THE DETERMINATION OF THE EFFECTS OF PHYSICAL AND CHEMICAL PARAMETERS ON THE COLUMN FLOTATION CELL, PERFORMANCE IN THE FLOTATION OF PYRITE

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DECLARATION

I hereby declare that the material incorporated in this dissertation is my own work except where indicated otherwise.

Signed by candidate

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April 1991

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SYNOPSIS

This masters dissertation on column flotation is to determine the effects of physical and chemical parameters on the column flotation cell performance in the flotation of pyrite. Hypotheses are also proposed to explain observed changes.

Chapter one gives a brief description of flotation columns, some applications of flotation columns and a literature survey that covers the effects of various physical and chemical parameters.

Chapter two states the objective of this research and the plan of action used to achieve this objective.

Chapter three describes the first part of the plan of action, namely to design a laboratory flotation column and to draw up an experimental procedure. A flotation column rig with an adequately repeatable experimental procedure is the result.

The repeatability and sensitivity of the experimental procedure is given in chapter four. This chapter also includes the results obtained when the physical and chemical parameters are varied. The effects on the flotation cell performance by varying parameters are summarised. The biggest changes observed in the flotation column cell performance are as follows:

1) An increased air rate yields an optimum sulphur recovery.

2) Concentrate sulphur grades decrease when the air rate is increased.
3) The concentrate grades increase when the froth depth is increased.
4) Increasing the feed solids percent to the flotation column has no effect on the concentrate grades and recoveries as long as the column is operated below its maximum carrying capacity.
5) The concentrate grades are improved by adding wash water.

Chapter five then discusses the changes observed in the flotation column cell performance.

Chapter six covers the design of a pilot plant flotation column rig, the results obtained on plant and the discussion of these results. It is found that the pilot plant rig can be effectively used for on site test work. The flotation column cell performed better than the conventional cells. The pilot plant test work showed that:

1) Increasing the air rate increases the recovery.
2) Increasing the wash water rate improves the concentrate grades.

Finally, in chapter seven, conclusions are drawn regarding the results and discussions. Some of these conclusions are:

1) The flotation column cell performes better than conventional flotation cells due to the column´s deep water washed froth and counter-current contact mechanism.
2) Increasing the air rate decreases the grades of the concentrate due to increased entrainment, while the recovery moves through an optimum.
3) Increasing the froth depth increases the concentrate grades due to a longer cleaning action. For the same reason the recovery decreases.

4) The feed solids percentage has no effect on the grade or recovery. Should the maximum carrying capacity however be reached the recovery will drop.

5) Froth stability is essential for good concentrate grades and recoveries.

6) Increased wash water rates increase the concentrate grades due to a stronger washing action and more stable froths.

7) The chemical parameters which are varied show the same trends as was observed in conventional flotation.

All data obtained during the test work is attached in the appendices.

LIST OF SYMBOLS

Ca	carrying capacity
D	column diameter (d_c)
E_k	collection efficiency
H	collection zone height
Jb	superficial bias rate
Jg	superficial air rate (actual average)
Jg^*	superficial air rate (atmospheric conditions)
Jpf	superficial floated particle rate
Js	superficial bubble surface rate
Jse	effective superficial bubble surface rate (assuming that each particle is shared by two bubbles)
Jsl	superficial slurry velocity
Jw	superficial wash water rate
K	constant
L	collection zone length
Nd	vessel dispersion number for the liquid
Np	vessel dispersion number for solid particles
Pc	pressure at concentrate lip
Pt	pressure at air input level
Rep	particle Reynolds number
S	bubble surface
Sr	bubble surface required per gram of solids
d80	80% passing size of feed
dc	column diameter
g	gravitational acceleration
k	collection zone rate constant
v_g	gas rate
v_{gmin}	minimum gas rate
μf	fluid viscosity
ρp	solids density
$\rho susp$	suspension density
τl	liquid residence time
τp	particle residence time

LIST OF TERMS USED

air sparger	bubble generator
bias rate	net flow of water through the cleaning zone (positive direction downwards).
carrying capacity	the maximum carrying rate, mass of concentrate solids recovered per unit time per unit column cross-sectional area.
cleaning zone	combination of the packed bubble bed and conventional froth; zone above interface.
collection efficiency	the effectiveness of the particle attaching (and to remain attached) to a bubble.
collection zone	the section from the sparger to the interface.
counter current	bubbles moving up and the pulp moving down.
displacement washes	the amount of wash water required to replace the concentrate slurry. ie. 1 displacement wash is equivalent to the same volume of concentrate.
flotation column	alternatively called the Canadian Flotation Column.
froth zone	cleaning zone
H/D ratio	the ratio of the collection zone length to the column diameter.
hydraulic entrainment	entrainment into the concentrate due to drag and frictional forces of the rising air bubbles.
hydrophilic	particle does not tend to attach to bubble.
hydrophobic	particle tends to attach to bubble.
positive bias	net downward flow of water in the cleaning zone.
pulp-froth interface	interface between the collection and cleaning zone.
slugging	the formation of large bubbles that rise much faster than the superficial air rate.
superficial air rate	the volumetric air flow rate per unit cross-section.
superficial bias rate	the bias rate per unit cross-section.

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CHAPTER ONE

INTRODUCTION

In order to appreciate present day froth flotation and especially column flotation a brief account of the history and development of column flotation is presented.

The earliest information on the difference in wettability of various minerals is recorded in the British patent 488/1860 to William Haynes. Ore was agitated with an oily agent in water during which the sulphide minerals and oily agents were separated. The Australian invention of the flotation process involving gas as the buoyant medium followed in the period 1901 to 1905.

The use of pine oil as a frother was discovered around 1909. This and other soluble collecting agents formed the basis of what is known today as froth flotation. In the early twenties the discovery of xanthates triggered off research into the fundamental chemical and physical principles of froth flotation.

In the early 1960's the **flotation column** was invented by D.A. Wheeler. The column differs significantly from conventional flotation in that no mechanical agitation is needed, the pulp moves counter current to the air bubbles, a deep froth bed is employed and the froth bed is washed.

1.1 DESCRIPTION OF THE FLOTATION COLUMN

Industrial **flotation columns** are typically 0.5 to 2.5 m in diameter and 10 to 12 m high (Table 1.1). The dimensions and the type of operations have been extensively revued. (Yianatos, J.B., 1987; Finch, J.A. and Dobby, G.S.,1988). The columns are either square or circular in shape.

The flotation column makes use of **counter current** contact between the mineral slurry and the air phase. Small air bubbles with a diameter of less than 2 mm are generated at the bottom of the column by an **air sparger**. The concentrate grades are improved with the use of a **deep froth zone (cleaning zone)** and wash water which is added at the top of the column.

A schematic diagram of a flotation column is shown in Figure 1.1. The column consists essentially of the **collection zone** and the **cleaning zone**.

The feed is introduced about two thirds of the way up the column and below the **pulp-froth interface**. The feed particles move down the collection zone and collide with the rising air bubbles in a counter current fashion. The **hydrophobic** particles attach themselves to the rising air bubbles. The **hydrophilic** particles are removed in the tailings at the bottom of the column.

The hydrophobic particles **attached** to the rising air bubbles are transported to the cleaning zone. Wash water is added near the top of the cleaning zone to maintain a **positive bias.** This positive bias prevents **hydraulic entrainment** of fine hydrophilic particles into the concentrate.

SOME COLUMN FLOTATION USERS	DIMENSIONS	OPERATION
Mt. Isa Mines (Australia)	2.4m * 10m	Zn Scavenging
	(3 in series)	
Woodcutters N.T. (Australia)	0.8m * 8m	Zn Cleaner
Bouganville (Australia)	1.2m * 13m	Cu Rougher Cleaner
	1.8m * 13m	Scavenger Cleaner
Gibraltar Mines (Canada)	(3@) 2m * 13m	Cu Cleaner
Inco, Sudbury (Canada)	(2@) 1.8m * 12m	Cu Matte
	0.9m * 12m	
Falconbridge, Sudbury (Canada)	0.9m * 14m	Cu-Ni Separation
Cominco, Polaris (Canada)	0.4m * 10m	Pb and Zn
	0.76m square	
Brenda Mines (Canada)	(5@) 1.7m * 12m	Cu and Mo
Highland Valley (Canada)	1m * 12m	Mo
Island Copper (Canada)	0.85m 1.3m 1.8m	Mo
	(2@) 2.4m	
Curraugh Resources (Canada)	1m * 12m	Zn
Magma Copper Co. (Pinto Vall.)	0.6m * 7.6m	Mo Scavenging
	0.6m * 7.6m	Mo Cleaner
Magma Copper Co. (San Manuel)	1.8m * 12.1m	Cu/Mo Cleaner
	1.5m * 12.1m	Cu/Mo Cleaner
	1.2*1.2*9.7m	Mo Cleaner
	0.8m * 3.6m	Mo Cleaner
Cyprus Minerals (Sierrite)	0.9m * 15m	Mo Cleaner
	2m * 15.1m	Cu/Mo Cleaner
Lornex Mining Co.	0.6m * 11m	Mo Cleaner
	0.9m * 12.8m	Mo Cleaner
Disputada Exxon (Chile)	0.9m * 12m	Cu Cleaner
	0.45m * 12m	Cu Cleaner
Codelco Chuquicamata	0.9m * 13.6m	Mo Cleaner
Harbour Lights (Australia)	2.5m * 12m	Bulk Sulphide Rougher
	1.2m * 12m	
Paddington (Australia)	(2@)3.25m * 10m	Bulk Sulphide Rougher

Table 1.1: A summary of some column flotation applications.

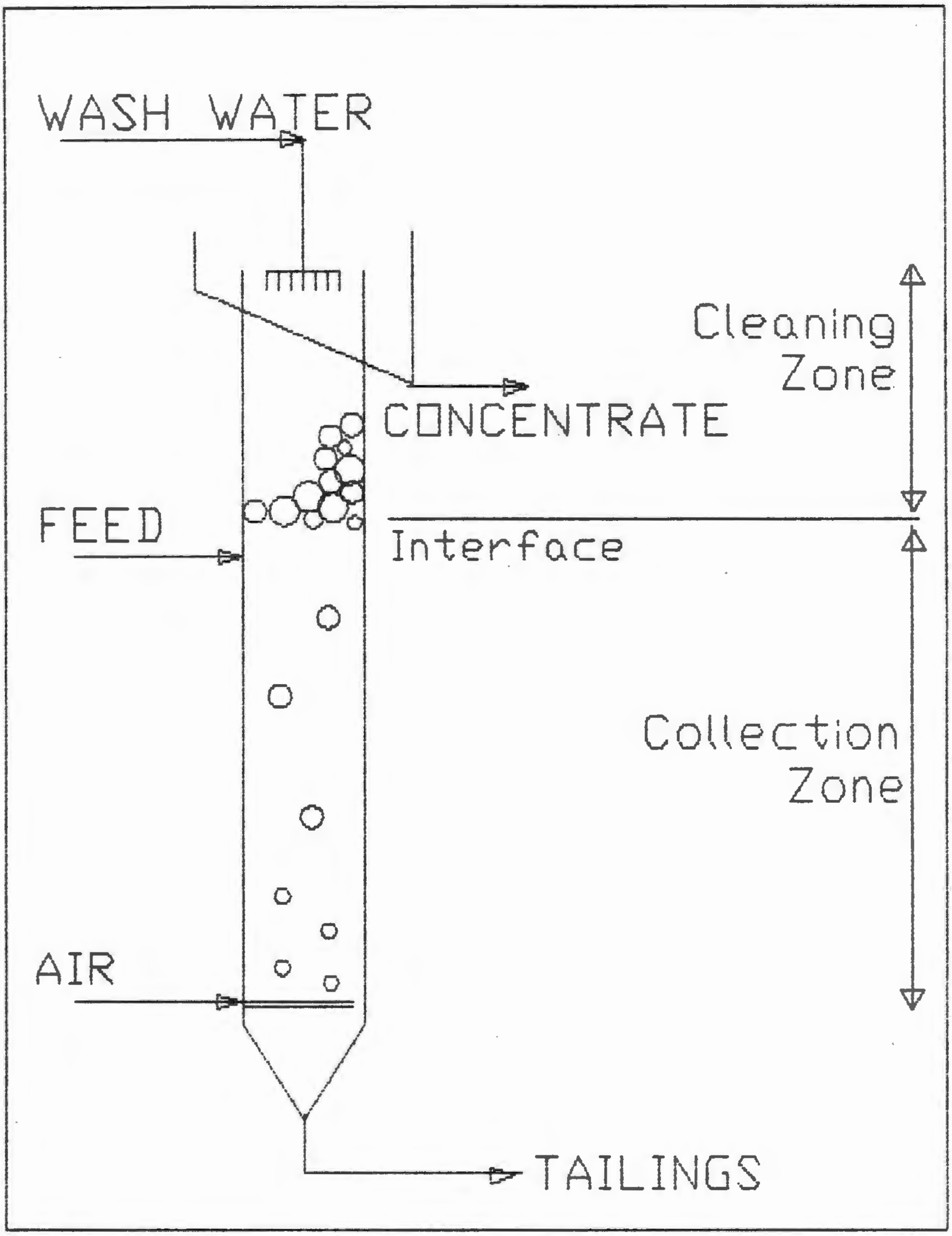

Figure 1.1: Schematic diagram of a flotation column.

1.2 THE USE OF COLUMN FLOTATION CELLS

Column flotation is widely used and studied in the flotation of sulphide ores, non-sulphide ores, precious metals, phosphates and coal. A number of typical **column flotation** applications are listed in Table 1.1. The application of column flotation in sulphide and non-sulphide flotation is summarised below.

1.2.1 GOLD-PYRITE FLOTATION

In sulphide flotation there are only 2 commercial installations mentioned in the researched literature. The first is at the Harbour Lights Mine in Lenora, Western Australia. A 2.5 m diameter by 12 m high column has been successfully used in a 100 tons/hour flotation plant to produce a final gold bearing sulfide concentrate in a single stage. Concentrates containing 120 g/ton of gold, 6% arsenic and 35 to 40% sulphur were obtained.

The second sulphide column flotation in use is at the Paddington mine in Western Australia. Two columns, each with a diameter of 3,25 m and a height of 11,45 m are used as a bulk sulphide rougher. The flotation feed is milled up to 80% passing approximately 75 μm. 80% of the total sulphides and 50% of the arsenopyrite is recovered in the columns.

Pilot plant tests with an 11 cm diameter column cell at the Harbour Lights Mine demonstrated that columns can be used in rougher, scavenger and cleaner flotation stages (Subramanian, et al., 1988).

1.2.2 NON SULPHIDE FLOTATION

From Table 1.1 it can be seen that most of the applications are for molybdenum. In these cases the column is mostly used in a cleaner mode.

In many cases column flotation has proved to be a superior method of flotation, but this technology has not yet been perfected. Research into this field of mineral processing is therefore still an ongoing process.

1.3 LITERATURE SURVEY

This literature survey is divided into 3 sections. The first two sections record the effects of physical and chemical parameters on the performance of the flotation column with respect to the concentrate grades and recoveries. The last section discusses models proposed and used for the scale up and design of columns. Physical parameters which have been studied are the following:

1) Air rate;
2) Collection zone length;
3) Froth depth;
4) Feed solids percentage;
5) Feed rate;
6) Wash water rate, position and temperature;
7) Interface level;
8) Bubble size;
9) Particle size.

The chemical parameters considered important are the following:

1) Reagent type and addition;
2) pH

1.3.1 THE EFFECT OF PHYSICAL PARAMETERS

1.3.1.1 Air Rate

The hydrophobic minerals collide with and attach to the rising air bubbles in the collection zone. The mineral is then transported to the froth by the air bubbles. Because the amount of hydrophobic mineral transported to the froth depends on the size and number of air bubbles available it is important to know the air rate on which these two variables depend.

The air rate in flotation columns is usually reported in terms of the **superficial air rate**, **Jg** (cm/sec). This superficial air rate is calculated by dividing the volumetric air flow rate by the column cross sectional area. **Jg** is the actual average air rate in the column and is related to the superficial air rate at atmospheric conditions, **Jg***, by the following relationship:

$$Jg = \frac{Pc\ Jg^{*}\ \ln(Pt/Pc)}{Pt - Pc}$$

Typical values of Jg are 1 to 3 cm/sec (Yianatos, J.B., 1987).

Studies done on the air rate show that there is an optimum air rate for maximum recovery. It has however been noted that this optimum air rate varies between about 1,2 and 2,5 cm/sec.

Using a laboratory column in a copper flotation, the maximum recovery occurred at **Jg** values of about 2,0 ± 0,5 cm/sec (Dobby, D.S. and Finch, J.A., 1986). In fine coal flotation an optimum **Jg** value of 1,70 cm/sec was found (Parekh, et al., 1988). In a pyrite flotation the maximum sulphur recovery was achieved at a **Jg** of about 1,2 cm/sec (Goodall, C.M. and O'Connor, C.T., 1989). In micro bubble column flotation of fine coal the recovery increased with an increasing air flow rate until an optimum was found at just less than 2,5 cm/sec. Above an air rate of 2,5 cm/sec **slugging** occurs (Luttrell, G.H., 1988). Egan et al (1988) found in their studies that the air rate should be kept below 3 cm/sec. They also found that as the ratio of the zinc in the feed to air decreased the zinc recovery increased.

The grade of the concentrate was shown to decrease with an increasing air flow rate (Goodall, C.M. and

O´Connor, C.T., 1989). In the flotation of fine coal from waste coal refuse an increase in the air flow rate had a detrimental effect on the quality of the concentrate (Misra, M. and Harris, R., 1988). A smaller decrease in the quality of the clean coal was observed with an increased air flow rate (Parekh, et al., 1988).

The effect of the gas rate on the cleaning of the froth is more significant than the **bias ratio**. Feed water is completely **rejected** from the concentrate at superficial gas rates less than 1,5 cm/sec with froth depths greater than 1m (Yianatos, J.B. et al., 1987).

1.3.1.2 Collection Zone Length

The collection zone height, **H**, is an important factor in column flotation. The mineral recovery increases with increasing **H/D ratio** for a constant volume and feed flow rate, while at the same time the concentrate grade decreases only to a minor extent (Yianatos, et al., 1988; Luttrell, et al., 1988).The **maximum H/D ratio** is determined by the **maximum carrying capacity**. This is because an increased collection zone height increases the recovery which could lead to the carrying capacity being the limiting factor.

Increasing the height of the collection zone increases the recovery until a **maximum** is achieved and maintained. At the same time the ash content of the concentrate passes through a **maximum** value (Parekh, et al., 1988).

In a study using fluorite and manganese, the recoveries decreased with a decrease in the column height. The fluorite grades increased with decreasing column length while the manganese grades remained unaffected (Ynchausti, et al., 1988).

1.3.1.3 Froth Depth

Froth depths in plant operations are typically between 0,5 and 1,5 m. **Hydraulic entrainment** appears to be eliminated close to the collection-froth zone interface when operating at moderate gas rates (Jg* < 1,5 cm/sec) (Yianatos, J.B., 1987). If selectivity between hydrophobic species is required or if high gas rates are to be used (Jg* > 2 cm/sec) then deep froths are desirable.

Using a laboratory flotation column it was shown that increasing the height of the froth zone increased the grade, whilst the recovery remained constant up to a point at which the recovery then decreased (Goodall, C.M. and O'Connor, C.T., 1989). It was also shown that the grade was linearly related to the mean residence time in the froth phase.

Ynchausti and co-workers (1988) showed that fluorite grades increased with increasing froth depth. For the manganese system considered it was not possible to increase the froth depth to more than 0.3 meters. No trend or relationship was predicted for the recovery. For coal flotation it was shown that the ash expulsion increased with increased froth depth. Again the recovery was affected only minimally (Parekh, et al., 1988).

To minimize entrainment in a column the superficial gas rates should be less than 1,5 to 2 cm/sec. The froth depth should be at least 1 m and the **superficial bias rates** should be 0,2 to 0,4 cm/sec (Yianatos, et al., 1987). These conditions should provide a good starting point for initiating column testing. Varying the froth depth could be used as a means of column control (Ynchausti, et al., 1988 b; Amelunxen, et al., 1988). By varying the froth depth the concentrate grades can be adjusted.

1.3.1.4 Feed Solids Percentage

High solids percentages of 30% to 50% can be used in the feed without affecting the concentrate grade (Feeley, et al., 1987). Kosick et al (1988) showed that the best operating feed density of galena at the Polaris Concentrator is between 45% and 50% solids. If the feed density exceeds 50% solids, the lead recovery decreases. Decreasing the feed density to 10% solids using a sulphide ore resulted in a decrease in recovery (Goodall, C.M. and O´Connor, C.T., 1989).

By increasing the solids density of the feed pulp the **maximum carrying capacity**, Ca, can be determined (Contini, N.J., 1988; Espinosa-Gomez, et al., 1988).

1.3.1.5 Feed Rate

By increasing the volumetric feed rate the solids through put can be increased. This results in a reduced mean solids residence time in the froth and collection zone (Goodall, C.M. and O´Connor, C.T., 1989).

Increasing the feed flow rate results in a decrease in recovery (Kosick, G.A. and Kuehn, L.,1988; Feeley, et al., 1987; Luttrell, et al., 1988). This reported decrease could be due to the solids residence time decreasing or the column being operated at its maximum carrying capacity. It has also been reported that the recovery passes through an optimum when the feed rate is increased (Goodall, C.M. and O´Connor, C.T., 1989). The recovery of an arsenopyrite system has been reported to be unaffected by a feed flow rate increase (Subramanian, et al., 1988). It seems that the solids residence time is longer than required and that the carrying capacity is not limiting.

The grade is reported to increase with increasing feed rate (Yianatos, et al., 1988), although this is then contradicted at other times when the grade is said to decrease with increasing feed rate (Goodall, C.M. and O'Connor, C.T., 1989; Subramanian, et al., 1988). It was also reported that the grade was unaffected by a change in the feed rate (Luttrell, et al., 1988; Kosick, G.A. and Kuehn, L.,1988).

1.3.1.6 Wash Water Rate, Position and Temperature.

It is known that wash water addition improves the grades of the concentrates (Kosick, G.A. and Kuehn, L., 1988; Nicol, et al., 1988; Ynchausti, et al., 1988; Parekh, et al., 1988). An optimum wash water addition rate is achieved in column cleaners at 1,2 to 1,5 **displacement washes** (Egan, et al., 1988).

It was reported that excess wash water decreases the recovery (Kosick, G.A. and Kuehn, L., 1988; Luttrell, et al., 1988). At a superficial wash water rate greater than 0,38 cm/sec the froth bed collapses. Parekh and co-workers (1988) found that for some coals the recovery decreases for J_w greater than 0,34 cm/sec. An optimum recovery is reported for a wash water addition rate of 0.23 cm/sec (Goodall, C.M. and O'Connor, C.T., 1989).

Kosick and Kuehn (1988) reported that the best position of the wash water distributor for the flotation of galena is between 7,5 cm and 10 cm below the surface of the froth bed. If the position is less than 7,5 cm below the froth bed surface, the grade decreases and the solids percentage will be lower. Misra et al. (1988) stated that 0,45 m needs to be available above the wash water distributor to provide a high grade clean coal. No loss in recovery was incurred. Hutchinson (1987) used tracer studies in a 2 phase study to determine the effect of the wash water distributor position. He found that at 0,19 m below the concentrate overflow lip the wash water effect is better than having the distributor closer to the concentrate overflow lip.

Reddy et al. (1988) state that superficial wash water velocities of 0,3 cm/sec give a steadier performance than at velocities of 0,18 cm/sec. For coal, using micro bubble flotation, ash rejection improved until **Jw** = 0,33 cm/sec. No further ash rejection was achieved by increasing the wash water rate (Luttrell, et al., 1988).

Wash water increases the froth stability and allows a deep froth bed to develop (Yianatos, J.B., 1987). The aim when adding wash water is to keep the superficial bias rate, **Jb**, positive. Typical wash water rates are 0,3 to 0,5 cm/sec.

In an investigation of the effect of temperature on the conventional flotation of pyrite it was found that the flotation rate of pyrite increases with an increasing temperature (O´Connor et al.,1984). This was ascribed to a reduction in the rate of mass transfer of pyrite from the pulp to the froth.

1.3.1.7 Interface level

Keeping the total column length fixed and lowering the interface level (increasing froth depth and at the same time reducing the collection zone length) improves the grade and also decreases the recovery (Amelunxen, et al., 1988). If the interface level is lowered to below 0.91 m the grade decreases rapidly.

For a galena float the grade does not improve if the interface level is lowered to below 40 cm, but the froth bed becomes unstable. If the interface level is raised too far (above about 20 cm) the grades decrease (Kosick, G.A. and Kuehn, L., 1988).

1.3.1.8 Bubble Size

The bubble size is a critical parameter in flotation. The size of the bubbles is a function of the air flow rate and chemical conditioning of the pulp. The bubble diameter in a column is typically 0.5 to 2 mm (Yianatos, J.B., 1987).

Bubble size can be estimated by using photographic techniques. This is not possible though for pulp. A method has been developed to estimate the bubble diameter using the gas holdup and phase velocities (Dobby et al., 1988)

A technique to measure the bubble size directly in 2 and 3 phase systems has been developed (Randall et al., 1989). Bubbles are sucked through a capillary tube and the size, velocity and number of bubbles are determined using optical sensors.

1.3.1.9 Particle Size

The particle size of the ore is important for the following reasons. Firstly, when operating the column with fine particles (< 20 μm) and with a high solids recovery into the concentrate, the bubbles carrying capacity could be a limitation (Espinosa, et al., 1988). Secondly, the particle residence time, τ_p, increases with decreasing particle size (Yianatos, J.B., 1987).

It is reported that column flotation is more suitable to fine particle flotation. For copper flotation the column achieved better recoveries than a conventional flotation cell in the particle size ranges smaller than 3 μm (Hu, W. and Liu, G., 1988). For molybdenum the flotation column recoveries were better for particle sizes smaller than 100 μm.

In column flotation of a flourite ore the grades were greater for all particle size fractions than for conventional flotation. For manganese the grades for the column flotation were better at $-105\ \mu m$, but worse at $+177\ \mu m$. The recoveries for manganese were also better for the minus 120 µm particles than in the conventional flotation cell (Ynchausti, et al., 1988). It was found that for coal flotation, particles of minus 75 µm are the ideal size (Misra, M. and Harris, R., 1988).

1.3.2 SUMMARY OF THE EFFECTS OF PHYSICAL PARAMETERS

Some effects of the physical parameters reported in the research literature is summarized in Table 1.2. This summary shows that there are some contradictions as to the effects of the physical parameters. At times the extent of the changes also differ.

	EFFECTS REPORTED	
	RECOVERY	GRADE
INCREASED AIR RATE	Optimum	Decrease
INCREASED COLLECTION ZONE	Increase (until max. is reached)	Decrease Optimum Unaffected
INCREASED FROTH DEPTH	Unaffected Decrease	Increase
INCREASED FEED SOLIDS %	Unaffected	Unaffected
INCREASED VOL. FEED RATE	Unaffected Decrease Optimum	Increase Decrease Unaffected
INCREASED WASH WATER RATE	Optimum	Increase

Table 1.2: A summary of the effects of some physical parameters.

1.3.3 THE EFFECT OF CHEMICAL PARAMETERS

Reagents are used in flotation to render the desired minerals hydrophobic and to improve bubble stability. Because the field of reagents has been extensively researched, this literature survey only covers experiences relating to column flotation.

1.3.3.1 Collector Addition

When adding collector to the feed of flotation columns the same trend (recovery increases while the grade decreases) as in conventional flotation is established but with superior results (Parekh, et al., 1988).

The addition of collector increases the recovery, especially that of the coarser size fraction. In the flotation of galena at the Polaris Concentrator (on Little Cornwallis Island, Northwest Territories, about 1450 km from the north pole) the collector addition enhances the grade by recovering the high grade coarse lead particles.

1.3.3.2 Frother Addition

The presence of frother in the column helped to produce a deep and stable froth, preventing coalescence of fine bubbles (Foot, et al., 1986; Kosick, G.A. and Kuehn, L., 1988).

It is also reported that the principal function of the frother is to optimize the bubble size relative to the mineral particle size (Egan, et al., 1988).

Experience with fine coal showed that at low air flow rates an increase in frother concentration caused a slight increase in ash content. At high air flow rates there was no change in ash content when the frother concentration was changed.

1.3.3.3 The Effect of pH

An important feature of pyrite flotation behavior is the observation that at an alkaline pH it undergoes extreme depression whereas in an acidic pH it is easily floated (Gaudin, 1932). This behavior of pyrite is closely connected with its susceptibility to rapid oxidation. When pyrite is oxidized in the presence of water, a film of ferric hydroxide forms on its surface, which leads to a high level of hydration.

Fuerstenau, et al, (1968) suggested that dixanthogen is the species which is primarily responsible for flotation. This followed on the pyrite recovery curves obtained from potassium ethyl xanthate and diethyl dixanthogen. The amount of dixanthogen that can form at high pH values is extremely low.

1.3.4 COLUMN DESIGN

In order to design a flotation column or to scale laboratory data up to plant scale, a model is needed. Some of the proposed models are given below.

1.3.4.1 Rate Constants

Dobby and Finch (1986) use the **collection efficiency**, E_k, which is directly related to the collection zone rate constant, **k**. The equation for the rate constant is given below:

$$k = \frac{1,5 \cdot v_g \cdot E_k}{d_b}$$

In order to get data for the determination of the rate constant the column was operated at a very high bias, (close to 100%) and the wash water was added half way between the column top and the feed entrance. This was to eliminate the cleaning zone so that the collection zone rate constant could be determined by assuming the recovery in the cleaning zone to be 100%. Problems with this method are that at high air rates the recovery by entrainment is higher than in normal column flotation (ie. column flotation with a cleaning zone) and entrainment is not completely eliminated.

Contini et al (1988) developed a method which enabled the collection zone rate constant and the froth zone recovery to be measured. This method used a specially modified flotation column in both counter-current and co-current mode.

At the Julius Kruttschnitt Mineral Research Centre Alford (1989) used a generalized form of the rate parameter relationship. This rate parameter is for the complete column and not only for the collection zone. This is because the pilot plant column was operated at a constant froth depth whilst maintaining a **positive bias**. The general rate parameter is given below:

$$k_i = \frac{C_i \cdot (v_g - v_{gmin}) \cdot a}{\mu}$$

1.3.4.2 Mixing Characteristics

The mixing in the collection zone of the column has been described in terms of the vessel dispersion number for solid particles, **Np**, which is summarized as follows (Yianatos, Column Flotation modelling and technology, 1989):

$$Np = \frac{Dp}{(u_i + u_p) \cdot L}$$

where $Dp = 2.98 \cdot dc \cdot 1,31 \cdot v_g \cdot 0,33 \cdot \exp(-0,025 \cdot S)$

The axial dispersion coefficient for solids was empirically determined.

The mixing in the collection zone of the column has also been determined using the vessel dispersion number for the liquid, **Nd** (Dobby, G.S. and Finch, J.A., 1986):

$$Nd = \frac{D}{u_i \cdot L}$$

Nd was determined by using fluorescein ($C_{20}H_{12}O_5$) dye and MnO_2 as tracers. The axial dispersion coefficient of the solid particles was reported to be the same as that of the liquid (Dobby, G.S. and Finch, J.A., 1986).

However, Goodall and O'Connor (1989) found that it was not possible to deduce the residence time distribution of the solids by studying the behaviour of the liquid. They also suggested that a tanks-in-series model for the collection zone would be an improvement on the use of vessel dispersion numbers.

1.3.4.3 Particle Residence Time

A method to determine the mean particle residence time is provided by Dobby and Finch (1986). The mean particle residence time is as follows:

$$\tau_p = \tau_l \frac{u_l}{(u_l + u_{sp})}$$

where u_{sp} is the particle slip velocity of the particles and is determined as follows:

$$u_{sp} = \frac{g \cdot d_p^2 \cdot (1 - \epsilon_g)^{2.7} \cdot (\rho_p - \rho_{susp})}{18 \cdot \mu_f \cdot (1 + 0.15 \cdot Re_p^{0.687})}$$

and τ_l is given by:

$$\tau_l = \frac{H \cdot (1 - \epsilon_g)}{J_{sl}}$$

1.3.4.4 Recovery Estimation

The recovery can be estimated in terms of the mixing characteristics, **Np**, the mean particle residence time, **τp**, and the rate constant, **k** (Dobby and Finch., 1986).

$$R = 1 - \frac{4 \cdot A \cdot \exp(0,5/Np)}{(1 + A)^2 \cdot \exp(0,5\ A/Np) - (1 - A)^2 \cdot \exp(-0,5\ A/Np)}$$

where

$$A^2 = 1 + 4 \cdot k \cdot \tau p \cdot Np$$

Goodall and O´Connor (1989) proposed a tanks-in-series model rather than the dispersion model used above.

1.3.4.5 Carrying Capacity Limitation

The **maximum carrying capacity** could be a limiting factor to a columns feed capacity. A theoretical relationship was developed by Yianatos (1987). The carrying capacity, in this case referred to as the superficial floated particle rate, J_{pf}, is estimated as follows:

$$J_{pf} = \frac{8.47 * 10^{-2} \cdot Jg^{*} \cdot \mu_p \cdot d_{80}}{d_b}$$

The maximum carrying capacity, **Ca** (g/min/cm2), can also be estimated from the following semi-theoretical relationship (Espinosa, et al, 1988).

$$Ca = \frac{60 \cdot K_1 \cdot \tau \cdot d_{80} \cdot \mu_p \cdot Jg}{d_b}$$

This relationship is essentially the same as the theoretical relationship mentioned above. From this equation it can be seen that the air flow rate (Jg) increases the carrying capacity proportionally.

The maximum carrying capacity can also be determined from pilot unit experiments. This is done by increasing the solids feed rate to the column until the concentrate mass rate reaches a maximum.

CHAPTER TWO

OBJECTIVES OF RESEARCH

The objective of this research programme was to determine the effects of physical and chemical parameters on the column flotation cell performance in the flotation of pyrite and then to propose hypotheses to explain the effects observed.

To achieve this objective this research was structured as follows:

1) Design and commission an instrumented laboratory column flotation cell;
2) Carry out a systematic study of the effect of the following physical parameters on the flotation of pyrite in the flotation column:
    (i) Air flow rate;
    (ii) Sparger type;
    (iii) Collection zone length;
    (iv) Froth depth;
    (v) Feed solids percent;
    (vi) Volumetric feed rate;
    (vii) Wash water addition rate;

(viii) Wash water distributor position;
(ix) Wash water temperature;
(x) Feed point.

3) Carry out a systematic study on the effect of the following chemical parameters on the flotation of pyrite in the flotation column:

(i) Leaching of the ore;
(ii) Conditioning procedure;
(iii) Collector dosage;
(iv) Frother dosage.

4) Design and commission an instrumented portable pilot scale column flotation cell for plant trials;

5) Carry out plant trials in order to evaluate the suitability of using a pilot plant column cell and to investigate the viability of column flotation.

CHAPTER
THREE

EXPERIMENTAL METHODS

In this chapter the details of the design of the laboratory flotation column, the layout of the experimental rig and the experimental procedures adopted for the test work are described.

3.1 DESIGN OF THE LABORATORY FLOTATION COLUMN CELL

To study the effects of physical and chemical parameters on column flotation a laboratory flotation column cell with all the auxiliary equipment had to be designed and built.

The objective was to design an experimental rig that could be run and monitored on a continuous basis. The design had to make provision for easy adjustment to physical and chemical parameters when required.

3.1.1 COLUMN SIZING

The sizing of the laboratory flotation column could not be done using the kinetic models proposed for column design since no kinetic data was available for the Unisel test ore. Therefore an estimate using typical column data was used for the design of the laboratory flotation column. The procedure for the design is described below.

3.1.1.1 Column Carrying Capacity

To initiate the laboratory column design a standard perspex tube with an inner diameter of 54 mm was chosen. This transparent perspex tube allowed the internal operation to be monitored visually. The next step was to calculate the maximum carrying capacity for a column of this size. The theoretical method proposed by Yianatos (1987) was used for this purpose. The maximum carrying capacity was then calculated to be 107 g/hr/cm2 (APPENDIX A). Typical design and operating conditions used for the calculations are listed in Table 3.1.

Superficial Gas Rate	1 - 3 cm/sec
Superficial Pulp Rate	1 - 2 cm/sec
Superficial Wash Water Rate	0.3 - 0.5 cm/sec
Superficial Bias Rate	0.1 - 0.2 cm/sec
Froth Depth	> 100 cm
Average Bubble Size	0.05 - 0.2 cm
Height to Diameter Ratio	> 10/1

Table 3.1: Typical Design and Operating Conditions. (Yianatos, J.B., 1987)

The volumetric feed rate required for the column was calculated to be about 1.7 l/min (APPENDIX A). A peristaltic feed and tailings pump capable of this flow rate was purchased.

3.1.1.2 Solids Residence Time

The recovery in the column flotation cell is a function of the solids residence time in the collection zone while the concentrate grade is a function of the solids residence time in the cleaning zone. The column design had to therefore allow the solids residence time in the collection and froth zone to be varied. This was made possible by the 50 cm long flanged perspex sections used for the column. The column length could therefore be increased from 1 m to 8 m. The feed could also be introduced at any point along the length of the column.

3.1.2 BUBBLE SIZE MEASUREMENT

The bubble sizing technique (Randall, E.W., et al. 1989) used in the experimental work is described below. The bubble sizing equipment is shown in figure 3.7.

The air bubbles generated in the flotation column are drawn into a capillary tube by means of a vacuum. The bubbles then pass two phototransistors. A signal is generated which leads to the production of two pulses per bubble. From these pulses the velocity and volume of the bubbles can be calculated.

A microprocessor is used to time the pulses, store the result in memory together with the real time of the event and then to transmit this data to a personal computer. The personal computer then performs the data analysis.

The volume of the bubbles is determined as a fraction of the total volume of bubbles collected. The assumption is made that the liquid film on the walls of the capillary tube is constant. At a constant pressure and temperature this assumption is valid.

The advantage of this system is that it is capable of directly measuring the size of bubbles in two- and three-phase systems. The results are highly reproducible and the system is easy to operate.

3.1.3 AIR SPARGER DESIGN

The **air sparger** must be able to produce small and uniform sized bubbles. Three different types of spargers were built. These were the **sintered glass disc** sparger, **SGDS**; the filter cloth sparger, **FCS**; and the United States, Bureau of Mines sparger, **USBM**. The designs of the spargers are discussed below. The results for the tests done on the spargers are given in section 4.3.1.1. The bubble data is recorded in APPENDIX B.

An air rotameter followed by a needle valve was initially used to control and monitor the air flow rate. The air pressure in the rotameter decreased as the air flow rate increased deeming this inadequate. A second needle valve was therefore installed in front of the air rotameter. The pressure in the air rotameter was monitored at the outlet of the rotameter. Now the air rotameter could be operated at a set pressure while the air flow rate could be varied. The air rotameter was calibrated at 400 kPa.

3.1.3.1 Sintered Glass Disc Sparger

The **sintered glass disc** sparger (SGDS) was made of 6 sintered glass discs (porosity number 4 and diameter 1 cm) which were fitted into holes in the wall of a tube. A sketch of the design is given in Figure 3.1. The total sintered disc surface area was 4,7 cm^2.

This provides a ratio of column cross sectional area to sparger area of 4,86.

The fine porosity glass discs were chosen so that small bubbles of less than 2 mm could be produced. The sintered glass discs were fitted at an angle of 45 degrees to prevent solids from settling on the discs and blocking them.

3.1.3.2 Filter Cloth Sparger

The filter cloth sparger (FCS) was a "sock" made from filter cloth which was pulled over a 2 cm diameter tube. The tube had 2 mm holes drilled into the walls (Figure 3.2). The total surface area was 23 cm^2. This gives a ratio of column cross sectional area to sparger area of 1.00.

The large filter cloth surface area was intended to keep the pressure drop across the filter cloth low. The large area was also to prevent the mean bubble size produced from increasing excessively when the air flow rate was increased.

3.1.3.3 U.S.B.M Sparger

The external bubble generator developed by the United States Bureau of Mines (U.S.B.M.) was used as a third type of sparger.

The U.S.B.M made for the laboratory column consisted of a mixing chamber packed with 1 mm and 2 mm glass beads. The mixing chamber was 30 mm in diameter and 250 mm long (Figure 3.3). Clean water at a desired frother concentration and air were introduced into the mixing chamber. The water-air mixture was then released into the column through two 0,9 mm nozzles facing downwards at 45 degrees.

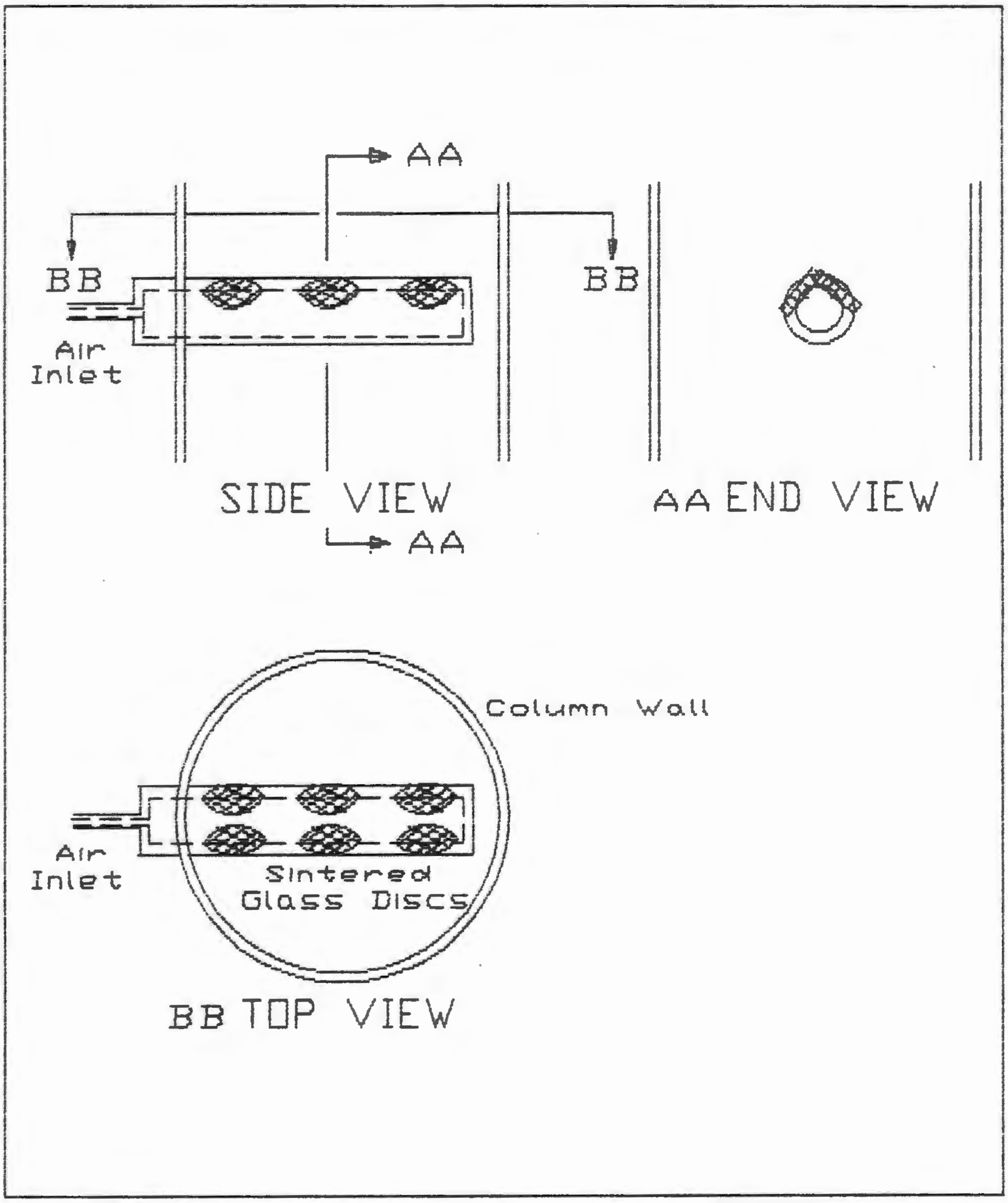

Figure 3.1: Sketch of the Sintered Glass Disc Sparger

The operation of this sparger was slightly different to the operation of the original U.S.B.M. sparger which operated at high pressures of up to 6 bar. The sparger operated at a pressure which was as low as possible (< 4 bar) in order to keep the volumetric air flow rate high. The high volumetric air flow rate ensured high velocities through the mixing chamber which facilitated the mixing of the air and water. In order to operate at a low pressure the number of nozzles had to be increased. The water flow rate required depended on the amount of frother used and the bubble size required.

3.1.4 LEVEL CONTROLLER

The level controller (A picture of the level controller prototype is shown in Figure 3.8) which was used to keep the pulp-froth interface at a set point was based on a system developed by Ormrod (1984). The system relies on the differences in the conductivity of the froth and the pulp.

A current is passed through a resistance wire which passes from the froth into the pulp. Parallel to the resistance wire is a common electrode. The two wires are therefore connected by the pulp conductivity. The froth conductivity is much higher and therefore negligible. The potential measured by the electrode thus increases linearly as the pulp-froth interface rises.

The level is then corrected by varying the tailings pump rate.

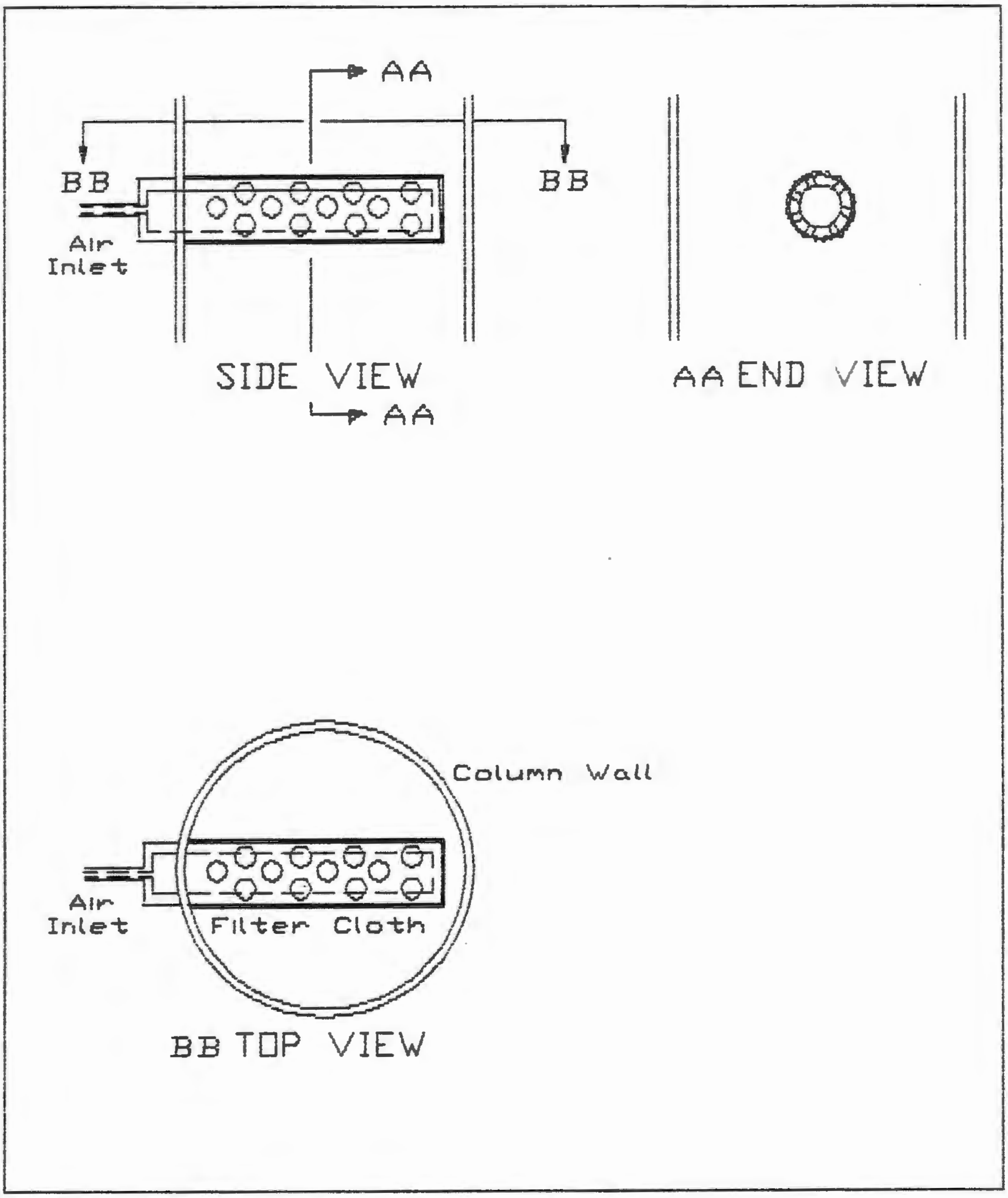

Figure 3.2: Sketch of the Filter Cloth Sparger.

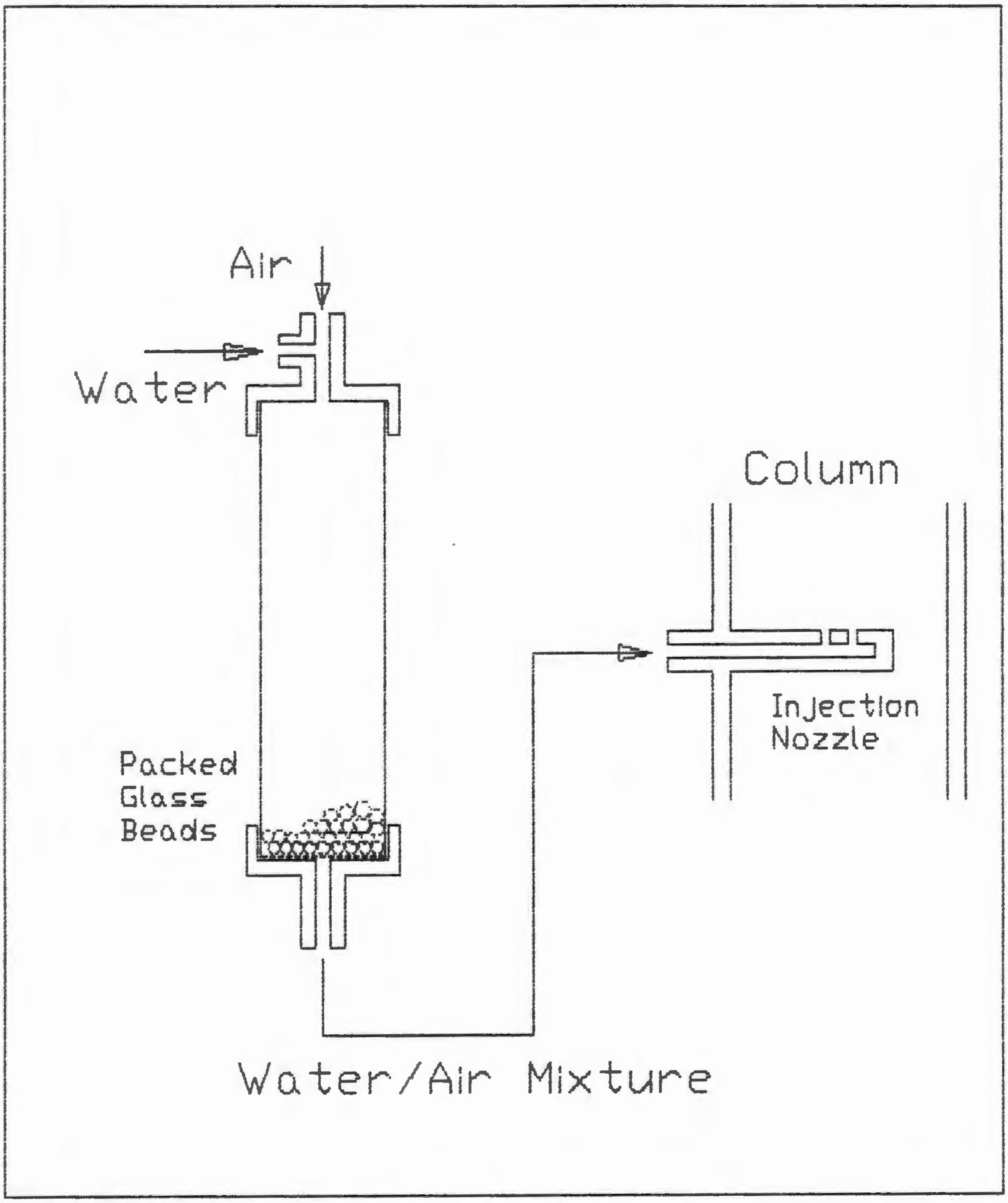

Figure 3.3: Sketch of the United Stated, Bureau of Mines Sparger.

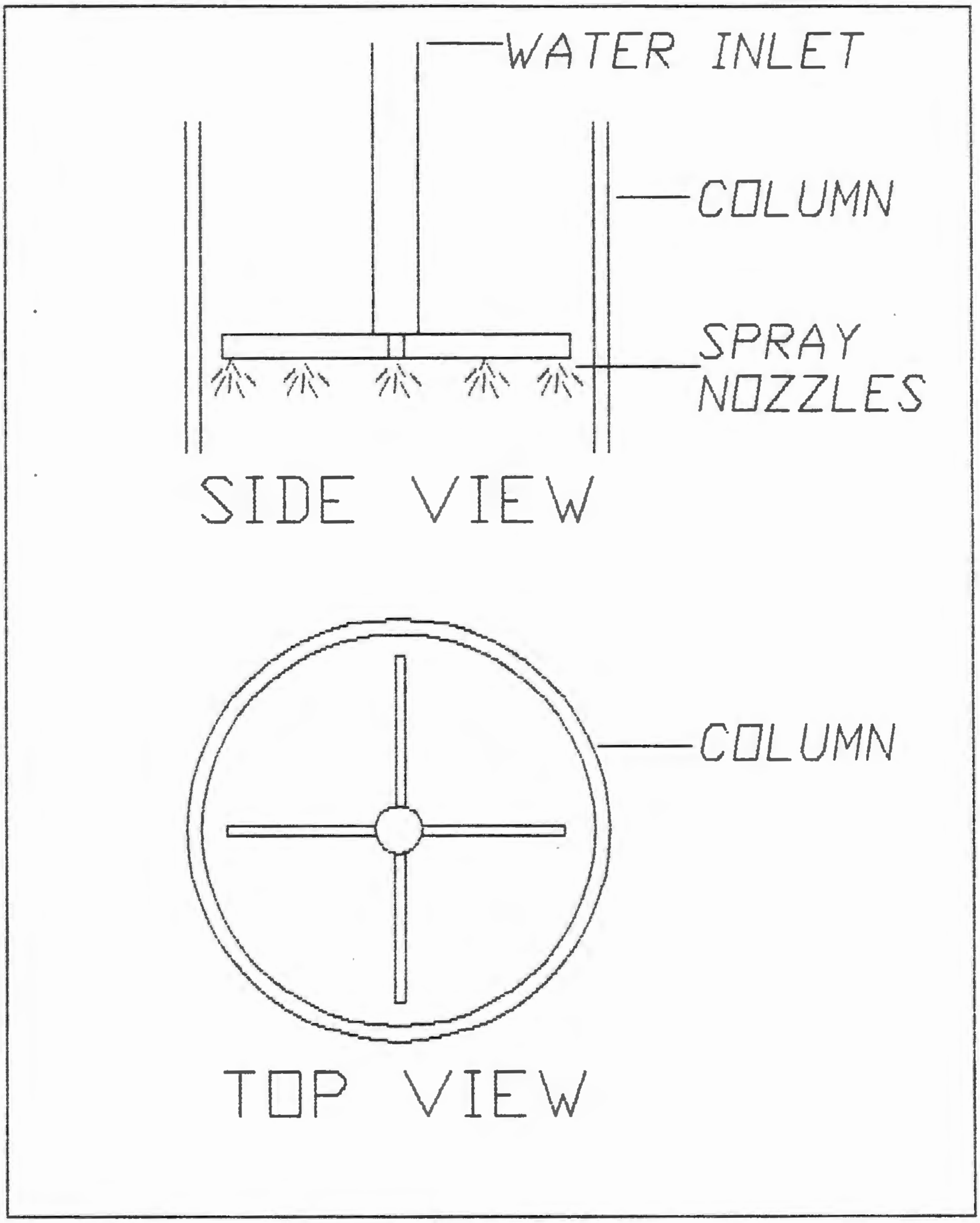

Figure 3.4: Sketch of the Wash Water Distributor.

3.1.5 WASH WATER DISTRIBUTOR

The wash water distributor was made of copper tubing which formed a cross (Figure 3.4) at the bottom of the supply tube. The distributor could be moved up or down by simply pushing the supply tube up or down within a tight fitting gland.

Holes were drilled into the cross to distribute the wash water evenly in the froth.

3.2 EXPERIMENTAL SETUP

The experimental rig essentially consists of 3 sections. These are the feed section, the auxiliary equipment section and the sampling and monitoring section. Each piece of equipment is discussed below. A schematic diagram of the experimental rig is given in Figure 3.5. Also included (Figure 3.6) is a photograph taken of the experimental rig.

3.2.1 Feed Section

The feed section consists of a 350 liter holdup tank fitted with a 1,5 hp three phase motor. The propeller used is a marine-type propeller set in an angular off-center position. No baffles were thus required. The agitator keeps the pulp well mixed.

The pulp is then pumped to the conditioning tank with a variable speed peristaltic pump. At the same time the reagents are added to the conditioning tank with the reagent peristaltic pump. The reagents are pumped from a beaker that is well mixed. The conditioning tank is a modified 35 liter Denver laboratory batch flotation cell. The air to the cell is not used.

The pulp is then pumped to the column with a second variable speed peristaltic pump. To prevent any pulsation in the column due to the peristaltic action a surge chamber was installed.

3.2.2 Auxiliary Equipment Section

The auxiliary section consists of the air, sparger water and the wash water supply, the level controller and the bubble sizing equipment.

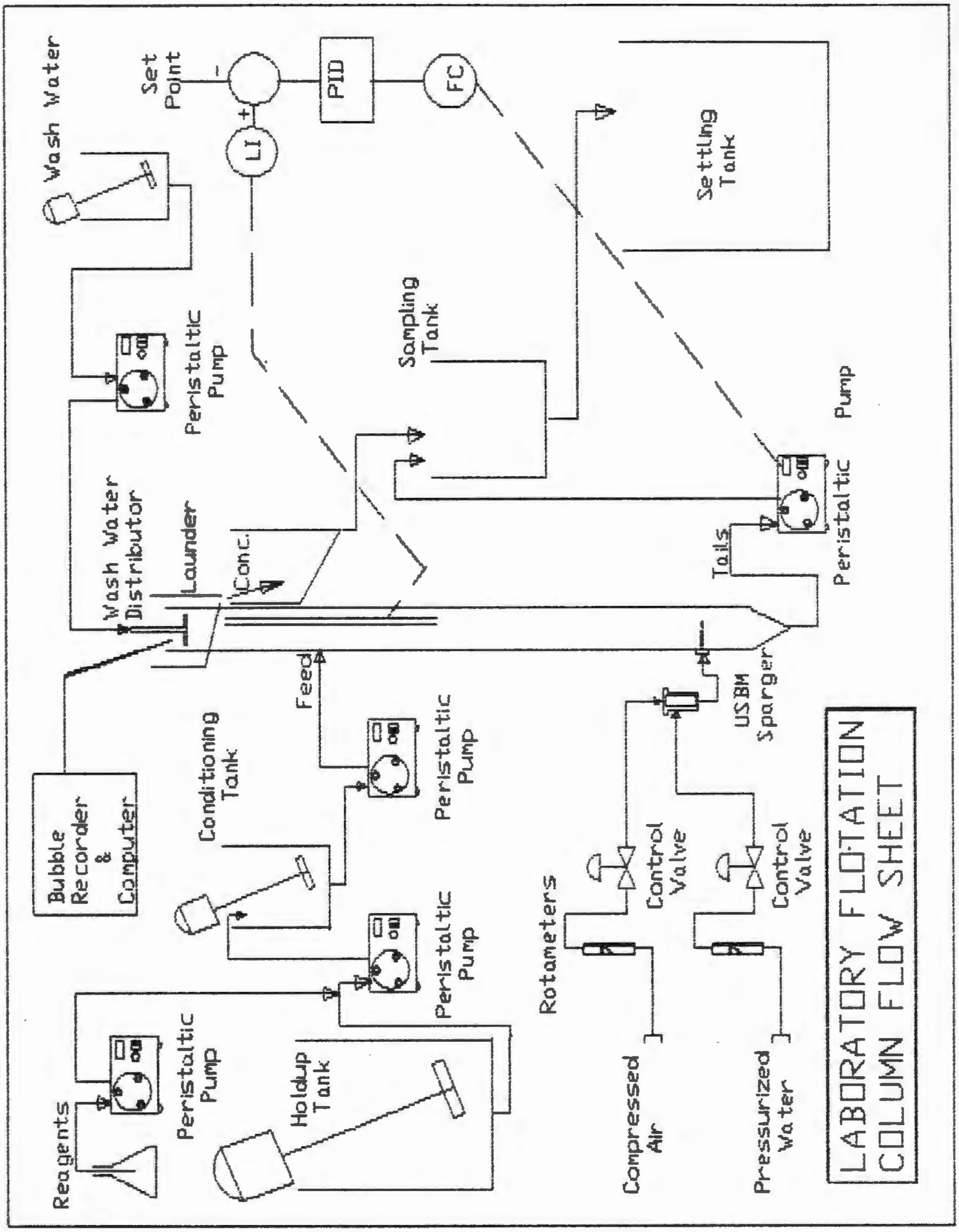

Figure 3.5: A schematic Diagram of the Experimental Rig

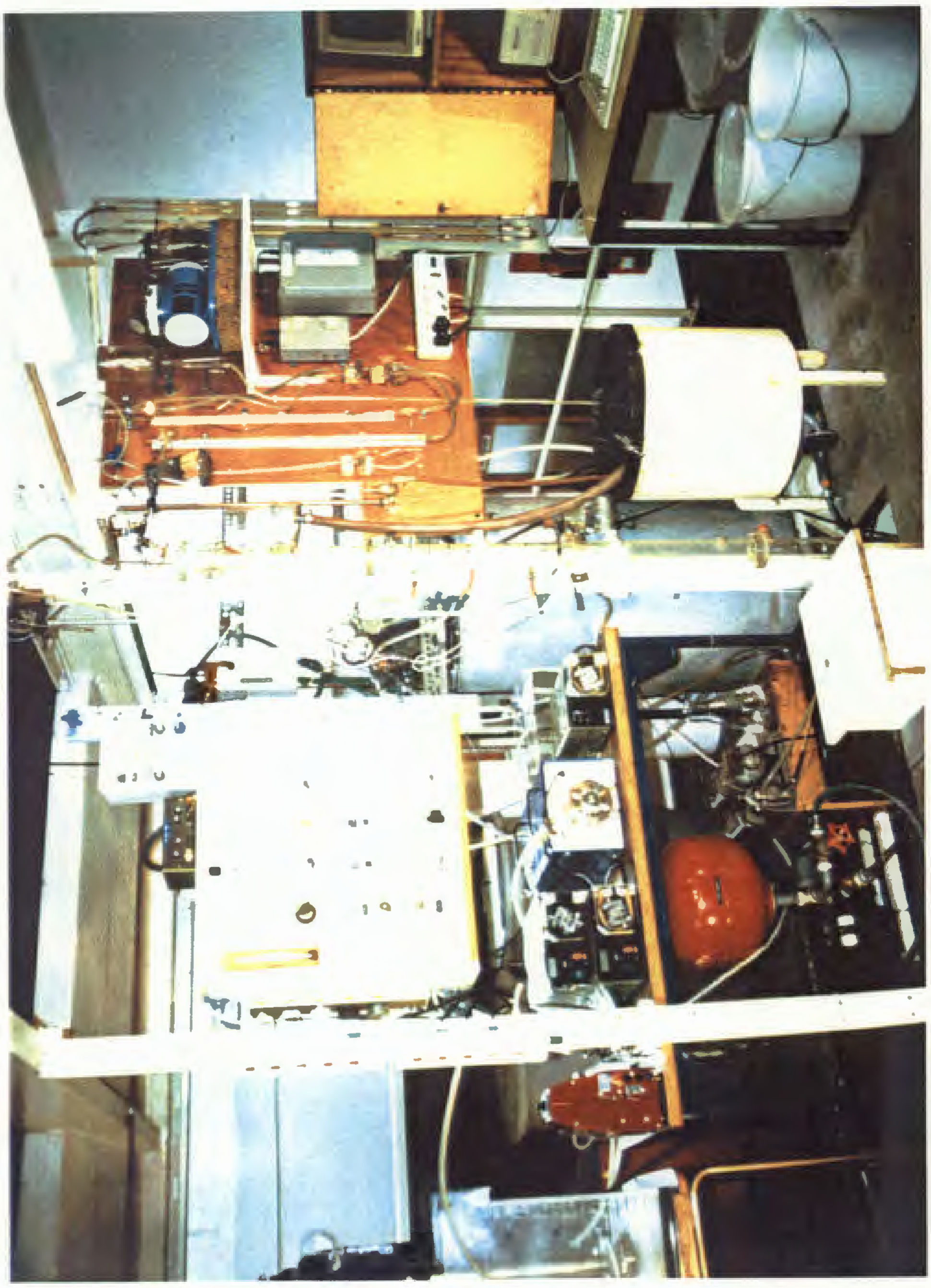

Figure 3.6: A picture of the complete laboratory rig

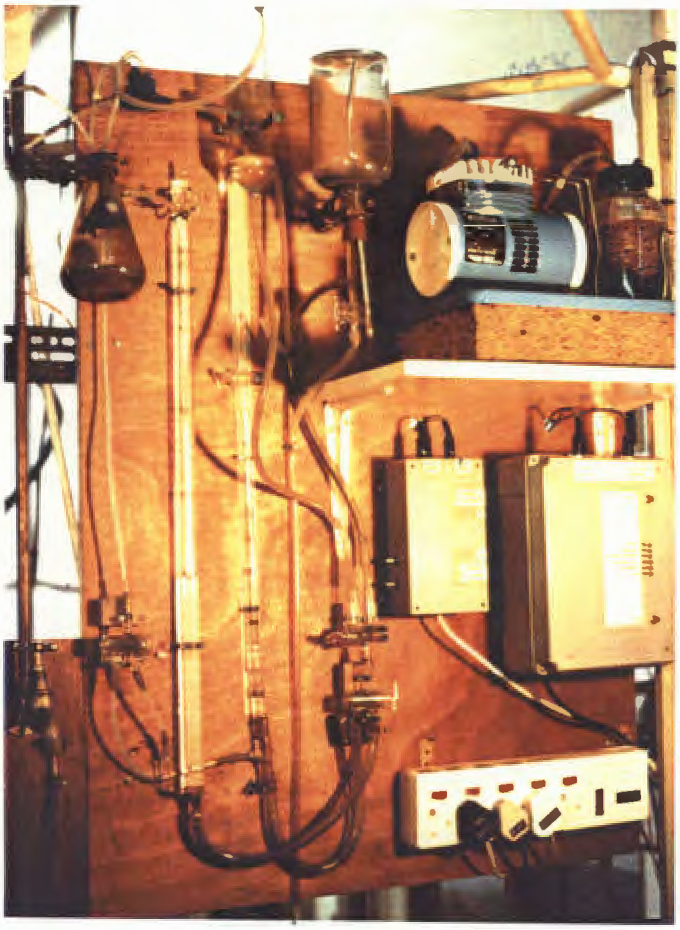

Figure 3.7: Bubble Sizing Equipment

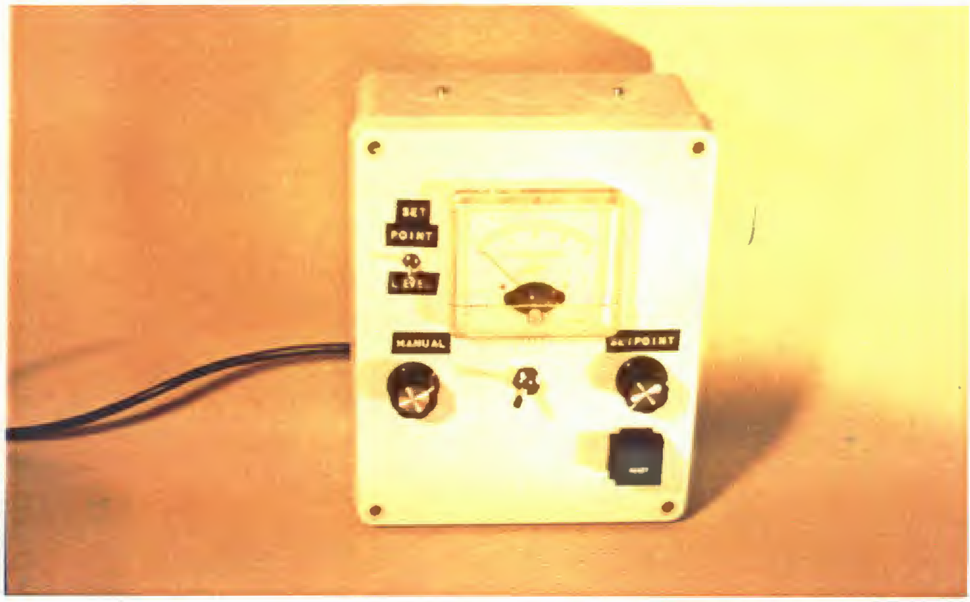

Figure 3.8: Level Controller

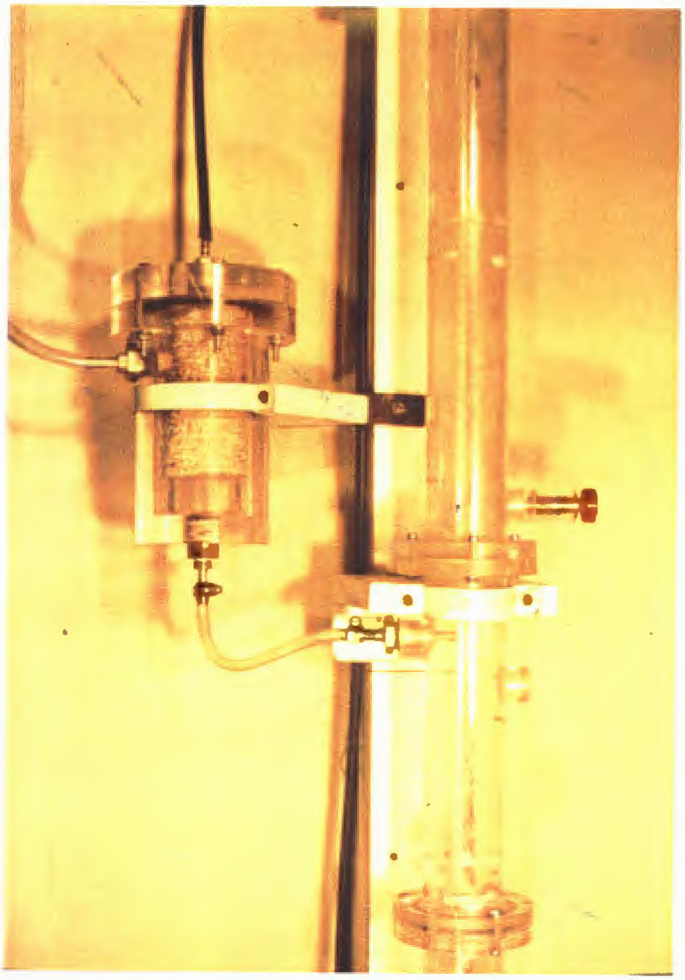

Figure 3.9: U.S.B.M. Type Sparger

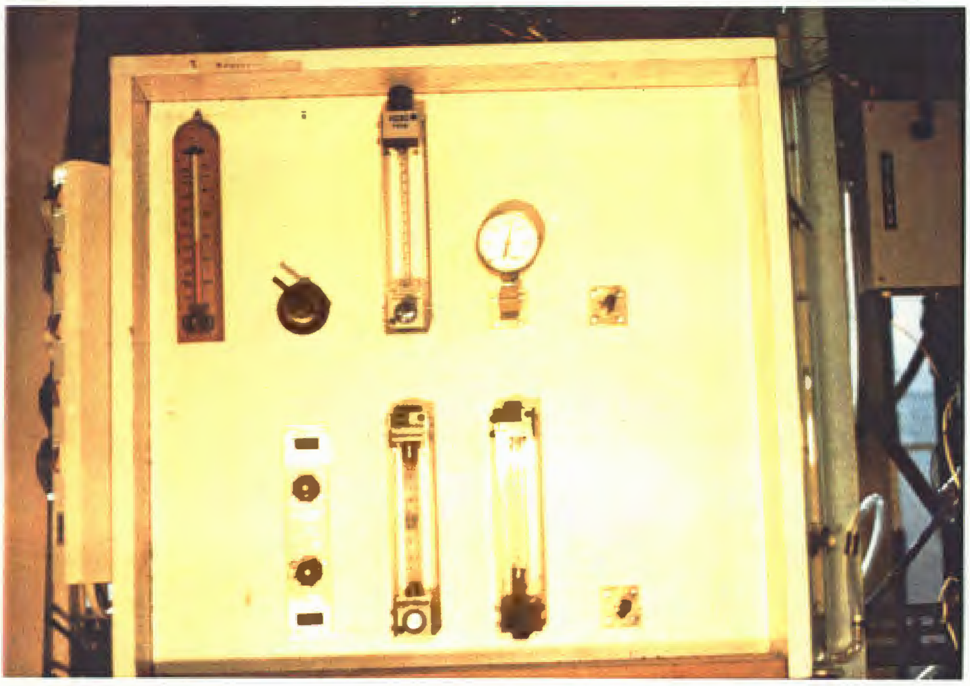

Figure 3.10: Air and Water Rotameter

The air is supplied from a compressed air line. The air flow is then controlled and monitored by two needle valves and an air rotameter (Figure 3.10). The air rotameter was calibrated at 400 kPa. The pressure at the outlet of the air rotameter was therefore always kept at 400 kPa. The rotameter could thus be used to monitor and adjust the air flow rate without the pressure affecting the rotameter reading.

A high pressure water pump is used to pump the sparger water to the USBM sparger. The water is pumped from a water tank which is also used to make up the frother concentration of the sparger water. The flow rate is controlled and monitored with a rotameter and a control valve.

The wash water is pumped from the wash water tank to the wash water distributor with a variable speed peristaltic pump. The desired amount of frother is added to the wash water in the tank.

The level controller described above can be used to vary the pulp-froth interface over a height of 150 cm. The level is controlled by varying the speed of the tailings peristaltic pump while keeping the feed pump at a set speed. A surge chamber is also included in the tailings line to prevent pulsating in the column due to the peristaltic action.

The bubble sizing equipment is installed next to the column so that the bubble sizes can be measured.

3.2.3 Sampling and Monitoring Section

Samples can only be taken from the feed line before and after the experimental run. The concentrate and tailings can be sampled at the same time. To make this task easier the tailings are pumped back up to the level of the launder into the sampling tank. The pulp not sampled drains to a settling tank.

The air flow rate is monitored by using an air rotameter as described above. The flow rates of the pulp are measured physically while the wash water and reagent pumps are calibrated. The pressure at the top and bottom of the column is monitored using mercury manometers.

3.3 EXPERIMENTAL PROCEDURE

The ore, UNISEL ore, is kept submerged under water in airtight containers to prevent excessive oxidation. The ore is then transferred to the holdup tank without prior drying. The holdup tank is filled before a particular set of experiments to ensure that the effect of a particular parameter can be tested with identical feed condition. The pulp density is established before each set of experiments and then adjusted if necessary.

Before starting up the flotation column the air rate is set. The column is then filled with water up to the height where the pulp-froth interface level is intended to be. The controller is then switched on.

The reagents decided upon for a specific experimental run are made up as well as the wash water with the desired frother concentration. The frother is only added to the wash water if it is required for additional froth stability.

The ore from the holdup tank is now pumped together with the mixed reagents to the conditioning tank. Once the conditioning tank is filled so that a specific conditioning time can be maintained the feed pump is started at a preset rate. Before feeding the ore to the column a sample is taken so that the exact feed rate can be established.

The wash water pump is started at the same time as the feed pump is started. This enables a deep and stable froth to develop as soon as possible after start up.

In the early stages of each run it may be necessary to adjust the set point, which is a voltage, to correct any changes in conductivity in the pulp or in the bubble bed.

Once the column has been operating for a few minutes the conductivity will not change any more and the level controller will keep the level at the set point.

The column is operated for 30 minutes to ensure that steady state is reached (Determined in section 4.2.2). After 30 minutes the concentrate is sampled for 5 minutes and the tailings for 30 seconds (Determined in section 4.2.1). At the same time the air flow rate and the pressure at the top and bottom of the column are recorded.

The pulp densities and solid mass flow rates of the feed, concentrate and tailings samples are then determined. The samples are also analysed for their sulphur content.

From this data the sulphur recovery is calculated. This calculation and any other calculation used are shown in APPENDIX C. The experimental data and the calculated results are given in APPENDIX D.

CONVENTIONAL BATCH FLOTATION PROCEDURE

For the batch flotation procedure 3 liters of feed to the column was taken and transferred to the conventional batch cell. The feed conditions were thus identical for the column and for the batch cell. The reagent dosage rates and conditioning times were therefore also the same.

The agitator was set at 1500 rpm. Concentrate samples were taken for the following time periods: First sample for 1 minute and the five consecutive samples for 2 minutes each.

CHAPTER FOUR

RESULTS

In this chapter the results obtained from the experimental test work conducted in the laboratory are presented. The first part describes the ore used. The second part presents the results of a sensitivity analysis as well as steady state and reproducibility data. The results of comparing a conventional batch laboratory cell and the column flotation cell are also presented in the second part. The third and fourth part of this section presents the results of the physical and chemical parameters studied.

The laboratory results and calculated values are tabulated in APPENDIX D. Particle size data is reported in APPENDIX E while bubble size data is reported in APPENDIX B.

4.1 DESCRIPTION OF ORE USED

The ore used in the test work was provided by GENCOR's Unisel mine. The ore is a pyrite ore with a sulphur content of 2% and a gold content of 10 g/ton. Table 4.1 gives the particle size distribution of the ore while Table 4.2 gives the percentage sulphur in different size ranges. The particle size analysis was done using a Malvern particle size analyser and sieves. The results obtained with the two methods compare well.

The sulphur content was determined for four size ranges. It was found that the coarser particles (+ 75 µm) had the highest sulphur grade. Most of the total sulphur content however occurs in the minus 38 µm size range. This information is presented graphically in Figure 4.1.

MALVERN DATA		SIEVE DATA	
Size *(Microns)*	**Weight below size** *(%)*	**Size** *(Microns)*	**Weight below size** *(%)*
188.0	100.0	188.0	100.0
87.2	99.5	75.0	96.1
53.5	91.2	53.0	90.0
37.6	80.7	38.0	75.0
28.1	68.4		
21.5	55.0		
16.7	42.2		
13.0	31.9		
10.1	22.7		
7.9	14.8		
6.2	9.7		
4.8	6.5		
3.8	3.4		
3.0	1.2		
2.4	0.6		
1.9	0.3		

Table 4.1: Particle size distribution for Unisel ore.

Size	% Sulphur below size
(Microns)	*(%)*
188	100.0
106	95.8
75	76.2
53	60.5
38	51.1

Table 4.2: Sulphur distribution for Unisel ore.

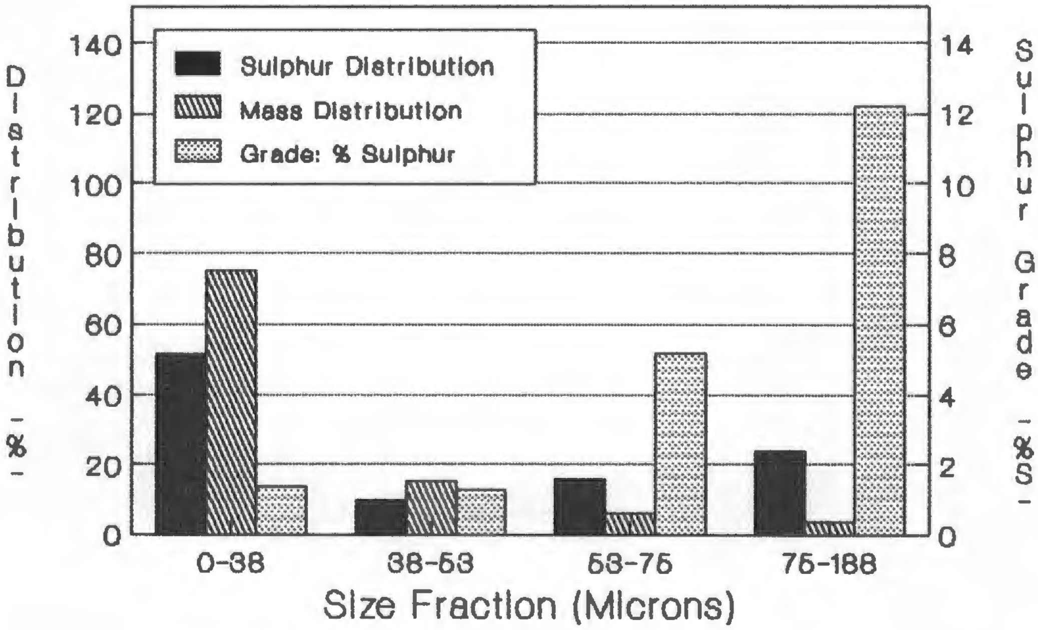

Figure 4.1: Percent sulphur, percent mass and sulphur grades for 4 size fractions of the feed.

4.2 REPRODUCIBILITY AND ANALYSIS OF DATA

4.2.1 Sensitivity Analysis

In order to determine the validity of the experimental method employed, a sensitivity analysis was done. This was divided into two parts.

The first part was the determination of the variance in the concentrate sulphur grades and recoveries. The 95% confidence interval and the "student" t distribution were used in this case. To determine the 95% confidence interval a run was done for three hours and the concentrate was sampled every 15 minutes (Figure 4.2).

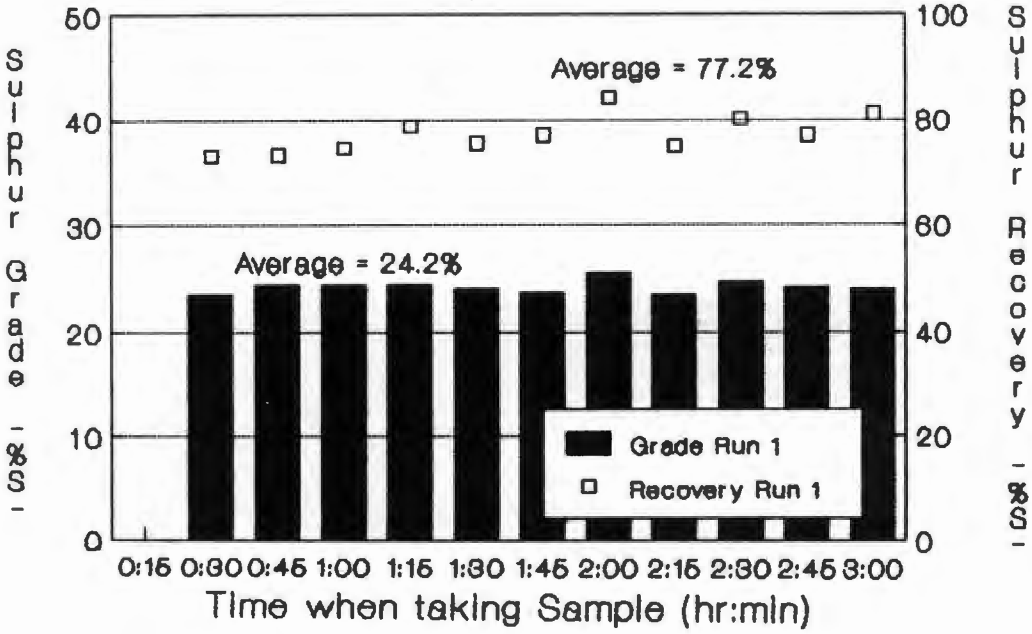

Figure 4.2: Sulphur grade and recovery steady state determination for times between 15 minutes and 3 hours (Run 1).

The arithmetic mean and sample standard deviation were then calculated for the sulphur grades and recoveries.

The 95% confidence limits for this small sample were calculated using the following equation.

$$95\%\ \text{confidence limits} = X \pm t_c \frac{S}{N-1}$$

$$\text{Therefore interval} = X_{max} - X_{min}$$

The values determined are listed in Table 4.3.

	Run	X	s	tc	Xmax	Xmin	Interval
Sulphur Grade	1	24.16	0.6	2.23	24.58	23.74	0.84%
Sulphur Recovery	1	77.17	3.45	2.23	79.60	74.74	4.86%

Table 4.3: The values used for the determination of the 95% confidence boundary.

From Table 4.3 it can be seen that the 95% confidence interval for the sulphur recovery was nearly six times larger than for the sulphur grade. Another run (Figure 4.3) was therefore done to reduce the 95% confidence interval for the sulphur recovery. The concentrate samples were taken for five minutes instead of one minute (Table 4.4). The 95% confidence interval for the recovery improved from 4.86% to 1.94%. It was thus decided to take the concentrate sample for five minutes (or more) to improve the accuracy of the experimental test work.

The second part of the sensitivity analysis determined the sensitivity of the calculated recovery as a result of deviations in the values used for the calculation. The equation used to calculate the recovery is given below.

$$\text{Percent Sulphur Recovery} = \frac{cC}{cC + tT} * 100$$

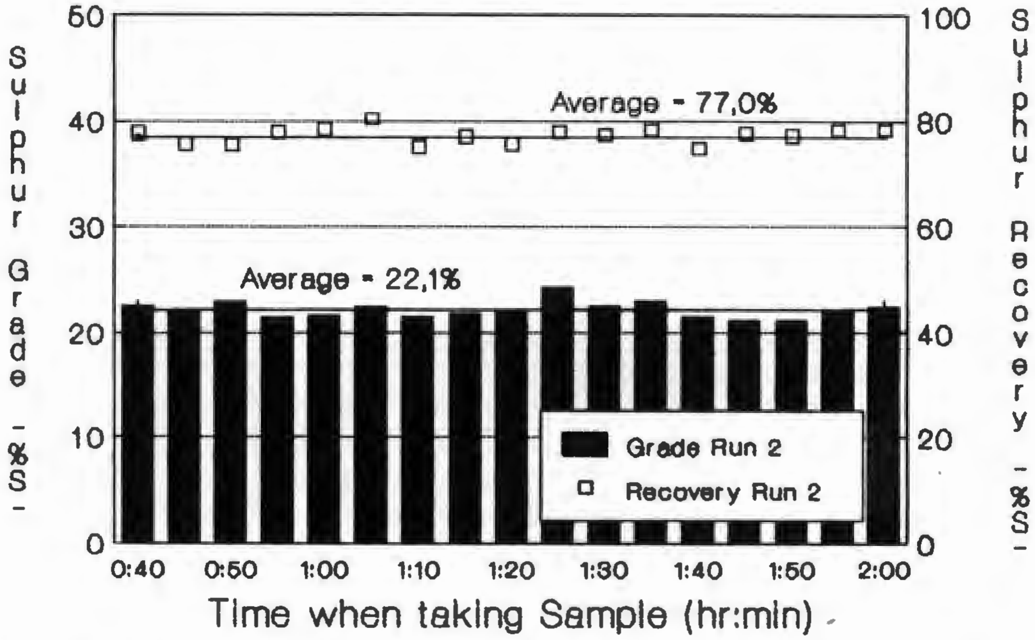

Figure 4.3: Sulphur grade and recovery sensitivity analysis for times between 40 minutes and 2 hours.

	Run	X	s	tc	Xmax	Xmin	Interval
Sulphur Grade	2	22.10	0.80	2.12	22.64	21.56	1.08%
Sulphur Recovery	2	77.04	1.45	2.12	78.01	76.07	1.94%

Table 4.4: 95% Confidence boundary values for improved sampling techniques.

The variances in the sulphur recovery were determined from the deviations in masses of samples collected, sample sulphur content and sulphur assaying inconsistencies. The results are given in Table 4.5.

The maximum deviation in the calculated recovery is only 1%. Any variances due to deviations in sulphur assays are thus less than the possible variations due to the systems sensitivity and are therefore negligible. It can also be seen that all the sample deviations have a similar influence on the calculated sulphur recovery. There is therefore no specific sample mass, or sample sulphur content or assaying technique that is more responsible than another for deviations calculated.

	Sample Deviation				Assay Deviation	
	C	c	T	t	c	t
Arithmetic Mean	8.7	22.0	106.5	0.54	22.0	0.54
Sample Standard Deviation	0.4	0.8	4.4	0.03	0.1	0.01
Variance in Recovery	0.8%	0.6%	0.7%	1.0%	0.1%	0.3%

Table 4.5: Variance in the recovery due to deviations in the samples and the sulphur assays.

4.2.2 Steady State Determination

In continuous column flotation test work it is essential to reach steady state before taking samples. This is unlike conventional laboratory batch flotation tests where the concentrate is collected from the beginning of the run. It was therefore necessary to determine how long after start up steady state is reached.

To determine when steady state is reached concentrate and tailings samples were taken continuously from the start of a run. This test was done in duplicate. The concentrate and tailings were sampled every five minutes from start up until 30 minutes elapsed. The samples were then weighed and analysed for their sulphur content. The sulphur grades and recoveries were then plotted against time (Figures 4.4 and 4.5) and used to determine when steady state was reached.

The steady state criterion decided upon was that the sulphur grade and recovery of samples taken at steady state must lie within the 95% confidence interval determined in section 4.2.1.

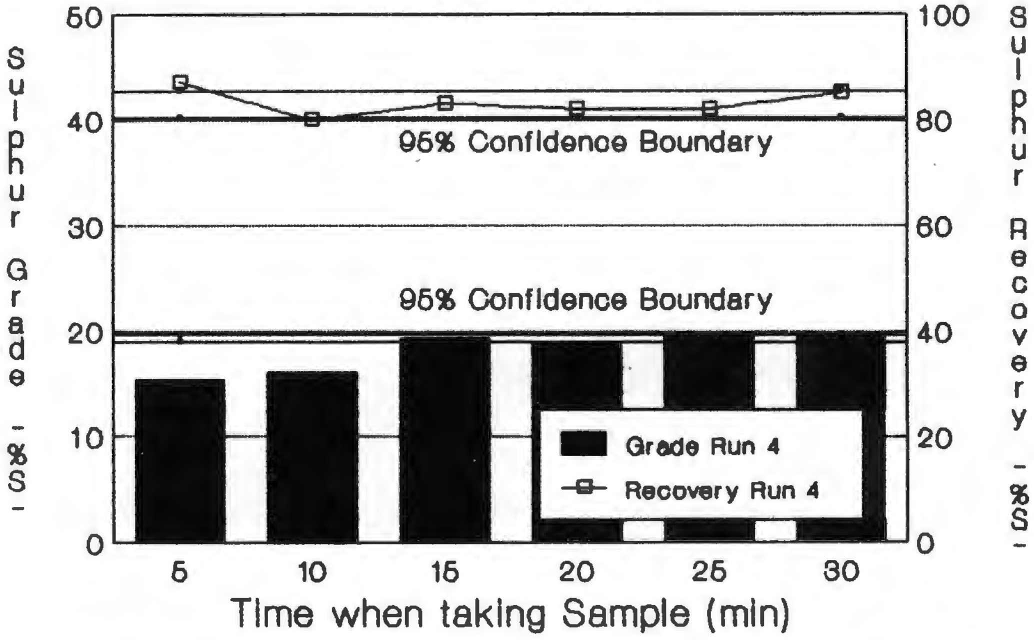

Figure 4.4: Sulphur grade and recovery versus time to determine when steady state is reached in the column (Run 3).

From figure 4.4 and 4.5 it can be seen that after fifteen minutes steady state was reached. This was confirmed by doing another run (Figure 4.6) and taking samples every minute from fifteen to twenty two minutes after start up. Again the sulphur grades and recoveries were within the 95% confidence interval determined.

Although only fifteen minutes were required to reach steady state it was decided to wait thirty minutes before taking samples.

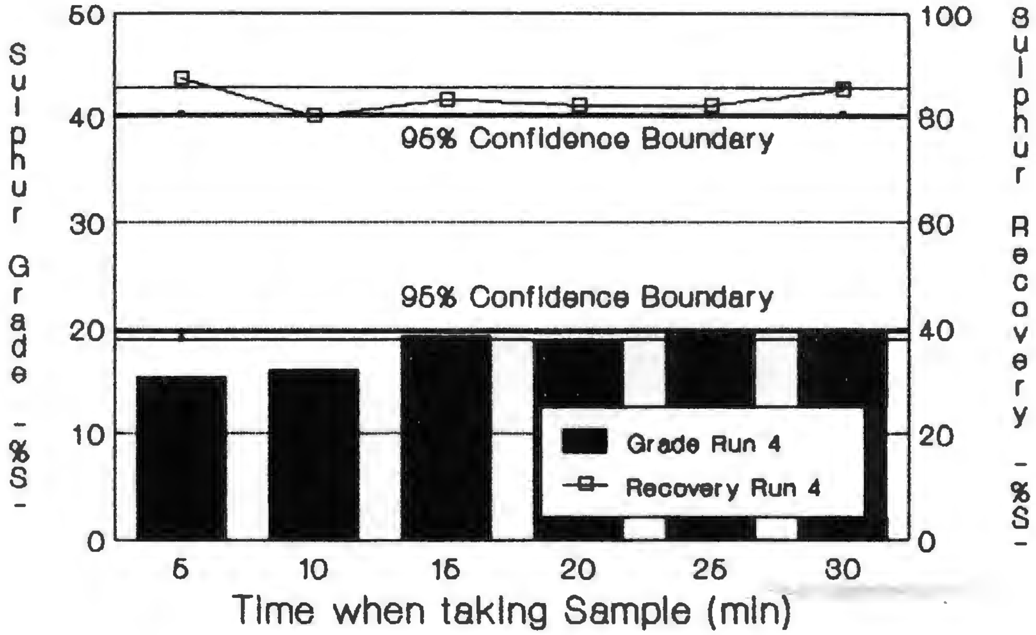

Figure 4.5: Sulphur grade and recovery versus time to determine when steady state is reached in the column (Run 4).

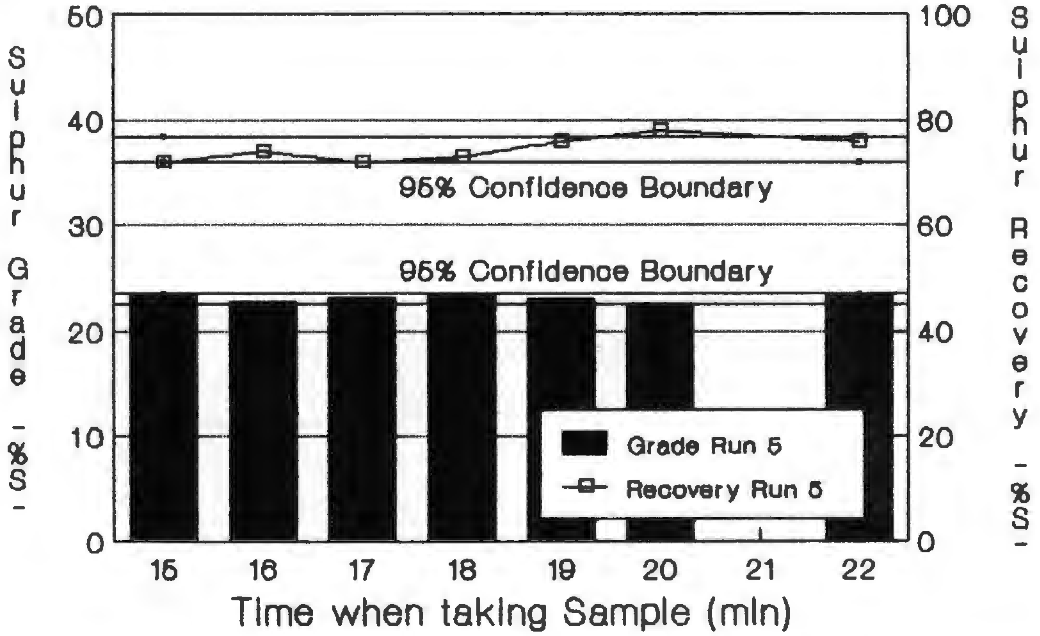

Figure 4.6: Sulphur grade and recovery steady state confirmation for times between 15 and 22 minutes (Run 5).

4.2.3 Repeatability

It is important to determine the repeatability of the experimental procedure. Firstly repeatability ensures that the grade and recovery trends observed for variations in the values of physical and chemical parameters are meaningful. Secondly the grades and recoveries obtained should be consistently achievable.

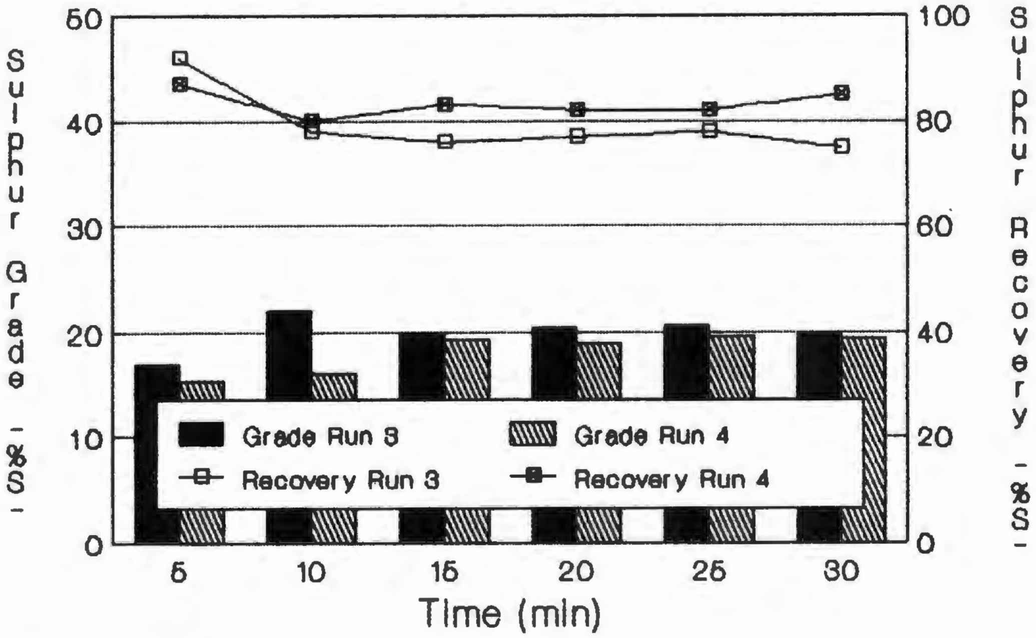

Figure 4.7: Sulphur grade and recovery versus time to determine the repeatability of the column (Run 3 & 4).

Figure 4.7 shows the repeatability in the grade and recovery values over the thirty minutes from start up. In both runs 3 and 4 the percent sulphur recovery dropped between five to ten minutes after start up and then remained essentially constant. The sulphur grades increased slightly over the first fifteen minutes. For runs 6 and 7 (Figure 4.8) good repeatability was again achieved for the grade and recovery trends. The wash water temperature was varied in runs 6 and 7.

The second part of the repeatability study was to achieve the same grades and recoveries consistently. Two runs were each done for eighty minutes. Figure 4.9 shows the consistency achieved. The sulphur grades differed on average by less than 1% and the sulphur recoveries differed on average by less than 2.5%.

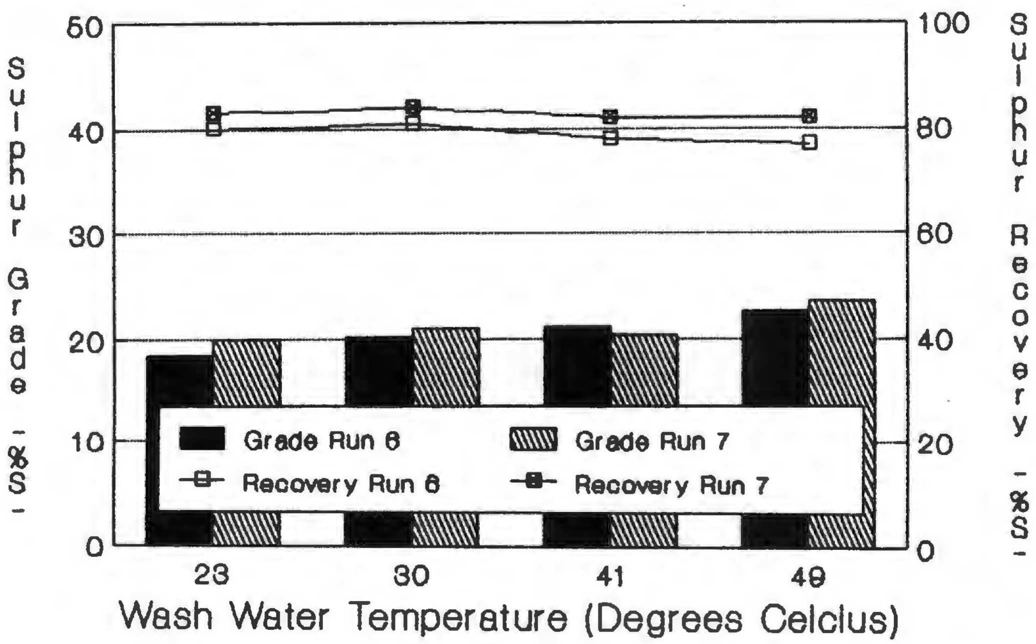

Figure 4.8: Sulphur grade and recovery versus wash water temperature to confirm the column repeatability (Run 6 & 7).

It was therefore concluded that the experimental procedure was adequately repeatability.

4.2.4 Comparison between conventional and column flotation.

The reason for doing a comparative study of a conventional laboratory batch flotation cell and a laboratory column flotation cell is that it is necessary to determine which type of flotation performs better. The performance is judged in terms of the concentrate grades and recoveries obtained.

A difference in the performance is expected because of the different types of pulp/air contact, the froth

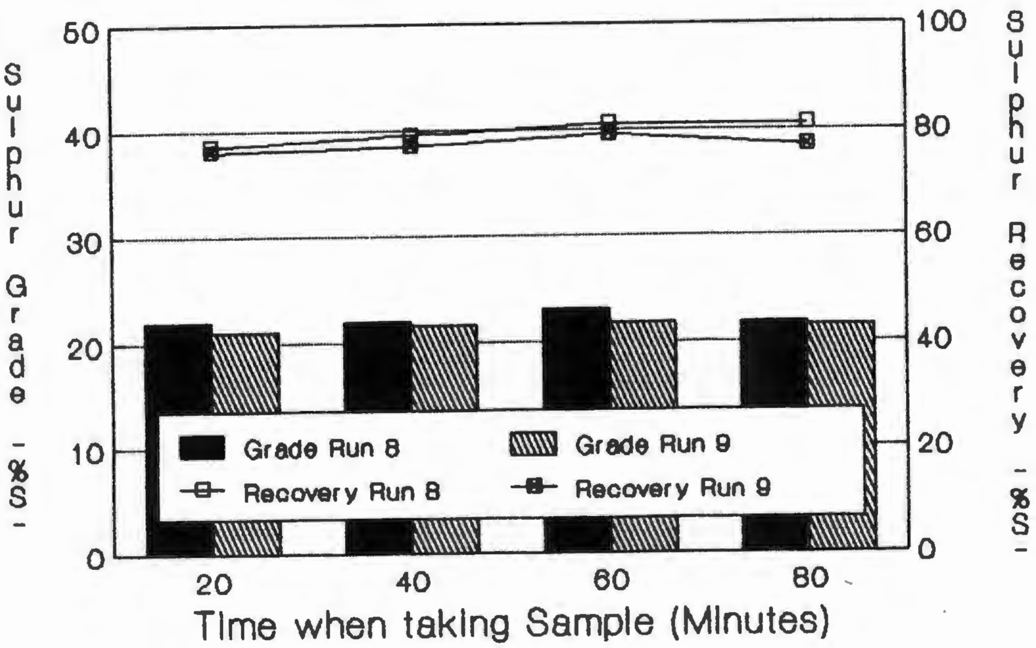

Figure 4.9: Sulphur grade and recovery versus time to confirm the column reproducibility (Run 8 & 9).

height and the presence or absence of wash water. An improved concentrate grade in the column flotation cell is expected due to the presence of wash water and the deep froth height.

To test the performance of each flotation system the column was used as described in the experimental procedure (Section 3.2). A batch of the feed sample to the column was then taken and added directly to the conventional laboratory batch flotation cell to ensure that identical feed conditions were used. This procedure allowed the conditioning time and reagent additions to be identical for both systems.

From the experiments conducted it was found that the column produced concentrates with higher sulphur grades, for similar recoveries, than possible with the batch cell (Figure 4.10). From Figure 4.10 it can also be seen that the sulphur recoveries were generally higher for the column. The higher sulphur recoveries obtained with the column were achieved at lower solids residence times than in the batch cell (Figure 4.11).

The column produced a gold concentrate with a grade of 86 g/ton compared to 140 g/ton for the batch cell (Column run 17 and batch test 1). The gold recoveries for both methods were however the same. The particle residence time in the batch cell was 14 minutes compared to the 2.9 minutes in the column.

From a size analysis it was found that the percent fines reporting to the concentrate (Figure 4.12) is lower for the column flotation cell than for the conventional batch cell.

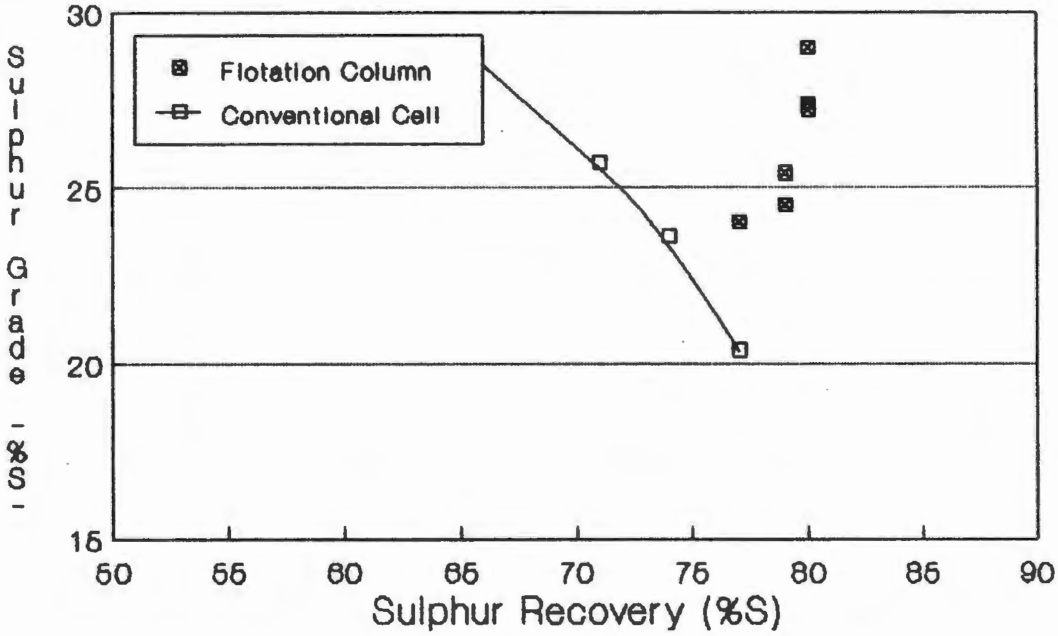

Figure 4.10: Sulphur grade versus sulphur recovery for column and conventional batch flotation comparison (Run 10).

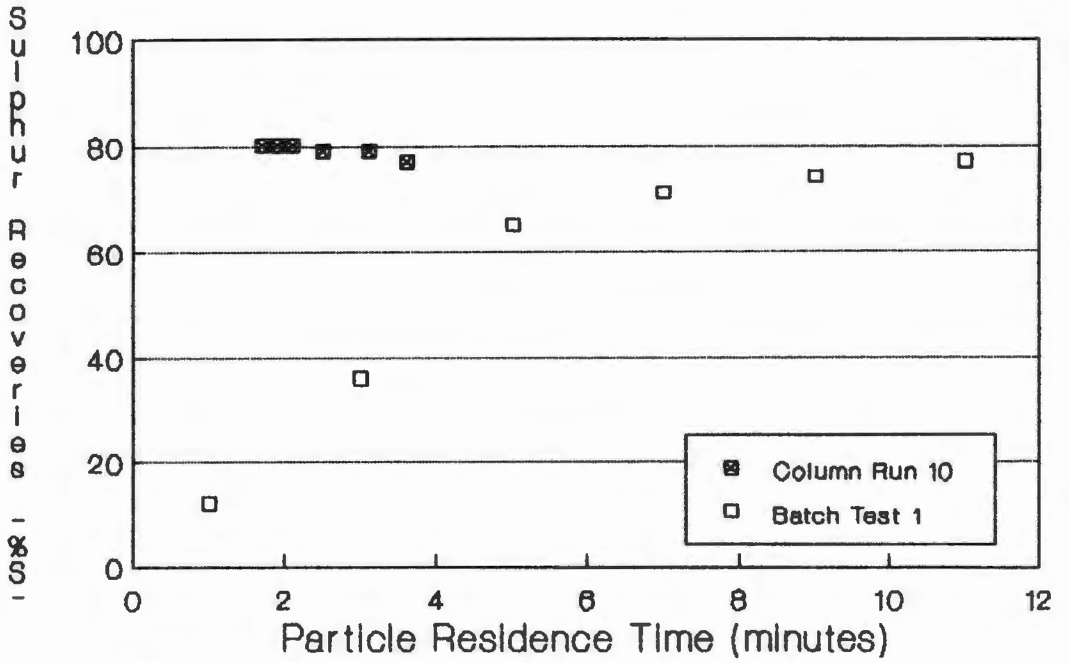

Figure 4.11: Sulphur recovery versus particle residence time for column and batch comparison (Run 10).

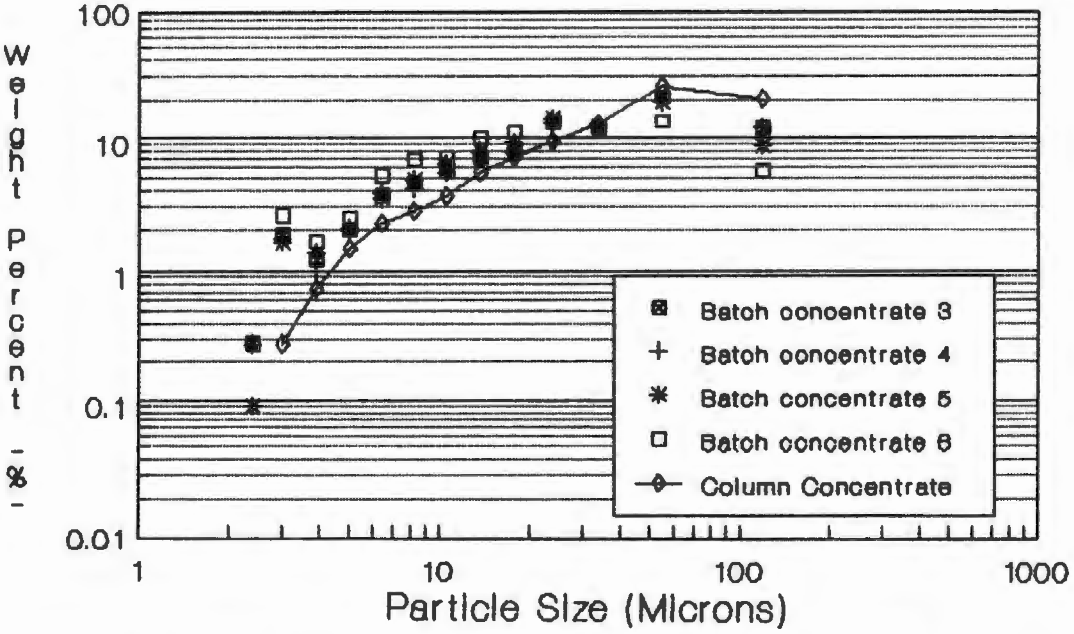

Figure 4.12: Particle size analysis for a column and conventional batch flotation comparison.

4.3 THE EFFECTS OF PHYSICAL PARAMETERS

To be able to model and design a column flotation cell an understanding of the effects of physical and chemical parameters on the performance of the cell is required. This understanding, together with the contradictions found in literature regarding the operation of a flotation column cell motivated the study of the effects of physical parameters. Each parameter was investigated while keeping all other parameters fixed. In order to improve the sulphur recovery of this ore the standard conditions were adjusted by changing them to the parameters which yielded the best recovery in the previous test done. The results of the study are reported in this section and in section 4.4.

4.3.1 Air Flow Rate

As the air flow rate to the flotation column was increased the sulphur recovery increased (Fig. 4.13; Standard conditions not the same. Only the trends are important). It was possible to increase the superficial

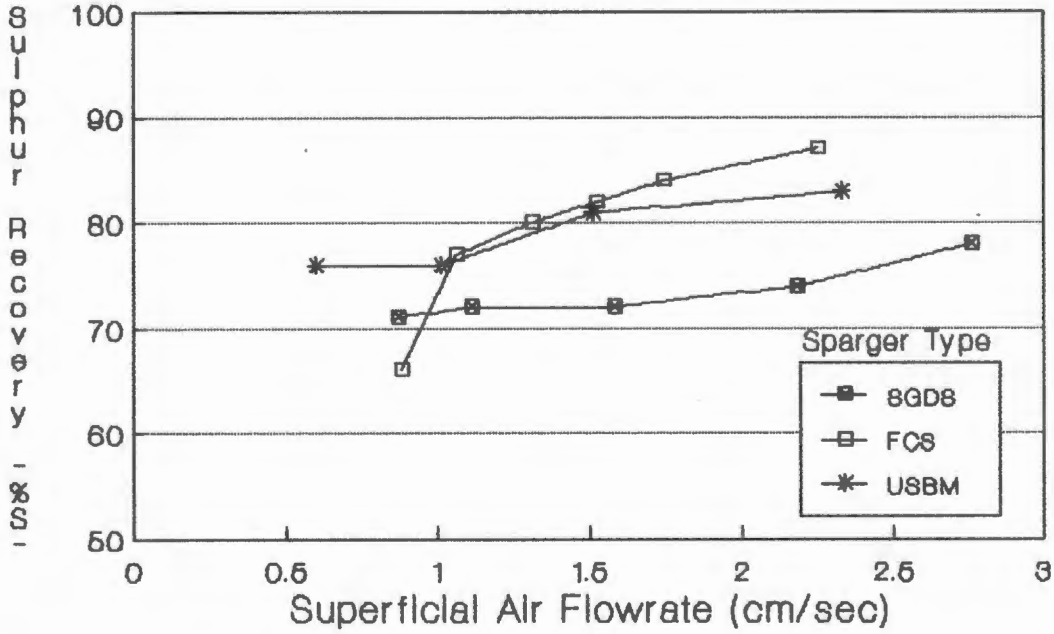

Figure 4.13: The effect of increasing the air flow rate on the sulphur recovery (Run 11, 12 & 13).

air rate up to 2,5 ± 0,3 cm/sec before slugging occurred. When **slugging** occurred the deep froth collapsed. Optimum air rates were therefore obtained which compare well with those reported in literature.

As expected, the sulphur grades decreased as the air rates were increased (Figure 4.14; Standard conditions not the same. Only the trends are important). An exception occurred when the USBM type sparger was used.

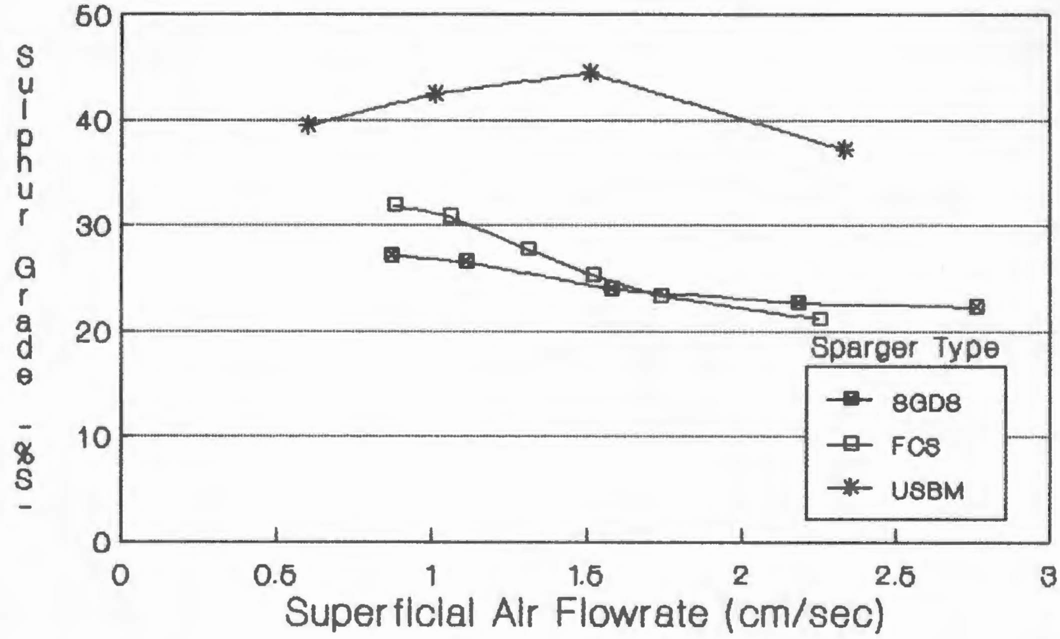

Figure 4.14: The effect of increasing the air flowrate on the concentrate sulphur grade (Run 11, 12 & 13).

When the air rate was increased the solids density in the concentrate decreased (Figure 4.15; Standard conditions not the same. Only the trends are important).

The mass of fine particles reporting to the concentrate generally increased as the air rate was increased (Figure 4.16, 4.17 and 4.18).

4.3.1.1 Sparger Type

When the sintered glass disc sparger, SGDS, is used bubbles with a mean diameter of 1,2mm can be produced.

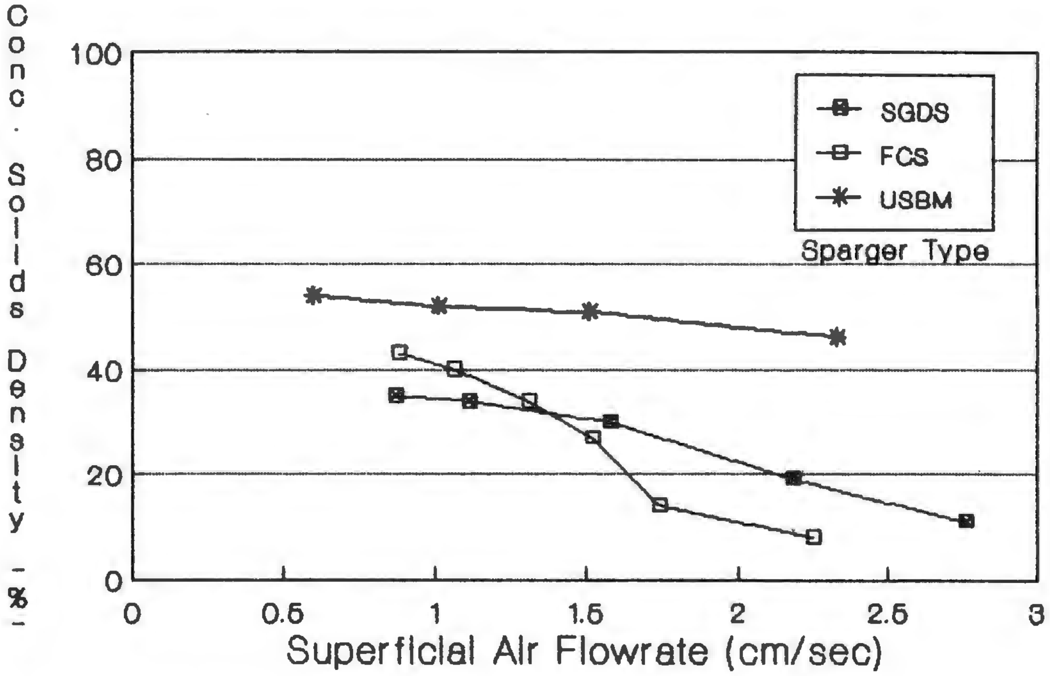

Figure 4.15: The effect of increasing the air flowrate on the concentrate solids density (Run 11, 12 & 13).

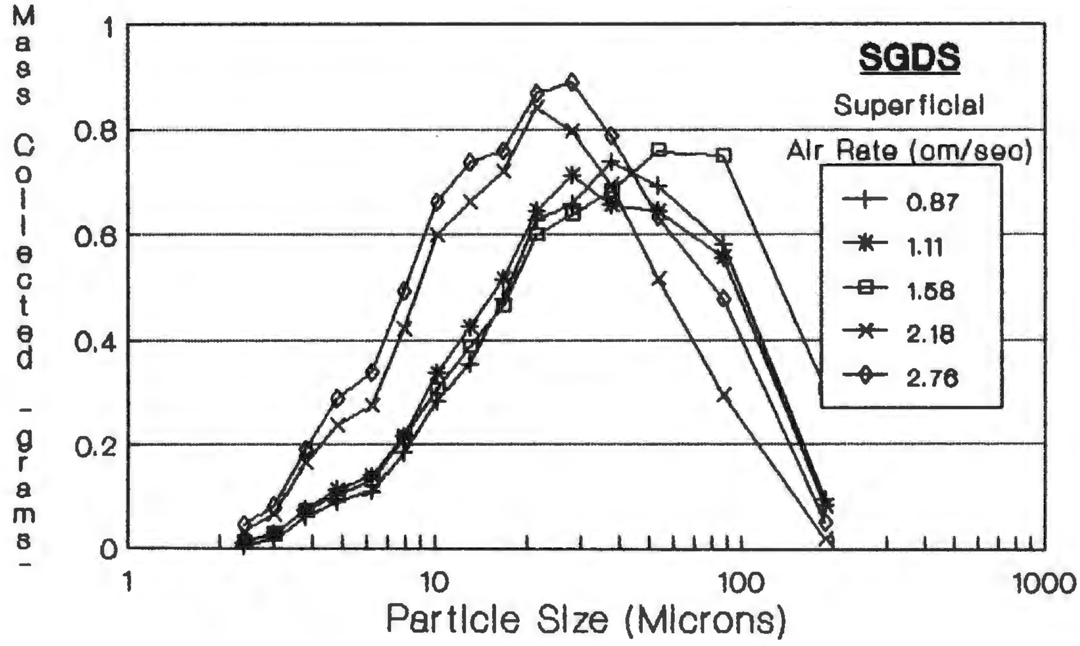

Figure 4.16: Particle size distribution for the concentrate when using the SGDS at varying superficial air rates.

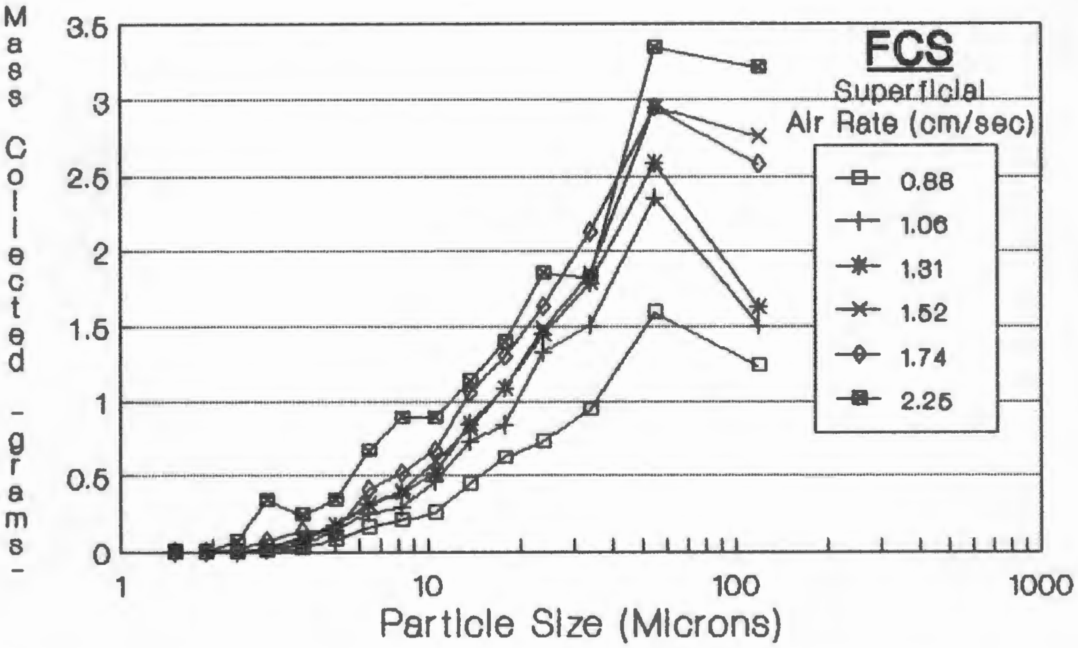

Figure 4.17: Particle size distribution for the concentrate when using the FCS at varying superficial air rates.

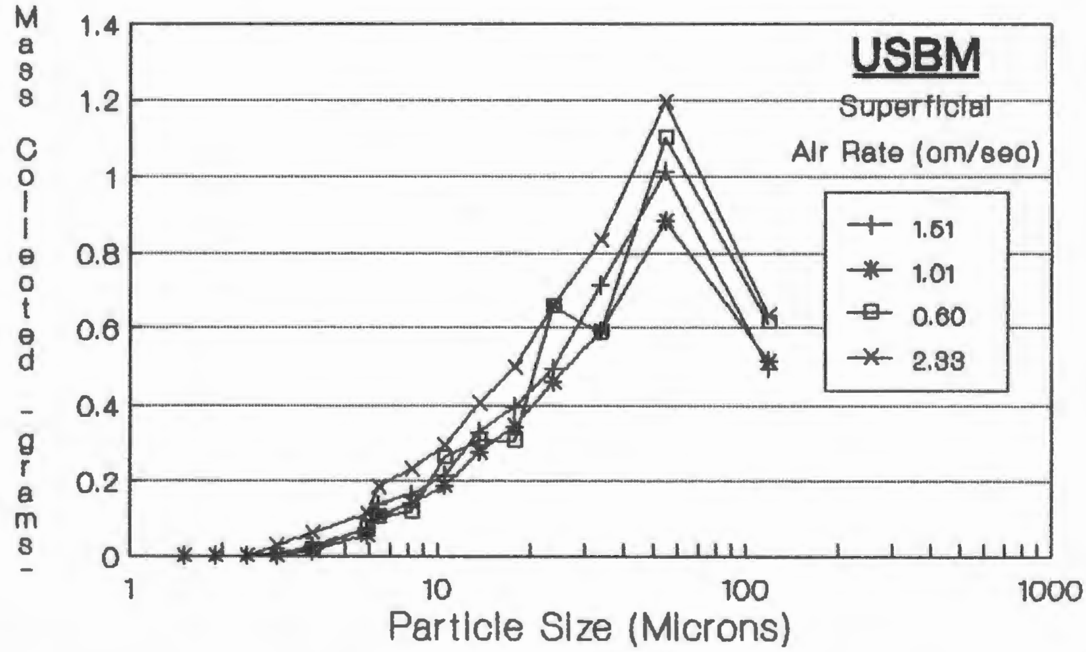

Figure 4.18: Particle size distribution for the concentrate when using the USBM sparger at varying superficial air rates.

The pressures required for the operation of the SGDS ranged between 4 and 6 bar. The bubble sizes obtained in a 2 phase study for the SGDS are shown in Figure 4.19.

When no frother was used the mean bubble diameter was between 1,8 mm and 2,1 mm. By adding 2 ppm by volume frother (Dow 200) the mean bubble size was reduced to about 1,2 mm. Adding more frother did not decrease the mean bubble size any further.

When no frother was used the mean bubble size decreased with an increasing air flow rate in contradiction with literature. This is because the air initially escapes through the larger pores and only at higher pressures (Higher air rates) through the smaller pores. Increasing the air flow rate when frother was used, caused the mean bubble size to increase.

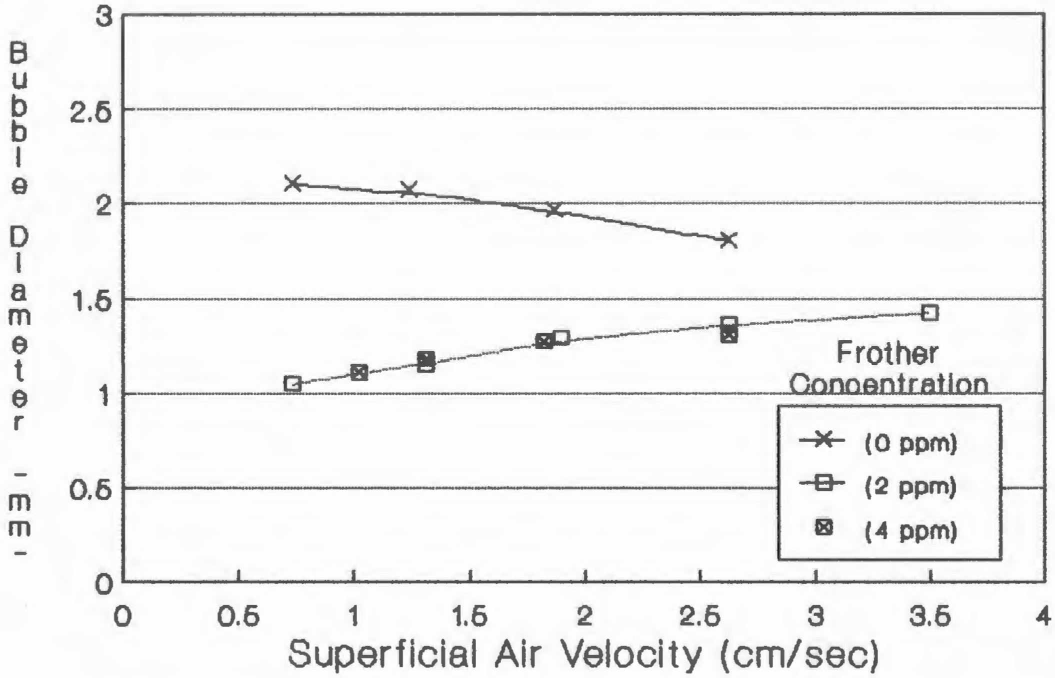

Figure 4.19: Bubble diameter versus superficial air velocity for the sintered glass disc sparger at different frother concentrations.

The filter cloth sparger, FCS, produced bubbles with a mean bubble diameter that decreased as the superficial air velocity was increased (Figure 4.20). This is in contradiction with literature. This is because the air initially escapes through the larger pores and only at higher pressures (Higher air rates) through the smaller pores. This is possible due to the large sparger surface area available (Ratio of column cross sectional area to sparger surface area equals 1). Adding 4 ppm Dow frother resulted in the mean bubble diameter only decreasing by about 0,1 mm.

The United States, Bureau of Mines type sparger, USBM, produced bubbles with a mean bubble diameter of about 2,3 mm (Figure 4.21) without the addition of frother in the sparger water. When a frother, Dow 200, was added it was possible to control the bubble size and reduce the mean bubble diameter to about 1 mm.

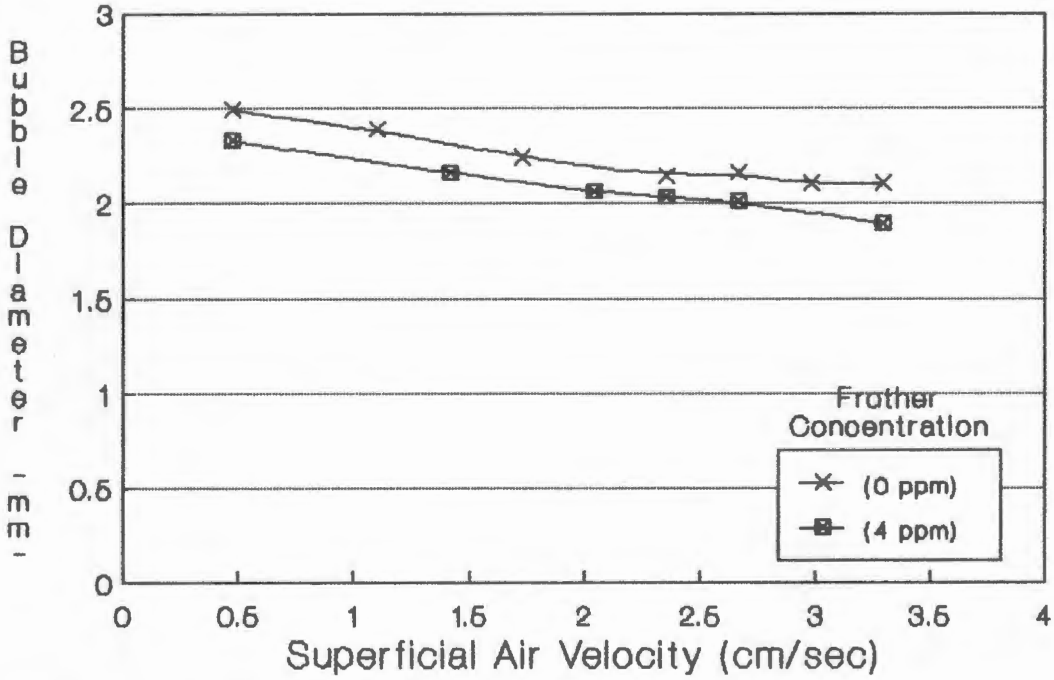

Figure 4.20: Bubble diameter versus superficial air velocity for the filter cloth sparger at different frother concentrations.

The standard deviation of the bubble diameter distribution was smallest for the USBM. The standard deviation was about 10 times smaller for the USBM than for the FCS (Appendix B; Run 10 and 11).

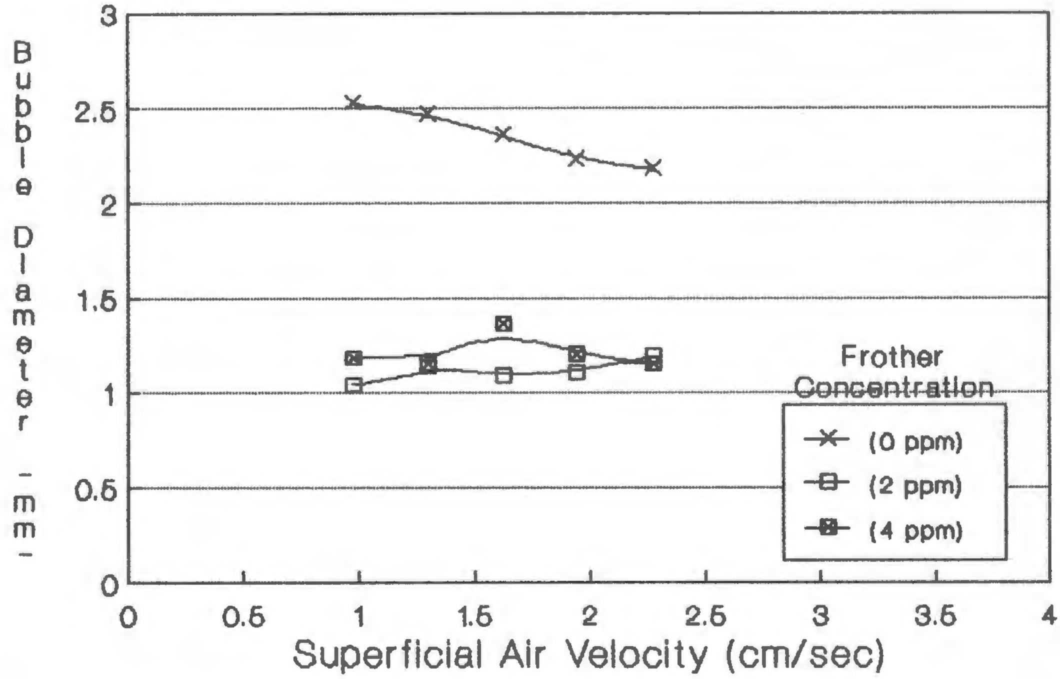

Figure 4.21: Bubble diameter versus superficial air velocity for the United States Bureau of Mines sparger at different frother concentrations.

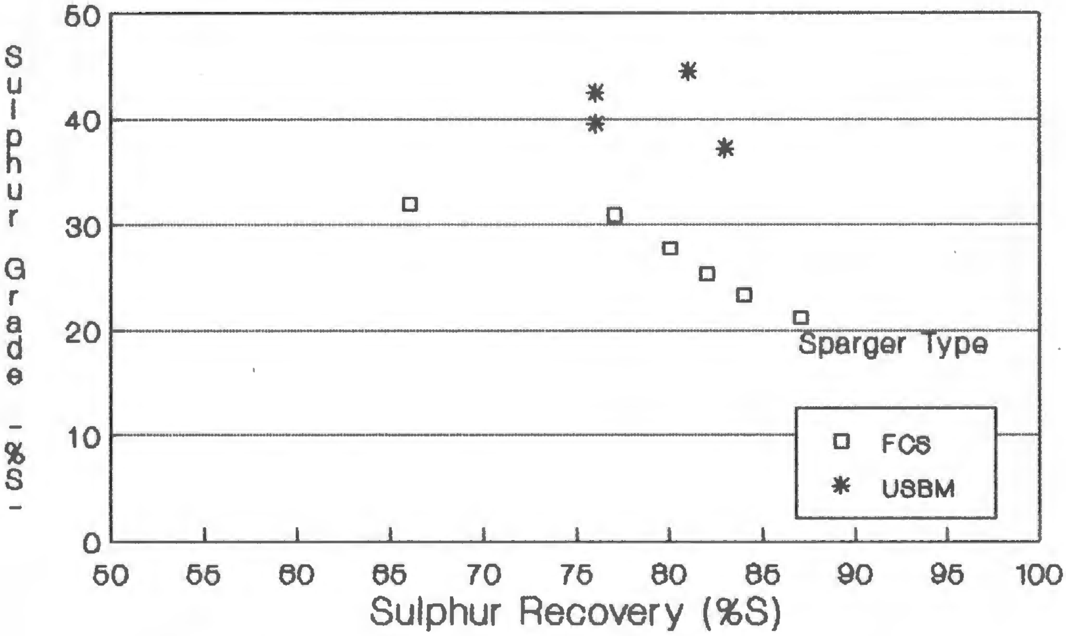

Figure 4.22: Comparison of the sulphur grade versus the sulphur recovery for the SGDS, FCS and the USBM spargers.

Comparing the sulphur grades and recoveries obtained with the filter cloth sparger, FCS, and the United States Bureau of Mines sparger, USBM, it was found that the USBM produced the best concentrate grades at a specific recovery (Figure 4.22). The frother concentration was not increased further because this increase does not significantly decrease the bubble size.

4.3.2 Collection Zone Length

By increasing the collection zone length from 1m to 4m the sulphur recovery increased from 84% to 89% (Figure 4.23). The sulphur recovery remained at approximately 89% when the collection zone was increased to 6m. The solids mass in the concentrate also increased as the collection zone length was increased. The maximum was reached at a length of 4 meters.

In general the concentrate sulphur grade remained constant as the collection zone length increased (Figure 4.23). There was also no change in the particle size distribution as the collection zone length was increased.

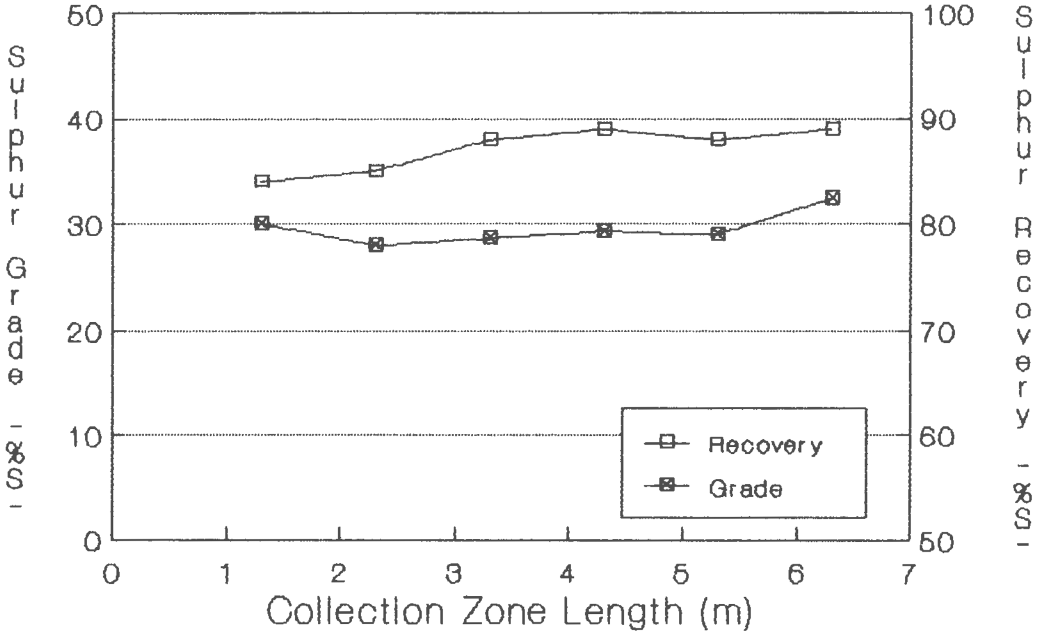

Figure 4.23: Sulphur recovery and grade versus collection zone length (Run 14).

4.3.3 Cleaning Zone Depth

The most obvious effect of increasing the cleaning zone depth was that the sulphur recovery decreased (Figure 4.24). When the froth depth was increased from 25 cm to 100 cm the recovery decreased from 83% to 73%.

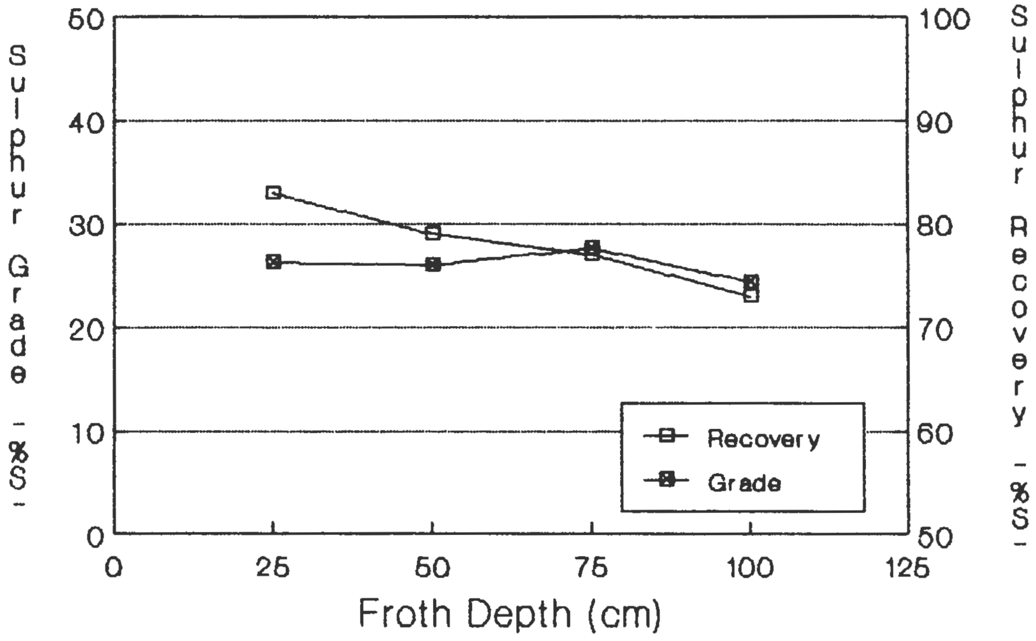

Figure 4.24: Sulphur recovery and grade versus froth depth (Run 15).

Unlike the improvement in concentrate grades that were reported in literature, it was found that the concentrate sulphur grades remained constant and then increased slightly at a depth of 75 cm (Figure 4.24). At a froth depth of 100 cm the sulphur grade decreased again.

From Figure 4.25 it can be seen that the amount of larger particles reporting to the launder increase with a decreased froth depth.

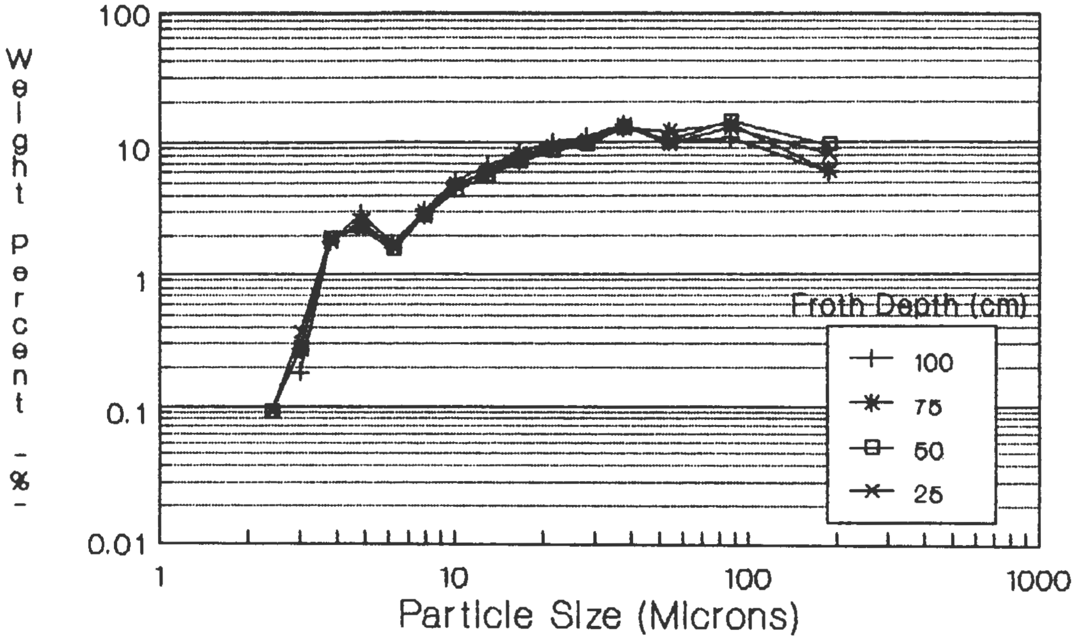

Figure 4.25: Particle size analysis for varying cleaning zone depth.

4.3.4 Feed Solids Percent

As the feed solids percent was increased the sulphur recovery decreased minimally up to a solids density of 25,3%. When the feed solids density was increased to 38,4% the sulphur recovery dropped dramatically to 59% (Figure 4.26).

The sulphur grade increased and then reached a maximum as the solids feed rate was increased (Figure 4.26).

By increasing the feed solids density the mass of the solids collected increased in all size ranges. The percent larger particles collected was decreased slightly. At a feed solids density of 38,4% however, the mass of larger particles collected decreased drastically (Figure 4.27).

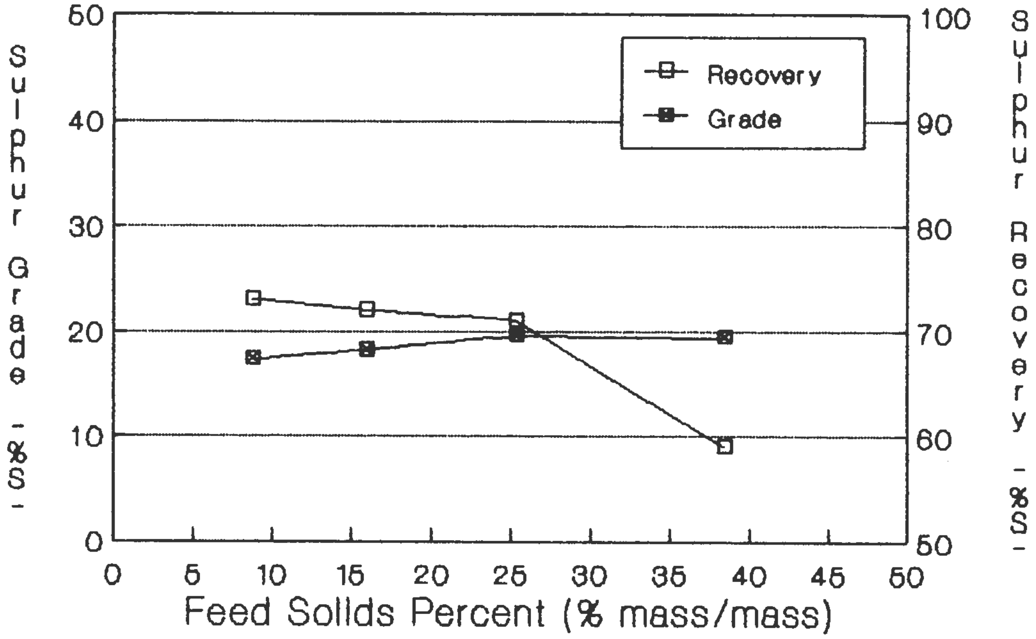

Figure 4.26: Sulphur recovery and grade versus feed solids percent (Run 16).

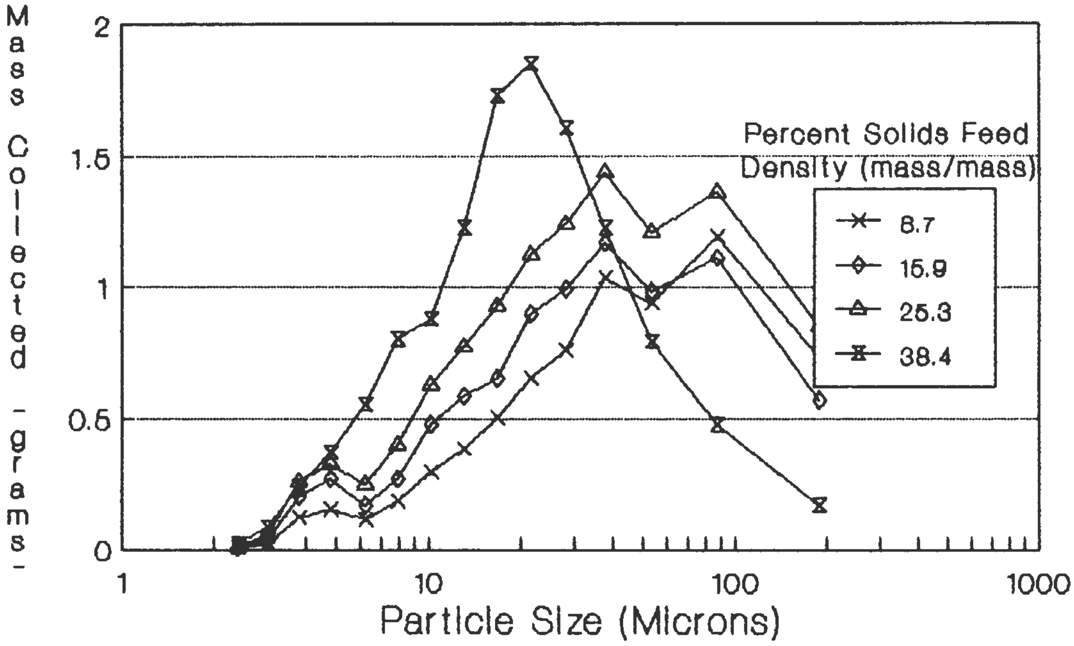

Figure 4.27: Particle size analysis for a varying feed solids density.

4.3.5 Volumetric Feed Rate

Increasing the volumetric feed rate from about 800 ml/min to about 2000 ml/min while keeping the percent solids constant at 10% solids had virtually no effect on the sulphur recovery. The sulphur recovery increased from 77% to 80% (Figure 4.28). The largest increase (2%) was found when the volumetric feed rate was increased from 800 ml/min to 1000 ml/min, while the other increases were not significant.

The sulphur grade also remained essentially constant. A possible optimum in grade can be seen at a volumetric feed rate of 1600 ml/min (Figure 4.28).

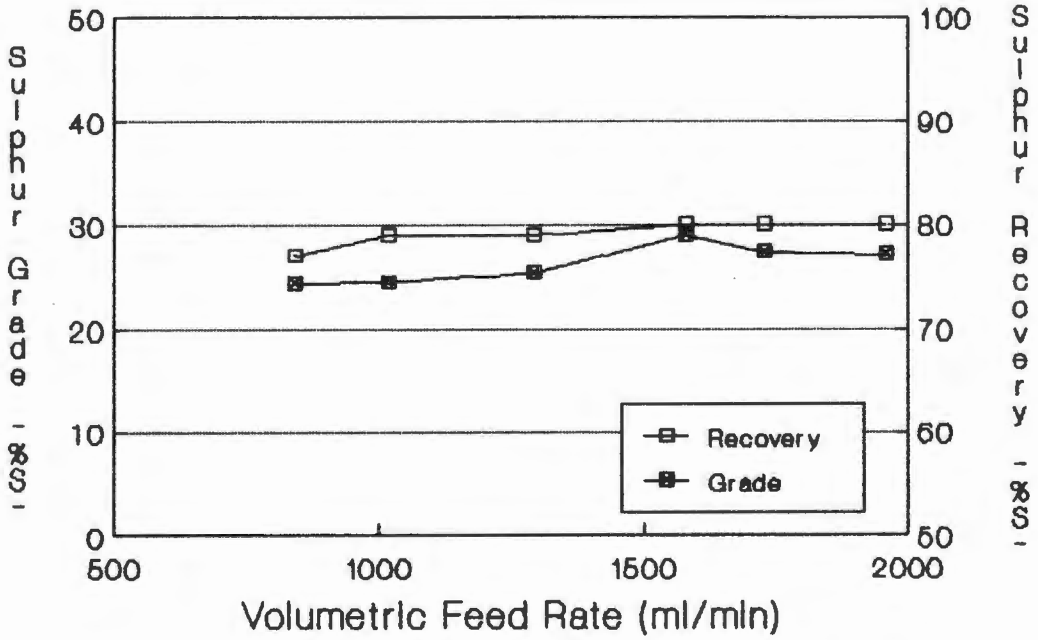

Figure 4.28: Sulphur recovery and grade versus volumetric feed rate (Run 10).

Increasing the volumetric feed rate increased the weight percent of larger particles collected in the concentrate. At a volumetric feed rate of 1958 ml/min the weight percent of larger particles collected decreased drastically (Figure 4.29).

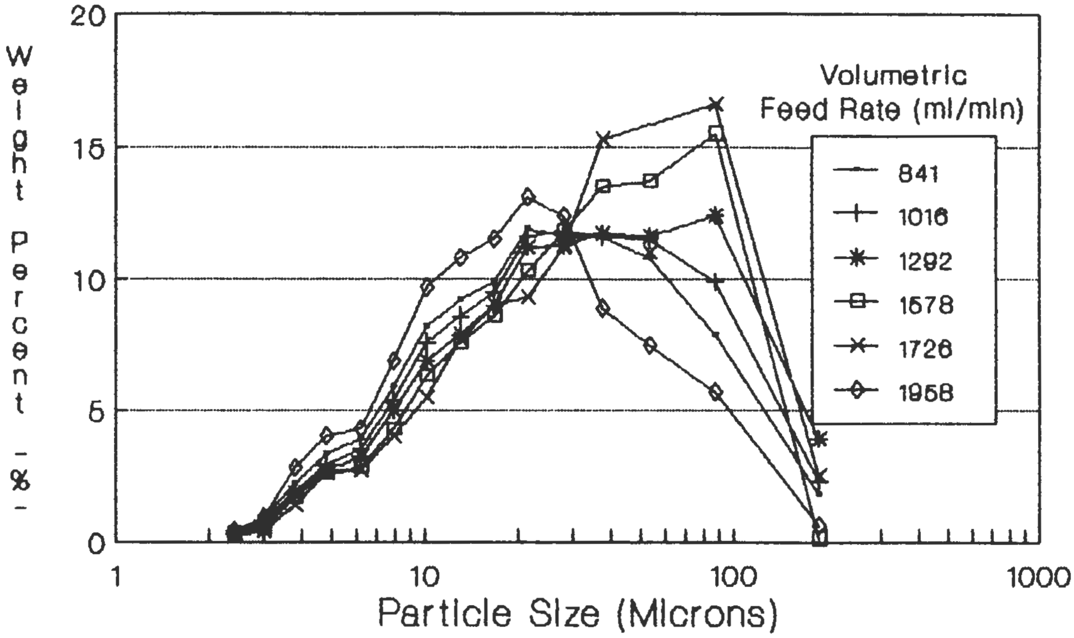

Figure 4.29: Particle size analysis for a varying volumetric feed rate.

4.3.6 Wash Water

4.3.6.1 Wash Water Rate

The sulphur recovery passed through an optimum (Figure 4.30) at a wash water rate of about 210 ml/min.

The concentrate sulphur grade decreased as the wash water rate increased from 118 ml/min to 212 ml/min (Figure 26). A further increase of the wash water rate to 355 ml/min caused the concentrate sulphur grade to increase again.

Increasing the wash water rate from 118 ml/min to 137 ml/min increased the mass of the coarse particles collected (Figure 4.31).

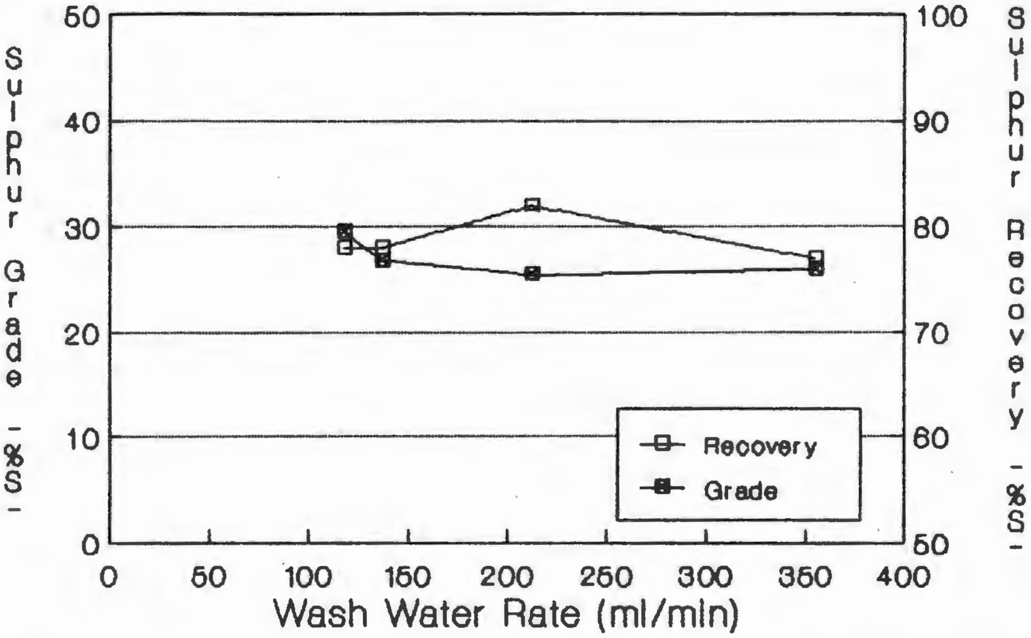

Figure 4.30: Sulphur recovery and grade versus wash water rate (Run 17).

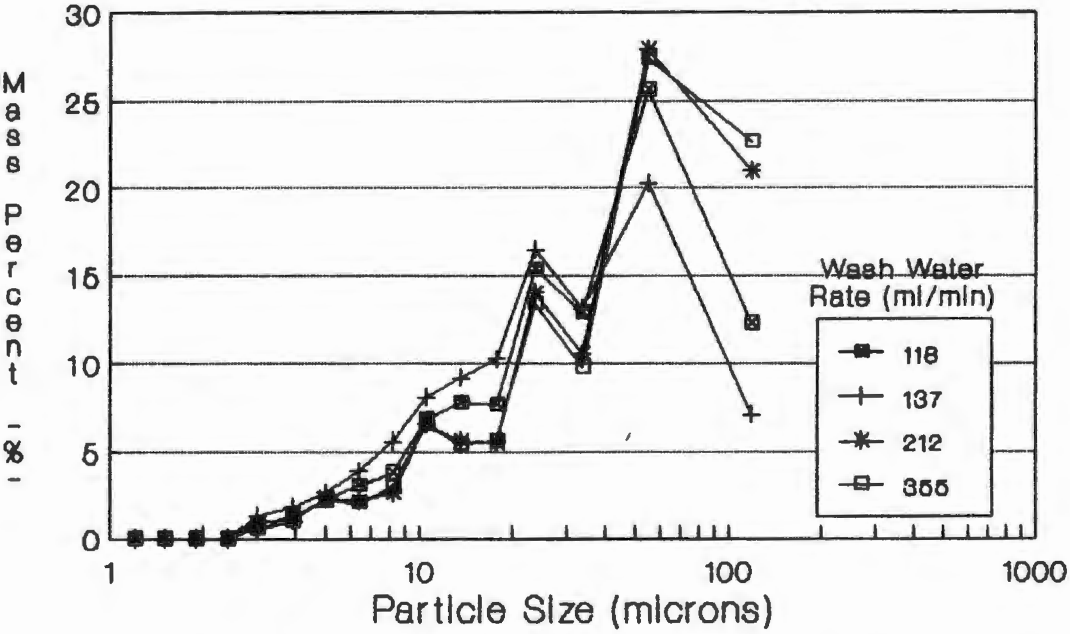

Figure 4.31: Particle size analysis for a varying wash water rate.

4.3.6.2 Wash Water Position

Lowering the wash water position from 5 cm to 20 cm below the concentrate overflow level did not affect the sulphur recovery or grade (Figure 4.32). The only effect that lowering the wash water position had on the column operation was that the percent solids in the concentrate was increased and then reached a maximum.

The particle size distribution did not change at all as the wash water position was changed (Figure 4.33).

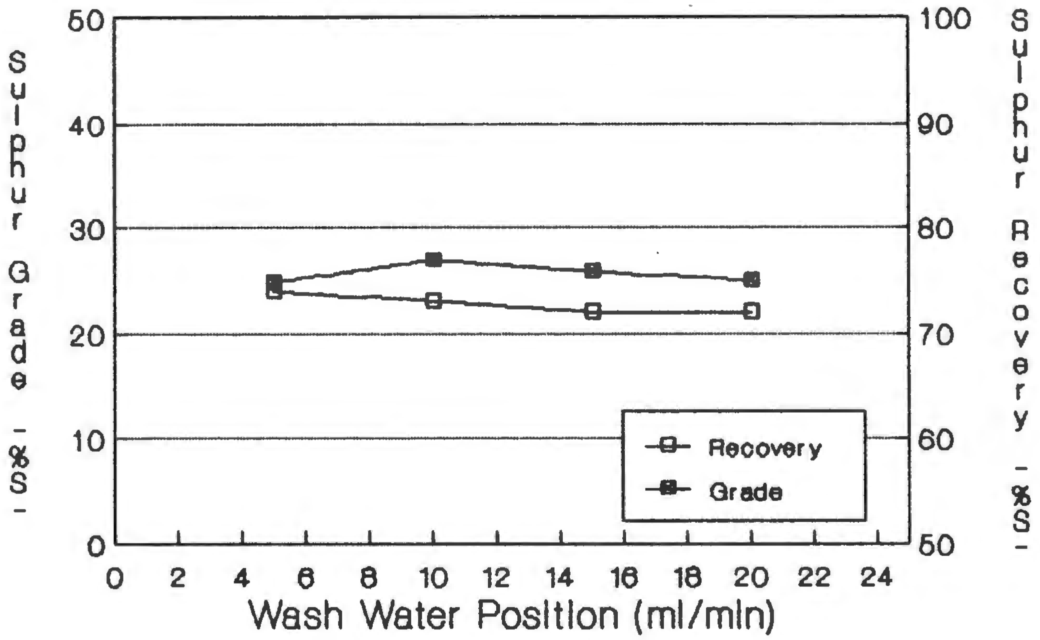

Figure 4.32: Sulphur recovery and grade versus wash water position (Run 18).

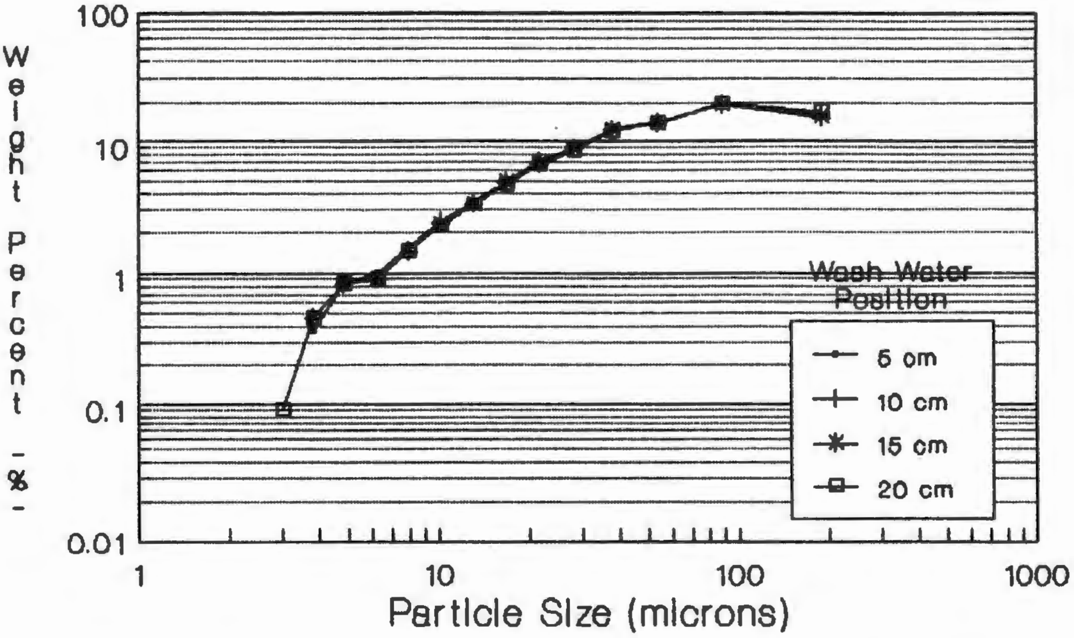

Figure 4.33: Particle size analysis for a varying wash water position.

4.3.6.3 Wash Water Temperature

The wash water temperature was increased from about 21°C to about 50°C. This decreased the sulphur recovery marginally while the grade increased by about 4% (Figure 4.8).

At higher wash water temperatures more particles larger than 100 microns were collected. This is clearly shown in Figure 4.34.

4.3.7 Feed Position

In the laboratory tests the feed position was lowered from 55 cm to 85 cm below the concentrate overflow while keeping the froth depth and the collection zone depth constant. The sulphur recovery decreased slightly while the concentrate sulphur grade improved (Figure 4.35).

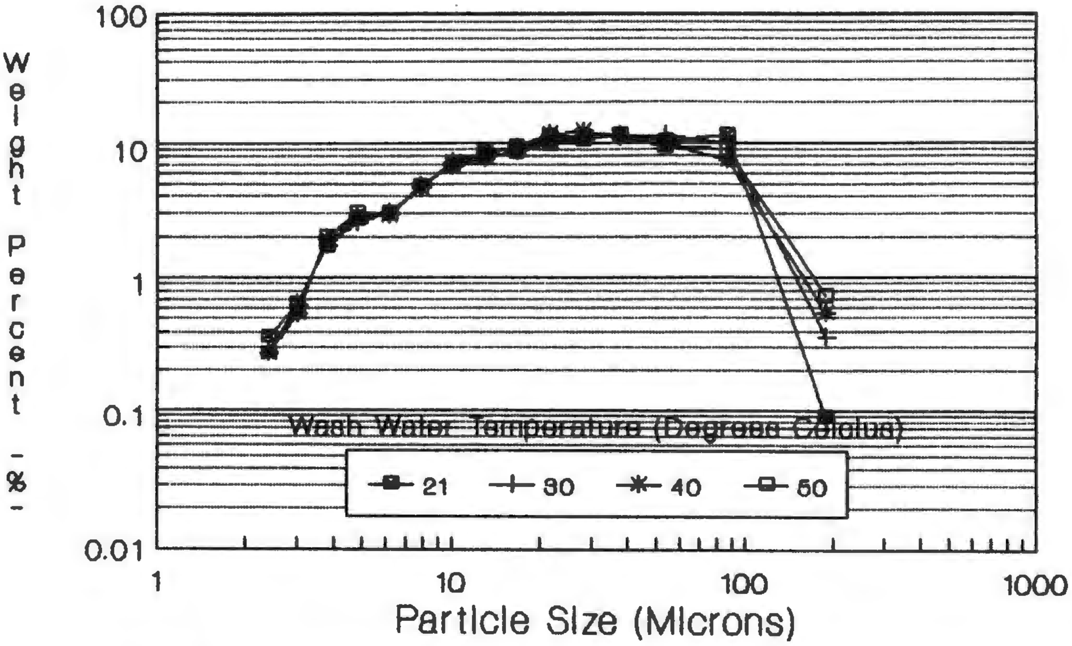

Figure 4.34: Particle size analysis for a varying wash water temperature.

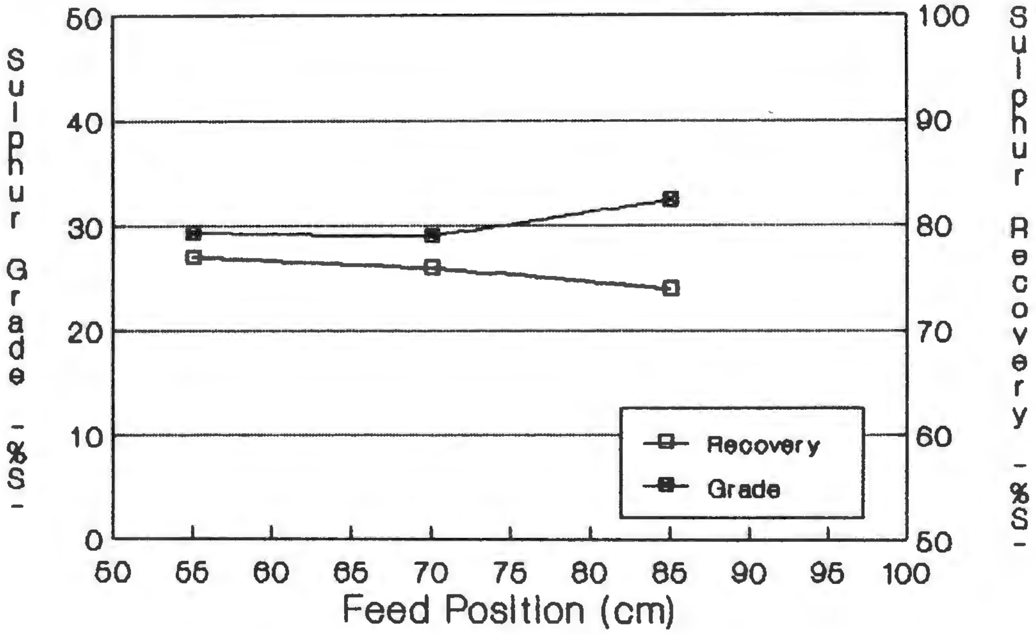

Figure 4.35: Sulphur recovery and grade versus feed position (Run 19).

4.4 THE EFFECT OF CHEMICAL PARAMETERS

4.4.1 Leaching the ore

The pulp was leached for 3 hours at a pH of 3.8 and then it was floated at the same pH to determine what the effect was on the sulphur recovery.

As a result of the lowered pH the sulphur recovery increased from about 85% to 88% (Run 20). The concentrate sulphur grade however decreased from about 22% to about 20% sulphur when the pH was lowered.

4.4.2 Conditioning Procedure

By increasing the conditioning time from about 25 to 30 minutes it was possible to increase the sulphur recovery

	Coll. Height	Froth Height	Solids Feed Rate	Frother Addition	Wash Water Rate	Jg	Sup. Bias Rate
	cm	*cm*	*g/min*	*g/ton*	*ml/min*	*cm/sec*	*cm/sec*
Run 18	225	25	176	137	200	1.0	0.13
Run 19	225	25	177	136	200	1.0	0.13
Run 22	225	25	160	270	200	1.5	0.07
Run 14	230	25	171	137	200	1.9	0.12

	Sparger Type	SIPX Addition Rate	AFLT Addition Rate	Cond. Time		**Sulphur Recovery**	**Sulphur Grade**
		g/ton	*g/ton*	*min*		*%S*	*%S*
Run 18	FCS	23	38	25		73	25
Run 19	FCS	22	37	30		76	30
Run 22	FCS	26	44	30		85	22.5
Run 14	FCS	18	30	10+10		85	30

Table 4.6: Details for different conditioning times and procedures.

from 73% to 76%. At the same time the concentrate sulphur grade was increased from 25% to 30% (Table 4.6).

It was observed that by adding the reagents step wise it was possible to increase the sulphur grade by 5% while keeping the recovery constant at 85% (Table 4.6). This improvement was possible even though the total conditioning time was reduced and only 70% of the reagent dosage was used.

4.4.3 Collector Dosage

By increasing the collector dosage the sulphur recoveries and the sulphur concentrate grades improved significantly.

An increase in reagent dosage from about 10 g/ton to 120 g/ton improved the sulphur recovery from 52% to 84%, while the sulphur grade improved from about 19% to 26% (Figure 4.36).

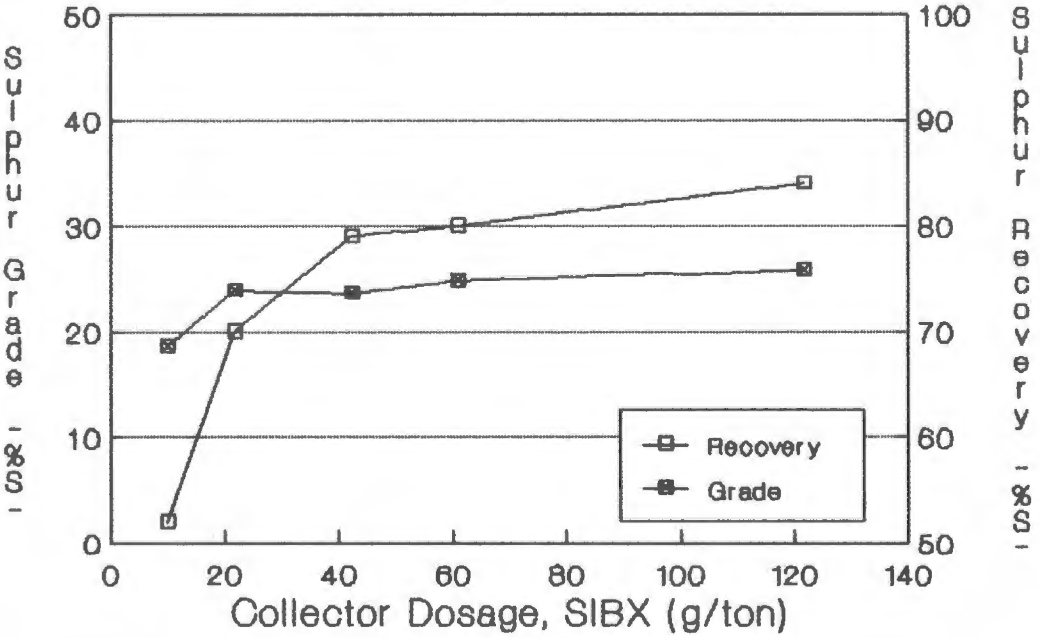

Figure 4.36: Sulphur grade and recovery versus collector dosage (Run 21).

As the collector dosage was increased the mass percent particles plus 100 microns collected increased (Figure 4.37).

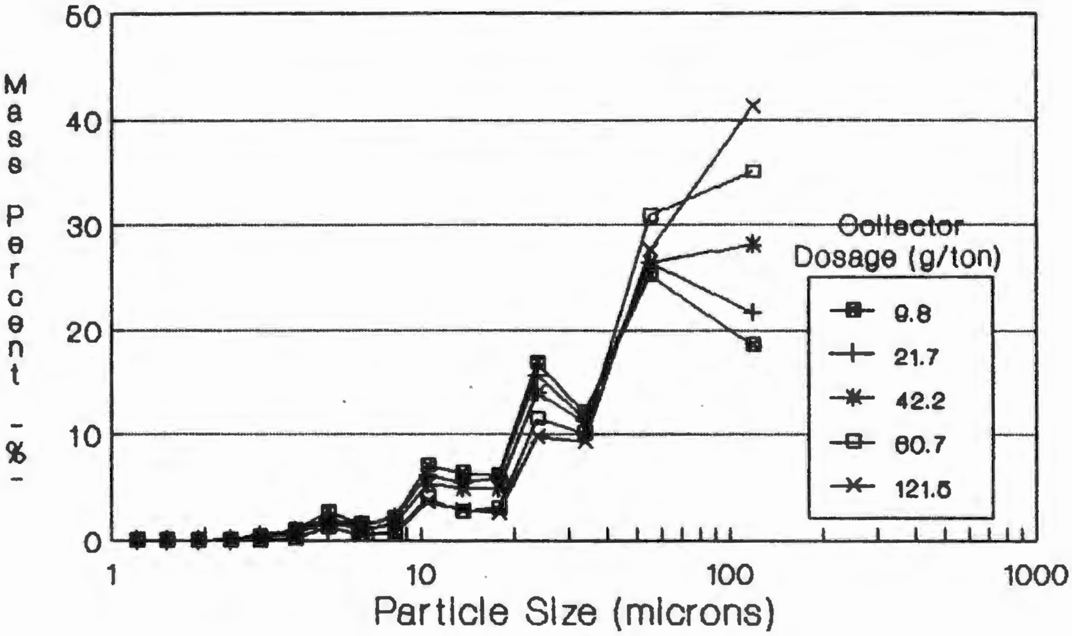

Figure 4.37: Particle size analysis for an increasing collector dosage.

4.4.4 Frother Dosage

The frother dosage was increased from 30 g/ton to 113 g/ton. At 30 g/ton frother addition the sulphur recovery was 81%. At frother addition rates above 57 g/ton the sulphur recovery remained constant at 85% (Figure 4.38). The concentrate sulphur grade did not vary with increasing frother addition.

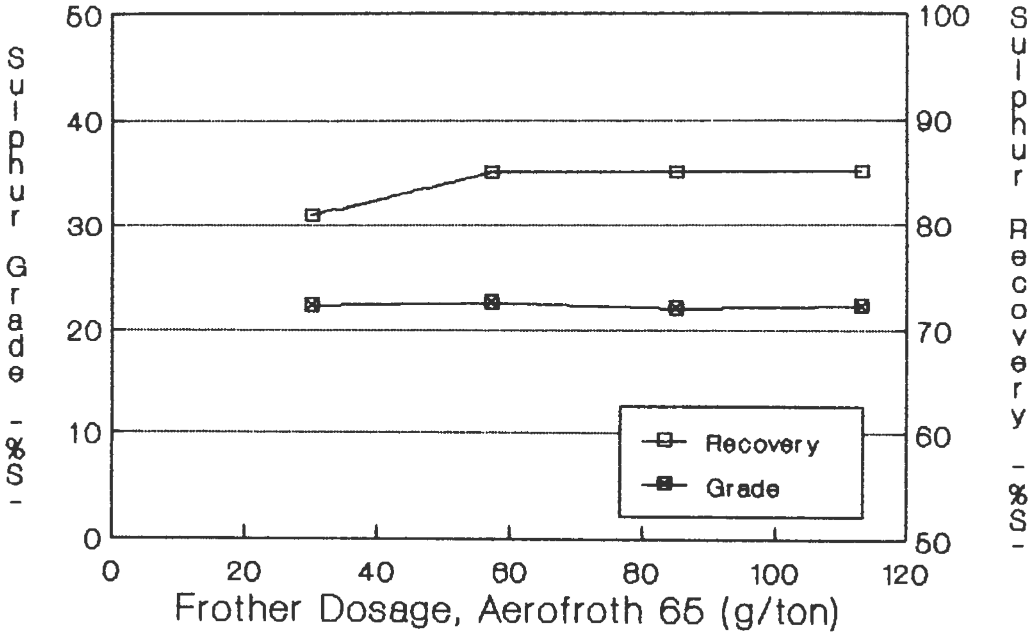

Figure 4.38: Sulphur recovery and grade versus frother addition rate (Run 22).

4.5 SUMMARY OF THE EFFECTS OF PARAMETERS VARIED

The effects of physical and chemical parameters are summarised in Table 4.7. The table also includes a number of other changes recorded (Appendix D) as a result of varying physical and chemical parameters.

The following symbols were used in table 4.7 :

– = No change
I = Small increase
II = Large increase
D = Small decrease
DD = Large decrease
OPT = Value passes through an optimum
MIN = Value passes through a minimum

PARAMETER INCREASED	Js	Bias Rate	Tp Coll.	Tp Froth	d50	%Sol Density	Water Rate	Mass Rate	%Sol. Rec.	Grade	S Rec
FCS Air Rate	II	DD	–	DD	I	DD	II	II	II	DD	II
SGDS Air Rate	II	DD	–	DD	I	DD	II	II	II	DD	II
USBM Air Rate	II	DD	–	DD	I	DD	I	I	I	OPT	I
Collect Zone	–	–	II	–	OPT	MIN	OPT	I	I	MIN	I
Froth Depth	–	II	D	II	–	II	DD	DD	D	–	DD
Feed % Solids	–	DD	I	–	* II	I	I	I	D	I	D
W/W Rate	–	II	D	–	–	D	II	I	–	D	–
W/W Temp	–	–	–	–	–	I	D	D	D	I	D
W/W Depth	–	–	–	–	–	I	D	–	–	–	–
Vol Feed Rate	–	DD	DD	D	I	II	II	II	–	–	–
Feed Depth	–	–	D	–	–	D	–	D	–	I	D
Collect Dosage	–	II	–	–	–	I	DD	I	I	I	D
Frother Dosage	–	DD	–	–	–	D	I	I	I	–	–

* Increase at maximum carrying capacity.

Table 4.7: Summary of the effects of physical and chemical parameters.

CHAPTER
FIVE

DISCUSSIONS

5.1 COLUMN VERSUS BATCH FLOTATION

The column flotation cell was capable of producing sulphur grades 5 % greater than those obtained in a conventional laboratory batch cell at similar recoveries. The column also achieved up to 15 % higher sulphur recoveries than the batch cell at similar grades. The residence times for the column were about 2 minutes compared to the 11 minutes that were required for conventional flotation to achieve similar recoveries.

The better concentrate sulphur grades produced by the column flotation cell were due to the better cleaning effect of the deep, water washed froth. The superior cleaning also reduced entrainment into the concentrate launder, since less (see figure 4.12) of the minus 10 micron particles, which are usually entrained instead of being collected by attachment, were collected. A more selective flotation of the finer size fractions (51.1 % sulphur was contained in the minus 38 µm size fraction) therefore resulted in better sulphur grades.

The improved recovery achieved with the column flotation cell is due to the particle-bubble counter current contact mechanism, the wash water and the control over the bubble sizes. The advantage of the counter current contact mechanism is that any hydrophobic particles that are not immediately attached to a bubble have the opportunity to attach themselves to a **clean bubble** further down the column without competing against other more hydrophobic particles. The wash water washes most fine (minus 10 µm) particles that tend to be entrained, out of the froth, while with a controlled bubble size the bubble surface rate can be optimised to yield an optimum recovery.

The column flotation cell is thus significantly better than the batch cell both in terms of kinetics as well as sulphur grades and recoveries obtained for the Unisel ore used. Another attractive feature is the possible reduction of capital outlay (when installing flotation columns versus conventional flotation cells) and a reduction of operating costs.

The improved performance with the flotation column cell confirmed reports in literature. The column is therefore an attractive alternative to conventional flotation and should always be considered as an option.

5.2 THE EFFECTS OF PHYSICAL PARAMETERS

5.2.1 AIR FLOW RATE

As the air flow rate was increased the sulphur recovery increased until an optimum was reached. This was true for the three types of spargers tested. The increased recovery due to an increased air flow rate was the result of a larger percentage of the solids being recovered in the concentrate. The experimental results (Table 4.7) prove that the main reason for an increased sulphur recovery, when increasing the air flow rate, was due to an increased mass pull. The sulphur grade decreased but the sulphur recovery increased further.

The question now arises as to how the increased air flow rate increased the percentage solids recovered. The answer is that the increased air rate increased the superficial bubble surface area which in turn increased the concentrate solids removal rate (increased mass pull).

When the filter cloth sparger, the sintered glass disc sparger (at 0 ppm frother) and the USBM sparger (at 0 ppm frother) were used (Figure 4.19, 4.20 & 4.21) the total superficial bubble surface area was increased which allowed more particles to be collected. The superficial bubble surface area increased with an increased air flow rate when the bubbles produced by the spargers became smaller in diameter (Figure 4.19, 4.20, 4.21).

An increased superficial bubble surface area will also be obtained when the bubbles do not increase in size, but in numbers. Similarly, when the USBM sparger and the sintered glass disc sparger (at 0 ppm frother) were used, the number of bubbles produced did not decrease (viz. no coalescence of bubbles) but the mean bubble diameter was increased instead (Figure 4.19 and 4.21). This resulted in an increased superficial bubble surface area.

The increased concentrate solids removal rate with an increased air rate or decreased bubble size is confirmed by the carrying capacity model published by Espinosa-Gomez, et al (1988). There is however a limitation to the maximum amount of solids that will float, and to the maximum superficial bubble surface rate that can be obtained.

Besides the increased concentrate solids removal rate experienced due to the increased air rate, the following two effects also increased the percentage solids recovered:

Firstly, the bubble residence time in the froth, and therefore the floated solids residence time in the froth, t^{froth} (Table 4.7) was reduced, allowing more of the less hydrophobic and also more finer particles (Figure 4.16 to 4.18) to pass into the launder.

Secondly, the collection rate was increased ($k_c = 1.5 \cdot J_g E_k d_b^{-1}$) by an increased air rate and a decreased bubble diameter.

When the air flow rates were increased further **slugging** occurred. The flotation process thus ceased to exist. The maximum in the recovery obtained was due to the following reasons. The total superficial bubble surface decreased rapidly when bubbles coalesced (causing slugging), decreasing the concentrate solids removal rate. In the collection rate equation, $k = 1{,}5 \cdot v_g \cdot E_k / d_b$, it is shown that the collection rate decreases when the bubble size increases. When slugging occurs this then becomes the overriding parameter.

In the experimental work the optimum air rates were about 2,5 cm/sec for the best sulphur recovery. These air rates compared well to optimum air rates reported in literature (Between 1.7 and 2.5 cm/sec).

The sulphur grades were generally observed to decrease as the air flow rates were increased. This was due to a higher percentage solids being collected. This decrease in grade was also reflected in the increased dilution of the concentrate and therefore more fines (minus 50 microns) were collected. Although 60.5 % of the sulphur was in the minus 53 µm size fraction, higher air rates entrain the gangue particles.

Increasing the air flow rate to its maximum, without the bubbles coalescing or the number of bubbles produced reducing, will be the best option as far as the recovery is concerned. In the literature this optimum air rate was recorded to be between 1.7 and 2.5 cm/sec. The concentrate sulphur grade may however be limiting if the wash water rate or the froth depth cannot be increased enough to achieve an adequate grade.

From the results for increased air rates it was observed that the concentrate grade decreased proportionally to the concentrate solids density. The percentage solids density can therefore be used to estimate the concentrate grade for this ore, and maybe for some other ores.

5.2.2 SPARGER TYPE

The USBM type sparger produced the best concentrate grades at a specific recovery. The better grades are due to a better cleaning action. (The slightly different reagents used for these test do not significantly affect the the results. This is shown in table 4.7. The physical parameters influence the sulphur recovery and grade much more than the chemical parameters).

The superior cleaning action of the froth, when using the USBM type sparger, is due to smaller bubbles beinggenerated with the USBM type sparger. The smaller bubbles in the froth cause the froth to be more stable, allowing the froth to be better washed.

Turbulence in the froth can be worsened if larger bubbles are only produced in a few sections of the column instead of being evenly distributed over the whole cross sectional area of the column. It is therefore desirable to produce bubbles with as narrow a size distribution as possible. Any turbulence in the froth due to the rising bubbles (or wash water) should be minimized.

5.2.3 COLLECTION ZONE LENGTH

By increasing the collection zone length the recovery increased up to 89% and then remained at about this value. The increased recovery was due to an increase in the solids residence time available for particle-bubble attachment.

The recovery increased up to 89% and then remained fairly constant due to an equilibrium being reached between the bubble-particle attachment rate and the bubble-particle separation in the froth due to the cleaning action. If the recovery zone length is increased after the maximum in recovery is reached any collected particles that are less hydrophobic (slower rate constant) and are not attached firmly to the bubble will thus be washed out of the froth. This was confirmed by the increase in the sulphur grade at a collection zone of 6 meters, and by the fact that the mass of solids collected remained constant after the maximum recovery was obtained.

Therefore, unlike conventional flotation where an increased residence time increases the recovery while the grade is decreased, the flotation column can maintain a superior grade (above conventional flotation) with an increased residence time due to the deep froth and the washing water.

5.2.4 CLEANING ZONE DEPTH

It is hypothesised that as the cleaning zone depth is increased the time for cleaning is also increased. The probability that particles are entrained or hydrophobic particles are collected in the concentrate are thus reduced. The grade is therefore expected to improve. The results from the laboratory test work (Figure 4.24) showed that the concentrate grade increased up to a froth depth of 75 cm. When the froth depth was increased to 100 cm the grade decreased. The reason for this decreased concentrate grade is due to the froth becoming unstable. At the same time the cleaning action was reduced due to bubble coalescence.

The recovery decreased with an increased froth depth. This decrease in recovery was due to the better grades obtained with a froth depth of up to 75 cm. At a froth depth of 100 cm the recovery decreased further due to the froth being unstable and the bubbles coalescing. Another factor which influenced the decreased recovery with a deeper cleaning zone was that fewer large particles report to the concentrate (Figure 4.25). This was also reported by Dobby and Finch (1990).

The increase in grade due to an increased cleaning action is well recorded in literature and was confirmed by the test work. An unstable froth was only reported by Ynchausti and co-workers (1988) for a manganese system. Froth instability is usually due to a lack of water or solids in the froth and can be overcome by increasing the wash water rate and by adding frother to the wash water.

For column flotation operations the cleaning zone depth can be adjusted to control the concentrate grade. An improved grade will however reduce the recovery.

5.2.5 FEED SOLIDS PERCENTAGE

When the feed solids percentage was increased the sulphur recovery remained nearly constant at about 72% (Figure 4.26) until a point was reached where the sulphur recovery decreased dramatically to 59%. The grade remained more or less constant at 18 to 19% sulphur.

Increasing the feed solids percentage, while keeping all other parameters constant, did not affect the recovery and the grade significantly because the particle residence time did not vary. The recovery suddenly decreased as the maximum carrying capacity was reached. This was due to the froth becoming overcrowded which resulted in the larger particles (about 35 μm) being washed out of the froth. This phenomenon was clearly illustrated in figure 4.28.

In the experiments conducted the decrease in recovery of large particles was proof that the maximum carrying capacity was exceeded. To determine the maximum carrying capacity for a specific flotation condition, the feed solids percentage can thus be increased while all other parameters are kept constant.

5.2.6 VOLUMETRIC FEED RATE

By varying the volumetric feed rate to the flotation column an optimum performance in terms of sulphur grade was achieved. It can be hypothesised that the increased feed gives rise to a higher hydrophillic solids rate in the froth which in turn stabilises the froth. The more stable froth is more suitable to an even washing effect

At volumetric feed rates that are too low the froth may become unstable due to not enough particles being present to stabilise the froth. The grade can therefore decrease due to short circuiting in an unstable froth (Goodall and O'Connor, 1991).

In the laboratory tests the maximum carrying capacity was just about reached because larger particles (Larger than 30 microns) were lost to the tailings instead of being floated (Figure 4.29). There was no loss in recovery yet. When the froth becomes overcrowded the larger particles are washed out of the froth. This was also noticed when the feed solids percentage was increased.

If the volumetric feed rate is however increased without the maximum carrying capacity being reached, the particle residence time in the collection zone may be decreased to such an extent that the recovery will be inhibited.

5.2.7 WASH WATER

By increasing the wash water rate from 118 to 355 ml/min the recovery passed through an optimum at 212 ml/min while the grade decreased from about 30 % sulphur to about 26 % sulphur.

The wash water, as the name suggests, has the function of washing the collected particles in the cleaning zone. The decrease in the sulphur grade with an increased wash water rate can only be explained as follows. At low wash water rates (118 and 137 ml/min) it was visually observed that the wash water had a stabilising effect on the cleaning zone. Because the cleaning zone was stable the cleaning action of the wash water was producing high sulphur grades. When the wash water was increased the cleaning effect should have been increased. It is postulated however that the higher wash water rates caused the froth to be turbulent, resulting in a reduced cleaning action of the froth.

Changing the wash water position had no effect on the recovery or grade. It did however have an influence on the water recovered in the concentrate. By lowering the wash water position the froth above the wash water distributor has more time to allow the water to drain out of the froth.

The position of the wash water distributor can therefore be varied to obtain a concentrate with a higher solids density, thus reducing the dewatering capacity required downstream of the flotation circuit.

An increase in the wash water temperature did not affect the sulphur recovery significantly (Figure 4.8). The sulphur grade did however increase by 4%. This increase was related to an increased solids density in the concentrate. The increased wash water temperature decreased the viscosity and therefore more wash water was drained from the cleaning zone which resulted in more fine entrained particles being washed out of the cleaning zone. Wash water with a higher temperature therefore also resulted in an increased solids density.

5.2.8 FEED POSITION

By lowering the feed position the sulphur grade increased while the recovery decreased. The sulphur recovery decrease is due to the collection zone length being decreased and the column above the feed point acting as part of the cleaning zone. The sulphur grade increased because of the collection zone above the feed point acting as a cleaning zone and thus preventing any entrainment into the froth.

5.3 CHEMICAL PARAMETERS VARIED

5.3.1 Leaching the Ore

The ore was leached for 3 hours at a pH of 3,8 in order to investigate to what extent the flotation performance could be affected as a result of slight oxidation. The 3 % increase in sulphur recovery and 2 % decrease in sulphur grade observed was not considered significant enough to warrant using a lower pH in the investigation.

This increased recovery does however correspond to reports by Gaudin (1932).

5.3.2 Conditioning Procedure

Increasing the conditioning time by 5 minutes improved the sulphur recovery and grade by 3% and 5% respectively. Since the research was mainly focused on the effect of physical variables on the column performance no further investigation was carried out on the conditioning time.

It should however be noticed that by conditioning the ore in stages it was possible to increase the grade while the recovery remained constant. The reagent consumption was also lower by using staged conditioning. The conditioning procedure can be of great importance and the good results obtained warrant future research into this field.

5.3.3 Reagent Dosage

As expected the sulphur recovery increased as the collector addition rate was increased. At the same time the sulphur grade increased. This once again illustrates the fact that the correct collector addition rate is of great importance.

An increased frother dosage did not affect the sulphur recovery or grade. Only at a low frother dosage did the recovery decrease slightly due to the froth becoming unstable. The function of the frother is to keep a stable froth and to control the bubble sizes.

During the testwork a high frother concentration was used so that a stable froth was produced. The higher frother concentration did not have an effect on the sulphur recovery and grade and therefore did not affect the investigation of the physical parameters.

CHAPTER
SIX

PLANT TRIALS

Plant trials were done at the Buffelsfontein Gold Mine. These trials motivated the design and commissioning of a portable pilot plant flotation column. The aim was to run the column in a rougher, cleaner and rougher-scavenger mode and to evaluate the effect of varying parameters such as wash water rate, air flow rate, mean liquid and solids residence time and the solids feed rate.

6.1 PILOT PLANT FLOTATION COLUMN DESIGN

Pilot plant flotation column trials have numerous advantages in that the tests can be carried out at conditions as close as possible to the actual flotation plant conditions. The pilot plant column can be run in parallel to the existing flotation plant making direct comparison possible. The feed chosen for the column does not need additional treatment as is often the case in laboratory flotation tests. The flotation column can also be run continuously for much longer than in the laboratory. It was thus decided to design a pilot plant flotation column.

At the outset it was decided to design a pilot plant flotation column which was:

1) transportable;
2) self supporting with regards to compressed air and water;
3) designed for quick and easy assembly;
4) capable of running as 2 columns in parallel or in series; and
5) capable of running continuously from a holdup tank and making use of a conditioning tank.

To make the flotation column transportable every part of the flotation column rig had to be made small and compact. The water rotameters, air rotameters and the level controller were placed into a single control unit (Figure 6.1). Figure 6.2 shows the quick release joints of the flotation column enabling the column to be disassembled into 1.5 m sections. The holdup tank, water tank and conditioning tank can be placed into each other to save space in transit. A cabinet was made to transport the slurry pumps and to provide protection.

To make the pilot flotation column self supporting a portable compressor and high pressure water pump was bought (Figure 6.3). The only service required was a 220 V power supply and potable water.

The facility of being able to assemble and disassemble the flotation column quickly was considered important. The column sections all had quick release joints and ˝clip on˝ brackets (Figure 6.2) for quick assembly. The air and water tubes were all connected with quick release connections (Figures 6.1, 6.2, 6.3 & 6.4).

To be able to run 2 columns, the control unit was duplicated (Figures 6.3, 6.4 & 6.5). The air compressor and the high pressure water pump was used for both columns simultaneously. The pilot flotation column rig also included a holdup tank with mixer, a metering pump to add reagents and a conditioning tank.

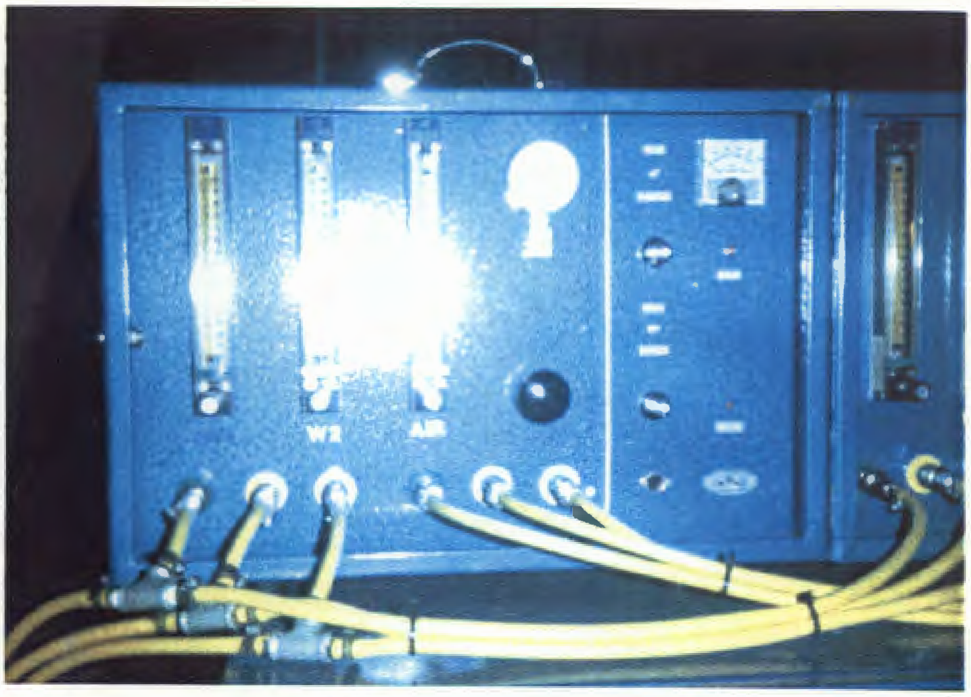

Figure 6.1: Control Unit

Fig 6.2: Quick release joints

Fig 6.3: Pilot plant equipment

Figure 6.4: Pilot column launder

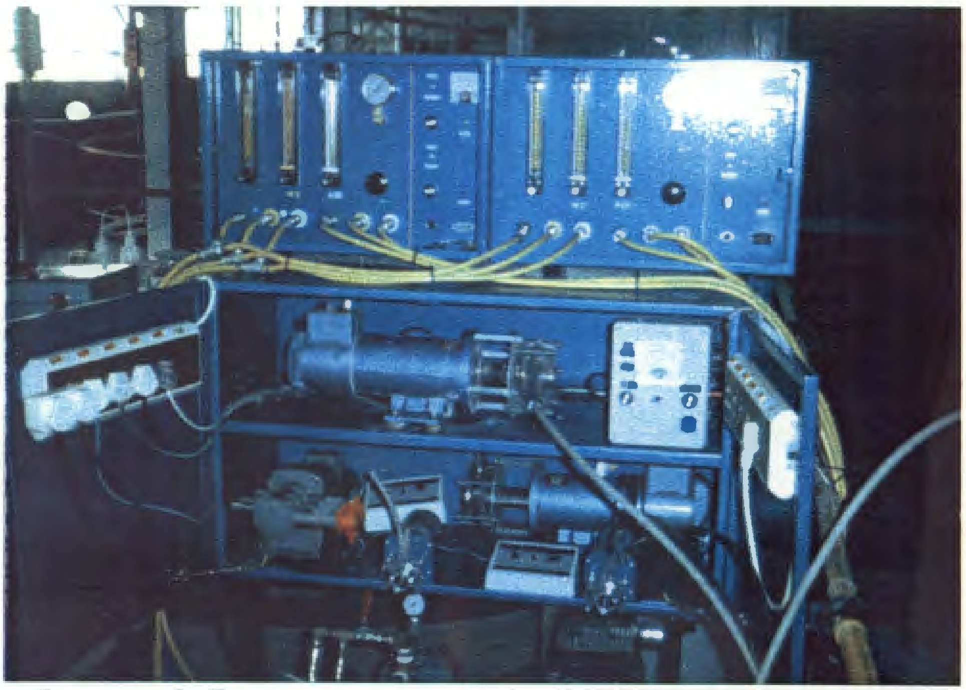

Figure 6.5: Complete pilot plant

6.2 EXPERIMENTAL PROGRAM

The plant trials were carried out over a period of two weeks at Buffelsfontein Gold Mine. The layout of the pyrite flotation plant is shown in Figure 6.6. Details for the pyrite floatation plant are given in Table 6.1. The values given are the averages for the 7 months between January 1989 to July 1989.

PLANT DETAILS

5 Banks of 16 cells
Volume per cell = 1,676 cubic meters

REAGENTS USED:
Senkol 12: 115 g/ton
Frother 6003: 36 g/ton
CuSO4: 45 g/ton

Feed = 1 352 000 tons per 7 months
(Up to 31/07/89)
Feed S.G. = 1.32 to 1.35
Solids Density = 2.73

Averages for 7 months up to 31/07/89.

Head Grade	1.35
Residue	0.48
Conc. Grade	35.19
Theo. Recovery	65.64
Actual Recovery	71.73

Table 6.1: Details of the Buffelsfontein Gold Mine Flotation Plant for 7 months up to 31/07/1989.

The feed for the rougher column flotation tests was taken from the feed to the rougher cell A. For the cleaner column flotation tests the feed was taken from the feed to the cleaner cells. No additional reagents were added to the pulp for the flotation column tests.

The experimental program was as follows:

WEEK 1: DAY 1: Setup flotation column
DAY 2: Do initial test run to determine suitable column rougher operating conditions.
DAY 3: Determine the effect of varying the wash water rate.
DAY 4: Determine the effect of varying the air flow rate.
DAY 5: Run the column as a cleaner.

WEEK 2: DAY 1: Run the small and large column in parallel for comparison.
DAY 2: Vary the collection zone length and the feed rate.
DAY 3: Run the column as a rougher, aiming for improved recovery.
DAY 4: Run the columns in series, viz. as a rougher and as a scavenger.
DAY 5: Disassemble column.

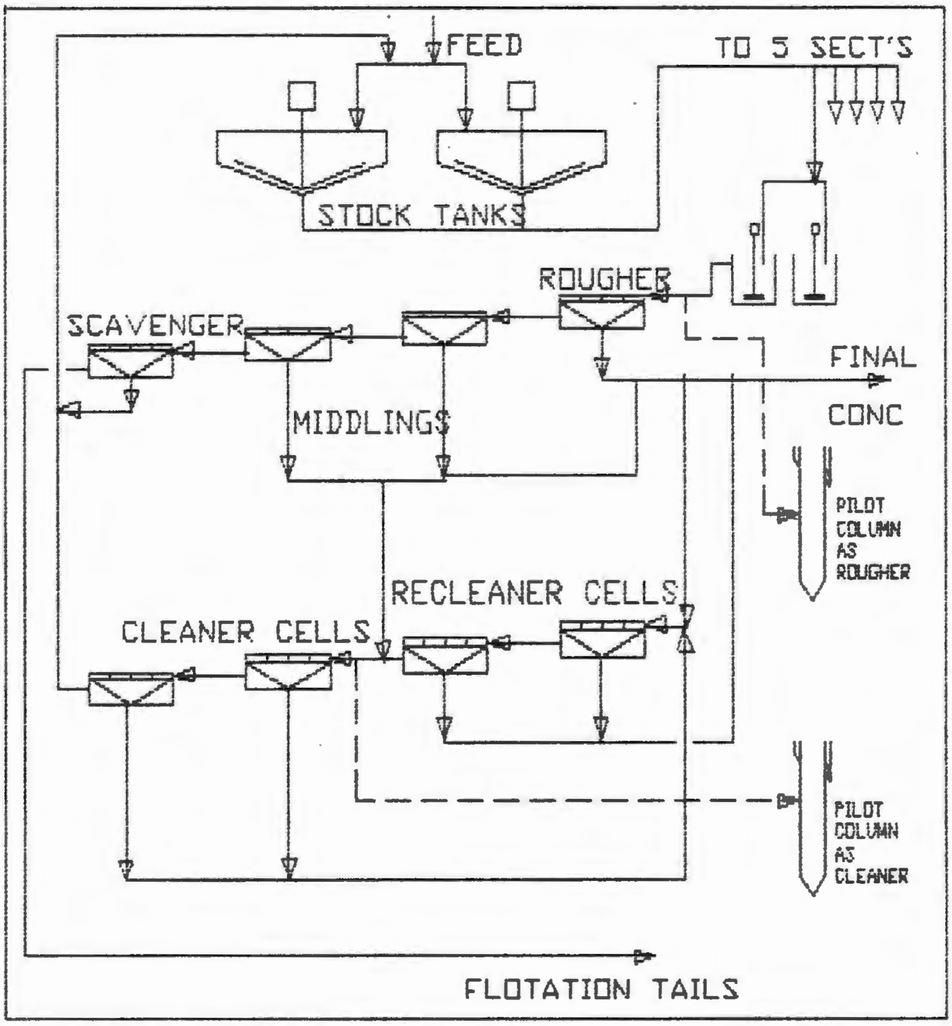

Figure 6.6: Buffelsfontein Gold Mine Flotation Plant layout.

6.3 RESULTS AND DISCUSSIONS

The detailed results obtained for the pilot plant testwork are tabulated in APPENDIX F.

6.3.1 Comparison of the column to the plant rougher

The performance of the pilot plant flotation column was compared to the performance of the plant rougher cell A and the total rougher bank. The best flotation column results for a specific day was compared to the plant results of that day. The feed for the rougher column flotation tests was taken from the feed to the rougher cell A. This ensures that the feed conditions are the same.

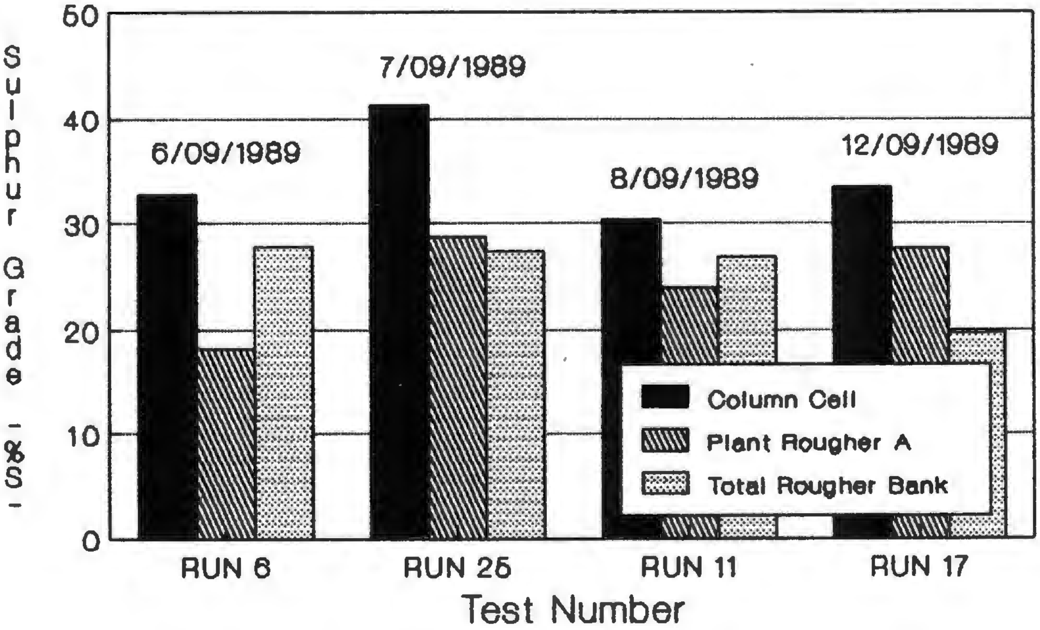

Figure 6.7: Comparing the concentrate grades of the rougher flotation column with the plant rougher A and total roughers.

The average concentrate sulphur grades for the pilot plant rougher flotation column were higher than the sulphur grades of the plant rougher A and sulphur grades from the total plant rougher bank (Figure 6.7). The flotation column can thus be used as a rougher to achieve higher concentrate sulphur grades than is possible with conventional flotation. This is due to the cleaning of the froth using wash water.

The average sulphur recoveries for the flotation rougher column were also higher than the sulphur recoveries calculated for the plant rougher (Figure 6.8). For run 17, on the 12/09/1989, the superficial air rate was 3,02 cm/sec which resulted in a recovery of 39% sulphur.

These improvements in recovery and grade over the plant confirm the results obtained in laboratory results.

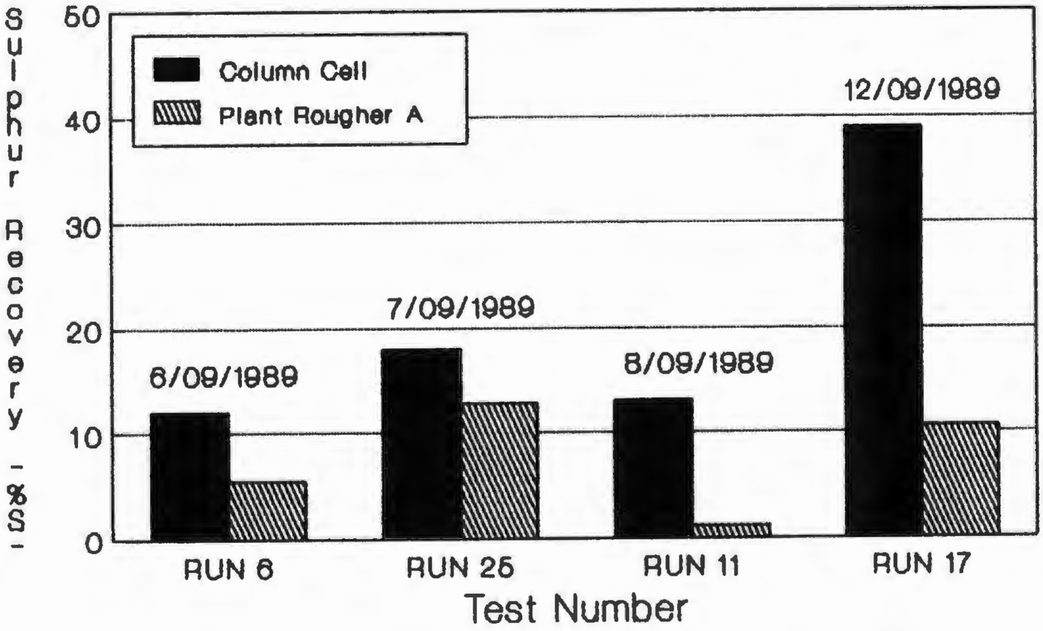

Figure 6.8: Comparing the sulphur recoveries of the rougher flotation column with the plant rougher A.

The gold recoveries of the rougher flotation column were compared to the plant rougher and the total plant gold recoveries (Figure 6.9). For run 19 on the 12/09/1989, the average gold recovery for the flotation column was about 61% compared to the plant rougher gold recovery of about 43% and the overall plant recovery of about 61%. This was only possible at superficial air flow rates greater than 3,02 cm/sec. This was probably because of the improved flotation of coarse pyrite particles (which contain more gold) at a high superficial air rate.

The results from run 19 again show that the flotation column can perform as good or better than conventional flotation. In this case (Run 19) the pilot plant column did the work the whole plant flotation circuit did.

Figure 6.10 shows that the gold grade was similar for run 19 and for the plant on the 12/09/1989. For run 11 the gold grade obtained with the column was lower than that obtained by the plant rougher and total rougher bank (11/09/1989).

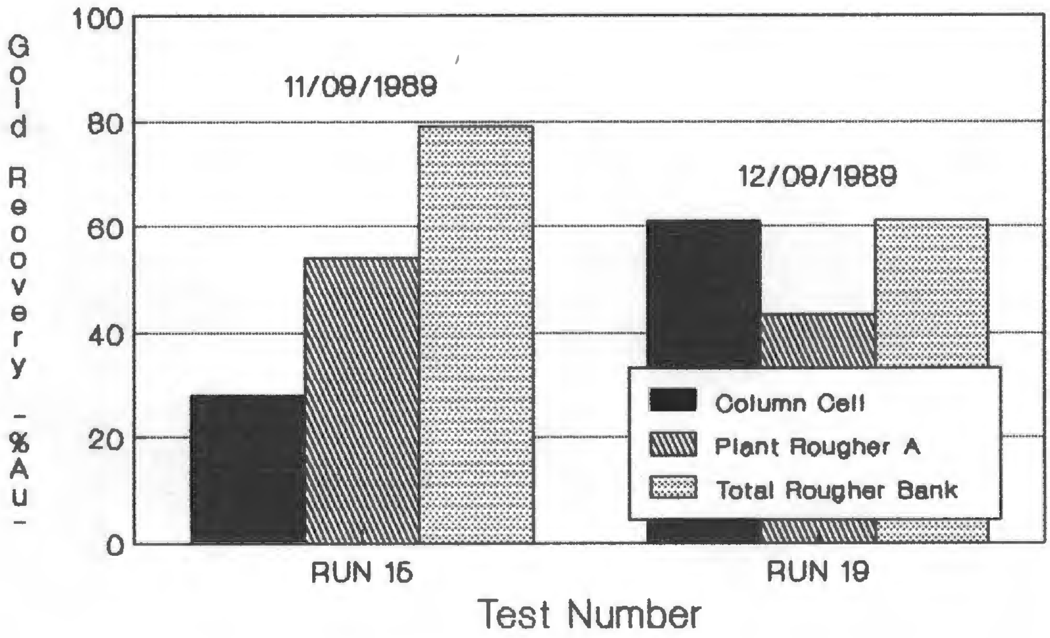

Figure 6.9: Comparing the gold recovery of the rougher flotation column with the plant rougher A and total plant roughers.

Figure 6.11 shows that the theoretical mean liquid residence time (Volume/Volumetric flow rate) for the pilot plant flotation column in rougher mode was nearly twice as long as that of the plant rougher. On the 7/09/1989 when the mean liquid residence time of the

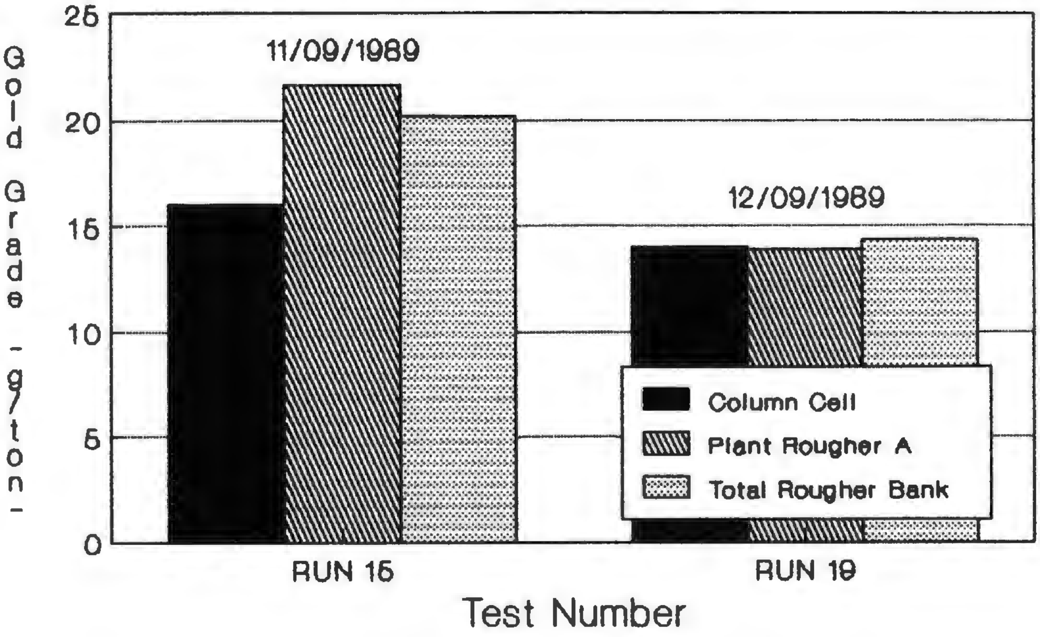

Figure 6.10: Comparing the concentrate grades of the rougher flotation column with the plant rougher A and total roughers.

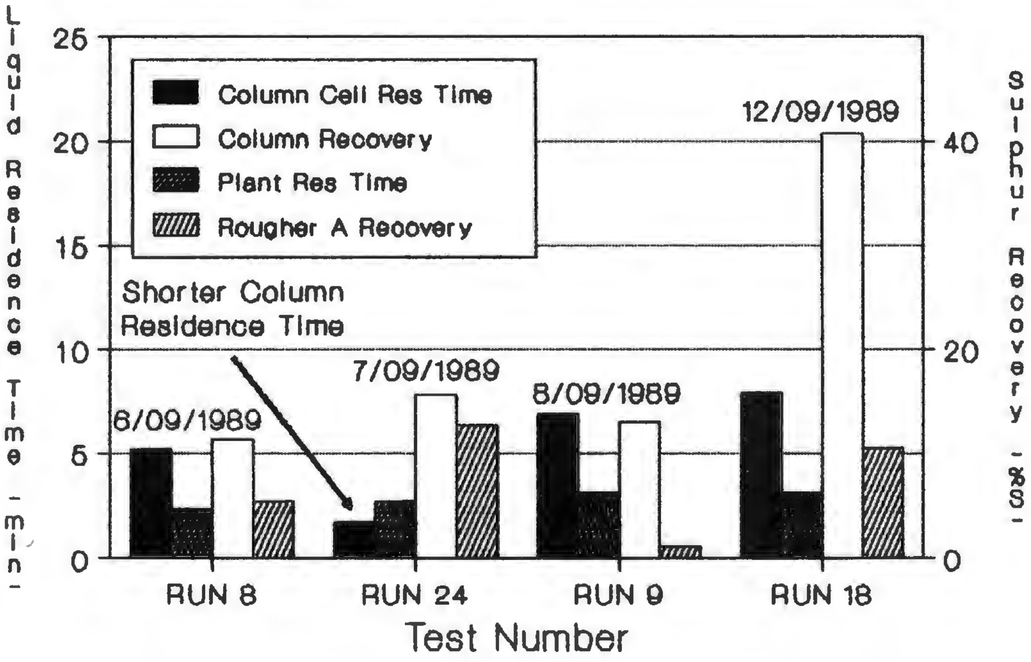

Figure 6.11: Comparing the mean liquid residence time of the rougher flotation column with the plant rougher A.

flotation column was less than that of the plant rougher cell due to a high volumetric feed rate being used. The sulphur recovery obtained with the column was still more than that of the plant rougher.

6.3.2 Comparing the column to the plant cleaner

The pilot plant flotation column was compared to the plant cleaner cells. For these cleaner column flotation tests the feed for the column was taken from the feed to the cleaner cells. A sample from the feed to the plant cleaner was taken and pumped into the holdup tank. From here the pulp was pumped into the flotation column.

The sulphur grades obtained with the cleaner flotation column are in the range of 15.9% to 19.6% compared to the 16.0% obtained with the plant cleaner (Figure 6.12). The sulphur recoveries obtained in the flotation column are much lower though than the theoretical recoveries of the plant cleaner cells. This low sulphur recovery was due to the column operating above the maximum carrying

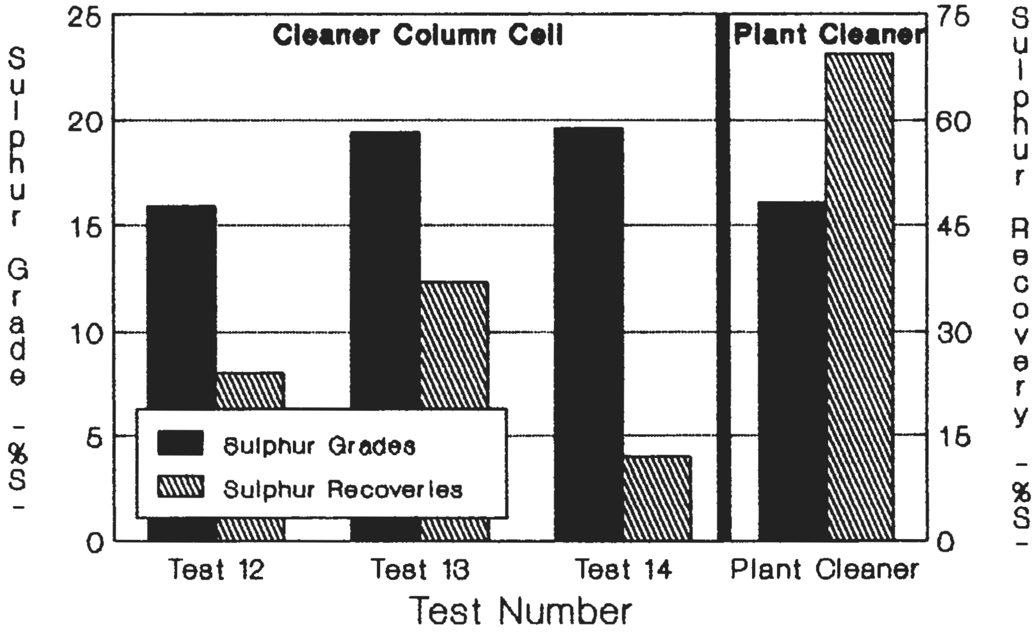

Figure 6.12: Comparing the sulphur grade and recoveries between the cleaner flotation column and the plant cleaner cell.

capacity. It was visually observed that the maximum carying capacity was reached as most of the collected solids were washed back into the collection zone. The feed to the column could not be reduced due to physical limitations of the solids settling out in the feed lines at lower feed rates.

6.3.3 Using the column in a rougher-scavenger mode.

The pilot plant flotation column was used in a rougher-scavenger mode. The feed for the flotation column was taken from the feed to the plant rougher A. This would determine the maximum sulphur recovery possible if no additional reagents are added. In the plant flotation circuit additional collector is added to the feed at the 2nd middlings float and frother is added to every flotation bank. No additional reagents were added to the column flotation circuit.

The sulphur grade from the rougher flotation column was 20.0%, and 11.5% from the scavenger flotation column. This resulted in a grade of 19.1% for the total concentrate. To obtain better grades the air rate would have to be reduced and the low grade sulphur concentrate from the scavenger flotation column should possibly be returned to the feed for the rougher column. The effect of varying the superficial air rate is discussed section 6.3.5.

The sulphur recovery in the rougher flotation column was 47% and in the scavenger column 5% (Of total sulphur feed to plant). This resulted in a total sulphur recovery of 52% for the column operating in a rougher-scavenger mode. This sulphur recovery is much lower than the theoretical plant recovery (Calculated by Buffelsfontein Mine) of 65.6% and the actual plant sulphur recovery of 71.7% (Plant values are average values for 7 months up to 31/07/1989). The reason for this low sulphur recovery with the flotation column setup was possibly due to less reagents being added to the column than to the overall flotation plant. Less reagents were added to the pilot plant flotation column because the plant has additional reagent feed points to the flotation cells.

6.3.4 Effect of varying the superficial wash water rate.

By increasing the superficial wash water rate from 0 cm/sec to 0.18 cm/sec the concentrate sulphur grade increased from 22.0% to 32.8%. This improved grade was also reported in the literature (Kosick, G.A. and Kuehn, L., 1988; Nicol, et al., 1988; Ynchausti, et al., 1988; Parekh, et al., 1988). The grade was improved because entrainment is reduced with increased wash water addition. Increasing the superficial wash water rate above 0.18 cm/sec did not further improve the

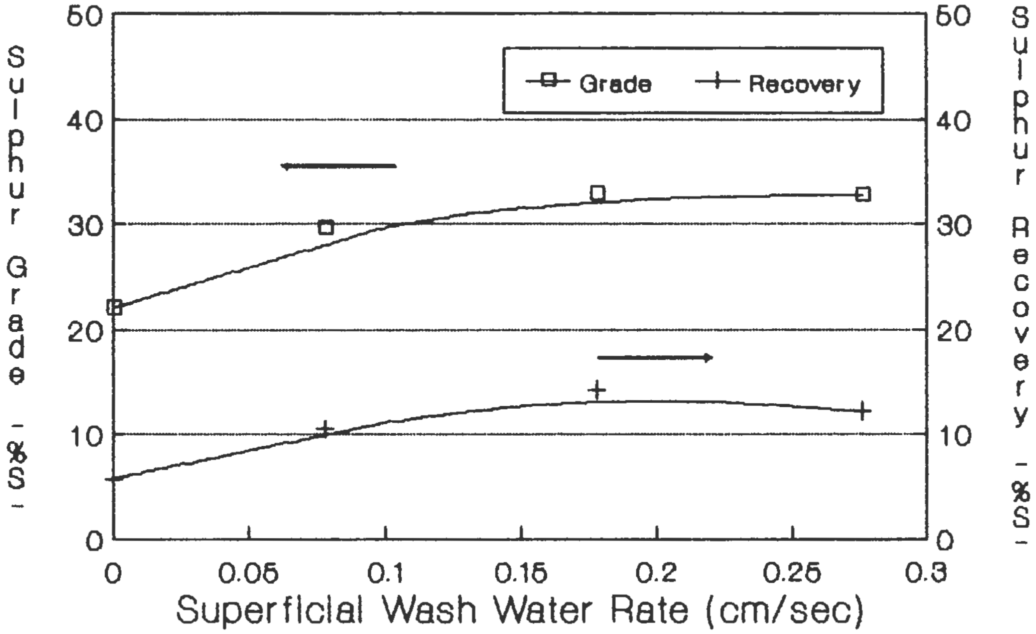

Figure 6.13: The effect of increasing the wash water rate on the sulphur grade and recovery for the flotation column.

grade (Figure 6.13). The maximum grade obtainable by varying the wash waster rate was thus 32.8% sulphur at a superficial wash water rate of 0.18 cm/sec or greater.

The sulphur recovery improved from 6% to 14% as the superficial wash water rate was increased from 0 cm/sec to 0.18 cm/sec (Figure 6.13).

The sulphur recoveries were so low because high feed rates were used. The column was thus operated above the maximum carrying capacity with the result that most of the floated material dropped back into the collection zone. This was observed visually.

Increasing the superficial wash water rate above 0.18 cm/sec decreased the sulphur recovery to 12%. That excess wash water decreases the recovery was reported by Kosick (1988) and Luttrell (1988). This decrease in sulphur recovery at high wash water rates was also observed in the laboratory tests.

The mass recovery reached a maximum of 0.69% at a superficial wash water rate of 0.18 cm/sec (Figure 6.14). This trend influenced the sulphur recovery as discussed above. The increased wash water addition rate increased the froth carrying capacity due to the froth becoming more stable. The stability prevents the bubbles from coalescing, and therefore reducing the carrying capacity. At excessively high wash water addition rates collected solids are washed out of the froth. The mean liquid residence time (Estimated) was also decreased which resulted in a lower recovery.

The solids density in the concentrate decreased as the superficial wash water rate was increased. The decrease in the solids density of the concentrate was much greater at superficial wash water velocities greater than 0.18 cm/sec. This increased water recovery in the concentrate was water from the additional wash water added and not from the feed pulp.

These trends were also observed in test carried out with the column operating as a cleaner (Figures 6.15 & 6.16).

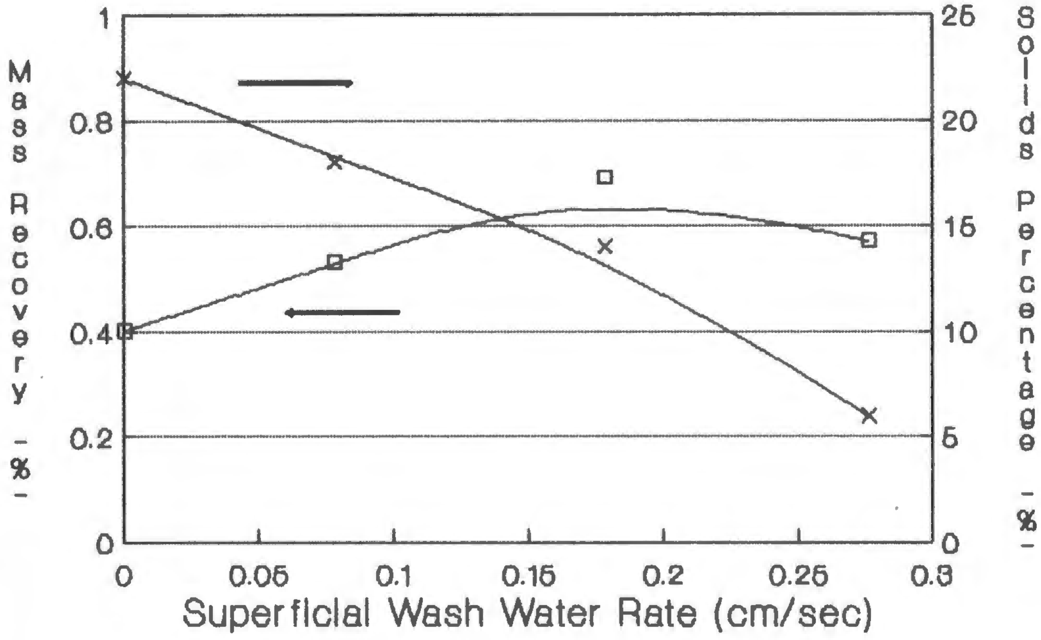

Figure 6.14: The effect of wash water rate on the percent solids recovered and the solids percent (mass/mass) in the concentrate.

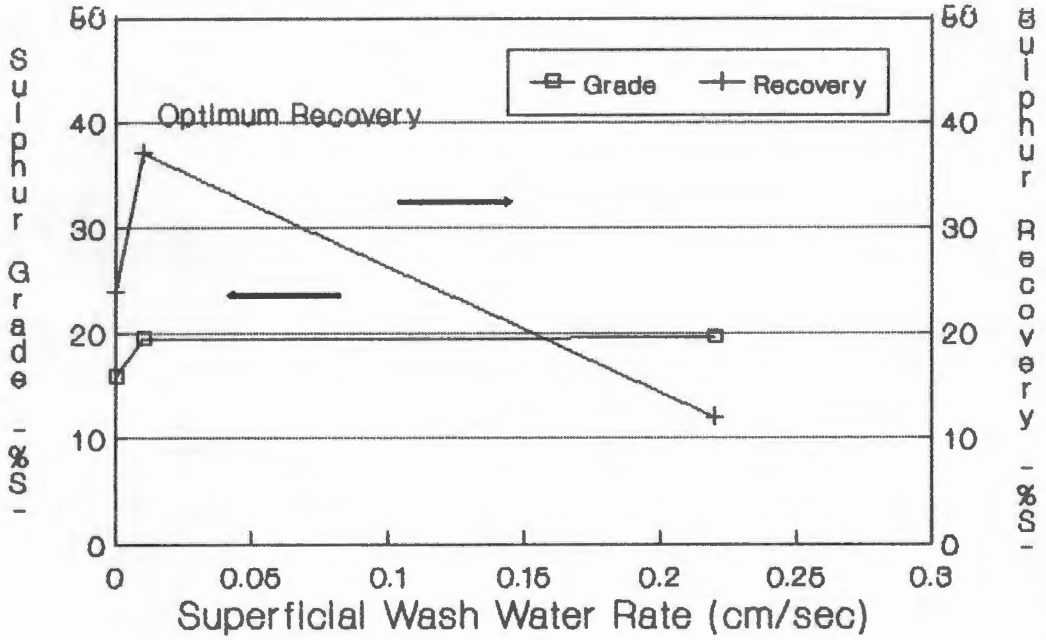

Figure 6.15: The effect of increasing the wash water rate on the sulphur grade and recovery for the cleaner column.

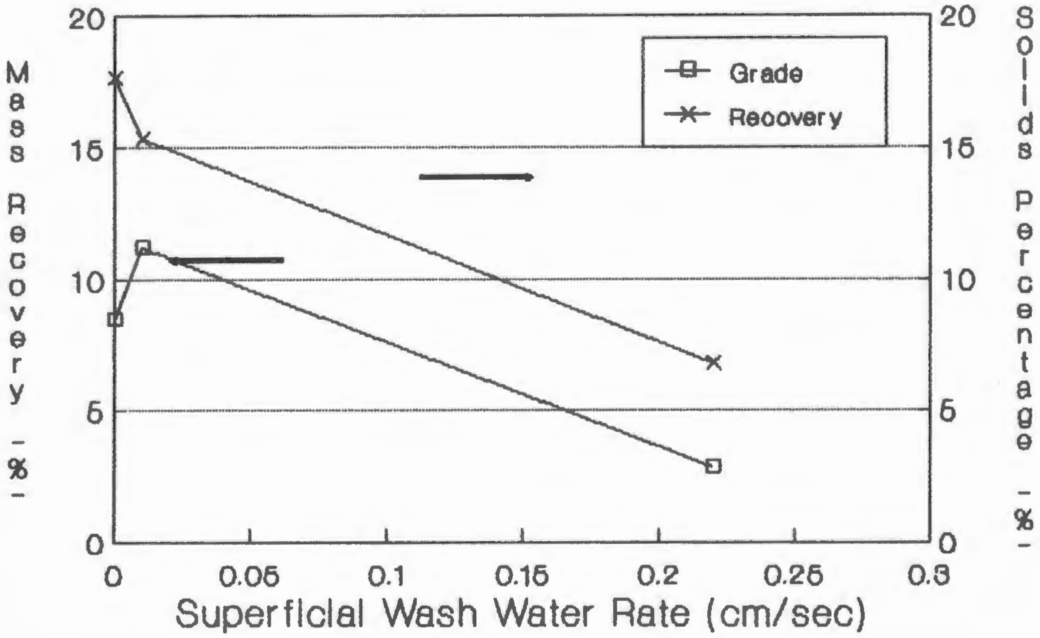

Figure 6.16: The effect of wash water rate on the percent solids recovered and the solids percent (mass/mass) in the column cleaner concentrate.

6.3.5 The effect of varying the superficial air velocity

Increasing the superficial air velocity from 1.22 cm/sec to 2.06 cm/sec decreased the concentrate sulphur grade from 38.3% to 30.8% (Figure 6.17). This decrease in the grade with increasing air rate has been shown by several researchers (Goodall, C.M. and O'Connor, C.T., 1989; Misra, M. and Harris, R.,1988; Parekh, et al., 1988). This decrease in the concentrate grade is due to increased entrainment caused by increased air velocities.

During a subsequent test the superficial air rate was increased from 3.02 cm/sec to 3.57 cm/sec. The concentrate sulphur grade decreased from 33.4% to 17.6%. When the superficial air rate was further increased to 4.19 cm/sec, the concentrate grade improved to 20%.

The sulphur recovery reached an optimum of 14% at a superficial air rate of 1.86 cm/sec (Figure 6.17). This optimum in recovery was repeatedly reported (Dobby, D.S. and Finch, J.A., 1986; Parekh, et al., 1988; Goodall, C.M. and O'Connor, C.T., 1989; Luttrell, G.H., 1988). The optimum in the sulphur recovery is due to the bubbles coalesing. Initially the carying capacity increases with an incresed air rate, but when the bubbles start to coalesce, the carrying capacity is reduced and therefore the recovery starts to decrease.

When the superficial air rates were increased during a subsequent test (from 3.02 cm/sec to 4.19 cm/sec), the sulphur recovery increased from 39% to 47%. This increase is due to a higher mass recovery at a high superficial air rate.

The percent solids recovered in the concentrate at low superficial air flow rates (up to 2.06 cm/sec) increased up to a maximum of 0.67% +-0.02% (Figure 6.18). At high superficial air rates (greater than 3.02 cm/sec) the percent solids reporting to the concentrate increased from 1.66% to 3.74% (Obtained in a subsequent test).

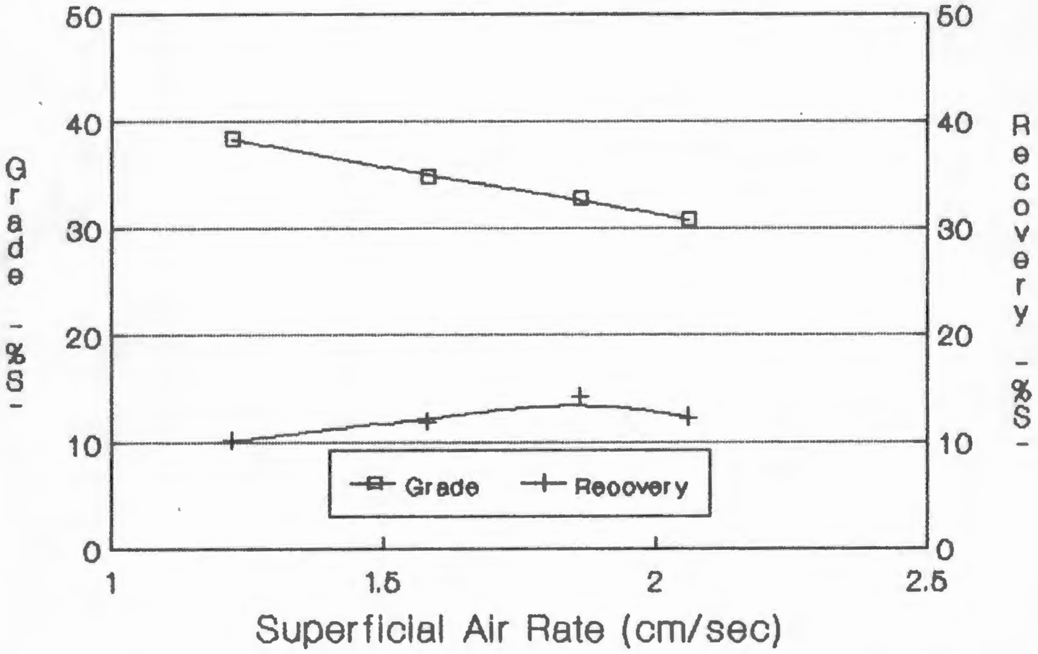

Figure 6.17: The effect of air flow rate on the grade and the recovery for the pilot plant flotation column.

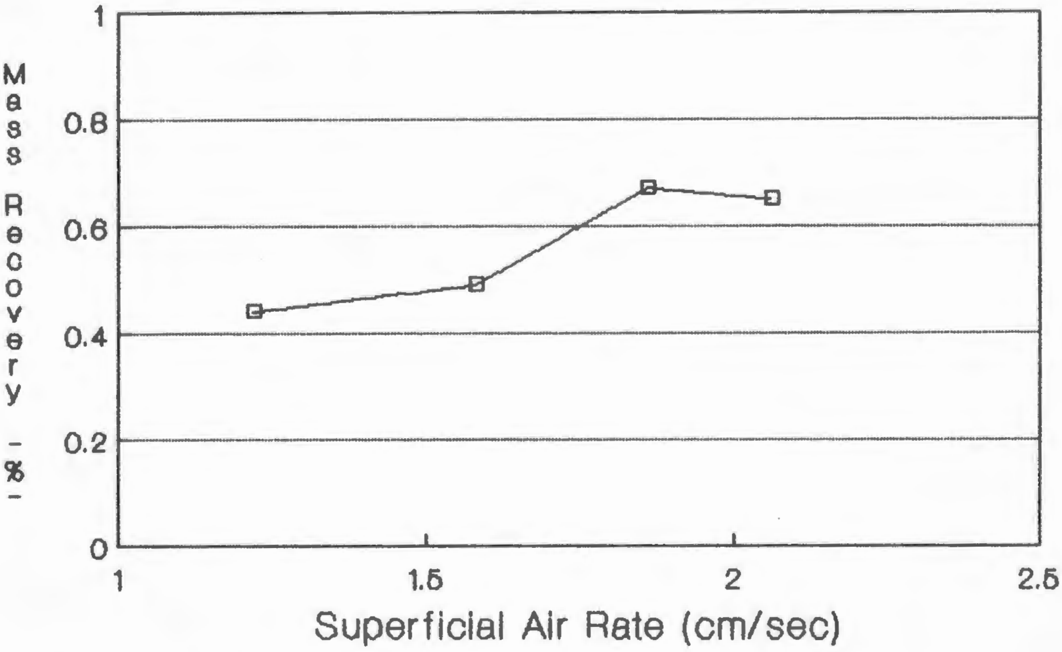

Figure 6.18: The effect of air flow rate on the percent solids recovered.

The effect of increasing the superficial air flow rate from 1.83 cm/sec to 3.60 cm/sec increased the gold recovery from 27% to 63% (Figure 6.19).

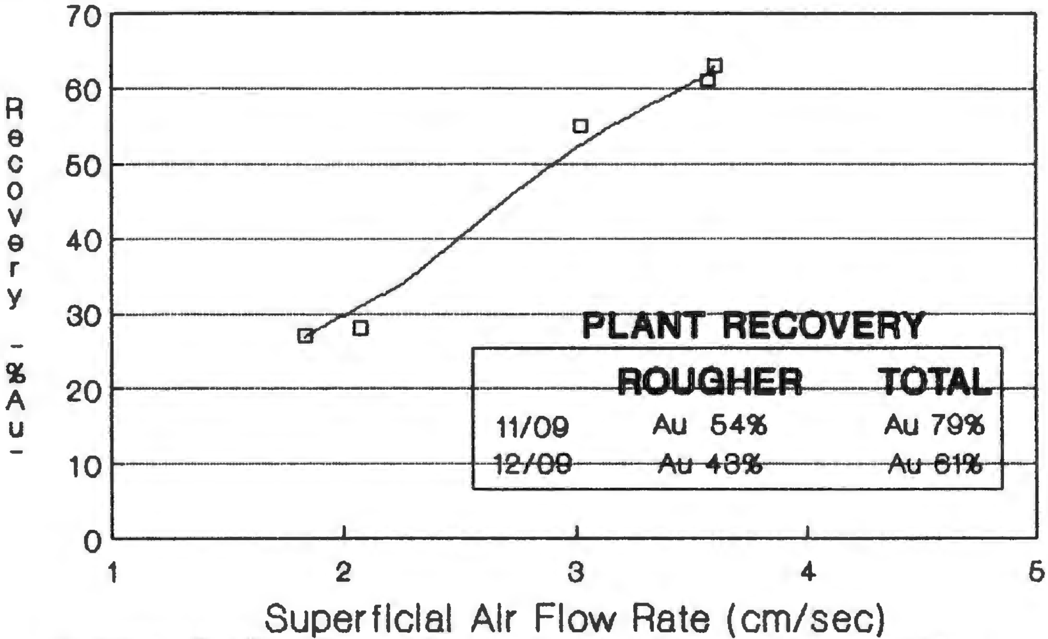

Figure 6.19: The effect of air flow rate on the total gold recovery.

6.4 SUMMARY OF THE PLANT TRIALS

1) The pilot plant column flotation cell can be used to do on site test work.

2) The column flotation cell was used on a pyrite flotation circuit as a rougher cell producing improved sulphur grades. Sulphur recoveries were however lower. The gold recoveries and grades were similar to those of the plant.

3) The pilot plant flotation column yielded improved sulphur recoveries at shorter residence times than the plant rougher cell.

4) By increasing the wash water rate the sulphur concentrate grade can be improved.

5) By increasing the air rate the gold recovery can be increased.

CHAPTER
SEVEN

CONCLUSIONS

For this research on the determination of the effects of physical and chemical parameters on the flotation column cell performance in the flotation of pyrite, a laboratory and a pilot flotation column were designed and built. Using the flotation columns with the experimental procedure developed produced highly repeatable results.

In view of the experimental test work conducted the following conclusions can be drawn:

7.1 The flotation column produced better sulphur grades than conventional flotation due to the flotation column´s deep, water washed froth, reducing entrainment into the concentrate. This was confirmed by plant test work.

7.2 The flotation columns yielded higher sulphur recoveries than conventional flotation at lower residence times due to the flotation columns counter-current contact mechanism of the bubbles and the particles.

7.3 An increased air rate yielded an optimum in sulphur recovery. The increase in recovery was due to an increased mass pull which resulted from an increased superficial bubble surface rate as the air rate was increased. The optimum in sulphur recovery was due to the superficial bubble surface rate decreasing at high air flow rates.

7.4 The concentrate sulphur grades decreased as the air rate was increased. This was because of increased entrainment.

7.5 Increasing the collection zone length increased the sulphur recovery to its maximum because of an increased solids residence time. Maximum recovery was obtained at a 4 meter collection zone length.

7.6 The sulphur grade decreased while the sulphur recovery increased when the collection zone length was increased.

7.7 Increasing the froth depth decreased the recovery. This decrease was due to an improved cleaning effect and froth instability at deep froth depths.

7.8 The grade increased as the froth depth was increased due to an increased cleaning action of the froth. The cleaning action was however reduced when the froth depth was increased too much, due to the froth collapsing at times.

7.9 Increasing the feed solids percentage to the flotation column had no effect on the grade or recovery. The recovery did however decrease when the maximum carrying capacity was reached.

7.10 The volumetric feed rate to the flotation column should be high enough so that there are enough particles to stabilise the froth. A high volumetric feed rate therefore yielded better concentrate grades than a low volumetric feed rate.

7.11 Increasing the volumetric feed rate resulted in operating at the maximum carrying capacity. A further increase in volumetric feed rate, or a too short solids residence time, would decrease the recovery.

7.12 Increasing the wash water rate increased the concentrate grades. This was due to an increased froth stability and a higher washing rate. When the wash water was increased too much, the froth became unstable, resulting in a decreased concentrate grade.

7.13 Lowering the wash water distributor decreased the water recovered into the concentrate. This was due to more time being allowed for drainage of the water.

7.14 Wash water at a high temperature increased the concentrate grades while less water was recovered. The wash water at a higher temperature has a lower viscosity and therefore it drains out of the froth easily, increasing the cleaning effect.

7.15 Lowering the feed position increased the grade while the recovery decreased. This was caused by an increase in the cleaning zone and a decrease of the collection zone.

7.16 When the flotation column was used the chemical parameters which were varied showed the same trends as conventional flotation. Leaching the ore and adding more collector increased the recovery.

7.17 At low frother dosage rates the froth became unstable and the recovery decreased. It is therefore important to maintain a stable froth.

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APPENDIX A

CALCULATIONS FOR LABORATORY FLOTATION COLUMN DESIGN

The method used to design the laboratory flotation column was based on work done by Yianatos (1987).

Using a typical superficial gas rate, Jg, of 2 cm/sec the superficial gas rate at standard conditions (atmospheric) can be calculated as follows:

$$Jg^{*} = \frac{(P_t - P_c)\cdot Jg}{P_c \cdot \ln(P_t/P_c)}$$

P_t is the pressure at the air input level and can be calculated as follows:

$$P_t = \mu \cdot g \cdot H + 101.3$$

μ For the pulp was estimated at 1.33 kg/m^3 and H, the height of the collection zone was taken as 2 meters. Using this collection zone height and a column diameter of 5.4 cm yields a height to diameter ratio, H/D, of 37.

$$\text{Thus} \quad P_t = 1.33 \cdot 9.81 \cdot 2 + 101.3$$
$$= 127 \text{ kPa}$$

P_c, the pressure at the concentrate lip is at atmospheric pressure and is taken as 101.3 kPa.

Therefore:

$$Jg^{*} = \frac{(127 - 101.3)\cdot 2}{101.3 \cdot \ln(127/101.3)}$$
$$= 2.25 \text{ cm/sec}$$

The bubble diameter, d_b, was estimated as 0.2 cm. The bubble diameter at the top of the froth, d_b*, is estimated by assuming a 50% increase in bubble diameter:

$$\begin{aligned} d_b* &= 2 \cdot d_b \\ &= 2 \cdot 0.2 \\ &= 0.4 \text{ cm} \end{aligned}$$

Now the superficial bubble surface rate, Js, can be calculated.

$$\begin{aligned} Js &= \frac{6 \cdot Jg*}{d_b*} \\ &= \frac{6 \cdot 2.25}{0.4} \\ &= 34 \text{ cm}^2\text{/s/cm}^2 \end{aligned}$$

The 80% passing particle size, d_{80}, was estimated to be about 50 microns.
Using the following equation to bias the smaller particles:

$$\begin{aligned} d_p &= 0.3 \cdot d_{80} \\ &= 0.3 \cdot 50 \\ &= 0.0015 \text{ cm} \end{aligned}$$

Assuming that each particle is spherical and occupies d_p^2 of the bubble surface, then the bubble surface required per gram of solids is:

$$\begin{aligned} Sr &= \frac{6}{Pi \cdot d_p \cdot \rho(p)} \\ &= \frac{6}{Pi \cdot 15 \cdot 2.23} \\ &= 571 \text{ cm}^2\text{/g} \end{aligned}$$

Assuming that each particle is shared by two bubbles at the overflow, the effective superficial bubble surface rate, Jse is:

$$\begin{aligned} Jse &= 0.5 \cdot Js \\ &= 17 \text{ cm}^2\text{/s/cm}^2 \end{aligned}$$

The superficial floated particle rate, Jpf, is therefore given by:

$$Jpf = \frac{3600 \cdot Jse}{Sr}$$

$$= \frac{3600 \cdot 17}{571}$$

$$= 107 \text{ g/hr/cm}^2$$

The mass flow rate of concentrate solids, C, therefore equals:

$$C = \frac{Jpf \cdot A_c}{1000}$$

$$= \frac{107 \cdot 22.9}{1000}$$

$$= 2.45 \text{ kg/hr}$$

$$= \underline{41 \text{ g/min}}$$

Assuming a feed grade, f, of 2%, a concentrate grade, c, of 30% and a tailings grade, t, of 0.2% the feed mass flow rate, F, is calculated as follows:

$$F = C \cdot \frac{(c-t)}{(f-t)}$$

$$= 41 \cdot \frac{(30 - 0.2)}{(2 - 0.2)}$$

$$= 679 \text{ g/min}$$

From the percent solids

$$\text{Mass of pulp per minute} = \frac{100 \cdot 679}{30}$$

$$= 2263 \text{ g/min}$$

Using a pulp density of 1.33 the volumetric pulp feed can be calculated:

$$\text{Volumetric pulp feed rate} = \frac{2263}{1.33}$$

$$= \underline{1.7 \text{ l/min}}$$

APPENDIX B

DATA FOR BUBBLE SIZING

The bubble size determination was done in the laboratory flotation column. In all cases the bubbles were measured 200 cm above the spargers. The type of spargers and the frother concentrations used are listed below:

Run 1:	Sintered Glass Disc Sparger	0 ppm
Run 2:	Sintered Glass Disc Sparger	2 ppm
Run 3:	Sintered Glass Disc Sparger	4 ppm
Run 4:	Filter Cloth Sparger	0 ppm
Run 5:	Filter Cloth Sparger	4 ppm
Run 6:	USBM Sparger	0 ppm
Run 7:	USBM Sparger	2 ppm
Run 8:	USBM Sparger	4 ppm
Run 9:	Filter Cloth Sparger	40 to 41 g/ton
Run 10:	USBM Sparger	74 to 107 g/ton

All the runs were done in a two phase system except for run 9 and 10. The frother concentration for run 9 and 10 are therefore given in g/ton.

Test No.	Data File	Sparger Type	2 or 3 Phase	Frother Conc.	Sup. Air Rate	Vol. Air Coll.	Bubble Diameter	Std Deviation
1-6	*name*			*ppm*	*cm/sec*	*ml*	*mm*	
1.1	KCS0P1	SGDS	2	0	0.73	6.4	2.1	--
1.2	KCS0P2	SGDS	2	0	1.24	8.6	2.1	--
1.3	KCS0P3	SGDS	2	0	1.86	6.5	2	--
1.4	KCS0P4	SGDS	2	0	2.62	4.3	1.8	--
2.1	KCS2P1	SGDS	2	2	0.73	2.8	1.1	--
2.2	KCS2P2	SGDS	2	2	1.31	2.9	1.2	--
2.3	KCS2P3	SGDS	2	2	1.89	3.7	1.3	--
2.4	KCS2P4	SGDS	2	2	2.62	5.0	1.4	--
2.5	KCS2P5	SGDS	2	2	3.49	3.5	1.4	--
3.1	KCS4P1	SGDS	2	4	1.02	2.7	1.1	--
3.2	KCS4P2	SGDS	2	4	1.31	2.2	1.2	--
3.3	KCS4P3	SGDS	2	4	1.82	3.5	1.3	--
3.4	KCS4P4	SGDS	2	4	2.62	4.1	1.3	--
4.1	KF0P1	FCS	2	0	0.47	30.4	2.5	--
4.2	KF0P2	FCS	2	0	0.79	29.0	--	--
4.3	KF0P3	FCS	2	0	1.1	28.2	2.4	--
4.4	KF0P4	FCS	2	0	1.41	27.2	--	--
4.5	KF0P5	FCS	2	0	1.73	24.8	2.2	--
4.6	KF0P6	FCS	2	0	2.04	23.9	--	--
4.7	KF0P7	FCS	2	0	2.35	22.8	2.1	--
4.8	KF0P8	FCS	2	0	2.66	23.2	2.2	--
4.9	KF0P9	FCS	2	0	2.98	22.1	2.1	--
4.10	KF0P10	FCS	2	0	3.29	22.5	2.1	--
5.1	KF4P10	FCS	2	4	0.47	29.0	2.3	--
5.2	KF4P11	FCS	2	4	0.79	28.8	--	--
5.3	KF4P7	FCS	2	4	0.79	29.0	--	--
5.4	KF4P6	FCS	2	4	1.09	27.4	--	--
5.5	KF4P4	FCS	2	4	1.41	23.6	2.2	--
5.6	KF4P5	FCS	2	4	1.73	25.6	--	--
5.7	KF4P3	FCS	2	4	2.04	21.2	2.1	--
5.8	KF4P2	FCS	2	4	2.35	20.0	2	--
5.9	KF4P1	FCS	2	4	2.66	19.4	2	--
5.10	KF4P8	FCS	2	4	2.98	17.7	--	--
5.11	KF4P9	FCS	2	4	3.29	16.4	1.9	--
6.1	KSUS0P1	USBM	2	0	0.98	1.8	2.53	--
6.2	KSUS0P2	USBM	2	0	1.30	5.0	2.47	--
6.3	KSUS0P3	USBM	2	0	1.62	4.7	2.36	--
6.4	KSUS0P4	USBM	2	0	1.94	3.6	2.23	--
6.5	KSUS0P5	USBM	2	0	2.27	4.0	2.18	--

Test No.	Data File	Sparger Type	2 or 3 Phase	Frother Conc.	Sup. Air Rate	Vol. Air Coll.	Bubble Diameter	Std Deviation
7–10	*name*			*ppm*	*cm/sec*	*ml*	*mm*	
7.1	KSUS2P1	USBM	2	2	0.98	1.50	1.0	--
7.2	KSUS2P2	USBM	2	2	1.30	0.80	1.1	--
7.3	KSUS2P3	USBM	2	2	1.62	0.55	1.1	--
7.4	KSUS2P4	USBM	2	2	1.94	0.90	1.1	--
7.5	KSUS2P5	USBM	2	2	2.27	0.70	1.2	--
8.1	KSUS4P1	USBM	2	4	0.98	0.50	1.2	--
8.2	KSUS4P2	USBM	2	4	1.30	0.60	1.2	--
8.3	KSUS4P3	USBM	2	4	1.62	0.85	1.4	--
8.4	KSUS4P4	USBM	2	4	1.94	0.75	1.2	--
8.5	KSUS4P5	USBM	2	4	2.27	0.55	1.2	--
9.1	r1	FCS	3	42	0.88	4.3	2.12	0.55
9.2	r2	FCS	3	42	0.88	0.7	2.05	0.60
9.3	r3	FCS	3	42	0.88	17.8	2.04	0.58
9.4	r4	FCS	3	41	1.06	17.1	2.04	0.63
9.5	r5	FCS	3	41	1.06	18.4	1.96	0.58
9.6	r6	FCS	3	41	1.06	15.6	2.08	0.59
9.7	r7	FCS	3	42	1.31	18.6	2.08	0.64
9.8	r8	FCS	3	42	1.31	19.6	2.07	0.63
9.9	r9	FCS	3	42	1.31	19.2	--	--
9.10	r10	FCS	3	40	1.52	19.1	2.13	0.64
9.11	r11	FCS	3	40	1.52	22.1	2.14	0.67
9.12	r12	FCS	3	40	1.52	21.4	1.95	0.65
9.13	r13	FCS	3	42	1.74	21.0	2.18	0.66
9.14	r14	FCS	3	42	1.74	21.9	2.16	0.66
9.15	r15	FCS	3	42	1.74	22.3	--	--
9.16	rr1	FCS	3	41	2.25	22.7	2.18	0.67
9.17	rr2	FCS	3	41	2.25	24.0	2.18	0.69
9.18	rr3	FCS	3	41	2.25	23.0	2.16	0.69
10.1	rr2	USBM	3	107	0.60	20.0	--	--
10.2	rr3	USBM	3	107	0.60	21.4	2.16	0.066
10.3	rr4	USBM	3	107	0.60	21.0	2.17	0.063
10.4	rr5	USBM	3	106	1.10	17.2	2.04	0.053
10.5	rr6	USBM	3	106	1.10	15.9	1.99	0.054
10.6	rr7	USBM	3	106	1.10	15.2	1.96	0.053
10.7	rr8	USBM	3	101	1.51	10.0	1.66	0.050
10.8	rr9	USBM	3	101	1.51	1.7	1.77	0.045
10.9	rr10	USBM	3	101	1.51	7.8	1.72	0.050
10.10	rr11	USBM	3	101	1.51	10.8	1.75	0.055
10.11	rr12	USBM	3	74	2.33	22.5	2.07	0.078
10.12	rr13	USBM	3	74	2.33	19.6	1.99	0.071
10.13	rr14	USBM	3	74	2.33	19.1	2.01	0.073

APPENDIX C

CALCULATIONS USED TO EVALUATE DATA

The calculations used to evaluate the data in Appendix D and Appendix F are given below. Both the recorded and calculated data for the laboratory test work are given in Appendix D. The plant trial test data and the corresponding calculated data is given in Appendix F.

The measured quantities were the following:

T - Mass of tailing solids;
T' - Total mass of tailings;
C - Mass of concentrate solids;
C' - Total mass of concentrate;
t - Tailings sulphur grade;
c - Concentrate sulphur grade;
f - Feed sulphur grade;
d_{80}- 80% Particle passing size;
d_{50}- 50% Particle passing size.

The calculated values are given below:

$$\text{Volumetric concentrate rate} = C' - C + C/2.23$$

A value of 2.23 g/ml was used as the solids density. The density of the water was assumed to equal unity in all cases.

$$\text{Feed solids mass rate, } F = C + T$$

$$\text{Percent concentrate solids density (mass/mass)} = (C/C')\cdot 100$$

$$\text{Percent solids recovered} = (C/F)\cdot 100$$

$$\text{Water recovered} = C' - C$$

$$\text{Sulphur recovery} = (C\cdot c/(C\cdot c + T\cdot t))\cdot 100$$

To check that the values recorded and calculated were acceptable the calculated and measured feed sulphur grades for each test run were compared to each other. The feed sulphur grades were calculated as follows:

$$f = \frac{(c\cdot C + t\cdot T)}{F}$$

To calculate the maximum carrying capacity the bubble size was estimated from bubble sizing done using the spargers in a 2 phase system. The bubble sizing for laboratory test runs 12 and 13 were done during the test runs and are thus exact values.

$$\text{Bubble surface rate, } Js = \frac{6 \cdot \text{Volumetric Air Rate}}{(\text{Bubble size} \cdot 60)}$$

$$C_a = \frac{60 \cdot \mathbf{k} \cdot 3.14 \cdot d_{80} \cdot (1/10000) \cdot 2.23 \cdot Jg(\text{actual}) \cdot 22.9 \cdot 60}{(\text{Bubble size} \cdot 1000)}$$

The liquid and solids residence time was calculated as follows:

$$E_g = 1 - \frac{(P_t - P_c) \cdot 133.23}{9.81 \cdot H \cdot (T'/ \text{ Volumetric tails rate})}$$

$$\tau_l = 22.9 \cdot H \cdot (1-E_g)/(\text{Volumetric tails rate})$$

$$U_p = \frac{g \cdot dp^2 \cdot (1-E_g)^{2.7} \cdot (Ro_p - Ro_{susp})}{18 \cdot \mu \cdot (1+0.15 \cdot Re_p^{0.687})}$$

$$\tau_p = \frac{(\text{Volumetric tails rate}/1374) \cdot \tau_l}{(\text{Volumetric tails rate}/1374) + U_p}$$

τ_p for the froth zone was estimated by assuming that the particles collected in the concentrate have the same residence time as the air in the froth. Therefore:

$$\tau_{froth} = \frac{22.9 \cdot \text{Froth Depth} \cdot (\text{Air rate}/(\text{Air rate} + \text{Conc. rate}))}{\text{Air rate}}$$

It should be noted that the above equation will not give the exact solids residence time in the froth, but was only used to get a rough estimate.

APPENDIX D

LABORATORY DATA

The data recorded for the laboratory test work together with the calculated data is given in this appendix. The run number and the parameter varied or the type of test done is listed below:

Run No.	Variable	Page
1.	Steady State Determination	
2.	Steady State Determination	D2-D6
3.	Steady State (0 - 30 minutes)	
4.	Steady State (0 - 30 minutes)	
5.	Steady State (15 - 22 minutes)	
6.	Wash Water Temperature	
7.	Wash Water Temperature	
8.	Reproducibility	D7-D11
9.	Reproducibility	
10.	Volumetric Feed Rate	
11.	Superficial Air Rate, SGDS	
12.	Superficial Air Rate, FCS	
13.	Superficial Air Rate, USBM	
14.	Collection Zone Height	
15.	Froth Depth	D12-D16
16.	Feed Solids Density	
17.	Wash Water Rate	
18.	Wash Water Position	
19.	Feed Point	
20.	pH 6 and 3,6	D17-D21
21.	Collector Dosage	
22.	Frother Dosage	

APPENDIX D

	COLUMN DIMENSIONS				FEED CONDITIONS				
Run No.	Sparger Type	Coll. Height	Froth Height	Feed Pt From Top	Vol. Rate	Solids Rate	Solids Density	Grade Measured	Grade Calc.
1–4		cm	cm	cm	ml/min	g/min	%(m/m)	%S	%S
1.1	SGDS	200	100	110	946	117.6	11.6	2.04	2.04
1.2	SGDS	200	100	110	943	117.3	11.6	2.04	2.04
1.3	SGDS	200	100	110	943	117.4	11.7	2.04	2.04
1.4	SGDS	200	100	110	942	117.8	11.7	2.04	2.03
1.5	SGDS	200	100	110	941	117.6	11.7	2.04	2.04
1.6	SGDS	200	100	110	942	117.9	11.7	2.04	2.04
1.7	SGDS	200	100	110	941	118.0	11.7	2.04	2.03
1.8	SGDS	200	100	110	941	117.8	11.7	2.04	2.03
1.9	SGDS	200	100	110	941	117.9	11.7	2.04	2.04
1.10	SGDS	200	100	110	941	117.8	11.7	2.04	2.03
1.11	SGDS	200	100	110	942	118.2	11.7	2.04	2.03
2.1	SGDS	205	50	55	1121	117.9	9.9	2.02	2.11
2.2	SGDS	205	50	55	1162	118.1	9.6	2.02	1.97
2.3	SGDS	205	50	55	1105	112.5	9.6	2.02	2.28
2.4	SGDS	205	50	55	1099	112.6	9.7	2.02	2.10
2.5	SGDS	205	50	55	1097	111.2	9.6	2.02	2.11
2.6	SGDS	205	50	55	1011	108.6	10.1	2.02	2.47
2.7	SGDS	205	50	55	1240	122.5	9.4	2.02	1.96
2.8	SGDS	205	50	55	1176	117.5	9.5	2.02	2.08
2.9	SGDS	205	50	55	1035	106.6	9.7	2.02	2.18
2.10	SGDS	205	50	55	1223	123.2	9.5	2.02	2.33
2.11	SGDS	205	50	55	1171	118.0	9.5	2.02	2.29
2.12	SGDS	205	50	55	1114	115.1	9.8	2.02	2.27
2.13	SGDS	205	50	55	1193	119.0	9.5	2.02	2.08
2.14	SGDS	205	50	55	1065	112.1	10.0	2.02	2.12
2.15	SGDS	205	50	55	1150	116.5	9.6	2.02	2.07
2.16	SGDS	205	50	55	1154	115.9	9.5	2.02	2.18
2.17	SGDS	205	50	55	1147	114.4	9.5	2.02	2.19
3.1	SGDS	300	100	110	952	66.5	6.7	1.42	3.03
3.2	SGDS	300	100	110	1023	105.1	9.7	1.42	1.84
3.3	SGDS	300	100	110	1026	104.3	9.6	1.42	1.63
3.4	SGDS	300	100	110	977	96.7	9.4	1.42	1.84
3.5	SGDS	300	100	110	1019	103.3	9.6	1.42	2.01
3.6	SGDS	300	100	110	1025	103.4	9.6	1.42	1.52
4.1	SGDS	300	100	110	1086	85.4	7.5	1.83	2.05
4.2	SGDS	300	100	110	1050	103.5	9.3	1.83	1.85
4.3	SGDS	300	100	110	1041	105.0	9.6	1.83	1.92
4.4	SGDS	300	100	110	1096	108.6	9.4	1.83	1.87
4.5	SGDS	300	100	110	1073	108.0	9.5	1.83	1.94
4.6	SGDS	300	100	110	1032	104.9	9.6	1.83	1.91

APPENDIX D

Run No.	REAGENT ADDITIONS						
	Collector one	Coll. one Dosage	Collector two	Coll. two Dosage	Cond. Time	Frother Type	Total Frother
1–4	*name*	*g/ton*	*name*	*g/ton*	*min*	*Name*	*ppm (vol)*
1.1	Senkol 50	25	–	–	20	DF 200	22.8
1.2	Senkol 50	25	–	–	20	DF 200	22.8
1.3	Senkol 50	25	–	–	20	DF 200	22.8
1.4	Senkol 50	25	–	–	20	DF 200	22.9
1.5	Senkol 50	25	–	–	20	DF 200	22.9
1.6	Senkol 50	25	–	–	20	DF 200	22.9
1.7	Senkol 50	25	–	–	20	DF 200	23.0
1.8	Senkol 50	25	–	–	20	DF 200	23.0
1.9	Senkol 50	25	–	–	20	DF 200	23.0
1.10	Senkol 50	25	–	–	20	DF 200	23.0
1.11	Senkol 50	25	–	–	20	DF 200	23.0
2.1	SIBX	17	AFLT 208	32	20	AF 65	23.3
2.2	SIBX	17	AFLT 208	32	20	AF 65	22.6
2.3	SIBX	17	AFLT 208	32	20	AF 65	22.6
2.4	SIBX	17	AFLT 208	32	20	AF 65	22.7
2.5	SIBX	17	AFLT 208	32	20	AF 65	22.5
2.6	SIBX	17	AFLT 208	32	20	AF 65	23.8
2.7	SIBX	17	AFLT 208	32	20	AF 65	21.9
2.8	SIBX	17	AFLT 208	32	20	AF 65	22.2
2.9	SIBX	17	AFLT 208	32	20	AF 65	22.9
2.10	SIBX	17	AFLT 208	32	20	AF 65	22.3
2.11	SIBX	17	AFLT 208	32	20	AF 65	22.4
2.12	SIBX	17	AFLT 208	32	20	AF 65	22.9
2.13	SIBX	17	AFLT 208	32	20	AF 65	22.1
2.14	SIBX	17	AFLT 208	32	20	AF 65	23.4
2.15	SIBX	17	AFLT 208	32	20	AF 65	22.5
2.16	SIBX	17	AFLT 208	32	20	AF 65	22.3
2.17	SIBX	17	AFLT 208	32	20	AF 65	22.1
3.1	Senkol 50	90	–	–	20	DF 200	12.8
3.2	Senkol 50	90	–	–	20	DF 200	18.8
3.3	Senkol 50	90	–	–	20	DF 200	18.6
3.4	Senkol 50	90	–	–	20	DF 200	18.2
3.5	Senkol 50	90	–	–	20	DF 200	18.6
3.6	Senkol 50	90	–	–	20	DF 200	18.5
4.1	Senkol 50	90	–	–	20	DF 200	14.4
4.2	Senkol 50	90	–	–	20	DF 200	18.1
4.3	Senkol 50	90	–	–	20	DF 200	18.5
4.4	Senkol 50	90	–	–	20	DF 200	18.2
4.5	Senkol 50	90	–	–	20	DF 200	18.5
4.6	Senkol 50	90	–	–	20	DF 200	18.6

APPENDIX D

Run No.	WASH WATER				AIR					
	Rate	Frother Addition	Temp	Position from Top	Vol. Rate	Press Top	Press Bott	Jg* STP	Jg Actual	Sup. Bias
1–4	*ml/min*	*ppm*	*Deg C*	*cm*	*ml/min*	*mm Hg*	*mm Hg*	*cm/sec*	*cm/sec*	*cm/sec*
1.1	140	20	21	5	1630	2.0	13.3	1.19	0.84	0.08
1.2	140	20	21	5	1630	2.0	13.3	1.19	0.84	0.09
1.3	140	20	21	5	1630	2.0	13.3	1.19	0.84	0.09
1.4	140	20	21	5	1630	2.0	13.3	1.19	0.84	0.09
1.5	140	20	21	5	1630	2.0	13.3	1.19	0.84	0.09
1.6	140	20	21	5	1630	2.0	13.3	1.19	0.84	0.09
1.7	140	20	21	5	1630	2.0	13.3	1.19	0.84	0.09
1.8	140	20	21	5	1630	2.0	13.3	1.19	0.84	0.09
1.9	140	20	21	5	1630	2.0	13.3	1.19	0.84	0.09
1.10	140	20	21	5	1630	2.0	13.3	1.19	0.84	0.09
1.11	140	20	21	5	1630	2.0	13.3	1.19	0.84	0.09
2.1	200	20	21	5	4250	0.8	13.2	3.09	2.21	0.10
2.2	200	20	21	5	4250	0.8	13.2	3.09	2.21	0.11
2.3	200	20	21	5	4250	0.8	13.2	3.09	2.21	0.11
2.4	200	20	21	5	4250	0.8	13.2	3.09	2.21	0.10
2.5	200	20	21	5	4250	0.8	13.2	3.09	2.21	0.10
2.6	200	20	21	5	4250	0.8	13.2	3.09	2.21	0.10
2.7	200	20	21	5	4250	0.8	13.2	3.09	2.21	0.11
2.8	200	20	21	5	4250	0.8	13.2	3.09	2.21	0.11
2.9	200	20	21	5	4250	0.8	13.2	3.09	2.21	0.12
2.10	200	20	21	5	4250	0.8	13.2	3.09	2.21	0.11
2.11	200	20	21	5	4250	0.8	13.2	3.09	2.21	0.11
2.12	200	20	21	5	4250	0.8	13.2	3.09	2.21	0.12
2.13	200	20	21	5	4250	0.8	13.2	3.09	2.21	0.12
2.14	200	20	21	5	4250	0.8	13.2	3.09	2.21	0.12
2.15	200	20	21	5	4250	0.8	13.2	3.09	2.21	0.12
2.16	200	20	21	5	4250	0.8	13.2	3.09	2.21	0.12
2.17	200	20	21	5	4250	0.8	13.2	3.09	2.21	0.12
3.1	140	20	21	5	1630	2.0	13.3	1.19	0.84	-0.02
3.2	140	20	21	5	1630	2.0	13.3	1.19	0.84	0.08
3.3	140	20	21	5	1630	2.0	13.3	1.19	0.84	0.07
3.4	140	20	21	5	1630	2.0	13.3	1.19	0.84	0.08
3.5	140	20	21	5	1630	2.0	13.3	1.19	0.84	0.07
3.6	140	20	21	5	1630	2.0	13.3	1.19	0.84	0.08
4.1	140	20	21	5	1630	2.0	13.3	1.19	0.84	-0.04
4.2	140	20	21	5	1630	2.0	13.3	1.19	0.84	-0.02
4.3	140	20	21	5	1630	2.0	13.3	1.19	0.84	0.01
4.4	140	20	21	5	1630	2.0	13.3	1.19	0.84	0.02
4.5	140	20	21	5	1630	2.0	13.3	1.19	0.84	0.02
4.6	140	20	21	5	1630	2.0	13.3	1.19	0.84	0.02

APPENDIX D

Run No.	CONCENTRATE							TAILINGS		
	Vol. Rate	Mass Rate	Solids Density	% Sol. Rec.	Water Rec	S Grade	S Rec.	Vol. Rate	Mass Rate	S Grade
1-4	ml/min	g/min	%(m/m)	%	g/min	%S	%S	ml/min	g/min	%S
1.1	26	7.5	25	6	22.8	23.5	73	1060	110.1	0.58
1.2	22	7.2	27	6	19.0	24.4	73	1060	110.1	0.58
1.3	22	7.3	28	6	19.0	24.4	74	1060	110.1	0.56
1.4	21	7.7	30	7	17.8	24.4	78	1060	110.1	0.47
1.5	20	7.5	31	6	17.0	24.0	75	1060	110.1	0.54
1.6	22	7.8	30	7	18.2	23.6	77	1060	110.1	0.51
1.7	21	7.9	31	7	17.2	25.5	84	1060	110.1	0.35
1.8	20	7.7	31	7	16.9	23.4	75	1060	110.1	0.54
1.9	21	7.8	31	7	17.7	24.6	80	1060	110.1	0.44
1.10	21	7.7	31	7	17.1	24.1	77	1060	110.1	0.49
1.11	21	8.1	31	7	17.7	23.9	81	1060	110.1	0.42
2.1	59	8.6	13	7	55.3	22.5	78	1262	109.3	0.51
2.2	54	8.0	14	7	50.0	21.9	75	1308	110.1	0.52
2.3	55	8.4	14	7	51.2	22.9	75	1250	104.0	0.61
2.4	56	8.6	14	8	52.3	21.4	78	1243	104.0	0.51
2.5	56	8.5	14	8	52.5	21.6	78	1240	102.7	0.50
2.6	58	9.6	15	9	53.4	22.4	80	1153	99.0	0.54
2.7	55	8.4	14	7	50.8	21.5	75	1385	114.1	0.53
2.8	55	8.6	15	7	50.9	21.8	77	1321	108.9	0.52
2.9	42	8.0	17	8	38.1	21.9	75	1194	98.6	0.58
2.10	42	9.2	19	7	38.1	24.3	78	1381	114.0	0.56
2.11	44	9.2	19	8	40.1	22.5	77	1327	108.7	0.57
2.12	40	9.3	21	8	35.4	22.0	78	1275	105.8	0.54
2.13	35	8.6	22	7	31.2	21.5	75	1358	110.4	0.57
2.14	35	8.7	22	8	31.0	21.1	77	1230	103.4	0.52
2.15	35	8.8	22	8	31.0	21.1	77	1315	107.7	0.52
2.16	35	8.9	22	8	31.2	22.0	78	1319	106.9	0.52
2.17	36	8.7	21	8	32.4	22.4	78	1311	105.7	0.52
3.1	163	11.0	6	16	157.7	17.0	92	929	55.6	0.28
3.2	37	6.8	17	6	33.5	22.0	78	1127	98.3	0.44
3.3	41	6.6	15	6	38.1	19.8	76	1125	97.7	0.41
3.4	34	6.7	18	7	30.6	20.4	77	1083	90.0	0.45
3.5	46	7.9	16	8	42.0	20.6	78	1114	95.4	0.47
3.6	32	5.9	17	6	29.2	19.8	75	1133	97.5	0.41
4.1	200	9.8	5	12	195.8	15.4	87	1025	75.6	0.31
4.2	173	9.5	5	9	168.5	16.0	80	1017	94.0	0.41
4.3	122	8.7	7	8	118.4	19.2	83	1059	96.3	0.36
4.4	114	8.8	7	8	110.0	18.9	82	1122	99.8	0.37
4.5	113	8.8	7	8	109.0	19.6	82	1100	99.2	0.37
4.6	106	8.8	8	8	102.4	19.4	85	1065	96.1	0.32

	CARRYING CAPACITY					SOLIDS AND LIQUID RESIDENCE TIME				
Run No.	Bubble Size	Js	d80	'Ca'	Carry. Const.	Eg	d50	TAU l	TAU p	TAU p Froth
1–4	cm	cm/sec	microns	g/min	K	(frac)	microns	min	min	min
1.1	0.20	807	–	–	–	0.27	–	3.1	–	1.38
1.2	0.20	807	–	–	–	0.27	–	3.1	–	1.39
1.3	0.20	807	–	–	–	0.27	–	3.1	–	1.39
1.4	0.20	807	–	–	–	0.27	–	3.1	–	1.39
1.5	0.20	807	–	–	–	0.27	–	3.1	–	1.39
1.6	0.20	807	–	–	–	0.27	–	3.1	–	1.39
1.7	0.20	807	–	–	–	0.27	–	3.1	–	1.39
1.8	0.20	807	–	–	–	0.27	–	3.1	–	1.39
1.9	0.20	807	–	–	–	0.27	–	3.1	–	1.39
1.10	0.20	807	–	–	–	0.27	–	3.1	–	1.39
1.11	0.20	807	–	–	–	0.27	–	3.1	–	1.39
2.1	0.20	2125	56.5	8.6	0.24	0.22	33.1	2.9	2.9	0.27
2.2	0.20	2125	56.5	8.0	0.22	0.21	33.1	2.8	2.8	0.27
2.3	0.20	2125	56.5	8.4	0.23	0.21	33.1	3.0	2.9	0.27
2.4	0.20	2125	56.5	8.6	0.24	0.21	33.1	3.0	3.0	0.27
2.5	0.20	2125	56.5	8.5	0.24	0.21	33.1	3.0	3.0	0.27
2.6	0.20	2125	56.5	9.6	0.27	0.22	33.1	3.2	3.2	0.27
2.7	0.20	2125	56.5	8.4	0.23	0.21	33.1	2.7	2.7	0.27
2.8	0.20	2125	56.5	8.6	0.24	0.21	33.1	2.8	2.8	0.27
2.9	0.20	2125	56.5	8.0	0.22	0.21	33.1	3.1	3.1	0.27
2.10	0.20	2125	56.5	9.2	0.26	0.21	33.1	2.7	2.7	0.27
2.11	0.20	2125	56.5	9.2	0.26	0.21	33.1	2.8	2.8	0.27
2.12	0.20	2125	56.5	9.3	0.26	0.21	33.1	2.9	2.9	0.27
2.13	0.20	2125	56.5	8.6	0.24	0.21	33.1	2.7	2.7	0.27
2.14	0.20	2125	56.5	8.7	0.24	0.21	33.1	3.0	3.0	0.27
2.15	0.20	2125	56.5	8.8	0.24	0.21	33.1	2.8	2.8	0.27
2.16	0.20	2125	56.5	8.9	0.25	0.21	33.1	2.8	2.8	0.27
2.17	0.20	2125	56.5	8.7	0.24	0.21	33.1	2.8	2.8	0.27
3.1	0.20	815	–	–	–	0.50	–	3.7	–	1.28
3.2	0.20	815	–	–	–	0.51	–	3.0	–	1.37
3.3	0.20	815	–	–	–	0.51	–	3.0	–	1.37
3.4	0.20	815	–	–	–	0.51	–	3.1	–	1.38
3.5	0.20	815	–	–	–	0.51	–	3.0	–	1.37
3.6	0.20	815	–	–	–	0.51	–	3.0	–	1.38
4.1	0.20	815	–	–	–	0.51	–	3.3	–	1.25
4.2	0.20	815	–	–	–	0.51	–	3.3	–	1.27
4.3	0.20	815	–	–	–	0.51	–	3.2	–	1.31
4.4	0.20	815	–	–	–	0.51	–	3.0	–	1.31
4.5	0.20	815	–	–	–	0.51	–	3.0	–	1.31
4.6	0.20	815	–	–	–	0.51	–	3.1	–	1.32

Run No.	COLUMN DIMENSIONS				FEED CONDITIONS				
	Sparger Type	Coll. Height	Froth Height	Feed Pt From Top	Vol. Rate	Solids Rate	Solids Density	Grade Measured	Grade Calc.
5-11		cm	cm	cm	ml/min	g/min	%(m/m)	%S	%S
5.1	SGDS	200	100	110	754	81.1	10.2	2.3	1.65
5.2	SGDS	200	100	110	758	81.1	10.1	2.3	1.72
5.3	SGDS	200	100	110	753	81.1	10.2	2.3	1.53
5.4	SGDS	200	100	110	754	81.1	10.2	2.3	1.68
5.5	SGDS	200	100	110	758	81.1	10.1	2.3	1.90
5.6	SGDS	200	100	110	762	81.1	10.0	2.3	2.06
5.7	SGDS	200	100	110	757	81.1	10.1	2.3	1.85
6.1	SGDS	200	50	55	1040	110.9	10.1	2.05	1.91
6.2	SGDS	200	50	55	1079	115.9	10.1	2.05	1.97
6.3	SGDS	200	50	55	1099	119.0	10.2	2.05	1.84
6.4	SGDS	200	50	55	1054	117.7	10.5	2.05	1.85
7.1	SGDS	200	50	55	1026	117.5	10.8	2.05	1.93
7.2	SGDS	200	50	55	1089	117.4	10.2	2.05	2.00
7.3	SGDS	200	50	55	1074	113.0	9.9	2.05	1.84
7.4	SGDS	200	50	55	1091	115.5	10.0	2.05	2.01
8.1	FCS	200	50	55	965	96.3	9.5	1.73	2.08
8.2	FCS	200	50	55	960	99.0	9.8	1.73	2.06
8.3	FCS	200	50	55	915	93.5	9.7	1.73	2.27
8.4	FCS	200	50	55	919	91.2	9.4	1.73	2.27
9.1	FCS	200	50	55	981	83.7	8.1	1.80	2.08
9.2	FCS	200	50	55	984	91.1	8.8	1.80	2.06
9.3	FCS	200	50	55	931	85.4	8.7	1.80	2.06
9.4	FCS	200	50	55	953	84.4	8.4	1.80	2.10
10.1	SGDS	200	100	110	841	90.8	10.2	2.30	1.70
10.2	SGDS	200	100	110	1016	108.7	10.1	2.30	2.18
10.3	SGDS	200	100	110	1292	140.3	10.2	2.30	2.21
10.4	SGDS	200	100	110	1578	170.6	10.2	2.30	2.22
10.5	SGDS	200	100	110	1726	186.4	10.2	2.30	2.26
10.6	SGDS	200	100	110	1958	210.0	10.1	2.30	2.26
11.1	SGDS	200	50	55	1119	117.6	9.9	2.09	1.90
11.2	SGDS	200	50	55	1135	119.3	9.9	2.09	1.90
11.3	SGDS	200	50	55	1101	116.8	10.0	2.09	1.82
11.4	SGDS	200	50	55	1084	115.8	10.1	2.09	1.94
11.5	SGDS	200	50	55	1080	117.3	10.2	2.09	2.11
11.6	SGDS	200	50	55			Slugging		

	REAGENT ADDITIONS						
Run No.	Collector one	Coll. one Dosage	Collector two	Coll. two Dosage	Cond. Time	Frother Type	Total Frother
5–11	*name*	*g/ton*	*name*	*g/ton*	*min*	*Name*	*ppm (Vol)*
5.1	Senkol 50	100	–	–	20	DF 200	19.7
5.2	Senkol 50	100	–	–	20	DF 200	19.6
5.3	Senkol 50	100	–	–	20	DF 200	19.7
5.4	Senkol 50	100	–	–	20	DF 200	19.7
5.5	Senkol 50	100	–	–	20	DF 200	19.6
5.6	Senkol 50	100	–	–	20	DF 200	19.5
5.7	Senkol 50	100	–	–	20	DF 200	19.6
6.1	Senkol 50	80	–	–	20	DF 200	12.7
6.2	Senkol 50	80	–	–	20	DF 200	12.8
6.3	Senkol 50	80	–	–	20	DF 200	12.9
6.4	Senkol 50	80	–	–	20	DF 200	13.3
7.1	Senkol 50	80	–	–	20	DF 200	13.6
7.2	Senkol 50	80	–	–	20	DF 200	12.8
7.3	Senkol 50	80	–	–	20	DF 200	12.5
7.4	Senkol 50	80	–	–	20	DF 200	12.6
8.1	SIBX	20	AFLT 208	20	20	AF 65	23.3
8.2	SIBX	19	AFLT 208	19	20	AF 65	23.3
8.3	SIBX	21	AFLT 208	21	20	AF 65	23.6
8.4	SIBX	21	AFLT 208	21	20	AF 65	23.6
9.1	SIBX	26	AFLT 208	26	20	AF 65	23.2
9.2	SIBX	24	AFLT 208	24	20	AF 65	23.2
9.3	SIBX	25	AFLT 208	25	20	AF 65	23.5
9.4	SIBX	26	AFLT 208	26	20	AF 65	23.3
10.1	Senkol 50	100	–	–	20	DF 200	20.4
10.2	Senkol 50	100	–	–	20	DF 200	22.1
10.3	Senkol 50	100	–	–	20	DF 200	21.0
10.4	Senkol 50	100	–	–	20	DF 200	20.7
10.5	Senkol 50	100	–	–	20	DF 200	21.5
10.6	Senkol 50	100	–	–	20	DF 200	20.8
11.1	Senkol 50	90	–	–	20	DF 200	21.7
11.2	Senkol 50	90	–	–	20	DF 200	21.6
11.3	Senkol 50	90	–	–	20	DF 200	21.8
11.4	Senkol 50	90	–	–	20	DF 200	21.7
11.5	Senkol 50	90	–	–	20	DF 200	21.8
11.6	Senkol 50	90	–	–	20	DF 200	Slugging

Run No.	WASH WATER				AIR					
	Rate	Frother Addition	Temp	Position from Top	Vol. Rate	Press Top	Press Bott	Jg* STP	Jg Actual	Sup. Bias
5–11	ml/min	ppm	Deg C	cm	ml/min	mm Hg	mm Hg	cm/sec	cm/sec	cm/sec
5.1	140	20	21	5	1710	1.6	13.3	1.24	0.89	0.09
5.2	140	20	21	5	1710	1.6	13.3	1.24	0.89	0.09
5.3	140	20	21	5	1710	1.6	13.3	1.24	0.89	0.09
5.4	140	20	21	5	1710	1.6	13.3	1.24	0.89	0.09
5.5	140	20	21	5	1710	1.6	13.3	1.24	0.89	0.09
5.6	140	20	21	5	1710	1.6	13.3	1.24	0.89	0.09
5.7	140	20	21	5	1710	1.6	13.3	1.24	0.89	0.09
6.1	140	20	21	5	2670	1.3	13.3	1.94	1.38	0.08
6.2	140	20	30	5	2670	1.3	13.3	1.94	1.38	0.08
6.3	140	20	40	5	2670	1.3	13.3	1.94	1.38	0.08
6.4	140	20	50	5	2670	1.3	13.3	1.94	1.38	0.09
7.1	140	20	21	5	2130	1.3	13.3	1.55	1.10	0.08
7.2	140	20	30	5	2130	1.3	13.3	1.55	1.10	0.08
7.3	140	20	40	5	2130	1.3	13.3	1.55	1.10	0.08
7.4	140	20	50	5	2130	1.3	13.3	1.55	1.10	0.09
8.1	200	20	21	5	3100	1.3	13.5	2.26	1.60	0.08
8.2	200	20	21	5	3100	1.3	13.5	2.26	1.60	0.08
8.3	200	20	21	5	3100	1.3	13.5	2.26	1.60	0.08
8.4	200	20	21	5	3100	1.3	13.5	2.26	1.60	0.08
9.1	200	20	21	5	3100	1.3	13.4	2.26	1.60	0.08
9.2	200	20	21	5	3100	1.3	13.4	2.26	1.60	0.08
9.3	200	20	21	5	3100	1.3	13.4	2.26	1.60	0.08
9.4	200	20	21	5	3100	1.3	13.4	2.26	1.60	0.07
10.1	140	20	21	5	1710	1.6	13.3	1.24	0.89	0.09
10.2	140	20	21	5	1710	1.6	13.3	1.24	0.89	0.08
10.3	140	20	21	5	1710	1.6	13.3	1.24	0.89	0.08
10.4	140	20	21	5	1710	1.6	13.3	1.24	0.89	0.08
10.5	140	20	21	5	1710	1.6	13.3	1.24	0.89	0.07
10.6	140	20	21	5	1710	1.6	13.3	1.24	0.89	0.07
11.1	140	20	21	5	1530	1.3	8.8	1.11	0.87	0.09
11.2	140	20	21	5	1940	1.3	8.6	1.41	1.11	0.09
11.3	140	20	21	5	2750	1.5	8.4	2.00	1.58	0.09
11.4	140	20	21	5	3740	1.5	7.9	2.72	2.18	0.08
11.5	140	20	21	5	4700	1.2	7.5	3.42	2.76	0.05
11.6	140	20	21	5	1630			Slugging		

Run No.	CONCENTRATE							TAILINGS		
	Vol. Rate	Mass Rate	Solids Density	% Sol. Rec.	Water Rec	S Grade	S Rec.	Vol. Rate	Mass Rate	S Grade
5–11	*ml/min*	*g/min*	*%(m/m)*	*%*	*g/min*	*%S*	*%S*	*ml/min*	*g/min*	*%S*
5.1	15	4.2	24	5	13.1	23.3	72	879	76.9	0.48
5.2	19	4.6	21	6	17.2	22.6	74	879	76.5	0.48
5.3	15	3.9	23	5	13.1	23.0	72	878	77.2	0.45
5.4	15	4.3	24	5	13.3	23.4	73	879	76.8	0.48
5.5	19	5.1	23	6	16.8	22.9	76	879	75.9	0.48
5.6	23	5.8	22	7	20.2	22.5	78	879	75.3	0.49
5.7	18	4.9	24	6	15.4	23.2	76	879	76.1	0.47
6.1	29	9.2	27	8	24.7	18.5	80	1151	101.7	0.41
6.2	31	9.1	25	8	27.1	20.2	80	1188	106.8	0.42
6.3	24	8.1	28	7	20.8	21.1	78	1215	110.9	0.43
6.4	21	7.4	29	6	18.0	22.6	77	1173	110.3	0.46
7.1	28	9.5	28	8	23.9	20.0	83	1138	108.1	0.35
7.2	27	9.4	29	8	23.2	21.0	84	1202	108.0	0.35
7.3	24	8.5	30	7	19.8	20.3	82	1190	104.5	0.35
7.4	18	8.1	36	7	14.6	23.6	82	1212	107.4	0.38
8.1	84	7.1	8	7	80.5	21.9	77	1082	89.2	0.51
8.2	86	7.4	8	7	82.6	21.9	79	1074	91.6	0.47
8.3	84	7.5	8	8	80.8	23.0	81	1031	86.0	0.47
8.4	89	7.6	8	8	85.5	21.8	81	1030	83.6	0.48
9.1	93	6.3	7	8	89.9	21.0	76	1088	77.4	0.53
9.2	89	6.7	7	7	85.8	21.6	77	1095	84.4	0.51
9.3	97	6.4	6	7	93.7	21.8	79	1034	79.0	0.47
9.4	106	6.3	6	7	102.8	21.5	77	1048	78.1	0.53
10.1	18	4.9	24	5	15.4	24.0	77	964	85.9	0.42
10.2	26	7.6	25	7	22.8	24.5	79	1130	101.1	0.49
10.3	33	9.7	25	7	28.7	25.4	79	1399	130.6	0.49
10.4	33	10.4	27	6	28.1	29.0	80	1685	160.2	0.48
10.5	40	12.3	27	7	34.0	27.4	80	1827	174.1	0.48
10.6	43	14.0	28	7	36.5	27.2	80	2055	196.0	0.48
11.1	14	5.9	35	5	11.0	27.1	71	1245	111.7	0.58
11.2	15	6.2	34	5	12.2	26.5	72	1260	113.1	0.56
11.3	18	6.4	30	5	14.9	24.0	72	1223	110.4	0.54
11.4	35	7.4	19	6	32.0	22.6	74	1189	108.4	0.54
11.5	77	8.6	11	7	73.2	22.3	78	1143	108.7	0.51
11.6					Slugging					

Run No.	CARRYING CAPACITY					SOLIDS AND LIQUID RESIDENCE TIME				
	Bubble Size	Js	d80	'Ca'	Carry. Const.	Eg	d50	TAU l	TAU p	TAU p Froth
5–11	*cm*	cm/sec	*microns*	*g/min*	*K*	*(frac)*	*microns*	*min*	*min*	*min*
5.1	–	–	–	–	–	0.24	–	4.0	–	1.33
5.2	–	–	–	–	–	0.24	–	4.0	–	1.32
5.3	–	–	–	–	–	0.24	–	4.0	–	1.33
5.4	–	–	–	–	–	0.24	–	4.0	–	1.33
5.5	–	–	–	–	–	0.24	–	4.0	–	1.32
5.6	–	–	–	–	–	0.24	–	4.0	–	1.32
5.7	–	–	–	–	–	0.24	–	4.0	–	1.33
6.1	0.15	1780	43.3	9.2	0.4	0.22	20.9	3.1	3.1	0.42
6.2	0.15	1780	42.6	9.1	0.4	0.22	20.9	3.0	3.0	0.42
6.3	0.15	1780	39.0	8.1	0.4	0.22	19.9	2.9	2.9	0.42
6.4	0.15	1780	36.5	7.4	0.4	0.22	19.7	3.0	3.0	0.43
7.1	–	–	–	–	–	0.23	–	3.1	–	0.53
7.2	–	–	–	–	–	0.22	–	3.0	–	0.53
7.3	–	–	–	–	–	0.22	–	3.0	–	0.53
7.4	–	–	–	–	–	0.22	–	2.9	–	0.53
8.1	–	–	–	–	–	0.21	–	3.4	–	0.36
8.2	–	–	–	–	–	0.21	–	3.4	–	0.36
8.3	–	–	–	–	–	0.21	–	3.5	–	0.36
8.4	–	–	–	–	–	0.21	–	3.5	–	0.36
9.1	–	–	–	–	–	0.21	–	3.3	–	0.36
9.2	–	–	–	–	–	0.21	–	3.3	–	0.36
9.3	–	–	–	–	–	0.21	–	3.5	–	0.36
9.4	–	–	–	–	–	0.21	–	3.5	–	0.36
10.1	0.15	1140	38.0	4.9	0.4	0.24	19.0	3.6	3.6	1.33
10.2	0.15	1140	42.6	7.6	0.5	0.24	20.3	3.1	3.1	1.32
10.3	0.15	1140	48.4	9.7	0.6	0.24	22.1	2.5	2.5	1.31
10.4	0.15	1140	48.4	10.4	0.6	0.24	24.1	2.1	2.1	1.31
10.5	0.15	1140	52.3	12.3	0.7	0.24	25.5	1.9	1.9	1.31
10.6	0.15	1140	30.9	14.0	1.3	0.24	16.1	1.7	1.7	1.31
11.1	0.15	1020	31.5	5.9	0.6	0.51	16.6	1.8	1.8	0.74
11.2	0.15	1293	52.7	6.2	0.3	0.53	25.7	1.7	1.7	0.59
11.3	0.15	1833	43.8	6.4	0.2	0.55	22.2	1.7	1.7	0.41
11.4	0.15	2493	46.1	7.4	0.2	0.59	24.2	1.6	1.6	0.30
11.5	0.15	3133	33.8	8.6	0.2	0.59	16.9	1.6	1.6	0.24
11.6	Slugging									

	COLUMN DIMENSIONS				FEED CONDITIONS				
Run No.	Sparger Type	Coll. Height	Froth Height	Feed Pt From Top	Vol. Rate	Solids Rate	Solids Density	Grade Measured	Grade Calc.
12–18		cm	cm	cm	ml/min	g/min	%(m/m)	%S	%S
12.1	FCS	225	25	55	944	158.5	15.4	2.02	1.93
12.2	FCS	225	25	55	943	162.7	15.7	2.02	2.33
12.3	FCS	225	25	55	930	159.0	15.6	2.02	2.37
12.4	FCS	225	25	55	964	166.7	15.8	2.02	2.29
12.5	FCS	225	25	55	942	160.4	15.6	2.02	2.37
12.6	FCS	225	25	55	939	164.2	16.0	2.02	2.39
13.1	USBM	225	25	55	905	148.7	15.1	2.14	2.15
13.2	USBM	225	25	55	828	141.4	15.6	2.14	1.99
13.3	USBM	225	25	55	799	136.4	15.6	2.14	2.22
13.4	USBM	225	25	55	1068	163.9	14.2	2.14	2.29
14.1	FCS	630	25	55	977	183.0	17.0	2.06	2.55
14.2	FCS	530	25	55	917	169.8	16.8	2.06	2.39
14.3	FCS	430	25	55	952	180.5	17.2	2.06	2.39
14.4	FCS	330	25	55	995	180.4	16.5	2.06	2.41
14.5	FCS	230	25	55	979	171.6	16.0	2.06	2.37
14.6	FCS	130	25	55	971	171.7	16.1	2.06	2.32
15.1	SGDS	200	100	105	1596	163.4	9.7	1.92	1.47
15.2	SGDS	200	75	80	1577	164.7	9.9	1.92	1.83
15.3	SGDS	200	50	55	1564	167.2	10.1	1.92	2.15
15.4	SGDS	225	25	55	1580	170.8	10.2	1.92	2.22
16.1	FCS	225	25	55	1058	97.1	8.7	2.00	1.94
16.2	FCS	225	25	55	1013	176.6	15.9	2.00	2.00
16.3	FCS	225	25	55	910	267.7	25.3	2.00	2.12
16.4	FCS	225	25	55	858	417.3	38.4	2.00	1.84
17.1	SGDS	200	50	55	1044	119.2	10.7	2.13	2.13
17.2	SGDS	200	50	55	1043	116.1	10.5	2.13	2.13
17.3	SGDS	200	50	55	1042	116.6	10.5	2.13	2.13
17.4	SGDS	200	50	55	1047	124.9	11.2	2.13	2.13
18.1	FCS	225	25	55	1011	177.0	16.0	1.85	1.99
18.2	FCS	225	25	55	1015	176.0	15.8	1.85	2.16
18.3	FCS	225	25	55	1023	175.0	15.6	1.85	2.07
18.4	FCS	225	25	55	1063	174.3	15.0	1.85	1.93

Run No.	REAGENT ADDITIONS						
	Collector one	Coll. one Dosage	Collector two	Coll. two Dosage	Cond. Time	Frother Type	Total Frother
12–18	*name*	*g/ton*	*name*	*g/ton*	*min*	*Name*	*ppm (Vol)*
12.1	SIPX	27	AFLT 208	45	30	AF 65	6.8
12.2	SIPX	26	AFLT 208	43	30	AF 65	6.8
12.3	SIPX	27	AFLT 208	44	30	AF 65	6.9
12.4	SIPX	25	AFLT 208	42	30	AF 65	6.7
12.5	SIPX	26	AFLT 208	44	30	AF 65	6.9
12.6	SIPX	26	AFLT 208	43	30	AF 65	6.9
13.1	SIPX	28	AFLT 208	47	30	DF 200	17.0
13.2	SIPX	30	AFLT 208	50	30	DF 200	17.5
13.3	SIPX	31	AFLT 208	52	30	DF 200	16.6
13.4	SIPX	26	AFLT 208	43	30	DF 200	11.0
14.1	SIPX	18	AFLT 208	30	10+10	DF 200	23.3
14.2	SIPX	18	AFLT 208	30	10+10	DF 200	23.6
14.3	SIPX	18	AFLT 208	30	10+10	DF 200	23.2
14.4	SIPX	18	AFLT 208	30	10+10	DF 200	23.1
14.5	SIPX	18	AFLT 208	30	10+10	DF 200	23.2
14.6	SIPX	18	AFLT 208	30	10+10	DF 200	23.2
15.1	Senkol 50	100	–	–	20	DF 200	20.9
15.2	Senkol 50	100	–	–	20	DF 200	21.0
15.3	Senkol 50	100	–	–	20	DF 200	21.0
15.4	Senkol 50	100	–	–	20	DF 200	21.0
16.1	SIPX	34	–	–	20	AF 65	22.9
16.2	SIPX	34	–	–	20	AF 65	23.2
16.3	SIPX	34	–	–	20	AF 65	23.6
16.4	SIPX	34	–	–	20	AF 65	23.9
17.1	Senkol 50	90	–	–	20	DF 200	21.5
17.2	Senkol 50	90	–	–	20	DF 200	21.9
17.3	Senkol 50	90	–	–	20	DF 200	23.2
17.4	Senkol 50	90	–	–	20	DF 200	25.9
18.1	SIPX	23	AFLT 208	38	25	AF 65	23.0
18.2	SIPX	23	AFLT 208	38	25	AF 65	22.9
18.3	SIPX	23	AFLT 208	38	25	AF 65	22.6
18.4	SIPX	23	AFLT 208	38	25	AF 65	21.8

	WASH WATER				AIR					
Run No.	Rate	Frother Addition	Temp	Position from Top	Vol. Rate	Press Top	Press Bott	Jg* STP	Jg Actual	Sup. Bias
12–18	*ml/min*	*ppm*	*Deg C*	*cm*	*ml/min*	*mm Hg*	*mm Hg*	*cm/sec*	*cm/sec*	*cm/sec*
12.1	173	20	21	5	1800	0.8	16.3	1.31	0.88	0.12
12.2	173	20	21	5	2150	0.8	16.1	1.56	1.06	0.11
12.3	173	20	21	5	2650	0.8	15.8	1.93	1.31	0.11
12.4	173	20	21	5	3070	0.8	15.6	2.23	1.52	0.10
12.5	173	20	21	5	3450	0.8	14.6	2.51	1.74	0.06
12.6	173	20	21	5	4400	0.8	14.0	3.20	2.25	-0.01
13.1	200	10	21	5	1230	0.8	16.8	0.90	0.60	0.82
13.2	200	10	21	5	2060	0.8	16.2	1.50	1.01	0.77
13.3	200	10	21	5	3030	0.8	15.5	2.21	1.51	0.71
13.4	200	10	21	5	4600	0.8	14.4	3.35	2.33	0.63
14.1	200	20	21	5	3800	0.8	42.0	2.77	1.31	0.13
14.2	200	20	21	5	3800	0.8	35.0	2.77	1.41	0.12
14.3	200	20	21	5	3800	0.8	28.0	2.77	1.55	0.12
14.4	200	20	21	5	3800	0.8	21.0	2.77	1.72	0.12
14.5	200	20	21	5	3800	0.8	14.8	2.77	1.91	0.12
14.6	200	20	21	5	3800	0.8	8.7	2.77	2.17	0.13
15.1	140	20	21	5	3000	2.0	8.5	2.18	1.72	0.09
15.2	140	20	21	5	3000	1.5	8.3	2.18	1.73	0.08
15.3	140	20	21	5	3000	1.1	8.1	2.18	1.74	0.07
15.4	140	20	21	5	3000	0.8	8.3	2.18	1.73	0.04
16.1	200	20	21	5	3070	0.8	14.1	2.23	1.57	0.05
16.2	200	20	21	5	3070	0.8	14.1	2.23	1.57	0.04
16.3	200	20	21	5	3070	0.8	14.1	2.23	1.57	0.02
16.4	200	20	21	5	3070	0.8	14.1	2.23	1.57	0.01
17.1	118	20	21	5	3740	1.1	8.0	2.72	2.17	0.07
17.2	137	20	21	5	3740	1.1	8.0	2.72	2.17	0.08
17.3	212	20	21	5	3740	1.1	7.8	2.72	2.18	0.13
17.4	355	20	21	5	3740	1.1	7.8	2.72	2.18	0.22
18.1	200	20	21	5	2030	0.8	15.9	1.48	1.00	0.13
18.2	200	20	21	10	2030	0.8	15.9	1.48	1.00	0.13
18.3	200	20	21	15	2030	0.8	15.9	1.48	1.00	0.13
18.4	200	20	21	20	2030	0.8	15.9	1.48	1.00	0.13

Run No.	CONCENTRATE							TAILINGS		
	Vol. Rate	Mass Rate	Solids Density	% Sol. Rec.	Water Rec	S Grade	S Rec.	Vol. Rate	Mass Rate	S Grade
12–18	ml/min	g/min	%(m/m)	%	g/min	%S	%S	ml/min	g/min	%S
12.1	11	6.3	43	4	8.4	31.9	66	1105	152.2	0.69
12.2	18	9.5	40	6	14.1	30.8	77	1098	153.2	0.57
12.3	26	10.9	34	7	20.8	27.7	80	1077	148.1	0.51
12.4	40	12.5	27	7	34.1	25.2	82	1097	154.2	0.44
12.5	88	13.7	14	9	82.2	23.3	84	1027	146.7	0.42
12.6	186	16.2	8	10	179.0	21.1	87	926	148.0	0.35
13.1	8	6.2	54	4	5.2	39.5	76	2027	142.6	0.53
13.2	7	5.1	52	4	4.6	42.4	76	1886	136.4	0.49
13.3	8	5.6	51	4	5.3	44.5	81	1771	130.8	0.43
13.4	14	8.4	46	5	10.0	37.2	83	1929	155.4	0.40
14.1	26	12.9	39	7.0	20.4	32.4	89	1151	170.1	0.29
14.2	29	12.3	34	7.3	23.9	29.0	88	1087	157.4	0.31
14.3	31	13.1	35	7.3	24.8	29.3	89	1121	167.4	0.28
14.4	33	13.3	33	7.4	27.4	28.7	88	1162	167.0	0.32
14.5	32	12.4	32	7.2	26.6	28.0	85	1147	159.2	0.37
14.6	27	11.2	34	6.5	22.1	30.0	84	1144	160.5	0.39
15.1	22	7.1	27	4	19.2	24.5	73	1713	156.3	0.42
15.2	28	8.4	26	5	23.9	27.6	77	1689	156.3	0.45
15.3	48	10.9	20	7	43.3	26.0	79	1656	156.3	0.49
15.4	90	12.0	12	7	84.7	26.3	83	1630	158.8	0.40
16.1	131	7.9	6	8	127.2	17.4	73	1127	89.2	0.57
16.2	152	14.1	9	8	145.4	18.2	72	1061	162.5	0.60
16.3	173	20.5	11	8	163.9	19.6	71	937	247.3	0.67
16.4	180	23.4	12	6	169.7	19.4	59	878	393.9	0.80
17.1	17	6.7	33	6	13.5	29.5	78	1145	112.5	0.49
17.2	21	7.3	29	6	17.7	26.8	78	1159	108.9	0.49
17.3	30	8.0	23	7	26.1	25.5	82	1225	108.6	0.41
17.4	48	7.9	15	6	44.4	26.0	77	1354	117.0	0.52
18.1	27	10.5	32	6	22.4	24.8	74	1184	166.4	0.55
18.2	23	10.4	37	6	18.1	26.9	73	1192	165.6	0.61
18.3	21	10.2	38	6	16.9	25.8	72	1201	164.9	0.61
18.4	20	9.7	38	6	15.8	25.0	72	1243	164.7	0.58

Run No.	CARRYING CAPACITY					SOLIDS AND LIQUID RESIDENCE TIME				
	Bubble Size	Js	d80	'Ca'	Carry. Const.	Eg	d50	TAU l	TAU p	TAU p Froth
12–18	*cm*	*cm/sec*	*microns*	*g/min*	*K*	*(frac)*	*microns*	*min*	*min*	*min*
12.1	0.21	870	54.5	6.3	0.47	0.13	30.2	4.1	4.0	0.32
12.2	0.20	1059	51.4	9.5	0.62	0.14	27.8	4.0	4.0	0.26
12.3	0.21	1280	50.3	10.9	0.60	0.16	26.7	4.0	4.0	0.21
12.4	0.21	1483	56.8	12.5	0.52	0.17	30.8	3.9	3.9	0.18
12.5	0.22	1590	53.7	13.7	0.55	0.22	27.4	3.9	3.9	0.16
12.6	0.22	2028	54.7	16.2	0.50	0.27	25.2	4.1	4.1	0.12
13.1	0.22	569	50.7	6.2	0.77	0.07	27.6	2.4	2.4	0.46
13.2	0.20	1030	50.0	5.1	0.35	0.11	27.0	2.4	2.4	0.28
13.3	0.17	1751	48.2	5.6	0.23	0.15	26.3	2.5	2.5	0.19
13.4	0.20	2277	47.9	8.3	0.26	0.21	24.8	2.1	2.1	0.12
14.1	0.20	1900	56.5	13	0.61	0.18	33.1	10.3	10.3	0.15
14.2	0.20	1900	49.0	12	0.60	0.19	24.1	9.1	9.1	0.15
14.3	0.20	1900	48.6	13	0.60	0.21	23.2	7.0	7.0	0.15
14.4	0.20	1900	57.0	13	0.46	0.23	33.7	5.0	5.0	0.15
14.5	0.20	1900	49.4	12	0.44	0.23	25.0	3.5	3.5	0.15
14.6	0.20	1900	55.8	11	0.31	0.23	33.2	2.0	2.0	0.15
15.1	0.15	2000	51.2	7.1	0.21	0.58	24.7	1.1	1.1	0.76
15.2	0.15	2000	57.3	8.4	0.22	0.56	27.3	1.2	1.2	0.57
15.3	0.15	2000	66.6	10.9	0.24	0.55	29.1	1.3	1.3	0.38
15.4	0.15	2000	60.7	12.0	0.30	0.57	27.0	1.4	1.4	0.19
16.1	0.20	1535	30.1	7.9	0.58	0.23	17.0	3.5	3.5	0.18
16.2	0.20	1535	53.9	14.1	0.58	0.26	25.1	3.6	3.6	0.18
16.3	0.20	1535	53.4	20.5	0.85	0.30	25.7	3.9	3.9	0.18
16.4	0.20	1535	67.9	23.4	0.76	0.36	31.3	3.8	3.8	0.18
17.1	0.15	2493	48.5	6.7	0.17	0.56	24.4	1.8	1.8	0.30
17.2	0.15	2493	41.4	7.3	0.21	0.55	20.3	1.8	1.8	0.30
17.3	0.15	2493	55.7	8.0	0.17	0.57	32.6	1.6	1.6	0.30
17.4	0.15	2493	57.0	7.9	0.17	0.57	33.8	1.5	1.5	0.30
18.1	0.20	1015	82.0	10.5	0.44	0.15	40.4	3.7	3.7	0.28
18.2	0.20	1015	81.8	10.4	0.44	0.15	40.7	3.7	3.7	0.28
18.3	0.20	1015	82.5	10.2	0.43	0.15	41.1	3.6	3.6	0.28
18.4	0.20	1015	84.4	9.7	0.40	0.15	42.4	3.5	3.5	0.28

Run No.	COLUMN DIMENSIONS				FEED CONDITIONS				
	Sparger Type	Coll. Height	Froth Height	Feed Pt From Top	Vol. Rate	Solids Rate	Solids Density	Grade Measured	Grade Calc.
19–22		cm	cm	cm	ml/min	g/min	%(m/m)	%S	%S
19.1	FCS	225	25	55	1011	177.0	16.0	1.85	2.25
19.2	FCS	225	25	70	1046	177.1	15.5	1.85	2.15
19.3	FCS	225	25	85	1054	176.4	15.3	1.85	2.19
20.1	SGDS	200	50	55	1082	122.1	10.6	2.65	2.30
20.2	SGDS	200	50	55	1024	115.8	10.6	2.11	1.72
20.3	SGDS	200	50	55	1010	117.4	10.9	2.11	1.97
21.1	FCS	225	25	55	1018	91.2	8.5	2.28	1.92
21.2	FCS	225	25	55	1109	102.5	8.8	2.28	2.14
21.3	FCS	225	25	55	986	88.7	8.6	2.28	2.27
21.4	FCS	225	25	55	1109	102.9	8.8	2.28	2.35
21.5	FCS	225	25	55	974	76.0	7.5	2.28	2.89
22.1	FCS	225	25	55	925	169.3	16.6	2.05	2.03
22.2	FCS	225	25	55	880	160.5	16.6	2.05	2.31
22.3	FCS	225	25	55	889	161.5	16.5	2.05	2.24
22.4	FCS	225	25	55	892	160.6	16.4	2.05	2.26

	REAGENT ADDITIONS						
Run No.	Collector one	Coll. one Dosage	Collector two	Coll. two Dosage	Cond. Time	Frother Type	Total Frother
19–22	*name*	*g/ton*	*name*	*g/ton*	*min*	*Name*	*ppm (Vol)*
19.1	SIPX	22	AFLT 208	37	30	AF 65	23.0
19.2	SIPX	22	AFLT 208	37	30	AF 65	22.2
19.3	SIPX	23	AFLT 208	38	30	AF 65	22.0
20.1	H SO	(pH=3.8)	Senkol 50	100	30	DF 200	21.8
20.2	H SO	(pH=8.0)	Senkol 50	100	30	DF 200	21.9
20.3	H SO	(pH=8.0)	Senkol 50	100	30	DF 200	22.0
21.1	SIBX	9.8	–	–	40	AF 65	23.1
21.2	SIBX	21.7	–	–	40	AF 65	22.7
21.3	SIBX	42.2	–	–	40	AF 65	23.2
21.4	SIBX	60.7	–	–	40	AF 65	22.8
21.5	SIBX	121.5	–	–	40	AF 65	23.3
22.1	SIPX	25	AFLT 208	42	30	AF 65	5.3
22.2	SIPX	26	AFLT 208	44	30	AF 65	10.0
22.3	SIPX	26	AFLT 208	44	30	AF 65	14.9
22.4	SIPX	26	AFLT 208	44	30	AF 65	19.6

Run No.	WASH WATER				AIR					
	Rate	Frother Addition	Temp	Position from Top	Vol. Rate	Press Top	Press Bott	Jg* STP	Jg Actual	Sup. Bias
19–22	ml/min	ppm	Deg C	cm	ml/min	mm Hg	mm Hg	cm/sec	cm/sec	cm/sec
19.1	200	20	21	5	2030	1.6	15.9	1.48	1.00	0.13
19.2	200	20	21	5	2030	1.6	15.9	1.48	1.00	0.13
19.3	200	20	21	5	2030	1.6	15.9	1.48	1.00	0.13
20.1	140	20	21	5	2000	1.6	8.7	1.46	1.14	0.07
20.2	140	20	21	5	2000	1.6	8.7	1.46	1.14	0.08
20.3	140	20	21	5	2000	1.6	8.7	1.46	1.14	0.08
21.1	200	20	21	5	3070	1.3	14.4	2.28	1.59	0.08
21.2	200	20	21	5	3070	1.3	14.4	2.28	1.59	0.08
21.3	200	20	21	5	3070	1.3	14.4	2.28	1.59	0.08
21.4	200	20	21	5	3070	1.3	14.4	2.28	1.59	0.09
21.5	200	20	21	5	3070	1.3	14.4	2.28	1.59	0.10
22.1	200	0	21	5	3000	1.3	15.2	2.18	1.50	0.11
22.2	200	0	21	5	3000	1.3	15.2	2.18	1.50	0.08
22.3	200	0	21	5	3000	1.3	15.2	2.18	1.50	0.07
22.4	200	0	21	5	3000	1.3	15.2	2.18	1.50	0.07

Run No.	CONCENTRATE							TAILINGS		
	Vol. Rate	Mass Rate	Solids Density	% Sol. Rec.	Water Rec	S Grade	S Rec.	Vol. Rate	Mass Rate	S Grade
19–22	ml/min	g/min	%(m/m)	%	g/min	%S	%S	ml/min	g/min	%S
19.1	27	10.5	32	6	22.38	29.3	77	1184	166.4	0.55
19.2	28	10.1	30	6	22.99	29.0	76	1219	167.0	0.54
19.3	26	8.8	28	5	22.19	32.4	74	1228	167.6	0.60
20.1	40	12.5	26	10	34.78	19.8	88	1182	109.6	0.30
20.2	24	7.4	26	6	20.83	22.6	84	1140	108.4	0.30
20.3	30	9.4	28	8	24.6	21.0	85	1120	108.0	0.32
21.1	93	4.9	5	5	90.68	18.6	52	1126	86.2	0.97
21.2	92	6.4	7	6	88.71	23.9	70	1218	96.1	0.68
21.3	83	6.7	8	8	80.44	23.6	79	1103	82.0	0.52
21.4	78	7.8	10	8	74.38	24.8	80	1231	95.1	0.50
21.5	67	7.2	10	9	63.78	25.8	84	1107	68.9	0.50
22.1	53	12.6	21	7	46.96	22.3	81	1072	156.8	0.41
22.2	84	14.0	15	9	77.84	22.5	85	996	146.4	0.38
22.3	100	13.9	13	9	93.29	22.0	85	989	147.6	0.38
22.4	107	14.0	12	9	101.17	22.2	85	985	146.7	0.37

	CARRYING CAPACITY					SOLIDS AND LIQUID RESIDENCE TIME				
Run No.	Bubble Size	Js	d80	'Ca'	Carry. Const.	Eg	d50	**TAU l**	**TAU p**	**TAU p Froth**
19–22	*cm*	*cm/sec*	*microns*	*g/min*	*K*	*(frac)*	*microns*	*min*	*min*	*min*
19.1	0.20	1015	–	–	–	0.20	–	3.5	–	0.28
19.2	0.20	1015	–	–	–	0.20	–	3.4	–	0.28
19.3	0.20	1015	–	–	–	0.20	–	3.4	–	0.28
20.1	0.20	1000	–	–	–	0.54	–	1.8	–	0.56
20.2	0.20	1000	–	–	–	0.54	–	1.8	–	0.57
20.3	0.20	1000	–	–	–	0.54	–	1.9	–	0.56
21.1	0.20	1535	–	–	–	0.24	–	3.5	–	0.18
21.2	0.20	1535	–	–	–	0.24	–	3.2	–	0.18
21.3	0.20	1535	–	–	–	0.24	–	3.6	–	0.18
21.4	0.20	1535	–	–	–	0.24	–	3.2	–	0.18
21.5	0.20	1535	–	–	–	0.24	–	3.6	–	0.18
22.1	0.20	1500	–	–	–	0.22	–	3.7	–	0.19
22.2	0.20	1500	–	–	–	0.22	–	4.0	–	0.19
22.3	0.20	1500	–	–	–	0.22	–	4.0	–	0.18
22.4	0.20	1500	–	–	–	0.22	–	4.1	–	0.18

APPENDIX E

PARTICLE SIZE DATA

The particle size data is recorded as a weight percent in each size fraction. There are however exceptions when the particle size data is recorded in terms of mass of solids collected in the concentrate.

The tables that use the mass collected instead of the weight percent in each size fraction are marked with a MC (g). The MC (g) stands for mass of solids, in gramms, collected per minute.

COMPARISON BETWEEN COLUMN AND CONVENTIONAL BATCH CELL

SIZE	CONVENTIONAL BATCH FLOTATION				RUN 10	
(Microns)	*Conc 3*	*Conc 4*	*Conc 5*	*Conc 6*	*Column*	
1.2	0.0	0.0	0.0	0.0	0.0	
1.5	0.0	0.0	0.0	0.0	0.0	
1.9	0.0	0.0	0.0	0.0	0.0	
2.4	0.3	0.3	0.1	0.3	0.0	
3.0	2.0	1.9	1.8	2.8	0.3	
3.9	1.3	1.1	1.5	1.8	0.8	
5.0	2.2	2.2	2.3	2.7	1.6	
6.4	4.0	4.2	3.9	5.7	2.5	
8.2	4.9	4.6	5.3	7.4	3.1	
10.5	6.2	6.0	7.0	7.7	4.0	
13.6	7.8	8.8	8.6	10.9	6.0	
17.7	8.8	8.6	10.1	12.1	7.9	
23.7	14.4	14.1	15.5	14.6	10.4	
33.7	13.1	14.0	13.6	13.1	14.2	
54.9	22.2	20.9	20.5	14.6	27.2	
118.4	12.9	13.4	9.7	6.2	22.0	

VARYING AIR RATE WHEN USING THE SGDS – RUN 11

SIZE	AIR RATE (cm/sec)					MC (g)
(Microns)	*0.87*	*1.11*	*1.58*	*2.18*	*2.76*	
2.4	0.005	0.010	0.011	0.032	0.044	
3.0	0.015	0.026	0.027	0.064	0.081	
3.8	0.060	0.072	0.071	0.166	0.191	
4.8	0.090	0.114	0.104	0.236	0.287	
6.2	0.110	0.139	0.131	0.274	0.338	
7.9	0.184	0.217	0.213	0.420	0.493	
10.1	0.284	0.336	0.305	0.599	0.662	
13.0	0.354	0.423	0.387	0.662	0.736	
16.7	0.478	0.516	0.464	0.719	0.758	
21.5	0.628	0.645	0.600	0.840	0.868	
28.1	0.658	0.713	0.638	0.796	0.890	
37.6	0.737	0.656	0.682	0.694	0.787	
53.5	0.693	0.645	0.758	0.516	0.633	
87.2	0.583	0.558	0.747	0.293	0.478	
188.0	0.095	0.083	0.305	0.019	0.051	

VARYING AIR RATE WHEN USING THE FCS – RUN12

SIZE	AIR RATE (cm/sec)					MC (g)
(Microns)	*2.5*	*3.0*	*3.5*	*4.0*	*4.5*	5.7
1.2	0.000	0.000	0.000	0.000	0.000	0.000
1.5	0.000	0.000	0.000	0.000	0.000	0.000
1.9	0.000	0.000	0.000	0.000	0.000	0.000
2.4	0.000	0.000	0.000	0.000	0.014	0.065
3.0	0.006	0.009	0.022	0.025	0.069	0.341
3.9	0.025	0.047	0.076	0.087	0.137	0.243
5.0	0.075	0.142	0.174	0.175	0.137	0.341
6.4	0.157	0.256	0.316	0.312	0.411	0.665
8.2	0.208	0.294	0.392	0.399	0.521	0.892
10.5	0.258	0.465	0.523	0.599	0.672	0.892
13.6	0.447	0.731	0.849	0.823	1.056	1.135
17.7	0.629	0.845	1.089	1.085	1.302	1.395
23.7	0.730	1.329	1.448	1.484	1.631	1.849
33.7	0.944	1.509	1.786	1.846	2.125	1.817
54.9	1.591	2.354	2.581	2.955	2.948	3.341
118.4	1.233	1.509	1.623	2.756	2.577	3.212

VARYING AIR RATE WHEN USING THE USBM – RUN13

SIZE	AIR RATE (cm/sec)					MC (g)
(Microns)	*1.51*	*1.01*	*0.60*	*2.33*		
1.2	0.000	0.000	0.000	0.000		
1.5	0.000	0.000	0.000	0.000		
1.9	0.000	0.000	0.000	0.000		
2.4	0.000	0.000	0.000	0.000		
3.0	0.004	0.000	0.000	0.026		
3.9	0.024	0.014	0.017	0.062		
5.9	0.073	0.054	0.075	0.113		
6.4	0.138	0.107	0.104	0.185		
8.2	0.163	0.140	0.120	0.231		
10.5	0.216	0.186	0.262	0.293		
13.6	0.330	0.275	0.307	0.405		
17.7	0.395	0.343	0.303	0.498		
23.7	0.496	0.461	0.661	0.662		
33.7	0.716	0.594	0.590	0.832		
54.9	1.013	0.884	1.101	1.196		
118.4	0.496	0.515	0.615	0.636		

VARYING THE FROTH DEPTH - RUN15

SIZE	FROTH DEPTH (cm)					
(Microns)	*100*	*75*	*50*	*25*		
1.9	0.0	0.0	0.0	0.0		
2.4	0.0	0.0	0.1	0.1		
3.0	0.2	0.3	0.3	0.4		
3.8	2.0	2.0	2.1	2.1		
4.8	3.2	2.8	2.5	2.6		
6.2	1.9	1.8	1.7	1.9		
7.9	3.4	3.1	3.1	3.3		
10.1	5.7	5.3	4.8	5.2		
13.0	7.4	6.5	6.2	6.8		
16.7	9.1	7.9	7.9	8.8		
21.5	11.0	10.2	9.4	9.8		
28.1	12.0	11.4	10.5	10.9		
37.6	14.6	13.9	14.0	14.5		
53.5	10.8	13.0	11.4	10.6		
87.2	11.8	15.0	15.6	14.1		
188.0	6.6	6.7	10.5	9.0		

VARYING THE PERCENT SOLIDS FEED RATE - RUN16

SIZE	% SOLIDS FEED DENSITY (mass/mass)					MC (g)
(Microns)	*8.7*	*15.9*	*25.3*	*38.4*		
1.2	0.000	0.000	0.000	0.000		
2.4	0.007	0.008	0.011	0.024		
3.0	0.021	0.034	0.055	0.084		
3.8	0.121	0.202	0.262	0.240		
4.8	0.156	0.269	0.327	0.372		
6.2	0.114	0.168	0.251	0.552		
7.9	0.185	0.269	0.403	0.804		
10.1	0.298	0.479	0.632	0.876		
13.0	0.383	0.588	0.774	1.224		
16.7	0.504	0.655	0.926	1.728		
21.5	0.653	0.899	1.123	1.848		
28.1	0.760	0.991	1.243	1.608		
37.6	1.037	1.168	1.439	1.224		
53.5	0.937	0.983	1.210	0.792		
87.2	1.193	1.109	1.362	0.480		
188.0	0.731	0.571	0.850	0.168		

VARYING THE VOLUMETRIC FEED RATE – RUN10

SIZE	VOLUMETRIC FEED RATE (ml/min)					
(Microns)	*841*	*1016*	*1292*	*1578*	*1726*	*1958*
1.9	0.0	0.0	0.0	0.0	0.0	0.0
2.4	0.4	0.4	0.3	0.3	0.2	0.4
3.0	0.9	0.8	0.7	0.5	0.4	0.9
3.8	2.2	1.9	1.8	1.7	1.4	2.8
4.8	3.3	2.9	2.7	2.6	2.6	4.0
6.2	3.9	3.5	3.2	2.8	2.7	4.3
7.9	5.9	5.4	5.0	4.3	4.0	6.9
10.1	8.2	7.6	6.9	6.3	5.5	9.7
13.0	9.2	8.6	7.9	7.6	7.7	10.8
16.7	9.9	9.5	9.0	8.6	9.0	11.5
21.5	11.9	11.6	11.2	10.3	9.9	13.1
28.1	11.6	11.8	11.3	11.8	11.2	12.4
37.6	11.6	11.6	11.7	13.5	15.3	8.9
53.5	10.8	11.5	11.6	13.7	--	7.5
87.2	7.9	9.9	12.4	15.5	16.6	5.7
188.0	1.8	2.3	3.9	0.1	2.5	0.6

VARYING THE WASH WATER RATE – RUN17

SIZE	WASH WATER RATE (ml/min)					
(Microns)	*118*	*137*	*212*	*355*		
1.2	0.0	0.0	0.0	0.0		
1.5	0.0	0.0	0.0	0.0		
1.9	0.0	0.0	0.0	0.0		
2.4	0.0	0.0	0.0	0.0		
3.0	0.9	1.3	0.6	0.6		
3.9	1.2	1.8	1.0	1.1		
5.0	2.3	2.7	2.4	2.3		
6.4	3.1	3.9	2.2	2.1		
8.2	3.8	5.6	2.7	3.0		
10.5	6.9	8.1	6.6	6.6		
13.6	7.8	9.2	5.6	5.4		
17.7	7.7	10.3	5.5	5.6		
23.7	15.4	16.5	14.0	13.4		
33.7	12.9	13.1	10.6	9.8		
54.9	25.6	20.3	27.9	27.4		
118.4	12.3	7.1	21.0	22.7		

VARYING THE WASH WATER POSITION – RUN18						
SIZE	WASH WATER POSITION (cm)					
(Microns)	*5*	*10*	*15*	*20*		
1.9	0.0	0.0	0.0	0.0		
2.4	0.0	0.0	0.0	0.0		
3.0	0.0	0.0	0.0	0.1		
3.8	0.4	0.5	0.5	0.5		
4.8	0.9	0.9	0.9	0.9		
6.2	1.1	1.0	1.0	1.0		
7.9	1.7	1.6	1.6	1.6		
10.1	2.7	2.7	2.6	2.5		
13.0	3.8	3.7	3.7	3.6		
16.7	5.4	5.4	5.4	5.2		
21.5	7.8	7.7	7.5	7.3		
28.1	9.8	9.8	9.8	9.5		
37.6	13.6	13.6	13.6	13.2		
53.5	15.3	15.3	15.2	15.1		
87.7	20.5	21.0	21.1	21.3		
188.0	16.8	16.7	17.1	18.2		

VARYING THE WASH WATER TEMPERATURE – RUN7						
SIZE	WASH WATER TEMPERATURE (Degrees Celcius)					
(Microns)	*21*	*30*	*40*	*50*		
1.2	0.0	0.0	0.0	0.0		
2.4	0.4	0.3	0.3	0.3		
3.0	0.7	0.7	0.6	0.6		
3.8	1.9	1.9	2.1	2.2		
4.8	3.0	2.8	3.1	3.3		
6.2	3.3	3.3	3.2	3.2		
7.9	5.2	5.2	5.1	5.2		
10.1	7.5	7.5	7.9	7.7		
13.0	8.6	8.4	9.2	9.1		
16.7	9.5	9.6	10.3	10.2		
21.5	10.9	11.6	12.9	12.2		
28.1	11.7	12.1	13.4	12.2		
37.6	12.7	12.4	12.3	12.7		
53.5	11.4	12.5	10.5	10.3		
87.2	12.6	11.1	8.3	9.9		
188.0	0.1	0.4	0.6	0.8		

VARYING THE COLLECTOR DOSAGE RATE – RUN21						
SIZE	COLLECTOR DOSAGE RATE (g/ton)					
(Microns)	*9.8*	*21.7*	*42.2*	*60.7*	*121.5*	
1.2	0.0	0.0	0.0	0.0	0.0	
1.5	0.0	0.0	0.0	0.0	0.0	
1.9	0.0	0.0	0.0	0.0	0.0	
2.4	0.0	0.1	0.0	0.0	0.0	
3.0	0.3	0.5	0.3	0.0	0.0	
3.9	0.9	0.9	0.6	0.2	0.2	
5.0	2.7	2.0	1.8	1.2	1.3	
6.4	1.5	1.4	1.1	0.5	0.5	
8.2	2.1	2.1	1.5	0.8	0.7	
10.5	7.1	6.2	5.3	4.0	3.6	
13.6	6.3	5.5	4.9	2.7	2.9	
17.7	6.2	5.9	4.9	3.0	2.6	
23.7	16.9	15.7	14.0	11.5	9.8	
33.7	12.1	11.6	11.2	10.1	9.4	
54.9	25.2	26.4	26.4	30.9	27.6	
118.4	18.6	21.7	28.1	35.0	41.3	

APPENDIX F

PILOT PLANT DATA

The data recorded for the pilot plant test work together with the calculated data is given in this appendix.

The fractional gas holdup was estimated by stopping the feed and tailings pumps, as well as the air to the column. The fractional gas holdup could only be estimated due to the pulp from the froth increasing the reading for the pulp in the collection zone.

The mean liquid residence time was calculated by assuming that the liquid flow was plug flow.

RUN NUMBER (Date)	COLUMN DIAMETER (cm)	COLLECTION HEIGHT (cm)	FROTH DEPTH (cm)	FRACTIONAL GAS HOLDUP (Estimated)
1 (05/09)	10.6	350	50	0.11
2 (06/09)	10.6	350	50	0.11
3 (06/09)	10.6	350	50	0.11
4 (06/09)	10.6	350	50	0.11
5 (06/09)	10.6	350	50	0.11
6 (06/09)	10.6	350	50	0.11
7 (06/09)	10.6	350	50	0.07
8 (06/09)	10.6	350	50	0.09
9 (06/09)	10.6	350	50	0.11
10 (06/09)	10.6	350	50	0.13
11 (08/09)	10.6	350	50	0.22
12 (08/09)	10.6	350	50	0.11
13 (08/09)	10.6	350	50	0.11
14 (08/09)	10.6	350	50	0.11
15 (11/09)	5.7	198	50	0.21
16 (11/09)	10.6	350	50	0.11
17 (12/09)	10.6	200	50	0.21
18 (12/09)	10.6	200	50	0.22
19 (12/09)	10.6	225	25	0.22
20 (13/09)	5.7	210	40	0.33
21 (13/09)	10.6	200	50	0.23
22 (07/09)	10.6	350	50	0.11
23 (07/09)	10.6	100	50	0.17
24 (07/09)	10.6	100	50	0.17
25 (07/09)	10.6	100	50	0.17

RUN NUMBER	AIR RATE (ml/min)	WASH WATER RATE (ml/min)	FEED RATE (ml/min)	TAILINGS RATE (ml/min)
1	11650	941		
2	11650	761	3660	4197
3	11650	0	3919	3884
4	11650	414	4160	4514
5	11650	941	3900	4748
6	11650	1462	4621	5860
7	7714	941	4252	5150
8	9906	941	4133	5020
9	11650	941	3900	4748
10	12857	941	3822	4645
11	12030	941	2704	3498
12	11650	0	2952	2590
13	11650	64	3194	2904
14	11650	1181	2607	3409
15	3445	277	521	742
16	11650	941	3227	3903
17	17352	1462	1346	2514
18	20773	1462	1299	1866
19	20773	1462	999	1436
20	5714	277	1703	1926
21	24127	1462	1082	1703
22	11650	941	2902	3697
23	11650	941	2916	3783
24	11650	941	3301	4162
25	11650	941	4086	4907

RUN NUMBER	CONCENTRATE RATE	FEED SOLIDS RATE	CONCENTRATE SOLIDS RATE	TAILINGS SOLIDS RATE
	(ml/min)	(g/min)	(g/min)	(g/min)
1				
2	224	1882	10	1872
3	35	2223	9	2214
4	60	2271	12	2259
5	93	2175	15	2161
6	223	2625	15	2610
7	43	2361	10	2351
8	53	2244	11	2233
9	93	2175	15	2161
10	118	2154	14	2140
11	147	1126	6	1120
12	362	851	72	779
13	354	535	60	474
14	379	953	27	926
15	56	227	7	220
16	265	1822	19	1803
17	294	569	9	559
18	894	577	20	557
19	1025	593	22	571
20	54	448	2	447
21	841	466	17	448
22	146	1472	11	1461
23	73	1292	6	1286
24	80	1713	9	1704
25	120	2080	13	2067

RUN NUMBER	SUPERFICIAL AIR RATE	SUPERFICIAL W/WATER RATE	SUPERFICIAL BIAS RATE	MEAN LIQUID RESIDENCE t
	(cm/sec)	(cm/sec)	(cm/sec)	(min)
1				
2	1.84	0.14	0.10	6.55
3	1.84	0.00	-0.01	7.08
4	1.85	0.08	0.07	6.09
5	1.86	0.18	0.16	5.79
6	1.86	0.28	0.23	4.69
7	1.22	0.18	0.17	5.58
8	1.58	0.18	0.17	5.60
9	1.86	0.18	0.16	5.79
10	2.06	0.18	0.16	5.79
11	1.97	0.18	0.15	6.89
12	1.88	0.00	-0.07	10.61
13	1.83	0.01	-0.05	9.47
14	1.88	0.22	0.15	8.06
15	2.07	0.18	0.14	5.38
16	1.83	0.18	0.13	7.04
17	3.02	0.28	0.22	5.55
18	3.60	0.28	0.11	7.38
19	3.57	0.28	0.08	10.78
20	3.42	0.18	0.15	1.86
21	4.19	0.28	0.12	7.98
22	1.85	0.18	0.15	7.44
23	1.88	0.18	0.16	1.94
24	1.87	0.18	0.16	1.76
25	1.86	0.18	0.16	1.49

RUN NUMBER	CONCENTRATE GRADE	TAILINGS GRADE	SULPHUR RECOVERY	GOLD RECOVERY
	%S	%S	%S	%Au
1	37.6	1.2	##	##
2	31.7	1.3	12	##
3	22.0	1.4	6	##
4	29.6	1.4	11	##
5	32.8	1.3	14	##
6	32.7	1.3	12	##
7	38.3	1.5	10	##
8	34.8	1.3	12	##
9	32.8	1.3	14	##
10	30.8	1.4	12	##
11	30.4	1.1	13	##
12	15.9	4.7	24	##
13	19.4	4.2	37	##
14	19.6	4.3	12	##
15	34.3	0.9	54	28
16	29.6	1.2	21	27
17	33.4	0.9	39	55
18	17.6	0.8	44	63
19	15.2	0.9	39	61
20	11.5	0.8	5	61
21	20.0	0.9	47	##
22	35.1	1.1	19	##
23	39.8	1.2	14	##
24	39.7	1.2	15	##
25	41.3	1.2	18	##

RUN NUMBER	FEED GRADE	PERCENTAGE SOLIDS REC	% SOLIDS IN CONC.	SUPERFICIAL BUB. SURFACE
	%S	%	%	Js
1	##	##	##	##
2	1.46	0.53	4.34	961.82
3	1.51	0.40	21.72	961.82
4	1.50	0.53	17.76	966.64
5	1.54	0.67	14.28	969.84
6	1.51	0.57	6.39	970.51
7	1.65	0.44	21.03	637.92
8	1.43	0.49	18.31	822.50
9	1.54	0.67	14.28	969.84
10	1.62	0.65	11.03	1073.45
11	1.28	0.53	3.97	1029.27
12	5.63	8.41	17.59	980.78
13	5.92	11.30	15.41	952.38
14	4.75	2.86	6.88	982.96
15	1.85	2.93	10.97	1079.42
16	1.45	1.03	6.78	956.11
17	1.42	1.66	3.15	1573.50
18	1.38	3.50	2.23	1877.06
19	1.44	3.69	2.11	1864.39
20	0.87	0.38	3.09	1784.63
21	1.58	3.74	2.04	2187.07
22	1.35	0.75	7.18	964.77
23	1.37	0.50	8.31	981.00
24	1.41	0.52	10.52	973.08
25	1.46	0.62	10.12	971.62