method of upgrading in conventional flotation of bubble coalescence, whereas upgrading in column flotation is achieved by froth washing and not coalescence. Therefore a particle attached to a bubble in a flotation column has a greater probability of being recovered, and froth phase recoveries in column flotation may be as high as 95 percent (Yianatos et al 1988). This has important implications for slow-floating fractions as a slow-floating particle (less hydrophobic) has a higher probability of rejection by coalescence.

The improved grade/recovery curve for the flotation column was also due to the column's upgrading obtained by washing the froth, because the feed water and its associated gangue minerals were displaced by wash-water which almost eliminated recovery of gangue by entrainment. As little as one percent of the feed water reports to the concentrate (Dobby and Finch 1986).

5.1.3 Column Flotation Results

The results are presented in Table 5.1 in chronological order. Tests CF1 and CF2 were excluded from parts of this discussion because their feed grades (21 percent sulphide sulphur and 8 g/t gold instead of the normal 6 percent sulphide sulphur and 3 or 4 g/t of gold) were so different, due perhaps to temporary disruption of the operation of the rougher cells, it was as if they had a different source.

The first column of results in Table 5.1 shows percentage mass pull, which varied between 8 and 31 percent (excluding tests CF1 and CF2). This variation was achieved by changing the gas and wash-water rates. An increase in either the gas rate or the bias ratio (a measure of washing in the column defined as volumetric tailings flowrate divided by volumetric feed flowrate) increased the mass pull.

Test BR3 had a low gas rate of 0,94 cm/sec, and a low bias ratio of 1,08 due to a low wash water flowrate of 3,7 ℓ/min, and, consequently, a low mass pull of 10 percent. Test TTL1, with a high gas rate of 1,24 cm/sec and a high bias ratio of 1,16 due to the higher wash water rate of 10,3 ℓ/min, had a mass pull of 23 percent. The higher mass pull yielded higher recoveries at lower grades.

The gold recoveries varied between 35 and 73 percent. Gold recovery was a function of many interacting variables: residence time, gas rate, wash water flowrate, bias ratio, carrying capacity etc. The results indicated that gold recovery increased with residence time (Test JB3) but that the optimum residence time for good efficiency was approximately 20 minutes (Test JB1). To achieve further gold recovery a second column could be installed in series.

The concentrate gold grades of Tests BR1, BR2, JB1 and JB2 were higher than those achieved in conventional flotation (Table 5.3) due to the method of upgrading used in column flotation.

Table 5.3

Conventional Laboratory Results

Test 3

Fraction	Mass		Sulphur				Gold				Uranium			
			Assay		Distribution		Assay		Distribution		Assay		Distribution	
	%	Cum. %	%	Cum. %	%	Cum. %	g/t	Cum. g/t	%	Cum. %	g/t	Cum. g/t	%	Cum. %
C1	7,3	7,3	34,75	34,75	59,7	59,7	12,10	12,10	34,5	34,5	514	514	17,8	17,8
C2	3,4	10,7	23,55	31,19	18,9	78,6	11,20	11,81	14,9	49,4	557	528	9,0	26,8
C3	2,6	13,3	11,74	27,39	7,2	85,8	9,53	11,37	9,7	59,1	603	542	7,4	34,2
C4	2,0	15,3	7,67	24,81	3,6	89,4	8,03	10,93	6,3	65,4	592	549	5,6	39,8
C5	3,4	18,7	4,35	21,09	3,5	92,9	7,61	10,33	10,1	75,5	621	562	10,0	49,8
C6	2,4	21,1	3,04	19,04	1,7	94,6	6,47	9,89	6,0	81,5	750	583	8,6	58,4
CT	78,9	100	0,29	4,25	5,4	100	0,60	2,56	18,5	100	111	211	41,6	100
ASSAY HEAD VALUES			4,33				3,19				212			

Pulp Conditions : pH = 3,96
14 % Solids

The gold tailings grades varied between 1,23 g/t (Test JB3) and 2,17 (Test TTL2) ignoring tests CF1 and CF2 due to their exceptionally high feed grades. This is a small variation for plant conditions and demonstrates how column flotation handles a wide variation of gold feed grades (2,9 to 6,4 g/t). In a month picked randomly ERGO's cleaner tailings' gold grades varied between 3,2 and 0,7 g/t.

The sulphide sulphur recoveries varied between 48 and 90 percent, and the sulphide sulphur grades varied between 21 and 33 percent. These variations were due to different operating conditions and varying feed grades. The sulphide sulphur tailings grades ranged from 2,2 to 0,74 percent. This was a larger range than the gold tailings range, and indicates that the column was more consistent in gold than sulphide sulphur recovery.

The residence time had a comprehensive range between 12 minutes and 43 minutes. The optimum residence time for this application was approximately 20 minutes.

Superficial gas velocity is one of the key parameters in column flotation. It was varied from 0,94 to 1,50 cm/sec. 1,5 cm/sec was higher than the optimum as it produced low concentrate grade (19 percent sulphide sulphur) and low gold recovery (37 percent). This result agrees with observations of the froth. Above a gas rate of approximately 1,2 cm/sec the froth-pulp interface underwent a transition from a thin plane to a thicker zone and increased mixing occured in the froth. At low gas velocities the froth behaved as an ordered bubble bed in a plug flow regime. At higher velocities the froth behaved in an agitated fashion. If gas rate is low, surface area flux is low, and recovery decreases. The optimum gas rate was found to be approximately 1,1 cm/sec (Test JB1). Above this, the concentrate grade dropped significantly without a corresponding increase in gold recovery.

The wash water rate and bias ratio may be discussed together. The wash water rate is simply the flowrate of water distributed to the top of the froth bed. The bias ratio is the volumetric ratio of

tailings to feed flowrate. Values greater than one indicate "a positive bias", and are achieved by the flow of wash-water increasing the down-flow through the column. As the wash-water flows downwards through the froth it displaces the feed water and entrained gangue from the concentrate. The column was operated at a positive bias, at wash-water rates varying from 3 to 10 ℓ/min, and the bias ratios varying from 1,1 to 1,8. The wash-water served two purposes. Firstly it stabilised the froth and made it more fluid, by acting as a lubricant between the bubbles, and preventing the bubble walls from thinning and the bubbles from coalescing. Secondly it washed the entrained gangue minerals from the froth. A high wash-water rate was thus required both for good recovery and good grades.

The importance of wash-water in stabilising the froth to produce high mass pulls was demonstrated in Test TTL1. The wash-water rate was set at 10 ℓ/min. The resulting mass pull was a high 23 percent. However, the bias ratio was about average at 1,16 because the high mass recovery also meant a high recovery of the wash water to the concentrate. So a high wash water rate may mean a lower bias rate due to the increase in mass pull. Test TTL3 was similar to test TTL1 (similar residence time and gas rate) but the wash-water rate was reduced from 10 to 8,5 ℓ/min. The mass recovery fell from 23 percent to 8 percent. (The high mass pull in CF2 with a wash-water rate of zero was due to the five-fold increase in feed grade).

The disadvantage of high mass pull was that grade tended to drop, and good wash-water rates and bias ratios were found to be in the region of 5 ℓ/min and 1,15 respectively.

Carrying capacity is defined as the maximum concentrate mass flowrate for a given material and flux of bubble surface area divided by the column cross-section. There is a limit to the concentrate mass flowrate and thus the carrying capacity in the column due to bubble surface limitations, but the limit is impossible to ascertain theoretically because not all the factors involved are yet known. However, it appears to be characteristic of a particular column application, and must be determined empirically. The carrying capacity for ERGO was taken as the maximum carrying capacity measured i.e. 1,85 g/min/cm^3.

The hold-up is the percentage of pulp displaced by air. The pressure gauge for the hold up measurement was fitted halfway through the testwork. The results ranged from 14 to 21 percent, the hold-up being proportional to gas rate (Table 5.1). Hold-up is also affected by slurry density, but the variation in slurry density was so small that this effect was not observed.Feed percent solids varied between 11 and 17 percent, and unfortunately this range was too narrow to enable conclusions to be drawn on the effect of feed percent solids on column performance.

The gold and sulphide sulphur feed grades varied considerably. Gold grade varied between 2,7 and 6,4 g/t and sulphide-sulphur between 3,8 and 7,3 percent. The ranges were even greater if tests CF1 and CF2 were taken into account.

Froth depth proved to be an important variable. The froth depth was deepest in the first four tests for which the sulphide-sulphur grades were remarkably low (approximately 20 percent). This result conflicted with observations by Yianatos et al (1988) who observed that concentrate grades increased with froth depth. Work done at Mintek has shown that the deeper froths have a higher drop-back of coarse pyrite (Ross 1989). This would seem to be the case at ERGO as the sulphide sulphur recovery tended to be higher with low froth heights.

The effect of reagent addition (see appendix II) was masked by froth recovery effects. Column performance was both good and poor with and without reagent addition, and no conclusions could be drawn on reagent addition from the results.

5.1.4 Statistical Analysis of Results

The data were interpreted in the previous sections on the basis of operating experience and visual observations. In this section a statistical analysis, using the computer package written by SAS Institute Inc. (1985), was performed to give a more objective view.

Definite conclusions could not be drawn from the analysis, due to the large number of variables associated with the ERGO column flotation pilot plant and the limited number of data points. Certain

variables, such as bubble size and reagent addition, were also not quantifiable in the commercial environment in which the testwork was done. Nevertheless, the analysis was useful for identifying trends.

An initial analysis by univariate statistics revealed no outliers among the data. Thereafter, regressions were done to investigate the effect of the independent on the dependent variables. The seven independent variables used were residence time, superficial gas velocity, wash-water flowrate, percent solids in the feed, feed gold grade, feed sulphide sulphur grade and froth height. The six dependent variables were gold recovery, gold concentrate grade, sulphide sulphur recovery, sulphide sulphur concentrate grade, bias ratio and carrying capacity. A regression was attempted for each dependent variable as a linear function of the seven independent variables.

Because of the small sample size, only residence time showed up as significant to the one percent level (Table 5.4). Next the ten percent significance level was used, along with the higher probability of false leads, to give pointers to potential areas of further research. On this basis the regressions for gold recovery and bias ratio revealed independent variables with statistical significance.

Gold Recovery Regression (Table 5.4)

Residence time was significant to the one percent level, with a coefficient (parameter estimate) of 1,3 $\pm$ 0,66 (i.e. 1,94 x std error) with 95 percent confidence. Thus a change in residence time from 20 to 25 minutes can be predicted with 95 percent confidence to improve gold recovery by 6,5 $\pm$ 3,3 percentage points.

The coefficient for superficial gas velocity was negative and large: -40 $\pm$ 36 to 95 percent confidence. This suggests that an increase in superficial gas velocity of 1,4 to 1,5 would decrease gold recovery 4 $\pm$ 3,6 percent. However, this result was influenced by the particularly high gas rates and low gold recoveries in tests

Table 5.4

REGRESSION RESULTS FOR GOLD RECOVERY

Model: MODEL1
Dependent Variable: GREC

Analysis of Variance

Source	DF	Sum of Squares	Mean Square	F Value	Prob>F
Model	7	1939.80073	277.11439	5.951	0.0225
Error	6	279.41355	46.56893		
C Total	13	2219.21429			

Root MSE	6.82414	R-square	0.8741
Dep Mean	54.64286	Adj R-sq	0.7272
C.V.	12.48863		

Parameter Estimates

Variable	DF	Parameter Estimate	Standard Error	T for H0: Parameter=0	Prob > |T|
INTERCEP	1	114.361157	27.55037001	4.151	0.0060
RESTIME	1	1.291329	0.32851288	3.931	0.0077
GASV	1	-40.497974	18.11798707	-2.235	0.0668
WWRATIO	1	5.370226	2.08983674	2.570	0.0424
FEEDSOLS	1	-2.593271	1.19156976	-2.176	0.0724
FEEDGG	1	-2.672587	2.24633925	-1.190	0.2791
FEEDSSG	1	-0.901927	0.92510637	-0.975	0.3672
FROTH	1	-0.037440	0.01591339	-2.353	0.0568

TTL3, TTL4 and CF1 (see Figure 5.4). The suggestion was made in section 5.1.3 that recovery improved with increasing gas velocity up to a maximum value, above which the froth became turbulent and recovery dropped. The regression was unable to account for this non-linear behaviour.

The wash water rate coefficient was large and positive. An increase of 6 to 8 in wash water rate is predicted to increase gold recovery by 11 $\pm$ 8 percentage points to 95 percent confidence which suggested a froth stabilising effect of wash water (see section 5.1.3).

The feed percent solids coefficient was negative, and an increase from 13 to 16 percent is predicted to reduce the gold recovery by 8 $\pm$ 7 percentage points to 95 percent confidence. A similar trend has been observed in conventional flotation.

The coefficients for the feed grades were not significant.

The froth height coefficient was negative, giving the prediction that a decrease in froth height from 800 to 400 millimetres would increase gold recovery 15 $\pm$ 13 percentage points with 95 percent confidence. This trend agreed with observations noted in section 5.1.3.

Bias Ratio Regression (Table 5.5)

This is an interesting regression, highlighting variables important for good washing in the column.

The residence time coefficient was positive for the prediction of bias ratio, and was strongly influenced by tests JB3 and JB4, which had long residence times and high bias ratios. Although the concentrate flowrate decreased with increasing residence time (due to lower feedrates), the wash water flowrate often remained constant and hence the bias ratio in the column tended to increase with residence time.

Figure 5.4

STATISTICAL ANALYSIS

Plot of Gold Recovery Vs Gas Velocity.

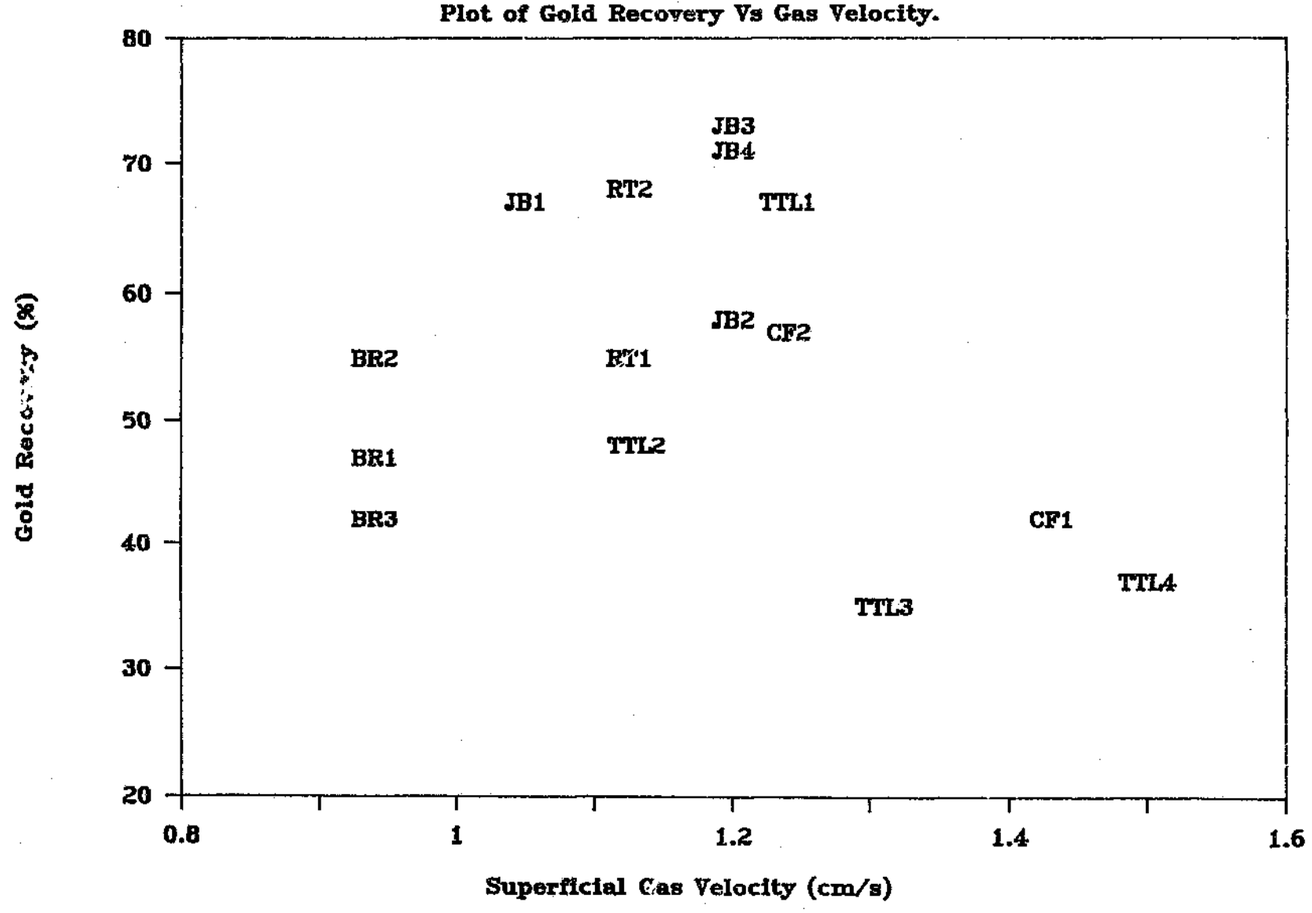


Table 5.5

REGRESSION RESULTS FOR BIAS RATIO

Model: MODEL5
Dependent Variable: BIASRAT

Analysis of Variance

Source	DF	Sum of Squares	Mean Square	F Value	Prob>F
Model	7	0.71692	0.10242	17.410	0.0014
Error	6	0.03530	0.00588		
C Total	13	0.75221			

Root MSE	0.07670	R-square	0.9531
Dep Mean	1.21393	Adj R-sq	0.8983
C.V.	6.31819		

Parameter Estimates

Variable	DF	Parameter Estimate	Standard Error	T for H0: Parameter=0	Prob > |T|
INTERCEP	1	0.733649	0.30964568	2.369	0.0556
RESTIME	1	0.028670	0.00369224	7.765	0.0002
GASV	1	-0.487329	0.20363271	-2.393	0.0538
WWRATIO	1	0.066732	0.02348821	2.841	0.0295
FEEDSOLS	1	0.010655	0.01339236	0.796	0.4566
FEEDGG	1	0.020494	0.02524718	0.812	0.4479
FEEDSSG	1	-0.035822	0.01039751	-3.445	0.0137
FROTH	1	0.000045497	0.00017885	0.254	0.8077

Gas velocity had a negative coefficient. As it increased, so the concentrate flowrate increased and the washing in the column dropped off.

Wash water rate had a positive coefficient, and an increase in wash water flowrate of 4 to 8 l/min is predicted to increase the bias ratio by 0,27 $\pm$ 0,18 to a 95 percent confidence limit.

The only other variable identified as significant to bias ratio was the feed sulphide sulphur grade. The coefficient was negative and largely influenced by tests CF1 and CF2 which had very high sulphide sulphur feed grades and a correspondingly high mass pull. The high mass pull resulted in low washing efficiency.

5.1.5 Routine Conventional Cell Laboratory Test Results

The recovery versus time curve is shown in Figure 5.5 for test number 3. The curve is typical for conventional flotation of Witwatersrand gold ore. The sulphide-sulphur is fast-floating and its recovery curve flattened off after 20 minutes of flotation. Gold flotation being slower, had not been completed at the end of the 40 minute test. The concentrate sulphide-sulphur grade was low at the end of the test i.e. 19 percent.

The results for the four tests conducted are tabulated in Tables 5.6, 5.7, 5.3 and 5.8 respectively. The first two tests were repeated due to a low sample pH of 1,56. Tests 2 and 4 (Tables 5.3 and 5.8) were conducted with the addition of reagent. The reagent was added as deemed necessary to improve the float. A comparison of test 1 and 2 (low pH) and tests 3 and 4 (correct pH) revealed the importance of good pH control for flotation. The low pH had a deleterious effect on the concentrate sulphide sulphur grades, and adversely affected the gold recovery, although not the sulphide sulphur recovery.

The low pH decreased the rate of flotation of both gold and sulphide sulphur. The ERGO management have since improved the pH control system at the flotation plant.

Figure 5.5

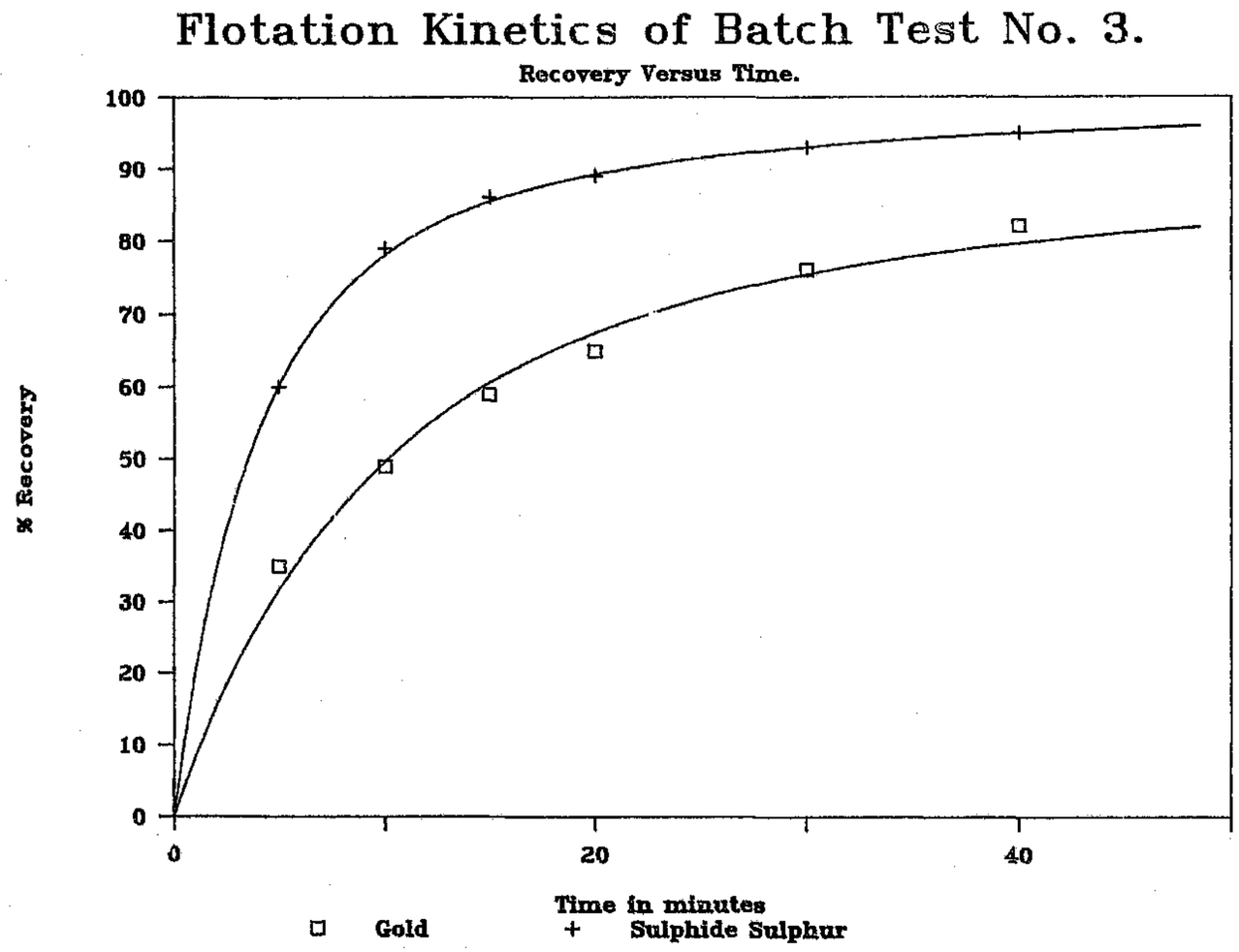

Flotation Kinetics of Batch Test No. 3.
Recovery Versus Time.
100
90
80
70
60
50
40
30
20
10
0
% Recovery
0
20
40
Time in minutes
Gold
Sulphide Sulphur

Table 5.6

Conventional Laboratory Results

Test 1

Fraction	Mass		Sulphur				Gold				Uranium			
			Assay		Distribution		Assay		Distribution		Assay		Distribution	
	%	Cum. %	%	Cum. %	%	Cum. %	g/t	Cum. g/t	%	Cum. %	g/t	Cum. g/t	%	Cum. %
C1	6,6	6,6	29,20	29,20	53,6	53,6	11,75	11,75	29,9	29,9	378	378	23,1	23,1
C2	3,6	10,2	19,30	25,71	19,3	72,9	10,01	11,14	13,9	43,8	215	320	7,2	30,3
C3	2,7	12,9	11,00	22,63	8,3	81,2	8,49	10,58	8,9	52,7	222	300	5,5	35,8
C4	2,5	15,4	7,60	20,22	5,4	86,6	7,01	10,00	6,8	59,5	246	291	5,7	41,5
C5	3,9	19,3	4,86	17,12	5,3	91,9	5,73	9,14	8,6	68,1	257	284	9,3	50,8
C6	3,1	22,4	2,65	15,11	2,3	94,2	4,59	8,51	5,5	73,6	214	275	6,1	56,9
CT	77,6	100	0,27	3,60	5,8	100	0,88	2,59	26,4	100	60	108	43,1	100
ASSAY HEAD VALUES			3,33				2,46				79			

Pulp Conditions : pH = 1.56
13 % Solids

Table 5.7

Conventional Laboratory Results

Test 2

Fraction	Mass		Sulphur				Gold				Uranium			
			Assay		Distribution		Assay		Distribution		Assay		Distribution	
	%	Cum. %	%	Cum. %	%	Cum. %	g/t	Cum. g/t	%	Cum. %	g/t	Cum. g/t	%	Cum. %
C1	8,4	8,4	24,20	24,20	53,0	53,0	10,60	10,60	32,1	32,1	326	326	27,3	27,3
C2	5,2	13,6	16,20	21,14	21,9	74,9	9,34	10,12	17,5	49,6	176	269	9,1	36,4
C3	3,9	17,5	10,20	18,70	10,4	85,3	7,39	9,51	10,4	60,0	208	255	8,1	44,5
C4	4,3	21,8	5,49	16,10	6,1	91,4	5,66	8,75	8,8	68,8	235	251	10,0	54,5
C5	4,0	25,8	3,42	14,13	3,6	95,0	5,01	8,17	7,2	76,0	208	244	8,3	62,8
C6	3,3	29,1	1,73	12,72	1,5	96,5	3,42	7,63	4,1	80,1	166	236	5,4	68,2
CT	70,9	100	0,19	3,84	3,5	100	0,78	2,77	19,9	100	45	100	31,8	100
ASSAY HEAD VALUES			3,33				2,46				79			

Pulp Conditions : pH = 1,56
13 % Solids

Table 5.8

Conventional Laboratory Results

Test 4

Fraction	Mass		Sulphur				Gold				Uranium			
			Assay		Distribution		Assay		Distribution		Assay		Distribution	
	%	Cum. %	%	Cum. %	%	Cum. %	g/t	Cum. g/t	%	Cum. %	g/t	Cum. g/t	%	Cum. %
C1	9,4	9,4	31,40	31,40	65,5	65,5	13,70	13,70	45,1	45,1	786	786	31,5	31,5
C2	4,4	13,8	17,70	27,03	17,3	82,8	10,85	12,79	16,7	61,8	621	733	11,6	43,1
C3	3,2	17,0	9,46	23,72	6,7	89,5	8,42	11,97	9,5	71,3	629	714	8,6	51,7
C4	3,3	20,3	5,11	20,70	3,8	93,3	6,56	11,09	7,6	78,9	544	686	7,6	59,3
C5	3,3	23,6	3,47	18,29	2,6	95,8	5,58	10,32	6,5	85,4	458	654	6,4	65,7
C6	2,0	25,6	2,50	17,06	1,1	96,9	4,16	9,84	2,9	88,3	372	632	3,2	68,9
CT	74,4	100	0,19	4,51	3,1	100	0,45	2,85	11,7	100	98	235	31,1	100
ASSAY HEAD VALUES			4,33				3,19				212			

Pulp Conditions : pH = 3.96
14 % Solids

The addition of reagent significantly improved the final recovery of gold from 82 to 88 percent. The sulphide sulphur recovery improved from 95 to 97 percent. However, the increase in the flotation rates is of greater significance. Because rate of gold flotation is slow, the flotation of gold is generally not completed on a plant. Therefore, if the rate of flotation of gold is increased the gold recovery in the plant will increase for the same residence time.

5.1.6 Release Analysis Results

The release analysis was conducted to establish the optimum grade/recovery curve for ERGO scavenger concentrate. This curve was plotted in Figure 5.6. The grade/recovery curves for the column flotation pilot plant and the conventional cell bench tests were included for comparison purposes. The release analysis indicated that the performance of the column was close to optimum in the operating range of 26 and 30 percent sulphide sulphur.

The release analysis gold results were far from satisfactory (Table 5.9). The procedure for the analysis relied on visually detecting the flotation end point. While this was straightforward for pyrite flotation it was impossible for gold flotation because the pyrite is visible, but not the gold. Gold flotation was slower and was not completed when the pyrite flotation end point was detected.

5.2 ACCURACY OF REPORTED DATA

The large number of variables in column flotation make the reproduction of a test difficult, and when columns are operated on a widely fluctuating feed reproducibility becomes even more difficult to check.

However, tests JB3 and JB4 were similar in all respects and both achieved similar recoveries. The gold recoveries were 73 and 71 percent and the sulphide sulphur recoveries were 90 and 87 percent respectively.

Figure 5.6

COMPARISON OF COLUMN AND BENCH CELL

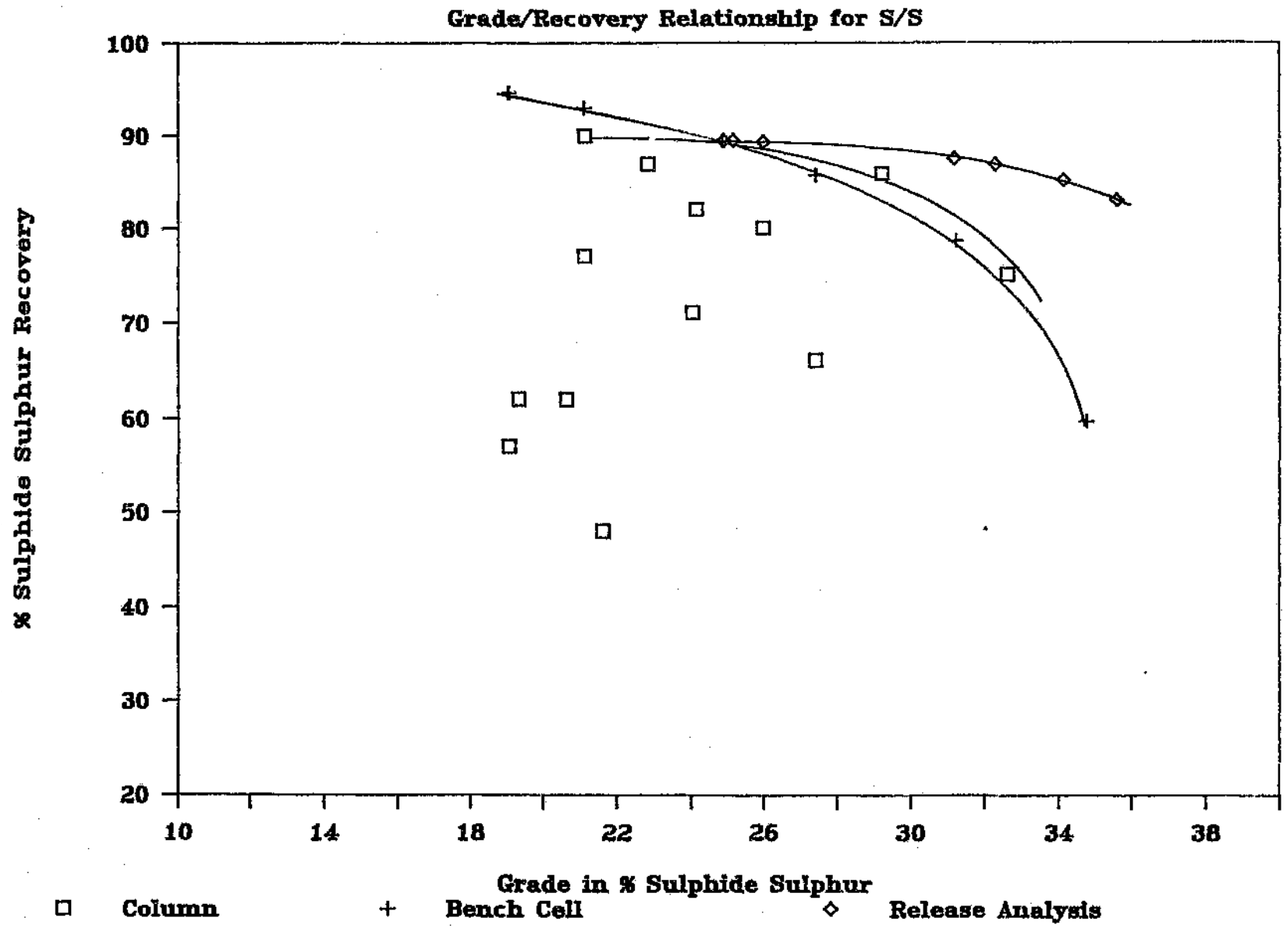


Table 5.9

COLUMN FLOTATION (R/88/322)

Release Analysis for ERGO Cleaner Feed.

	MASS		GOLD				SULPHIDE SULPHUR				PYRITE				GANGUE			
	Dist. %	Cum. Dist. %	Assay g/t	Cum. Assay g/t	Dist. %	Cum. Dist. %	Assay %	Cum. Assay %	Dist. %	Cum. Dist. %	Assay %	Cum. Assay %	Dist. %	Cum. Dist. %	Assay %	Cum. Assay %	Dist. %	Cum. Dist. %
Plant Cleaner Feed Head Sample	100.0%	100.0%	4.67	4.67	100.0%	100.0%	7.93	7.93	100.0%	100.0%	14.96	14.96	100.0%	100.0%	85.04	85.04	100.0%	100.0%
First Recleaning																		
Concentrate	28.5%	28.5%	7.03	7.03	43.0%	43.0%	24.85	24.85	89.5%	89.5%	46.89	46.89	89.5%	89.5%	53.11	53.11	17.8%	17.8%
Tailings	71.5%	100.0%	3.73	4.67	57.0%	100.0%	1.17	7.93	10.5%	100.0%	2.21	14.96	10.5%	100.0%	97.79	85.04	82.2%	100.0%
Final Recleaning																		
C1	10.0%	10.0%	7.04	7.04	15.2%	15.2%	43.90	43.90	55.7%	55.7%	82.83	82.83	55.7%	55.7%	17.17	17.17	2.0%	2.0%
C2	4.4%	14.5%	9.40	7.76	8.9%	24.1%	32.40	40.30	18.1%	73.8%	61.13	76.19	18.1%	73.8%	38.87	23.81	2.0%	4.1%
C3	4.0%	18.4%	9.78	8.20	8.4%	32.4%	18.20	35.58	9.2%	83.0%	34.34	67.14	9.2%	83.0%	65.66	32.86	3.1%	7.1%
C4	1.3%	19.8%	9.05	8.25	2.5%	35.0%	13.40	34.12	2.2%	85.2%	25.28	64.37	2.2%	85.2%	74.72	35.63	1.2%	8.3%
C5	1.6%	21.3%	7.64	8.21	2.6%	37.5%	8.70	32.26	1.7%	86.9%	16.42	60.86	1.7%	86.9%	83.58	39.14	1.5%	9.8%
C6	0.9%	22.3%	6.22	8.13	1.2%	38.8%	5.50	31.13	0.6%	87.6%	10.38	58.74	0.6%	87.6%	89.62	41.26	1.0%	10.8%
C7	5.0%	27.2%	3.54	7.29	3.8%	42.6%	2.81	25.95	1.8%	89.3%	5.30	48.96	1.8%	89.3%	94.70	51.04	5.6%	16.4%
C8	0.9%	28.2%	1.79	7.10	0.4%	42.9%	0.99	25.12	0.1%	89.4%	1.87	47.40	0.1%	89.4%	98.13	52.60	1.1%	17.5%
Tailings	0.3%	28.5%	0.55	7.03	0.0%	43.0%	0.38	24.85	0.0%	89.5%	0.72	46.89	0.0%	89.5%	99.28	53.11	0.4%	17.8%
Recleaning	28.5%	28.5%	7.03	7.03	43.0%	43.0%	24.85	24.85	89.5%	89.5%	46.89	46.89	89.5%	89.5%	53.11	53.11	17.8%	17.8%

It is also difficult to evaluate the accuracy of the data by mass balances because of the continuous variation in feed density. The volumetric balance is the best measure of accuracy as it will only vary due to changes in feed viscosity. (The feed flowed by gravity from the feed sampler to the column and a change in viscosity would affect the flowrate). Table 5.10 shows the mass and volumetric balances for each test. The "IN" (measured) and "OUT" (calculated) volumetric flowrates agree, the average deviation being four percent and the maximum being 12 percent (test BR2). The average values for the IN and OUT volumetric flowrates are in close agreement, being 27,4 and 27,6 respectively. The mass flowrate pairs vary more widely due to variations in feed densities. The volumetric balances for some tests are perfect but the mass balance is poor e.g. test JB2. However, the average mass flowrate figures for the In and the Out agree and are 2,82 and 2,92 respectively.

It should be noted that feed samples were taken directly after the test unlike the concentrate and tailings samples, which were taken during the test. This was done because sampling the feed interrupted the flow of slurry to the column, thus interfering with the column's performance during the test. The feed line also tended to block when being sampled.

Table 5.11 compares the actual feed gold and sulphide sulphur grades, with the feed grades calculated from the tailings and concentrate assays. The pairs of results differ widely, indicating that the feed grades varied continuously, though the sulphide sulphur grades varied less than the gold. The average of the actual (IN) grades was higher than the average of the calculated (OUT) grades. The average of the mass and volumetric flowrates IN were less than the OUT; an opposite trend.

In some tests e.g. TTL1 the calculated gold grade was greater than the actual gold grade, while the calculated sulphide sulphur grade was less than the actual sulphide sulphur grade. These observations suggested that the differences in the grades and balances were due to random fluctuations and not a systematic error in the testwork. The average actual sulphide sulphur grade agreed

Table 5.10

Mass and Volumetric Balances

Test	Mass Balance kg/min		Volumetric Balance ℓ/min	
	In	Out	In	Out
TTL1	3,1	3,5	36	37
TTL2	2,9	3,2	32	32
TTL3	3,7	4,2	34	36
TTL4	3,8	4,1	35	36
BR1	3,6	4,2	29	30
BR2	4,0	3,7	29	26
BR3	4,3	4,5	30	30
CF1	3,9	3,6	31	31
CF2	-	3,5*	-	22*
RT1	2,7	2,9	30	32
RT2	2,9	2,4	32	32
JB1	1,9	1,7	22	22
JB2	1,3	1,8	20	20
JB3	0,73	0,53	12	11
JB4	0,68	0,59	12	11
$\bar{x}$	2,82	2,92	27,4	27,6
σ_{n}	1,17	1,28	7,59	8,25
σ_{n-1}	1,22	1,32	7,88	8,56
n	14	14	14	14

* Not included in averages.

Table 5.11

Actual and Calculated Feed Grades

Test No	Gold Grades g/t		Sulphide Sulphur %	
	Actual (In)	Calculated (Out)	Actual (In)	Calculated (Out)
TTL1	2,9	3,7	7,3	6,5
TTL2	4,7	3,6	5,6	4,9
TTL3	2,7	2,4	3,4	3,5
TTL4	3,8	3,0	4,0	2,8
BR1	3,7	3,5	4,6	4,2
BR2	3,5	5,3	4,7	3,7
BR3	5,3	4,5	4,8	4,5
CF1*	7,8	6,4	20,8	15,1
CF2*	N/A	7,9	N/A	19,9
RT1	6,4	3,9	4,2	6,0
RT2	4,6	3,4	3,8	5,0
JB1	3,4	3,7	4,9	6,3
JB2	3,3	2,7	5,6	3,6
JB3	3,1	3,1	5,0	7,3
JB4	3,6	3,3	5,2	6,9
$\bar{x}$	4,20	3,75	5,99	5,74
σ	1,40	1,01	4,21	2,93
σ_{n-1}	1,45	1,05	4,37	3,04
n	14	14	14	14

* Not included in averages.

with the average calculated grade, but the agreement was not as good for the gold grade averages being 4,20 and 3,75 for actual and calculated grades respectively, which represents a difference of 12 percent.

5.3 PILOT PLANT OPERATION

5.3.1 Visual Assessment of Performance

Like conventional flotation, the operation of a flotation column may be assessed visually. At the beginning of each test run one or two parameters would be set to predetermined values, and the others would be adjusted to give the best operation, determined visually. The procedure used is described below.

Gas Rate

At low gas rates the bubbles behave in an orderly fashion and rise in a plug to overflow at the concentrate lip. The optimum gas rate was found by increasing the gas rate until the froth/pulp interface expanded from a thin distinct line to a thick blurred line. The gas rate was than reduced slightly so that the froth/pulp interface was a fine line. This visual practice was based on the idea that increasing the gas rate increases gold recovery until turbulence and axial mixing in the froth caused a loss of valuable mineral. This tendency was confirmed in the statistical analysis.

Wash Water Flowrate

Moys (1978) states "well-drained froths are generally unstable and do not flow easily". The wash water flowrate used was the minimum flowrate required to make the froth stable enough to eliminate coalescence. This froth stability was achieved by separating the individual bubbles with interstitial wash water, which made the froth more bulky and less viscous. The additional volume of the water in the froth zone reduced the bubble residence time in the froth phase. Coalescence was a function of the residence time of the bubbles in the froth phase because a long bubble residence time increases the probablity of coalescence.

Froth Height

The froth height was set at the minimum height to keep the bubble residence time low, while still allowing the entrained feed water to be washed out. Moys (1984) demonstrated with conventional flotation that froth stability was increased in shallow froths with high gas rates. This corresponds to a low residence time in the froth phase.

Wash Water Distributor Depth

The wash water stabilises the froth below the distributor. Above the distributor conventional froth, with froth drainage and coalescence, occurs. The wash water distributor was set below the froth at all times, and its depth was adjustable. The distributor was raised when too much bubble coalescence was observed. It was lowered when the concentrate grade was too low to allow just enough coalescence for the grade to improve.

5.3.2 Operation of Pilot Plant

In general the pilot plant operation was satisfactory. It appeared to be possible to operate the column continuously for indefinite periods of time, although the longest period it was operated was a working day.

The pilot plant required regular maintenance. The bubble injector was cleaned with hydrochloric acid and a rifle barrel cleaning brush. The water filters for the bubble generator and wash water were cleaned. The wash water distributor holes were checked for blockages and washed with hydrochloric acid. The manometer lines were washed out with fresh water before each reading.

The Bredel pumps performed reliably; the sampling system proved to be simple and effective; the three interchangeable diameters of pipe were essential for differing feed flowrates. The control system was reliable and effective, although the detection of the froth/pulp

interface level with the differential pressure cell had a minor drawback. When operating variables such as gas rate and wash water rate were changed, the bulk density of the froth, and hence the level of the froth/pulp interface, also changed. When these variables were set the level was controlled adequately.

The only other problem encountered with the pilot plant was the tendency of the line between the feed sampler and the column to block. The blocking was due to the build-up of solids at the bottom of the sampler because of a low hydrostatic head between the sampler and the column. The bottom of the sampler was a 30° cone which fed the feed line, but this angle should have been greater to avoid the blockages. To facilitate unblocking of the line without a shutdown, air and water lines were fitted to the slurry line. If the line blocked during a test the samples were discarded.

The wash water steady-head tank operated satisfactorily and the Hanna pH controller was reliable.

5.4 CARRYING CAPACITY AND FLOTATION KINETICS

5.4.1 Discrimination between Carrying Capacity and Flotation Kinetic Limitations

It is important to distinguish, when studying the performance of a flotation column, between the results limited by carrying capacity and the results limited by flotation kinetics. In conventional flotation, bubble carrying capacity is rarely a limitation. The open area of a column is approximately a fifth that of the conventional cell, and bubble surface area is often the parameter limiting a column's performance.

5.4.2 Carrying Capacity

A semi-empirical relationship is discussed in Section 2.5.1 for the estimation of carrying capacity.

$$C_a = \frac{60\pi K_1 \; d_p \rho_p J_g}{d_b} \qquad \text{Equation 2.5.1}$$

In this testwork the average bubble size at the top of the froth d_b was kept constant, so the equation may be simplified to

$$C_a = \text{Const.}\; d_p \rho_p J_g.$$

This is a linear equation and the straight line passes through the origin. The best measure of d_p (mean particle diameter) is discussed in depth in Section 2.5. It is suggested that the best mean diameter for bubble loading is the diameter of a sphere whose cross-sectional area is the average cross-sectional area of the particles i.e. the $d_{(3,2)}$, while the diameter used in the literature is the d_{80} and is equal to the size of aperture through which 80 percent of the distribution passes.

Table 5.12 summarises the relevant information. The right hand side of Equation 5.4.1 was calculated for each test, using both measures of particle diameters. The data is plotted in Figure 5.7 using the $d_{(3,2)}$, and Figure 5.8 using the d_{80}. On the assumption that at least one or two tests in the upper range of the graphs were carrying capacity limited, a line was drawn separating these from the rest of the data. The lines divide the results into two sets: the carrying capacity limited results on the left hand side and the flotation kinetic limited results on the right hand side.

Although CF1 and CF2 clearly have the highest concentrate mass flux, and were obviously limited by carrying capacity, they were excluded from consideration. The column was operating under very different conditions with this material than it was in the other tests. The very high quantity of hydrophobic material made the froth exceptionally stable and fluid, and the bubble residence time in the froth low. While no wash water was used in CF2, it had a higher concnetrate mass flux than CF1, whereas with the usual feed froth without wash water would be unstable and have very low concentrate mass flux and recovery. So the carrying capacities of

Table 5.12

Concentrate Mass Flux and Particle Size

Test No	Conc. Mass Flux g/min/cm^2	$d_{(3,2)}$ um	d_{80} um	J_g cm/sec	Particle Density g/cm^3	$d_p \rho_p J_g$ / $d_{(3,2)}$	$d_p \rho_p J_g$ / d_{80}
TTL1	1,85	12,6	72,8	1,24	3,38	52,8	305
TTL2	1,06	10,5	57,3	1,13	3,38	40,1	219
TTL3	0,73	8,4	37,9	1,31	3,38	37,2	169
TTL4	0,77	7,5	27,6	1,50	3,38	38,0	140
BR1	0,79	11,7	73,7	0,94	3,65	40,1	253
BR2	1,36	12,5	71,9	0,94	3,58	42,1	242
BR3	1,05	13,0	61,1	0,94	3,79	46,3	218
RT1	1,16	12,6	76,1	1,13	3,52	50,1	303
RT2	0,86	11,5	77,1	1,13	3,38	43,9	294
JB1	0,71	11,9	68,8	1,05	3,70	46,2	267
JB2	0,48	7,7	35,9	1,20	3,52	32,5	152
JB3	0,37	8,5	45,6	1,20	3,40	34,7	186
JB4	0,34	10,2	42,6	1,20	3,50	42,8	179

Figure 5.7

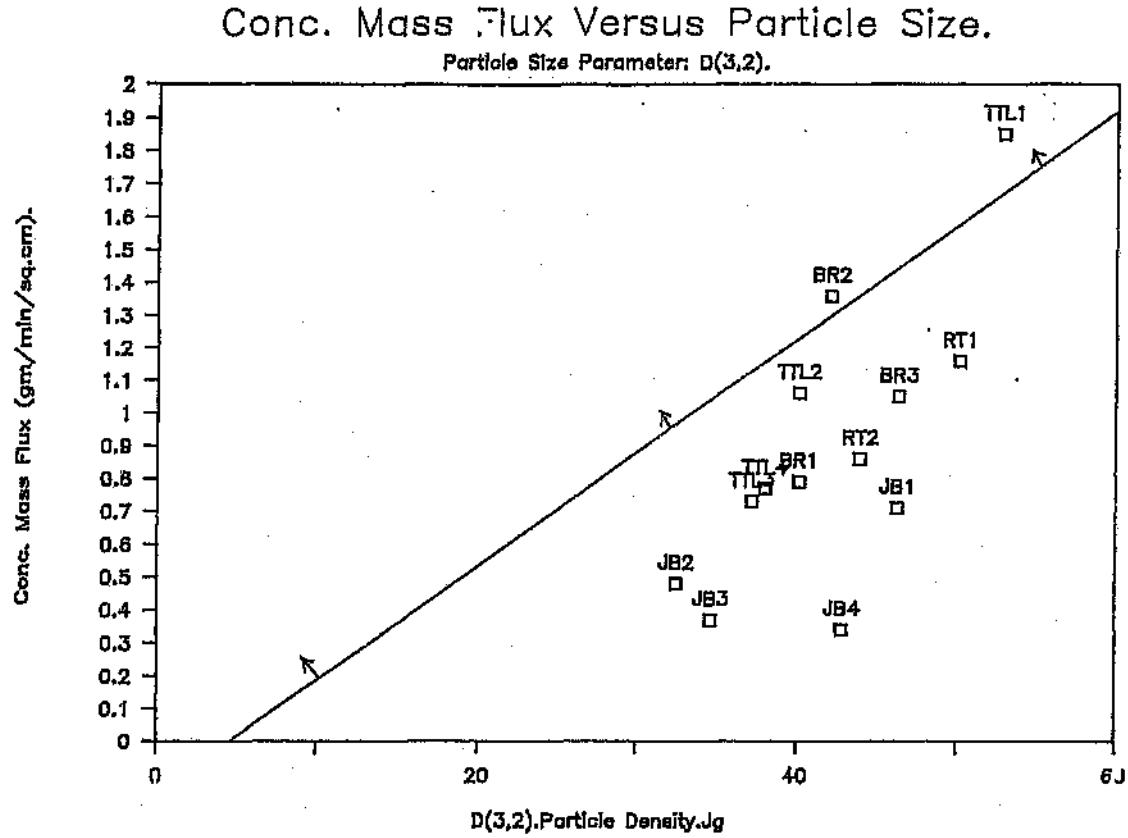
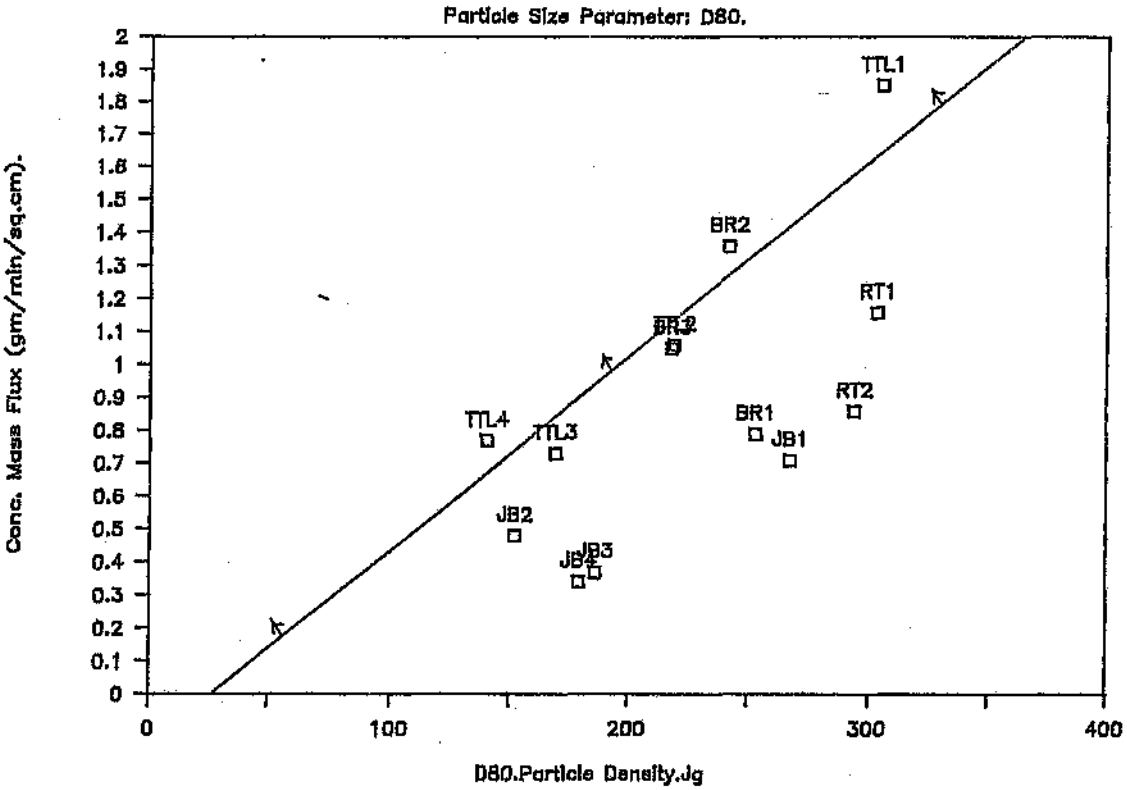


Figure 5.8



these tests were not considered true of the column under normal conditions.

Both measurements of particle size segregated the individual test results, except for test TTL4, into the same categories i.e. the carrying capacity limited category. TTL1 had the highest concentrate mass flux 1,85 g/min/cm^2, after the CF1 and CF2 tests, probably due in part to a slightly higher sulphide sulphur feed grade of 7,3 percent. BR2 was probably also limited by carrying capacity, as it had the next highest concentrate mass flux of 1,36 g/min/cm^2, although it had a normal sulphide sulphur feed grade of 4,7 percent. It is unlikely that TTL4 was limited by carrying capacity, as it had a low concentrate mass flux of 0,77 g/min/cm^2. The reason it was identified by the d_{80} analysis as being carrying capacity limited was that the d_{80} indicated it as very fine, finer than the $d_{(3,2)}$ indicated it to be.

The best straight line fit through the origin was determined by linear regression for each measure of diameter (Figures 5.9 and 5.10). The $d_{(3,2)}$, however, is a better measure of mean particle diameter for carrying capacity, and was therefore the measurement used in this discussion.

The following relationship is proposed for the estimation of carrying capacity for Witwatersrand gold ores:

$$C_a = 0,034\ d_{(3,2)}\ \rho_p\ J_g \qquad \text{Equation 5.4.2}$$

The bubble diameter is not included in this relationship as it was not measured. From visual estimation of the bubble size the diameter appeared to be two millimetres at the top of the froth.

The empirical equation for carrying capacity based on operating data (Section 2.5.1) estimates the carrying capacity for ERGO scavenger concentrate to be:

Figure 5.9

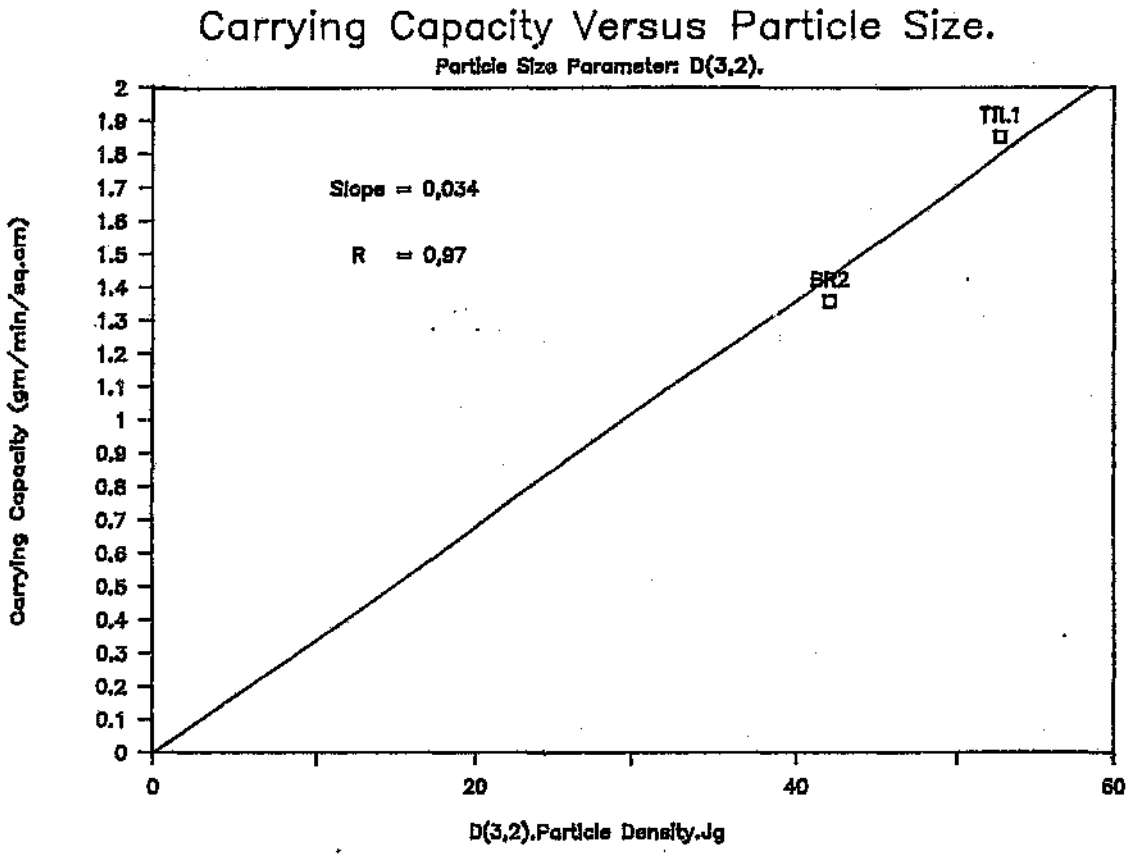
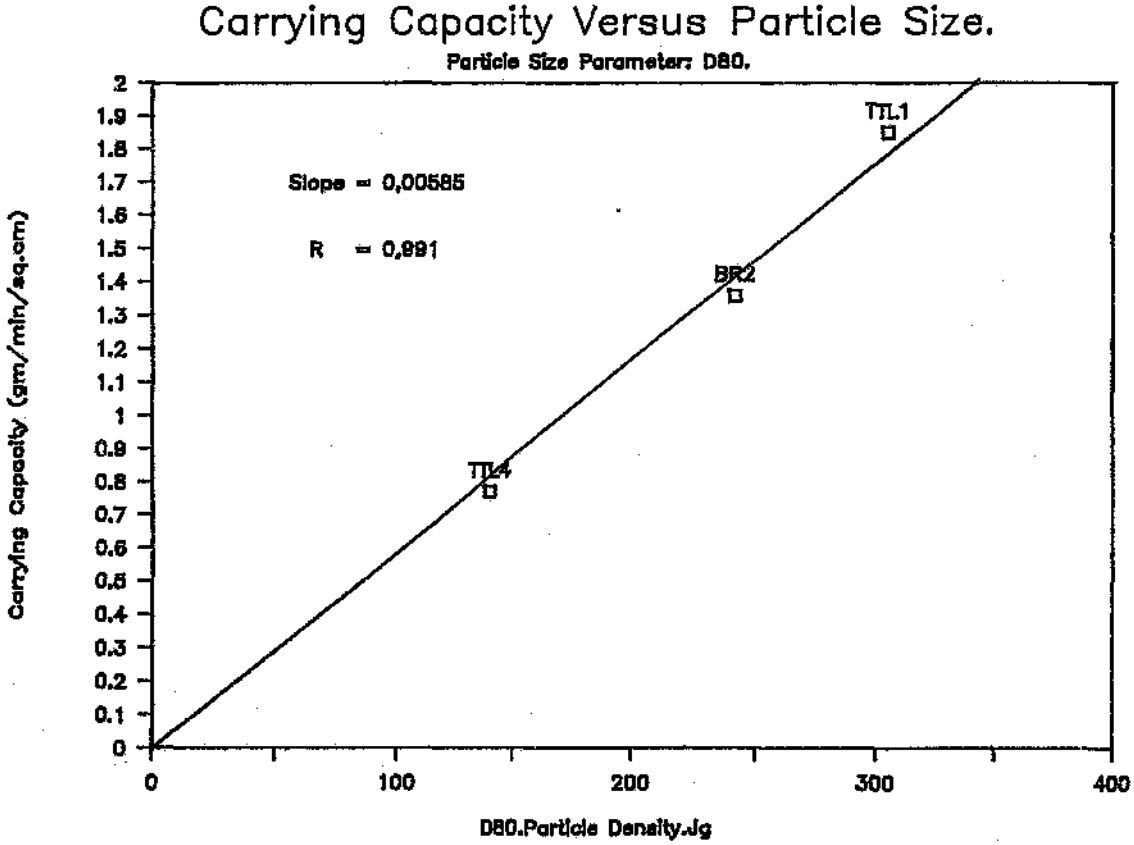


Figure 5.10



$$C_a = 0,0682\ (d_{80} \cdot \rho_p) \qquad \text{Equation 2.5.2}$$

$$= 0,0682 \cdot 57,3 \cdot 3,38$$

(data from test TTL2)

$$= 13,2\ \text{gm/min/cm}^2$$

The actual value for test TTL2 was 1,06 gm/min/cm². The carrying capacity for ERGO material was approximately a magnitude less than predicted by equation 2.5.2. The ERGO value of carrying capacity for Witwatersrand gold ore has been confirmed by results from another column flotation pilot plant operating on similar material (Moys 1989A). The reason for the poor prediction may be the use of the d_{80} to quantify the particle size distribution. While the d_{80} quantified the coarse range of the size distribution adequately, it is the ultra-fine material that consumes bubble surface area, and hence determines the column's carrying capacity for the material.

5.4.3 Flotation Rates

On a conventional flotation plant it is relatively simple to obtain kinetic data by sampling each cell lip along a bank of cells. The column on the other hand, had to be run at different feed rates to establish kinetic data. However, it is not possible to keep important variables, like gas rate and bias ratio, constant for the different tests. Bias ratio is a function of mass pull, which changes with residence time and wash water rate, and the optimum gas rate changes with feed grade and residence time. The variables in column flotation are interdependant, and a change in one may have ramifications for several others.

The column results not governed by the carrying capacity limitation were governed by other limitations e.g. flotation kinetics. The gold and sulphide sulphur recoveries were plotted versus residence time in Figure 5.11. The liquid residence times were adjusted to allow for settling of the solids (see Appendix III). With residence times in excess of 20 minutes, settling became an important limitation. For instance, in test JB2, a liquid residence time of 23

Figure 5.11

E R G O Flotation Column Kinetics.

Recovery Versus Residence Time.

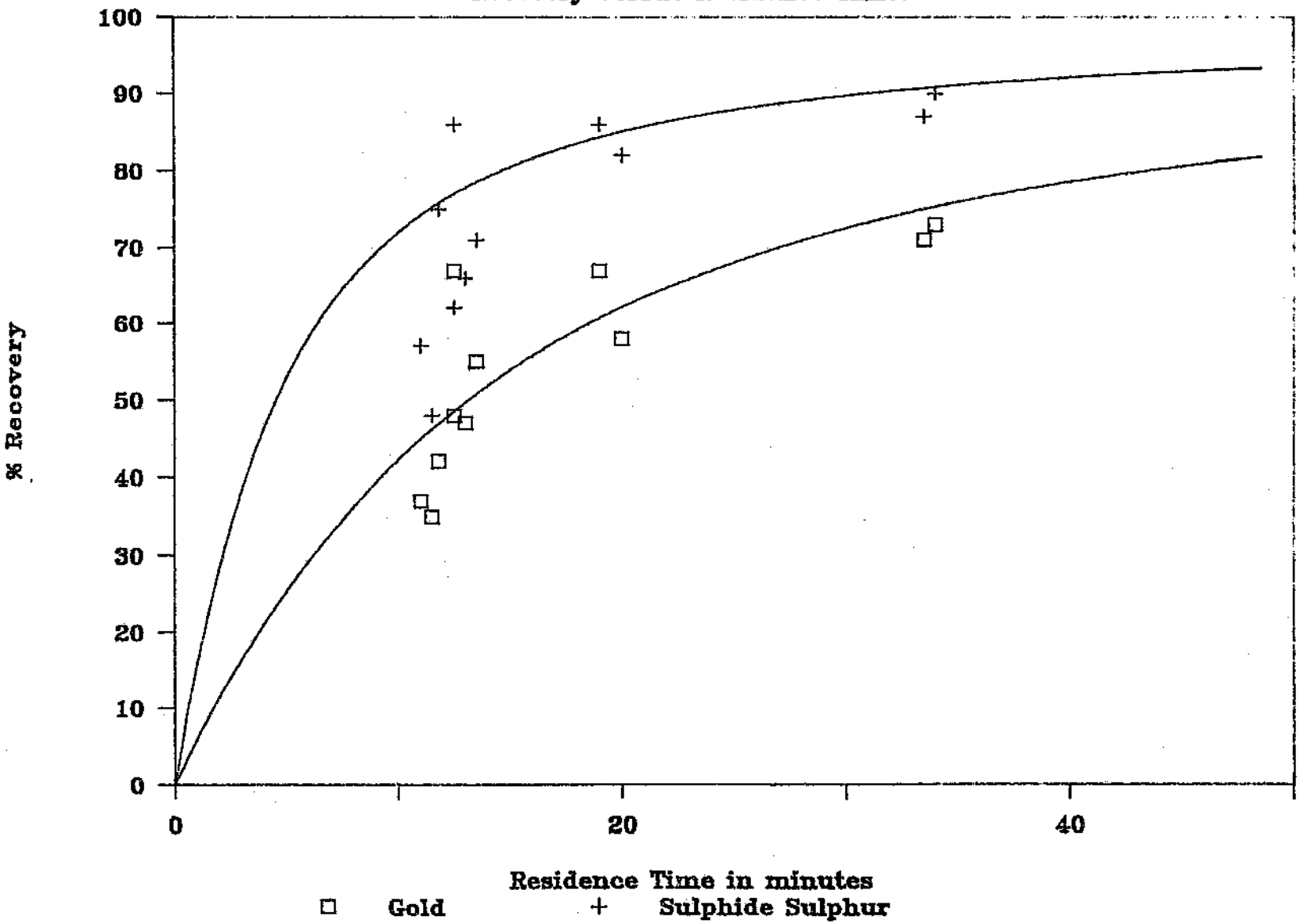
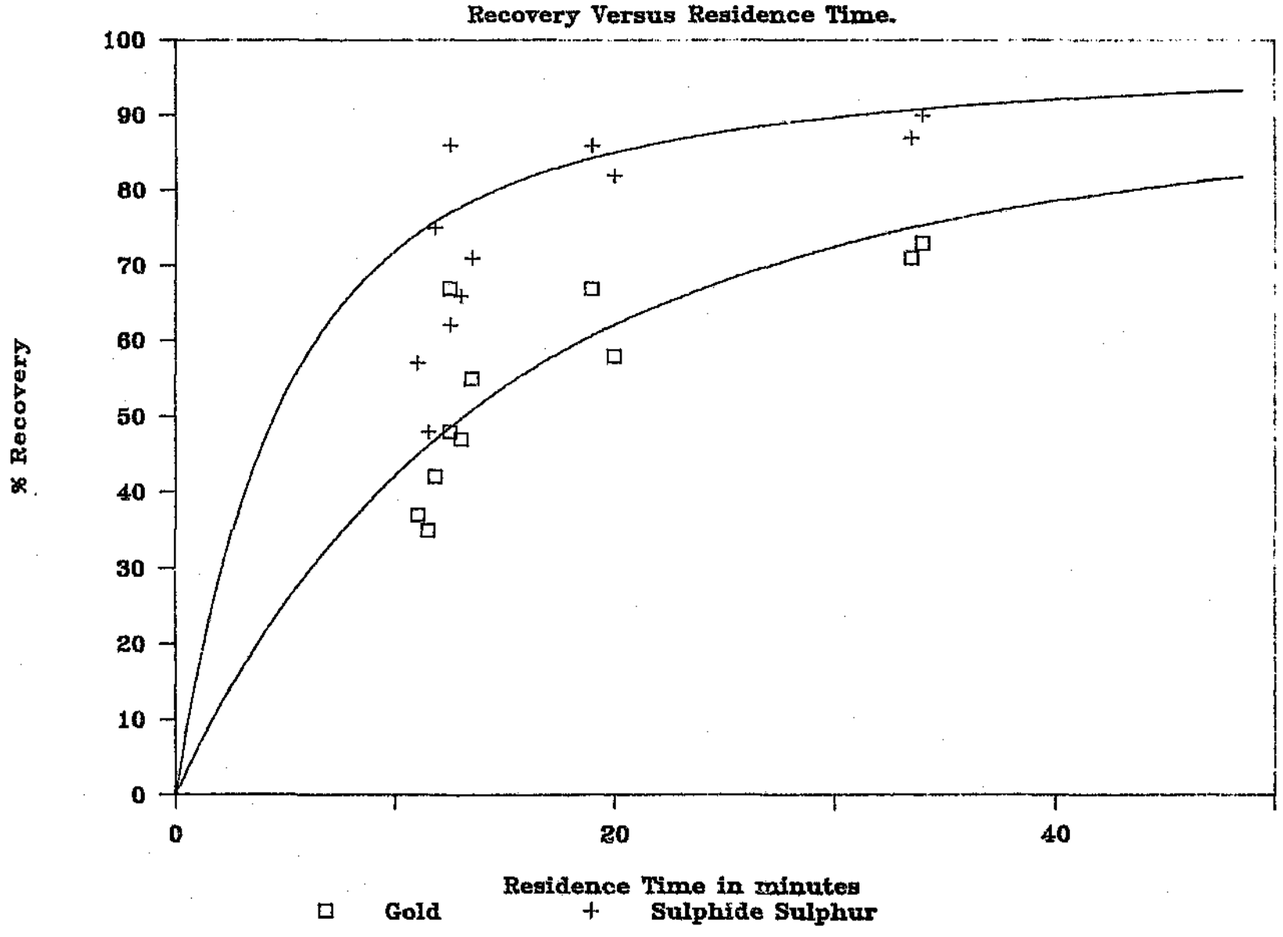


Figure 5.11

E R G O Flotation Column Kinetics.

Recovery Versus Residence Time.



minutes was calculated to have a mean flotable sulphide sulphur particle residence time of 20 minutes. The equivalent figures for test JB4 were 43 and 34 minutes.

Data were scattered because of different froth stabilities in different tests. While collection zone conditions were probably similar for most of the tests, the froth conditions were not. Column performance is sensitive to froth recoveries, in turn sensitive to froth conditions, so the column performance varied for similar residence times.

The criteria for choosing results approximately on the same kinetic curve are given below:

a) The concentrate grade must decrease with time.
b) The recovery must increase with time.

The tests which approximate to the same kinetic curve are tests BR3, JB1 and JB3 (Table 5.13). The feed grades for these tests were close to the mean values, except for the gold result for BR3. However the gold recovery for this test was not abnormally high.

The three results are plotted in Figure 5.11. The best-fit curves were determined using the Eureka P.C. package (1987) for the least - squares fit of

$$r = R\ (1-1/(k_u t)(1-\exp -k_u t)) \qquad \text{Equation 2.6.7}$$

and the results are tabulated in Table 5.14 for both column and conventional flotation.

Although the infinite time recovery of gold for column flotation was higher than for conventional flotation, the difference was within experimental error.

The rate constants for column and batch flotation are not really comparable, because they are a function of bubble size, bubble concentration, froth recovery etc., and not just flotability.

Table 5.13

RESULTS APPROXIMATING TO THE SAME KINETIC CURVE

Test No.	Residence Time	Gold Recovery %	Sulphide Sulphur Recovery %	Sulphide Sulphur Grade
BR3	11,8	42	75	32,6
JB1	18,8	67	86	29,2
JB3	34,1	73	90	21,1

Table 5.14

Comparison of the Parameters of Equation 2.6.7
for Column Flotation and Conventional Flotation

	Infinite Time Recovery R %	Flotation Rate Constant k_u t^{-1}	Time Lag Minutes	Max.* Error
Column Flotation				
Gold	98	0,13	0	6,7
Sulphide Sulphur	99	0,36	0	1,7
Conventional Flotation (Batch)				
Gold	92	0,18	0	2,92
Sulphide Sulphur	101	0,44	0	0,44
Column Flotation Assuming a Time Lag				
Gold	77	0,87	9,7	-
Sulphide Sulphur	92	1,17	7.4	-

*The maxium error is the largest difference between the data and the curve fit.

Kinetic data collected in other work (Moys 1989) indicated a time lag for gold flotation. The term (t-c) was substituted for t in Equation 2.6.7 and the equation solved for the three constants R, k, and c (Table 5.14). While the results have no statistical significance, the time lag c for gold flotation was 2,3 minutes greater than the time lag for sulphide sulphur flotation which may be attributed to the lower flotability of gold relative to sulphide sulphur. In the flotation column, where bubble surface area is limited, it is hypothesised that the gold tended to be crowded off the bubbles by the more flotable sulphides at lower residence times.

The infinite time recovery results for both cells were within limits of experimental error (Table 5.12), and the column recoveries showed no significant improvement over conventional flotation. The metallurgical benefit of column flotation for ERGO is an improved grade/recovery relationship.

CHAPTER 6

CONCLUSIONS AND RECOMMENDATIONS

6.1 SUMMARY AND CONCLUSIONS

a) A column flotation pilot plant was operated "live" in the ERGO flotation plant building. It experienced the normal plant variations and was large enough to experience industrial scale problems like axial mixing.

b) The gold recovery by column flotation may be improved 13 percentage points to 67 percent compared to a conventional bench scale test at a sulphide sulphur grade of 29 percent. Subsequent testwork (not discussed here) on rougher feed material at ERGO showed a significant improvement in performance by the single-stage column over the three-stage plant.

c) The column's recovery of coarse material was superior to that of the plant. The coarse material recovered by the column was probably coarse pyrite.

d) The USBM bubble generator performed well and produced fine (ca. one millimetre diameter) uniform bubbles. However, process water caused clogging in the bubble generator and scale formation in the injector holes, and potable water, which is expensive in South Africa, had to be used to operate it.

e) The overall performance of the column was dependant on the froth phase. Coalescence caused loss of recovery and it was necessary to add sufficient wash water to maintain a fluid froth.

f) The performance of the column was a complex function of the many variables associated with column flotation e.g. gas rate, residence time, froth height, bias rate, wash water rate and the position of the wash water distributor.

g) The many parameters in column flotation are difficult to keep constant over varying residence times. For example, bias ratio varies with residence time if all other variables are constant.

h) The results were separated into two categories: those results assumed to be limited by carrying capacity and those with other limitations such as flotation kinetics, gas rate, wash-water rate, etc. When plotted as recovery versus time, the data were scattered (Figure 5.11) and in order to best identify those results which were flotation rate limited and on the same flotation rate curve two criteria were used.

   i) The gold and sulphide sulphur recoveries must increase with residence time; and

   ii) the sulphide sulphur concentrate grade must decrease with residence time.

i) The mean Sauter diameter or $d_{(3,2)}$ is the correct measure of particle size for carrying capacity correlations. It is defined as the diameter of a sphere whose cross-sectional area is the mean sectional area of the particles and is easily determined on a laser particle size analyser. The use of other size distribution parameters e.g. d_{80} may lead to incorrect conclusions and unreliable carrying capacity correlations.

j) The carrying capacity of the ERGO column was a tenth that predicted by d_{80} literature correlations. The correlation suggested for Witwatersrand gold ore is:

$$C_a = 0{,}034 . d_{(3,2)} . \rho_p . J_g .$$

k) The efficiency of the column dropped off at residence times in excess of 20 minutes due to settling in the column.

1) The differential pressure cell performed adequately as a level transducer for steady-state operation of the ERGO pilot plant. However, differential pressure cell level control is vulnerable to variations in column parameters such as gas rate, and is subject to large errors where operating parameters are expected to vary.

6.2 RECOMMENDATIONS

a) The performance of the flotation column should be evaluated on different feeds at ERGO e.g. rougher feed, plant tailings, etc.

b) Column flotation should be evaluated on other Witwatersrand flotation plants.

c) A feasibility study should be conducted for column flotation of Witwatersrand gold ore.

d) A commercially viable sparger which can use dirty plant water should be developed.

e) A fundamental investigation should be conducted into the low carrying capacities associated with Witwatersrand gold ore.

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APPENDIX I

DEMONSTRATION OF THE EQUIVALENCE OF THE d_p, THE DERIVED PARTICLE DIAMETER FROM SECTION 2.5.2, AND THE $d_{(3,2)}$, THE SAUTER MEAN DIAMETER

$$d_{(3,2)} = \frac{\int_{D_1}^{D_2} D^3 n(D)\,dD}{\int_{D_1}^{D_2} D^2 n(D)\,dD} \qquad \text{(Malvern 1987)}$$

In the same notation used previously to derive Equation 2.5.5 (Section 2.5)

$$d_{(3,2)} = \frac{\int_{l_1}^{l_2} l^3 n(l)\,dl}{\int_{l_1}^{l_2} l^2 n(l)\,dl} \qquad \text{Equation I}$$

In this study the whole size range of the sample is of interest and the integral limits are:

$$l_1 = 0 \quad \text{and} \quad l_2 = \infty$$

Equation I then becomes

$$d_{(3,2)} = \frac{\int_0^{\infty} l^3 n(l)\,dl}{\int_0^{\infty} l^2 n(l)\,dl} \qquad \text{Equation II}$$

$n(l)dl$ = the fraction by number of particles in the interval l to $l + d(l)$.

$$n(l)dl = \frac{\text{Mass of particles in the interval } dl}{\text{Mass of a particle of diameter } l}$$

$$\text{Total number of particles}$$

$$= \frac{\dfrac{p_c(l)dl}{\rho_p \dfrac{\pi l^3}{6}}}{\displaystyle\int_0^{\infty} \frac{p_c(l)dl}{\rho_p \dfrac{\pi l^3}{6}}} \qquad \text{Equation III}$$

However, for any particular sample the total number of particles is constant.

$$\int_0^{\infty} \frac{p_c(l)dl}{\rho_p \dfrac{\pi l^3}{6}} = N_T$$

Equation III becomes

$$n(l)dl = \frac{\dfrac{p_c(l)dl}{\rho_p \dfrac{\pi l^3}{6}}}{N_T}$$

$$n(l)\,dl = \frac{6\,p_c(l)\,dl}{\rho_p\,\pi l^3 N_T} \qquad \text{Equation IV}$$

Substituting Equation IV in Equation II

$$d_{(3,2)} = \frac{\int_0^{\infty} l^3 \cdot \dfrac{6\,p_c(l)\,dl}{\rho_p \pi l^3 N_T}}{\int_0^{\infty} l^2 \cdot \dfrac{6\,p_c(l)\,dl}{\rho_p \pi l^3 N_T}}$$

$$d_{(3,2)} = \frac{\int_0^{\infty} p_c(l)\,dl}{\int_0^{\infty} \dfrac{p_c(l)\,dl}{l}}$$

$$d_{(3,2)} = \left(\int_0^{\infty} \frac{p_c(l)\,dl}{l} \right)^{-1}$$

$$d_{(3,2)} = d_p$$

APPENDIX II

ERGO COLUMN FLOTATION RESULTS
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX

	Date		15-May	18-May	18-May	18-May
	Stream (A,B,C)		A	A	A	A
	Expt no		TTL1 PBi0	TTL2	TTL3	TTL4
MEASUREMENTS:						
	FEED					
	SAMPLE: Volume	l	15.000	14.250	15.000	16.500
	Wet Mass	kg	15.900	15.150	17.100	17.400
	Dry Mass	kg	2.064	1.908	2.495	2.503
	Time	sec	40.000	40.000	40.000	40.000
	ASSAYS: Sulphide Sulphur	%	.300	5.600	3.370	3.960
	Gold	g/t	2.870	4.650	2.710	3.770
	Uranium	g/t		317.000	212.000	130.000
	pH		3.800	3.460	3.600	3.000
	CONCENTRATE					
	SAMPLE: Volume	l	4.340	6.000	6.780	8.600
	Wet Mass	kg	4.580	6.500	7.090	8.990
	Dry Mass	kg	0.549	0.625	0.432	0.458
	Time	sec	40.000	80.000	80.000	80.000
	ASSAYS: Sulphide Sulphur	%	21.400	20.600	21.600	19.300
	Gold	g/t	10.660	11.700	10.600	13.200
	Uranium	g/t		724.000	570.000	441.000
	pH			3.800	3.860	3.550
	TAILS					
	SAMPLE: Volume	l	0.500	18.500	15.340	20.000
	Wet Mass	kg	21.400	19.000	16.500	21.260
	Dry Mass	kg	1.785	1.803	1.889	2.514
	Time	sec	40.000	40.000	40.000	40.000
	Volumetric Flowrate	l/min	30.800	27.600	31.200	30.000
	ASSAYS: Sulphide Sulphur	%	1.950	2.200	2.010	1.340
	Gold	g/t	1.620	2.170	1.670	2.050
	Uranium	g/t		193.000	165.000	125.000
	pH			3.580	3.780	3.430
OPERATING CONDITIONS:	BUBBLE GENERATOR					
	Gas Rate	l/min	33.000	30.000	35.000	40.000
	Gas Pressure	kPa	630.000	600.000	600.000	600.000
	Water Flowrate	l/min	2.750	1.940	1.750	1.600
	Water Pressure	kPa	590.000	560.000	560.000	550.000
	Water pH		3.300	3.200	3.150	3.400
	Frother Concentration	ppm	165.000	165.000	165.000	165.000
	B. Generator Pressure	kPa		600.000	600.000	600.000
	WASH WATER					
	Wash Water Flowrate	l/min	10.300	8.500	8.500	8.500
	Frother Concentration	ppm	165.000	165.000	165.000	165.000
	pH		3.300	3.200	3.150	3.400
	Distributor above or below Froth Surface		Below	below	below	below
	FROTH [illegible]					
	Froth Height	mm	710.000	870.000	850.000	820.000
	Controller Set Point		65.000	60.000	60.000	60.000
	Conductivity Reading					
REAGENT ADDITION:	Copper Sulphate	ppm	10.000	0.000	0.000	0.000
	Senkol 50	ppm	20.000	0.000	0.000	0.000

CALCULATIONS:					
FEED					
Volumetric Flowrate	l/min	22.500	21.375	24.000	24.750
Mass Flowrate	kg/min	3.095	2.862	3.743	3.755
% Solids	%	12.978	12.594	14.591	14.385
R.D.	kg/l	1.060	1.063	1.069	1.055
Superficial Velocity	cm/s	0.844	0.802	0.900	0.928
RECONSTITUTED FEED					
Mass Flowrate	kg/min	3.505	3.159	4.166	4.115
ASSAYS:Sulphide Sulphur	%	6.516	4.931	3.534	2.839
Gold	g/t	3.742	3.584	2.365	2.981
Uranium	g/t	0.000	271.802	196.498	151.381
CONCENTRATE					
% Recovery:Mass	%	23.477	14.840	7.777	8.349
S/Sulphur	%	77.100	62.002	47.541	56.747
Gold	%	66.874	48.442	34.865	36.969
Uranium	%	ERR	39.530	22.560	24.321
Volumetric Flowrate	l/min	6.510	4.500	5.085	6.450
Mass Flowrate	kg/min	0.823	0.469	0.324	0.344
% Solids	%	11.978	9.615	6.093	5.095
R.D.	kg/l	1.055	1.083	1.046	1.045
TAILS					
Volumetric Flowrate	l/min	30.800	27.600	31.200	30.000
Mass Flowrate	kg/min	2.682	2.690	3.842	3.771
% Solids	%	8.343	9.489	11.448	11.825
R.D.	kg/l	1.044	1.027	1.076	1.063
CRITICAL C.F. PARAMETERS					
Residence Time	min	12.411	13.592	12.052	12.579
Superficial Gas Velocity	cm/s	1.238	1.125	1.313	1.500
Concentrate Mass Flux	g/min/sq.cm	1.852	1.055	0.729	0.773
Bias Ratio		1.156	1.185	1.131	1.078
REAGENT CONSUMPTION					
Frother	g/t	614.301	545.363	405.962	443.867
Copper Sulphate	g/t	11.412	0.000	0.000	0.000
Senkol 50	g/t	22.823	0.000	0.000	0.000
VOLUMETRIC BALANCE					
IN	l/min	35.550	31.815	34.250	34.850
OUT	l/min	37.310	32.100	36.285	36.450
MASS BALANCE					
IN	kg/min	3.095	2.862	3.743	3.755
OUT	kg/min	3.505	3.159	4.166	4.115

E R G O FLOTATION COLUMN
Test No. TTL1

	Mass kg/hr	Mass %	Gold g/t	S/S %	U3O8 g/t	% Distribution Gold	S/S	U3O8	% -32 um
Cons	49.4	23	10.66	21.40	0	67	77	ERR	58.3
Tails	160.9	77	1.62	1.95	0	33	23	ERR	91.9
Calc Head	210.3	100	3.74	6.52	0	100	100	100	84.7
Meas'd Head	185.7		2.87	7.30	0				

E R G O FLOTATION COLUMN
Test No. TTL2

	Mass kg/hr	Mass %	Gold g/t	S/S %	U3O8 g/t	% Distribution Gold	S/S	U3O8	% -32 um
Cons	28.1	15	11.70	20.60	724	48	62	40	65.0
Tails	161.4	85	2.17	2.20	193	52	38	60	92.0
Calc Head	189.5	100	3.58	4.93	272	100	100	100	90.6
Meas'd Head	171.7		4.65	5.60	317				

E R G O FLOTATION COLUMN
Test No. TTL3

	Mass kg/hr	Mass %	Gold g/t	S/S %	U3O8 g/t	% Distribution Gold	S/S	U3O8	% -32 um
Cons	19.4	8	10.60	21.60	570	35	48	23	76.6
Tails	230.5	92	1.67	2.01	165	65	52	77	89.8
Calc Head	250.0	100	2.36	3.53	196	100	100	100	89.9
Meas'd Head	224.6		2.71	3.37	212				

E R G O FLOTATION COLUMN
Test No. TTL4

	Mass kg/hr	Mass %	Gold g/t	S/S %	U3O8 g/t	% Distribution Gold	S/S	U3O8	% -32 um
Cons	20.6	8	13.20	19.30	441	37	57	24	82.9
Tails	226.3	92	2.05	1.34	125	63	43	76	94.4
Calc Head	246.9	100	2.98	2.84	151	100	100	100	90.8
Meas'd Head	225.3		3.77	3.96	130				

ERGO COLUMN FLOTATION RESULTS
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX

	Date		08-Jun	08-Jun	08-Jun
	Stream (A,B,C)		A	A	A
	Expt no		BR1	BR2	BR3
MEASUREMENTS:					
	FEED				
	SAMPLE: Volume	l	13.200	14.550	15.400
	Wet Mass	kg	14.800	16.200	17.400
	Dry Mass	kg	2.386	2.656	2.893
	Time	sec	40.000	40.000	40.000
	ASSAYS: Sulphide Sulphur	%	4.580	4.720	4.760
	Gold	g/t	3.680	3.490	5.340
	Uranium	g/t	158.000	193.000	147.000
	pH		2.700	3.200	3.200
	CONCENTRATE				
	SAMPLE: Volume	l	2.600	2.750	2.300
	Wet Mass	kg	3.000	4.400	2.600
	Dry Mass	kg	0.470	0.808	0.623
	Time	sec	80.000	80.000	80.000
	ASSAYS: Sulphide Sulphur	%	27.100	26.300	32.600
	Gold	g/t	14.800	13.600	9.940
	Uranium	g/t	856.000	633.000	422.000
	pH		3.320	4.080	4.150
	TAILS				
	SAMPLE: Volume	l	19.000	16.600	18.250
	Wet Mass	kg	20.600	17.900	20.000
	Dry Mass	kg	2.626	2.146	2.545
	Time	sec	40.000	40.000	40.000
	Volumetric Flowrate	l/min	27.600	24.200	28.600
	ASSAYS: Sulphide Sulphur	%	1.270	1.280	1.260
	Gold	g/t	1.570	2.150	1.590
	Uranium	g/t	119.000	141.000	61.000
	pH		4.370	3.410	3.520
OPERATING CONDITIONS:	BUBBLE GENERATOR				
	Gas Rate	l/min	25.000	25.000	25.000
	Gas Pressure	kPa	620.000	600.000	600.000
	Water Flowrate	l/min	2.750	2.750	2.750
	Water Pressure	kPa	540.000	540.000	540.000
	Water pH		3.860	3.470	3.800
	Frother Concentration	ppm	165.000	165.000	165.000
	B. Generator Pressure	kPa	570.000	570.000	580.000
	WASH WATER				
	Wash Water Flowrate	l/min	6.100	4.800	3.700
	Frother Concentration	ppm	165.000	165.000	165.000
	pH		3.860	3.470	3.800
	Distributor above or below Froth Surface		Below	Below	Below
	FROTH PHASE				
	Froth Height	mm	410.000	390.000	410.000
	Controller Set Point		75.000	75.000	75.000
	Conductivity Reading				
REAGENT ADDITION:	Copper Sulphate	ppm	10.000	10.000	10.000
	Senkol 50	ppm	20.000	20.000	20.000

CALCULATIONS:				
FEED				
Volumetric Flowrate	l/min	19.800	21.825	23.100
Mass Flowrate	kg/min	3.579	3.984	4.340
% Solids	%	16.122	16.395	16.626
R.D.	kg/l	1.121	1.113	1.130
Superficial Velocity	cm/s	0.743	0.818	0.866
RECONSTITUTED FEED				
Mass Flowrate	kg/min	4.167	3.735	4.456
ASSAYS:Sulphide Sulphur	%	3.455	5.340	4.547
Gold	g/t	2.689	4.008	2.466
Uranium	g/t	181.344	220.837	98.858
CONCENTRATE				
% Recovery:Mass	%	8.459	16.227	10.487
S/Sulphur	%	66.351	79.920	75.193
Gold	%	46.556	55.062	42.277
Uranium	%	39.930	46.513	44.766
Volumetric Flowrate	l/min	1.950	2.063	1.725
Mass Flowrate	kg/min	0.353	0.606	0.467
% Solids	%	15.667	18.364	23.962
R.D.	kg/l	1.154	1.600	1.130
TAILS				
Volumetric Flowrate	l/min	27.600	24.200	28.600
Mass Flowrate	kg/min	3.815	3.129	3.988
% Solids	%	12.748	11.989	12.725
R.D.	kg/l	1.084	1.078	1.096
CRITICAL C.F. PARAMETERS				
Residence Time	min	14.334	16.384	13.832
Superficial Gas Velocity	cm/s	0.938	0.938	0.938
Concentrate Mass Flux	g/min/sq.cm	0.793	1.364	1.051
Bias Ratio		1.200	1.146	1.083
REAGENT CONSUMPTION				
Frother	g/t	350.423	333.578	238.858
Copper Sulphate	g/t	9.599	10.711	8.978
Senkol 50	g/t	19.198	21.422	17.955
VOLUMETRIC BALANCE				
IN	l/min	28.650	29.375	29.550
OUT	l/min	29.550	26.263	30.325
MASS BALANCE				
IN	kg/min	3.579	3.984	4.340
OUT	kg/min	4.167	3.735	4.456

E R G O FLOTATION COLUMN
Test No. BR1

	Mass kg/hr	Mass %	Gold g/t	S/S %	U3O8 g/t	% Distribution Gold	S/S	U3O8	% -32 um
Cons	21.2	8	14.80	27.10	856	47	66	40	55.5
Tails	228.9	92	1.57	1.27	119	53	34	60	90.6
Calc Head	250.0	100	2.69	3.45	181	100	100	100	97.3
Meas'd Head	214.7		3.68	4.58	158				

E R G O FLOTATION COLUMN
Test No. BR2

	Mass kg/hr	Mass %	Gold g/t	S/S %	U3O8 g/t	% Distribution Gold	S/S	U3O8	% -32 um
Cons	36.4	16	13.60	26.30	633	55	80	47	57.4
Tails	187.7	84	2.15	1.28	141	45	20	53	89.7
Calc Head	224.1	100	4.01	5.34	221	100	100	100	86.7
Meas'd Head	239.0		3.49	4.72	193				

E R G O FLOTATION COLUMN
Test No. BR3

	Mass kg/hr	Mass %	Gold g/t	S/S %	U3O8 g/t	% Distribution Gold	S/S	U3O8	% -32 um
Cons	28.0	10	9.94	32.60	422	42	75	45	57.7
Tails	239.3	90	1.59	1.26	61	58	25	55	88.2
Calc Head	267.3	100	2.47	4.55	99	100	100	100	85.2
Meas'd Head	260.4		5.34	4.76	147				

SIZE ANALYSES
BR2 Concentrate

Screen Size um	Mass %	Cum. Mass %	Cum. Mass %
+150	0.3%	0.3%	100.0%
-150+106	6.6%	6.9%	99.7%
-106+75	22.0%	28.9%	93.1%
-75+53	7.0%	35.9%	71.1%
-53+45	0.5%	36.3%	64.1%
-45	63.7%	100.0%	63.7%
Total	100.0%	100.0%	100.0%

SIZE ANALYSES
BR2 Head

Screen Size um	Mass %	Cum. Mass %	Cum. Mass %
+150	4.5%	4.5%	100.0%
-150+106	2.6%	7.0%	95.5%
-106+75	4.6%	11.6%	93.0%
-75+53	1.7%	13.3%	88.4%
-53+45	0.2%	13.5%	86.7%
-45	86.5%	100.0%	86.5%
Total	100.0%	100.0%	100.0%

SIZE ANALYSES
BR2 Tailings

Screen Size um	Mass %	Cum. Mass %	Cum. Mass %
+150	5.4%	5.4%	100.0%
-150+106	2.3%	7.7%	94.6%
-106+75	1.9%	9.6%	92.3%
-75+53	0.7%	10.3%	90.4%
-53+45	0.0%	10.3%	89.7%
-45	89.7%	100.0%	89.7%
Total	100.0%	100.0%	100.0%

ERGO COLUMN FLOTATION RESULTS
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX

	Date		23-Jun	23-Jun
	Stream (A,B,C)		A	A
	Expt no		CF1	CF2
			CLEANER FEED	
MEASUREMENTS:				
	FEED			
	SAMPLE: Volume	l	12.500	
	Wet Mass	kg	14.000	
	Dry Mass	kg	2.439	
	Time	sec	40.000	
	ASSAYS: Sulphide Sulphur	%	20.825	
	Gold	g/t	7.840	
	Uranium	g/t		
	pH		3.600	
	CONCENTRATE			
	SAMPLE: Volume	l	15.500	13.000
	Wet Mass	kg	16.585	14.630
	Dry Mass	kg	1.432	2.289
	Time	sec	80.000	80.000
	ASSAYS: Sulphide Sulphur	%	24.905	27.920
	Gold	g/t	8.972	9.238
	Uranium	g/t		
	pH		3.770	3.060
	TAILS			
	SAMPLE: Volume	l	14.750	8.500
	Wet Mass	kg	15.940	9.325
	Dry Mass	kg	1.935	1.276
	Time	sec	40.000	40.000
	Volumetric Flowrate	l/min	18.900	12.000
	ASSAYS: Sulphide Sulphur	%	10.955	12.215
	Gold	g/t	5.325	6.660
	Uranium	g/t		
	pH		3.700	3.010
OPERATING CONDITIONS:	BUBBLE GENERATOR			
	Gas Rate	l/min	38.000	33.000
	Gas Pressure	kPa	620.000	600.000
	Water Flowrate	l/min	3.400	3.450
	Water Pressure	kPa		560.000
	Water pH		3.800	3.800
	Frother Concentration	ppm	0.000	0.000
	B. Generator Pressure	kPa	600.000	580.000
	WASH WATER			
	Wash Water Flowrate	l/min	9.000	0.000
	Frother Concentration	ppm	60.000	0.000
	pH		4.600	
	Distributor above or below Froth Surface		below	above
	FROTH PHASE			
	Froth Height	mm	150.000	270.000
	Controller Set Point		77.000	77.000
	Conductivity Reading			
REAGENT ADDITION:	Copper Sulphate	ppm	10.000	10.000
	Senkol 50	ppm	20.000	20.000
HOLDUP:	Manometer 7	mm	1905.000	1897.000
	Manometer 6	mm	1935.000	1925.000
	Manometer 5	mm	1980.000	1950.000
	Manometer 4	mm	ERR	ERR
	Manometer 3	mm	ERR	ERR
	Manometer 2	mm	ERR	ERR
	Gauge Reading	kPa	-8.800	-5.606

CALCULATIONS:				
	FEED			
	Volumetric Flowrate	l/min	18.750	ERR
	Mass Flowrate	kg/min	3.659	ERR
	% Solids	%	17.421	ERR
	R.D.	kg/l	1.120	ERR
	Superficial Velocity	cm/s	0.703	ERR
	RECONSTITUTED FEED			
	Mass Flowrate	kg/min	3.553	3.518
	ASSAYS:Sulphide Sulphur	%	15.171	19.879
	Gold	g/t	6.427	7.918
	Uranium	g/t	0.000	0.000
	CONCENTRATE			
RECOVERY	% Recovery:Mass	%	30.224	48.797
	S/Sulphur	%	49.616	68.537
	Gold	%	42.190	56.933
	Uranium	%	ERR	ERR
	Volumetric Flowrate	l/min	11.625	9.750
	Mass Flowrate	kg/min	1.074	1.717
	% Solids	%	8.634	15.646
	R.D.	kg/l	1.070	1.125
	TAILS			
	Volumetric Flowrate	l/min	18.900	12.000
	Mass Flowrate	kg/min	2.479	1.801
	% Solids	%	12.124	13.684
	R.D.	kg/l	1.082	1.097
CRITICAL	CRITICAL C.F. PARAMETERS			
PARAMETERS	Residence Time	min	21.544	33.486
	Superficial Gas Velocity	cm/s	1.425	1.238
	Concentrate Mass Flux	g/min/sq.cm	2.417	3.863
	Bias Ratio		0.855	0.467
	HOLDUP			
	Collection Zone	%	22.424	19.408
	Manometers (1375-1175)	%	21.444	21.609
	Manometers (1175-975)	%	28.376	20.241
	Manometers (975-775)	%	ERR	ERR
	Manometers (775-575)	%	ERR	ERR
	Manometers (575-375)	%	ERR	ERR
	REAGENT CONSUMPTION			
	Frother	g/t	151.966	0.000
	Copper Sulphate	g/t	11.257	11.370
	Senkol 50	g/t	22.513	22.739
	VOLUMETRIC BALANCE			
	IN	l/min	31.150	ERR
	OUT	l/min	30.525	21.750
	MASS BALANCE			
	IN	kg/min	3.939	ERR
	OUT	kg/min	3.553	3.518

E R G O FLOTATION COLUMN
Test No. CF1

	Mass kg/hr	Mass %	Gold g/t	S/S %	U3O8 g/t	% Distribution Gold	S/S	U3O8	% -32 um
Cons	64.4	30	8.97	24.91	0	42	50	ERR	
Tails	148.8	70	5.33	10.96	0	58	50	ERR	
Calc Head	213.2	100	6.43	15.17	0	100	100	100	
Meas'd Head	236.3		7.84	20.83	0				

E R G O FLOTATION COLUMN
Test No. CF2

	Mass kg/hr	Mass %	Gold g/t	S/S %	U3O8 g/t	% Distribution Gold	S/S	U3O8	% -32 um
Cons	103.0	49	9.24	27.92	0	57	69	ERR	
Tails	108.1	51	6.66	12.22	0	43	31	ERR	
Calc Head	211.1	100	7.92	19.88	0	100	100	100	
Meas'd Head	ERR		0.00	0.00	0				

ERGO COLUMN FLOTATION RESULTS
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX

	Date			29-Jun	29-Jun
	Stream (A,B,C)			A	A
	Expt no			RT1	RT2
				CLEANER FEED	
MEASUREMENTS:					
	FEED				
	SAMPLE: Volume	l		13.240	14.150
	Wet Mass	kg		14.400	15.400
	Dry Mass	kg		1.787	1.950
	Time	sec		40.000	40.000
	ASSAYS:Sulphide Sulphur	%		4.155	3.765
	Gold	g/t		6.350	4.590
	Uranium	g/t		298.000	255.000
	pH			4.910	4.820
	CONCENTRATE				
	SAMPLE: Volume	l		4.910	4.820
	Wet Mass	kg		5.400	5.200
	Dry Mass	kg		0.885	0.508
	Time	sec		80.000	80.000
	ASSAYS:Sulphide Sulphur	%		24.800	19.300
	Gold	g/t		12.000	14.500
	Uranium	g/t		668.000	874.000
	pH			4.240	3.650
	TAILS				
	SAMPLE: Volume	l		19.000	19.750
	Wet Mass	kg		20.000	20.600
	Dry Mass	kg		1.581	1.383
	Time	sec		40.000	40.000
	Volumetric Flowrate	l/min		28.235	28.371
	ASSAYS:Sulphide Sulphur	%		2.100	2.220
	Gold	g/t		2.170	1.290
	Uranium	g/t		204.000	170.000
	pH			4.060	3.170
OPERATING					
CONDITIONS:	BUBBLE GENERATOR				
	Gas Rate	l/min		30.000	30.000
	Gas Pressure	kPa		600.000	600.000
	Water Flowrate	l/min		3.750	3.750
	Water Pressure	kPa		560.000	565.000
	Water pH			3.800	3.800
	Frother Concentration	ppm		20.000	20.000
	B. Generator Pressure	kPa		580.000	580.000
	WASH WATER				
	Wash Water Flowrate	l/min		6.700	6.700
	Frother Concentration	ppm		20.000	20.000
	pH			3.000	3.710
	Distributor above or				
	below Froth Surface			Below	Below
	FROTH PHASE				
	Froth Height	mm		315.000	290.000
	Controller Set Point			77.000	77.000
	Conductivity Reading				
REAGENT					
ADDITION:	Copper Sulphate	ppm		0.000	0.000
	Senkol 50	ppm		0.000	0.000
ROLDUP:	Manometer 7	mm		1900.000	1917.000
	Manometer 6	mm		ERR	1937.000
	Manometer 5	mm		ERR	1965.000
	Manometer 4	mm		ERR	1985.000
	Manometer 3	mm		ERR	2010.000
	Manometer 2	mm		ERR	ERR
	Gauge Reading	kPa		-2.900	-3.900

CALCULATIONS:				
	FEED			
	Volumetric Flowrate	l/min	19.860	21.225
	Mass Flowrate	kg/min	2.681	2.925
	% Solids	%	12.410	12.662
	R.D.	kg/l	1.088	1.088
	Superficial Velocity	cm/s	0.745	0.796
	RECONSTITUTED FEED			
	Mass Flowrate	kg/min	2.863	2.382
	ASSAYS:Sulphide Sulphur	%	6.030	4.952
	Gold	g/t	3.934	3.403
	Uranium	g/t	287.256	282.618
	CONCENTRATE			
RECOVERY	% Recovery:Mass	%	17.943	15.997
	S/Sulphur	%	71.421	62.343
	Gold	%	54.735	88.158
	Uranium	%	41.726	49.470
	Volumetric Flowrate	l/min	3.683	3.615
	Mass Flowrate	kg/min	0.514	0.381
	% Solids	%	12.685	9.769
	R.D.	kg/l	1.100	1.079
	TAILS			
	Volumetric Flowrate	l/min	28.235	28.571
	Mass Flowrate	kg/min	2.349	2.001
	% Solids	%	7.905	6.714
	R.D.	kg/l	1.053	1.043
CRITICAL	CRITICAL C.F. PARAMETERS			
PARAMETERS	Residence Time	min	14.161	14.033
	Superficial Gas Velocity	cm/s	1.125	1.125
	Concentrate Mass Flux	g/min/sq.cm	1.156	0.857
	Bias Ratio		1.141	1.142
	HOLDUP			
	Collection Zone	%	12.545	13.267
	Manometers (1375-1175)	%	ERR	13.714
	Manometers (1175-975)	%	ERR	17.549
	Manometers (975-775)	%	ERR	13.714
	Manometers (775-575)	%	ERR	16.110
	Manometers (575-375)	%	ERR	ERR
	REAGENT CONSUMPTION			
	Frother	g/t	72.995	87.752
	Copper Sulphate	g/t	0.000	0.000
	Senkol 50	g/t	0.000	0.000
	VOLUMETRIC BALANCE			
	IN	l/min	30.310	31.675
	OUT	l/min	31.918	32.186
	MASS BALANCE			
	IN	kg/min	2.681	2.925
	OUT	kg/min	2.863	2.382

E R G O FLOTATION COLUMN
Test No. RT1

	Mass kg/hr	Mass %	Gold g/t	S/S %	U3O8 g/t	% Distribution Gold	S/S	U3O8	% -32 um
Cons	30.8	18	12.00	24.00	668	55	71	42	37.2
Tails	141.0	82	2.17	2.10	204	45	29	58	89.8
Calc Head	171.8	100	3.93	6.03	287	100	100	100	
Meas'd Head	160.8		6.35	4.16	298				88.1

E R G O FLOTATION COLUMN
Test No. RT2

	Mass kg/hr	Mass %	Gold g/t	S/S %	U3O8 g/t	% Distribution Gold	S/S	U3O8	% -32 um
Cons	22.9	16	14.50	19.30	874	58	62	49	56.7
Tails	120.0	84	1.29	2.22	170	32	38	51	91.3
Calc Head	142.9	100	3.40	4.95	283	100	100	100	
Meas'd Head	175.5		4.59	3.77	255				94.0

E R G O FLOTATION COLUMN
Test No. RT1

Screen Size um	HEAD Mass %	Gold g/t	S/S %	Gold %	S/S %	CONCENTRATE Mass %	Gold g/t	S/S %	Gold %	S/S %	TAILINGS Mass %	Gold g/t	S/S %	Gold %	S/S %
+75	12.50	6.54	13.50	19.65	26.79	28.30	18.30	33.40	41.94	37.05	10.80	2.08	1.01	10.80	5.01
-75	87.50	3.82	5.27	80.35	73.21	71.70	10.00	22.40	58.06	62.95	89.20	2.08	2.32	89.20	94.99
Overall	100.00	4.16	6.36	100.00	100.00	100.00	12.35	25.51	100.00	100.00	100.00	2.08	2.18	100.00	100.00

E R G O FLOTATION COLUMN
Test No. RT2

Screen Size um	HEAD Mass %	Gold g/t	S/S %	Gold %	S/S %	CONCENTRATE Mass %	Gold g/t	S/S %	Gold %	S/S %	TAILINGS Mass %	Gold g/t	S/S %	Gold %	S/S %
+75	17.90	6.10	7.90	28.36	33.69	33.60	15.80	30.70	40.19	54.63	11.80	1.81	0.66	8.78	5.94
-75	82.10	3.36	3.39	71.64	66.31	66.40	11.90	12.90	59.81	45.37	88.80	2.50	1.39	91.22	94.06
Overall	100.00	3.85	4.20	100.00	100.00	100.00	13.21	18.88	100.00	100.00	100.80	2.43	1.31	100.00	100.00

	Date		03-Jul	03-Jul	07-Jul	07-Jul
	Stream (A,B,C)		A	A	A	A
	Expt no		JB1	JB2	JB3	JB4
			CLEANER FEED		12h00	12h20
MEASUREMENTS:						
	FEED					
	SAMPLE: Volume	l	8.000	7.330	5.020	5.150
	Wet Mass	kg	8.800	8.000	5.800	5.900
	Dry Mass	kg	1.274	0.871	0.974	0.909
	Time	sec	40.000	40.000	80.000	80.000
	ASSAYS: Sulphide Sulphur	%	4.900	5.560	4.970	5.150
	Gold	g/t	3.375	3.345	3.070	3.560
	Uranium	g/t	201.000	226.000	212.000	253.000
	pH		3.790	4.570	3.730	3.680
	CONCENTRATE					
	SAMPLE: Volume	l	3.600	4.650	5.160	4.050
	Wet Mass	kg	4.000	4.800	6.000	4.400
	Dry Mass	kg	0.420	0.287	0.438	0.401
	Time	sec	80.000	80.000	160.000	160.000
	ASSAYS: Sulphide Sulphur	%	29.200	24.100	21.100	22.800
	Gold	g/t	13.500	12.700	7.450	9.010
	Uranium	g/t	707.000	358.000	573.000	638.000
	pH		4.280	4.360	3.920	4.020
	TAILS					
	SAMPLE: Volume	l	13.050	11.400	13.000	11.650
	Wet Mass	kg	13.500	12.100	13.300	12.000
	Dry Mass	kg	0.941	1.047	0.520	0.549
	Time	sec	40.000	40.000	80.000	80.000
	Volumetric Flowrate	l/min	19.355	16.783	9.231	9.262
	ASSAYS: Sulphide Sulphur	%	1.090	0.940	1.100	1.140
	Gold	g/t	1.480	1.280	1.230	1.270
	Uranium	g/t	105.000	99.000	97.000	106.000
	pH		4.050	4.180	3.950	3.940
OPERATING CONDITIONS:	BUBBLE GENERATOR					
	Gas Rate	l/min	28.000	32.000	32.000	32.000
	Gas Pressure	kPa	510.000	510.000	510.000	510.000
	Water Flowrate	l/min	3.900	3.800	3.800	3.800
	Water Pressure	kPa	470.000	475.000	460.000	460.000
	Water pH		3.800	3.800	4.050	4.030
	Frother Concentration	ppm	20.000	20.000	20.000	20.000
	B. Generator Pressure	kPa	480.000	480.000	480.000	480.000
	WASH WATER					
	Wash Water Flowrate	l/min	6.500	4.800	4.000	4.000
	Frother Concentration	ppm	20.000	25.000	25.000	25.000
	pH		3.520	3.500	3.560	3.600
	Distributor above or below Froth Surface		Below	Below	Below Low Down	Below Low Down
	FROTH PHASE					
	Froth Height	mm	500.000	490.000	420.000	420.000
	Controller Set Point		70.000	70.000	69.000	69.000
	Conductivity Reading				830.000	800.000
REAGENT ADDITION:	Copper Sulphate	ppm	0.000	0.000	0.000	0.000
	Senkol 50	ppm	0.000	0.000	0.000	0.000
HOLDUP:	Manometer 7	mm	1795.000	1796.000	1807.000	1807.000
	Manometer 6	mm	1820.000	1817.000	1838.000	1836.000
	Manometer 5	mm	1840.000	1842.000	1868.000	1865.000
	Manometer 4	mm	1867.000	1865.000	1897.000	1893.000
	Manometer 3	mm	1890.000	1894.000	1931.000	1923.000
	Manometer 2	mm	ERR	ERR	ERR	ERR
	Gauge Reading	kPa	ERR	ERR	-6.900	-6.500
BUBBLE SIZE:	Visual estimation	mm	1.000	1.000	1.000	1.000

CALCULATIONS:						
	FEED					
	Volumetric Flowrate	l/min	12.000	10.995	3.765	3.863
	Mass Flowrate	kg/min	1.911	1.307	0.731	0.682
	% Solids	%	14.477	10.888	16.793	15.407
	R.D.	kg/l	1.100	1.091	1.155	1.146
	Superficial Velocity	cm/s	0.450	0.412	0.141	0.145
	RECONSTITUTED FEED					
	Mass Flowrate	kg/min	1.711	1.757	0.533	0.587
	ASSAYS:Sulphide Sulphur	%	6.266	3.602	7.258	6.688
	Gold	g/t	3.693	2.679	3.145	3.253
	Uranium	g/t	215.854	130.491	243.553	242.274
	CONCENTRATE					
RECOVERY	% Recovery:Mass	%	18.414	12.253	30.788	25.615
	S/Sulphur	%	85.808	81.975	89.510	87.321
	Gold	%	67.307	58.081	72.932	70.956
	Uranium	%	60.314	33.429	72.435	67.455
	Volumetric Flowrate	l/min	2.700	3.488	1.935	1.519
	Mass Flowrate	kg/min	0.315	0.215	0.164	0.150
	% Solids	%	10.500	5.979	7.300	9.114
	R.D.	kg/l	1.111	1.032	1.163	1.086
	TAILS					
	Volumetric Flowrate	l/min	19.355	16.783	9.231	9.266
	Mass Flowrate	kg/min	1.396	1.541	0.369	0.437
	% Solids	%	6.970	8.653	3.910	4.575
	R.D.	kg/l	1.034	1.061	1.023	1.030
CRITICAL	CRITICAL C.F. PARAMETERS					
PARAMETERS	Residence Time	min	20.233	23.359	42.809	42.644
	Superficial Gas Velocity	cm/s	1.050	1.200	1.200	1.200
	Concentrate Mass Flux	g/min/sq.cm	0.709	0.484	0.370	0.338
	Bias Ratio		1.326	1.112	1.614	1.831
	HOLDUP					
	Collection Zone	%	ERR	ERR	14.137	14.187
	Manometers (1375-1175)	%	15.417	15.678	17.406	16.994
	Manometers (1175-975)	%	13.000	17.562	16.917	16.994
	Manometers (975-775)	%	16.383	16.620	16.429	16.508
	Manometers (775-575)	%	14.450	19.446	18.872	17.479
	Manometers (575-375)	%	ERR	ERR	ERR	ERR
	REAGENT CONSUMPTION					
	Frother	g/t	121.593	111.576	329.909	299.804
	Copper Sulphate	g/t	0.000	0.000	0.000	0.000
	Senkol 50	g/t	0.000	0.000	0.000	0.000
	VOLUMETRIC BALANCE					
	IN	l/min	22.400	19.595	11.565	11.663
	OUT	l/min	22.055	20.271	11.166	10.785
	MASS BALANCE					
	IN	kg/min	1.911	1.307	0.731	0.682
	OUT	kg/min	1.711	1.757	0.533	0.587

E R G O FLOTATION COLUMN
Test

	Mass kg/hr	Mass %	Gold g/t	S/S %	U3O8 g/t	% Distribution Gold	S/S	U3O8	% -32 um
Cons	18.9	18	13.50	29.20	707	67	86	60	59.0
Tails	83.7	82	1.48	1.09	105	33	14	40	92.7
Calc Head	102.6	100	3.69	6.27	216	100	100	100	
Meas'd Head	114.7		3.38	4.90	201				93.2

E R G O FLOTATION COLUMN
Test No. JB2

	Mass kg/hr	Mass %	Gold g/t	S/S %	U3O8 g/t	% Distribution Gold	S/S	U3O8	% -32 um
Cons	12.9	12	12.70	24.10	356	58	82	33	77.7
Tails	92.5	88	1.26	0.74	99	42	18	67	92.0
Calc Head	105.4	100	2.68	3.60	130	100	100	100	
Meas'd Head	78.4		3.35	5.56	226				91.1

E R G O FLOTATION COLUMN
Test No. JB3

	Mass kg/hr	Mass %	Gold g/t	S/S %	U3O8 g/t	% Distribution Gold	S/S	U3O8	% -32 um
Cons	9.9	31	7.45	21.10	573	73	90	72	71.0
Tails	22.2	69	1.23	1.10	97	27	10	28	94.6
Calc Head	32.0	100	3.15	7.26	244	100	100	100	
Meas'd Head	43.8		3.07	4.97	212				90.7

E R G O FLOTATION COLUMN
Test No. JB4

	Mass kg/hr	Mass %	Gold g/t	S/S %	U3O8 g/t	% Distribution Gold	S/S	U3O8	% -32 um
Cons	9.0	26	9.01	22.80	638	71	87	67	72.7
Tails	26.2	74	1.27	1.14	106	29	13	33	93.4
Calc Head	35.2	100	3.25	6.69	242	100	100	100	
Meas'd Head	40.9		3.56	5.15	253				92.3

E R G O FLOTATION COLUMN
Test No. JB1

Screen Size um	HEAD Mass %	Gold g/t	S/S %	Gold %	S/S %	CONCENTRATE Mass %	Gold g/t	S/S %	Gold %	S/S %	TAILINGS Mass %	Gold g/t	S/S %	Gold %	S/S %
+75	7.40	9.82	21.70	22.72	30.24	22.00	18.40	27.30	31.28	22.13	4.30	1.45	0.26	5.61	1.39
-75	92.60	2.67	4.00	77.28	69.76	78.00	11.40	27.10	68.72	77.87	93.70	1.64	1.24	94.39	98.61
Overall	100.00	3.20	5.31	100.00	100.00	100.00	12.94	27.14	100.00	100.00	100.00	1.63	1.18	100.00	100.00

E R G O FLOTATION COLUMN
Test No. JB2

Screen Size um	HEAD Mass %	Gold g/t	S/S %	Gold %	S/S %	CONCENTRATE Mass %	Gold g/t	S/S %	Gold %	S/S %	TAILINGS Mass %	Gold g/t	S/S %	Gold %	S/S %
+75	6.60	11.70	28.00	20.69	30.63	5.50	19.50	36.30	7.86	7.54	5.00	1.36	0.15	4.61	0.92
-75	93.40	3.17	4.48	79.31	69.37	94.50	13.30	25.90	92.14	92.46	95.00	1.48	0.85	95.39	99.08
Overall	100.00	3.73	6.03	100.00	100.00	100.00	13.64	26.47	100.00	100.00	100.00	1.47	0.82	100.00	100.00

E R G O FLOTATION COLUMN
Test No. JB3

Screen Size um	HEAD Mass %	Gold g/t	S/S %	Gold %	S/S %	CONCENTRATE Mass %	Gold g/t	S/S %	Gold %	S/S %	TAILINGS Mass %	Gold g/t	S/S %	Gold %	S/S %
+75	5.20	9.60	13.00	16.02	10.70	11.80	16.10	18.40	25.33	14.51	2.31	1.90	0.61	3.47	1.13
-75	94.80	2.76	5.95	83.98	89.30	88.20	6.35	14.50	74.67	85.49	97.69	1.25	1.24	96.53	98.87
Overall	100.00	3.12	6.32	100.00	100.00	100.00	7.50	14.96	100.00	100.00	100.00	1.27	1.24	100.00	100.00

E R G O FLOTATION COLUMN
Test No. JB4

Screen Size um	HEAD Mass %	Gold g/t	S/S %	Gold %	S/S %	CONCENTRATE Mass %	Gold g/t	S/S %	Gold %	S/S %	TAILINGS Mass %	Gold g/t	S/S %	Gold %	S/S %
+75	7.30	10.70	ERR	21.48	ERR	15.29	17.90	27.00	29.35	19.28	3.64	1.54	0.51	4.48	1.44
-75	92.70	3.08	ERR	78.52	ERR	84.71	7.78	20.40	70.65	80.72	96.36	1.24	1.32	95.52	98.56
Overall	100.00	3.64	ERR	100.00	ERR	100.00	9.33	21.41	100.00	100.00	100.00	1.25	1.29	100.00	100.00

APPENDIX III

CORRECTION OF RESIDENCE TIMES FOR PARTICLE SETTLING VELOCITIES

A suitable particle size was needed to calculate settling velocity, and it was decided to use the screen size which retains 50 percent of the flotable sulphide sulphur.

Ideally, to determine this size a head sample for each test would have been floated to completion in a batch cell (as described in section 4.2.1) and a screen analysis, together with a sulphide sulphur size analysis, would have been done for each head sample.

In the plant environment however, this type of attention to detail was not really feasible, so a reasonable estimate of the screen size was attempted using the available data.

In the last six tests done the concentrate was screened at 75 micrometres, and the fractions were assayed for sulphur. The average concentrate mass of the plus 75 micrometre fraction was found to be 22,2 percent, containing 35,5 percent of the sulphide sulphur. Test JB2 was excluded from consideration because the concentrate screen analysis disagreed with the head and tails analyses.

Assuming the assay on the next size fraction down to be the same, it was estimated that the coarsest 31,2 percent of the average concentrate mass would contain 50 percent of the flotable sulphide sulphur. The theoretical screen size that would retain 31,2 percent would be the particle size used for the settling velocity analysis. So the screen size which would retain the 22,2 percent, and an additional 9 percent of the next coarsest material, would give the particle size for 50 percent of the floated sulphide sulphur.

A full screen analysis done on the concentrate in test BR2 showed that 7,5 percent of the concentrate mass was in the minus 75 plus 45 micrometre size fraction. Because 45 micrometres was the

smallest screen size available, and because of the approximate nature of the estimate, it was d[illegible]d that 29,7 was close enough to 31,2 percent to justify using 45 micrometre as the screen size which retained 50 percent of the flotable sulphide sulphur.

The settling velocity was estimated using figure 5-80 in Perry (1984) for terminal velocities of spherical particles settling in water under the action of gravity. It was assumed that the feed pulp was dilute enough for the continuous phase to be considered pure water. A particle specific gravity of 3 was used and the terminal velocity of a 45 micrometre particle was estimated as 0,240 cm/sec.

The corrected residence time was estimated by the addition of the superficial feed velocity and the mean settling velocity to yield the mean solids velocity. The collection zone height was assumed to be 7,8 metres. The effective residence times are tabulated below.

Table A.1

CORRECTIONS FOR SETTLING VELOCITIES
IN RESIDENCE TIME ESTIMATES

Test No.	Superficial Feed Velocity cm/s	Mean Settling Velocity cm/s	Corrected Residence Time min
TTL1	0.844	0.240	12.0
TTL2	0.802	0.240	12.5
TTL3	0.900	0.240	11.4
TTL4	0.928	0.240	11.1
BR1	0.743	0.240	13.2
BR2	0.818	0.240	12.3
BR3	0.866	0.240	11.8
CF1	0.703	0.240	13.8
CF2	0.703	0.240	13.8
RT1	0.745	0.240	13.2
RT2	0.796	0.240	12.5
JB1	0.450	0.240	18.8
JB2	0.412	0.240	19.9
JB3	0.141	0.240	34.1
JB4	0.145	0.240	33.8

A P P E N D I X I V

EVALUATION OF THE DIFFERENTIAL PRESSURE CELL FOR LEVEL CONTROL

The differential pressure cell is known to be very sensitive to variation in column parameters (Moys and Finch 1988). During the testwork at ERGO the column variables were controlled closely. Visual assessment of the resulting level control showed it to be satisfactory.

However, in the results tabulated in Table A.2 and plotted in Figure A.1, it can be seen that one control setting, 77, gave different levels, 150 and 315 millimetres for tests CF1 and RT1 respectively. This difference in level was due to gas rate and gas hold-up being high in CF1 and low in RT1.

Table A.2

INTERFACE LEVEL MEASUREMENT BY DIFFERENTIAL PRESSURE CELL

Table of Interface Depths and Controller Setpoints

Test No.	Interface Depth mm	Gas Rate cm/sec	Controller Setting %
TTL1	710	1,24	65
TTL2	870	1,13	60
TTL3	850	1,31	60
TTL4	820	1,50	60
BR1	410	0,94	75
BR2	390	0,94	75
BR3	410	0,94	75
CF1	150	1,43	77
CF2	270	1,24	77
RT1	315	1,125	77
RT2	290	1,125	77
JB1	500	1,05	70
JB2	490	1,20	70
JB3	420	1,20	69
JB4	420	1,20	69

Where variations in operating parameters like gas rate can be expected, level control with the differential pressure cell will not be satisfactory.

An increase in gas rate caused an increase in gas hold-up in the pulp phase. The hydrostatic pressure at the Dp cell was maintained by increasing the height of the pulp phase and therefore reducing the height of the froth phase.

A multilinear regression with setpoint and gas rate as independent variables was done, and gave a higher correlation coefficient than when only setpoint was correlated (see Table A.3). The gas rate coefficient was negative , indicating that an increase in gas rate caused a decrease in froth depth.

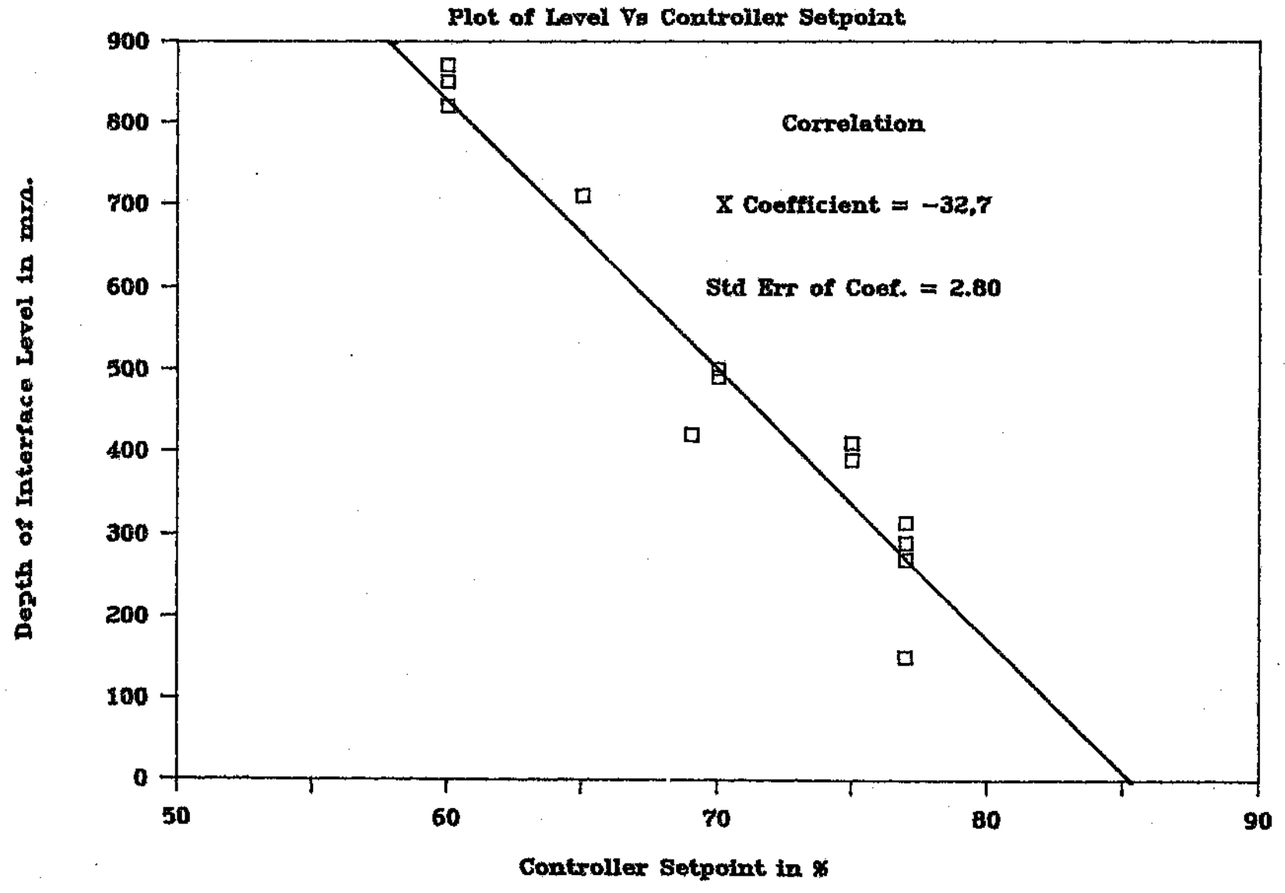
Table A.3

REGRESSION ANALYSIS

	Setpoint and Gas Rate	Setpoint Only
Coefficients		
Setpoint	-36	-33
Gas Rate	-270	
Correlation Coefficient		
R Squared	0.95	0,91

Figure A.1

INTERFACE LEVEL MEASUREMENT BY Dp CELL



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