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The Development of a Flotation Simulation Methodology Towards an Optimisation Study of UG2 Platinum Flotation Circuits

T.A.Harris BSc (Chem Eng)

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Requirements for the Degree of
Doctor of Philosophy**



**Mineral Processing Research Unit
Department of Chemical Engineering
University of Cape Town**

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I hereby certify that the work embodied in this
thesis is the result of original research and has
not been submitted for a higher degree to any other
university or institution.

ABSTRACT

The design of new flotation plants is still based substantially on experience obtained from prior plant designs. Design criteria are typically based on rule of thumb procedures that often prove to be highly ineffective. The development of appropriate models of the flotation process is a means of addressing these problems so that flotation plant simulations can be performed to identify the true potential of existing plants as well as in the design of new flotation plants. This thesis addresses the complex issue of developing a robust methodology for use in flotation circuit modelling and simulation, based on the operation of pilot-scale flotation equipment. The investigation is focussed on the processing of platinum bearing ores from the UG2 reef in South Africa. The UG2 reef is of considerable economic importance, containing a substantial proportion of the world wide reserves of platinum group metals. However, UG2 ore has very complex mineralogy, and a high degree of association with chromite, which renders processing of this ore relatively difficult. Consequently, optimising the design of flotation circuits to treat this ore has proved challenging, and has been the focus of much attention by major platinum producers over the last 10 to 20 years.

The primary aim of this thesis is to identify, evaluate and refine appropriate flotation models and modelling procedures reported in the literature, and use these models to develop a flotation circuit simulation procedure that can be used to provide a sound basis for the evaluation of current UG2 circuit design trends, and as an optimisation tool for the design of UG2 circuits in the future.

The processing of UG2 ore from three different locations is evaluated, using two continuous pilot plants of different size, each operated in a variety of different configurations. The investigation required extensive modification of the pilot plant facility that was used, to facilitate the acquisition of the data required by the selected models. The basis and implications of these modifications with respect to the design and operation of pilot plants as a tool for design and optimisation is extensively covered.

The modelling approach adopted in this study is based on the separation of the factors affecting flotation into those arising from the characteristics of the ore (floatability) and those arising owing to the hydrodynamics in the equipment, in both the pulp and froth zones of the cell. It is proposed that the pulp zone performance can best be related to the bubble surface area flux based on the work of Gorain et al (1997), while ore floatability is modelled using a distributed floatability model, based on the approach developed by Runge et al (1997) and Harris (1997). The performance of the froth phase with respect to minerals recovered by true flotation is modelled based on the approach proposed by Harris (1997) and Gorain et al (1998), while entrainment in the froth is modelled using the approach proposed by Savassi et al (1998).

This thesis emphasises the central role of a water recovery model when attempting to simulate the performance of flotation circuits. Water recovery models reported in the literature are evaluated, and their limitations for use in this type of study are highlighted. A new, albeit fully empirical, model is developed for use in this work, based on the froth residence time of air, which is able to account for the effect of cell operating conditions such as gas rate and froth height during simulation. One of the conclusions in this study is that this very important aspect of flotation modelling deserves and requires more attention from flotation researchers in the future.

A step by step algorithm for the modelling and simulation of a pilot plant is presented, showing in detail how the individual component models can be combined into a structure that is relatively simple and computationally robust, based on implementation in an Excel spreadsheet. The models that are used are shown to fit the data derived from the operation of both pilot plants very well, for a wide range of cell operating conditions and circuit configurations. The use of the models for predictive simulation is also shown to be very encouraging, within the constraints imposed by limitations with respect to the predictive modelling of the water recovery, and the use of appropriate boundary conditions for the component models.

An extensive analysis of the parameters in the various models is presented with respect to their correlation with measured process variables, based on ore characteristics such as particle size, mineralogy and liberation, and cell operating conditions such as froth height and gas rate.

The pilot plant simulation model is used to analyse and evaluate the appropriate use of recycle in the design of flotation circuits, with particular reference to the design of circuits to process UG2 ore. Current “evolutionary” trends in UG2 circuit design are validated by simulation results. In addition, the simulation results provide a strong indication that further improvements are possible in the design of these circuits. The results of this investigation suggest that it should be possible to achieve either higher grades at the same recoveries, or higher recoveries at the same grade, than is currently typically the case on plant operations treating UG2 platinum ores.

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NOMENCLATURE

Subscripts

d'	size fraction
i	floatability component
j	mineral
k	flotation cell or bank
s	stream
z	circuit node

Arabic Symbols

A	cross section area (m^2)
a	empirical parameter in S_b model
a'	water recovery slope parameter in King's model
$a_{p,q}$	fraction of tailings re-circulated from cell (p) to cell (q)
A_s	impeller aspect ratio
$a_{u,j,s}$	assay of mineral (j) in concentrate (u) during flotation of stream (s)
b	empirical parameter in S_b model
b'	water recovery exponent parameter in King's model
$b_{p,q}$	fraction of concentrate re-circulated from cell (p) to cell (q)
c	empirical parameter in S_b model
C_∞	portion of feed that will not float even at infinite flotation time
$C(g)$	connection matrix of tailings flow rates
c_b	the concentration of bubbles in the flotation unit at time (t)
c_m	the concentration of mineral particles in the flotation unit at time (t), (g/l)
d	empirical parameter in S_b model
D	flotation cell depth (cm)
d_b	sauter mean bubble diameter (mm)
d_{calc}	model predicted data point
d_{exp}	experimentally measured data point
e	empirical parameter in S_b model
$E(t)$	residence time distribution function
$E_{j,d'}$	degree of entrainment of mineral (j) in size class (d')
$F_{i,j,s}$	mass flow rate of floatability component (i) of mineral (j) in stream (s), (g/min)
$F_{i,j,s}^{in}$	mass flow rate of mineral (j) in floatability class (i) in stream (s) entering a circuit node (g/min)
$F_{i,j,s}^{out}$	mass flow rate of mineral (j) in floatability class (i) in stream (s) exiting a circuit node (g/min)
$g_{i,j,k}$	enhancement factor of floatability component (i), of mineral (j), in node or cell (k)
$G_{j,s}$	grade of mineral (j) in stream (s)
H	flotation cell height (cm)
h	number of batch floatability tests performed on a flotation circuit

l_H	Impeller stator height (cm)
l_R	Impeller stator radius (cm)
J_g	superficial gas velocity (cm/sec)
$k_{c,i,j}$	collection zone first order rate constant of floatability component (i), mineral (j) (1/min)
$k_{o,i,j}^{batch}$	overall first order flotation rate constant, of floatability component (i), mineral (j) in the batch test, (1/min)
L	flotation cell length (cm)
$m_{i,j,k}$	mass fraction of floatability class (i), of mineral (j), in stream (s)
N	Impeller speed (1/sec)
N	number of experimental data points
n	number of floatability classes
N_s	Impeller peripheral speed (m/sec)
P	number of parameters
P_{80}	80 % passing feed size (microns)
$P_{i,j}$	inherent floatability of component (i) of mineral (j)
Q_{air}	volumetric air flow rate (l/min)
$Q_{i,j,s}$	volumetric flow rate of component (i), of mineral (j), in stream (s), (l/min)
$Q_{i,j,s}^{total}$	total volumetric flow rate of mineral (j) in stream (s) (l/min)
r	dimensionless exponent in Gaudin-Meloy distribution function
R_{∞}	recovery at infinite flotation times
$R_{i,j,s}^{batch}$	is the batch recovery at time (t) of mineral (j) from stream (s) in the v^{th} concentrate
$R_{f,j,k}$	froth recovery of attached particles of mineral (j), in bank (k) in the flotation circuit
$R_{i,j,k}$	continuous flotation cell recovery of floatability component (i) of mineral (j) from flotation unit (k)
$R_{ent,i,j,k,d'}^o$	total entrained recovery of mineral (j) in unit (k) of size fraction (d')
$R_{i,j,k}^o$	total (overall) recovery of floatability component (i) of mineral (j) from continuous flotation unit (k)
R_w^o	overall water recovery
S	percentage solids (%)
$S_{b,k}$	bubble surface area flux in unit (k), ($mm^2/mm^2 \cdot sec$)
$Sg_{j,s}$	specific gravity of mineral (j) in stream (s)
$S_{j,d'}$	arithmetic mean particle size of mineral (j) in size class (d')
S_r	stator shaft radius (cm)
t	batch floatability test cumulative flotation time (min)
v	number of concentrate samples taken during a batch test
$V_{f,k}$	volume of froth in unit (k), (l)
$V_{pulp,k}$	volume of pulp in unit (k), (l)
$w_{u,s}$	weight % recovered in concentrate (u) during batch flotation of stream (s)
X	particle diameter (μm)
x	total number of streams composing the combined streams
X_i	parameter with dimension of size in Gaudin-Meloy distribution function (μm)
Y	cumulative fraction weight undersize in Gaudin-Meloy distribution function
y	total number of samples collected during batch flotation test, including tailings

y	total number of data points
$Y_{d',j}$	cumulative fraction weight undersize in size interval (d') of mineral (j)

Greek Symbols

$\alpha_{j,k}$	derived froth drainage parameter of mineral j in unit or bank k
ε_g	gas hold-up (%)
δ_j	drainage parameter in hyperbolic secant function for mineral (j)
$\phi_{act,i,s}$	activated floatable fraction of mineral (j) in stream (s)
ξ_j	entrainment parameter in hyperbolic secant function for mineral (j)
γ_j	fraction of mineral j in feed
Ω_k	empirical water froth recovery parameter in unit or bank (k)
σ_k	empirical water froth recovery parameter in unit or bank (k)
χ_k	empirical water froth recovery parameter in unit or bank (k)
$\tau_{air,k}$	mean froth residence time based on volumetric air flow rate in unit k (min)
$\tau_{c,k}$	mean flotation cell residence time based on tailings flow rate in unit k (min)
$\tau_{f,k}$	mean froth residence time based on concentrate volume flow rate in unit k (min)
μm	micrometer (dimension of size)

Abbreviations

ADJ	exponent contribution in the empirical hyperbolic secant entrainment function
Cr	chromite
FH	froth height
Ga	gangue (total dry mass less chromite and total PGM and Au)
HGC	high grade cleaner bank
KW	kilo watt of power
LGC	low grade cleaner bank
MF	model fit coefficient
PGM	platinum group metal
r.p.m.	revolutions per minute
StDev	standard deviation of an experimental data point
TSSE	total sum of error squared
W	water

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Chapter 1

INTRODUCTION

1.1 Background to flotation modelling and simulation practices

Flotation, the practice of concentrating minerals from an ore has established itself as an extremely useful and effective process in the metallurgical industry. Hydrophobic particles are separated by floating them to the top of a liquid suspension with the help of air bubbles, while particles that are naturally hydrophilic, or are made to be by means of suitable chemicals, remain with the liquid and are left un-recovered into the tailings stream of a flotation cell.

The effectiveness of this process is clearly evident in the platinum industry where ores containing an average feed grade of 4 g/t can be concentrated up to as much as 1000 g/t, achieving recoveries as high as ninety percent. Small quantities of reagents are required and process water can be recycled making this process extremely effective and viable. However, while the process itself is relatively simple to implement, describing the process on the basis of fundamental relationships is extremely difficult, as there are a very large number of variables, many of which interact with one another. The number of variables affecting the kinetics of a floating particle have been considered by Arbiter and Harris (1962), who have reported that as many as 100 variables contribute to the effective flotation of a particle in a flotation cell. Despite these difficulties, many approaches have been used to describe the different sub processes in order to model the kinetics of this process. Some models lump all the sub processes together while others try to describe each sub process individually. In the literature there are several examples of relatively fundamental models of the flotation process. However, the difficulty of measurement of many of the physical process variables used in these models, and the use of many unrealistic assumptions, render the application of these models on industrial plants impossible at this time. The general consensus is that the most useful flotation models are semi-empirical in nature. The importance in developing an effective model of the flotation process is to be able to make realistic simplifications which will reduce, to within reasonable limits the amount of experimental work required to derive the model parameters, but will not cause gross inaccuracies in its predictive abilities, (Lynch et al, 1981).

In addition to describing the separation of minerals in an individual cell, plants invariably employ a number of flotation cells to increase the separation efficiency of the process. Multistage circuits with recycle streams are employed to achieve satisfactory separation of the valuable minerals from the gangue. The performance of these circuits is controlled by the arrangement of the cells into a network. The design of new flotation circuits can, in most instances be considered an art based on experience, rather than a science. At present, the use of simulation to specify cell sizes, operating setpoints and

circuit layout for a new flotation plant cannot be undertaken with any confidence. This is due to a lack of practical, user friendly models describing the kinetics of each component in the circuit. Generally the level of complication of a network of flotation cells depends on the grade requirements as well as the characteristics of the feed material entering the plant. It is often not known how many stages of cleaning are required and many very different circuit configurations can achieve a satisfactory separation. As a result, many new flotation plants are over designed by as much as one hundred percent of the estimated residence times calculated from laboratory and pilot plant tests, as a safety cushion to ensure the required recoveries are met. Obviously these problems lead to excessively large plant capital costs, a wasted installed capacity as well as exceptionally long plant commissioning times.

Simulations of various aspects of mineral processing operations has been practiced in one form or another for many years. Primarily, the emphasis has been on empirically based experimental methods, for instance, batch locked cycle tests, that culminate in pilot plant testing. From the observations of the difficulties encountered in commissioning new plants it might be concluded that these methods could be improved. The increased emphasis on mathematical simulation techniques has enhanced interests in this general area of mineral processing. Often the results from the batch locked cycle tests can only provide information as to the "equilibrium" recoveries that can be achieved, and do not provide information of the effect of recycling streams and short circuiting of particles as a result of the continuous nature of operating plants. In addition, conditions differ greatly between laboratory and plant flotation cells. To be truly useful it is essential that concentrate grade and mineral recovery yield accurate estimates obtained in a continuous plant operation. This all highlights the advantages of using pilot plants that can be used for investigating the effect of recycle streams, number of flotation stages and possible circuit configurations.

1.2 The problem behind UG2 circuit designs

The issues discussed above are of particular relevance to the design of flotation circuits treating platinum bearing ores. The platinum group metals (PGM's) that are mined in South Africa are found in a number of different layers of the Bushveld Igneous Complex, of which the Merensky Reef and the Upper Group 2 chromitite seam (UG2 reef) are of greatest commercial importance, (Corrans et al, 1982). Owing to the high chromite content of UG2 ores, the treatment of this material often necessitates the use of very complicated flotation circuits that typically employ multiple cleaning stages to keep the chromite content within smelting specifications (typically < 3%). The optimisation of the performance of these circuits has proven to be very difficult because the run-of-mine ore is extremely variable. Batch flotation tests on several ores from various localities in the Bushveld Complex can show marked differences in response to concentration by flotation as a result of markedly different mineralogy between the samples. Different flotation characteristics can even be found between samples obtained from the same area of the complex, (Corrans et al, 1982).

This thesis was initiated in order to gain a greater understanding as to the reasons behind the recent circuit design trends of many new industrial platinum producers treating UG2 ores. Over a period of time the typical circuits have evolved from a conventional rougher/cleaner/re-cleaner circuit to what has been called the “Mintek” circuit in this thesis. These circuits are relatively complicated having multiple concentrate product streams. The reason for this is that it has been found to be beneficial to separate the particles that float very quickly and are easy to concentrate from the slower floating particles. It has been found that higher concentrate grades can be achieved by doing this. Re-grind stages are often employed after a primary flotation stage to increase the recovery. The concentrate from the re-grind rougher bank is also cleaned in a separate cleaning stage, without combining all the primary and re-grind concentrates into a single cleaning stage. While no scientific basis has been provided as to the reasons for these designs, they have been found to provide superior circuit performances based on pilot plant trials and have therefore become the “norm” for many of the new UG2 circuit designs. It was considered necessary to look at this problem from a simulation point of view to understand the mechanism behind this design trend in order to understand why experience tends to point in this direction. In addition, a simulation model could be used to exploit this principle to its full capability and could possibly be used to design even better circuits. Figure 1.1 shows a typical flotation circuit treating UG2 platinum ores, based on the “Mintek” design principle.

In order to design optimum circuits for the treatment of UG2 ores, the above problems can best be addressed by the development of suitable flotation circuit models capable of predicting the performance of a circuit in a variety of configurations. The use of a well controlled pilot plant facility where process disturbances and operating variables can be adequately measured is a means of obtaining enough information to develop these models. By ensuring the feed is homogeneous in composition, consistent, as well as having a controlled environment allows one to obtain accurate experimental data in order to determine the influence of operating variables during the processing of this ore.

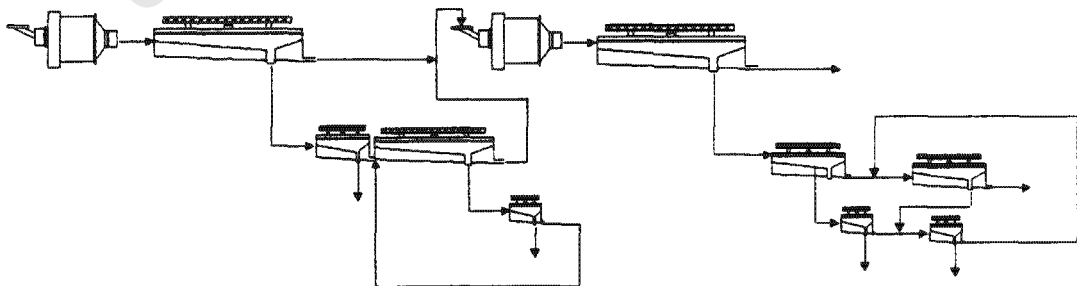


Figure 1.1: The typical design trend of newer UG2 flotation circuits producing multiple concentrate product streams

The primary objective of this thesis was to investigate whether a methodology could be developed to simulate the performance of a well controlled flotation pilot plant facility at the Council for Mineral Technology (Mintek) situated in Johannesburg, South Africa. The emphasis of the work was to determine whether existing models in the literature could be used to simulate the flotation kinetics of UG2 ore. A collaborative research venture between the Julius Kruttschnitt Mineral Research Centre (JKMRC) in Brisbane, Australia and the Mineral Processing Research Unit (MPRU), at the University of Cape Town, South Africa has been involved with the development of such models of the flotation process. The chief objective of this research group is to develop a methodology for modelling, simulating and optimizing existing flotation plants with confidence. An aim is to develop methodologies to improve the operation and the design of existing plants with an emphasis on a practical approach that is able to deliver something to the industry of practical use. Presently, much of the work done to date has focussed on practical models describing the kinetics of the floatable mineral particles, (Gorain et al, 1997). A methodology of obtaining enough information to derive the large number of parameters in the models has also been developed, (Runge et al, 1997), (Harris, 1997). Recently an extensive amount of work has been done to incorporate entrainment into this model as well as the influence of the froth phase on the overall kinetics, (Savassi et al, 1998), (Mathe et al, 1998). Much of this previous work forms the basis of the work presented in this thesis.

1.3 Thesis objectives

The primary objectives of this thesis can be divided into two main sections:

1. To identify, evaluate and refine appropriate flotation models and modelling procedures reported in the literature and to develop a flotation circuit simulation procedure that can be used to provide a sound basis for the evaluation of current UG2 circuit design trends.
2. To use this simulation methodology to attempt to understand and develop sound, scientific procedures for the design and optimization of industrial scale flotation circuits for platinum ores based on pilot plant investigations.

The first objective has been divided into the following categories in this thesis in order to test and validate the proposed simulation methodology.

- Inherent in any model, especially ones describing complicated systems such as flotation are assumptions that simplify the process in order to make predictions. The basis of the modelling methodology assumes that the inherent flotation rate of a particle in a flotation circuit remains constant throughout the circuit i.e.: the floatability (inherent flotation rate) of a particle is conserved around units and nodes within a circuit of flotation cells. This assumption is tested to determine whether the modelling methodology could be applied to the UG2 platinum ore, using a technique developed by Runge et al (1997).

- The flotation rate of a particle depends on many factors. Imaizumi and Inoue (1963) demonstrated that particles of a particular mineral within a flotation cell do not all react with the bubbles in the cell with the same rate. The rate of flotation of the particle depends on its particle size, reagent coverage, degree of liberation, surface composition, particle shape and its density. In order to model the flotation rates of a real ore that is made up of distributions of all these variables, a distributed floatability model is derived for the feed material entering the feed to a flotation circuit. This method lumps together all the factors affecting the flotation rates of the particles into a single distribution of floatabilities. A problem with this method is that the floatability distribution of a feed is not amenable to any direct measuring procedure. A method of regressing this distribution has been developed using laboratory rate batch floats on samples obtained from a continuous plant in combination with plant mass balance data assuming the flotation rates of the floatable particles remain the same throughout the circuit, (Harris, 1997). An objective of this thesis is to determine the floatability distribution of the Platinum Group Metals (PGM) in the ore using this methodology by developing an optimization subroutine in an excel spreadsheet to obtain the various parameters describing this distribution. This method leads to a highly constrained non-linear optimization problem where many possible parameters could give a satisfactory fit to the data. Methods of testing the parameters obtained from the optimizations are used to determine the robustness of this technique. A statistical approach using Monte Carlo Simulation Techniques is used as well as indirect methods of inferring the distribution by knowing characteristics of the feed material via QEMSEM (quantitative evaluation of minerals using a scanning electron microscope) analysis.
- In addition to true hydrophobic mineral flotation, a considerable proportion of the material that reports to the concentrate of a flotation cell is as a result of entrainment of hydrophilic particles with water. It is very important to characterize the chromite recovery kinetics in UG2 ore flotation, as chromite causes problems during the later smelting stages of the concentration process. In fact, the overall design of UG2 flotation circuits is aimed specifically at minimizing the recovery of chromite into the final concentrate. The recovery of chromite is generally considered to occur via an entrainment mechanism. Assuming that the entrainability of chromite varies as a function of particle size, a hyperbolic secant equation developed by Savassi et al (1997), is applied to the pilot plant facilities to determine how well this method is capable of empirically parameterising the entrainment of chromite. A combination of this model and a floatability distribution model is tested to determine whether the kinetics of the unwanted gangue material can be predicted in order to determine accurate grade predictions from the pilot plant tests.
- In order to simulate a flotation circuit it is necessary to have a water recovery model to predict the water flows into the concentrate of a flotation cell. The prediction of water flows determines the residence time calculations used in the entrainment model and is thus the basis of the entire simulation methodology. However, no satisfactory model is yet available that allows this. Harris and Rimmer (1966) stressed the importance of the dependence of the hydrophilic gangue recovery rate

on the water recovery rate in a flotation cell. The partial dependence of the recovery rate on hydrophobic gangue and valuables on the water recovery rate was outlined. The complicated nature of this inter-dependence means that a detailed mechanistic model for the water recovery is necessary or that a consistent methodology is required to overcome the need for a complicated water recovery prediction. A purely empirical model describing the water kinetics, as a function of cell operating parameters such as froth height and volumetric airflow rate is needed that is based exclusively on cell operating conditions. Therefore, a large thrust in the development of a simulation methodology for the pilot plant is to develop an empirical water model of this nature without a need for a comprehensive model.

- A need exists to combine all the models, including the true flotation of valuable minerals, entrainment of gangue and minerals as well as the water model into a simulation algorithm. This simulation algorithm needs to be robust and stable and needs to converge rapidly to a solution. In addition, the simulation algorithm needs to be extended to include flotation circuit simulations. Investigation is required of the overall form of the models to determine whether the combination of these models make physical sense with respect to the resulting grade recovery curves as a function of cell operating parameters such as froth depth and volumetric air flow rates. The limitations of the simulations and the sensitivity of the model parameters to the form of the model need to be investigated to determine the validity of the resulting simulations. A comparison of the simulator's grade-recovery curve to experimental data for a unit cell would provide a basis to determine the accuracy of the combined model prediction. In addition, the output of the simulator can be used to obtain insight into the mechanism behind the flotation of UG2 ores particularly with respect to the froth phase performance and the effective use thereof.

The development of sound kinetic models of flotation provide the basis for predictive simulation. This can then be used as a powerful tool for process optimization as well as in the design of new flotation plants. The second objective of this thesis is to use the results obtained from the above work and to use the simulator to explore some of the various questions asked when designing and operating flotation circuits with particular reference to platinum flotation circuits. An objective of the development of the simulator is to determine whether simulations can be used to exploit the principle behind the "Mintek" circuit to its full capability (see figure 1.1). An opportunity exists to use the simulator to try see if even better plants can be designed once an understanding of the mechanisms in these designs have been developed.

1.4 Structure of the thesis

This section presents the general layout of the chapters in this thesis to facilitate an understanding of the continuity and structure.

Chapter 2: Literature Review

Chapter 2 presents a review of the literature covering all aspects related to the development of a simulator. The aspects pertinent to the understanding of the mathematical tools, models and the work relating to previous work is described.

Chapter 3: Preliminary model development

This chapter presents the initial work using a mini plant facility. This work was not excluded in this thesis as it clearly highlights the progression of the development of the final model presented in chapter four. This chapter logically leads the reader through to the next chapter highlighting the need to upgrade the pilot plant facilities at Mintek in order to obtain accurate measurements of the variables in the model as well as to obtain more reliable mass balance data. The need to include the contribution of entrainment to the overall flotation of the PGM minerals in the ore is also clearly evident from the work done using the mini plant facility.

Chapter 4: Upgraded pilot plant testwork

This chapter describes the work done on the pilot plant facilities at Mintek. The existing pilot plant facilities at Mintek were upgraded in order to make the necessary measurements to satisfactory model and test the proposed modelling methodology. The chapter starts by giving a description of the pilot plant facilities and then discusses the work done in order to obtain accurate control and measurement of the froth height and volumetric air flow rates into each flotation cell. Steps taken to run the pilot plant under a more controlled environment is discussed. This then leads to a description of the different pilot plant campaigns that were carried out in order to model the flotation kinetics of the ore as well as the work done to test some of the assumptions in the modelling methodology. The pilot plant campaigns discussed in the chapter are divided into two phases of work, each phase having their own aims and objectives.

Chapter 5: Pilot plant simulation

This chapter presents the development of an empirical water model which is essentially a curve fit to the observed trends obtained from the froth characterisation tests in chapter 4. After a discussion of the water model used in this chapter it is possible to present the simulation algorithm for the pilot plant treating UG2 ores. This algorithm combines each of the four models describing the PGM, chromite, gangue and water kinetics into a mathematical form that is solved via a nested iteration sequence of successive approximations of the cell and froth residence times for each unit in the simulated circuit. The nested iteration sequence is discussed with reference to its stability and rates of convergence. After developing the algorithm, the chapter progresses to present the output of the pilot plant simulation algorithm using the

preliminary parameters obtained in chapter 4 as well as the water model parameters obtained in this chapter. The output of the simulator is presented for a unit rougher cell, showing the resulting grade and recovery predictions as a function of the froth height and volumetric air flow rates to this cell. The simulator output is discussed with respect to the boundary conditions as well as its form.

Chapter 6: Overall model validation

This chapter is devoted primarily to test the simulation and modelling methodology presented in this thesis. The chapter is divided into three sections, the first of which is a comparison of the simulator's prediction to the two alternative circuit configurations of the work in chapter 4. The next section is an attempt to validate the parameters of the PGM regression methodology. The objective of this was to determine if there is any physical significance in the regressed parameters or whether they merely represent a fit to the experimental data. The final section makes use of the QEMSEM data obtained from the feed stream to the pilot plant during the pilot plant work of chapter 4. This data is used to infer some significance on the PGM floatability mass fractions regressed for the Impala UG2 ore.

Chapter 7: An optimisation study using the simulator

The main emphasis of this chapter is to show some results obtained from the simulator which clearly show the benefits obtained from flotation circuits designed to produce multiple product streams as well as making effective use of the recycle mechanism in re-circulating separation systems. The application of the multiple product recycle mechanism is discussed with reference to common platinum flotation circuit designs.

Chapter 8: Concluding remarks

This chapter presents the major findings of this study.

Chapter 9: Recommendations

This chapter presents recommendations for future work based on the findings and conclusions obtained in this thesis

Chapter 2

LITERATURE REVIEW

2.1 Introduction

The main purpose of flotation circuit simulation and modelling is to be able to predict what the response of a circuit of cells will be if any of the variables that effect the kinetics of the flotation process are changed. The response of a circuit usually includes the change in water, mineral and gangue flows through each cell which are governed by the residence time of the elements flowing through the circuit. Knowing the flows of gangue and mineral, the prediction of mineral recovery and grade is made possible.

There are a very large number of variables that affect the flotation process and many of these interact with one another, hence modelling of a flotation circuit generally encompasses a large number of assumptions in order to simplify the actual mechanisms involved, so as to make accurate predictions. Accurate flotation circuit modelling is important in the modern era as grades of valuable mineral are decreasing in mining, new technologies are being incorporated and economic costs are rising. It is therefore an advantage to be able to use models to predict circuit responses without the need to make physical changes on a real plant. In addition, the ability to predict circuit response will make it possible to optimise existing plants so as to achieve the best recovery and grades from existing infrastructure. This could also greatly facilitate improved control of flotation circuits. Finally, the development of effective predictive simulation procedures would greatly improve the design of new flotation plants, and substantially reduce the time required to commission them.

The structure of this review attempts to address every aspect which has bearing on a flotation circuit simulator. The first section discusses the flotation process in general, discussing the variables affecting the flotation process and the sub-process involved in the effective recovery of a floated particle. The kinetic modelling of the flotation process is also discussed in this section, and this is divided into the various sub-process and component recovery mechanisms inherent in the flotation process, namely: collection zone, froth zone, entrainment, and water recovery models. The next section of the review discusses flotation cells in general describing the mixing regimes and scale-up criteria. This is followed by a discussion of flotation circuits. This section describes the general design of flotation circuits as well as focussing on the modelling and simulation of flotation circuits. Included in this section is a review on the various measurement and circuit analysis techniques used in developing a simulation algorithm of the flotation process. Finally, the last section of the review focuses on platinum ore specifically, the literature associated with the mineralogy and processing of the UG2 reef is described. A brief overview of common flotation circuits and previous studies on the modelling and simulation of platinum ores is presented. This chapter is concluded with a summary of

the findings obtained from the literature as well as the modelling approach selected for use in this study.

2.2 The flotation process

2.2.1 Variables that affect the flotation process

The aim of this section is to illustrate the difficulty in quantitatively describing the flotation process. There are a very large number of variables that affect the flotation process and many of these interact with one another, making a comprehensive model incorporating all the sub-processes extremely difficult, if not virtually impossible to achieve. Table 2.1 was constructed by Sutherland (1955) giving an indication of the number of sub-processes involved in the effective recovery of a particle in a flotation cell. Even the contents of this table cannot be considered comprehensive; Arbiter and Harris (1962) considered that there may be as many as 100 variables affecting the flotation process. In this paper, they encapsulate the difficulties inherent in this process very well when they state:

"We do not intend to be comprehensive in our analysis, but rather to highlight the challenges that face the scientific masochist who chooses to understand froth flotation"

Table 2.1: The variables of flotation, (Sutherland and Wark, 1955)

Raw Materials	Grinding and Classification	Conditioning	Flotation
• Nature of valuable mineral • Secondary valuable minerals (including noble metals) • Associated gangue • Soluble constituents of ore • Degree of oxidation of ore • Oxidation during mining • Quantity of make-up water available	• Fineness of particles and closeness of association • Degree of hardness of minerals and gangue • Relative time for different minerals in grinding circuit • Type of machine used • Type of grinding media used- whether pebbles, steel balls, or steel	• Pulp density • Chemicals added • Order and form of adding chemicals • Temperature • Mean time allowed • Intensity of agitation	• Particle density • Particle size • Pulp density • Temperature • pH value • Circulating load • Time in machine • Geometry of machine • Power intensity • Impeller speed • Degree and type of aeration • Height of froth • Specific chemicals added

-
- Soluble impurities in the water available, especially heavy metal salts
 - Presence of dissolved gases, e.g., H₂S or SO₂ in mine water, if this must be used
 - Degree of alkalinity or acidity of water
 - rods
 - Oxidation during grinding
 - Chemicals added during grinding
 - Pulp density during grinding
 - Rapidity of circulation
 - Time
-

2.2.2 Flotation sub-processes

Froth flotation is dependent on the successful completion of a large number of interdependent sub-processes as described in table 2.1, all of which are required for a mineral particle to report to the flotation concentrate. The overall process of froth flotation will be discussed under the broad terms of particle preparation, particle-bubble interaction and particle transport in the following sections.

2.2.2.1 Particle preparation

Mineral beneficiation processes are based on breaking a mineral aggregate to liberate the mineral grains and then making use of their physical properties to separate the mixture into its components. Therefore, before a desired mineral can be separated from the associated gangue, the ore must be reduced in size to obtain sufficient release or liberation of the desired mineral. Froth flotation is a means of treating a pulp of finely ground ore so that the valuable or desired mineral is obtained in a concentrate that will be amenable to further processing.

Sulphide minerals are naturally less hydrophilic than oxide minerals which constitute the gangue particles, (Fuerstenau, 1982b). For efficient separation of the sulphides from the gangue, the sulphides have to be made hydrophobic. In order to do this, collectors are added which interact with the surface of the mineral. A collector is a molecule consisting of a polar end that attaches to the mineral surface and a non-polar end that attaches to an air bubble, although other mechanisms of collection can occur. Some minerals require activation by metal cations before suitable hydrophobicity is created upon addition of a collector. A common activator used in the flotation of platinum ores is copper sulphate (CuSO₄). Common plant practice is to add CuSO₄ as the first reagent

to contact the freshly milled ore before addition of collectors which results in better separation of the minerals from the gangue.

The nature of the gangue minerals in the flotation system often promotes the recovery of these minerals either by true flotation or by entrainment, causing lower grades of valuable minerals in the concentrate. Chemicals called depressants are frequently required to change the nature of the gangue in order to minimise gangue recovery to the concentrate. In general, two different classes of depressants exist, those that enhance the hydrophilic nature of the gangue by preventing the formation of hydrophobic species on the gangue surface and those that cause the gangue minerals to aggregate together causing them to be too large to be entrained and therefore remain in the pulp. Typical examples of depressants are starch derivatives, guar gums, inorganic salts, carboxymethylcellulose (CMC) and polysaccharides.

Finally, chemicals called frothers are added to the pulp of a flotation cell in order to maintain the size distribution of bubbles. The frother reduces the surface tension of the pulp, facilitating the formation of a stabilised froth. A frother is usually a heteropolar surface active agent that adsorbs onto air bubbles with the non-polar end of the molecule while the polar end extends into the liquid surrounding the bubble. The adsorbed molecules are in dynamic equilibrium with the surrounding liquid, and the degree of adsorption is a function of the decrease in surface tension of the liquid, (Booth and Freyberger, 1962).

2.2.2.2 Particle-bubble interaction

After a suitable conditioning period air is introduced into the flotation cell and dispersed into small bubbles by the impeller. In order to be collected, hydrophobic particles have to collide with air bubbles resulting in the thinning and rupture of the liquid film between the particle and the bubble. The effective collision between a particle and an air bubble is affected by the size of the particle, the size of the bubble and aeration rate, and the degree of mixing in the cell.

Gaudin (1957) was the first to state that the probability of a direct encounter between a particle and a bubble varied according to the particle size. Particle size plays a large role in the selectivity of the flotation process. Trahar (1981) stated that the significance of particle size and its consequences in flotation research, in plant operations and in control schemes has been underrated. Coarse particles in a flotation pulp tend to be purely dependent on hydrophobic adhesion as they are too heavy to be entrained into the concentrate. These particles show a lower tendency to be transported to the froth by gas bubbles due to their larger weight. Sutherland (1948) argued that fine particles tend to be less dependent on surface properties, having a lower probability of effective collision with a bubble. These particles also have a greater tendency to be entrained into the concentrate, (Smith and Warren, 1989). Therefore, it is widely accepted that there is an intermediate size interval for which the flotation rate reaches a maximum, (Lynch et al, 1981). Particle size recovery curves have a characteristic maximum

describing the average particle size for maximum recovery from a flotation cell. Figure 2.1, from Trahar (1981) illustrates how the recovery-size curve can be divided into three regions. The region below typically $5\text{--}10\text{ }\mu\text{m}$ consists of fines which are difficult to float and more difficult to separate. The region from typically $10\text{ }\mu\text{m}$ - $70\text{ }\mu\text{m}$, known as the intermediate region, contains the particles which are usually the most floatable. The region above typically $70\text{ }\mu\text{m}$, defined as the coarse particle region is where flotation may be easy or difficult according to the mineral and conditions employed. It is not usually the case to assign exact particle size ranges for the different regions but to rather refer to fine, intermediate and coarse particle behaviour due to the ill-defined and variable locations of these three regions, (Trahar, 1981).

The efficient flotation of fine particles has probably received the most attention with respect to particle size effects on flotation. Many researchers have investigated methods of increasing the recovery of this "slimes" fraction, eg. Meloy (1962), Mehrotra and Kuper (1974), Trahar and Warren (1976), Jameson (1977), Ahmed and Jameson (1985), Flynn and Woodburn (1986, 1987), Subrahmanyam and Forssberg (1990) and more recently Deglon (1997). The general consensus reached is that the flotation rate of the fines fraction is much slower than the other size fractions, primarily as a result of the decreased probability of collision between particles and air bubbles as the particle size is reduced.

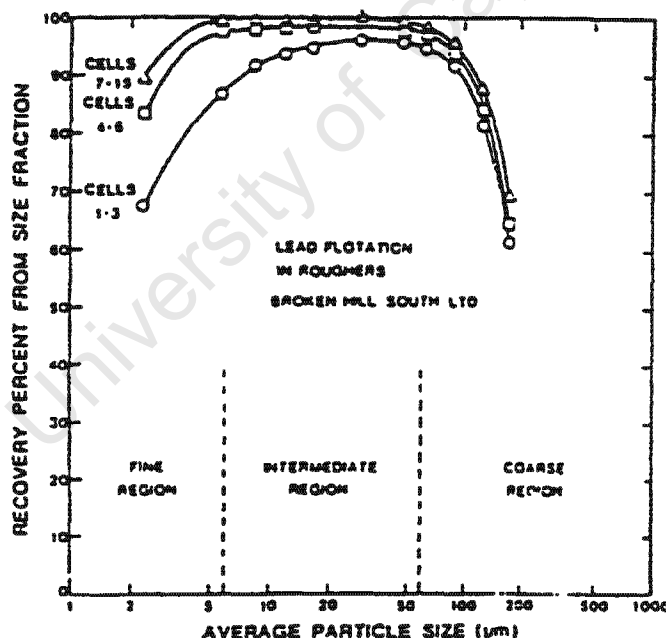


Figure 2.1 : Recovery as a function of particle size, (Trahar, 1981)

The creation of bubbles of a specific size in a flotation cell is very important. Gaudin et al (1931) realised that large bubbles would have less surface area to which particles can attach. However, the creation and dispersion of small bubbles is expensive in terms of mixing energy requirements. It is also possible that bubbles could be too small to provide a sufficiently large buoyancy force required to lift the particles out of the pulp. Thus, an optimum bubble size to particle size relationship exists in the pulp of a flotation

cell. The degree of mixing also needs to be at an optimum. Enough energy is required to disperse the particles and bubbles and to bring them into constant contact with each other with a force large enough to overcome the energy barrier caused by the liquid lamella surrounding the bubbles. However, the mixing must not be too vigorous as this may cause particle-bubble aggregates to break-up. Over vigorous mixing may also destabilise the froth zone, promoting bubble break-up in the froth zone and causing particles to detach from bubbles in the froth.

2.2.2.3 Particle transport

The recovery of particles into the concentrate of a flotation cell can be considered to occur via two distinct mechanisms, namely true flotation and entrainment. These two mechanisms are described below.

2.2.2.3.1 Attached particles (true flotation)

True flotation occurs when a particle attached to a bubble surface is transported from the pulp into the concentrate launder, without the particle detaching from the bubble. After a successful collision, particle-bubble aggregates have to be both sufficiently buoyant to rise through the pulp and sufficiently stable to endure any disturbances in the pulp. The aggregate must not be broken up by collision with other particles, or with other aggregates. When the aggregates and unladen bubbles reach the pulp-froth interface, some water is drained out of the bubble lamellae. The attachment must be strong enough to withstand this mechanical draining. As the aggregates reach the pulp-froth interface, elutriation can take place. That is, the aggregate could break and the mineral could fall back into the pulp. It is possible that a detached hydrophobic mineral could re-attach to another air bubble in the froth and be collected successfully. A detailed analysis of the froth phase on the recovery of attached particles is described in section 2.2.3.2.

2.2.2.3.2 Unattached particles (entrainment)

Simultaneously, particles not actively collected by bubbles can pass into the froth as a result of entrainment with water. This occurs when small particles of unwanted hydrophilic gangue can be entrained between the bubbles and carried into the froth phase. The entrainment of particles is unselective and it is possible to recover poorly liberated or fine valuable (hydrophobic) minerals by entrainment. The degree of entrainment increases with the recovery of water (Wrobel, 1953). Particles entrained in the bubble lamella could also drop back into the pulp phase with the drainage of water from the froth phase. The models used to describe the entrainment mechanism in the flotation process are discussed later on in section 2.2.3.3.

2.2.3 Modelling the flotation process

Modelling of any system can be divided into four distinct approaches, namely mechanistic, semi-empirical, empirical and analogue. The first approach attempts to describe analytically the actual phenomenon of the process. For complex processes such as flotation, the use of this method is generally only useful in describing highly simplified systems under controlled laboratory conditions. An example is the modelling of single particle bubble interactions in a quiescent environment. Empirical methods are generally based on fitting observed data from an industrial plant to an equation of a preconceived form. This method is often used primarily to control or optimise an industrial scale process but nothing in this approach gives any information or understanding of the actual mechanisms involved in the process. Hence the use of such a model is limited to an individual process and a new model would have to be developed every time the process is changed. The semi-empirical method is generally a combination of the analytical and empirical approaches with parameters containing some physical significance, combined with some empirical parameters derived purely from experimental data. This method is the most widely used, particularly for complex systems such as flotation. Finally, the analogue approach to kinetic modelling involves applying methods used in other processes to describe the system under consideration, such as the use of principles derived from chemical kinetics to describe the flotation process. To some extent, this method can be considered to be an extension of the semi-empirical method described above. Many approaches to flotation modelling have been attempted to date. They include all of the above categories with particular attention being given to the analogue and semi-empirical approaches. Few attempts have been made with the analytical approach of flotation modelling due to the large number and extreme interaction of the process variables. These different attempts to describe the flotation phenomenon will be discussed under the relevant categories. The simplification of the models that are described should be appreciated, with many of the models describing the overall process using only single parameter models.

2.2.3.1 Collection zone models

The following overview of the collection zone kinetics in a flotation cell is based on review papers by Inoue et al (1984) and Fichera and Chudacek (1992).

2.2.3.1.1 Probability model

This analysis of the flotation system considers the micro-processes in a flotation cell as the basis of a probabilistic model. The removal rate of particles or rate of flotation from the pulp has already been discussed and was considered to be a consequence of four sub-processes;

1. Collision between particles and bubbles (P_c)
2. Adhesion of particles to bubbles (P_a)

3. Detachment of particles from bubbles (P_d)
4. Chances of recovery in the froth phase (P_f)

If all collisions between valuable minerals and bubbles resulted in the transport of particles to the concentrate then the probability of collision would represent the fractional recovery of the valuable mineral. It is logical to assume that not all collisions result in the particles adhering to the bubbles due to insufficient energy requirements for attachment. Hence only a fraction of the collisions P_a results in adhered particles to bubbles, ie: $P_c \cdot P_a$. Now, of the adhered particles, some particles have a probability of detaching from the bubbles due to turbulence effects, and hence $(1 - P_d)$ represents the fraction of particles that do not detach. Therefore the total recovery becomes $P_c \cdot P_a \cdot (1 - P_d)$. Statistically, the chances of a particle in a flotation pulp to be recovered to the concentrate launder and surviving the froth phase is given by the multiplication of the probabilities for the different sub-processes:

$$R = P_c P_a P_f (1 - P_d) \dots\dots\dots 2.1$$

This analysis was first put forward by Schuhmann (1942) who considered P_a and P_c , later Sutherland (1948) included P_d and Tomlinson and Flemming (1963) introduced the probability of being retained in the froth phase. Various attempts at giving a quantitative description of these probabilities have been made. None of the models developed have been used in real flotation plants due to the complexity implicit in all their derivations.

2.2.3.1.2 Single rate kinetic models

The analogy of relating the kinetics of a chemical reaction to flotation kinetics was first attempted by Zuniga (1935). Chemical reactions are concerned with the interaction of molecules and atoms, while flotation concerns the interaction of bubbles and particles. Assuming that this can be considered a first order process, the rate of disappearance of a valuable component in a cell can be described by equation 2.2 according to the chemical kinetics analogy.

$$\frac{dC}{dt} = -kC \dots\dots\dots 2.2$$

Schuhmann (1942), as well as other researchers such as Imaizumi and Inoue (1965) showed that the flotation rate in a continuous cell can be well described using first order kinetics. It was later shown that implicit in the assumption of this model is that the feed solids in the cell are all homogeneous, or equally floatable. Experience with using this model has shown that the time-recovery profile predicted by this model gives a good fit to experimental data in the first minute, but deviates from the first order assumption after this time. Arbiter and Harris (1962) believed this was due to a portion of the feed

that will not float over prolonged periods of time and included a C_{∞} parameter into equation 2.2 to correct for this phenomenon.

2.2.3.1.3 Distributed rate kinetic models

The simple first order kinetic model discussed above defined a rate constant which was relatively crude, with the assumption that the feed solids are all identical and therefore all float at the same rate. Obviously this is an extreme simplification as the feed solids to a flotation cell consist of particles with varying sizes, densities and extents of liberation. As a first attempt to account for this, the C_{∞} parameter was introduced. It was never really concluded whether C_{∞} actually represented the un-floatable fraction of particles in a cell and many researchers questioned this assumption but still used it as a "fudge factor" to represent their results. The next section describes the different attempts to describe the flotation rate as a more complex function incorporating the variability of particle flotation rates in the feed.

a. Distributed rate models based on size

Particle size models account for the differences in the flotation rate constants for particles of different size by assuming a different rate constant for each particle size fraction. Imaizumi and Inoue (1965) were the first to introduce the concept of a distribution of first order rate constants. This arose from a consideration of the probable causes of observed deviations from first order kinetics as discussed in section 2.2.3.1.2. The range of recoveries was therefore distributed according to the size distribution of the feed. For (m) discrete size fractions, they proposed the following equation:

$$\frac{dC_i}{dt} = -\sum_{i=1}^m k_i C_i \dots\dots\dots 2.3$$

where

- C_i is the concentration of each size fraction (i)
- k_i is the first-order rate constant associated with each size fraction (i)
- m is the number of fractions used to describe the size distribution
- t is the cumulative flotation time

This model is superior to the simple first order model which did not account for size effects. Tomlinson and Flemming (1963) were the first to experimentally test and confirm the accuracy and validity of the size distributed model. They concluded that individual size fractions followed first order kinetics, but a combined mixture of different size fractions gave an impression of non first-order kinetics. Instead of using discrete size classes, it is possible to use a continuous distribution function to describe the

range of size classes. Sutherland (1981) introduced the concept of a distribution function of particle size (r) and described the distribution by a density function $F(r)$:

$$\int_0^{\infty} F(r) dr = 1$$

$$\frac{dC_r}{dt} = \int_0^{\infty} k(C_{\infty} - C_r) F(r) dr \quad \dots\dots\dots 2.4$$

However, owing to the difficulty of finding a function to describe the size distribution it was considered simpler to calculate the flotation rate by splitting the feed into size fractions as described by equation 2.3, (Sutherland, 1981). The assumption of independent rate constants for particles of different size for flotation is reasonable as, in the absence of complex interactions or dependencies in the flotation cell, particles should respond as separate and independent species.

b. Distributed rate models based on floatability

With the advent of models using distributed first order rate constants, important characteristics of a flotation feed material have received considerable attention. In flotation, the property that is exploited to separate a mineral from an ore is the ability of a mineral particle to form a stable attachment to a bubble. This "ability" is commonly referred to as the floatability of the mineral particle. Particles of a particular mineral within a flotation feed would not be expected to attach to bubbles in a flotation cell at the same rate. The flotation rate of a particle would be expected to be influenced by its particle size, reagent coverage, degree of liberation, surface composition, shape, and density. It can therefore be appreciated that the response of a heterogeneous feed to a flotation circuit must consist of a distribution of individual flotation rate constants which represent complicated interactions of size and liberation. Figure 2.2 shows a theoretical floatability distribution of a feed to a flotation circuit. The flotation rate of particles range from zero to a maximum, with the proportion of material with each flotation rate being a function of the area under the curve.

Therefore, the concept of a distribution of floatabilities lumps together all factors affecting the floatability of a particle and proportions the mass of the feed into various fractions depending on the average floatability of that fraction. The overall flotation rate is then calculated as the sum over the feed of all particle floatabilities. Figure 2.3 gives an indication as to how the ore can be split into different floatability classes showing the underlying principles of this model.

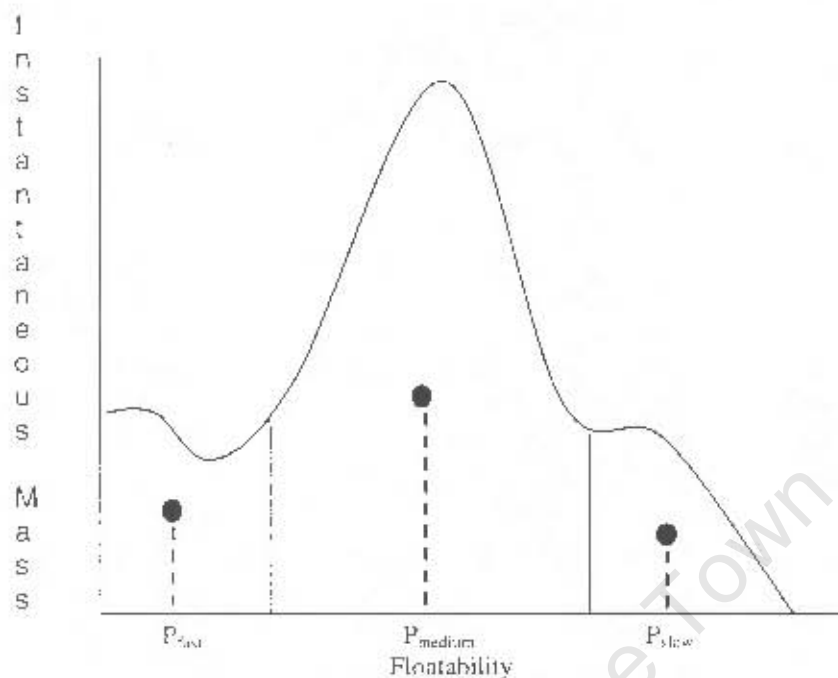


Figure 2.2: Theoretical floatability distribution

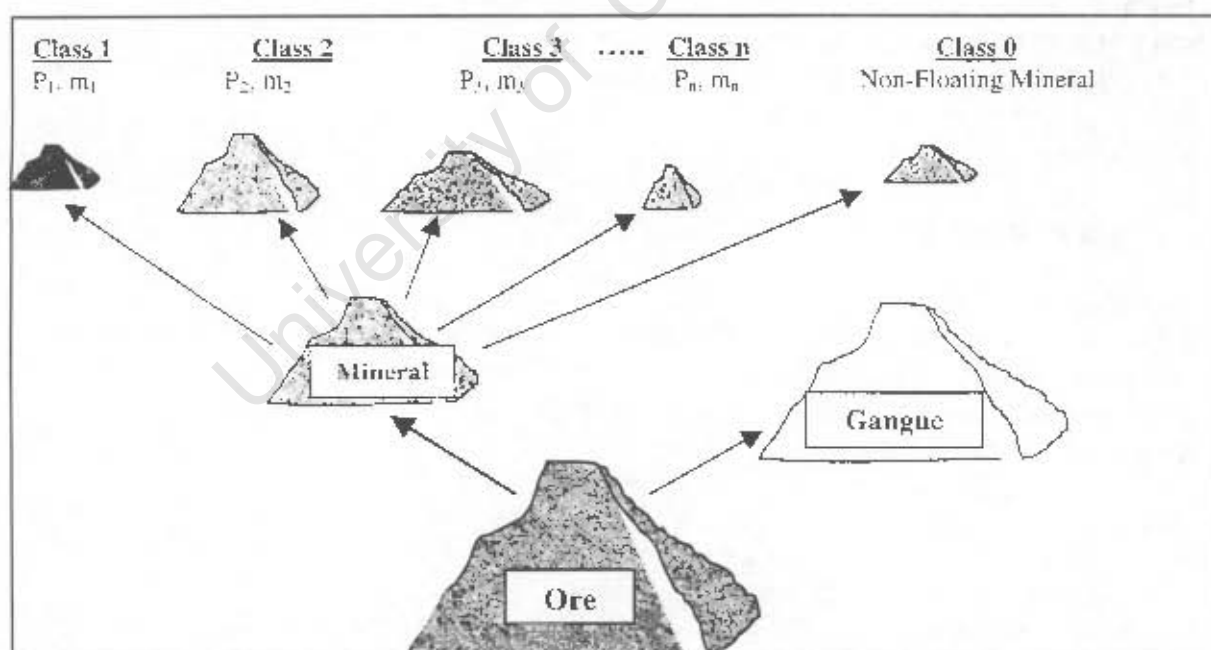


Figure 2.3: The separation of a mineral in an ore into floatability components

The simplest form of this model is the application to a binary system where the distribution of floatabilities can be represented by two particle floatability classes such as floatable/non-floatable or fast/slow floating. Imaizumi and Inoue (1965) applied a graphical solution to systems with distributions of floatabilities while Woodburn and

Loveday (1965) proposed a parametric rather than a graphical solution to describe the distribution of rate constants. The assumption made was that the distribution of rate constants (floatabilities) has the form of a gamma function (Γ) described by equation 2.5.

$$G(k) = \frac{b^{a+1} k^a e^{-bk}}{\Gamma(a+1)} \dots\dots\dots 2.5$$

This method is purely empirical and reduces to a curve fitting exercise, somewhat limiting its usefulness. Inoue and Imaizumi (1968) and Harris and Chakravarti (1970) later considered different forms for the gamma function while Fichera and Chudacek (1992) found that the discrete form to have greater physical significance than either the continuous or gamma functional forms. Woodburn et al (1984) criticised the floatability method as being too simple as it cannot differentiate between or determine the effects of the influencing parameters on the flotation rate such as particle size and liberation. It must be remembered that these equations hold for discrete components of the same floatability fraction, and the overall recovery would be the summation of all floatability components in the system. An assumption that the different components do not interfere with one another during flotation is made and is the norm at present as no means of accounting for an interaction between the different floatability fractions have been identified to date. In conclusion, particle floatability models allow for flexibility in the analysis of flotation kinetics but are applicable to complex systems where extreme variations in mineralogy and surface properties justify their usage.

c. Distributed rate models based on liberation

This type of model includes ore liberation characteristics to differentiate between the floatability components. This is considered to be the most accurate means of characterising the feed to a flotation circuit as the flotation phenomenon is a surface process which should be able to be defined explicitly by liberation data. The models discussed previously only make the distinction between mineral by size, with the mineral component obtained from chemical assays. This gives the total amount of the mineral present, including the fraction of mineral locked within the ore particles, ignoring the fact that only the minerals liberated on the surface of the particles contribute to the hydrophobicity of the particle. Unfortunately the use of liberation based kinetic models is restricted as the required data is both difficult and very expensive to obtain.

Zaidenberg et al (1964) derived a model which considered mineralogical composition of particles in addition to size and floatability distribution. The particles were separated into discrete size classes, each of these were further split into sub-classes determined by mineralogical composition. The floatability distribution was then determined for each mineralogical sub-class. Woodburn et al (1970) noted that this method was not readily amenable for the analysis of flotation experiments. This was due to the increased

complexity of this model over the more simpler models described above. Another problem was that at the time of Zaidenberg's work there was no method readily available for the quantitative determination of the mineralogical composition within each size class.

King (1973) presented a complex mechanistic model for the design and control of flotation plants based on Zaidenberg's distributed parameter model. The model allows for a variation in feed properties and grind. In addition, the model provides for a distribution in rate constants, particle size and mineral fraction. Three distinct properties of the particles that influence the rate of flotation significantly were considered; particle size (D), mineral composition of the particle (g) and the surface activity (k) and were distributed over the individual particles in the feed. The mineral composition (g) is a vector of values representing the fraction of each mineral type in the particle. Microscopic counting of particles to quantify their mineralogical distribution was done using QEM*SEM, a system whereby particles are characterised by scanning electron microscope image analysis allowing for the quantitative determination of the distribution of particle characteristics such as mineralogical composition and degree of liberation of particles within size classes.

The active sites model described by Savassi (1995) is another version of this type. The main feature of this model is that the probability of attachment is split up into chemical and mineralogical effects. The chemical effects are included through the use of an activation parameter which gives an indication of the strength of a particle-bubble bond. This parameter is purely empirical and has meaning only to a fixed chemical environment, although the model does allow for the possibility of modelling the changes in surface chemistry with time in a circuit. The main assumption of the active sites model is that the most liberated particles of a given size interval should float faster and have a higher ultimate recovery. The feed to a circuit is therefore divided into a number of floatability classes, where particles with the same surface grade and size are assigned the same class. Particles in each class are assigned the same probability of attaching to a bubble. The approach further assumed that, thermodynamically there exists a minimum surface grade for a particular size particle below which the particle cannot be floated. This gives rise to an inactive fraction of the feed (equivalent to the C_{∞} parameter described previously). In addition, any change in surface chemistry (eg. by oxidation) across a bank of cells is considered to be evenly distributed amongst the floatability components and independent of particle size.

The above methods of modelling the flotation process definitely show considerable potential but unfortunately require considerable time, effort and expense. Methods of obtaining quicker and more inexpensive liberation results are being developed at present which, once established, could greatly facilitate the use of models of this type. It should be noted, however, that there are still many unresolved issues in terms of the application of this approach to highly complex ores, where a large number of minerals occur in complex associations, such as is the case with platinum bearing ores.

2.2.3.1.4 Higher order models

Higher order kinetic models consider that the flotation rate can be described by an equation of the form:

$$\frac{dC}{dt} = -kC^n, \quad n > 1 \dots\dots\dots 2.6$$

where n = order of flotation rate

The order of the rate equation (n) has been a subject of debate over many years and has been found to vary between 0 and 2.8 depending on both the researcher and the system being investigated. The arguments for non first-order expressions have been found to be largely empirical and have generally been rejected in favour of the first-order expression, (Fichera and Chudacek, 1992). Volkova (1946a) suggested that at high pulp densities multi-collisions occurred so that flotation would tend towards a higher order process. Yet later this was rejected in favour of an argument originated by Tomlinson and Flemming (1963) who suggested that at high pulp densities the order would vary from one to zero (transitional and inhibited flotation) due to overloading of bubbles. This led to their eventual rejection of higher order models in favour of the first order rate. The next section presents a model used to describe inhibited flotation kinetics as a means to provide an explanation to experimental studies that showed deviations from the first order rate assumption.

2.2.3.1.5 Inhibited flotation kinetics

Previous researchers noted a zero order flotation rate believed to be attributed to the lack of adequate bubble surface which resulted in a constant flotation rate, independent of the concentration of valuables in the cell. It was this observation that made researchers realise that realistic models should allow for an equilibrium bubble surface loading and that the kinetics of flotation could be inhibited under certain conditions.

As air bubbles pass through a pulp, the particles attach and detach from a bubble and an equilibrium loading can be reached. At this equilibrium the bubble surface might be saturated with particles. Under bubble surface area limiting conditions, the valuable recovery rate becomes independent of the concentration of mineral in the pulp. At this point the bubble concentration or bubble surface area determines the recovery rates. Tomlinson and Flemming (1963) introduced the concept of free and inhibited flotation. Under conditions of high pulp density, high feed grade or low aeration rate, inhibited flotation can occur, particularly in rougher or cleaner cells. Conditions in a flotation circuit have been found to change from free to inhibited flotation across a bank in a continuous plant or with time in a batch cell. It has been shown that the rate of bubble loading undergoes an exponential decay as the free bubble surface area decreases, (King, 1973). Thorne et al, (1976) noted however that the concept of bubble overloading should not be confused with froth overloading effects. Froth overloading can arise for a

variety of reasons, such as insufficient frother and air to support the froth. Under such conditions the froth volume to launder transfer becomes the rate determining step. This type of phenomenon was found to occur in the first flotation cells of the rougher and cleaner banks while attempting to model an industrial UG2 platinum circuit, (Theron, 1998).

The simplest attempt to describe froth overloading was by Whiten (1972), using a general first order flotation rate assumption, but adjusted the rate constant to give an effective rate constant ($k_{i,eff}$) defined as:

$$k_{i,eff} = \frac{k_i}{h} \dots\dots\dots 2.7$$

$h \rightarrow 1$ at low mineral loads

$h \rightarrow \infty$ as $\sum C_i$ increases

This value of h is defined as an overloading factor, equivalent to a resistance (analogy to mass transfer resistance):

$$h = \frac{1}{(1 - \beta \frac{\sum C_{vi}}{V})} \dots\dots\dots 2.8$$

V effective volume of the cell

β parameter which limits the maximum flow rate of valuables to the concentrate,
ie: $\beta = V / \sum C_{vi}$

C_{vi} concentration of valuable component (i) in the pulp

This model has been extended to modelling gangue recovery as well as defining an overloading factor to discrete size fractions, (Whiten, 1972). Frew and Trahar (1982) showed that the rate coefficient varied not only from bank to bank in a flotation circuit but also from cell to cell, which was attributed to the variation in pulp density (especially in cleaner cells), causing a change in the bubble loading. A more thorough approach to modelling this effect was undertaken by King and coworkers (1973, 1978). Generally the flotation rate was shown to be first order up until a saturation point. On this basis, it is therefore reasonable to expect a relationship between the flotation rate constant and the amount of bubble loading. To describe this, the degree of bubble loading was based on Langmuir's theory of gas adsorption onto solids, except in reverse ie: adsorption of solids onto gas bubbles. This theory has a number of assumptions:

- The surface of a solid is made of spaces. All the spaces are identical
- Gas molecules next to each other do not affect one another
- Rate of desorption is directly proportional to the concentration on the surface

Without going into the details of this model which can be obtained elsewhere (Smith, 1984) the fraction of the bubble surface loaded with solids is given by:

$$\theta = \frac{K_{ore} * X}{1 + K_{ore} * X} \dots\dots\dots 2.9$$

where

- X concentration of solids in the pulp
 K_{ore} k/r
 r rate of detachment from a completely covered surface
 k rate constant for attachment of particles to bubbles

For a saturated froth, which is equivalent to having a maximum flow rate of concentrate ($Q_{l_{sat}}$);

$$Q_{l_{sat}} = N_b \theta \eta = kX \dots\dots\dots 2.10$$

- η maximum volume of solids a bubble can carry
 θ bubble loading
 N_b number of bubbles

From this an approximation of the rate constant for attachment of particles to bubbles is given by equation 2.11.

$$-\frac{\partial X}{\partial t} = X * k = \frac{N_b * \eta}{\left(\frac{1}{K_{ore}} + X\right)} * X \dots\dots\dots 2.11$$

Using this model, at low solids concentrations; $X \ll 1/K_{ORE}$ equation 2.11 reduces to a first order equation. At high mineral concentrations $X \gg 1/K_{ORE}$, equation 2.11 reduces to a zero order function: $-dX/dt = N_b \eta$. The maximum volume of solids a bubble can carry (η) does not necessarily occur at total surface coverage and is known to be a function of bubble retention time, degree of agitation, bubble diameter, mineral type and size. The problem with this model is that it contains a large number of parameters that are difficult to obtain in an industrial environment and hence it is impractical to use in its present form.

2.2.3.1.6 Decoupled kinetic models

In order for a model to be used to simulate the flotation process, the rate constant must be calculated based on practical, measurable separation criteria. The models described previously all relate the flotation rate to the mineral properties such as the number of

size fractions or floatability classes. The flotation rate is also dependent on the characteristics of the flotation unit, including cell and impeller geometry, power input, gas dispersion and froth depth. The flotation rate constant would therefore vary from cell to cell and even from particle to particle in a flotation circuit. The models described below attempt to decouple the flotation rate constant into independent cell variables and mineral variables for application during simulation.

a. Air factor

The simplest attempt to decouple the rate constant was to separate the contribution of the air flow rate on the overall flotation kinetics. Woodburn and Wallin (1984) proposed a model in which the first order rate constant can be decoupled into kinetic effects on the basis of aeration rate (modelled by a multiplier χ) and equilibrium effects (modelled by K) that represents the inherent distribution of a solid species between an air interface and a continuous pulp phase:

$$k = \chi K \dots\dots\dots 2.12$$

where

- K is an empirical factor that accounts for the particle packing on the bubble surface
 χ is a multiplier described by:

$$\chi = \alpha Q \dots\dots\dots 2.13$$

- α is a parameter describing the froth stability effects
 Q is the volumetric air flow rate

Although partitioning the flotation rate constant into aeration effects, froth stability and flotation characteristics of the solid species is sound, it is often impossible to study the effect of operating conditions and froth phase variables, unless these parameters are expressed in terms of measurable flotation variables. Jameson (1985) attempted to describe the influence of the variables affecting the flotation rate constant and proposed the following equation:

$$k = \frac{3}{2} E_k \frac{6J_g}{d_b} \dots\dots\dots 2.14$$

where

- J_g is the superficial gas velocity (cm/sec)

d_b is the bubble size (mm)
 E_k is the collection efficiency

Yoon and Luttrell (1989) defined collection efficiency, E_k as:

$$E_k = A \left(\frac{d_p}{d_b} \right)^n \dots\dots\dots 2.15$$

where

d_p is the particle size
 A, n are empirical parameters

Although this model is based on measurable variables, it does not separate the influence of the froth phase on the overall rate constant.

b. Bubble surface area flux

The most recent advancement made to quantitatively relate flotation kinetics to flotation cell hydrodynamics was the work of Gorain et al, (1995). Investigations conducted at Hellyer, Scuddles and Mt Isa mines in Australia showed a strong linear correlation between the collection zone flotation rate constant (k_c) and the bubble surface area flux (S_b) in a flotation cell.

$$k_c \propto S_b \dots\dots\dots 2.16$$

$$S_b = 6J_g/d_b \quad \text{(for spherical bubbles)}$$

where

S_b bubble surface area flux or interfacial area rate; *(which indicates the availability of air in the collection zone ($\text{mm}^2/\text{mm}^2.\text{sec}$))*
 J_g ratio of air flow rate to cell cross sectional area or superficial gas velocity (cm/sec)
 d_b average bubble diameter (mm)

Also of considerable interest in this work was their observation that this relationship seemed to be independent of the type of impeller used, the type of flotation cell, the size of the flotation cell as well as the impeller speed. The collection zone rate constant represents the rate constant when no froth phase is present. This relationship between k_c and S_b indicates that the greater the available surface area of bubbles, the greater the chance of particle-bubble contact and therefore the higher the flotation rate constant. If the bubble surface area flux is a good measure of the gas dispersion or hydrodynamics in a flotation cell, particles in a cell should behave the same way when presented with the same S_b , irrespective of the size of the cell in question. Particles in a

cell are only concerned with the amount of available bubble surface area. Therefore if S_b is the same in two cells of different sizes treating the same ore in the same chemical environment, the collection zone flotation rate would be expected to be the same in both cells, (Gorain et al, 1995). This implies that it should be possible to determine the collection zone rate constant of a larger flotation cell treating the same ore in the same chemical environment if the bubble surface area flux is known in the larger cell. The value of the floatability, P (or slope), depends on the mineralogy of the ore, the liberation of the minerals, the size and degree of collector coating of the particles etc. The value of S_b depends mainly on the gas dispersion in the cell. The mineral variables have been effectively separated from the cell variables in the evaluation of the factors contributing to the flotation rate. If the change in slope of this relationship can be quantified with the variation in chemical conditions, this could be a starting point for modelling the influence of chemical additions on the flotation rate, (Gorain et al, 1995). Figure 2.4 describes the linear relationship between the collection zone rate constant and bubble surface area flux in a flotation cell for different mineral floatabilities (P_i).

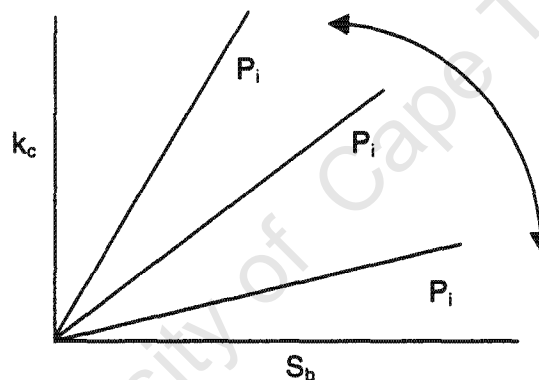


Figure 2.4: Linear relationship between the collection zone rate constant and bubble surface area flux in a flotation cell for different mineral floatabilities (P_i)

So far the kinetics of the collection zone have only been considered. The effect of the froth phase on the overall kinetics is generally to decrease the overall rate constant of flotation due to drop back of particles back into the collection zone. The next step Gorain et al (1997) took was to study the effect of froth depth on the above relationship. When the same tests were performed at varying froth heights, it was noticed that the mineral floatability, P , was not constant, even if the chemical environment in the flotation cell had not been altered. P decreased as the froth depth was increased which was clearly contrary to the definition of P (indicative of the minerals inherent floatability). It was therefore considered necessary to include an additional parameter correcting for the froth phase characteristics on the overall flotation rate constant. Equation 2.17 shows how the rate constant can be decoupled into mineral, hydrodynamic and froth effects.

$$k_{o,i} = P_i S_b R_f \dots\dots\dots 2.17$$

- P_i is the mineral floatability of component (i)
 S_b is the bubble surface area flux (*bubble surface/time/cell area*)
 R_f is the froth recovery (*ratio of the mass of mineral reporting to the concentrate to the mass of mineral reporting to the froth-slurry interface*)
 $k_{o,i}$ is the overall flotation rate constant associated with floatability component (i)

It must be emphasized that the linear relationship is only considered to be valid for the collection zone of the flotation cell. The effect of the froth on the overall performance of the cell is included through the froth recovery parameter in the above equation. Research into the effects of the froth zone on this relationship is still undergoing investigation and the literature associated with this is discussed in section 2.2.3.2. Deglon et al (1999) pointed out that the relationship between k_c and S_b presented above implies that the flotation rate constant should be inversely proportional to the bubble size to the power of 1 which is not consistent with the flotation literature which predicts a power of between 1.5 and 3. Finally, the effect of turbulence or power input is an important variable that is known to effect this relationship. This area is still relatively poorly understood, and no correlation has yet been established to define the relationship between floatability and turbulence effects.

c. Attachment-detachment model

A new flotation kinetic model has been proposed by Deglon, (1997) in order to describe certain consistent trends in flotation rate data for mechanically agitated flotation cells, termed the attachment-detachment model. This model decouples the two independent processes of particle-bubble attachment and detachment in the pulp zone of a flotation cell. The attachment rate constant was defined so as to correspond directly to the flotation rate constant while the detachment rate constant was defined so as to be independent of the gas phase. It was concluded in this work that using separate rate constants improves the flexibility and insight into the analysis of flotation kinetics but can lead to the misinterpretation of simple kinetics. Another advantage of using separate rate constants is that it is possible to determine the relative contributions of the impeller zone and the bulk tank to the overall attachment and detachment rate constants.

2.2.3.1.7 Discussion

Provided the heterogeneous characteristics of mineral particles to a flotation cell are accounted for, evidence from the literature suggests that flotation can best be considered to be a first order process. The effect of particle size is the most dominant factor influencing the rate of flotation. The influence of other variables on the floatability of a particle can be accounted for collectively in the form of a floatability spectra as a function of mineral or mineral by size class. This method is often used as the factors

influencing the floatability of a particle are difficult to isolate and can therefore not be accounted for individually. The distributed rate constant based on liberation shows considerable promise if the requisite data is available, as flotation is a surface process which could well be explicitly related to liberation data. However, as Inoue et al (1965) remarked, it is doubtful, due to the complexity of the flotation process, that all parameters can be completely understood in the goal of creating the complete flotation model. For fairly pure mineral systems, the double distributed parameter model (floatability by size class) is the most useful method from a practical point of view. The most recent advancement to modelling the flotation process is the recognition of the importance of the froth phase as a distinct and separate entity from the pulp zone in a flotation cell. It is believed that the complex interactions occurring within the froth phase is one of the major contributors to the incomplete understanding of the flotation process. Hence, methods for accounting for the froth phase are discussed in the next section.

2.2.3.2 Froth zone models

It is widely accepted that the flotation process involves two distinctly different zones, namely the pulp phase and froth phase. They influence the overall performance of the flotation process in different ways but are highly inter-connected via a continuous recycle of material between the two phases. Many of the earlier attempts at modelling the flotation process neglected the effects of the froth zone, rendering the scale-up of bench and pilot scale data to large scale cells extremely unreliable. Such models are also highly limited in terms of their use in circuit design and optimisation. Addressing this limitation has been the focus of many recent studies reported in the literature. Studies by Contini et al (1988), Falutsu and Dobby (1989), Finch and Dobby (1990), Savassi (1998) and Mathe et al (1998) have highlighted the importance of decoupling the influence of the pulp phase performance on the overall flotation recovery from that of the froth phase. It is in this light that the froth zone will be discussed in this section. The aim is to recognise the different methods that have been adopted to incorporate the effects of the froth zone on the overall performance of the flotation process. This section will begin by giving a brief qualitative description of the different sub-processes occurring in the froth phase and then some of the different attempts to model the froth phase will then be considered.

2.2.3.2.1 Qualitative description of the froth

The main function of the froth phase in the flotation cell is its ability to act as a separator of hydrophobic particles attached to bubbles from entrained gangue and hydrophilic mineral particles, with water acting as a transportation medium. Three major forces compete within the froth, namely buoyancy forces, gravitational forces and surface tension. The performance of the froth is a complex function of froth structure (determined by sub-processes in the froth which include; detachment, drainage, entrainment and reattachment), froth stability (which depends on the reagents, solids

concentration, particle size and aeration rate) and froth mobility (which is a function of froth height and froth removal techniques).

Bubbles within the pulp phase pick up mineral particles and carry them up to the pulp-froth interface of the flotation cell. The bubbles, with their load, crowd together at the base of the froth. Associated with each bubble and attached particles there is an envelope of slurry liquid which starts to drain with the particles that are held in suspension as a result of a sudden deceleration of the aggregate due to a buoyancy reduction. The relative velocity of the falling particles depend on the distribution of bubbles, the hold-up of slurry in the froth and the size, shape and density of the particles. At the same time, due to the continuous arrival of bubbles at the interface, the froth region experiences a net upward velocity. A velocity differential between the bubbles and slurry liquid causes a steady decrease in the slurry volume fraction with height causing the bubbles to move closer together, forming a true froth consisting of polyhedral bubbles separated by thin films with plateau borders at the intersection of the films. The bubbles in the upper regions of the froth coalesce due to thinning of these films as a result of drainage and thus the bubble surface area per unit froth volume decreases with height, (Moys, 1978).

It is clear from the above description that the froth phase is very complex with many interrelated variables that ultimately influence its performance as a separating medium. Therefore modelling of the froth phase, as will be addressed next is often neglected as a result of an incomplete understanding of the processes occurring in the froth phase. This is further complicated by the lack of quantitative froth models for describing industrial froth behaviour.

2.2.3.2.2 Froth recovery models

The most common approach to account for the froth zone in the kinetics of continuous flotation cells is the assumption that whatever happens in the froth is directly related to the response of the collection zone. This method of modelling the effects of the froth zone falls into the macro-approach of froth modelling, whereby all the variables influencing the froth performance are lumped together with only the froth transfer rate constants being used to describe the main transfer processes, ie.

$$k_o = k_c * \text{constant} \dots \dots \dots 2.18$$

where

- k_o overall rate constant describing the transfer rate including froth effects (min^{-1})
- k_c collection zone rate constant (min^{-1})

The proportionality constant in this equation is often referred to as the froth recovery factor, (R_f) defined by Finch and Dobby (1990). Froth recovery of attached particles is

associated only with those particles attached to bubbles and is defined as the overall recovery to the concentrate launder of particles entering the froth zone of the flotation cell from the pulp zone. A problem with this approach is that the froth recovery calculated for a particular cell is specific to that cell and it does not provide a method of determining the change in froth recovery when the circuit is operated under different conditions, or in different configurations.

Little attention has been given to the drop-back of particles from the froth phase into the pulp. Arbiter and Harris (1962) postulated that the contents of a flotation cell should be partitioned between the pulp and the froth phase and the interaction of both phases should be included to incorporate the effects of drop-back. They assumed first order kinetics in both the pulp and froth phases, assuming that both operate as perfect mixers as depicted in figure 2.5.

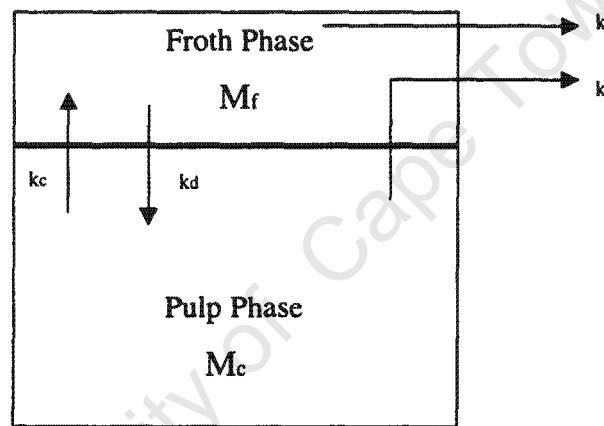


Figure 2.5: Two phase model of flotation, (Arbiter and Harris, 1962)

A mass balance equation for a batch operated flotation system, based on the assumption that the froth and pulp phases are perfectly mixed, is given by the following equations:

$$\frac{dM_f}{dt} = k_c M_c - k_d M_f - k' M_f \dots\dots\dots 2.19$$

$$\frac{dM_c}{dt} = k_d M_f - k_c M_c \dots\dots\dots 2.20$$

- k_c first order pulp rate transfer coefficient (min^{-1})
- k_d first order froth dropback rate transfer coefficient (min^{-1})
- k' first order froth transfer coefficient (min^{-1})
- k overall first order rate transfer coefficient including pulp and froth phases (min^{-1})
- M_f mass of valuable component in the froth phase
- M_c mass of valuable component in the pulp phase

Harris and Rimmer (1966) justified the use of this model experimentally using a continuous flotation cell. Mathe (1998) pointed out that this is based on a more sound theoretical basis than the first approach, distinguishing both zones of the flotation cell. This model assumes that the rate of rejection of particles from the froth phase is directly proportional to the mass of particles in the froth, which implicitly assumes that the probability of rejection is independent of the nature of the particles (k) and the grade of the froth is constant over the height range (no secondary enrichment occurs in the froth zone due to the perfectly mixed phase assumption). However, it is important to recognise that the last assumption is known not to be true, (Moys, 1978).

As an alternative to this approach, Falutsu and Dobby (1989), Contini et al (1988) and Finch and Dobby (1990) proposed a decoupled model of the flotation cell. The approach involved the same formulation of material balance equations around both the pulp and froth phases. They assumed that the dropback of material from the froth to the pulp can be treated as a recycle mechanism between the two phases. Equation 2.21 and figure 2.6 describes this model which predicts the overall flotation recovery (including pulp and froth phases) in a countercurrent flow system.

$$R_o = \frac{R_c R_f}{1 - R_c + R_c R_f} \dots\dots\dots 2.21$$

where

R_c is the collection zone recovery
 R_f is the froth recovery factor

For completely mixed pulp conditions and assuming first order kinetics, the overall recovery equations reduces to equation 2.22.

$$R_o = \frac{k_c \tau_c R_f}{1 + k_c \tau_c R_f} \dots\dots\dots 2.22$$

where

τ_c is the mean residence time within the collection zone

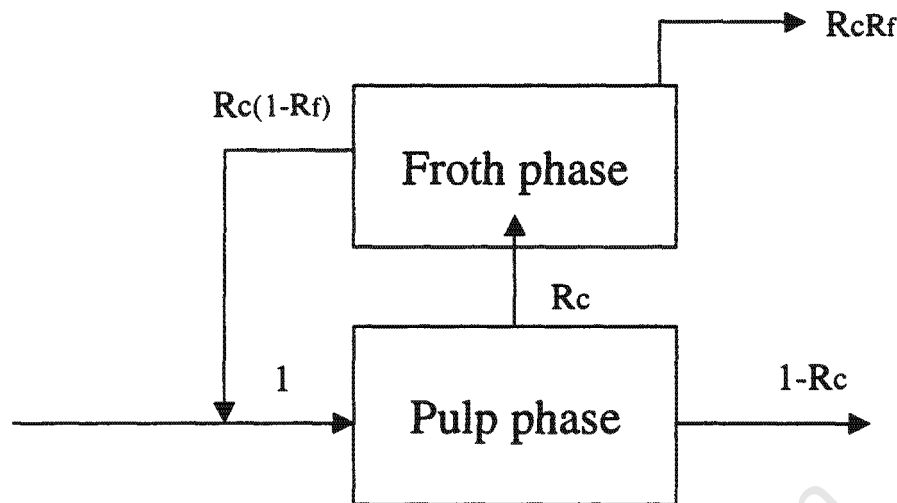


Figure 2.6: Decoupled flotation model, (Finch and Dobby, 1990)

As an alternative, Klimpel (1988) proposed an analytical solution of the differential equation for the analogy of a two stage chemical reaction in a batch flotation cell to include the influence of the froth zone, which can be described as;



k_3 collection zone flotation rate constant (min^{-1})

k_{-3} rate constant describing the drainage from the froth to the pulp (min^{-1})

k_3^* froth transfer rate constant (min^{-1})

The assumption that $k_{-3} \ll k_3$ was made to arrive at the following expression for the recovery by equating the rates at steady state.

$$r = R_{\infty} \left\{ \left(\frac{k_3}{k_3 - k_3^*} \right) (1 - e^{-k_3^* t}) \right\} - \left(\frac{k_3^*}{k_3 - k_3^*} \right) (1 - e^{-k_3 t}) \right\} \quad \dots\dots\dots 2.24$$

This is a three parameter model assuming first order kinetics in the two phases. This model also assumes that a mono-disperse feed exists which ignores the effects of different particle floatabilities within the pulp phase. It should be noted that this equation reduces to a first order equation as the limit of k_3 approaches infinity. Dowling et al (1985) claim that the k_3 parameter is unjustified as k_3^* is the rate limiting step. This model has been found to fit an experimentally determined recovery curve badly during the later stages of a float, underestimating R_{∞} , just as the single phase model does, probably due to the mono-dispersed feed assumption.

Several workers have investigated the influence of flotation variables such as froth height and particle size on froth recovery, (Falutsu and Dobby, 1989, Laplante et al, 1989, Flynn and Woodburn, 1987). It has been argued by several investigators, (Imaizumi and Inoue, 1965, Woodburn and Loveday, 1965) that for a sufficiently fast rate of froth removal, the return rate of particles to the collection zone can be considered negligible and therefore a two phase model is not necessary. This is disputed by Maksimov and Khainman (1965) who showed the importance of the return rate from the froth to the pulp and claimed that as much as an 80% return rate can occur. The differences in these two arguments is probably a result of different operating conditions employed during the test work such as rate of froth removal, froth depth, agitation speed, particle size, size of operating equipment, as well as a number of other variables affecting the dropback of particles into the collection zone.

2.2.3.2.3 Measurement of froth recovery

A number of approaches have been adopted in studies reported in the literature in an attempt to obtain a measure of froth recovery, which can be broadly classified as direct and indirect methods. Direct methods are based on physical measurements within a froth using a variety of specially developed measuring devices. Investigations have been conducted in laboratory columns by Falutsu and Dobby (1989) and Contini et al (1988). More recently Vera (1995) used Contini's approach on industrial flotation cells to study the effect of operating conditions such as air flow rate and froth depth on the froth recovery. Cutting et al (1986, 1989) sampled a froth column using a multi-point sampling lance in order to measure the density and grade with froth depth. An important conclusion from this work was that perfect mixing appeared to occur at the lower levels of the froth and plug flow behaviour occurred in the upper levels of the froth. In addition, significant changes in mineral grade was observed with froth height and the work implied that froth recovery is a function of particle size and that this varied with lateral location in a flotation cell. Moys (1978) attempted to model the froth by incorporating drainage and detachment in a plug flow based model. Ross (1991) in work involving sampling the froth into different compartments in an attempt to simulate the concentration profiles of mineral species and water in froths using a model based on the sub-processes of the froth zone, is believed to be the most comprehensive model based on the froth sub-processes. Froth recovery of attached particles has been estimated in both industrial and pilot scale cells by a number of techniques, with values in the range of 20-50%, (Savassi et al, 1998). A technique developed to measure froth recovery, where no sample of the froth is taken, instead the pulp region is sampled with a funnel connected to a vacuum has shown that froth recovery drops from 70% at the head of a rougher to 10% at the tailings end of the plant, (Savassi et al, 1998). The above technique was developed so that the froth phase would not be disturbed during measurements, limiting the interaction between the process and the measurement.

Indirect measurements of froth recovery have been reported by several workers in the literature. This method typically involves investigating the variation of the overall flotation rate constant (k_o) as a function of froth depth. Mula and Musara (1991)

reported a linear decrease in k_o with increasing froth height, while Laplante et al (1989) showed a curved relationship. Mathe et al (1998) stated that the froth recovery versus froth depth relationships developed in the above approaches was not capable of determining the dominant processes within the froth nor was it capable of providing a scale-up criterion for the froth phase. These two can only be achieved if the froth recovery is described in terms of measurable parameters that do not depend on the geometry of the flotation cell. Recently, Gorain et al (1998) used froth recovery values derived from plant data to show that a relationship exists between froth recovery and froth residence time based on concentrate flow rate. This relationship appeared to be exponential. Harris (1998) modified Gorain's original equation and proposed the relationship described in equation 2.25, this was shown to fit industrial data very well, (Harris, 1998).

$$R_{fk} = (1 - \alpha) \exp(-\beta * \tau_{fc,k}) + \alpha \quad \dots\dots\dots 2.25$$

where

- R_{fk} froth recovery in unit k
- β derived circuit (or bank) froth stability parameter
- α derived froth parameter related to the amount of non-draining mineral
- $\tau_{fc,k}$ froth retention time in unit k based on concentrate flow rate (volume of the froth divided by the concentrate volumetric flow rate)

The beta parameter (froth stability) determines the shape of the exponential decrease with τ_{fc} . It is important to note that the above analysis of the froth performance assumes that the froth recovery does not change as a function of the different floatability classes. This essentially means that any mineral particle entering the froth zone has an equal chance of being recovered, ie: it is non selective. It is known that this assumption is not entirely true and that there should be a change in froth recovery with particle size, density and hydrophobicity but this should only be relevant for circuits treating relatively coarse ores, (Harris, 1998). It must be noted that the effect of turbulence or power input is an important variable that is known to affect this relationship. This area is still relatively poorly understood and no correlation has been established to define the relationship between floatability, froth recovery, bubble surface area flux and turbulence effects.

The bubble surface area flux (S_b) in a flotation cell has been shown to be an effective approach to quantitatively describe the gas dispersion and hydrodynamics in a flotation cell, (Gorain, 1995), as discussed in section 2.2.3.1.6. It has been shown that S_b can be a useful tool for scaling-up flotation cell performance of the collection zones in different cells. Unfortunately, S_b cannot be expected to be a sufficient criterion for scale-up of the overall performance because the overall kinetics are determined by both the collection zone and the froth zone, (Finch and Dobby, 1990). Cell manufacturers usually include the effect of the froth zone during scale-up by incorporating a constant froth residence time (τ_{fg}), based on air flow rate in both cells, defined as the ratio of the froth depth and

superficial gas velocity (J_g). If both cells have the same froth residence times then it is expected that the froth zones in both cells behave similarly. Work at the Broken Hill Concentrator (JKMRC/AMIRA P9L, 1997) showed that the k_c - S_b relationship was able to scale-up metallurgical performance successfully from a 60 litre cell to a 100m³ cell at similar froth residence time values. Research into understanding the limitations of the k_c - S_b relationship in scale-up, as well as understanding the effect of froth characterisation on froth residence time have been investigated, (Harris, 1997). In this work it was found that the overall rate constant (k_o) decreases exponentially with τ_{fg} but the relationship is dependent on cell size and therefore not directly useful for scale-up. When τ_{fg} is divided by a typical cell dimension to include the effect of froth transportation distance in cells of different sizes, the relationship between k_o and the "specific froth residence time" fits the data better and is independent of cell size. Figure 2.7 depicts the variation of k_o/S_b with τ_{fw} for a 2.8m³ and a 0.25m³ cell for a zinc cleaner feed ore. The relationship between the froth residence time on the basis of the concentrate slurry flow rate τ_{fw} , defined by Lynch et al (1981) on the basis of slurry flow through the froth, ie: as the ratio of froth volume to concentrate flow rate, has been found to fit the exponential relationship much better than both the τ_{fg} and specific froth residence times. This suggests that the k_o/S_b - τ_{fw} plot is very useful for scale-up but it is necessary to predict τ_{fw} which in turn needs prediction of the concentrate flow rate from the cell operating conditions (impeller speed, air flow rate and froth depth).

The usefulness of these indirect methods of froth recovery measurement is often very questionable, and are merely based on observations from experimental data. Froth models based on fundamental principles are very difficult to formulate due to the complex nature of the processes involved in the froth phase. The usefulness of the indirect methods, particularly Harris's model has been demonstrated by the successful application of a wide range of data obtained from different plants operating under different conditions. However, their use at the moment is limited to describing existing plant flotation circuits or data.

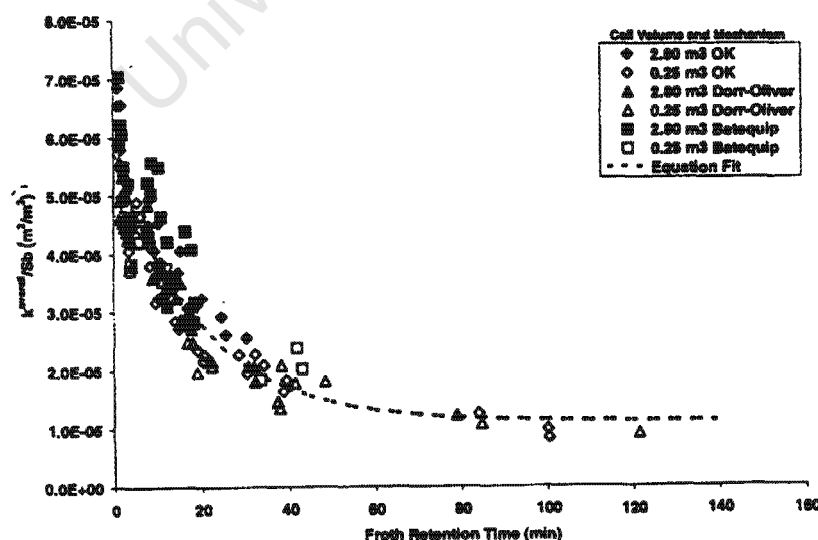


Figure 2.7: The effect of the froth phase on the flotation kinetics, (Harris, 1997)

2.2.3.2.4 Discussion

The role played by the froth phase has been noted to be very important by a number of researchers, and a number of attempts have been made to describe it mathematically. However, little work has been done to identify the various processes occurring within the froth phase, and to use this knowledge to improve the design of flotation machines. The two-phase model of Arbiter and Harris (1962) has formed the basis of much of the work done to describe the phenomena occurring within the froth zone because of its mathematical tractability. The approaches of extracting the required kinetic parameters by fitting performance equations to overall recovery data, Gorain (1998), Harris (1997) are limited as they are empirical in nature and it is only assumed that the parameters have physical meaning. The most important thing to note is that all these models tend to indicate that froth recovery decreases exponentially with an increase in froth residence time. None of these models explicitly includes the effect of bubble coalescence, particle detachment, drainage as well as the effect of chemistry on froth recovery. The complexity of the froth phase has made it almost impossible to develop a global model that can be used to quantify the influence of physical and chemical parameters on froth performance. Linking all the factors and processes governing the froth behaviour when mathematically describing the influence of the froth on flotation performance is a difficult task. As a matter of fact, none of the research reported in the literature incorporates all these factors into a comprehensive model. Given this situation, the influence of the froth zone on the overall flotation process is often accounted for by use of empirically derived parameters and functions when designing, simulating and optimising flotation cells, (Mathe et al, 1998). Much other work has been done which highlights the need for a deeper understanding of the processes occurring in the froth phase, (King, 1973).

2.2.3.3 Entrainment

Johnson et al (1974) defined entrainment in flotation as the undesirable and unpreventable classification of ore particles based exclusively upon size and density which happens simultaneously to true flotation. The entrainment phenomenon is independent of particle surface properties and is highly detrimental to the selectivity of the flotation process, (Trahar, 1981). On the other hand, according to Savassi (1998), the principle factors affecting entrainment of minerals in a flotation cell are the recovery of water, the percentage solids in the pulp, the size distribution in the feed, the froth structure, the residence time in the froth and the mineral specific gravity. Fine gangue material may enter the froth phase via a number of mechanisms, namely:

- as a specifically collected entity
- as a slime coating on valuable particles
- entrapped in froth mass
- entrained in water carry over in the froth

The last of these will be focused on in this section. Smith and Warren (1989) described three mechanisms whereby the entrainment process of gangue in water might occur.

- Water dragged along by bubble walls or the bubbles hydrodynamic layer, (Gaudin, 1957).
- Water transported in the wake of an ascending bubble, (Smith, 1984).
- Water mechanically pushed up into the froth by ascending swarms of air bubbles, commonly known as the bubble swarm theory, (Smith and Warren, 1989).

The primary cause of entrainment is the tendency of low inertia particles to follow water stream lines. Therefore the recovery of fine particles by entrainment in a flotation froth has received sporadic attention over a number of years, (Gaudin et al, 1942, Sutherland, 1948). Low inertia particles have a small chance of collision with bubbles as they tend to move in the same direction as the bubbles wake, therefore entrainment increases with decreasing particle inertia.

2.2.3.3.1 Entrainment models

Due to the homogeneous suspension of low inertia particles in the top levels of the pulp zone (quiescent zone) the ascending bubbles carry along water and entrained particles into the froth. The simplest attempt to model gangue mineral recovery was made by Johnson et al (1972) who showed that the recovery rate of gangue increased with the recovery rate of water, and that this relationship is approximately linear for fine particles regardless of whether froth height or aeration rate was used as the control variable. This was supported by the findings of Engelbrecht and Woodburn (1975). The proportionality constant later became known as the degree of entrainment after Subrahmanyam and Forssberg (1988).

$$R_g = e_g R_w \dots\dots\dots 2.26$$

R_g gangue recovery

R_w water recovery

e_g constant for a particular size particle and specific gravity

Johnson et al (1974) took the above equation one step further and showed that for a particular size fraction of gangue the equation below described his experimental results while working with a porphyry copper ore.

$$\text{Recovery rate of free gangue of size (i) by entrainment} = \frac{RRW * CF_i * MFGP_i}{MWATP} \dots\dots 2.27$$

RRW is the rate of water recovery
 MFGP_i is the mass free gangue in pulp
 MWATP is the mass of water in pulp

$$CF_i = \frac{(\text{Mass free gangue / unit weight water})_{\text{concentrate}}}{(\text{Mass free gangue / unit weight water})_{\text{pulp}}} \dots\dots\dots 2.28$$

The above equations only hold with the assumption that the pulp volume is perfectly mixed for all solids in all size fractions. CF_i is defined as a measure of the efficiency of transportation of gangue in each size fraction through the froth. Lynch et al (1981) showed how CF varies with froth height, particle size and specific gravity of particles. Another conclusion in this work was that the recovery of the non-floatable gangue particles increases with an increase in the percentage solids in the pulp at the same water recovery. This concept was taken one step further by Bisshop and White (1976) who showed that CF is dependent on froth residence time (λ) based on concentrate flow rate according to equation 2.29, (α, β being empirical constants being dependent on the mineral species and the operating conditions of the system being investigated).

$$CF_i = \left(\frac{1 + \alpha\lambda}{1 + \beta\lambda} \right) \dots\dots\dots 2.29$$

They also concluded that the linear relationship between gangue recovery rate and water recovery rate is dependent on the fineness of grind and the specific gravity. The gradient of the relationship increases for a decrease in particle density, (Maachar and Dobby, 1992). If other variables in the flotation cell are changed (collector, modifier, frother, air flow rate and impeller speed) it results in a movement of this linear trend into a more complex relationship, (Johnson et al, 1974, Engelbrecht and Woodburn, 1975). These all affect the froth structure in a flotation cell which is a complex function of many interrelated factors such as frother concentration, particle size, level of surface hydrophobicity, pulp hydrodynamics and chemical environment in the froth. Laplante et al (1989) showed that hydraulic entrainment is mostly dependent on particle size and feed water recovery. They used Z which is equivalent to CF_i defined previously and stated that z→1 for particles less than 5μm and that it decreases with increase in particle size and retention time in the froth. For a perfectly mixed cell, they developed the following relationship to describe gangue mineral recovery.

$$R_i = \frac{Z_i R_w}{1 + R_w (Z_i - 1)} \dots\dots\dots 2.30$$

where

R_i is the gangue recovery

Figure 2.8 depicts how the above model accounts for different values of Z_i .

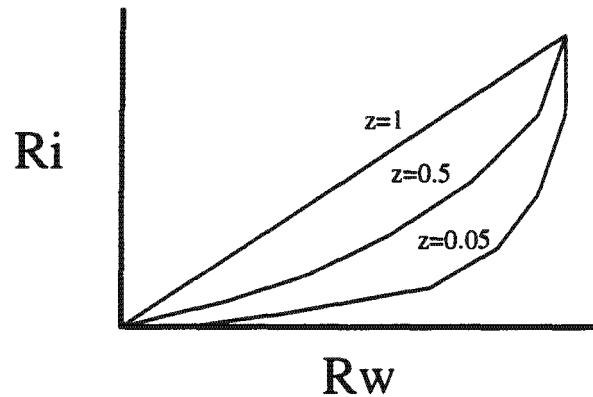


Figure 2.8 : Gangue mineral recovery versus water recovery as a function of Z_i

The variation of the degree of entrainment was found to resemble a classification function whose form depended on the operating conditions of the flotation cell, (Johnson, 1972). Coarse particles were shown to have faster drainage rates than finer particles and therefore the degree of entrainment approaches one for fine particles and approaches zero for larger particles, (Bisshop and White, 1976). Savassi (1998) presented a methodology for quantifying the mass transfer of entrained particles in a flotation cell. He extended Johnson's (1972) approach which described the change in the degree of entrainment (E_i) with particle size via a set of empirical equations. An empirical partition curve was used to describe the relationship between the degree of entrainment and particle size in a flotation cell based on a hyperbolic secant function described in equations 2.31 and 2.32.

$$E_i = \frac{2}{\exp(2.292(\frac{x_i}{\xi})^{ADJ}) + \exp(-2.292(\frac{x_i}{\xi})^{ADJ})} \dots\dots\dots 2.31$$

$$ADJ = 1 + \frac{\ln(\delta)}{\exp(\frac{x_i}{\xi})} \dots\dots\dots 2.32$$

where

- x_i is the average particle size in the i^{th} size interval (microns)
- ξ is the entrainment parameter, particle size for which the degree of entrainment is 20%
- δ is the drainage parameter, related to preferential drainage of the coarse particles from the froth

The constant 2.292 is associated with the choice of the entrainment parameter at a 20% cut size. This was chosen to allow a better account for the entrainment of relatively coarse particles. The drainage parameter accounts for the sharpness of size classification of entrained particles due to drainage in the froth. Increasing values of the ξ parameter increase the degree of entrainment for all particle size classes. Increasing values of the δ parameter increase the rate of drainage of the coarse particles in relation to the fine particles. This equation was found to model experimental data from various sources in the literature, (Bisshop, 1974, Trahar, 1976, Maachar and Dobby, 1992). The values of the two parameters in equations 2.31 and 2.32 need to be chosen carefully as certain combinations of these causes a "degeneration of the empirical partition curve" as stated by Savassi (1998). It must be ensured that the degree of entrainment decreases continuously with increasing particle size and that the curvature of the empirical partition curve is always positive in the region of the coarse particles for the form of the model to be feasible. Savassi found that the two parameters in this model varied linearly with froth residence time based on air (froth volume divided by the volumetric air flow into the flotation cell), this change was relatively small over the entire range of operating air froth residence times however.

Recently, Savassi (1998) proposed an equation that recognises that true flotation and entrainment of non-floating particles occurs simultaneously in a flotation cell. Equation 2.33 is regarded as a significant step towards developing a comprehensive flotation model for predicting the overall recovery of minerals in a flotation cell. No methods in the literature, to the authors knowledge, has included both aspects influencing the ultimate recovery of minerals in a flotation cell in a proper manner.

$$R_o = \frac{R_c R_f (1 - R_w) + E_i R_w (1 - R_c)}{(1 - R_w)(1 - R_c + R_c R_f) + E_i R_w (1 - R_c)} \quad \text{.....2.33}$$

where

- R_o is the overall recovery from the flotation cell, including pulp, froth and entrainment contributions
- R_c collection zone recovery
- R_f froth recovery factor
- E_i degree of entrainment as a function of particle size class (i)

Moys (1978) developed an alternative expression for the entrainment of particles in a flotation froth based on a discrete model of the froth phase. The mass flow rate of entrained solids is related to the gas flow rate into the flotation cell and the mass of entrained solids at a vertical point in a flotation froth.

$$\frac{\partial M_{ei}(z)}{\partial z} = -k_{ei} A_c(z) \frac{M_{ei}(z)}{G} \quad \text{.....2.34}$$

Integrating this expression indicates that the mass of entrained particles decreases exponentially with an increase in froth height.

$$M_{ei}(z) = M_{ei}(0) \exp\left[-\frac{k_{ei}}{G} \int A_{(z)} dz\right] \dots\dots\dots 2.35$$

A_c cross sectional area of flotation cell
 k_{ei} rate constant for transfer of entrained particles to the flowing stream
 G gas rate
 $M_{ei}(0)$ entrained mass flow rate at zero froth height
 z froth height

Another approach to quantifying the entrainment process was by Jowett (1966) who accepted that the mass flow of gangue from the pulp to the froth followed a first order rate according to equation 2.36.

$$\frac{\partial M_p}{\partial t} = a V_p C_p^g \dots\dots\dots 2.36$$

M_p mass of gangue in the pulp
 V_p pulp volume
 a rate coefficient for gangue
 C_p^g concentration of gangue in the pulp

It is reasonable to assume that only a portion of this entrained mass will ultimately land up in the concentrate due to classification in the froth zone. The flow rate of concentrate was related to the fraction of recovered gangue in the froth (an increase in flow rate causes an increase in the fraction of gangue recovered in the froth).

$$\frac{\partial M_g(\text{conc't})}{\partial t} = a V_p C_p^g * b Q_c \dots\dots\dots 2.37$$

Q_c concentrate volumetric flow rate
 b proportionality constant

This was taken further by recognising that this equation is only strictly valid for particles which are completely liberated. A proportion of gangue in the concentrate originates from an association of the gangue with valuable mineral in the pulp. Jowett (1985) split gangue in the pulp into two fractions, gangue which is fixed to valuable particles and free gangue particles which are completely liberated from the valuable particles. The fraction of gangue fixed to valuables was denoted (i). The modified equations were as follows:

$$\frac{dM_g}{dt} \text{ (free gangue to concentrate)} = aV_p * bQ_c * (C_p^g - iC_p) \dots\dots\dots 2.38$$

$$\frac{dM_g}{dt} \text{ (fixed gangue to concentrate)} = Q_c i C_c \dots\dots\dots 2.39$$

Another approach to modelling entrainment is that proposed by Kirjavainen and Laapas (1988) who included the effects of particle mass, shape, size, water recovery and slurry viscosity. They proposed that the concentration of hydrophilic mineral in the concentrate of a flotation cell is described by equation 2.40. It is evident that the degree of entrainment, E_i (or e_i) defined previously varied between nil and one, which indicated that it is actually a probability factor as described in equation 2.40.

$$C = C_o \int_0^1 \exp(-PR_w) f(P) dP \dots\dots\dots 2.40$$

C concentration of hydrophilic mineral in the concentrate

P probability factor

f(P) probability factor distribution function

The boundary conditions on the entrainment factor were quantified. For very fine particles, the probability of entrainment should approach unity, then the recovery of solids is equal to the recovery of water. If the water recovery is low then the probability of entrainment should approach zero. An equation describing these boundary conditions is:

$$P = \frac{W^a}{W^a + bM^c} \dots\dots\dots 2.41$$

The coefficient b in equation 2.41 was found to include the effect of viscosity on the entrainment parameter, hence b was defined as a function of slurry viscosity according to equation 2.42. This is the only equation that explicitly includes the effect of slurry viscosity on the probability of entrainment.

$$P = \frac{W^a}{W^a + f(\eta)M^c} \dots\dots\dots 2.42$$

2.2.3.3.2 Discussion

A reliable simulator for industrial flotation cannot be developed without including a model for entrainment. Models that do not include entrainment predict higher concentrate grades and lower concentrate recoveries compared to the true flotation process. All the models presented in this section rely on the water recovery prediction in a flotation cell. Engelbrecht and Woodburn (1975) found that the same relationship between the recovery of water and the recovery of free gangue applied regardless of whether they varied the aeration rate or froth depth to control the recovery of water. Therefore the gangue recovery models do not need to explicitly include froth height and air flow rate effects if these are incorporated into a water recovery model. Even though an entrainment model should explicitly include the effect of changing operating conditions, it can be indirectly included via the water recovery calculations if this takes into account the operating conditions. The equation developed by Savassi (1998) which recognises that true flotation and entrainment of non-floating particles occurs simultaneously in a flotation cell is regarded as a significant step towards developing a comprehensive flotation model for predicting the overall recovery of minerals in a flotation cell. Due to the dependence of the water recovery on the entrainment of gangue and valuable minerals, the next section describes the various attempts at predicting the water recovery from a flotation cell.

2.2.3.4 Water recovery

The content of water in a flotation froth is an important aspect affecting the degree of upgrading with froth depth as a result of the drainage of gangue particles with the water in the froth. It is also an important component to the overall stability, mobility and finally the ultimate recovery of valuable minerals from a flotation cell. The volume of water in every concentrate stream of a plant constitutes over sixty percent of the total volume, often much more, and therefore accounts for over half the volume flows in the circuit. The volume flows are used to calculate the residence times in the flotation cells of the circuit. These residence times are used in the mineral recovery models. Obviously, an error in the water model results in errors in the floatable mineral kinetics as well as the entrainment kinetics. Therefore, the importance of accurately describing the water kinetics cannot be underestimated. The purpose of this section of the literature review is to briefly describe the water models used previously in the literature particularly with respect to their application in a simulation algorithm.

Ideally, a simulation model should be able to predict the water flows in the concentrate and tailing streams, as well as the amount and location of water additions. However, no satisfactory model is yet available that allows this. Harris and Rimmer (1966) stressed the importance of the dependence of the hydrophilic gangue recovery rate on the water recovery rate in a flotation cell. The partial dependence of the hydrophobic gangue and valuables recovery rate on the water recovery rate was also highlighted in this work. The complicated nature of this interdependence means that a detailed mechanistic model for the water recovery is necessary or that a consistent methodology is required

to overcome the need for a complicated water recovery prediction. It is clearly evident that an attempt to analytically describe the water recovery rate cannot be considered in isolation from the recovery rates of the hydrophobic and hydrophilic particles, as these influence the stability and water content of the froth. Recent investigators have begun to recognise the importance of including water behaviour in flotation models, (Woodburn et al, 1984). The empirical entrainment models for gangue recovery developed by Moys (1978) and Lynch et al (1981) have resulted in the first approximate models for the distribution of water in a flotation cell. One of the most important factors that controls the water distribution in a flotation cell is the froth. Unfortunately there has been very little consideration of water and particle transport in the froth, even though the froth and associated drainage phenomena are considered to be the principle reasons for the cleaning action of the cell.

Langberg (1988) examined the factors which affect the water recovery rate from a theoretical and experimental point of view using the Ergun equation. He used fundamental chemical engineering principles relating to phenomena occurring in the flotation froth but still found it necessary to employ semi-empirical expressions. In order to develop a fundamental water model, the microstructure of the flotation froth needs to be understood. The microstructure of a flotation froth depends on the liquid holdup (ϵ_L). According to Langberg (1988), if ϵ_L is greater than 0.26 then bubbles remain spherical and form what is called a bubble bed. When ϵ_L approaches zero the bubbles deform to create a polyhedral shape and the dispersion of gas in the froth is then called a cellular foam. Bubble beds are easy to describe mathematically but cellular foams are more complex structures. They appear as very thin liquid films joined together by a network of capillary channels known as plateau borders. Leonard and Lemlich (1965) related mathematical equations to describe these different gas dispersion states of mineralised flotation froths and found that they could be characterised by the liquid hold-up, bubble size, hydraulic radius (size of interstitial channels through which water drains), the plateau border as well as the thickness of the film of liquid surrounding the bubbles. De Vries (1972) showed that most of the liquid in a solid free two phase cellular foam is found in plateau borders, while most of the gas/liquid interfacial area is associated with liquid films. The top of a flotation froth consists of cellular foams with bubble sizes in the range of 1-20mm and have polyhedral shapes. Jameson and Allum (1984) showed that bubbles in most flotation pulps are approximately 0.5-1.0mm in diameter and are spherical. This observation highlights the considerable amount of coalescence and drainage of interstitial liquid that takes place as bubbles rise through the froth. However, Gaudin (1957), Klassen and Mokrousov (1963) were the first to recognise that particles within a three phase froth can exert a profound influence upon its properties and behaviour. They provided only a qualitative interpretation of these influences and recognised that particle shape, size and hydrophobicity as well as the concentration of particles (mineralisation) are important variables affecting the stability, and water content of flotation froths.

2.2.3.4.1 Water recovery models

The most general method used to model the behaviour of water in a flotation cell assumes that water obeys first order flotation kinetics in a manner similar to that used to describe hydrophobic particles, (Imaizumi and Inoue, 1965, Scrimgeour et al, 1970, Thorne et al, 1976). A weak assumption in these models is that flotation of the minerals is independent of the water removal rate. Smith (1984) points out that this approach is limited as the water behaviour along an entire bank cannot be simulated. This approach was adopted by Lynch et al (1981) in their modelling of flotation circuits. They assigned a two component water model using fast and slow rate coefficients and found this to adequately reproduce their experimental data. Again Smith (1984) points out that this approach has no physical significance and only serves to highlight the lack of knowledge in this area. Woodburn and Wallin (1984) assumed a zero order water removal rate with respect to the water concentration in the pulp while using a decoupled kinetic model for simulation of flotation networks. They found that the water removal rates in semi-batch flotation tests showed an exponential decrease of the proportionality constant with time which they regarded as being a measure of the time dependence of the stability of the froth during the experiments. The following equation shows the model describing the water removal rate from their flotation tests.

$$Q_w = \alpha_o Q_A K_w \exp(-\varpi t) \dots\dots\dots 2.43$$

where

Q_w is the specific water removal rate
 α_o is a constant
 Q_A is the specific cell aeration rate
 K_w is the zero order rate constant
 t is the flotation time
 ϖ symbolizes the time-dependency of the stability of the froth and was found to vary with Q_A

Loveday and Marchant (1972) measured the residence time in a flotation cell and calculated the water flow rate knowing the cell volume and pulp density. King (1973) related the water flow rate to mass flow rate described by equation 2.44, provided the specific gravity of the pulp in the concentrate can be predicted or measured. This fixes the water flow rates for the whole plant. It was assumed that this relationship was independent of operating conditions. This is known not to be true for all cells in a circuit, especially between the roughers and cleaners with thicker, more stable froths typically found in the latter. They did however note the crucial role played by water in the flotation models and stated that ideally, the flotation rate of water should be quantified and treated independently from the solids and stressed that models predicting the flotation rates of water were unavailable. This is the water model that is currently being used by the UCT/JKMRC research collaboration in their simulation package, JKSIMFLOAT. It appears to fit most flotation plant data very well, (Runge, 1997).

However it is specific to a flotation cell in a fixed circuit configuration as it does not relate to any physical variables in the flotation cell and is therefore not appropriate for simulating alternative circuit configurations.

$$Q_w = aF_s^b \dots\dots\dots 2.44$$

where

- Q_w concentrate volume flow of water (l/hr)
- F_s total concentrate mass flow rate (kg/hr)
- a water recovery slope parameter
- b water recovery exponent parameter

Woodburn et al (1965) took water as the reference species throughout the circuit and defined enhancement ratios for all other species based on what happens to water across a cell. Sutherland (1981) using data from a lead flotation circuit assumed a fixed pulp density in the tailings stream. This assumption leads to large errors in the cleaning and recleaning stages of a circuit as the pulp density often decreases down a bank of cells. More recently, Kawatra et al (1982) assumed that the water recovery is directly proportional to the valuable mineral recovery.

Frew and Sutherland (1984) tested the following four methods of empirically modelling the water recovery in banks of sulphide flotation circuits.

- Constant pulp density in the flotation concentrate
- Constant first order rate process
- Constant pulp density in the flotation cells
- Constant ratio in the concentrate of water flow to values flow

The first two attempts seemed to be the best, but none of them could adequately describe the water flow rates due to the empirical nature of these approaches. Bisshop and White (1976) found that the flow of water in the concentrate of a flotation cell is a direct function of the residence time of air in the froth, approximated as the volumetric flow rate of air into the flotation cell divided by the volume occupied by the froth phase. This approach was used by Savassi (1998) who suggested that a power relationship between the recovery of water and the froth residence time of air could be used for the system investigated. Figure 2.9 shows the variation of the water recovery with froth residence time of air in this work and the relationship obtained is shown in equation 2.45 below. It was emphasised that the form of this equation is strongly dependent on the structure of the froth phase however and cannot be used to predict the water recovery in different cells with different froth structures.

$$R_w = 0.243\lambda_{air}^{-0.942} \dots\dots\dots 2.45$$

where

λ_{air} is defined as the residence time in the froth in terms of the volumetric flow rate of air into the flotation cell

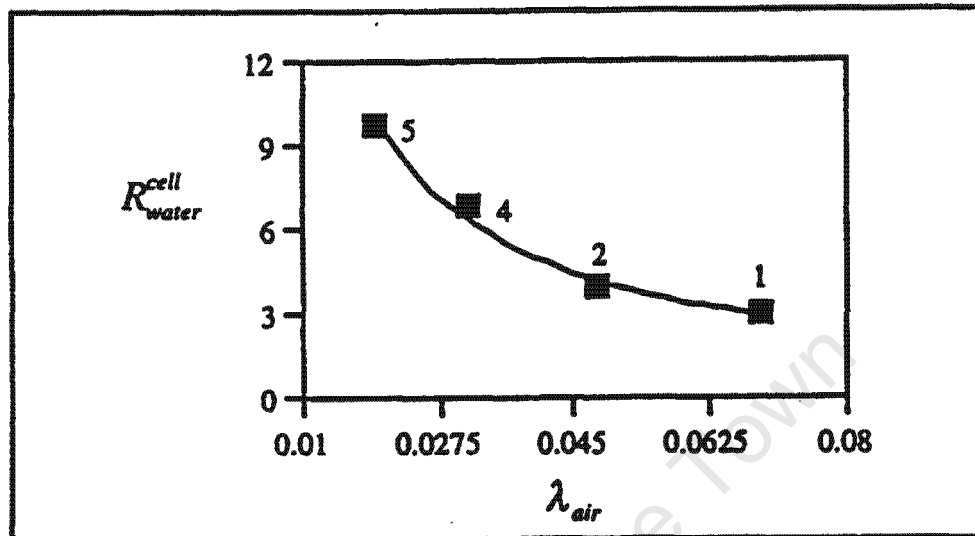


Figure 2.9: The variation of the water recovery with the froth residence time of air, (Savassi, 1998)

Smith (1984) attempted to model water recovery and made the following observations:

- Water is recovered to the concentrate when no solids are present
- The percentage solids in a flotation concentrate tends towards a maximum
- Water recovery down a set of rougher/scavenger banks can be modelled by a two component model
- A relationship exists between water recovery and mineral recovery

Smith (1984) considered water recovery due to lifting in the wake of ascending air bubbles and therefore water flow to the froth is due to air. Two proportionality constants were used to describe water recovery, one for transfer of water from the pulp to the froth due to air and a second for the transfer of water from the pulp to the froth due to the action of solids. This model seems to be the most comprehensive empirical model in the literature for describing the process of water recovery, including the different effects of particles and air in the process of water recovery.

Langberg (1988) stated that while techniques such as the ones described above may be useful and convenient when modelling complex flotation circuits, they offer no insight into the mechanisms of water recovery, and how it may be controlled. A number of mechanistic models have been published, which go some way towards this goal and are briefly described below. The general form of these models are divided into two categories. A well mixed froth phase assumption of Mika and Fuerstenau (1969) and

Bascur and Herbst (1985) and distributed models of the froth phase presented by Cooper (1966), Cutting and Devenish (1975) and Moys (1978).

Mika and Fuerstanau (1969) developed a model based on a population balance equation for each species in the pulp and froth phases. The froth was considered completely mixed and homogeneous in composition. The water was assumed to be carried into the froth as a result of water surrounding the sheath of rising bubbles into the froth. The water was split into two categories, boundary layer water associated with the sheath of water which does not drain, and free water which is associated with the water in the plateau borders and is free to drain. These authors admitted to the intuitive appeal of a distributed model rather than the well mixed assumption but rejected it in the grounds of it being too complex to analyse sufficiently accurately. Bascur and Herbst (1985) extended the above model by incorporating a drainage model to provide a more accurate description of the froth phase. They still assumed a well mixed froth phase however. The equation below describes the water flow over the cell lip developed in this work.

$$Q_c = k_c L \Delta H^{1.5} (1 - \varepsilon_G) \dots\dots\dots 2.46$$

where

k_c is a constant
 ε_G is the average gas holdup in the froth
 ΔH is the height which the froth column rises above the lip
 L is the lip length

On the other hand, distributed models of the froth phase presented by Cooper (1966), Cutting and Devenish (1975) and Moys (1978) predicted an approximately exponential decrease in the recovery rate with increasing froth height. The equation below shows the model developed by Moys (1978) in his plug flow analysis of the froth phase.

$$M_w = \left(\frac{6\alpha\delta Q_G}{d_b} \right) \exp\left(-\frac{k_w A_c H_F}{Q_G}\right) \dots\dots\dots 2.47$$

where

M_w water recovery rate
 α is the fraction of air which reaches the concentrate
 δ is the thickness of the sheath surrounding the bubbles entering the froth
 A_c is the cross sectional area of the column
 k_w is the rate constant for the transfer of water from the up flowing to the down flowing stream within the froth
 H_F is the froth height
 d_b is the diameter of bubble entering the froth
 Q_G is the gas volumetric flow rate

This equation was able to predict the water concentration with froth height by selecting suitable rate constants. However the rate constants were not related to any physical properties of the components or operating conditions in the cell. It is important to note that the above relatively complex equations were abandoned in favour of much simpler linear transfer functions as described previously.

The last model considered is that of Flynn and Woodburn (1987) who focussed attention on the top layer of the froth to describe the water recovery. The equation is descriptive rather than predictive however. They considered the upper layer of bubbles in the froth only, and described them as having a prismatic base and a hemispherical dome. The dome was considered fully loaded with minerals, while the sides of the upper bubbles contained water and entrained gangue. The following equation was derived for the recovery rate of water.

$$M_w = \frac{3\alpha Q_G (1 - x_s) \rho_{pulp} \delta}{2d_b} \dots\dots\dots 2.48$$

where

- δ film thickness
- d_b diameter of the bubble at the top of the froth column
- x_s is the mass fraction of solids in the froth
- ρ_{pulp} is the pulp density
- Q_G is the gas volumetric flow rate
- α is the fraction of air which reaches the concentrate

2.2.3.4.2 Discussion

The simple water models relating water to either the mineral flow and solids flow are extremely crude while the mechanistic models require the need to measure difficult process variables that are often not easily obtainable in operating industrial flotation cells. These models have often been replaced by much more simpler functions. Obviously a considerable amount of effort is still required in order to derive a practical, yet sufficiently descriptive water recovery model. The lack of the existence of a fundamental water recovery model in the literature is highlighted as the most important "weak link" in the chain of work currently being undertaken in the field of flotation circuit simulation.

2.2.4 Flotation cells

The design of flotation cells is often a matter of experience combined with judgement and, perhaps even, a certain amount of prejudice, rather than design from a detailed analysis from first principles, (Young, 1982). The ideal flotation machine would be able to produce maximum recovery of valuables at a high product grade and at the lowest possible cost. Machines designed to meet this criteria must achieve the following objectives simultaneously:

- Keep particles in suspension
- Disperse air into small bubbles in the slurry
- Create adequate conditions for bubble-particle contact
- Create quiescent conditions in the froth zone of the cell
- Ensure that particles entering the cell are removed relatively easily into the concentrate or tailings streams
- Ensure easy start-up after stoppages due to breakdowns, power outages, etc.
- Ensure that cell geometry fit easily into circuits
- Provide for easy removal of froth

The conventional mechanical cell is the most common type of flotation machine used in the industry. These cells are characterised by a mechanically driven impeller which agitates the pulp and disperses air into it. The high degree of turbulence, caused by the impeller is probably the main characteristic feature of this type of flotation cell design. The high intensity of contacting between particles and bubbles promotes the recovery of relatively fine particles. The good solids suspension on the other hand, results in high recovery of coarse material. It is normal to have a number of impellers in series in one vessel. This type can have different pulp flow regimes as well as different aeration mechanisms. The machines can have an "open" flow of pulp between each impeller or a "cell-to-cell" concept where there are weirs between each impeller. The "open flow" type has traditionally been the most commonly used, (Wills, 1988) although the tendency in recent times has been towards designing larger individual cells (tank cells), up to sizes of 130 m³. The introduction of air into the pulp can be by "self aeration", where the machine uses the depression created by the impeller to induce air, or where the machine receives air from a blower.

2.2.4.1 Mixing regimes and non-ideal flow

The recovery in a cell does not only depend on the distribution of rate constants but also on the distribution of particle residence times in the cell. This residence time distribution depends on the mixing regimes characteristic in a cell. The plug flow model assumes that all particles spend an equal residence time (τ) in the pulp of a flotation cell. Therefore the residence time of all the components can be described by equation 2.49.

$$\tau = \frac{V}{v} \dots\dots\dots 2.49$$

V pulp volume

v volumetric flow rate of tailings

This ideal flow mixing regime can be approximated for use in column flotation cells (as well as batch flotation cells). Substituting equation 2.49 into the first order equation for flotation (equation 2.2) and integrating the expression, the recovery is given by Finch and Dobby (1990) described in equation 2.50.

$$R = 1 - \exp(-k\tau) \dots\dots\dots 2.50$$

This equation is of most direct relevance to laboratory flotation cells. Industrial flotation columns operate under conditions between plug flow and perfectly mixed regimes, (Finch and Dobby, 1990). The second mixing model that is most often considered is the perfect mixing reactor model. This model assumes that all particles in the cell are perfectly mixed and therefore complete homogeneity exists throughout the cell contents. In such an instance it would be reasonable to assume that all material in the reactor then has an equal probability of leaving the reactor. For continuous reactors, the residence time distribution $E(t)$ can be described by equation 2.51.

$$E(t) = \frac{1}{\tau} e^{-\frac{t}{\tau}} \dots\dots\dots 2.51$$

Substituting equation 2.51 into the first order rate equation;

$$T_i = F_i \int_0^{\infty} e^{-k_i t} \frac{v}{V} e^{-\frac{v}{V} t} dt \dots\dots\dots 2.52$$

Integrating this expression by parts and using a mass balance equation around a flotation cell ($C_i = F_i - T_i$) the expression representing the recovery to the concentrate for the particular component can be represented by equation 2.53.

$$R_i = \frac{k_i \tau}{1 + k_i \tau} \dots\dots\dots 2.53$$

Most researches conducting tracer tests have concluded that the pulp volume of most flotation cells can be adequately represented by the perfect mixing model, (Jowett, 1966, Imaizumi and Inoue, 1965, Woodburn and Loveday, 1965). Deviations from this observation are most distinct in the coarse size fractions.

2.2.4.2 Scale-up

Fundamental knowledge of the micro-mechanisms of the flotation process is still very limited. As a result, flotation "machine design" scale-up has generally been carried out using similitude considerations and dimensionless group analysis. The same principles have been applied as used successfully in the design of chemical reactors and rotating machinery such as pumps and mixers. Table 2.2 describes the various bulk dimensionless numbers used to understand the hydrodynamics and gas dispersion properties in flotation cells, (Mavros, 1992). The physical interpretation of the dimensionless numbers are included in this table to highlight their apparent significance. The effects of these scale-up (or down) parameters on the metallurgical performance (grade and recovery) are used by several flotation cell manufacturers for scale-up and flotation circuit design. The most common dimensionless group considered is the power number, which describes the amount of pulp moving capacity in the cell. Comparing two impellers of different design which have the same diameters and agitation speed, a higher power number reflects a greater pulp moving capacity due to a greater effective blade area. An effective cell design for the flotation process should be aimed at imparting a high power dissipation intensity (power per unit mass of pulp). Therefore, the higher the power number imparted to the pulp, the greater the intensity of the flotation conditions. When considering scale-up of flotation cells, it is also important that the same power numbers are used if the model and prototype are to be scaled properly. The air flow number is believed to be a useful criterion for similar gas dispersion for different cell sizes, but this measure of gas dispersion is very generic and does not describe the dispersion of gas in terms of bubble size and gas holdup which are very critical in the flotation process. Arbiter and Harris, (1969) showed that the air flow number is an important criterion for obtaining similar metallurgical performances in different cell sizes of similar design. However, experiments by Fallenius (1979) did not support the contention that the air flow number is useful as a similitude criterion in flotation cell scale-up. However, due to the lack of a suitable alternative for scale-up, the air flow number continues to be used by cell manufacturers.

Deglon (1997) has pointed out that these relationships do not consider the fundamental sub-processes in a flotation cell such as bubble breakup and particle-bubble contacting occurring in the flotation micro-environment and can therefore only be used to a limited extent. Gorain (1997) pointed out that these numbers should only serve as criteria for similar particle suspension and agitation. With the recent trend of increasing flotation cell sizes (100 m³ and larger), the problem of scale-up has become even more acute. The scale-up features that may have been tolerated on smaller cells are not applicable to these newer, larger cells. Arbiter et al (1980) pointed out that if scale-up is attempted by rule-of-thumb and trial-and-error, then the probability of failure grows with scale. A

clear understanding of the conditions inside a flotation cell, such as the cell hydrodynamic conditions is critical to the development of reliable solutions to the problems associated with flotation scale-up.

Table 2.2 Dimensionless numbers used in characterising mechanical flotation cells

DIMENSIONLESS NUMBER	EXPRESSION	PHYSICAL INTERPRETATION
Power Number (N_{PW})	$\frac{P}{\rho N^3 D^5}$	Ratio of actual to theoretical power input
Air Flow Number (N_{AF})	$\frac{Q_g}{ND^3}$	Ratio of air flow rate to impeller pumping rate
Froude Number (N_{Fr})	$\frac{N^2 D}{g}$	Ratio of inertial to gravitational forces
Reynolds Number (N_{Re})	$\frac{N^2 D \rho}{\mu}$	Ratio of inertial to viscous forces
Weber Number (N_{We})	$\frac{N^2 D^2}{\sigma}$	Ratio of inertial to surface tension forces

2.2.5 Flotation circuits

Flotation plants always consist of several stages which are usually broadly categorised as roughing, scavenging and cleaning stages, depending on their position in the plant. Each stage usually consists of several flotation cells, or flotation banks, connected so that the tailings pass sequentially through the cells of each stage. The collection of stages with their interconnecting feed, concentrate and tailing streams is called the plant or circuit configuration.

2.2.5.1 Flotation circuit design

A flotation circuit or flowsheet can be defined as a set of unit operations, mechanically and chemically incorporated into a system, to concentrate minerals from a given ore. The layout of a circuit will therefore depend on the hydrodynamic and chemical properties needed to overcome the constraints of the ore to be treated. Flowsheet configuration plays a significant role in the determination of metallurgical performance, (Bulatovic and Wyslouzil, 1988).

Traditionally, flowsheet selection has been based solely on the sequential flotation of two or more valuable minerals, where complete circulation of the intermediate products within the circuit is practiced (Bulatovic et al, 1993). This conventional sequential flowsheet works well if the ore being treated is a coarse-grained ore with a simple

mineralogical composition. The circulation of the intermediate products in this case has little or no effect on the overall metallurgical performance. Bulatovic et al (1993) found that the conventional sequential flowsheet rarely provides optimum metallurgical results for sulphide ores with a complex mineralogical composition and poor intermediate selectivity. The design of alternative flowsheets then becomes an important task in the overall process development. In spite of the availability of several simulation models in the literature, design of flotation circuits for optimum performance is, by and large, based on:

- 1) Laboratory scale tests in batch and/or locked cycles
- 2) Pilot plant tests on a limited number of circuit configurations
- 3) Past experience and trial and error techniques

The major difficulties in using laboratory data for flotation circuit design include:

- Ore sample representivity
- Feed preparation procedures
- Laboratory procedures (type of machine, operation, experimental conditions)
- Flotation machine scale-up
- Flotation kinetic scale-up
- Flotation circuit design using laboratory data.

Even though all these complications arise when attempting to scale-up flotation circuit performance, attempts at developing small scale continuous flotation machines have been made over the years. Raffinot and Formanek (1963) were the first to present a small continuous flotation plant. Recently a smaller unit was introduced by the Dowa Mining Company Ltd of Japan (Dowa, 1984) which enables a few 10's of kilograms of ore to be processed continuously. The results from the tests conducted indicates that it is feasible to operate these small scale units and that the results obtained compare to industrial data on similar feeds.

A method widely used in the design of flotation circuits is to use batch flotation tests, performed either on the feed ore, or on selected circuit process streams. While such tests can be useful for the design of circuits and for the selection of flotation reagents, they do not give quantitative information about the number of stages and configuration of the circuit. The batch procedure of a rougher float followed by a cleaner float (in open circuit) can be a useful way of determining the final concentrate grade that may be achieved, but the recovery is usually significantly lower than that obtained in a three or four stage closed circuit operation. The differences in impeller (speed and size), power input, cell dimensions and water behaviour in a batch cell and plant cell can be considerable. The use of an intermediate step to bridge this gap, using pilot plant facilities, is therefore prevalent. A homogeneous feed composition can be ensured as well as a controlled circuit that can enable the design variables to be measured effectively. The effect of recirculating streams as well as the influence of the different pulp mixing regimes in continuous operation can be assessed. The mixing regime in batch cells is very different from the near perfect mixing characteristics in continuous

plants and pilot plants. The influence on final recovery due to the possibility of larger particles short circuiting to the tailings streams can be diagnosed. In addition, a large amount of data can be gathered over a very short period of time as pilot plants reach steady state a lot quicker than industrial plants and many different circuit structures can be assessed with relative ease. Full scale plant testing can be very expensive as any loss in performance during the testing could cost millions in lost revenue. In addition, the outcome of full scale testing may not always be clear as random variations in the process feed or equipment operation may interfere with the results. Many of the factors that influence the flotation performance interact with each other, and the right course of action to rectify poor performance can be unclear, Sweet (1999). Again the advantage of pilot plant testing is that negative results do not impact as greatly on lost revenue when sampling using a full scale plant to conduct test trials. The shortfalls of pilot plant testing is that the equipment necessary to set up a pilot plant can be very expensive. Also, pilot plants are not always easy to operate as they can be very sensitive to changes in operating variables. It is not always easy to reproduce plant chemical and milling operating conditions in a pilot plant, and obtaining a good, representative sample for such tests can be difficult. In addition, the applicability of the results to test work on a pilot plant to the operation of a full scale plant is not well understood.

2.2.5.2 Flotation circuit modelling and simulation

This section of the review will attempt at giving a brief description on the approaches taken to simulate circuits as opposed to individual cells. The assumptions made by previous researchers and the relevance of these assumptions will be discussed with regard to circuit simulation. The previous sections of the review have discussed the various models that have been used to model individual cells. Single stage separations in froth flotation are often not effective enough to achieve satisfactory separation of the valuable minerals from the gangue. Therefore multi-stage circuits with recycle streams are employed to increase the efficiency of the separation process. The performance of these circuits are controlled by the arrangements of the cells into a network and the various chemical and physical variables employed in the pulp environments.

2.2.5.2.1 Introduction

When flotation circuit re-arrangements are proposed or a new operational strategy is decided on, it is often very difficult to obtain conclusive evidence as to the effectiveness of the change. This is as a result of the variable nature of a flotation plant operation caused by changing feed composition as well as disturbances from routine maintenance and milling conditions masking whether an increase in recovery or grade has indeed been effected by the change. Simulation methods can be used to assist in the above problems, enabling the metallurgist on the plant to compare numerous operating and design strategies without needing to carry out costly and complicated test work. However, the conclusions reached by simulation are sensitive to the accuracy with which the system can be described by a mathematical model. As a result, the best

approach is usually a combination of simulation and experimental studies, with each contributing to a better understanding of the process. The next section discusses the various methodologies reported in the literature in an attempt to simulate flotation circuits.

2.2.5.2.2 Simulation models and procedures

Most early models that were developed limited the simulation of a circuit to an individual cell, with the dominating assumption that the flotation behaviour of any species remains constant irrespective of the position of the cell in a circuit. However, Frew and Trahar (1982) noted from tests conducted on industrial plants, that the characteristics of flotation in a cleaner bank were often substantially different from rougher banks, operated at the same conditions. This led to the necessity of using different methods to describe the rate behaviour in different sections of a circuit. Imaizumi and Inoue (1965) assumed that the rate coefficients are the same in all sections of the plant. Neimi and Paakkinen (1968) acknowledged that the rate coefficients are different in different sections of the plant but still used uniform values for the rate coefficients throughout the circuit. Loveday et al (1972) realised that the aeration and agitation intensity would be expected to affect the flotation rate, although the ratio of the rates of all the floating species would be expected to remain constant. Combining all the differences of the variables in each cell such as residence time, frother concentrations and aeration rates they produced a modified residence time to account for the differences in the variables in each stage of the plant. They then assigned a constant rate coefficient to all floating species across the bank. Loveday and Marchant (1972) claimed that using a continuous distribution of rate constants as described in section 2.2.3.1.3 gave a better prediction of performance when attempting to link the performance of a cleaner to operating conditions in a rougher bank.

Frew and Trahar (1982) conducted surveys in numerous plants using batch tests to evaluate each of the above methods. They found that the flotation behaviour in the cleaners can differ substantially from that of the roughers, and concluded that in order to simulate complete flotation plants one needs to ascertain these differences. The approach of Imaizumi and Inoue (1965) was found to give incorrect estimates of plant performance. The approach of Loveday et al (1972) allowing for the differences in rate coefficients by adjusting them by a constant factor proportional to the physical variables was also shown to be erroneous in some systems. They concluded that the decrease in the rate coefficients in the cleaner banks was as a result of weak mineral-collector interactions in the systems they studied and hypothesized that collector desorption from particle surfaces occurred in the cleaning circuit. They also concluded that the major differences in the rate coefficients occurred in the size fractions exhibiting substantially less than complete recovery, that is in the coarse and fine size fractions.

Another approach that has been proposed is to assign individual rate constants to each section of the flotation circuit, based on a steady state mass balance around each of the

relevant sections in the plant, (Sutherland, 1977). However, this does not allow for a means of predicting the rate constants during simulation for design and optimisation.

The earliest and most tedious method of attempting to simulate a flotation circuit was through the use of a 'Locked Cycle Batch Tests'. This test is a laboratory batch simulation of the performance of a continuous circuit. Successive rougher batch tests are carried out on replicate feed samples. The cleaner tail from the previous test is added to the following rougher flotation. The recleaner tail from the previous test is added to the cleaner flotation of the following test. Cycles are repeated until the analysis of rougher tails and recleaner concentrates stabilise. These tests are labour intensive, and the results obtained are often inconclusive in terms of the prediction of full-scale plant performance. However, Agar (1989) stressed that a lock cycle test program is an essential part in the design of a new flowsheet synthesis, and proposed a technique for simulating a continuous flotation circuit from batch test data. This was achieved by assigning "split factors" to each of the components in a separation stage obtained from the batch test. The split factors were defined as the distribution of a component in the non-float portion, being based on the feed to each separation stage in the circuit. The split factors were assumed to be functions of flotation chemistry, mechanical conditions and flotation time and fixing these would fix the split factors for a given separation stage in a circuit. Using the split factors in a set of algebraic equations defining the mass balance, it was possible to simulate a flotation circuit. This modelling approach can be put onto a firmer base by applying one of the flotation models discussed previously to predict the split factors across a separation stage. The complexity of the model used will determine the extent to which the simulation is able to predict changes in the variables of each cell. Agar (1989) attempted this procedure and assigned a modified first order rate equation to the batch test data rather than treating the separators as "black boxes" in which a split is made. Fitting a first order rate equation to the batch test data, it was demonstrated that the optimum time in a given separation stage could be specified. This procedure was confirmed by comparing the simulated prediction using the above technique and comparing it to an experimental locked cycle batch test results of the simulated circuit. This simulation method unfortunately cannot predict changes in the process variables of the individual cells in the simulated circuit.

King (1973) has presented a complex mechanistic model for the design and control of flotation plants. The model allows for a variation in feed properties and grind and can be used to evaluate the effects of a variation of the control variables on plant performance, viz. reagent addition, aeration rate and pulp level. In addition, the model provides for a distribution of rate constants, particle size and mineral fraction. The main assumption in the model development is that there are three distinct properties of the particles that influence the rate of flotation significantly, particle size (D), mineral composition of the particle (g) and the surface activity (k) and these properties are distributed over the individual particles in the particle population. The mineral composition (g) is a vector of values representing the fraction of each mineral type in the particle. On this basis the equation that describes the flotation response is:

$$M_F f_F(k, g, D) = k \phi(D) A S_{av} W f_T(k, g, D) \dots\dots\dots 2.54$$

where

M_F	is the total mass flow rate of floated material (kg/s)
$f_F(k, g, D)$	distribution of particle types in the floated material
k	specific rate constant (m/s)
$\phi(D)$	fractional efficiency of impaction, adhesion and levitation for particles of size D
A	bubble surface area per unit volume pulp (1/m)
S_{av}	average fraction of bubble surface that is uncovered
W	weight of solid material in the flotation cell (kg)
$f_T(k, g, D)$	distribution of particle types in the tailings stream

The flow rates of water in the simulated streams were noted as being important for the calculation of residence times, which were needed to calculate the material flows. The water flow rates were related to the mass flow rates assuming that the specific gravity of the concentrate streams from all the cells could be predicted or measured. The computational scheme can be found in the literature, (King, 1973). The main drawback of this approach is the difficulty of measuring many of the parameters in the model, although this type of model could certainly become more relevant when liberation analyser's become more accessible and widely used.

Kawatra et al, (1982) developed a computer simulation of an iron ore flotation circuit from data measured on the circuit. The flotation rate behaviour of each significant component was considered separately in the model. The mineral flotation model was based on a first order rate equation which was distributed by size class similar to that proposed by Tomlinson and Flemming (1963) presented in section 2.2.3.1.3., with each size fraction having its own first order rate constant. The water recovery from each flotation cell was linearly related to the valuable mineral recovery rate. The gangue minerals (non-floating) were considered to be recovered purely by entrainment with water as a function of size class. The degree of entrainment was modelled discretely using separate rate constants for each entrained particle size fraction. The flotation cells were approximated as perfectly mixed vessels and the retention time in the pulp was assumed to be approximated by the volume of the flotation cell divided by the tailings volume flow rate from each cell in the circuit. The distinguishing feature in this model is the inclusion of a froth overloading factor for both the mineral and gangue models, limiting the rate of the concentrate flow from a cell caused by the saturation of bubbles with solid particles, similar to that discussed in section 2.2.3.1.5. Each flotation cell in the circuit was assigned a unique froth overloading factor. The residence times were calculated iteratively in each flotation cell to provide a solution to the model mass balance around the circuit. The froth overloading model parameters as well as the rate constants for the minerals and gangue were determined empirically from experimental data. It was acknowledged that using a constant froth overloading factor for each flotation cell limited the accuracy of the model. It is also believed to limit the simulation

of alternative flowsheets. The valuable mineral rate constants were assumed to be constant throughout the circuit and it is possible that the froth overloading factors masked the differences in the rate constants for different units in the circuit. The correlation of the valuable mineral recovery rates to experimental data was reasonably accurate, but the model was much less accurate at simulating the water and gangue mineral recovery rates. The changes in the froth characteristics from cell to cell was highlighted as one of the major reasons for the large fluctuations in the water recovery rates observed in the circuit. It was concluded that froth depth measurements in each flotation cell had to be incorporated into the model to explain the unexpected deviations in the water recovery rates from the experimental data. The mineral rate constants were not decoupled and therefore this model is not able to distinguish between mineral, flotation machine as well as froth effects.

In 1984, Woodburn and Wallin proposed a decoupled kinetic model to simulate the tailings flow in a complex network of flotation cells. Their model was based on the assumption that the floatability of solids species in the feed was distributed according to experimentally determined functions. For example, the distribution of valuable minerals was represented by a continuous function whereas the gangue could be modelled by discrete classes. Water was regarded as a separate species. The model suggested used the simultaneous calculation of a mass balance of all species over the circuit to determine the flow rates of the tailings streams. In order to achieve this, they made use of a matrix of enhancement factors, g_{ir} , which were defined as the ratio of concentrate flow of sub-class i of a particular species r from cell l to the tails flow of that sub-class of that species from the same cell. The enhancement factors were calculated, assuming first order kinetics to hold, similar to the approach of Agar (1987) except the split factors were obtained directly from the plant and not from batch test data.

$$g_{ir} = k_{ir} \tau_l \dots\dots\dots 2.55$$

where

k_{ir} is the first order rate constant of sub-class i of mineral r in cell l
 τ_l is the residence time of pulp in cell l

Water removal was assumed to be zero order. The first order rate constant for mineral flotation was divided into two non-interacting components: kinetic effects and equilibrium effects. Kinetic effects were taken as those related to the specific aeration rate in the cells, and equilibrium effects as those related to the distribution of a solid species between the pulp phase and the liquid-air interface. In other words, the kinetic effects were characteristic of the cell, and the equilibrium effects represented the floatability of the mineral. It was assumed that the pulp phase behaved as a perfectly mixed reactor. The enhancement factor was thus decoupled as discussed in section 2.2.3.1.6 and represented by:

$$g_{ifl} = K_{ifl} \chi_l \tau_l \dots\dots\dots 2.56$$

where

K_{ifl} is the floatability of sub-class i in cell l

χ_l is the kinetic effect related to the specific aeration rate of cell l

The concentration of frother in the system was found to affect the bubble size and thus the specific aeration rate. χ was thus related to frother concentration. The rate of froth flowing over the concentrate launder, which was used to calculate the enhancement factor, was dependent on the froth stability. It was also dependent on the specific aeration rate of the cell. χ was thus also affected by froth stability. Using data generated in a laboratory batch flotation cell, Woodburn and Wallin found that the χ values depended on the amount and nature of solids in the froth. This finding was not consistent with the model, which assumed that χ would be constant for all species in a cell. Their model thus could not take into account the dependence of the froth stability on the nature of the solids in the froth. They found that no simple relationship existed between frother concentration and the froth stability parameter incorporated in χ , and indeed between frother concentration and χ itself. This modelling methodology could be useful for predicting the performance of a flotation circuit that does not treat a complex ore. However, for the flotation of complex sulphide ores the effect of the mineral on the froth would necessitate the use of a better defined froth factor.

Loveday and Raghubir (1995) looked at the design and optimisation of flotation circuits using simulation methods. In a review of current methodologies, they concluded that, although batch testing could give one a good indication of the effect of chemistry on flotation, batch testing alone was not suited for use in the design of complex circuits. They also recognised that pilot plants were an improvement on bench scale tests for complex ores, but were still not the ultimate test vehicle. An ideal situation, they concluded, would be for test work to be done on the circuit in question. This would allow results to be extrapolated to other circuit configurations with confidence. From their test work on a multi-stage flotation circuit, they proposed a model for circuit simulation in which the recovery of each mineral from each stage was a function of the residence time. It must be noted that, although bubble loading, froth depth, bubble size and frother concentration among others would also influence the recovery, it was assumed that all particles of a particular mineral would be affected by these variables to the same degree. This would be true if the minerals were divided into narrow size fractions and each size fraction treated individually. The variation in floatability of the mineral species must be taken into account. The flow of pulp in a bank of flotation cells was assumed to approximate plug flow conditions if the bank consisted of many cells. For a single cell, the pulp was assumed to behave as a perfectly mixed vessel. These assumptions could be checked by performing a residence time distribution study on the circuit. Loveday and Raghubir performed batch flotation tests on samples gathered from each stream in the circuit. This gave a good indication of the performance of the various stages in the circuit. They assumed a rectangular distribution of floatabilities (Klimpel model) to fit their batch flotation data of the mineral, linking all the curves into a single model. Once the flotation rate constants were calculated by the Klimpel model, they used the

residence time to predict the volume of the cell needed for the given throughput. Changes in circuit configuration could thus be modelled using the ratio of batch flotation time and cell residence time as a scale-up factor. However, the model does not allow for changes in grind or circuit chemistry to be predicted without further extensive and expensive test work. The modelling approach, although easily adaptable and practical, does not explicitly incorporate the influence of the froth phase kinetics and is thus limiting in being fully predictive for design purposes.

The flotation circuit simulation methodology proposed by Harris (1997) in collaboration with other researchers assumes that each particle maintains its characteristic floatability irrespective of the location of the particle in the circuit and is therefore independent on circuit configuration. The methodology is based on the scale-up of flotation variables using batch flotation and plant data. Extensive mass balance campaigns of mineral species and water, as well as batch flotation tests of the feed, concentrate and tailings of each cell in the circuit are required for the model development. In addition, the gas dispersion characteristics in each cell, the froth height and the residence time of pulp in each cell must be measured. A distributed floatability parameter model is used to describe the kinetic response to flotation of the floatable minerals, similar to the approaches of King (1973) and Woodburn et al (1984) discussed previously. The rate constant is decoupled according to Gorain et al (1995), (discussed in section 2.2.3.1.6) including the effect of the froth phase, ore characteristics and flotation machine hydrodynamics (S_b). The model for a continuous plant cell with perfect mixing in the pulp is expressed as:

$$R_{j,k} = \sum_{i=1}^n m_{i,j,k} \left(1 - \frac{1}{1 + P_{i,j} S_{b,k} R_{f,k} \tau_k} \right) \dots\dots\dots 2.57$$

and that of a batch cell, assuming plug flow conditions as:

$$R_j = \sum_{i=1}^n m_{i,j} (1 - e^{-P_{i,j} S_b t}) \dots\dots\dots 2.58$$

where

- $R_{j,k}$ is the recovery of mineral j in cell k
- $m_{i,j,k}$ is the mass fraction of floatability class i of mineral j in cell k
- $P_{i,j}$ is the floatability of class i of mineral j
- $S_{b,k}$ is the superficial bubble surface area flux in cell k
- $R_{f,k}$ is the froth recovery factor in cell k
- τ_k is the residence time of the pulp in cell k
- n is the number of floatability components

The relation between the batch and plant floatability rate constants is usually achieved using a scale-up parameter. The basis of this is similar to the ratio approach of Loveday et al (1972). The kinetics of the froth phase is incorporate using the froth recovery factor which is usually considered to be 1 in the batch flotation cell. This is because the batch

flotation tests are carried out with a shallow froth depth. This is not the case for the plant flotation cells and a method developed by Harris (1998) suggests that froth recovery is dependent on the froth retention time (FRT) as described by equation 2.25. In modelling the gangue mineral behaviour, the entrainment model proposed by Savassi (1995) as discussed in section 2.2.3.3.1 is used. The behaviour of water is modelled using King's model described in section 2.2.3.4.1.

In equations 2.55 and 2.56, FRT, S_b and τ can be measured on the plant. The remaining variables are determined by regression techniques, minimising the model output to experimental data. The froth retention time is obtained by measuring the concentrate flow from the unit or bank in question, and the superficial bubble surface area flux found by measuring the superficial gas velocity and bubble size in the cell. The residence time, τ , can be determined by direct experimentation on the plant by performing a residence time distribution study, or a mass balance around the circuit. This simulation methodology has been shown to work quite well for fixed circuit configurations or for minor adjustments to the circuit used to develop the simulation model. This is due to limitations of the water recovery model in accurately predicting the water kinetics for a change in circuit configuration due to the empirical nature of this model. The empirical nature of the froth recovery model also restricts the application of this simulation methodology in accurately predicting alternative circuit configurations but is capable of predicting the change in froth phase performance as a function of operating variables in a particular flotation cell in a circuit.

2.2.5.2.3 Discussion

The literature associated with flotation plant simulations is not extensive, although several groups of workers have been active in the development of simulation programs. In contrast, the literature associated with modelling of the flotation process is extremely extensive. The approach of King (1973) seems to be the most detailed method of simulating a flotation circuit but unfortunately is not very practical, being sometimes difficult to obtain the necessary parameters in this model. The modelling methodology of Harris (1997) seems to be more practical in this respect as all the variables in the model are based on easily measurable parameters that can be obtained directly off operating industrial plants. Woodburn and Wallin (1984) simulation methodology recognises the influence of the froth phase performance but they could not explicitly include the dependence of the froth stability on the nature of the solids in the froth, while Harris' (1997) methodology has been shown to incorporate this effect through the use of a froth stability parameter as a function of froth residence time. The development of recent models incorporating the influence of the froth phase more accurately will hopefully resolve the controversy around the different rate constants in different parts of a flotation circuit. In addition, it is clear that difficulties associated with the prediction of water recovery are a substantial obstacle to further progress in this area.

2.2.5.2.4 Measurement and circuit analysis techniques

This section briefly presents the most important measurement and circuit analysis techniques used when modelling a flotation circuit in an attempt to apply a simulation methodology to an existing plant. This section briefly covers aspects relating to sampling, cell hydrodynamics and residence time distribution in a flotation cell as well as testing the conservation of floatability via a nodal analysis, and the extraction of kinetic parameters from circuits.

a. Sampling

The basic rule for correct sampling is that each particle of ore or concentrate must have an equal probability of being collected and becoming part of the final sample for analysis. For effective representivity a complete cross-section of the ore stream should be taken with each sample cut. The size of the representative sample (M) to give the required accuracy can be calculated using Gy's method (Gy, 1982). Gy's basic sampling equation can be written as:

$$M = \frac{Cd^3}{s^2} \dots\dots\dots 2.59$$

where

- C is a sampling constant (g cm^{-1})
- d is the diameter of the biggest particle in the ore to be sampled (cm)
- s is a measure of the statistical error committed by sampling

Assuming a normal distribution, 67 out of 100 assays of samples would lie within $\pm 2s$ of the true assay, and 99 out of 100 would be within $\pm 3s$ of the true assay. As sampling is a statistical problem, there can never be complete confidence in the result of a sampling exercise, and for most practical purposes, a 95% confidence level is generally adopted as an acceptable probability level (Wills, 1988).

The sampling constant C is specific to the material being sampled, viz.

$$C = fgl\left(\frac{\rho}{a}\right) \dots\dots\dots 2.60$$

where f is a shape factor, which is taken as 0.5 and g is a factor that depends on the particle size range. If approximately 95% of the sample mass contains particles of size less than d_z cm and 95% of size greater than d_x cm, then if:

$d_z/d_x > 4,$	$g = 0.25$
d_z/d_x is 2-4	$g = 0.5$
$d_z/d_x < 2$	$g = 0.75$
$d_z/d_x = 1$	$g = 1$

where p is the relative density of the precious mineral, a is the grade of the precious mineral in grams per ton of ore and l is a liberation factor which can be calculated from the expression:

$$l = \left(\frac{d_l}{d}\right) \dots\dots\dots 2.61$$

where d_l is the liberation size of the precious mineral in centimeters.

b. Cell operating characteristics

The different modelling and simulation procedures presented in section 2.2.5.2.2 required the measurement of certain cell operating variables in the simulation models. The most common measurable cell operating variables include the gas rate, bubble size, gas hold-up and superficial air flow rate (J_g). These measurements are used to calculate the bubble surface area flux (S_b) which has been described as a measure of the hydrodynamic conditions in a flotation cell and can be effectively used in a kinetic model of a flotation cell. Methods of obtaining these measurements are presented below.

The air flow rate into a flotation cell is most commonly measured using a standard variable area air flow meter connected to the flotation cells air inlet pipe. The variable flow meter consists of a tapered metering tube and a float which is free to move up and down inside it. The tube is mounted vertically with a small opening at the bottom. At any flow rate within the range of the meter, air entering the bottom of the tube raises the float increasing the area between the float and the tube until the upward hydraulic forces are balanced by the downward force of the float. The air flow rate is indicated by the position of the float on a graduated scale. The measurement of bubble size can be obtained using the UCT Bubble Size Analyser. A detailed description of the Bubble Size Analyser can be found in Tucker et al, (1994). In principle, the liquid containing the bubbles is sucked through a capillary. Two electronic eyes on the capillary give an indication of the bubble sizes and the number of bubbles passing per unit time. The recorded bubble size data are processed statistically by a computer connected to the instruments.

The gas holdup in a flotation cell is a measure of the fractional volume occupied by the air bubbles in a flotation cells pulp. Gas holdup in flotation cells can be measured using different techniques, (Finch and Dobby, 1990). Most of these techniques measure the

“overall holdup” in a flotation cell or the “local holdup” over a given section in a column cell and are not quite suitable to measure “local holdup” in different locations in a mechanical cell. However, the technique of encapsulating a sample of aerated slurry, developed by Jameson and Allum (1984), is a practical and useful technique to measure gas holdup in different locations in an industrial size mechanical flotation cell. This instrument consists of a copper cylinder with two plungers attached to a central rod. The plungers have O-rings for air tight fitting when they move inside the cylinder. The cylinders are actuated by using a vacuum pump. The vacuum draws the plungers into the cylinder, encapsulating a volume of aerated pulp in the space between the plungers. The aerated pulp encapsulated between the plungers is emptied into a measuring cylinder at which point the air escapes into the atmosphere. The volume of the space between the cylinder is determined by calibration with water. The volume of air in the aerated pulp is calculated as the difference between the measured volume of slurry and the calibrated volume of water. The gas holdup is then the volume fraction of air in each sample.

A device used to measure the superficial air flow rate (J_g) consists of an extendible graduated perspex tube with a pneumatically controlled pinch valve and water inlet valve at the one end and an air outlet valve at the other. The tube is placed inside a flotation cell and the pinch valve opened to allow the air bubbles in the cell to move up into the tube. The time taken for water to traverse a known distance is measured, from which the value of J_g can be calculated in centimetres per second. Another method of approximating J_g is to assume that the air bubbles are evenly distributed over the cross section of the entire flotation cell. J_g can then be approximated as being the total volumetric air flow rate introduced into the flotation cell divided by the cross-sectional area of the flotation cell.

The superficial bubble surface area flux (S_b) represents the gas dispersion characteristics in a cell as discussed in section 2.2.3.1.6. S_b can be determined from measuring the average bubble size and the superficial air flow rate in a cell. Recent investigations have been conducted to develop a model that can be used to predict S_b in a cell as function of its operating conditions. The advantages of such a model is that it would be useful for cell characterisation, (Gorain, 1997).

$$S_b = a * N_s^b * \left(\frac{Q}{A}\right)^c * A_s^d * P_{80}^e \dots\dots\dots 2.62$$

where

- N_s impeller peripheral speed
- Q/A air flow rate per unit cell cross-sectional area
- A_s impeller aspect ratio (impeller diameter to impeller height)
- P_{80} 80% passing feed size

This model has been empirically tested from data obtained at the Hellyer and Scuddles Concentrator, (Gorain, 1995). In total, 100 sets of experimental values of S_b , N_s , Q/A , A_s and P_{80} were obtained over a wide range of operating conditions to obtain the parameters in the above model. The final form of the model, with the empirically fitted parameters was found to be:

$$S_b = 123 * N_s^{0.44} * \left(\frac{Q}{A}\right)^{0.78} * A_s^{-0.1} * P_{80}^{0.42} \dots\dots\dots 2.63$$

The model was validated on a 100 m³ Outokumpu tank cell treating zinc ore in scavenger operation at the Pasminco Broken Hill Concentrator. A restriction on the above model is that the fit to experimental data for S_b values higher than approximately 100 m²/m².sec was found to be poor. This corresponds to cell operating conditions for which the cell contents were visually found to be flooding, outside the normal range of operating conditions of flotation cells.

c. Residence time distribution

The residence time distribution (RTD) of a reactor is a characteristic of the mixing that occurs in a chemical reactor. The RTD can be determined experimentally by the injection of an inert tracer and measuring the tracer concentration in the effluent stream as a function of time. RTD studies are extremely useful for diagnosing malfunctioning or non-ideal flows in flotation cells (Mavros, 1992). RTD studies provide a sound base for the metallurgist to detect inefficiencies in large continuous machines, inevitable day-to-day circuit disruptions and gradual equipment deterioration. The application of RTD studies in the evaluation of conventional flotation machines is not well documented in the literature. Nasset (1988) applied RTD studies with great success in the analysis and improvement of existing flotation equipment. The author highlights the fact that although the techniques of RTD analysis have been well developed by the chemical engineering profession, it has not been applied to the same extent by mineral process engineers. The method outlined in the paper permits the mineral process engineer to evaluate various process and economical trade-offs when selecting the number of cells and the size. The detailed discussion on tracer characteristics and selection is probably the most valuable information in the paper.

d. Nodal analysis

Most simulation models assume, of necessity, that under conditions of steady state with no regrind stages or reagent addition points in the flotation circuit, that the floatability of the ore should not change with time. The extent to which the floatability of the mineral components in the ore remains constant can be evaluated by considering the results of batch floatability tests conducted around nodes in the circuit. A circuit node represents any part of the circuit where two or more streams are combined, or separated. If

floatability is indeed conserved around a node, then the response of one of the streams around the node can be determined mathematically by combining the data from the other streams around that node, (Runge et al, 1997). This is done according to equation 2.64. The calculated recovery time curve can then be compared to the actual experimental data for that stream. This is illustrated in Figure 2.10. For the circuit under investigation, the recovery versus time curve for the feed can be calculated from the data from the final concentrate and final tailings, and this compared to the actual recovery data.

$$R_j(v) = \frac{\sum_{s=1}^x \sum_{u=1}^y w_{u,s} a_{u,j,s} f_s}{\sum_{s=1}^x \sum_{u=1}^y w_{u,s} a_{u,j,s} f_s} \dots\dots\dots 2.64$$

where

- $R_j(v)$ is the cumulative recovery of the j^{th} mineral in the v^{th} concentrate in the combined stream
 $w_{u,s}$ is the weight % recovered in concentrate u during flotation of stream s
 $a_{u,j,s}$ is the assay of mineral j in concentrate u during flotation of stream s
 f_s is the mass flow rate of stream s
 x is the total number of streams comprising the combined stream
 y is the total number of samples collected during the flotation, including the tailings

It is necessary to ensure that all the batch tests are conducted at exactly the same set of operating conditions ie: same air flow rates, froth heights, power inputs and agitation speeds. All these ensure that the flotation rate constants are the same (or universal) for every batch float and the only factor affecting the floatability response between the different tests are the relative fractions of the floatability constituents in each stream. A shallow froth height should be used as this minimizes the influence of the froth zone on the batch tests. The samples taken should be taken "hot" from the plant, and not be allowed to stand for prolonged periods of time. This eliminates possible sample degradation by oxidation and allows a truly representative sample to be taken from the plant and tested. No chemicals should be added to the batch tests.

This method of nodal analysis assumes that the various floatability components do not interfere with one another in various parts of the plant as well as within every batch floatability test. This assumption has still not been validated, but should not be of major concern with very dilute mineral pulps ie: when only a small amount of mineral actually is available to get recovered as in the case with PGM ores. In pulps where there is a high concentration of recoverable mineral, sometimes a phenomenon known as bubble overloading occurs where all the available bubble surface area is used up by the floating minerals, this slows down the flotation rate as then it is not a function of how much mineral is in the cell but rather a function of the maximum available bubble

surface area, (Tomlinson and Flemming, 1963), (Whiten et al, 1972), (King et al, 1973), (Thorne, 1976). For PGM ores this is not a major concern as the concentration of floatable minerals are very small, (Liddell et al, 1986). The only possible occurrence of a slight bubble overloading would be in the first cleaner concentrate where there is a higher concentration of fast floating mineral in the cell.

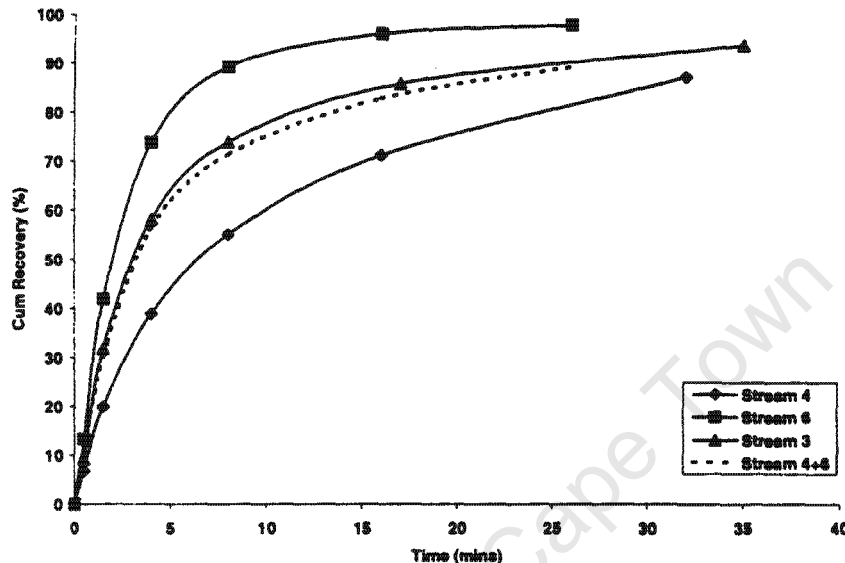


Figure 2.10: Comparison of a typical mineral versus time curve around a plant node, (Runge et al, 1997)

e: Extension of nodal analysis to kinetic parameter estimation

The graphical technique of determining the conservation of floatability components around a stream mixing node as outlined in the previous section can be extended to analysing the conservation of floatability components around all nodes in a circuit of flotation cells. Harris (1997) uses this assumption to determine the flotation model parameters for a feed stream to a circuit. In a batch laboratory flotation test in which the feed contains n floatability components, the flotation recovery achieved in any specified time period assuming a distributed floatability modelling approach (for an ideal batch system) is given by equation 2.65.

$$R_{j,s} = \sum_{i=1}^n m_{i,j,s} [1 - \exp(-k_{i,j}^{\text{batch}} t)] \dots\dots\dots 2.65$$

where

$R_{j,s}$ is the batch recovery at time t of mineral j from stream s

n is the number of floatability classes

$m_{i,j,s}$ is the mass fraction of mineral j in floatability class i in stream s

$k_{i,j}^{\text{batch}}$ is the batch rate constant associated with floatability class i of mineral j

The problem with this distributed floatability parameter model is the large number of parameters that have to be fitted to the flotation data, even if the number of floatability classes are small, i.e.:

$$P = 2n - 2 \quad \text{where one floatability class is non-floating (i.e. } k=0)$$

or

$$P = 2n - 1 \quad \text{where there are no non-floating mineral particles}$$

where

P is the number of floatability model parameters

It has been observed by a number of authors that most minerals typically require either 3 or 4 floatability classes to adequately describe their flotation behaviour, (Harris, 1997). It is clearly not possible to extract such a large number of parameters from experiments conducted on a single ore sample. There would be insufficient information to ensure that the determined parameters were unique, rather than just one set of many possible combinations of parameters that fit the data, but lack any physical significance. The problem can be overcome by considering the response of an ore in a flotation circuit. The mineral content of the various streams within a flotation circuit represents the outcome of a fractionation process of the mineral originally in the feed to the circuit, based on its original floatability distribution under the physical and chemical conditions that exist within the circuit. This implies that if the circuit is operating under constant, steady state conditions, and the floatability of the ore is not changing with time, then the floatability distribution of all the streams within a flotation circuit can be directly related to the "parent" floatability distribution of the mineral in the feed ore, (Harris, 1997). This fact can be used to provide the information required to extract the floatability model parameters, by performing "bench mark" floatability tests on the streams within a flotation circuit, using a batch flotation cell. Data obtained on a variety of flotation circuits from all over the world, treating a variety of ores, suggests that the conservation of floatability appears to be far more the rule than the exception, (Harris, 1997, Runge et al, 1997, Alexander et al, 1997, Theron et al, 1998).

As an example, consider the case where there are four floatability components, the parameters required are, fast fraction (k_f, m_f), medium fraction (k_m, m_m), slow fraction (k_s, m_s) and a non-floating fraction where $k_o = 0$ and $m_o = 1 - m_f - m_m - m_s$. This requires six parameters to determine (or fit) the recovery time profile response of a batch flotation test. Assume that concentrate samples are taken every 1, 3, 5, 7, 10 and 15 minutes in the batch test. This produces six data points to fit six parameters for every batch test, therefore relatively little confidence can be placed on the parameter set obtained and their significance would be questionable.

Number of data points	(N)	=	6
Number of parameters	(P)	=	6

If two different streams from the circuit are subjected to batch floatability tests, there would be 12 parameters necessary to predict the recovery time response of both profiles. However, the assumption that the floatability components rates are constant throughout the circuit reduces the necessary parameters by three because the $k_{i,j}$ are the same for both recovery expressions. Table 2.3 shows how the ratio between the number of parameters and the number of data points decrease as the number of batch floatability tests (h) are increased around a circuit. Equation 2.66 describes this relationship.

Table 2.3: Increase in the parameter confidence level with increase in the number of batch tests conducted

NUMBER OF BATCH TESTS	1	2	3	4	5	6
Data Points	6	12	18	24	30	36
Parameters	6	9	12	15	18	21
Ratio (P/N)	1	0.75	0.67	0.63	0.6	0.58

$$\left(\frac{P}{N}\right)_h = \frac{3(h+1)}{hc} \quad \dots\dots\dots 2.66$$

c number of concentrate samples taken during each batch test

The problem can be constrained even further by remembering that the floatability component flow rates have to comply with the mass balance constraints within every cell of the flotation circuit. Another constraint is that the fractions of floatability components within a stream must add up to one. Equations 2.67 to 2.69 show the mathematical equivalent of the constraints described above.

$$\sum_{i=1}^n m_{i,j,s} = 1 \quad \dots\dots\dots 2.67$$

and

$$M_{i,j,s} = f_s a_{j,s} m_{i,j,s} \quad \dots\dots\dots 2.68$$

and

$$M_{i,j,z}^{in} = M_{i,j,z}^{out} \quad \dots\dots\dots 2.69$$

where

- $M_{i,j,s}$ is the mass flow rate of mineral j in floatability class i in stream s
 $M_{i,j,z}^{in}$ is the total mass flow rate of mineral j in floatability class i in all streams entering circuit node z
 $M_{i,j,z}^{out}$ is the total mass flow rate of mineral j in floatability class i in all streams exiting circuit node z

Therefore, once it has been established that the floatability of the mineral components in the circuit does not change with time, the flotation response of all the streams can be combined to determine the required modelling floatability parameters. This is done by fitting the stream batch floatability data to equation 2.65, subject to the constraint of equation 2.67, using an appropriate non-linear optimisation procedure, based on sum-of-error-squared minimisation. The determination of the model fit parameters using equations 2.65 to 2.69 can be difficult, particularly for complex circuits, and larger values of n , as the sum-of-error-squared minimisation procedure frequently requires good initial parameter estimates in order to achieve convergence. In addition, the determination of an acceptable value of n must be established on a statistically sound basis. This problem has been addressed by Alexander and Morrison (1997), who have developed a relatively simple, linear mathematical approach capable of providing good initial parameter estimates for use in the above modelling procedure. Their method also provides a useful statistical basis to establish the number of floatability fractions, n , that can be justified.

2.3 Platinum ores

This section of the review discusses the mineralogy of the UG2 platinum reef from the Bushveld Igneous Complex in South Africa as well as an overview on the process routes used to concentrate the ore. Common industrial platinum flotation circuit designs are briefly described. Finally, the literature associated with previous modelling and simulation attempts on platinum ores is presented.

2.3.1 Background and overview

There are many rich reserves of precious and base metals that are recovered by various process routes in South Africa. Some 2000 million years ago, enormous volumes of molten magma burst into the upper layers of the earth's surface in an area covering most of South Africa's present-day North West, Gauteng, Mpumalanga and Northern Province. This magma cooled slowly allowing the formation of different mineral zones in the form of layered saucer-shapes to accumulate in a process known as magmatic differentiation. These layers are known today as the Bushveld Igneous Complex. This batholith of igneous rock contains a rich supply of metals, most importantly, more than eighty percent of the world's known reserves of platinum group metals (PGM). The PGM comprise platinum, palladium, rhodium, ruthenium, iridium and osmium. The PGM in the Bushveld Complex are found in a number of different layers, of which the Merensky and the Upper Group 2 chromitite seam (UG2 reef) are of

greatest economic importance. The Merensky reef is estimated to contain some 17 million kg of PGM's, while the Upper Ground 2 Reef contains reserves of at least 31 million kg of PGM's, (Corrans et al, 1982). The mines in the area have exploited the Merensky Reef for the past sixty years or so, while the UG2 reef was not exploited until recently because it generally contains less PGM per ton than ore from the Merensky Reef as well as a large percentage of chromite. The chromite can cause problems during the subsequent smelting process. At a certain stage, mines had to compare the economics involved in deeper development of the Merensky Reef with economics involved in horizontal development of the UG2 Reef. The latter choice became more attractive with time, as suitable treatment routes for UG2 ore were identified.

2.3.2 Mineralogy of UG2 reef ore

The UG2 reef, a platiniferous chromitite seam, lies 150 metres below the Merensky Reef in the western Bushveld and 330 meters below it in the east, (Liddell et al, 1986). This reef is between 15 and 255 centimetres thick, and consists primarily of chromite (30 to 90%), gangue silicates (5 to 25%), orthopyroxene and plagioclase. Trace amounts of base metal sulphides (BMS) occur mainly interstitial to the chromite grains (<1%).

Table 2.4: PGM minerals present in the UG2 Reef according to Louwrens and McLaren (1990)

MINERAL NAME	CHEMICAL COMPOSITION
Cooperite	PtS
Braggite	(Pt,Pd,Ni)S
Laurite	RuS ₂
Sperrylite	PtAs ₂
Pt-Ir Alloy	Pt-Fe
Moncheite	PtTe ₂
Merenskyite	PdTe ₂
Os-Ir alloy	
Potarite	PdHg
Genkinit	Pd-Pt-Sb
Vysotskite	PdS
Pt-Pd alloys	
Telluride	Pt-Pd
Plumbopalladinite	Pd ₃ Pb ₂
Majakite	PdNiAs
Stibiopallandite	Pd-Sb
Pt-Ag-Se Alloy	
Pt-Rh sulphide	

In this ore, the major BMS present are pyrrhotite, pentlandite, chalcopyrite with millerite and pyrite found in minor amounts, (Liddell et al, 1986). The BMS grain sizes are typically around 30µm in diameter. Mineralogical studies have indicated that 70 to 80 percent of the PGM in UG2 ore are typically associated with the BMS as occlusions and intergrowths, 10 to 20 percent are disseminated in silicate gangue and of the order of 5 percent in chromite chats. The PGM minerals may also occur as single grains and aggregates along the grain boundaries of two or more chromite grains, or along chromite grains in contact with gangue particles. The grain size of PGM minerals is smaller than that of the BMS, typically 15µm. The head grade of UG2 ores typically varies from 4 to 10 g/t.

2.3.3 Processing of the UG2 Reef

2.3.3.1 Overview

A typical process route to remove PGM and base metal sulphides (BMS) from the UG2 ROM ore is described below. The ore is finely ground, preconcentrated by flotation, the flotation concentrate is smelted and the base metals removed. The noble metals are then separated from each other and the individual metals refined. In the comminution stage, rod mills, ball mills or autogenous mills are used to grind the ore until there is sufficient liberation of the desired minerals for them to be amenable to preconcentration. Flotation is used in the preconcentration stage of PGM ores, unlike the gold-bearing ores that employ a direct leaching stage. In order to leach PGM ores, the ore would have to be chlorinated before being subjected to a leach under very harsh conditions. This process would be neither economically viable nor environmentally safe (Liddell et al, 1986). The flotation of the fine ore typically concentrates the PGM from a grade of about 4 g/t to about 100 g/t. After smelting it reaches approximately 2000 g/t or 0.2 percent. Removal of the base metals and refining results in a product that is typically between 99.95 and 99.99 percent pure, depending on the metal.

2.3.3.2 Platinum concentration circuits

Extensive work was done by the Council for Mineral Technology (Mintek) in the late seventies and eighties in comminution, flotation and smelting of UG2 ores. There are two types of comminution circuits used in the processing of UG2 ore today. Some mines use crushing and milling circuits while others mill run-of-mine ore. The circuits with crusher plants use either a ball or rod mill as a primary mill (Goodall, 1995). Rod mills have the advantage of a low steel consumption and a narrow product size distribution. The major drawback of rod milling is that it cannot produce a grind finer than 40% passing -75µm when it is used as a primary mill (Goodall, 1995). The newer UG2 plants tend to mill run-of-mine ore because it is cheaper. A grind of 80% - 75µm, necessary to liberate the PGM, can be achieved by using one or more milling stages.

Goodall (1995) reports that single stage, two stage and even three stage milling are in use in UG2 comminution plants.

The flotation circuits employed for the UG2 reef are designed essentially for the flotation of base-metal sulphides (BMS). Flotation takes place at a natural pH, between 7.5 and 9. A xanthate collector such as isobutyl xanthate or normal propyl xanthate is used. The general practice on most mines is the addition of another collector, usually dithiophosphate, which is mixed with the xanthate (Liddell et al, 1986). Copper sulphate is typically added as an activator, either to the milling circuit or at the head of the float. Hochreiter et al (1985) report the use of cresylic acid, alone or in mixtures with alcohol-base frothers, as the most common frother. The current tendency is to move away from cresylic acid to frothers that are easier to handle and more consistent in composition. Owing to the talcose nature of the ore, use is made of gangue depressants, which is normally an organic colloid. There are different types of depressants used in the flotation of UG2 ores, the choice of which depends on the primary gangue mineral in the ore. The flotation circuit layout for the treatment of UG2 ores is much more complex than for the Merensky ore where only a single flotation stage is often required. The large amount of chromite in UG2 ores is a problem as the content of chromite in the flotation concentrate needs to be minimised (typically < 3%) to prevent problems during smelting. The selection of appropriate frothers and depressants help maintain this level, but more importantly the design of complicated flotation circuits are aimed specifically at reducing the amount of chromite into the concentrate while still maintaining high PGM recoveries. It is widely believed that the chromite is recovered primarily through physical entrainment with water. Therefore, a rougher and cleaning circuit with a long residence time to enable flotation of the slow floating PGM's and multi stage cleaning with a brittle froth, to reduce the amount of physical entrained particles are typically used. The fine sulphides in UG2 ores have been found to have slower flotation rates, therefore the cleaner circuits need longer residence times to prevent recirculation to the roughers and consequently, preventing valuables lost to the tailings stream.

There are several different milling and flotation configurations used on the platinum concentrators today. Goodall (1995) describes four configurations, viz.:

1. Mill-Float (MF1)
2. Mill-Mill-Float (M2F)
3. Mill-Float-Mill-Float (MF2)
4. Mill-Float-Mill-Float-Mill-Float (MF3)

The most common of the four is the MF2 configuration. The primary milling circuit would usually produce a grind of 40-50 % - 75µm to enable the flotation of already liberated particles before over-grinding occurs, particularly with respect to the chromite. A rougher and scavenger with two or three stage cleaning is common practice for the primary float. The tails from the primary section are usually reground to 80% - 75µm to liberate the PGM's in the fine sulphide grains (< 15µm). The secondary float usually has

more cleaner capacity than the primary since the presence of slower floating minerals is higher than in the primary float.

The problem with attempting to simulate UG2 flotation circuits is that the assumption is made that the floatability rates are maintained throughout the flotation circuit. This is obviously not true if the circuit contains regrind stages and reagent addition points within the circuit that tends to change the floatability distribution of a stream. From the discussion above, most UG2 circuit designs contain regrind stages as well as the addition of depressants in the cleaning circuits, limiting the accuracy of simulations using methodologies that can't account for these phenomena. Runge et al (1997) assumed that a regrind mill or reagent addition points would not affect the floatability rates of the individual floatability classes but rather redistribute the mass fraction of each class. If this is a feasible assumption, then one way of simulating a complete UG2 flotation circuit would be to re-characterise the floatability distribution after a regrind stage or reagent addition point and treat it as a "transformation box" in the simulation algorithm. The assumption made by Runge et al (1997) has not been validated and this approximation is still very much open to question however.

2.3.4 Modelling and simulation of platinum circuits

De Beer (1996) simulated a two stage UG2 platinum flotation circuit by sampling the streams around the circuit and then performing batch flotation experiments on the samples. The batch flotation tests were modelled using a distributed floatability approach. The model yielded parameters that were associated with three floatability classes, viz. a fast class, medium floating and a slow floating class as well as a class for valuable minerals that did not float. The actual unit residence times were determined using residence time distribution studies. The rate constants associated with each floatability class and the unit residence times were used in association with the actual recoveries from the several units to simulate the performance of the circuit. Simulation was done by determining a scale-up factor for the rate constants for every unit in the circuit. De Beer assumed that the variables that affect flotation (gas dispersion characteristics, froth depth, aeration rate, etc.) were accounted for in the mixing parameters and the calculated scale-up factors. The effects of changes in unit residence times (changes in circuit layout) were accounted for by calculating new residence times with known volumetric feed flow rates and unit volume ratios.

Theron (1998) characterised and modelled the primary flotation circuit at Impala Platinum's UG2 concentrator. The circuit was characterised with respect to the platinum group elements, base metal sulphides and chromite content in the feed ore, as well as the water behaviour in the respective units in the circuit. A combination of the distributed floatability model and distributed size parameter model was used to describe the kinetic response to flotation of the platinum group elements (PGE) and base metal sulphide minerals by size. The kinetic response to flotation was modelled in terms of the contribution of the froth phase, the inherent ore floatability characteristics and the flotation machine to the recovery of the floatable minerals in the circuit. Mineral

floatability was assumed to be a function of mineral liberation, based on QEM*SEM analysis. The chromite was characterised by size to describe its response to entrainment with water. However, a portion of chromite was assumed to be recovered by true flotation. The water behaviour was modelled empirically as a function of unit residence time. The derived and measured characteristics of the feed ore, flotation units, and water behaviour were then used to simulate the performance of the circuit with respect to PGM and chromite recoveries under a variety of operating conditions. However, the water model that was used in this study limited the simulation exercise to recovery predictions only and the prediction of stream grades was not included. It was recommended that future circuit simulation research efforts should focus on the development of improved methods for modelling water behaviour in flotation circuits.

Sweet (1999) focussed on developing a procedure that could be used to evaluate the effect of changes in process chemistry on the performance of a flotation circuit treating a Merensky Platinum ore. A sixty litre continuous flotation cell was used to test the effect of changing the dosages of copper sulphate and dithiophosphate on the flotation response. The effect of the different reagent dosages was quantified using the pulp zone model of Gorain et al (1997) as well as the froth zone model of Harris (1998) described previously. It was found that it was not possible to apply these models under conditions of limiting froth conditions, ie. where a change in reagent dosage caused a complete breakdown of the froth phase of the pilot cell. It was concluded that the models used in this work were developed on operating plants at conditions that were reasonably conducive to stable operation, and it was thus not equipped to take bubble or froth overloading conditions into account. Also, it was concluded that the models used required considerable development if they are to be effective in the evaluation of changes in circuit chemistry.

2.4 Summary

This review has focussed particularly on the aspects of modelling and simulation pertinent to platinum flotation circuits, describing the various modelling approaches adopted by previous researchers. The double distributed parameter model (floatability by size class) seems to be the most useful and flexible method available to characterise the floatability distribution of particles in a flotation feed. The single rate kinetic model developed in the early stages of flotation research are extremely crude. The model first proposed by Zaidenberg (and subsequently investigated by King), in which the floatability of a mineral is distributed by liberation class by size class, is clearly of considerable theoretical interest. The main problem with this approach is the tremendous time (and cost) involved in generating the required information, which has greatly limited its application with respect to the analysis of industrial flotation processes. In addition, a direct, predictive relationship between floatability and liberation has still to be developed, so this method still requires correlation to actual operating plant (or pilot plant) information before it can be applied to flotation process design problems. Nonetheless, this method can be expected to become increasingly important in the future, as it should provide a basis for linking the processes of comminution and

flotation. This would provide the means to model, and optimise, both processes simultaneously, which could have substantial economic implications with respect to the design and operation of many industrial flotation plants.

The recent trend in the simulation of flotation circuits is to decouple the effects contributing to the overall flotation rate coefficient in terms of the factors affecting the cell hydrodynamics, mineral floatability as well as froth effects. This offers considerable advantages for use in simulation algorithms where cell operating characteristics need to be included in the kinetic model. The most promising advancement to describe the cell hydrodynamics as a separate entity from the mineral contribution on the overall flotation rate in the collection zone of a flotation cell is the work of Gorain et al (1997), using the bubble surface area flux to describe the cell hydrodynamics. In combination with work of Harris (1997) and Runge (1997) where nodal analysis is used to extract meaningful kinetic parameters from plant scale flotation circuits. The incomplete characterisation of the froth phase kinetics has been highlighted and is believed to be the major cause for the incomplete development of a comprehensive flotation simulation methodology. The influence of the froth phase needs to be explicitly included in the models describing the flotation kinetics. The most accessible means of accounting for the froth zone kinetics at present is through the use of empirically derived relationships obtained from pilot plant or industrial plant data as a function of the variables affecting the froth recovery of particles in the froth phase. The concentrate froth residence time based on the volumetric flow of concentrate has been highlighted as the most important and practical (measurable) variable to effectively model the froth phase performance on an empirical basis. Unfortunately this method is only suited to a fixed flotation cell in a fixed chemical environment in a flotation circuit and is still far removed from a comprehensive froth recovery model. As an alternative, the methods presented in the literature that enable direct measurements of the froth recovery (Savassi, 1997) need to be used to develop an understanding of the froth recovery mechanism from a more fundamental perspective. A reliable simulator for industrial flotation cannot be developed without including a model for entrainment. Models that do not include entrainment predict higher concentrate grades and lower concentrate recoveries compared to the true flotation process. The equation developed by Savassi (1998) which recognises that true flotation and entrainment of non-floating particles occurs simultaneously in a flotation cell is regarded as a significant step towards developing a comprehensive flotation model for predicting the overall recovery of minerals in a flotation cell. As a result of the dependence of the entrainment models on the water recovery models, as well as the dependence of the water recovery models on the gangue and minerals makes modelling of the water recovery from a flotation cell in a circuit extremely complicated. This interdependence means that a consistent methodology is required to overcome the need for a complicated water recovery prediction or otherwise a detailed mechanistic model is necessary. The lack of the existence of either of these two approaches is highlighted as the most important advancements required for the effective simulation of flotation circuits in the future.

There are still a number of inconsistencies in the literature regarding an effective procedure to simulate flotation circuits. The most recognised problem is that in some

systems the flotation rate coefficients seem to vary in different locations of a flotation circuit. A number of attempts have been made to try understand why this is the case. However, the methodology used by the UCT/JKMRC has shown that the floatability rates of components are maintained throughout a circuit of flotation cells in most instances reported in the literature, and a method of testing this assumption shows that this is more often than not an appropriate assumption, (Runge et al, 1997). It is believed that the observed variation in the rate constant in the previous modelling attempts are primarily as a result of the simplistic models used in these previous studies. Using models that incorporate the differences in the froth phase characteristics (particularly between the rougher and cleaners) as well as entrainment of minerals and gangue should resolve these issues.

2.5 Formulation of modelling approach

The primary objective of this thesis is to develop a simulation methodology to simulate the flotation kinetics of the Mintek pilot plant facilities treating a UG2 platinum ore. Based on the literature presented in this chapter, the following approach to model the Mintek pilot plant facilities has been selected to be the most appropriate for this study. The flotation kinetic modelling approach as well as the method selected for extracting the required parameters is presented. The selection of a suitable froth kinetic model as well as an entrainment model is presented. Finally the approach to modelling the water kinetics will be described.

2.5.1 Choice of characterising the UG2 ore sample

The mineralogy of the UG2 ore sample was briefly discussed in section 2.3.2. The PGM particles occur in a number of composite associations with other minerals in the ore. These valuable composite minerals need to be modelled independently from the particles that do not contain PGM's (gangue). The approach selected to model the PGM particles is to assume that all composite particles containing PGM's can be lumped into a single component for modelling purposes. The influence of all possible variables affecting the flotation rate of the PGM particles can then be accounted for collectively in the form of a floatability distribution. Due to the high percentage of chromite in UG2 ores (being the dominant gangue mineral) as well as the need to minimise the chromite content in the final concentrate of UG2 circuits, chromite can be modelled as a separate component. It is necessary to model the gangue kinetics in order to obtain PGM grades from the simulations. The remainder of the UG2 ore can be considered gangue, which can be calculated as the total dry mass less all the PGM and chromite particles. This can then be treated as an individual component in the simulation algorithm. Therefore, as a simplification to modelling the UG2 ore, the feed has been chosen to consist of three broad components namely; PGM, chromite and gangue.

2.5.2 Choice of modelling and simulation methodology

Flotation kinetic model selection

The distributed first order floatability model of Imaizumi and Inoue (1965) has been found to be appropriate for describing complex ores with extreme variations in mineralogy and surface properties. The mineralogy of the PGM's in UG2 ore is typical of such a complex ore and therefore this approach has been selected as the most suitable method of modelling the floatable kinetics. It has been shown in the literature review (section 2.2.4.1) that the recovery time profile from a batch test can be represented by a plug flow model. Assuming no interaction of the floatability components, the total recovery from a particular cell in a steady state circuit would be the summation of the various floatability component recoveries within that cell and can be represented by equation 2.70.

$$R_{j,s} = \sum_{i=1}^n m_{i,j,s} [1 - \exp(-k_{i,j}^{batch} t)] \dots\dots\dots 2.70$$

Assuming a perfectly mixed pulp for a continuous flotation cell on the pilot plant, the recovery of a component has been shown to be represented by equation 2.71.

$$R_{j,k} = \sum_{i=1}^n m_{i,j,k} \left(1 - \frac{1}{1 + k_{i,j,k} \tau_k}\right) \dots\dots\dots 2.71$$

where

- $R_{j,s}$ is the recovery at time (t or τ_k) of mineral j from stream s or cell k
- n is the number of floatability classes
- $m_{i,j,s}$ is the mass fraction of mineral j in floatability class i in stream s
- $k_{i,j}^{batch}$ is the rate constant associated with floatability class i of mineral j in unit or cell k
- t batch cell cumulative residence time
- τ_k continuous flotation cell residence time in unit k in the circuit

A decoupled rate constant was selected, according to the approach of Gorain et al (1995). Decoupling the rate constant in this manner enables the separate effects of the flotation cell hydrodynamics (S_b), the inherent mineral floatability (P_i) as well as the froth phase kinetics (R_f) to be explicitly included into the simulation algorithm so that their individual contributions to the overall flotation rate can be assessed. S_b can be measured directly in each cell by measuring the sauter mean bubble size in each flotation cell as a function of the air flow rate into the cell and the chemical and pulp physical conditions, using the UCT bubble size analyser. The cell residence times can be approximated as the cell volume divided by the volumetric tails flow rates in each cell measured directly from a circuit mass balance. To determine the froth recovery in a

flotation cell, the indirect empirical method developed by Gorain et al (1998) and extended by Harris (1998) has been selected as the most appropriate method. The reason for this is that this method is relatively simple and is based on easily measurable parameters that can be obtained directly from the operating pilot plant without needing complicated measurements of physical variables. In addition, this method has been demonstrated to be successful in application to a wide range of data obtained from different operating plants under different conditions. The froth phase can then be characterised by regressing the parameters described in equation 2.72 to experimental data by varying the froth height and air flow rate.

$$R_{fk} = (1 - \alpha) \exp(-\beta * \tau_{fc,k}) + \alpha \quad \dots\dots\dots 2.72$$

where

- R_{fk} froth recovery in unit k
- β derived circuit (or bank) froth stability parameter
- α derived froth parameter related to the amount of non-draining mineral
- $\tau_{fc,k}$ froth retention time in unit k based on concentrate flow rate (volume of the froth divided by the concentrate volumetric flow rate)

Characterising the froth phase as a function of froth depth and air flow rate restricts the use of this model to a fixed flotation cell in a fixed chemical and pulp environment in a circuit of flotation cells. It would be expected that the exponential decrease in froth recovery with froth residence time would change for different flotation cells in a circuit and therefore each flotation cell, or as an approximation, each flotation bank in the circuit needs to be characterised individually. An assumption that the froth recovery is not floatability specific and that all floatability classes will be subjected to the same froth recovery kinetics governed by equation 2.72 will be adopted. This is in order to limit the number of parameters in the simulation algorithm. It is acknowledged that this assumption is not entirely true as it was clearly stated in section 2.2.3.2 that froth recovery is dependent on particle size and mineral specific gravity. However, due to the very fine mineral pulps produced in UG2 flotation circuits, it is proposed that the differences in froth recovery due to these variables is minimal and can adequately be assumed equal for all floatability classes.

Most simulation models assume, of necessity, that under conditions of steady state with no regrind stages or reagent addition points in the flotation circuit, that the floatability of the ore should not change with time. The extent to which the floatability of the PGM's in the UG2 ore remains constant can be evaluated by considering the results of batch floatability tests conducted around nodes of the Mintek pilot plant operating at steady state, as discussed in the literature review (section 2.2.5.2.4), using the approach of Runge et al, (1997). This graphical technique of determining the conservation of floatability components around a stream mixing node can then be extended to analyse the conservation of floatability components around all nodes of the pilot plant. Once it is

established that the floatability of the PGM's in the circuit does not change with time, the flotation response of all the streams can be combined to determine the required modelling floatability parameters for the PGM's in the UG2 feed sample to the pilot plant using the approach developed by Harris (1998). This can be done by fitting the stream batch floatability data to equation 2.70, using an appropriate non-linear optimisation procedure, based on sum-of-error-squared minimisation. This will be subjected to the constraint that the PGM floatability component flow rates have to comply with the mass balance constraints within every cell of the flotation circuit. In addition, the fractions of PGM floatability components must add up to one within all streams.

Entrainment model selection

In order to model the entrainment kinetics for chromite and gangue particles it is assumed that particle size is the dominant variable affecting the entrainment of these particles with water. The variation of the degree of entrainment was found to resemble a classification function whose form depended on the operating conditions in a cell (Johnson, 1972). The empirical partition curve developed by Savassi (1998), which is an extension of Johnson's approach has been selected as the most suitable method to describe the variation of the degree of entrainment with particle size. Equation 2.73 and 2.74 shows the entrainment model selected from the literature.

$$R_{ent,j,k}^o = \sum_{i=1}^d E_{j,d} * R_{w,k} \quad \dots\dots\dots 2.73$$

where

$R_{ent,j,k}^o$ is the total recovery by entrainment in all the size fractions of mineral j in unit k
 $E_{j,d}$ is the degree of entrainment of mineral j for a particle in size class d
 $R_{w,k}$ is the measured water recovery from flotation cell k

$$E_{j,d} = \frac{2}{\exp(2.292 * (\frac{s_{j,d}}{\xi_j})^{ADJ}) + \exp(-2.292 * (\frac{s_{j,d}}{\xi_j})^{ADJ})} \quad \dots\dots\dots 2.74$$

$$ADJ = 1 - \frac{\ln(\frac{1}{\delta_j})}{\exp(\frac{s_{j,d}}{\xi_j})}$$

where

-
- $s_{j,d}$ is the average particle size in the i^{th} size class d (in microns) of mineral j
- ξ_j is the entrainment parameter for mineral j (given by the particle size for which the degree of entrainment is 20 percent)
- δ_j is the drainage parameter of mineral j (related to the preferential drainage of coarse particles from the froth)

This model is selected as it has been found to accurately model the entrainment for a wide range of minerals from various sources in the literature, (Savassi, 1998). The model is also relatively simple requiring only two parameters to describe the degree of entrainment as a function of particle size class for each entrained mineral in the circuit (chromite and gangue). All that is required to fully characterise the entrainment of chromite and gangue is a measure of the water recovery which can be obtained directly from experimental measurement as well as the particle size distribution of each component feeding the flotation circuit. Each particle size fraction can be assessed individually throughout a circuit assuming that they do not interfere with one another in the circuit. In addition, the water recovery measurement indirectly incorporates the influence of the entrained froth recovery and the parameters in equation 2.74 were found to vary only slightly with a change in froth height and air flow rate, (Savassi, 1998).

Water model analysis

The review of the literature concerning water recovery models has highlighted that the simpler approaches of relating the water recovery to either the mineral flow or solids flow are extremely crude while the mechanistic models are not practical and are difficult to adapt to processes other than in well controlled laboratory conditions. As a result of the absence of a practical and sufficiently meaningful water modelling approach in the literature, this thesis will analyse a number of the simpler approaches presented in the literature review and assess how well these various approaches are capable of being incorporated into a simulation algorithm. If none of the approaches prove to be useful, an alternative methodology that can describe the water recovery kinetics as a function of cell operating parameters will need to be developed.

Chapter 3

PRELIMINARY MODEL DEVELOPMENT

This chapter presents the initial work which was performed using a mini-pilot plant facility (Mintek mini-plant). The work highlights the progression of the development of the final model which is presented in chapter four. In particular, the shortcomings of the system used to perform the first phase of the work are identified, highlighting the need to upgrade the pilot plant facilities at Mintek in order to obtain accurate measurements of the variables in the model as well as to obtain more reliable mass balance data.

The chapter starts by briefly presenting the mini plant facility at Mintek, describing the equipment used as well as the experimental work performed. A detailed experimental procedure is presented. The next section focuses in detail on the PGM flotation kinetics in this system. A simple three step algorithm is developed as a method of obtaining a reliable parameter set describing the floatability distribution of the PGM minerals in the feed. The fit of the model to the experimental data is presented. The significance of the regressed PGM floatability distribution is tested using Monte Carlo simulation techniques and this is used to provide insight into the sensitivity of the regressed parameters in the model. The chapter concludes by discussing the limitations in the preliminary development of the models leading through to the next chapter which presents the main phase of the model development work where these limitations are addressed.

3.1 Description of the Mintek mini plant facility

A diagram of the Mintek mini plant used for the preliminary model development test work is shown in figure 3.1. The circuit was selected to be simple enough to facilitate steady state operation of the mini plant, but have sufficient units to provide the interaction needed to adequately assess the applicability of the proposed models. The flotation cells used for the testwork are commonly known as Sala flotation cells designed by Svedala (PTY) LTD. The rougher consisted of four thirty litre flotation cells joined together to form a bank. The scavenger consisted of two thirty litre flotation cells and the cleaners consisted of four fifteen litre flotation cells joined together to form a bank. All the thirty litre cells were square and manufactured from stainless steel. A common motor was used for two cells on a bank. An Agitar agitation mechanism was used with a 0.8 KW motor for all the flotation cells. The power drawn by the motors was not adjustable, and the agitation speed of the impellers was fixed. No froth scrapers were used, the froth was allowed to flow naturally over the cell lip into the launder. The froth produced flowed over only one side of each cell, but the launders were designed so that individual concentrate flows from each cell could be sampled during the tests. Table 1 gives the measured dimensions of the individual flotation cells, as well as the agitation speeds used in each individual cell.

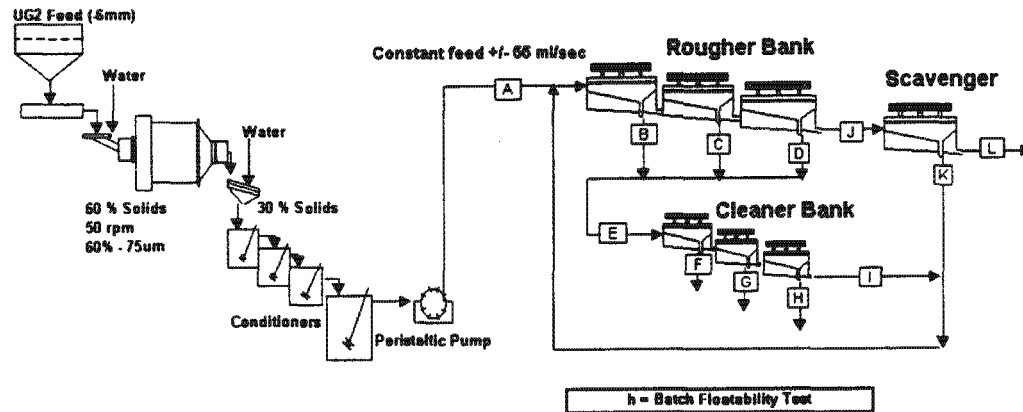


Figure 3.1: Mini plant circuit configuration depicting the feed setup as well as sample points for batch floatability tests

Table 3.1: Dimensions and agitation speeds of all the flotation cells used on the mini plant

CELL NAME	LENGTH (CM)	HEIGHT (CM)	DEPTH (CM)	R.P.M.
Rougher A	29.6	32.0	30.0	1095
Rougher B	29.6	32.1	29.9	1092
Rougher C	29.8	32.2	29.9	1095
Scavenger A	29.6	32.0	29.6	1071
Cleaner A	21.5	37	**	1262
Cleaner B	21.8	37	**	1266
Cleaner C	21.3	37	**	1268

** variable with cell height

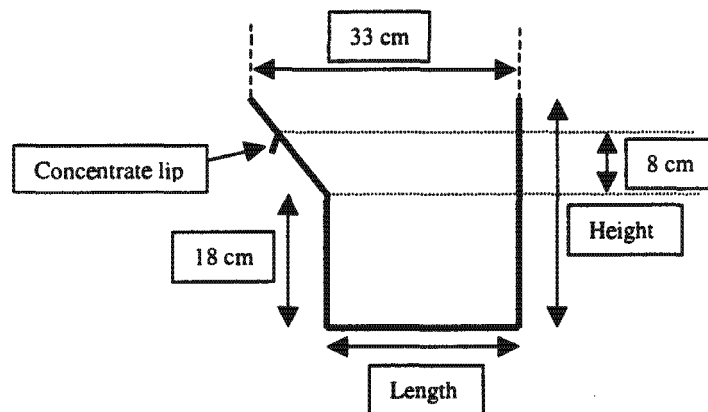


Figure 3.2: Denver Sala Flotation System – Design of the cleaner cells

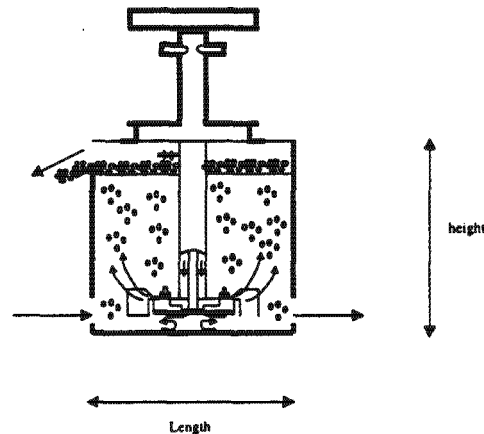


Figure 3.3: Denver Sala Flotation System AS – Design of the rougher and scavenger flotation cells

The cleaner bank had a variable cross sectional area as depicted in figure 3.2. This variation was taken into account when calculating the volume occupied by the froth and pulp phases. Figure 3.3 is a schematic representation of the SALA flotation cells of the rougher and scavenger banks. The froth height could not be varied independently of the pulp height, and therefore changing the froth height changed the pulp level. The volume occupied by the impeller/stator was also taken into account during the analysis of the mini plant data.

3.2 Experimental test work on the mini plant facilities

A platinum ore obtained from the Upper Group 2 chromitite seam (UG2 LONMIN, Eastern Platinum Mine) of the Bushveld Igneous Complex, in the North West Province of South Africa was used for the test work. This ore was considered an average one in terms of a small extent of oxidation and not too many occlusions or intergrowths of the PGM minerals into the gangue.

3.2.1 Circuit feed characterisation test

The main objective of this test was to characterise the feed to the mini plant. The outcome of this test was also used to assess the steady state operation of the mini plant. The methodology employed for this test was as follows:

- Characterise all the stream flows within the circuit with respect to volume, mass, mineral and gangue content.

- Characterise all the flotation cells in the circuit with respect to the gas dispersion in the cells as a function of the operating conditions employed (chemistry, gas rate, impeller speed, froth height).
- Characterise the floatability distribution of the ore used, using recovery-time profiles obtained from batch flotation experiments conducted on material from all streams in the circuit.

With the proposed circuit at a feed flow rate of 80 kg/h and 60% solids, there was approximately 40 minutes residence time in the rougher and scavenger banks combined. Air flow rates and froth heights for each cell were set at an average value to maintain a good froth flow rate from every cell. These settings were kept constant over the duration of the mass balance, and were measured before each mass balance was conducted. Three hours were allowed for the circuit to reach steady state. Samples of all the streams were taken for flow rate measurements. This was repeated four times with each balance one hour apart to assess the steady state of the circuit as well as to obtain standard deviations on the stream flows. The sample bags were kept for filtering and sample preparation after the mass balancing campaign. Each sample was tested for PGM and Cr_2O_3 content so that standard deviations on the PGM and Chromite results could also be obtained.

It was necessary to conduct reproducibility tests on the batch floats to confirm the experimental accuracy of the method employed. Reproducibility tests on the feed stream were used to determine the variation of the feed floatability and to test the consistency of the batch flotation technique. In addition, two reproducibility batch rate floats were done on the final tailings stream (stream L). The results of these tests are presented in figure 3.4. It can be seen that very good reproducibility was obtained.

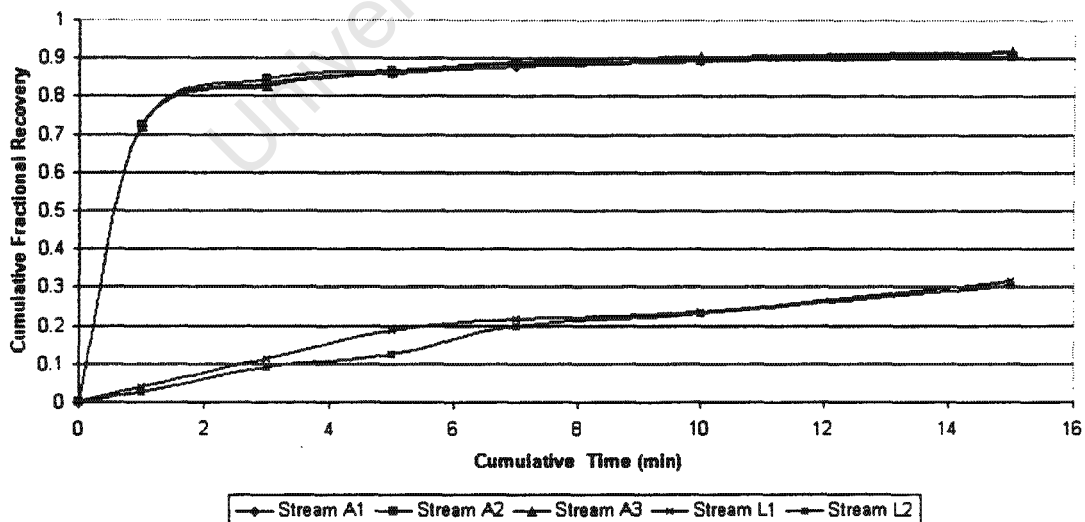


Figure 3.4 Batch reproducibility tests on the feed and tailings streams of the mini plant

3.2.2 Froth height and air flow rate characterisation tests

The aim of this test was to assess the various boundary conditions with which each bank could operate. The aim was to determine the overall mass flow rate relationships as a function of the froth phase and the pulp phase operating conditions of the cell. These tests provided an indication of the extent these variables could be changed independently from one another, and the impact of these variations on the circuit balance. The variables that were manipulated to vary the performance of the bank was the individual volumetric air flow rates into each cell as well as the pulp level for the entire bank. To characterize the rougher cell with respect to froth height and air flow rates the bank was operated on its own without the other cells in the circuit. This was to eliminate the effect of disturbances from the recycling streams. It was also necessary to characterize the cleaner cell with respect to these variables as the pulp density and floatability characteristics were substantially different from those in the rougher bank. The froth heights chosen at each air flow rate for the rougher tests represented those giving rise to a low, intermediate and high concentrate flow rate, in effect exploring the boundary conditions at which the bank could operate. The same approach was used for the cleaner test, except various air flow rates were chosen at constant froth heights as can be seen in Tables 3.2 and 3.3 below.

Table 3.2: Various combinations of volumetric air flow rates and froth heights for the rougher A test

VOLUMETRIC AIR FLOW RATE IN EACH CELL (L/MIN)	FROTH HEIGHT (CM)
10	(3), (3.5)
15	(3), (4.5), (5)
20	(6), (6.5), (7), (8)
25	(8), (8.5), (9), (10)
30	(11), (11.5), (12)

Table 3.3: Various combinations of volumetric air flow rates and froth heights for the cleaner A test

VOLUMETRIC AIR FLOW RATE IN EACH CELL (L/MIN)	FROTH HEIGHT (CM)
(12.5), (15), (17.5), (20)	5
(10), (12.5), (15)	3.5
(7.5), (10), (12.5)	2

The samples collected were all analysed for PGM, chromite, water and gangue content.

3.2.3 Circuit configuration changes

Two alternative circuit configurations were tested in order to assess the proposed models capacity to predict changes in the circuit structure. The configurations were chosen so that a different separation would be obtained from each circuit.

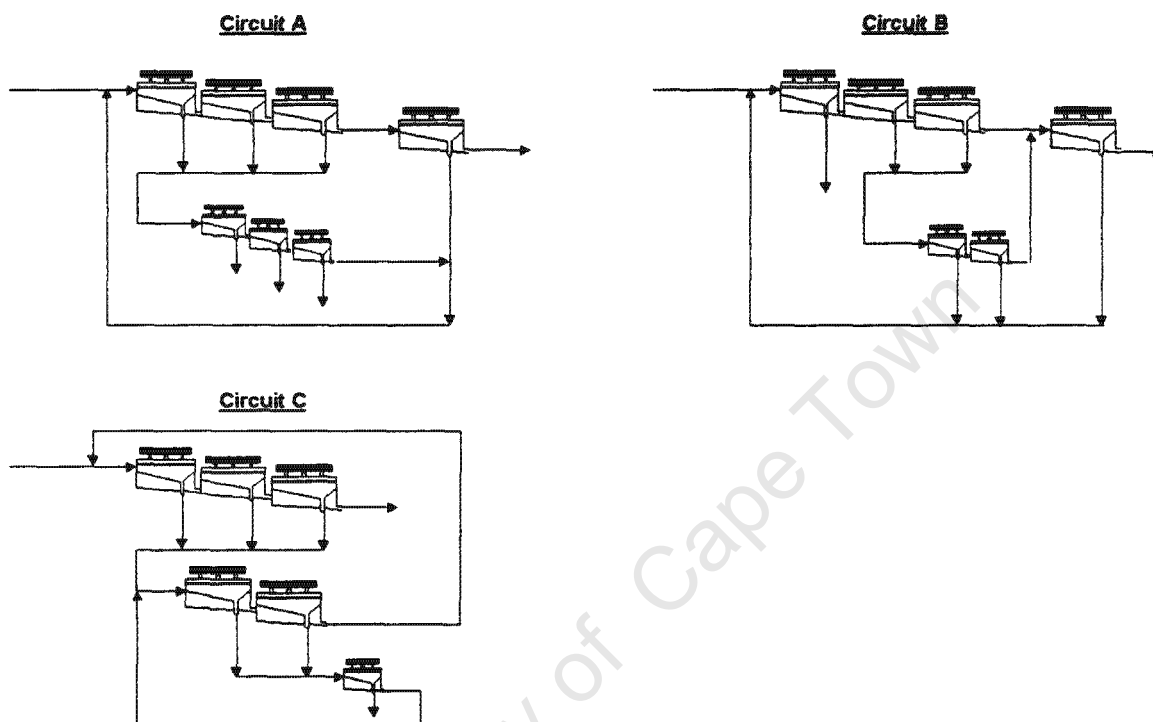


Figure 3.5 Alternative circuit configurations of the mini plant

Circuit A is the standard circuit that was used for the development of the preliminary model. The other circuit configurations were assessed with reference to this standard configuration and used to determine whether simulations using the model could accurately predict a change in circuit configuration. It was expected that circuit B should have a final concentrate with a lower grade than circuit A due to a higher entrainment probability of gangue in the rougher cell. Circuit C was expected to have the highest final concentrate grade due to the recleaning of the cleaner concentrate, but also to have the lowest PGM recovery due to a smaller residence time in the rougher flotation bank. The aim of this qualitative reasoning was to assess whether the circuits would respond in the expected manner and to determine whether the circuit simulations using the proposed model would be able to quantitatively predict these differences in the different configurations. Identical froth heights and air flow rates were used for the rougher cells in each circuit. This was to ensure that no bias in recovery or grade were due to changes in these variables. A single mass balance was conducted for both of the alternative circuit arrangements.

3.3 Experimental techniques used in obtaining experimental data

3.3.1 Run of mine sample preparation

A 20 ton sample of the UG2 ore was delivered as run-of-mine ore and dried on a drying pad. The ore was then crushed to – 6 mm in diameter. After the ore was crushed it was blended three times to ensure the sample was well mixed and completely homogeneous in composition. The blending technique consisted of layering the ore over an area of 10 x 2 metres in even strips (30 cm wide) along the 10 metre length using a blending table. The ore was then shoveled at ninety degrees to the 10 metre length of the sample, back into sample bags. This process ensured effective mixing of the sample. The blended sample was placed in storage bags that could be hoisted on top of the pilot plant mill feed hopper.

3.3.2 Preparation of feed to the mini plant

Exactly the same procedure of producing feed pulp to the mini plant was followed for every test described in section 3.2. This was so that a direct comparison between the alternative circuit configurations could be made, as well as to determine if the regressed floatability distribution of the model could accurately predict the kinetics of the floatable particles under the different conditions tested.

3.3.2.1 Milling conditions

A grind of 60% -75 μm was chosen as this was considered an appropriate grind achievable by the mill and is also easily amenable to flotation, liberating most of the base metal sulphides in the sample. It was decided to use an open circuit mill which was easier to operate and control compared to a closed mill/cyclone network. The advantage of an open circuit mill is that it reaches steady state a lot quicker and eliminates the complications arising from recycling streams in the mill circuit. It must be stressed that the intention of the test work was not to produce optimum feed conditions for flotation, but rather a constant feed composition that was easy to maintain.

Individual storage bags filled with the crushed and blended UG2 ore were hoisted on top of the feed hopper by means of a crane as feed to the hopper. Ore was introduced from the feed hopper onto a variable speed conveyor belt that fed directly into the mill feed port mill at the set feed rate of 80 kg/h. The milling medium consisted of steel rods. A variable speed dial together with an electronic mill speed indicator was used in order to set and monitor the rotation of the mill. Water was added into the head of the mill and controlled on a rotameter (which was recorded) to produce a mill product of sixty percent solids. The percentage solids was measured and maintained using a marcy scale, calibrated to a feed with a specific gravity of 3.7 corresponding to the average density of the ore. The marcy scale was calibrated before the start of every mini plant

run. The flow rate of dry ore from the conveyor into the mill was measured by cutting the entire stream for a duration of fifteen seconds with a sampling bucket. The mass collected was measured and converted to a flow rate. The collected sample was returned to the mill after the measurement.

A milling curve test was performed in order to obtain the desired grind of 60% -75 μ m. This was done by feeding the mill at a constant feed flow rate of 80 kg/h while maintaining sixty percent solids in the mill exit stream. The mill was initially set at the lowest feasible rotation speed of 15 rpm. After one hour the mill was considered to be operating at steady state and a fifteen second sample of the discharge stream was taken. After this the mill speed was increased in incremental steps of 5 rpm to 55 rpm while still maintaining the feed flow rate and percentage solids and waiting an hour before each sample was taken. The samples obtained from the milling test were wet screened over a vibrating 75 μ m sieve for ten minutes. The split samples were filtered and dried and the percentage smaller than 75 μ m was calculated. From this it was calculated that a mill speed of 50 rpm would give the required grind. The discharge from the mill flowed onto a vibrating 1mm screen to remove unwanted wood chips as well as to remove the occasional large particle to prevent pipe and pump blockages. Wash water was sprayed over the screen decreasing the percent solids to thirty percent which represented the desired flotation density. The percent solids of the screen undersize stream was also measured every 30 minutes during all the tests. The feed flow rate, reagent flow rates as well as the percentage solids of the mill exit stream and plant feed stream were all measured and recorded every 15 minutes for each sampling campaign. These were all adjusted if they deviated from their desired set points. A sample of the mill exit stream was also taken every hour to determine the grind and to monitor the performance of the mill. This stream was then pumped directly to the head of the reagent addition section of the plant. In order to approximate steady state conditions within the flotation circuit it was imperative that the feed stream remained as constant as possible. With a varying feed flow rate, percentage solids or feed grade, it is impossible to run the plant at steady state. It is for this reason that great care was taken in order to ensure that all the variables described above affecting the characteristics of the feed to the flotation circuit were maintained as constant as possible.

3.3.2.2 Reagent preparation and addition rates

The reagents used for the tests were those suggested by Mintek and were based on the standard reagent suite for UG2 ores of this type. All the reagents were pumped via peristaltic pumps from pre-prepared samples stored away from the plant to eliminate contamination of the reagent by pulp. The reagents were all freshly prepared at the start of a campaign to ensure no degradation with time. The reagents were mixed to a concentration of one percent by mass with water.

The reagent addition points consisted of three one litre conditioning tanks which were each mixed by a rushton impeller. The pulp overflowed each tank and cascaded into the next. Complete mixing was achieved in all three conditioners, without any settling of

the ore particles. A conditioning residence time of approximately 20 seconds was maintained in each mixer. The pulp from the last conditioner overflowed into a buffer tank. This consisted of an agitated tank of 75 litres with an exit port at the bottom connected to a variable speed constant head peristaltic Bredell pump. This unit was used to buffer out any flow rate disturbances from the milling and conditioning sections of the plant. This proved to be very effective allowing a constant feed flow rate to be maintained for the flotation circuit. The buffer tank also increased the conditioning time of the pulp before being subjected to flotation (additional 22.7 min).

Copper sulphate was added to the first conditioning tank at a flow rate of 2 ml/min (15 g/t). This was used as an activator to enhance the attachment of the collector onto the surface of the mineral particles. A xanthate collector, SIBX, was pumped into the second conditioning tank at 20 ml/min (150g/t). The third conditioning tank had both the frother (Dowfroth200) and depressant (KU-5, a carboxymethyl cellulose depressant) added to it at flow rates of 10 ml/min (75g/t) and 14 ml/min (105g/t) respectively. It must be noted that the KU-5 depressant was not very soluble in water and was allowed one hour of mixing with the water to ensure the solution was well mixed and free of agglomerates. The selected reagent suite and their respective flow rates were those recommended by Mintek for this type of ore. Evaluation of the reagent suit was not within the scope of this study.

3.3.3 Bubble size measurements

The UCT bubble size analyser was used as a direct method of obtaining the sauter mean bubble diameter within each cell of the mini plant, including the batch flotation cell used for the floatability tests. The mean bubble size was used in the model to calculate the bubble surface area flux in the flotation cell, which was a parameter directly influencing the flotation kinetics within the plant. The bubble size measurements were done with the plant operating at steady state in a closed loop, re-circulating the final concentrate and tailings streams to the feed of the pilot plant. This allowed all the measurements to be made without having to continually provide fresh feed to the plant. A detailed description of the UCT bubble size analyser can be found in Tucker et al (1994).

The air flow rates into each cell were set to a constant value during the bubble size measurements. Four bubble size tests were measured in each cell of the mini plant as well as in the batch cell in order to obtain a standard deviation of the measurements. The results obtained from these tests can be seen in figure 3.6 which gives the bubble sizes and standard deviations for each cell in the circuit as well as for the batch cell. It was necessary to determine to what extent the bubble size varied with air input to the cell. Bubble size measurements were made in each cell of the circuit at different air flow rates, over the range employed for each cell in the rougher and cleaner tests. Figure 3.7 shows the relationship between bubble surface area flux and volumetric air flow rate in the rougher and cleaner cells.

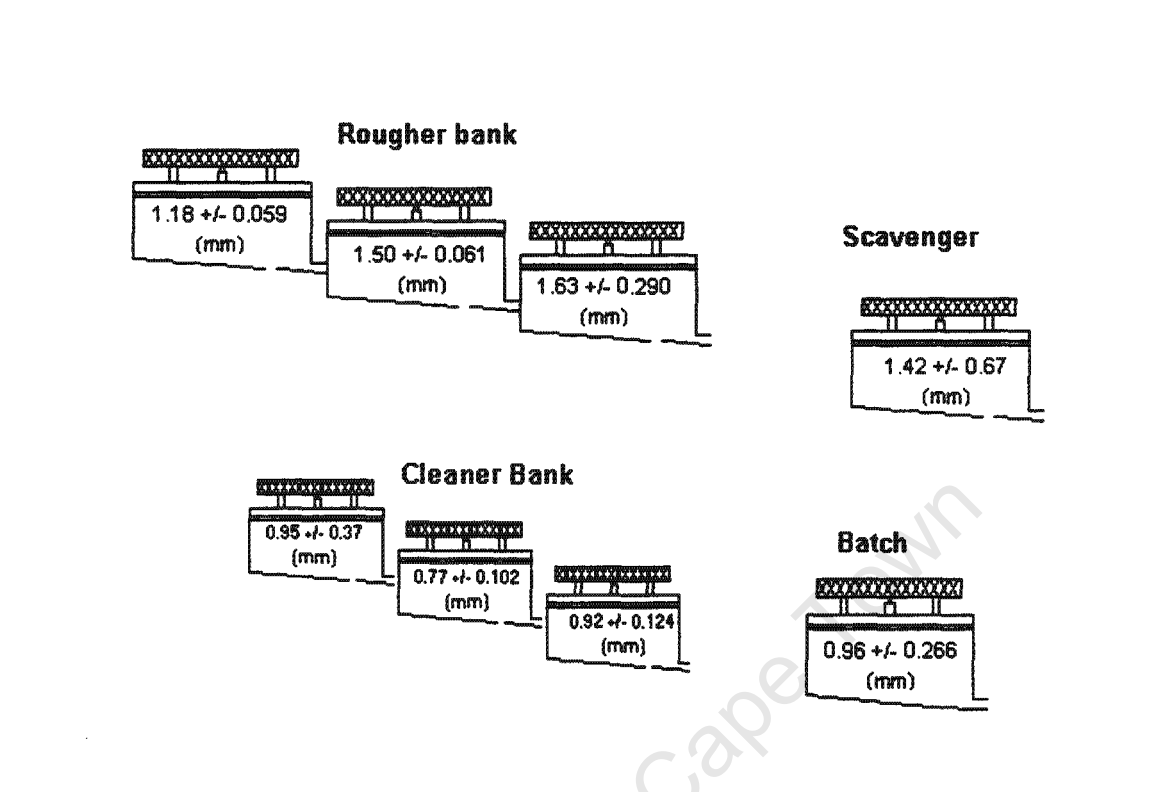


Figure 3.6: Bubble size measurements and standard deviations in each cell of the mini plant and the batch floatability cell

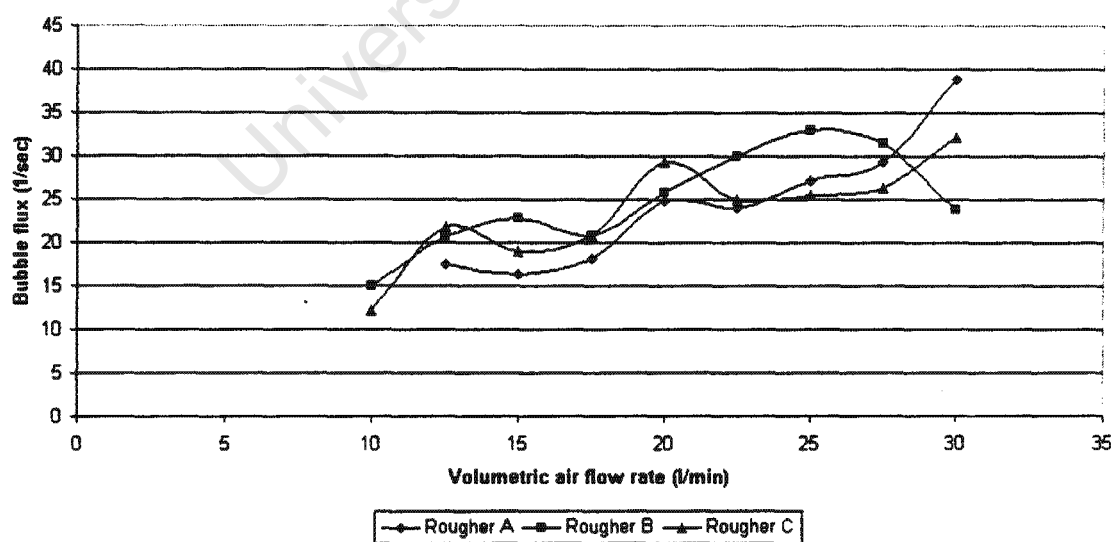


Figure 3.7: The variation of bubble surface area flux with volumetric air flow rate for the rougher cells on the mini plant

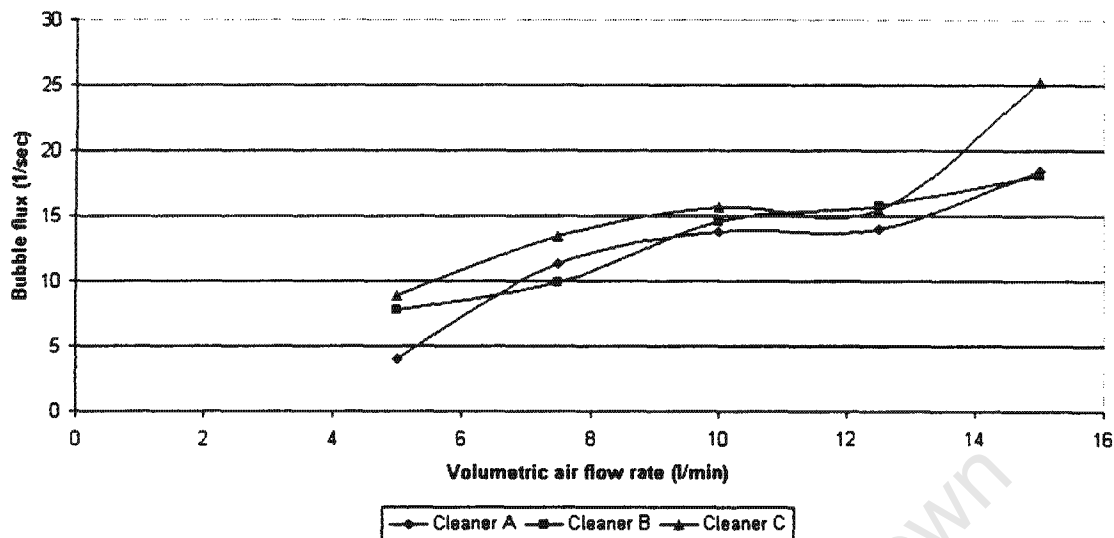


Figure 3.8: The variation of bubble surface area flux with volumetric air flow rate for the cleaner cells on the mini plant

3.3.4 Stream sampling methodology

A systematic procedure was followed in taking samples from the mini plant. This procedure minimized the disturbances created on other streams while sampling a particular stream in the circuit. Final concentrates and final tailings were sampled first as these streams had no direct effect on other stream flow rates in the plant. After these samples were taken, usually cleaner tailing streams were sampled next as this had a smaller flow rate than the rougher concentrates feeding the cleaner, and therefore considered to have less effect on disturbing the system. A few minutes was allowed before samples were taken from the rougher bank in order for the recycled cleaner tailings stream to settle down again. The scavenger concentrate streams were sampled next followed by the rougher tailings stream. Only after all other streams had been sampled was the feed stream sample taken as this stream affected the performance of the entire circuit. A similar method of sampling the streams was employed for all the circuit structures tested in order to reduce the disturbances made during sampling so that accurate and representative samples could be obtained.

Timed samples were collected into pre-weighed industrial A3 sized sampling bags. The typical duration of a concentrate sample was one to two minutes; tailings samples were typically collected for about fifteen seconds. Once the sample had been taken, it was sealed with a cable tie. After all the streams were sampled, the wet mass (including the bag and cable tie) were measured immediately to ensure no water losses occurred due to leakages or evaporation. After the campaign was completed, the samples were vacuum filtered and then dried in an oven.

3.3.5 Batch floatability tests

As discussed in section 2.5, the approach proposed by Runge et al (1997) and Harris (1997) was identified as the best method for the determination of kinetic parameters in circuits. Their approach requires that streams in the circuit be subjected to rate tests in a batch cell.

A standard Denver batch flotation machine was used with a flotation cell of three litres in volume. All the rate tests were conducted under identical conditions. Samples of the streams were obtained by cutting the stream with the flotation cell and allowing it to fill up to a marker on the inside of the cell wall. This was two centimeters from the cell lip. It was necessary to ensure that the effects of the froth zone during the kinetic rate floats were minimal without allowing the pulp to overflow the cell lip. The sample taken from the plant was floated immediately in order to obtain the same floatability response of the material in the batch cell relative to the plant, minimising the possibility of sample degradation and oxidation with time. No chemicals were added to these tests. The collected sample was floated using a standard Denver batch flotation rig, mixed at an agitation speed of 1000 rpm. This agitation speed provided enough energy to suspend all the solids in the flotation cell while providing a stable slurry/froth interface. To initiate a test, the air flow rate was turned on, at 10 l/min, and the test was then timed on a stopwatch. Froth scrapers were used to scrape the froth every ten seconds. Water was added from pre-weighed wash bottles to the sides of the cell walls as well as to wash the froth scrapers, using enough water to maintain the pulp volume inside the cell. The froth was collected into separate trays at 1, 3, 5, 7, 10, and 15 minutes respectively while still maintaining a constant scraping interval, froth height, air flow rate and agitation speed. After collecting the last sample, the mass of each tray filled with collected concentrate was measured. The mass of the batch cell with the remaining tailings as well as the wash bottles with remaining water were measured. The concentrates and tailing were filtered and dried. The masses of the dried filter cakes were measured and prepared for PGM analysis.

3.3.6 Sample preparation and analytical measurement techniques

Every filtered and dried sample collected during the pilot plant runs were subjected to a standard sample preparation procedure. Initially, all the high grade PGM concentrate streams were separated from the low grade tailings streams in order to avoid contamination during sample preparation and were prepared separately from one another. Each sample was then crushed through a 1 mm screen to break down any agglomerated particles from the vacuum filtration and drying. A representative sample was then taken by spreading the sample evenly onto a sample table and collecting small amounts over the entire area of the sample with a spatula. A 100 gram representative sample was then transferred into a Ziebtechnik pulveriser. All the samples were pulverised for one minute. Pulverising the sample ensured complete liberation of the mineral species so that accurate chemical assays could be obtained. The pulverising bowl was cleaned after every sample using silica sand and blow air.

The pulverised samples were then transferred into sample bags and sent to the analytical department at Mintek. The samples were subjected to a chemical analysis for PGM and gold content, and XRF analysis for the determination of chromite content. Details of the analytical procedures used are presented in Appendix A. The maximum variation between duplicate samples did not exceed 5% for either method of analysis.

3.3.7 Measurement of froth heights and air flow rates during the mini plant test work

The cells on the mini plant were not equipped with any means to measure air flow rate, so in turn, a portable rotameter was used to measure the air flow rates in each flotation cell. The air feed pipe feeding a flotation cell was disconnected and reconnected onto the rotameter for each measurement. The air flow rates into each flotation cell was measured after each mass balance was taken and therefore the interrupted air supply to the flotation cells had no effect on the samples taken during the mini plant tests.

The froth heights were measured using an inverted, plastic 500ml measuring cylinder. A zero level mark was made on the measuring cylinder at the point of the pulp froth interface level when no air was feeding the flotation cell. With the air flow rate on, the froth height could be measured from the difference of the measured zero level point and the top of the froth/air interface. The measurement of the froth depth was taken at the centre of the flotation cell between the impeller/stator stand pipe and the concentrate lip of the flotation cell for all measurements. This was to prevent a bias in the measurement if the measuring cylinder was placed elsewhere in the flotation cell as a result of variable mixing intensities over the cross sectional area of the flotation cell. The method of obtaining the froth heights was considered relatively crude but no other means was available at the time to obtain an accurate froth depth measurement. The standard deviation of the measured froth depth was $\pm 0.5\text{cm}$ of the measured froth height.

3.4 PGM modelling methodology

This section will describe the methodology used to model the PGM data obtained from the mini plant experiments. The basis of the modelling approach assumes that if one knows the characteristics of the feed ore coming into a circuit (floatabilities, P_i and fraction of each, m_i) together with a model that will predict the split across a separation stage (enhancement factor) and knowing the operational characteristics of each cell such as its froth height and bubble surface area flux, then it should be possible to predict the PGM flow rates in every stream that makes up the circuit.

3.4.1 Conservation of floatability verification

The first assumption in the modelling methodology that was selected in this study is that a particle maintains its inherent floatability independent of its location in a circuit. This

can be determined by performing batch rate tests on the streams in the circuit. The method used to conduct these tests is presented in section 3.3.5.

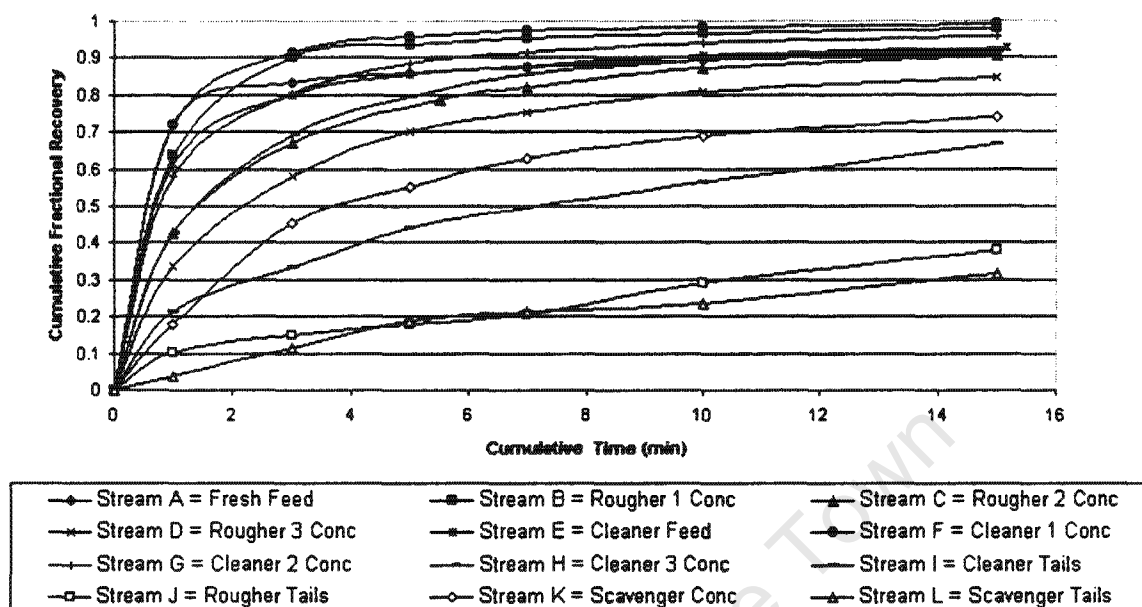


Figure 3.9 : The batch cumulative recovery time profiles of the streams in the mini plant circuit characterisation test of figure 3.1

The mini plant configuration used for this evaluation was as presented in figure 3.1. Figure 3.9 shows the experimental batch recovery time profiles that were obtained. It can be seen that the recovery time profiles obtained from the concentrate streams indicate much faster kinetics than the tails stream recovery time profiles, as expected. The data presented in figure 3.9 can be subjected to the nodal analysis procedure described in section 2.2.5.2.4. These results are presented in figures 3.10 and 3.11 for two of the nodes in the circuit, the cleaner feed node, and the entire plant node, respectively.

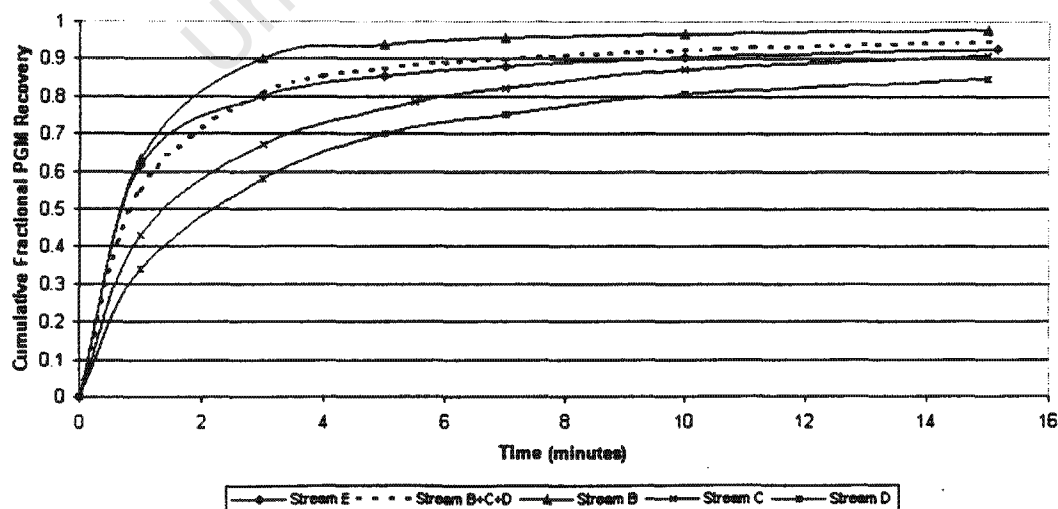


Figure 3.10: Predicting the recovery time profile around the cleaner feed mixing node showing the conservation of the PGM floatability components

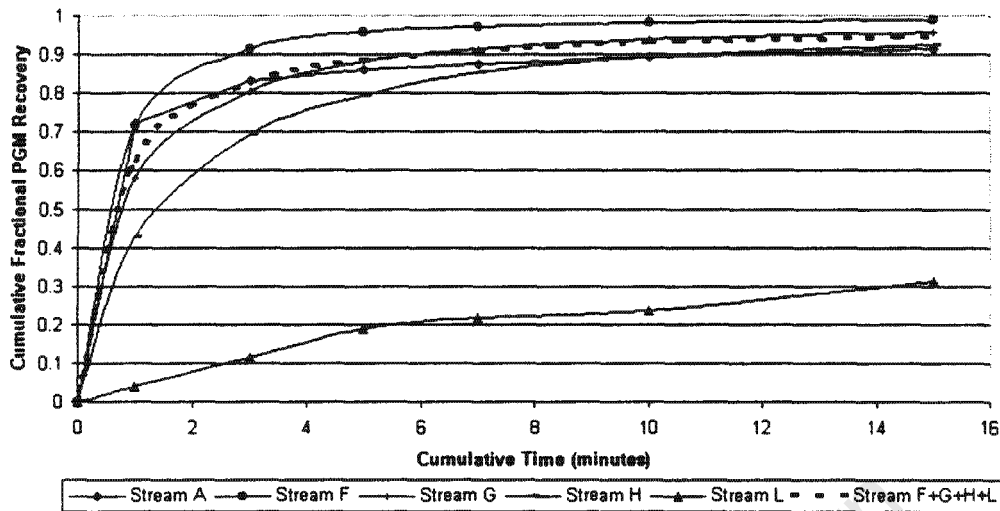


Figure 3.11: Predicting the feed stream recovery time profile from the final product streams of the mini plant showing the conservation of PGM floatability components over the whole circuit

In figures 3.10 and 3.11 it can be seen that good correspondence was obtained between the calculated and experimental results. The results obtained for the other nodes in the circuit was equally good. This indicates that the floatability of the PGM particles was not changing with time in the mini plant. On this basis it is possible to combine all the stream floatability tests, with all of them linked by the same set of floatability rate constants subjected to the constraint that the floatability component flow rates comply with the mass balance constraints within the flotation circuit.

3.4.2 Initial PGM parameter estimation algorithm – using only batch floatability data

This section describes the method that was used to derive the parameter set to predict the PGM flow rates around the mini plant. It has been demonstrated that the various streams within a flotation circuit represents the outcome of a fractionation process of the mineral originally in the feed to the circuit. This implies that the floatability distributions of all the streams within a circuit can be linked back to the “parent” floatability distribution of the mineral in the feed to the circuit.

The Non-Linear Optimisation Procedure

The basis of this section comes from a paper written by Woodburn and Wallin (1984), “Decoupled kinetic model for simulation of flotation networks”. This section presents a method of modelling any network of flotation cells, using a spreadsheet to ensure the mass balance constraints of all the components in the circuit are maintained. The method presented here is as an example applied to the circuit configuration used for the mini plant test work depicted in figure 3.12. It is important to note that all tailing streams

In any flotation network consist of individual streams and are never a combination of two or more streams and are always unique to an individual unit in a particular circuit. The material balance (tailings flows) are calculated by the solution of simultaneous mass balances of an individual subclass of a species over all the cells in a circuit. An individual subclass, (i), of species (j) is defined by its mass fraction, m_{ij} , in the feed solids, and in terms of its first order rate constant, k_{ij} .

Example: For PGM's, $m_{f,PGM}$ = fast floating fraction of PGM in the feed
 $P_{f,PGM}$ = fast floating floatability of PGM
 For Cr_2O_3 , m_{d,Cr_2O_3} = fraction of Cr_2O_3 in size class (d) in the feed
 E_{d,Cr_2O_3} = entrainability of a Cr_2O_3 particle of size class (d)

The calculated set of tailings flows over all the cells in the network is given by the vector δT_{ij}^T of each individual subclass (i) of species (j). The total species flow (ie Cr_2O_3 or PGM) is obtained by summing the flows of the sub-classes (n).

$$T = \sum_{j=1}^{i=n} \delta T_{ij}^T \dots\dots\dots 3.1$$

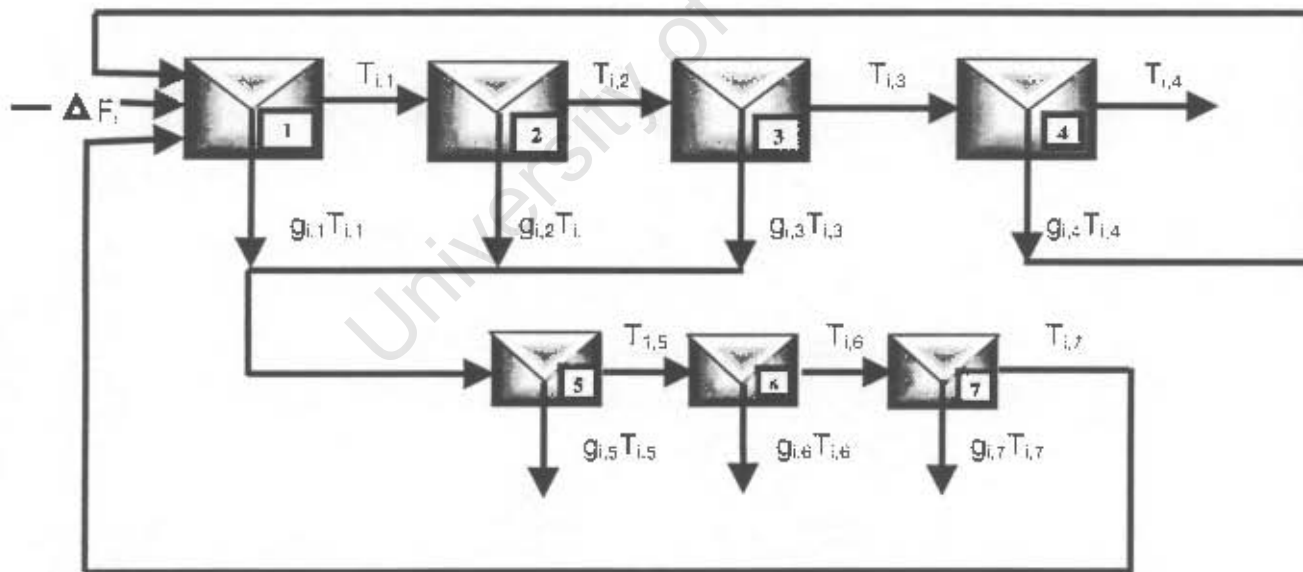


Figure 3.12: Flowsheet illustrating the formulation of a connexion matrix

The material balance is formulated in terms of a set of enhancement factors, g_{ij} , which is specific to a particular material (PGM, Cr_2O_3) and to each sub-class of that material (fast PGM, slow PGM, etc.). It is also defined by the set of operating conditions within a particular cell. A single enhancement factor is defined as the ratio of the flow rate of a

specific solids fraction in the concentrate to the flow of the same material in the tailings. Consider a single cell (k) in the circuit of figure 3.12 as shown in figure 3.13 below. Consider the mass balance in terms of a sub-class (i) of species (j).

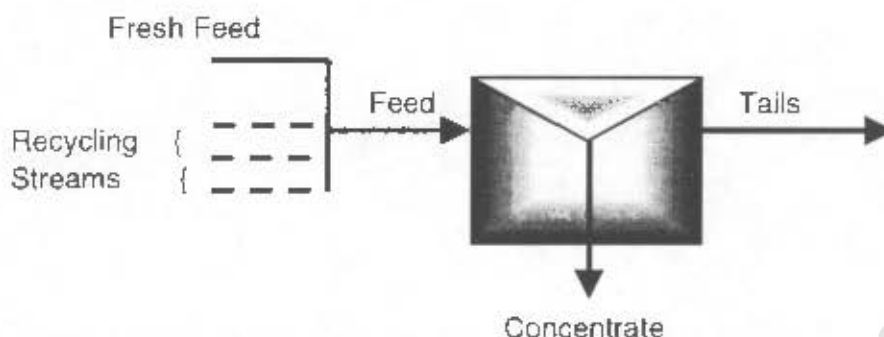


Figure 3.13: A single flotation cell showing all the possible streams entering and leaving the system

If the fraction of fresh feed of species (j) fed to cell (k) is $\gamma_{j,k}$ and the fraction of subclass (i) is $m_{i,j,k}$ then the flow rate of the total fresh feed (ΔF_k) of species (j) entering cell (k) is;

$$\gamma_{j,k} m_{i,j,k} \Delta F_k \dots\dots\dots 3.2$$

The rest of the feed to cell (k) is a combination of recycling streams from other cells in the circuit.

Recycling Tails Flows:
$$\sum_{n=1}^{n=7} a_{kh} \delta T_{i,j,n} \dots\dots\dots 3.3$$

Recycling Concentrate Flows:
$$\sum_{n=1}^{n=7} b_{kh} g_{i,j,n} \delta T_{i,j,n} \dots\dots\dots 3.4$$

where

$a_{k,n}$, $b_{k,h}$ fractions of tailings and concentrate re-circulated from cell h to cell k for the seven nodes in figure 3.12.

A common assumption of most modelling methodologies is that the sub-classes of each mineral species float independently from one another. Also, the individual flow rates of each sub-class are conserved around stream nodes and flotation cells in the circuit. Therefore the mass balance of a flotation cell (k) operating at steady state is:

Mass of sub-class (i), species (j) IN = Mass of sub-class (i), species (j) OUT

Feed = Concentrate + Tailings

$$\gamma_{j,k} m_{i,j,k} \Delta F_k + \sum_{h=1}^{h=7} (a_{k,h} + b_{k,h} g_{i,j,h}) \delta T_{i,j,h} = (1 + g_{i,j,k}) \delta T_{i,j,k} \dots\dots\dots 3.5$$

Note, the equation above can be defined for every cell in the circuit, generalising this as a set of seven equations for the whole circuit;

$$C(g) \delta T_{ij}^T = -\gamma_{j,k}^T m_{ij} \Delta F_k \dots\dots\dots 3.6$$

Solving this for the set of tailings flows of species (j) for all cells in the circuit (h) with (z) nodes;

$$T_{jk}^T = -\Delta F_k \sum_{h=1}^{h=z} m_{ij} C^{-1}(g) \gamma_{j,k}^T \dots\dots\dots 3.7$$

Therefore the tailings flows of species (j) in subclass (i) of unit (k) can be calculated in terms of a connexion matrix $C(g)$ that is a function of a set of enhancement factors. Table 3.4 below gives an example of the connexion matrix applied to the circuit in figure 3.12 for an individual species (j) in subclass (i).

Table 3.4: Matrix formulation for mass balance of species (j) in sub-class (i)

MASS FLOW OF TAILINGS T SPECIES (j), T_i^T	FRESH FEED FRACTION ADDED TO CELL, γ^T	MATERIAL RECEIVED BY	MATERIAL CIRCULATED FROM						
			CELL 1	CELL2	CELL3	CELL4	CELL5	CELL6	CELL7
T_{i1}	1	Cell 1	$-(1+g_{i1})$	0	0	g_{i4}	0	0	0
T_{i2}	0	Cell 2	1	$-(1+g_{i2})$	0	0	0	0	0
T_{i3}	0	Cell 3	0	0	$-(1+g_{i3})$	0	0	0	0
T_{i4}	0	Cell 4	0	0	0	$-(1+g_{i4})$	0	0	0
T_{i5}	0	Cell 5	g_{i1}	g_{i2}	g_{i3}	0	$-(1+g_{i5})$	0	0
T_{i6}	0	Cell 6	0	0	0	0	0	$-(1+g_{i6})$	0
T_{i7}	0	Cell 7	0	0	0	0	0	0	$-(1+g_{i7})$

A stream matrix as described above is formed with all the streams in the entire circuit linked to the feed of the circuit via the mass balance equations and the enhancement parameters.

Example applied to the PGM floatability components:

- The mass flow rate in the first rougher concentrate stream for a floatability class is calculated as the mass flow rate in the rougher tails multiplied by the enhancement parameter in the matrix.
- The mass flow rate in the first rougher feed stream for the same floatability class is calculated as the summation between the mass flow rates in the rougher tails and the calculated rougher concentrate stream due to the mass balance constraint.
- The same is done for each floatability component, and the sum of all the floatability components is equal to the overall stream recovery based on the rougher feed.
- The second rougher cell is treated in the same manner, except the concentrate and tails streams from this cell are linked back to the first rougher feed stream etc.

Initially the batch floatability tests were used as floatability component mass balance tools in order to determine an initial estimation of the mass fractions of the feed floatability components. Enhancement factors describing the split of the PGM floatability components around each node in the circuit were set as parameters during the non-linear regression procedure. This ensured that the mass balance constraints imposed around every unit and node were maintained for each floatability component. The only additional parameters required pertained to the floatability distribution of the feed stream to the mini plant. The information available to conduct this analysis can be summarised as follows:

- Number of batch tests = 15
- Number of concentrate samples per batch test = 6
- Therefore the number of experimental data points = $15 \times 6 = 90$

The number of floatability components used to describe the feed floatability distribution was initially assumed to be four, including a non-floating mineral fraction. Expanding equation 2.70 highlights the six parameters required to describe the feed floatability distribution, based on this assumption:

$$R_{feed}(t) = m_{feed,f}(1 - e^{-k_f t}) + m_{feed,m}(1 - e^{-k_m t}) + m_{feed,s}(1 - e^{-k_s t}) \dots\dots\dots 3.8$$

- Number of enhancement factors = (4-1) fractions * 7 nodes = 21
- Therefore the total number of parameters = $21 + 6 = 27$

The solver subroutine in excel was used to regress the first approximation of the feed stream floatability distribution by varying the six model parameters and the 21 enhancement parameters. The regression subroutine minimised the total sum of error squared (TSSE) between the calculated and the 90 experimental batch floatability test data points.

$$TSSE = \sum_{x=1}^y \left[\frac{(d_{\text{experimental}} - d_{\text{predicted}})}{(StDev)} \right]^2 \dots\dots\dots 3.9$$

where

$d_{\text{experimental}}$ is the value of the experimental data point
 $d_{\text{predicted}}$ is the predicted value of the same data point using the regressed flotation model parameters
 $StDev$ is the standard deviation of the experimental data point
 y is the total number of data points used to derive the floatability component parameters

One of the difficulties in conducting this analysis is that the problem is highly non-linear and highly susceptible to the identification of false minima. It is therefore necessary to initiate the solution with starting values of the parameters. There are a lot of parameters defining the problem, but the solution to the parameters are highly constrained as the optimisation could only fit the data while the components were conserved around nodes and while the floatability component mass fractions added up to one in every stream within the circuit. The fit to all the batch floatability curves was excellent as would have been expected as the number of parameters used was large. It must be emphasised however, that this first step was done to obtain an initial approximation of the feed floatability mass distribution.

3.4.3 Using the mini plant PGM model in conjunction with the mass balance campaign data

The next step of the parameter estimation algorithm was to use the plant PGM flow rate data, obtained from the sampling campaign at steady state, together with the batch data in order to define the floatability distribution of the feed ore (in terms of the floatability parameters P_i). Equation 3.10 was used to describe the recovery of each floatability component in each flotation cell of the mini plant, assuming perfect mixing in the pulp (as referred to in section 2.5).

$$R_{j,k} = \sum_{i=1}^n m_{i,j,k} \left(1 - \frac{1}{1 + ((1 - \alpha_j) \exp(-\beta_{k,j} * \tau_{f,k}) + \alpha_j) P_{i,j} S b_k \tau_{c,k}} \right) \dots\dots\dots 3.10$$

where

$R_{j,k}$ is the recovery of mineral j from flotation unit k
 n is the number of floatability classes
 $m_{i,j,k}$ is the mass fraction of mineral j of floatability class i in the feed to unit k
 $\beta_{k,j}$ derived circuit (or bank) froth stability parameter of mineral j

α_j	derived froth parameter related to the amount of “non-draining” mineral j
$\tau_{f,k}$	froth retention time in unit k based on concentrate flow rate
$P_{i,j}$	is the inherent floatability of mineral j in floatability class i
Sb_k	is the bubble surface area flux (bubble surface/time/cell area) in unit k
$\tau_{c,k}$	is the flotation cell pulp residence time in unit k

The cell residence time (volume pulp / tailings flow rate) as well as the froth residence time based on concentrate flow rate (volume froth / concentrate volumetric flow rate) was calculated from the experimental data. Using this model it is possible to predict all the enhancement factors which were set as parameters in section 3.4.2. This reduces the number of model parameters from 27 to 12 as described below:

$k_{fast,batch}$	} batch flotation rate constants	$m_{fast,PGM}$	} mini plant feed PGM floatability distribution
$k_{medium,batch}$		$m_{medium,PGM}$	
$k_{slow,batch}$		$m_{slow,PGM}$	
P_{fast}	} plant mineral floatabilities	α	} plant froth parameters
P_{medium}		$\beta_{rougher,PGM}$	
P_{slow}		$\beta_{cleaner,PGM}$	

It can be argued that the mineral flotation rate parameters used in the batch tests should be directly related to the plant floatability parameters based on the bubble surface area flux in the batch cell. In this study, it was decided not to adopt this assumption as it was considered uncertain to what extent the different pulp turbulent effects could have affected the absolute values of the floatability parameters between the batch and mini plant flotation cells owing to the much higher specific energy input in the batch cell. One method of linking the floatabilities between both units is to incorporate a scale-up number that can be multiplied to all the batch floatability parameters as reported by Runge et al (1997). For the results presented here, it was assumed that the floatabilities between both units were completely independent from one another and separate parameters were used for both systems. Essentially this means that the batch tests were only used to help constrain the problem of determining the mass fractions of each floatability component of the feed to the mini plant.

It was assumed that the froth stability parameters were constant for all cells in the rougher and scavenger banks. The cleaner cells however were given an independent froth stability parameter assuming that the characteristics of the froth phase would be substantially different from the rougher bank. In order to define the froth stability parameters for the rougher and cleaner banks, the froth characterisation test data (section 3.2.2) was included into the optimisation procedure. This added a further 130 data points to the regression analysis, at the cost of introducing a further 6 parameters (plant mineral floatabilities and plant froth parameters). This greatly contributed to the number of data points driving the regression of the parameters. The same model parameters derived in section 3.4.2 were used to predict the PGM flow rates for these tests.

- Number of plant data points = 44 from the feed characterisation tests
= 130 from the froth characterisation tests
- Number of batch test data points = 90
- Total number of data points (N) = 264

- Number of parameters (P) = 12

When the model was applied to the full data set, the mass fractions of the feed PGM floatability distribution derived in section 3.4.2 were initially kept constant. Using the solver subroutine in Excel, the TSSE between the calculated and measured data points from all the batch tests as well as the plant mass balance PGM flow rates were minimised. This was done by varying the mini plant floatabilities and the three froth parameters. Once the best set of parameters had been obtained, all the parameters, including the feed PGM floatability mass fractions and the batch floatability rate constants were fitted simultaneously. This parameter estimation algorithm shows how the problem of deriving the twelve parameters depicted in table 3.5 can be obtained by splitting the problem into three stages. Firstly deriving the feed floatability mass fraction parameters using only the batch tests subject to the mass balance constraints imposed by the circuit on which batch tests were conducted, followed by incorporating the plant kinetic model to derive the floatability parameters of the mini plant while keeping the floatability mass fraction parameters constant. The final step was then to regress all the parameters together to obtain the minimum TSSE. This method is believed to be more accurate and provide a reliable initial approximation of the feed floatability component mass fractions and therefore the subsequent final parameter set for this highly non-linear problem. The parameters obtained in this analysis are presented in Table 3.5.

Table 3.5: Model parameters for the PGM kinetics of the mini plant

PARAMETER		VALUE
Fast Floatability Parameter	$P_{fast,PGM}$	5.06e-1
Medium Floatability Parameter	$P_{medium,PGM}$	5.46e-4
Slow Floatability Parameter	$P_{slow,PGM}$	1.03e-4
Feed Fraction Fast Floating	$m_{fast,PGM}$	0.212
Feed Fraction Medium Floating	$m_{medium,PGM}$	0.643
Feed Fraction Slow Floating	$m_{slow,PGM}$	0.104
Rougher Froth Stability	$\beta_{PGM,rougher}$	0.0593
Cleaner Froth Stability	$\beta_{PGM,cleaner}$	0.0484
Derived Froth Parameter	α_{PGM}	9.54e-17
Batch Fast Floating Kinetic Parameter	$k_{PGM,f}^{batch}$	1.18e+1
Batch Medium Floating Kinetic Parameter	$k_{PGM,m}^{batch}$	8.67e-1
Batch Slow Floating Kinetic Parameter	$k_{PGM,s}^{batch}$	7.90e-2

Equation 3.11 is a measure of how well the model that is being used is able to predict the experimentally determined data points relative to their known standard deviations. A model fit of one essentially means that all the data points were predicted within the measured standard deviations of the experimental data points.

$$MF = \sqrt{\frac{TSSE}{(N - P)}} \dots\dots\dots 3.11$$

where

TSSE is the total sum of the error (calculated data point – measured data point) squared defined by equation 3.9

N is the total number of experimental data points

P is the total number of parameters used in the model

A model fit of $MF = 1.41$ was obtained using the parameter set in table 3.5. This shows that the fit to the data was relatively good and within the variability of the PGM flow rates of the mini plant. Figure 3.14 compares the calculated versus experimentally determined PGM flow rates (mg/hr) from the four mass balances. It is clearly evident that the PGM flow rate predictions get progressively worse down the rougher bank with the model under predicting the PGM flow rates in rougher C and the scavenger cell. The same trend is seen down the cleaner bank. It is hypothesized that this is as a result of not including an entrainment contribution to the overall PGM recovery. It would make sense that the model prediction would get worse when progressing down a bank of cells as the contribution to entrainment starts dominating once most of the hydrophobic particles have floated in the first few cells of a bank.

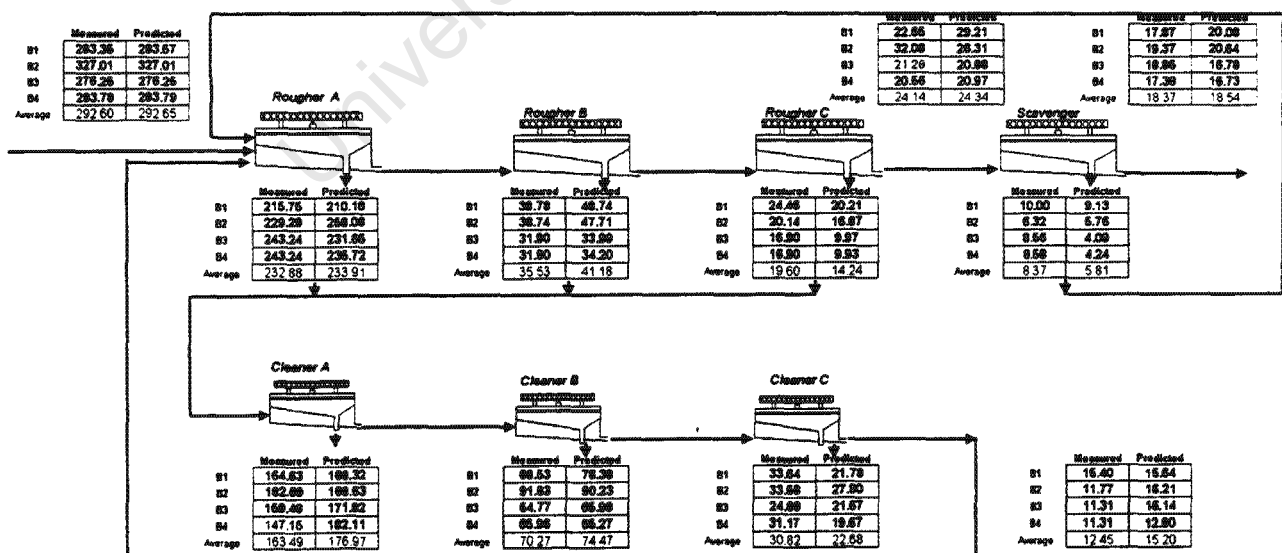


Figure 3.14: Flow sheet of the mini plant showing the measured and predicted PGM flow rates (mg/h) from the four mass balances

Figure 3.15 shows the comparison of the model fit to the batch floatability cumulative PGM recoveries. The calculated PGM recovery time profile was consistently over predicted for the cleaner tailings stream (stream I). It is again hypothesized that this is due to the lack of an entrainment contribution in the PGM model. In reality, a lower proportion of floatable components would be present in the cleaner tailings stream due to some being recovered by entrainment with water into the final concentrate. The model did not account for the decrease in floatable components in this stream due to entrainment causing it to predict a higher batch floatability response.

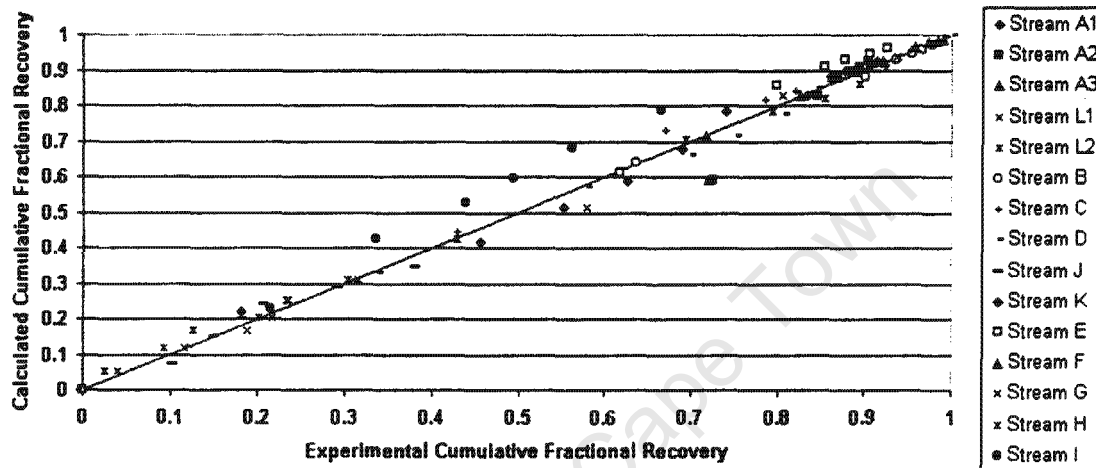


Figure 3.15: The PGM model fit to the batch floatability tests of the mini plant

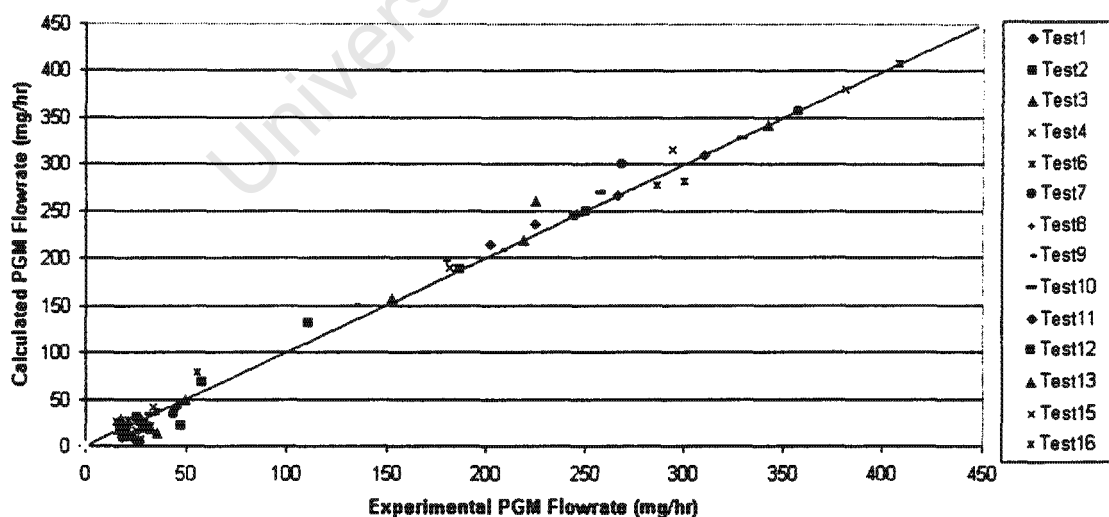


Figure 3.16: The PGM model fit to the rougher froth characterisation tests of the mini plant

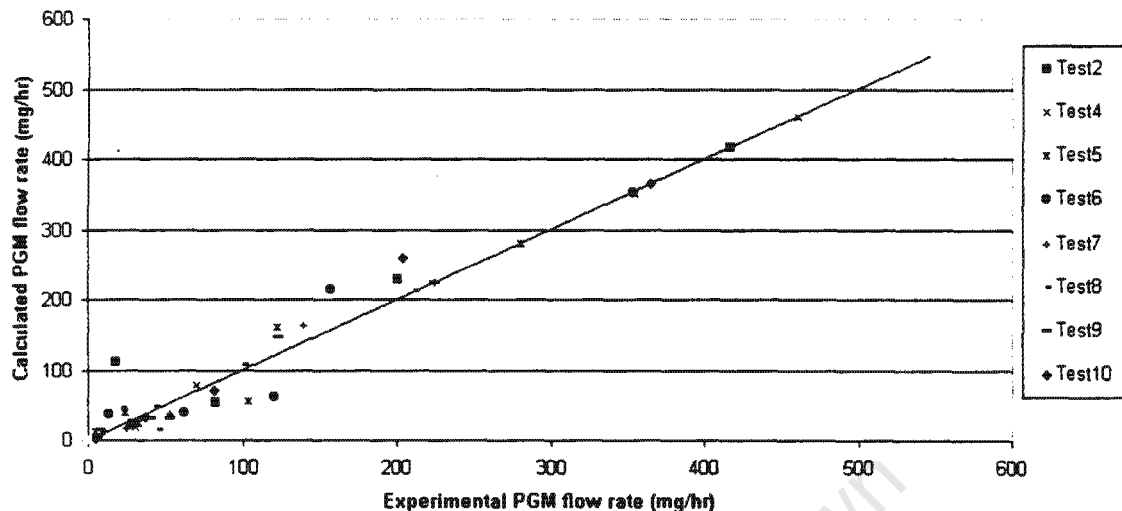


Figure 3.17: The PGM model fit to the cleaner froth characterisation tests of the mini plant

Figures 3.16 and 3.17 show the model fit for the froth characterisation tests. Again the fit was considered good showing that the froth stability parameters in the froth model were accurately predicting the PGM kinetics as a function of froth depth and volumetric air flow rates in the first rougher and cleaner flotation cells. This indicates that the froth model used in the analysis of this data is able to characterise the froth performance by allocating new froth recovery parameters under different conditions. This provides a significant improvement to most of the existing models in the literature that does not provide a basis for the allocation of new froth recovery parameters, and treats the froth recovery as constants during circuit simulation which is clearly unrealistic. As mentioned earlier, the froth model assumes a constant froth recovery in a particular unit as a function of floatability class. This assumption is known not to be true for different particle sizes, and it is believed that the froth model can be refined even further by incorporating a size dependent froth model into this analysis.

3.4.4 Monte Carlo analysis on the preliminary regressed parameters

It has been stated that the regression method is highly non-linear and that the problem is highly susceptible to finding false minima during the parameter regression analysis. While the method presented previously represents a highly constrained non linear regression, particularly due to the fact that there are many experimental data points driving the regression analysis, there is still no guarantee that the resulting parameters are in fact producing the absolute minimum TSSE. Another problem with non-linear regressions, particularly ones with a large number of parameters is to determine the sensitivity of the error to the various parameters. The parameter set defining the mini plant PGM kinetics was presented in table 3.5. A method is required to quantify the error associated with the parameters from the floatability component modelling exercise.

The parameter set that was obtained can be considered to be from a distribution of possible parameter sets. The Monte Carlo Boot Strap technique can be used to evaluate and quantify the error associated with the model parameters. The principle of this method is summarised in Appendix B. A program was written in Visual Basic to produce random numbers that were normally distributed between -3 and $+3$. Every random number generated was multiplied with a measured standard deviation from the experimental data obtained from the mini plant test work. The resulting number was then added onto the experimental data point to produce an artificial data point that was within three measured standard deviations. A macro was written to repeat the same calculation on every experimental data point generated by the mini plant tests. This produced a completely new artificial data set. Using the newly created data set, a new parameter set could be regressed. The program was allowed to run for 1000 iterations, producing 1000 artificial data sets (distributed about their standard deviations) which were used to derive 1000 model parameter sets. The average and standard deviations of all the parameters were determined. The standard deviation of each parameter is a measure of how well defined the parameter is. The results of the Monte Carlo analysis can be seen in table 3.6. The frequency distribution of the feed stream PGM flow rate (mg/hr) is shown in figure 3.18. Each point shows the number of times the value of the feed stream was randomly chosen about its standard deviation of 283.35 ± 23.2 mg/hr. This figure illustrates the normal distribution that was used on the feed stream data point from mass balance 1.

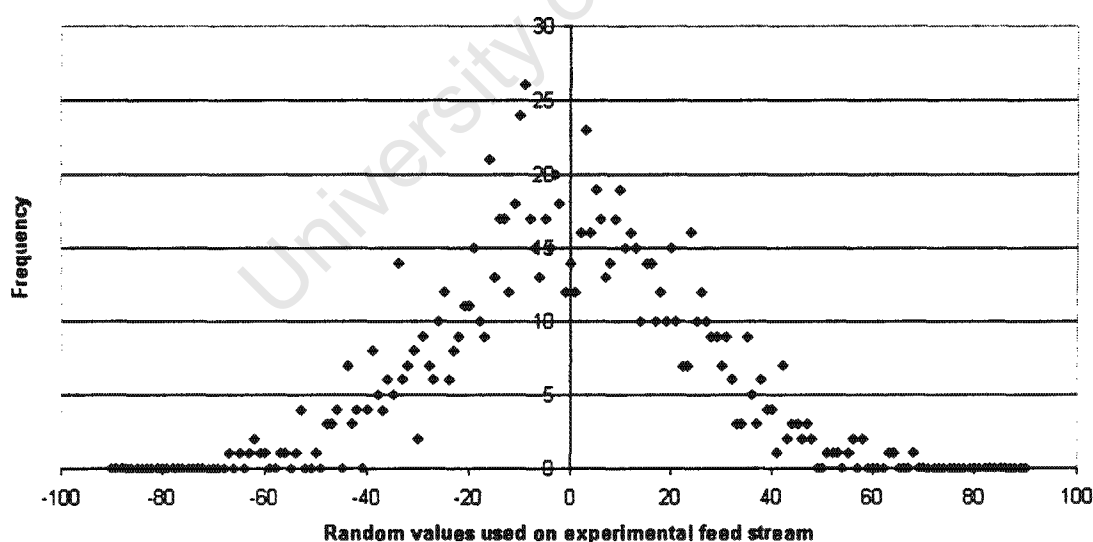


Figure 3.18: Normal frequency distribution of random standard deviations used in the Monte Carlo analysis of the feed stream PGM flow rate from mass balance 1

Table 3.6: Statistical analysis of the Monte Carlo results derived from 1000 parameter sets

PARAMETER	"BEST FIT"	AVERAGE	STD DEV.	DEVIATION (%)	95 % CONFIDENCE INTERVAL
$P_{fast,PGM}$	5.06e-01	7.72e-01	1.37e-01	17.69	1.04 – 0.504
$P_{medium,PGM}$	5.46e-04	6.96e-04	1.14e-04	16.34	9.18e-4 – 4.73e-4
$P_{slow,PGM}$	1.03e-04	1.30e-4	2.17e-05	16.45	1.72e-4 – 8.8e-5
$m_{fast,PGM}$	0.211	0.192	0.0397	20.68	0.270 – 0.114
$m_{medium,PGM}$	0.644	0.663	0.0363	5.48	0.735 – 0.592
$m_{slow,PGM}$	0.104	0.107	0.00129	12.07	0.132 – 0.0818
α_{PGM}	9.54e-17	9.54e-17	1.59e-23	0.00	9.54e-17 – 9.54e-17
$\beta_{rougher,PGM}$	0.0594	0.0817	0.0164	20.018	0.114 – 0.0497
$\beta_{cleaner,PGM}$	0.0484	0.067	0.0147	21.99	0.0959 – 0.0381
$k_{fast,PGM}^{batch}$	11.77	11.78	0.636	5.40	13.025 – 10.532
$k_{medium,PGM}^{batch}$	0.871	0.923	0.116	12.57	1.151 – 0.696
$k_{slow,PGM}^{batch}$	0.0791	0.0778	0.0072	9.19	0.0918 – 0.0638

The Monte Carlo analysis was able to give insight into the interactions of the model parameters. Very high standard deviation percentages were calculated for both of the froth stability parameters indicating that these two parameters had the greatest influence on the regression. It is proposed that this is due to the fact that single froth stability parameters were assumed for each bank on the mini plant. It was evident that the froth stability was not identical for every rougher cell and that perhaps different froth stability parameters were needed for every flotation cell in the circuit. This is however not considered truly necessary as a satisfactory model fit was still obtained using single froth stability parameters for each bank in the circuit. It does however give a possible reasoning for the higher standard deviations of these parameters. The derived froth parameter representing the amount of non-draining mineral was essentially zero for all the optimisations indicating that the use of this parameter is not necessary for the mini plant data set. This was possibly due to the fact the mini plant flotation cells had only single sided concentrate launders and the froth was therefore well drained having virtually zero non-draining PGM minerals in the froth.

All the parameters defining the PGM flow rates have a standard deviation of smaller than 21%. This may be considered relatively high, but it must be remembered that the standard deviations of the experimental data points were also high, owing to the inherent variability of the flotation process. No anomolous parameter sets were obtained for all the various combinations of data sets produced. It is for this reason that it is believed that the parameter set defined in table 3.6 is robust enough to give a reasonable quantitative description of the PGM flotation kinetics of the mini plant.

3.5 Model used to predict the chromite recovery on the mini plant

It is very important to characterise the chromite recovery kinetics in UG2 ore flotation as chromite causes problems during the later smelting stages of the concentration process, (Liddell et al, 1986). In fact, the overall design of UG2 flotation circuits are aimed specifically at minimising the recovery of chromite into the final concentrate. In UG2 ore flotation, the recovery of chromite is generally considered to occur via an entrainment mechanism. This implies that the chromite particles are carried over into the concentrate of the flotation cell in the wake of water surrounding the ascending air bubbles. Figure 3.19 shows the trend obtained with respect to water and chromite for all the data obtained from the circuit characterisation test as well as the froth characterisation tests. What is important to note is that the data points in figure 3.19 were obtained from different cells in the circuit operated at very different froth depths and volumetric air flow rates, and that these all fell onto the same trend. Simply using this relationship to predict the recovery of chromite into the concentrate could be considered reasonable but this would be purely empirical and would not provide any insight on the mechanism of chromite entrainment into the concentrate. This is particularly relevant when considering the influence of particle size on recovery by entrainment. Figure 3.20 shows the relationship for the recovery of chromite versus the recovery of water obtained from the froth characterisation tests.

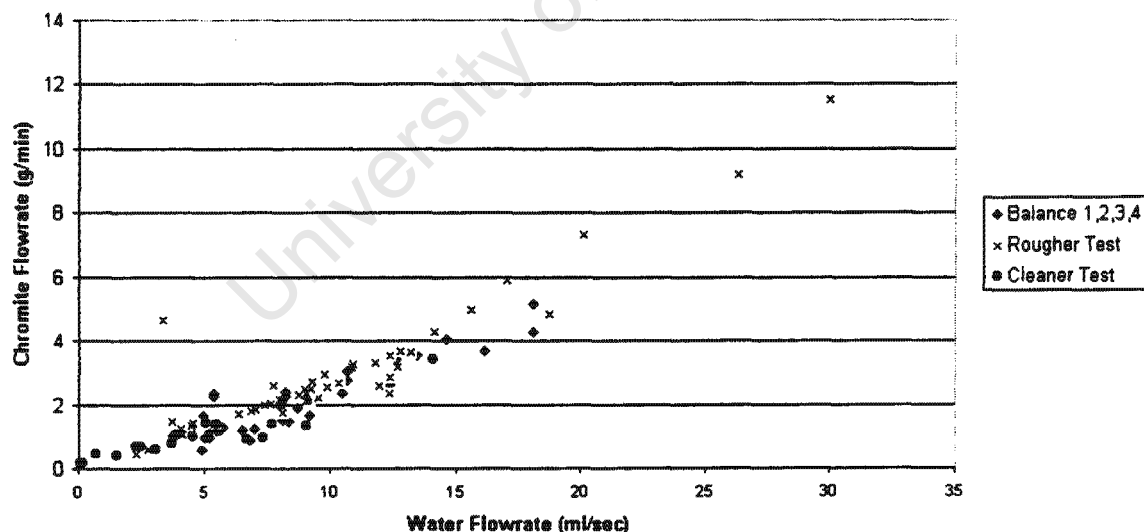


Figure 3.19: Relationship describing the entrainment of chromite with water from the mini plant tests

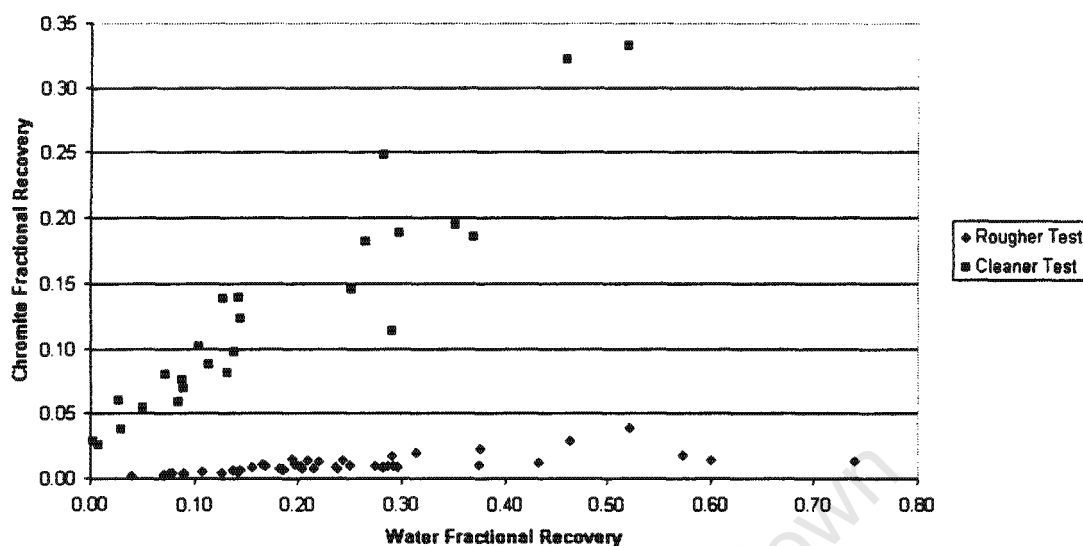


Figure 3.20: The linear relationship obtained between the chromite and water recovery from the froth characterisation tests on the mini plant

It can clearly be seen that two distinct entrainability functions (or slope parameters) are obtained for both the rougher and cleaner cells. A much smaller chromite recovery is obtained in the rougher cells than in the cleaner cells at the same water recoveries in each bank. To try and understand why this is the case, consideration should be given to the amount of entrainable chromite in the pulp zones (collection zones) within each bank. The rougher contains much more chromite than the cleaner bank, but only a small fraction of this is entrainable due to a difference in particle entrainability with size. The chromite in the cleaner cells constitute small entrained chromite particles with a higher entrainability and hence higher recovery. The extent of entrainment of a chromite particle varies with size, a large chromite particle would have a smaller probability of being entrained by water than a smaller chromite particle. The overall recovery of chromite would represent a summation of the particle size classes with different degree of entrainments for each particle size class.

An important point worth noting in figure 3.20 is that the recovery of chromite intercepts the y-axis at zero water recovery. Work done by Warren (1985) considered such a phenomenon on a system of de-slimed sulfide-rich cassiterite ore. Laboratory batch flotation tests were carried out keeping the chemical conditions constant but the water recovery was varied by changing the height of the froth column and the rate and depth of removal of froth. The recovery of the mineral was found to be linearly related to the water recovery. The intercept of the regression line on the mineral recovery axis, where the water recovery was zero, was interpreted as the recovery due to true flotation. The entrainment contribution was proportional to the slope of the line. In this way the contributions of entrainment and true flotation to the overall recovery was separated. On this basis, the results presented in figure 3.20 suggest that the contribution of true flotation to the chromite recovery was minimal. However, in order to account for the

possibility that some chromite was recovered by true flotation (owing to locking for example) a fraction of the chromite in the feed to the circuit was included as a floating fraction, using a standard first order rate equation to describe the kinetics of the floatable chromite fraction. On this basis, the following equation was used to model the chromite:

$$R_{Cr,k}^o = (1 - \phi_{act,Cr}) * \sum_{i=1}^d (E_{Cr,d} * R_{w,k}) + \phi_{act,Cr} * \left(1 - \frac{1}{1 + P_{Cr} S_{b,k} \tau_{c,k} * ((1 - \alpha_{Cr}) * \exp(-\beta_{Cr,k} \tau_{f,k}) + \alpha_{Cr})}\right) \dots\dots\dots 3.12$$

where

$R_{ent,Cr,k}^o$ is the total chromite recovery by entrainment in all the size fractions unit k
 $E_{Cr,d}$ is the degree of entrainment of a chromite particle in size class d
 $R_{w,k}$ is the measured water recovery from flotation cell k
 $\phi_{act,Cr}$ represents the fraction of chromite in the feed which floats by true flotation

$$E_{Cr,d} = \frac{2}{\exp(2.292 * (\frac{S_{Cr,d}}{\xi_{Cr}})^{ADJ}) + \exp(-2.292 * (\frac{S_{Cr,d}}{\xi_{Cr}})^{ADJ})} \dots\dots\dots 3.13$$

$$ADJ = 1 - \frac{\ln(\frac{1}{\delta_{Cr}})}{\exp(\frac{S_{Cr,d}}{\xi_{Cr}})}$$

where

$S_{Cr,d}$ is the average chromite particle size in the i^{th} size class d (in microns)
 ξ_{Cr} is the entrainment parameter for chromite (given by the particle size for which the degree of entrainment is 20 percent)
 δ_{Cr} is the drainage parameter of chromite (related to the preferential drainage of coarse particles from the froth)

If the particle size distribution of chromite in the feed to a circuit is known, then the feed chromite can be separated into different particle size fractions. The chromite recovery of each fraction can then be assessed individually throughout a given circuit. The overall recovery in a stream would then be the summation of each size fraction's recovery as well as the floatable chromite contribution in each stream. In this phase of the study, the chromite particle size distribution of the feed to the mini plant was not measured. The

Gaudin-Meloy particle size distribution function¹ was used to approximate this by regressing the parameters that describe this function from the mini plant data. This distribution function was used as it has been found to fit many particle size distributions found in practice. Knowing the particle size distribution of the feed to the circuit it is possible to calculate the mass fraction of chromite (m) in each size class (d).

$$Y = 1 - \left(1 - \frac{X}{X_d}\right)^r \quad \dots\dots\dots 3.14$$

where

- Y cumulative fraction weight undersize
- X particle diameter (μm)
- X_d parameters with dimension of size (μm)
- r dimensionless exponent

The data from the circuit characterisation, rougher, cleaner, and alternative configuration tests were used together with the model developed in equation 3.12 to minimise the sum of differences between the experimental chromite recoveries and the model chromite recoveries for each stream in each test. The water recovery, flotation cell and froth residence times were determined from experiment. A relatively good fit was obtained from this regression. Figure 3.21 shows the particle size distribution curve regressed from the optimisation for the mini plant feed stream as well as for the rougher concentrate stream (cleaner feed). The derived entrainability curves are also shown.

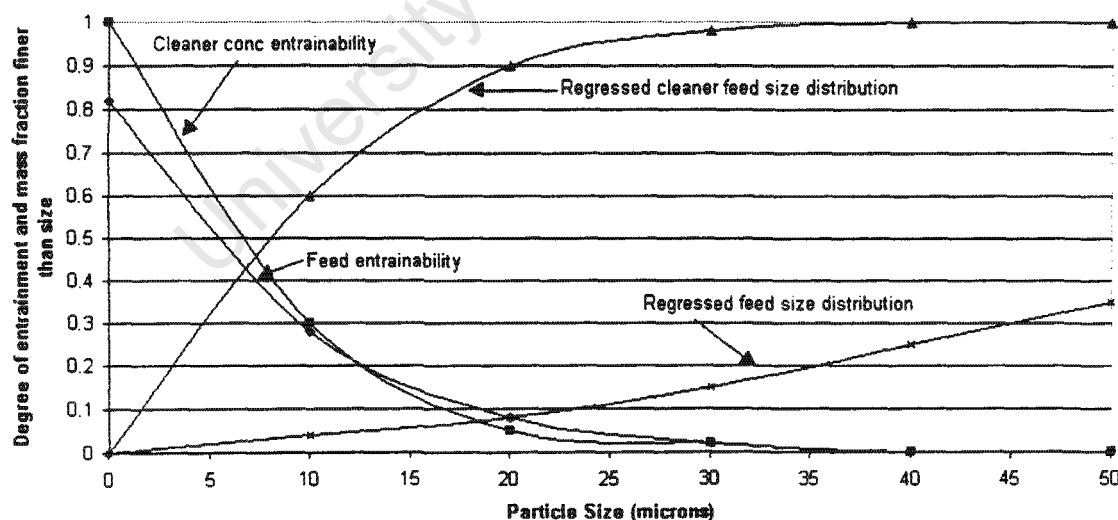


Figure 3.21: The mini plant entrainability function and particle size distributions of the feed stream and the rougher concentrate stream

¹ See Chemical Engineers Handbook, Chapter 8, Section 5 (1995 ed.)

The entrainability curves of the feed and cleaner feed streams in figure 3.21 imply that chromite particles greater than approximately 40 μ m have a zero probability of being entrained into the concentrate of any cell. This fits in well with the fact that the particle size distribution for the cleaner feed stream derived from the optimisation shows that all the particles in this stream are smaller than approximately 40 μ m. Confirmatory tests were conducted in order to verify the results obtained from the modelling exercise. A sample of the mini plant feed stream and a combined sample of each rougher concentrate were split into different size fractions. The percentage chromite was then measured in each size fraction, the results are presented in table 3.7. These results show that over 90% of the chromite particles in the rougher concentrate are smaller than 53 μ m, supporting the prediction of the modelling. The regressed particle size distribution of the feed stream compared well to the experimental feed particle size distribution. The experimental chromite particle size distribution of the mini plant feed stream was used in place of the derived size distribution, to determine how well the proposed model could predict the chromite recoveries from the mini plant tests. This reduced the number of parameters in the optimisation by two providing a better model fit to the data. The regressed parameters providing the smallest TSSE are listed in table 3.8 and the experimental chromite particle size distributions are plotted in figure 3.22.

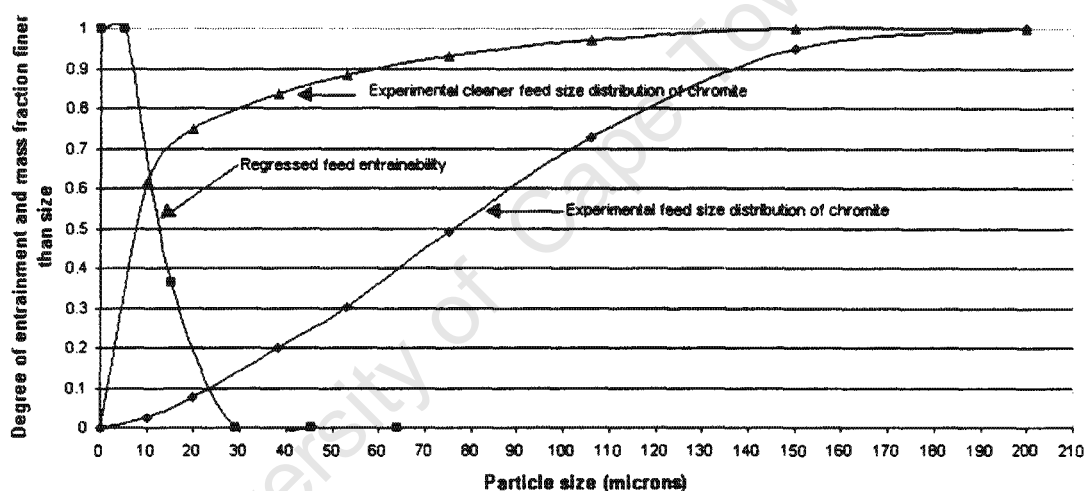
Table 3.7: The chromite particle size distributions of the feed and rougher concentrate streams from the mini plant circuit characterisation test

Particle Size Class (μ m)	FEED STREAM			ROUGHER CONCENTRATE STREAM		
	Dry Mass (g)	Chromite (%)	Cum. % Less Than	Dry Mass (g)	Chromite (%)	Cum. % Less Than
-38	74.8	24.41	25.44	102.4	12.76	88.40
+38-53	36	24.08	37.52	18.7	2.94	92.12
+53-75	61.2	22.83	56.99	21.7	2.74	96.14
+75-106	85.9	19.7	80.56	16	2.01	98.32
+106-150	66.2	16.35	95.64	10.1	2.46	100
+150	15.4	20.32	100			
Total	339.5			168.9		

A total of eight parameters were regressed from 240 experimental data points producing a model fit of 1.44. A comparison of the model fit to the experimental data is plotted in figures 3.23-25. Only 0.3 percent of the feed material was calculated to be activated to float, but this amounts to 9.5 percent in the cleaner feed material. Ignoring the small amount of activated chromite in the feed would have produced significant errors in the cleaner bank model predictions. The very small froth stability parameters that were regressed for both the rougher and cleaner froths could possibly infer that the fraction of floatable chromite is very floatable and if it gets a chance of reporting to the froth phase it will be recovered into the concentrate launder.

Table 3.8: Chromite model parameters derived from the regression of the mini plant data

PARAMETER	REGRESSED VALUES
$\phi_{act,Cr,feed}$	0.003
$\phi_{act,Cr,cleaner\ feed}$	0.095
ξ_{Cr}	16.41
δ_{Cr}	0.821
P_{Cr}	1.62e-4
$\beta_{Cr,rougher}$	0.00
$\beta_{Cr,cleaner}$	0.0042
α_{Cr}	1.0e-6

**Figure 3.22: Experimental size distribution for the mini plant feed and cleaner feed streams and the regressed feed entrainability curve for chromite**

A comparison of the measured and calculated chromite flow rates in figure 3.23-25 show that the chromite recovery model of equation 3.12 can adequately predict the flow rates of chromite for the different circuit configurations as well as for the froth characterisation tests in the rougher and cleaner banks. The froth characterisation test was conducted over the entire range of feasible operating conditions for these cells (as a function of froth height and volumetric air flow rate) indicating that this model well describes the recovery behaviour of the chromite particles. It should be noted that the entrainment model does not explicitly account for changes in operating conditions such as froth residence time as a function of froth stability. These factors would have to be addressed in the associated water recovery model, which would be required if any attempt was made to use the model derived in this chapter to perform predictive simulation. This is addressed in chapter 5.

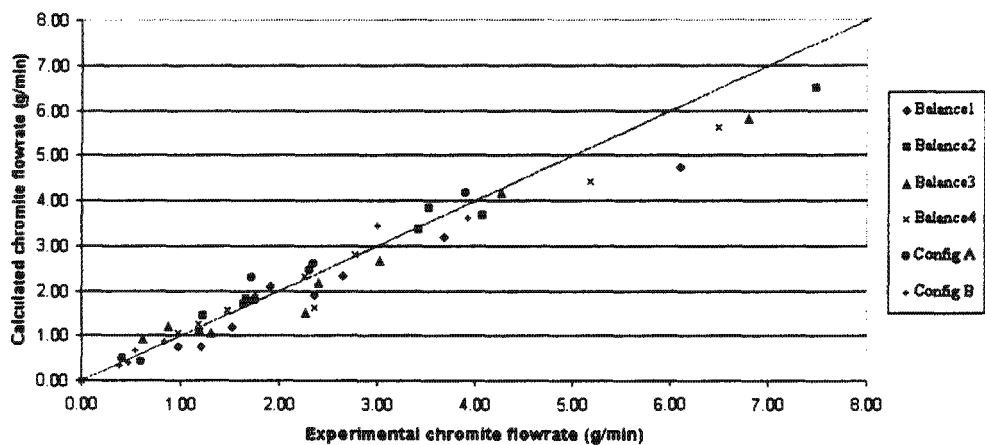


Figure 3.23: Calculated versus experimental chromite flow rates for the mini plant tests

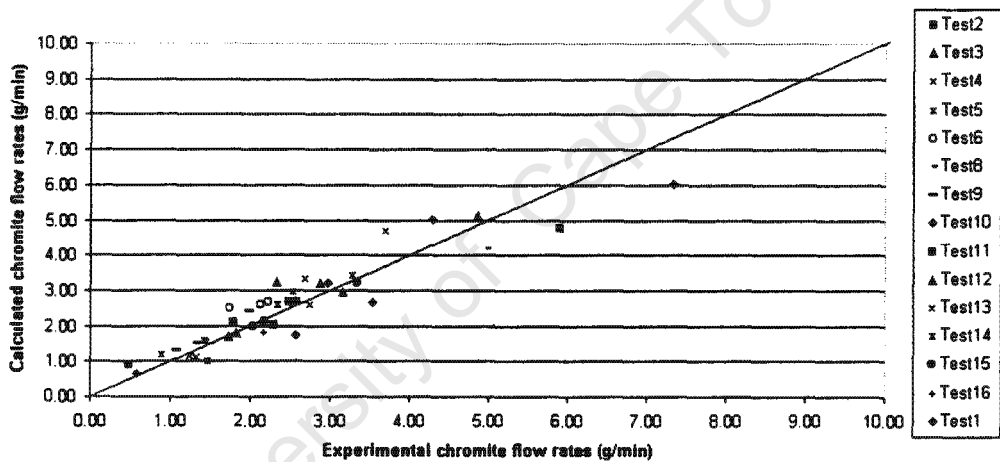


Figure 3.24: Calculated versus experimental chromite flow rates for the mini plant rougher froth characterisation test

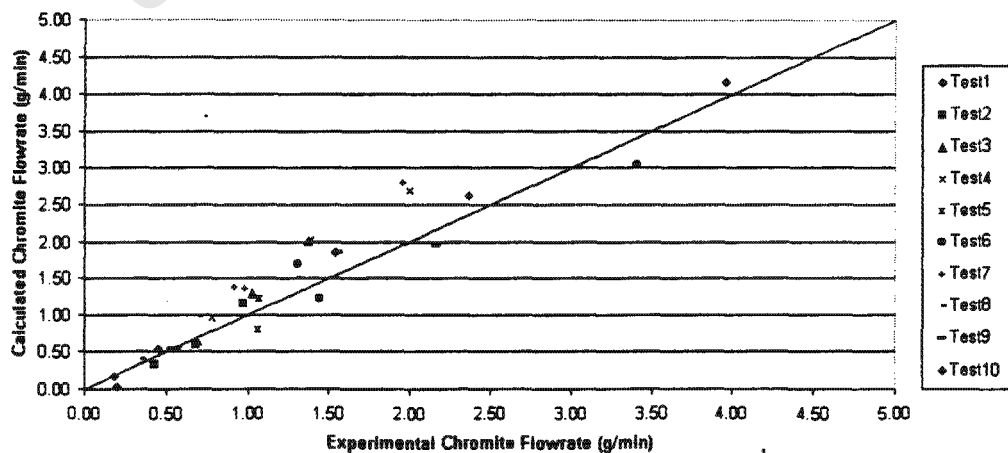


Figure 3.25: Calculated versus experimental chromite flow rates for the mini plant cleaner test

3.6 Model used to predict the gangue recovery on the mini plant

It is necessary to predict the gangue flow rates in a circuit in order to determine the grade of the final concentrate stream. For the purpose of modelling the mini plant, gangue is considered the total dry mass less the chromite and PGM carrying particles. The mechanism by which gangue is recovered into a flotation concentrate is more complex than the recovery mechanisms of the PGM and chromite particles discussed already. The gangue particles have a range of particle sizes but also have varying densities and other physical properties. Some gangue in PGM ores are naturally floatable, such as talc and pyroxene. A large portion of the gangue can be considered to be recovered by entrainment while an appreciable proportion is recovered as composite particles with floatable sulphide minerals in the ore. The only way to accurately model gangue behaviour is to use extensive information such as mineral by particle size by liberation data, and then to model the kinetics of each of these classes individually depending on whether the particles float or are entrained into the concentrate. Detailed information such as this is not always readily available and the mathematical complexity of modelling all these components individually around a circuit can be difficult. Obviously methods are required to approximate the kinetics of the gangue particles in a circuit using methods similar to the approaches taken to model the PGM and chromite flotation kinetics.

The flotation kinetics of the gangue particles were assumed to follow the same mechanism of recovery as the chromite particles discussed in section 3.5. The resulting parameters from the regression are depicted in table 3.9. The entrainability curve and the experimental gangue particle size distribution of the mini plant feed and cleaner feed streams are presented in figure 3.26. The model fit for the gangue flow rates of the mini plant data is presented in figures 3.27-3.29. A total of eight parameters were regressed from 240 experimental data points producing a model fit of 2.12. This was not considered satisfactory as would have been expected from the simplistic modelling approach used in this regression of the complex gangue material.

Table 3.9: Gangue model parameters derived from the regression of the mini plant data

PARAMETER	VALUE
$\phi_{act.Ga,feed}$	0.016
$\phi_{act.Ga,cleaner\ feed}$	0.85
ξ_{Ga}	22.91
δ_{Ga}	425
P_{Ga}	$7.3e-5$
$\beta_{Ga,rougher}$	0.00
$\beta_{Ga,cleaner}$	0.048
α_{Ga}	0.21

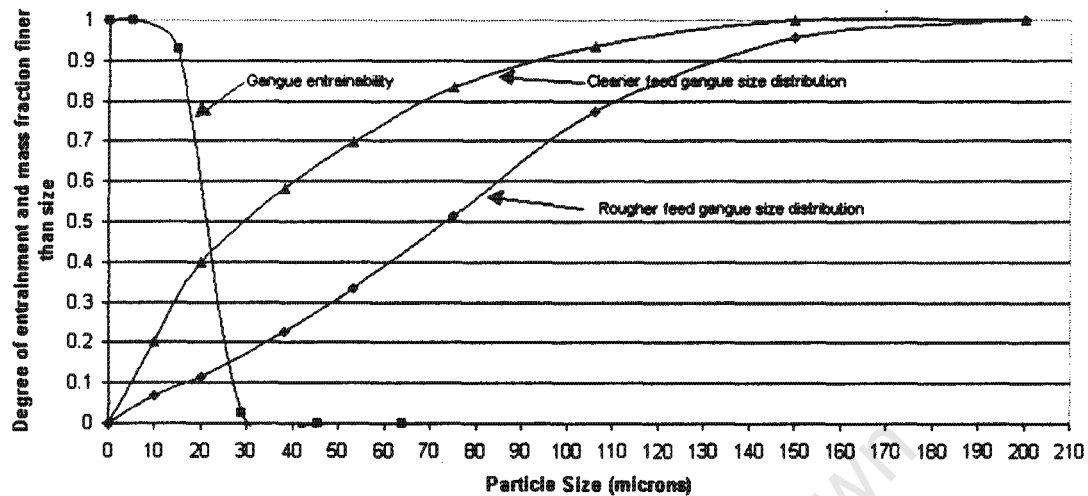


Figure 3.26: Experimental gangue particle size distribution for the mini plant feed and cleaner feed streams, as well as the regressed feed entrainability curve for gangue

A larger entrainment parameter was regressed for the gangue particles compared to the chromite. This makes sense as the average specific gravity of the gangue is approximately 3.2 while that of the chromite is approximately 4.2. The lower density increases the entrainment function across all size fractions for the gangue particles. This is evident by the larger area under the entrainment curve in figure 3.26. The fraction of floatable gangue material in the feed to the mini plant was regressed to be 1.6 percent, which seems to correspond well with the known sulphide content of the ore, (Liddell et al, 1986). The very small froth stability parameter that was regressed for the rougher froth could possibly infer that the rougher froth was very stable, restricting the amount of hydrophobic gangue (ie. recovered attached to air bubbles) returning to the pulp from the froth phase.

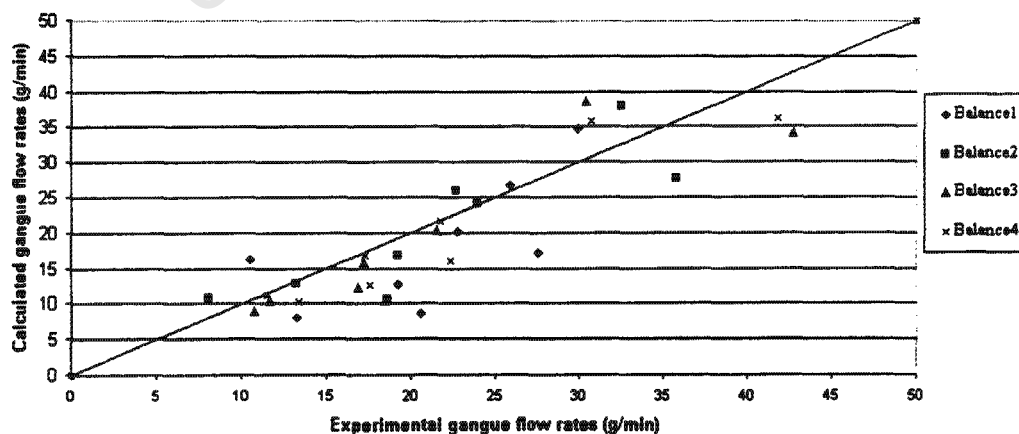


Figure 3.27: Calculated versus experimental gangue flow rates for the mini plant tests

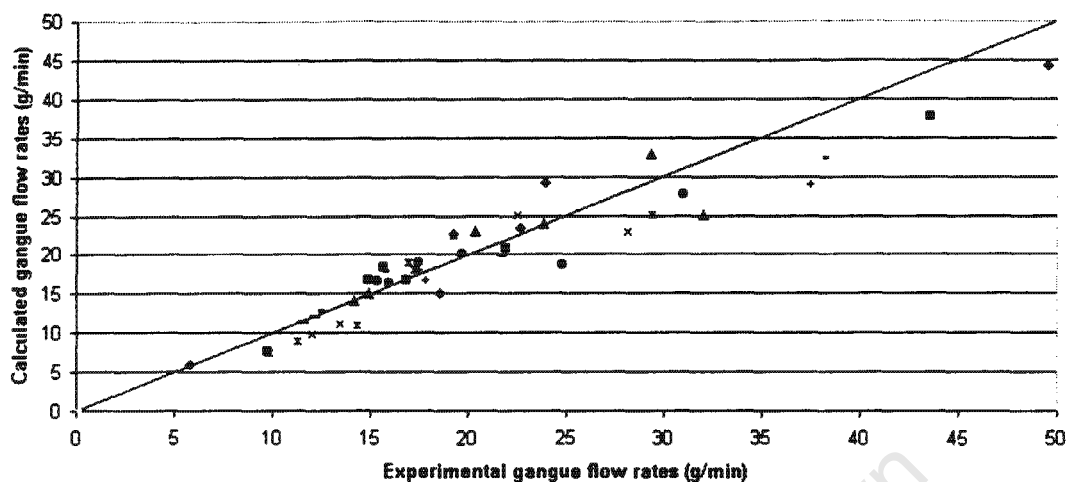


Figure 3.28: Calculated versus experimental gangue flow rates for the mini plant rougher froth characterisation tests

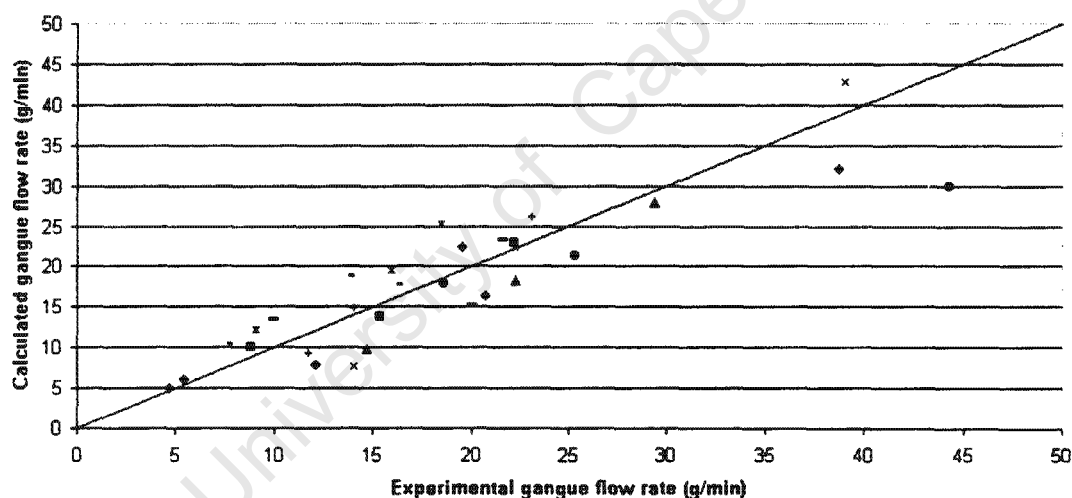


Figure 3.29: Calculated versus experimental gangue flow rates for the mini plant cleaner froth characterisation tests

3.7 Conclusions drawn from the preliminary model development

This chapter has presented the results of the preliminary model development on the Mintek mini plant facilities. There are a number of conclusions that can be drawn from this work and these can be summarised as follows:

- The conservation of the PGM floatability components around a stream mixing node as well as around the entire circuit have been verified using the nodal analysis

technique proposed by Runge et al (1997). It was important to test this assumption on this ore as it forms the basis of the entire simulation methodology. A three step algorithm has been presented whereby the parameters describing the PGM flotation kinetics can be effectively regressed. This comprises using only the batch floatability test initially as mass balance tools in order to get a first approximation of the feed stream floatability mass fractions. The next step is to use the derived mass fraction parameters from step one while using the plant mass balance data to regress the plant floatability and froth characterisation parameters. The third step is then to regress all the parameters together to obtain the minimum TSSE. A satisfactory model fit was obtained for the PGM flow rates from the feed and froth characterisation test work as well as for all the batch floatability data. Analysing the fit to the data indicated the need to include an entrainment contribution to the overall recovery of the PGM minerals as a result of the under prediction of the PGM flow rates in the last cells of each bank. This is addressed in Chapter 4. The Monte Carlo Boot Strap analysis on the regressed PGM parameters provided some insight into the sensitivity of the model parameters. The regression was shown to be the most sensitive to the froth stability parameters. This highlighted the need for a more comprehensive froth characterisation test. Chapter four presents the work on the pilot plant facilities where this problem is addressed.

- The entrainment equation developed by Savassi (1998) proved to be very effective in describing the degree of entrainment as a function of particle size for chromite, although it was found necessary to allow a small proportion of the chromite in the feed to be recovered by a true flotation mechanism. It can be concluded that realistic model parameters were regressed as a result of the verification between the regressed and the experimentally determined chromite feed particle size distributions.
- The preliminary model describing the recovery of gangue was found to be relatively poor in terms of the fit to the experimental data. This is owing to the complicated mechanism by which gangue is recovered into a flotation concentrate for this ore. Even though this was the case, the regression showed that a higher degree of entrainment (ξ_{Ga}) was determined for gangue compared to chromite. This was as expected as the gangue particles had a lower specific gravity compared to the chromite particles. The gangue particles have a range of particle sizes, densities and other varying physical properties influencing the entrainment of these particles with water. On this basis, a more comprehensive approach to modelling the gangue is presented in chapter 4.
- A few conclusions can be drawn based on the operation of the mini plant facilities at Mintek. The method of controlling the pulp level in the flotation cells was not satisfactory. A steel probe placed in the feed launder box in each bank was earthed every time the pulp level rose high enough to touch the base of the probe. This caused a pneumatic actuator to open the tailings valve on the bank. When the pulp dropped below the probe the valve closed again. This type of control system together with the fact that the dart valves on the tailings valves were too large

caused a pulsating tailings flow rate from all three banks in the circuit. This increased the standard deviations of all the flow rates in the plant. The method of measuring the volumetric air flow rates into each cell was tedious preventing many measurements to be made and also increased the level of disturbance on the plant. An inverted, buoyant measuring cylinder was used to measure the froth depths in all the flotation cells during the mini plant test work. This method was not accurate enough to develop a comprehensive froth characterisation model that can distinguish the effects of froth depth and volumetric air flow rate on the froth performance. A more sensitive instrument was needed to measure the froth depths. Chapter 4 describes the work that was done to upgrade the pilot plant facilities at Mintek in order to overcome these limitations of the mini plant test work. Invariably, the smaller the flotation cell, the more sensitive it gets to process disturbances. Larger flotation cells are more capable of damping out small process disturbances such as small variations in the feed grade, froth depths, air flow rates, pulp densities etc. It is for this reason that the larger pilot plant flotation facilities at Mintek were chosen to be upgraded for subsequent use in this study.

- It should be noted that this chapter related only to model fitting. No attempt was made to perform predictive simulation with the developed model. As such, no water model was required. This is also addressed in chapter 5.
- The mini plant test work highlighted the need for a detailed mineralogical analysis of the feed material as a means of evaluating the regressed PGM floatability distribution parameters. QEMSEM analysis (quantitative elemental measurement using a scanning electron microscope) was selected for this purpose on selected samples in subsequent work, combined with information on the mode of occurrence of the PGM minerals. This is also discussed in chapter 4.

Chapter 4

UPGRADED PILOT PLANT TESTWORK

This chapter describes the work done on the main pilot plant facilities at Mintek. Existing pilot plant facilities at Mintek were upgraded in order to make the necessary measurements to satisfactory model and test the proposed modelling methodology. The chapter starts by giving a description of the pilot plant facilities and then it discusses the work done in order to obtain accurate control and measurement of the froth height and volumetric air flow rates into each flotation cell. Steps taken to improve the control of the pilot plant are discussed. This is followed by a description of the different pilot plant campaigns that were carried out in order to model the flotation kinetics of the ore used in this phase of the work as well as to test some of the assumptions in the modelling methodology. The pilot plant campaigns discussed in this chapter are divided into two phases of work, each phase having its own aims and objectives.

Phase I work is essentially a continuation or reproducibility of the mini plant test work in chapter 3. The objectives of this work was to apply the same modelling methodology onto the pilot plant facilities with the new, improved data set generated on the pilot plant and to apply the different unit models based on the conclusions drawn in chapter 3 with the aim of refining the modelling procedure. This phase forms the bulk of the work of this chapter. The ore used during phase I consists of a sample of Impala UG2 ore. This ore had been standing at Mintek for a long time and had started to show signs of oxidation. This ore was chosen as it offered a wide distribution of floatabilities providing a larger fraction of slower floating components due to some oxidation on the surfaces of the minerals. This had an advantage over using an ore with a large fraction of fast floatable PGM particles as it provided more scope for the optimisation work presented in chapter 7. This chapter presents results of a detailed analysis of the feed material using the QEMSEM facilities at Mintek providing information on the mode of occurrence of the PGM minerals in this sample of ore.

Phase II work was carried out using a sample of Lonmin Karee UG2 ore. This ore was considered an extremely good ore having a large fraction of fast floating components. The aims of this work were primarily to tests the PGM regression technique. Feed material obtained from two distinctly different milling conditions were characterised using the method presented in chapter 3. The resulting parameter sets were analysed in comparison to known, measured differences in the feed material. This work provides some basis for extending the modelling methodology to include regrind stages in simulating more complicated circuit structures.

After presenting the details of the two phases of the pilot plant campaigns, the next section of this chapter presents the model fit to the experimental data and gives an interpretation of the pilot plant results. Included in this section is the fit of the pilot plant data to the PGM, chromite and gangue models from phase I work. An empirical bubble surface area flux model based on work developed by Gorain (1997) is applied to the

results obtained from the pilot plant and presented next. The chapter ends with conclusions drawn based on the development of the work presented in this chapter.

4.1 Description of the pilot plant facilities at Mintek

The pilot plant flotation cells at Mintek are commonly known as SALA flotation cells designed by Svedala (PTY) LTD. The rougher and scavenger banks each consisted of six eighty litre flotation cells joined together to form a bank. The cleaner cells consisted of four thirty litre flotation banks joined together to form a bank. All the cells were square and manufactured from stainless steel. A common motor was used for two cells on a bank. An Agitar agitation mechanism was used, driven by a 1.1 KW and 0.8 KW motor for the eighty litre and thirty litre cells respectively. The power drawn from the motors was not adjustable, nor was the agitation speed of the impellers. No froth scrapers were used to facilitate the movement of froth into the concentrate launders. The froth produced flowed over only one side of each cell, but the launders were redesigned so that individual concentrate flows from each cell could be sampled during the tests. Table 4.1 gives the measured dimensions of the individual flotation cells, as well as the agitation speeds of each individual cell^{4.1}.

Table 4.1: Dimensions and agitation speeds of the pilot plant flotation cells

CELL NAME	LENGTH (CM)	HEIGHT (CM)	WIDTH (CM)	R.P.M.
Rougher A	41.4	42.0	41.9	832
Rougher B	42.1	42.1	42.0	833
Rougher C	41.7	42.2	42.1	831
Rougher D	42.1	42.2	42.2	833
Rougher E	41.9	42.3	42.2	833
Scavenger A	41.9	41.7	42.3	838
Scavenger B	42.4	41.6	42.2	840
Scavenger C	41.4	41.6	42.1	840
Scavenger D	41.9	41.7	42.2	834
Scavenger E	42.0	42.0	42.2	833
HGC A	29.6	32.0	30.0	1106
HGC B	29.6	32.1	29.9	1105
HGC C	29.8	32.2	29.9	1105
LGC A	29.8	31.9	29.8	1112
LGC B	29.6	32.0	29.6	1109
LGC C	29.7	32.0	29.6	1113

^{4.1} HGC = high grade cleaner bank, LGC = low grade cleaner bank

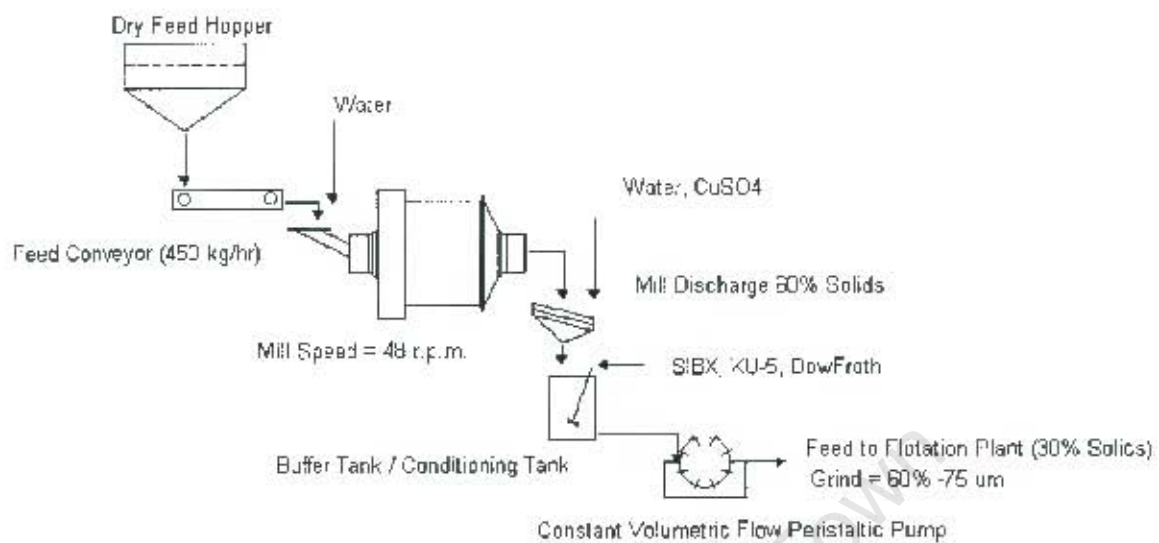


Figure 4.6: Flowsheet of the pilot plant feed preparation setup



Figure 4.7: Photograph showing the mill and mill feed assembly for the test work of phase I

between the alternative circuit configurations could be made as well as to determine if the regressed floatability distribution of the model could accurately predict the kinetics of the floatable particles under the different conditions tested.

4.4.2.1 Milling conditions used during phase I

An open circuit ball mill was used to grind the crushed -6 mm ore to an acceptable size distribution suitable for flotation. A grind of 60% -75 μm was chosen as this was considered an appropriate grind achievable by the mill which was also appropriate for flotation, liberating most of the base metal sulphides in the sample. It was decided to use an open ball milling circuit which was easier to operate and control compared to a closed mill/cyclone network. It must be stressed that the intention of the test work was not to produce optimum feed conditions for flotation, but rather a constant feed composition that was easy to maintain.

Three ball sizes, 50, 30 and 20 mm in diameter were used in the mill. Chrome-steel balls were used, as is typically the case in the industrial treatment of this type of ore. Characteristics pertaining to the mill and its operation are presented in table 4.2 below.

Table 4.2: Mill ball charge distribution

Total Charge Volume (%)		30
Average Specific Gravity of Chrome-Steel Balls		7.75
Percentage Solids in Mill (%)		60
Mill Length (mm)		1240
Mill Diameter (mm)		770
Ball Diameter	Number Balls	Total Mass Balls
(mm)		(kg)
30	1684	183.6
40	1232	326.4
50	986	510.0
Total Mass		1020

A milling curve was constructed to determine the mill speed required to obtain a feed grind of 60 % -75 μm at 60% solids in the mill at a feed rate of 450 kg/hr. This was calculated to be 48 rpm which was approximately 75% of the mill's critical speed. This dry mass feed flow rate was chosen as it allowed a residence time across the rougher and scavenger bank of approximately 30 minutes (based on tailings flow rates) which was typical of conditions employed in the treatment of this type of ore in industry. Figure 4.6 gives a schematic representation of the flotation feed preparation setup used during the pilot plant campaigns in phase I. Figure 4.7 is a photograph of the mill used during the test work of phase I.



Figure 4.5: Photograph showing the position of the digilevel in a flotation cell on the pilot plant

4.4 Pilot plant operating conditions and experimental technique

This section describes the operating conditions and experimental techniques used in obtaining the data for the pilot plant campaigns. The milling conditions and preparation of the feed to the pilot plant is also discussed.

4.4.1 Experimental techniques used during the pilot plant campaigns

The run-of-mine sample preparation was identical to the method described in section 3.3.1 in chapter 3 for both ores used during phase I and II in this chapter. The method of preparing the reagents, measuring bubble sizes, sampling streams, preparing samples as well as the batch floatability techniques were all identical to the procedures described previously in chapter 3.

4.4.2 The preparation of feed to the pilot plant

This section describes the feed preparation for the work in phase I. Phase II feed preparation conditions were different and these are discussed later in the section describing this work. Exactly the same procedure of producing feed pulp to the pilot plant was followed for every test during phase I. This was so that a direct comparison

The tailings flow rate is controlled using a pneumatic controller that receives an electronic control action from Floatstar which is converted into a pneumatic signal (via the I/P transducer) that opens or closes a dart valve at the discharge end of the flotation bank. The existing dart valves had to be re-designed as they were very flat. A small movement of the dart created a huge change in the tailings discharge flow rate, this, converted into an electronic measurement meant that the controller could only control the tailings flow rate over a 4 mA range (0 mA = closed, 4 mA = fully open). The normal control signal works over 12 mA range indicating that much better control could be achieved. New dart valves were designed that had a much finer taper, allowing a larger linear control action to be made resulting in a smaller change in the discharge tailings flow rate. This worked well, giving the Floatstar controller much better control of the level set points as well as producing a continuous tailings discharge from each bank. This was much superior than the pulsating flow rates of the previous level control setup described in chapter 3.

A Digilevel measurement sensor, designed by Mintek, was used to measure the pulp levels in the flotation banks. The digilevel-2 consists of a 1.3 metre UPVC pipe (50 millimetres diameter) with 96 measuring sensors mounted 5 millimetres apart occupying the bottom 0.5 metres of the probe. All the electronics reside inside the probe. A 10 metre length of 15 core 0.5 millimetre shielded cable exits the top of the probe via a gland and provides the plant interface with the level measurement signal. The digilevel measures the resistances ($1/\text{conductivity}$) across each sensor pair. From this it is able to calculate the resistivity profile from the bottom of the probe, through the pulp froth interface to the top of the froth. A computer program controlling the digilevel is able to calculate the position of the pulp/froth interface as well as the froth/air interface. The Floatstar controller maintains the level of the pulp/froth interface, while the depth of the froth phase can be recorded. Interesting information can also be obtained from the resistivity profiles obtained from the digilevel. The mixing characteristics within the pulp zone can be diagnosed (a linear profile below the pulp froth interface indicates good mixing of the pulp, while an uneven or sloped profile would indicate zones of higher solids concentrations in the pulp). The slope of the profile through the froth zone can also give information on the concentration profiles within the froth zone. The profiles can also be logged to a historical trend file that can be accessed after the pilot plant campaign is finished. This work was the first attempt to use the Mintek designed digilevels and they were shown to be very effective.

A digilevel probe was placed at the head of every flotation bank. This allowed the controller to operate with a feed-forward control response ie: if a disturbance was measured at the start of the bank, the control valve was adjusted before the disturbance was propagated down the bank. Figure 4.5 shows the position of the digilevel in the rougher bank of the pilot plant. During the pilot plant campaigns, the first cell to each bank was used purely for level control (ie. no flotation occurred in this cell) to prevent the froth from building up around the digilevel which could disturb the measurement. This also served to buffer out possible flow rate disturbances from the pumps feeding each bank of flotation cells in the pilot plant.

4.3 Floatstar level control

The preliminary work conducted on the mini plant highlighted the importance of accurate control of the pulp level in each flotation cell. A commercial level control system designed by Mintek, known as Floatstar, was installed onto the pilot plant in order to control the flow rates of pulp throughout the circuit. Floatstar controls the tailings flow rates (therefore pulp levels) of the whole circuit and not only an individual flotation cell. It uses a matrix method to gain information on the circuit configuration as well as information on the flotation bank cross sectional areas to calculate the volume hold-ups in various parts of the circuit. The controller automatically tunes itself in order to obtain the optimum gain and integral time controller parameters for every bank in the circuit. It does this by making step changes to one bank at a time, keeping all the other banks in the circuit constant and determines how the disturbance affects the level set points in other banks of the circuit. The result is that if a disturbance is made somewhere in the circuit, the controller automatically corrects for the disturbances made to the other banks in the circuit, dissipating the disturbance very rapidly allowing for a more controlled plant. Figure 4.4 below is a schematic representation of the Floatstar control setup. The dashed lines indicate the digital signal sent to the Floatstar controller from the level measurement device (discussed below). The solid lines show the control action signals sent to the I/P (current to pressure) transducers which convert the electronic signal into a pneumatic action for opening and closing the tailings control valves.

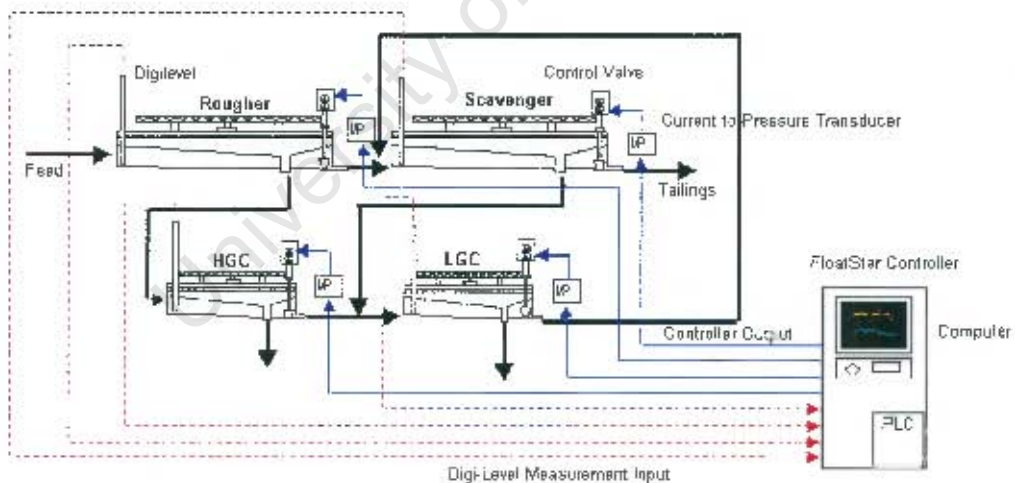


Figure 4.4: Floatstar control on the pilot plant

In addition to controlling the levels for the whole circuit, a historical trend database is created during the pilot plant run that records the actual levels in every bank, the level set points of every bank as well as the controller action sent to the level control probes. This data can be used to assess how well the circuit was operating during the campaign and can also quantify the amount of disturbance created while taking samples from the plant.

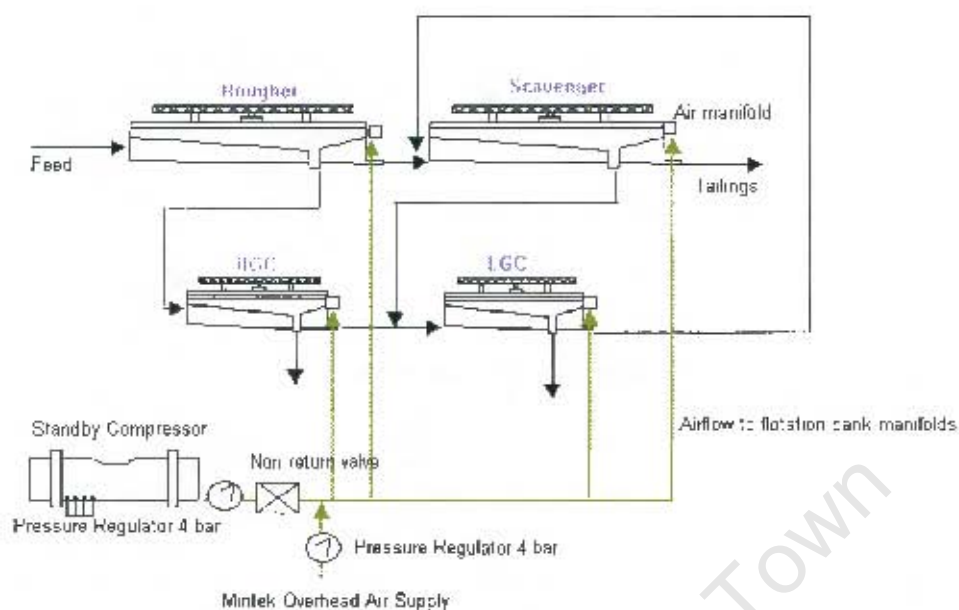


Figure 4.3: Feed air supply setup to the pilot plant

Compressed air generated from Mintek's main compressor was used as the primary source of air. The supply from this source was prone to fluctuation and sometimes varied from a maximum pressure of 6 bar to as low as 4 bar (S.T.P.). This problem was overcome by installing a pressure regulator at the head of the air supply and setting it to 4 bar. In addition, a standby compressor was linked to the supply source with a controller that started the compressor if the pressure feeding the pilot plant dropped below 4 bars ensuring that the air supply was always constant.

The air measurement manifold was designed so that the pipe feeding the rotameter was not at ninety degrees to the flow rate of air in the manifold in order to prevent a pressure swing from occurring within the manifold. The pressure regulators on the flotation cells were all set to 1.5 bar pressure. This was the minimum air pressure required to overcome the static head exerted by the pulp above the impeller stator assembly. This low pressure was favorable as it reduced the possibility of a change in air supply with varying overhead pressures as it was set much lower than the feed pressure supply of 4 bars. In addition, a lower pressure eliminated the possibility of disturbing the bubble breakup mechanism at the impeller stator. If the pressure was set too high, by-passing of air around the impeller could result in larger than desired bubble sizes in the flotation cells.



Figure 4.1: The Mintek flotation pilot plant at Eastern Platinum Mines

Figure 4.1 is a picture of the Mintek flotation pilot plant facilities. All the concentrate and tailings streams were pumped around the plant via high speed SALA pumps.

4.2 Air flow rate measurements and control

The original pilot plant flotation cells used an induced air mechanism to aerate the cells with no direct method of measuring the quantity of air induced by the agitator. The flow rate of air into the cells were controlled using a manually adjustable gate valve, which was difficult to control as the amount of air induced was extremely sensitive to the valve setting. An accurate and easy method of controlling and measuring the volumetric air input to the flotation cells was required. Figure 4.2 shows the air supply and measurement mechanism designed for the upgraded pilot plant facilities, showing the assembly of the valve connection to each flotation cell. Compressed air in the manifold flowed into an independent air pressure regulator, a pneumatic air needle valve followed by a 2-way T piece ball valve which could either direct the air flow into the flotation cell or otherwise into the air measurement manifold. The pneumatic needle valve allowed a much finer control of the air input to the flotation cell compared with the gate valves of the previous design. A pressure regulator was installed onto each cells air inlet to eliminate any pressure fluctuation resulting from air flow rate adjustments made on other cells in the circuit. The purpose of the ball valve is to redirect the air flow from a flotation cell to the metering manifold which feeds the air to a rotameter for measuring the volumetric air flow rates of all the flotation cells on the bank. The rotameters were sized to measure mid-range for an average concentrate flow so that

the entire spectrum of feasible air flow rates to the flotation cells could be measured during the froth characterisation work. This design was much less expensive than installing individual rotameters onto each cell across the bank, but it had a disadvantage that the air input to the cell had to be interrupted while measuring the flow rate. Fortunately the measurement only took a few seconds and was not considered to disturb the plant to any extent while taking air flow rate measurements. The air flow rates were measured after the streams were sampled during all the tests and therefore the interrupted air supply had no effect on the samples taken from the plant.

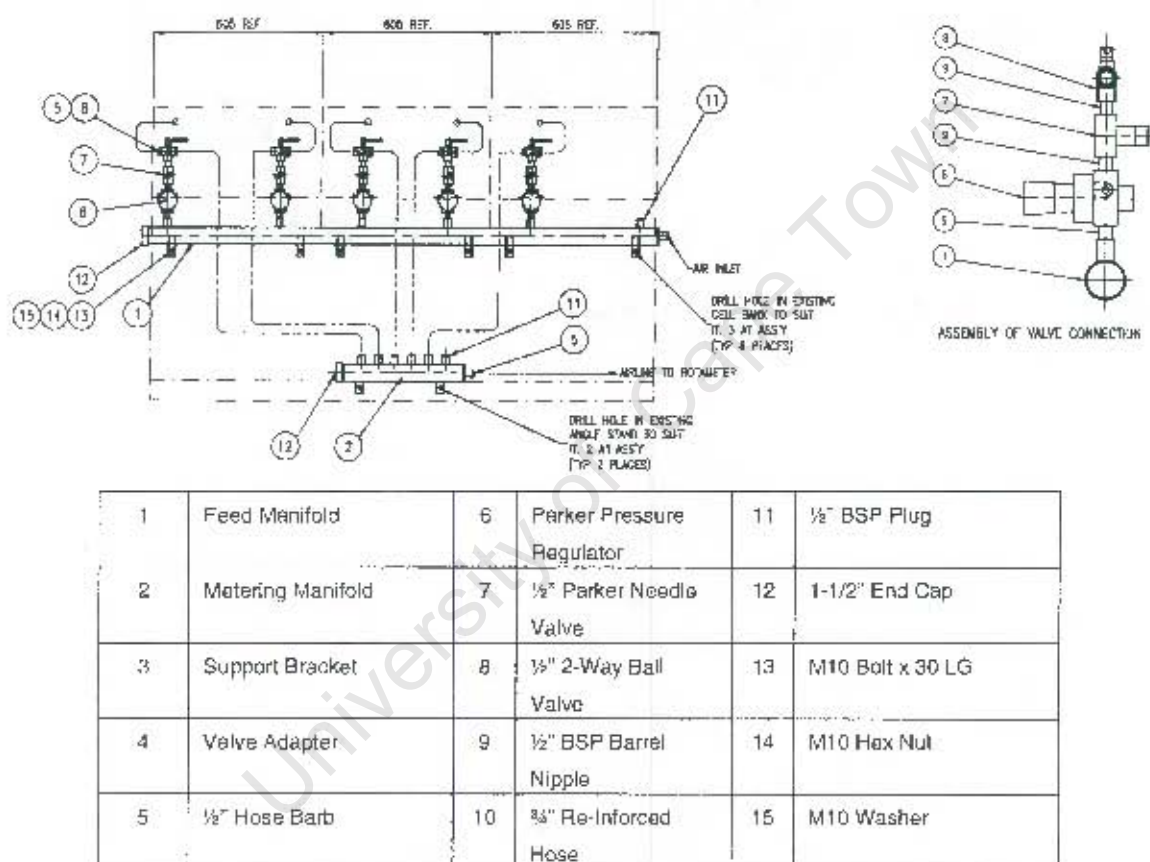


Figure 4.2: Schematic showing the design of the air supply and measurement mechanism on the flotation banks

A compressed air supply had to be used because the small air needle valve (valve 7 figure 4.2) on the flotation cells restricted the induced air flow rate considerably. Figure 4.3 shows the air supply setup feeding the flotation banks.

Ore was introduced from the feed hopper onto a variable speed conveyor belt that fed directly into the mill feed port. Water was added into the head of the mill, controlled by rotameter (which was recorded) to produce a mill product of sixty percent solids. The percentage solids was measured and maintained using a marcy scale, calibrated to a feed with a specific gravity of 3.7 corresponding to the average density of the ore. The flow rate of dry ore from the conveyor into the mill was measured by cutting the entire stream for a duration of fifteen seconds with a sampling bucket. The mass collected was measured and converted to a flow rate. The collected sample was returned to the mill after the measurement. The dry mass flow rate was measured every thirty minutes during a test to ensure it remained at its set point. The discharge from the mill flowed onto a vibrating 1mm screen to remove unwanted wood chips as well as to remove the occasional large particle to prevent pipe and pump blockages. Wash water was sprayed over the screen decreasing the percent solids to thirty percent. Copper sulphate activator was added on top of the screen, ahead of the other reagents in order for the activation of the sulphide particles to take place before collector addition. The screen undersize stream was pumped into a 200 litre, well mixed buffer tank. The percent solids of the screen undersize stream was also measured every 30 minutes during all the tests. A sample of the mill discharge stream was taken every hour during the tests to determine whether the grind was still being maintained. The buffer tank was used to buffer out any disturbances from the mill, and also served as a conditioning tank providing approximately 12 minutes of conditioning time. A constant volumetric flow rate peristaltic pump (bredell pump) pumped the pulp from the buffer tank to the first flotation cell on the rougher bank at 18 l/min. The peristaltic pump proved to be very effective in eliminating any disturbances from the mill and a constant feed flow rate was maintained to the pilot plant.

4.4.2.2 Reagent addition rates used during phase I

The reagents used for the tests were those suggested by Mintek and were based on the standard reagent suite for UG2 ores of this type. All the reagents were pumped via peristaltic pumps from pre-prepared samples stored away from the plant to eliminate contamination of the reagent by pulp. The reagents were all freshly prepared at the start of campaign to ensure no degregation with time. The reagents were mixed to a concentration of 1 % by mass with water. Copper sulphate was added first at an addition rate of 20 g/t. This was used as an activator to enhance the attachment of the collector onto the surface of the sulphide minerals. A xanthate collector SIBX (sodium isobutyl xanthate) was added at 150 g/t. Dowfroth 200 was used as a frother, KU-5 (a carboxymethyl cellulose) was added as a depressant at a rate of 40g/t and 100 g/t respectively. The selected reagent suite and flow rates were those used commonly by Mintek for the treatment of this type of ore.

4.5 Pilot Plant Campaigns

This section describes the details and objectives of every test carried out on the pilot plant facilities. The section is divided into two, the first section describes the work of phase I using an Impala UG2 ore. The second section describes the test work using the Lonmin Karee UG2 ore.

4.5.1 Phase I tests on the Mintek pilot plant

4.5.1.1 Feed characterisation test (Test1)

The main objective of this test was to obtain enough information about the kinetics of the floated PGM particles in order to characterise the floatability distribution of the feed to the pilot plant. Enough information had to be obtained in order to regress all the parameters necessary to model the kinetics of water, PGM, chromite and gangue. Figure 4.8 shows the circuit configuration chosen for this test, indicating the streams that were subjected to batch floatability tests.

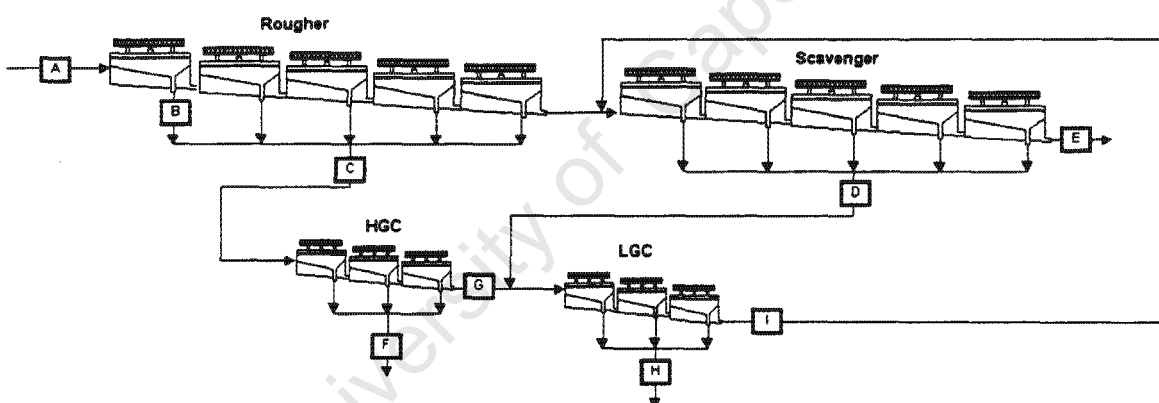


Figure 4.8: Flowsheet of the feed characterisation test indicating streams sampled for batch floatability tests

The circuit configuration in figure 4.8 is the standard “Mintek” circuit design used to treat most UG2 ores. Only single stage cleaning is employed however. The circuit produces two separate high and low grade concentrates from the two cleaner banks. This circuit was chosen as it is relatively simple, with only one recycle stream affecting only the low grade scavenger and low grade cleaner section of the plant. Both the rougher and high grade cleaner were essentially run in open circuit. This had the advantage that it was easy to maintain at steady state, but still complicated enough to test some of the assumptions made in the flotation kinetic model. This circuit was chosen as it uses the

same concept of concentrating the fast and slow floating particles separately as done by most of the industrial plants treating UG2 ores.

The methodology employed for this test is detailed below:

- All the streams in the plant were sampled in order to characterise the flow rates with respect to volume, mass, mineral and gangue content. The pilot plant was initially allowed three hours to reach steady state, keeping the feed composition as well as the operating variables within each cell (froth height and air flow rates) constant. Three hours ensured all the flotation banks were replaced with pulp three times as the summation of every cells residence time in the circuit was approximately 55 minutes. This also allowed approximately six complete cycles in the recycle stream, ensuring this stream had approached steady state. After steady state had been reached, samples were collected from all the streams every hour for four hours. From this, four independent mass balances could be obtained. This information was used to obtain standard deviations on all the flow rates with respect to volume, mass, mineral and gangue content.
- After every mass balance, all the flotation cell operating parameters were recorded. This included the froth depth and volumetric air flow rates. In addition, the air pressures into every cell as well as the regulator supplying the entire plant was recorded. A sample of the mill discharge was also taken to determine the grind of the feed during every mass balance.
- Samples of the plant feed, tailings and combined concentrates from each bank were taken for batch kinetic rate float tests. This data was used to obtain information on the distribution of the PGM fractions around the plant. This data allowed all the independent fractions to be linked to the feed floatability distribution to the plant. The samples were taken over the entire days run while recording the time that the sample was taken. Reproducibility tests were performed on the feed and final tailings streams. Combined concentrates from each bank were taken during this test as opposed to every stream to limit the amount of disturbance to the plant as individual concentrate streams flowed relatively slowly and a long time would have been required to obtain enough sample to conduct such tests.
- A representative, 1.5 kg composite sample of the feed stream, taken from all four of the plant mass balance samples was taken to determine the complete particle size distribution. The sample was wet screened to remove the $-38\mu\text{m}$ particles. The $+38\mu\text{m}$ fraction was dry screened and split into the following fractions, $+212\mu\text{m}$, $-212\mu\text{m}+150\mu\text{m}$, $-150\mu\text{m}+106\mu\text{m}$, $-106\mu\text{m}+75\mu\text{m}$, $-75\mu\text{m}+53\mu\text{m}$, $-53\mu\text{m}+38\mu\text{m}$, $-38\mu\text{m}$. The remaining $-38\mu\text{m}$ sample was further split into $-10\mu\text{m}$, $+10\mu\text{m}-25\mu\text{m}$, $+25\mu\text{m}-38\mu\text{m}$ size fractions using an ultrasonic bath and specially designed cloths to separate the fine fractions. The $-38\mu\text{m}$ fractions were not vacuum filtered, but allowed to dry by evaporation to avoid losing the small particles through the filter paper. These samples were then analysed for total chromite and PGM content.

- All the flotation cells in the circuit were characterised with respect to the gas dispersion in the cells as a function of the operating conditions employed (chemistry, gas rate, impeller speed, froth height). This was done using the UCT bubble size analyser. Four measurements were taken in each cell.

Table 4.3 is an averaged mass balance calculated from the four sets of samples for the feed characterisation test. Included are the standard deviations of every data point based on the four individual mass balances. Table 4.4 shows the measured cell operating variables including the measured standard deviations from the feed characterisation test.

Table 4.3 Averaged mass balance for the pilot plant feed characterisation test (test1)

	PGM mg/hr	Stdev	Chromite g/min	Stdev	Gangue g/min	Stdev	Water ml/sec	Stdev	Dry Mass kg/hr	Stdev	Vol Flow l/min	Stdev
Feed	1551.37	44.48	1345.06	66.22	5253.34	362.96	271.82	13.66	401.49	24.85	18.17	0.92
RA	618.04	78.12	10.26	2.19	146.51	12.17	13.67	4.49	11.63	0.89	0.87	0.27
RB	268.39	11.72	10.34	2.88	120.53	16.19	15.88	5.35	8.82	1.17	0.98	0.33
RC	80.13	19.40	7.34	1.73	73.83	7.60	13.91	2.58	5.16	0.52	0.86	0.16
RD	61.98	4.57	7.54	0.85	64.79	3.35	13.29	1.51	4.56	0.24	0.82	0.09
RE	48.61	8.58	7.37	0.73	56.71	7.05	13.62	1.90	4.02	0.49	0.84	0.12
RT	345.77	43.23	1276.22	93.32	4776.72	389.73	196.28	25.73	364.42	28.42	13.46	1.66
SA	66.59	11.82	9.86	3.18	73.67	16.67	21.97	6.97	5.25	1.23	1.34	0.42
SB	59.59	8.89	12.18	2.70	81.47	17.98	26.51	6.99	5.83	1.26	1.62	0.43
SC	31.11	0.42	7.09	0.94	47.02	4.94	15.42	2.70	3.36	0.35	0.94	0.16
SD	37.76	1.63	12.02	2.83	74.91	11.67	25.35	5.20	5.35	0.85	1.55	0.32
SE	37.72	8.31	14.68	4.34	66.55	20.67	29.14	7.78	6.21	1.53	1.78	0.47
ST	311.96	22.08	1319.42	103.37	4863.20	390.53	223.17	18.94	372.08	28.95	15.11	1.25
HGCA	585.60	11.32	4.34	0.58	86.59	3.90	3.93	0.17	7.56	0.23	0.27	0.01
HGCB	290.30	26.10	3.80	0.33	67.46	5.25	7.11	0.51	5.32	0.40	0.45	0.03
HGCC	181.19	9.37	4.53	0.39	70.67	2.40	10.75	0.45	5.09	0.15	0.67	0.03
HGCT	152.24	25.53	31.10	4.78	236.69	26.98	50.53	10.97	16.61	1.97	3.11	0.67
LGCA	95.76	12.98	4.20	0.52	49.92	3.81	10.98	0.91	3.59	0.24	0.68	0.05
LGCB	59.59	5.54	3.13	0.31	35.96	3.30	8.67	1.25	2.56	0.23	0.53	0.08
LGCC	55.72	6.36	3.90	0.29	42.38	2.31	12.80	1.20	2.98	0.16	0.78	0.07
LGCT	201.12	44.97	79.86	14.15	482.48	84.24	142.27	36.20	34.46	6.04	8.70	2.20

Table 4.4: Feed characterisation test operating set points. (test1)^{4.2}

	FH cm	StDev cm	Air Flow l/min	StDev l/min	db mm	StDev mm
Feed						
RA	7.13	0.29	47.69	1.77	0.89	0.038
RB	7.55	0.29	48.45	0.85	1.05	0.095
RC	7.93	0.25	53.70	2.97	1.22	0.048
RD	8.20	0.29	52.79	3.30	1.23	0.048
RE	8.43	0.51	57.68	3.01	1.18	0.019
RT						
SA	8.13	0.76	33.66	0.00	1.52	0.137
SB	8.10	0.83	31.26	0.35	1.14	0.059
SC	8.20	0.97	30.29	0.24	1.10	0.013
SD	8.00	1.17	30.86	0.20	1.00	0.017
SE	7.85	0.98	35.45	0.31	0.95	0.052
ST						
HGCA	7.20	0.81	18.56	5.99	0.59	0.048
HGCB	6.75	0.87	16.50	3.72	0.84	0.028
HGCC	6.75	1.00	21.18	2.62	0.65	0.064
HGCT						
LGCA	7.05	1.51	23.41	5.49	1.06	0.064
LGCB	8.00	0.58	19.53	3.54	0.63	0.095
LGCC	7.83	0.61	23.63	2.81	0.78	0.034
LGCT						

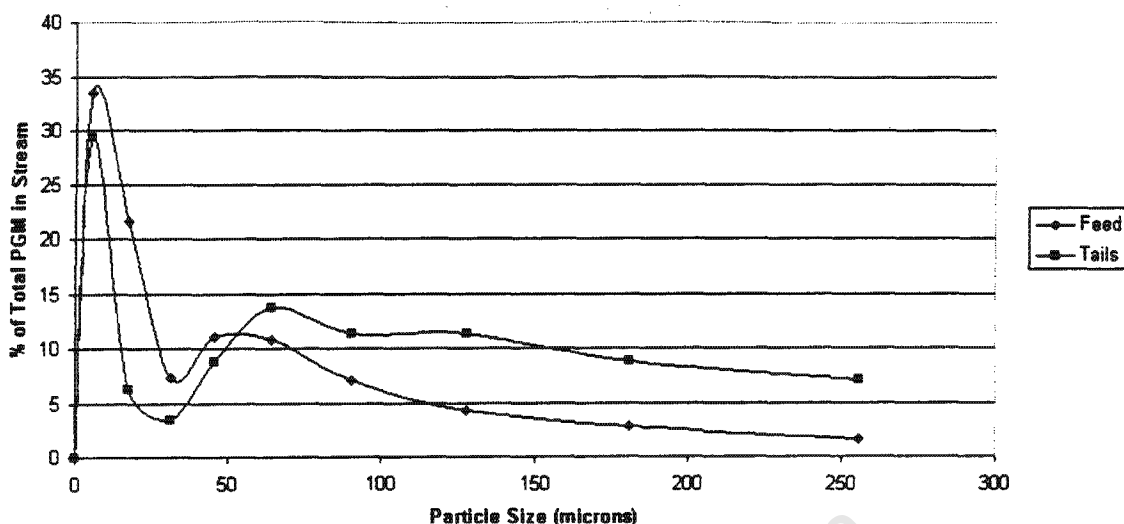


Figure 4.9: The distribution of total PGM mass in the feed and scavenger tailing streams of the pilot plant feed characterisation test

Figure 4.9 is included to give an indication of the total PGM mass fractions that were recovered from the pilot plant. A bi-model distribution of PGM particles occur with particle size showing a large fraction of $\sim 10 \mu\text{m}$ PGM particles in the feed. This is a very common occurrence in UG2 ores, (Liddell et al, 1986). Figure 4.9 shows that a large proportion of the ultrafine PGM's were not recovered. It can be hypothesised that these, probably fully liberated particles, have very slow flotation rates due to insufficient inertial forces to become attached to bubbles. This is supported by the need, identified on many industrial UG2 plants, for high pulp mixing energy requirements to achieve high recoveries. The high surface area of these particles suggest that they are probably also prone to oxidation.

Figure 4.10 shows the stability of the circuit during the feed characterisation test. These plots show the effect of the recycling streams on the stability of the scavenger and low grade cleaner banks. The rougher and high grade cleaner bank levels remained at their setpoints throughout the test due to the fact that they were run in open circuit. The plots also indicate the time that the samples were taken during the campaign showing that sampling the plant caused very little disturbance on the level setpoints. The control action (lower line) shows the position of the dart valve (0 = fully open, 100 = fully closed) it can be seen that the dart valves were open throughout the duration of the campaign showing the continuous tailings discharge flow rate as opposed to the pulsing flow rate of the mini plant. It is believed that the good control that could be achieved on the redesigned pilot plant contributed very significantly to the ability to obtain reliable and meaningful results amenable to the derivation of a process model.

^{4.2} FH = froth height, StDev = standard deviation, d_b = sauter mean bubble diameter

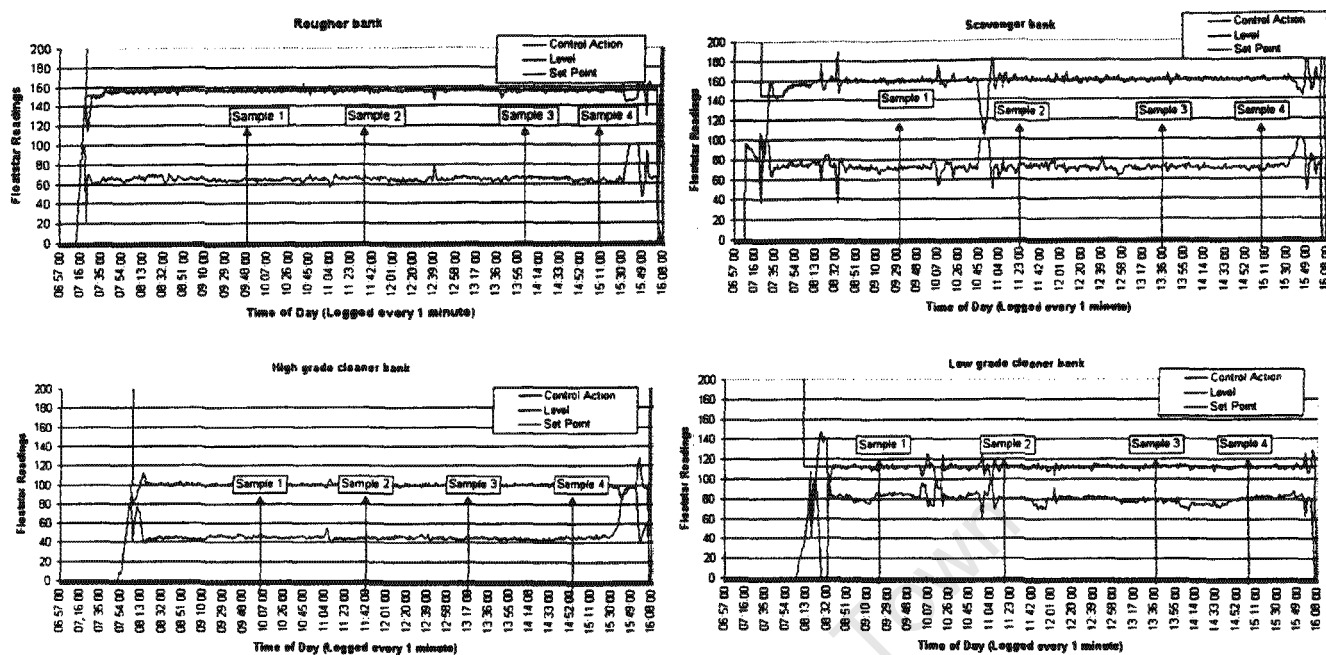


Figure 4.10: Figures showing the floatstar level control for the feed characterisation test

The batch test recovery time profiles obtained during the feed characterisation test (see figure 4.8) are presented in figure 4.11. The high kinetics of the concentrate streams relative to the feed and tails streams is once again very apparent.

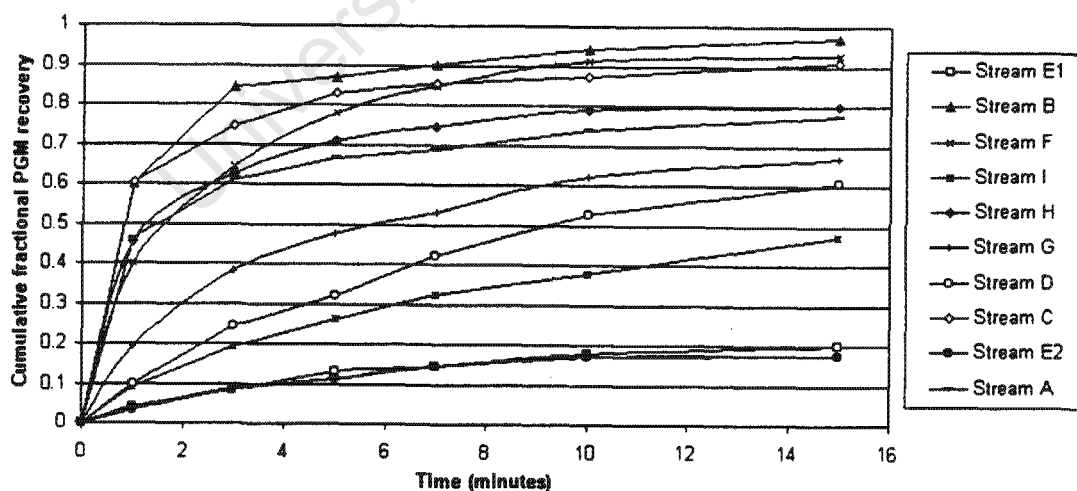


Figure 4.11: The batch PGM recovery time profiles from the feed characterisation test on the pilot plant

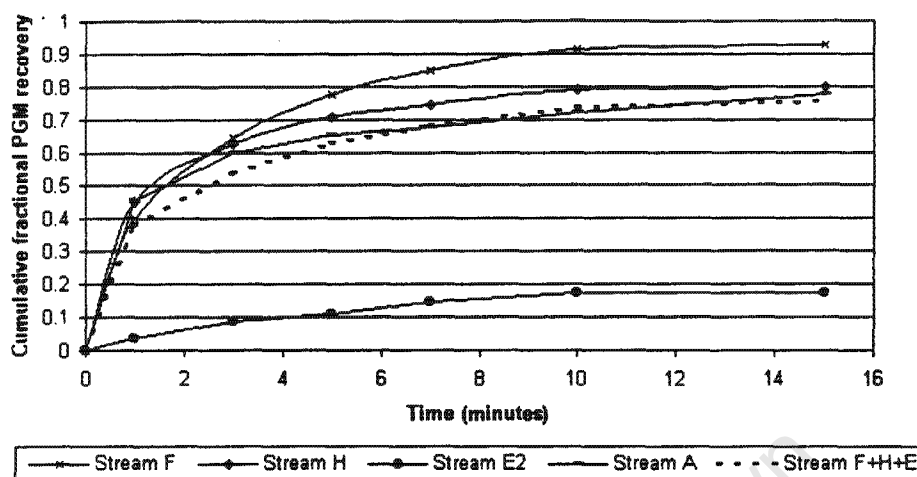


Figure 4.12: Predicting the feed stream recovery time profile from the final product streams of the pilot plant showing the conservation of PGM floatability components over the whole circuit

The data in figure 4.11 was subjected to the nodal analysis procedure described in section 2.2.5.2.4. In figure 4.12 it can be seen that good correspondence was obtained between the calculated and experimental results for the entire plant node. The results obtained for the other nodes in the circuit was equally good. This indicates that the floatability of the PGM minerals was not changing with time in the pilot plant.

4.5.1.2 Results of QEMSEM analysis of phase I feed material

A representative, 1.5 kg composite sample of the feed stream, taken from all four of the plant mass balance samples was taken and subjected to mineralogical analysis using QEMSEM. Four resin embedded polished sections of each fraction were prepared to determine the mineralogical composition and mode of occurrence of the PGM minerals and their associated minerals. These tests were done by the Mineralogy Department at Mintek. Optical microscopy and scanning electron microscope (SEM) based image analysis techniques were used. The results of this work is summarised in the tables and graphs below with a brief summary at the end discussing the findings of this work. The measured raw data from the QEMSEM search is presented in appendix F.

The mass distribution, total platinum-group metals (PGM) chemical analysis (obtained from the samples sent for analysis in section 4.5.1.1) and the grade derived from the PGM-search are shown in table 4.5. Fifty percent of the PGM's in the sample occurred in the $-25\mu\text{m}$ size fractions. In table 4.5 it can be seen that the grade determined from the mineralogical analysis was higher than the chemical assay in most cases. In these fractions, PGM grains larger than $10\mu\text{m}$ in diameter were detected which contributed approximately 50 percent of the total PGM volume in the fraction (table 4.6). The

QEMSEM search is based on the analysis of a very small sample of the entire feed stream to the pilot plant. The poor correspondence of the assay and mineralogical analysis is a consequence of the relatively small sample used during the QEMSEM search. Larger samples could have been used to obtain a more representative sample but this would have been extremely time consuming and possibly without any more accuracy than that obtained in table 4.5. It is emphasised however that this mineralogical data is still useful in terms of quantifying the extent of liberation of the PGM particles as well as assisting in giving an indication as to whether the regressed PGM floatability distribution of the feed sample has any physical significance. This is addressed in chapter 6.

Table 4.5: Total PGM in the size fractions of the feed to the pilot plant, (test1)

SIZE FRACTION (μm)	MASS (%)	TOTAL PGM	
		By chemical analysis (g/t)	By mineralogical analysis (g/t)
-10 μm	16.4	7.7	6.7
+10-25 μm	9.5	8.5	17.9
+25-38 μm	5.4	5.2	2.4
+38-53 μm	11.9	3.5	2.3
+53-75 μm	15.1	2.7	4.2
+75-106 μm	14.1	1.9	1.9
+106-150 μm	11.8	1.3	4.6
+150-212 μm	9.9	1.1	0.5
+212 μm	5.9	1.0	0.5
Calculated head grade:		3.8	4.7

The data on the quantification of the discrete PGM's are shown in tables 4.6 to 4.10. The combined size distributions shown in the last column were calculated using the mass distribution (table 4.5) of the sample based on mineralogical examination. The PGM's that occur sub-microscopically (e.g. in solid solution in base-metal sulphides) are not included in the investigation and consequently, base-metal sulphides (BMS) losses might be accompanied by platinum-group metal (PGM) losses. The size, composition and mode of occurrence of the individual PGM grains are shown in Appendix F.

PGM size distributions

Table 4.6 shows the PGM size distribution in the size fractions. The grain sizes reported refer to equivalent circle diameter (ECD) which is the diameter of a circle with an area equal to that of the measured particle. The PGM's were mostly smaller than 10 μm in diameter in most of the size fractions. A few relatively large (~18 μm in diameter) PGM's were detected in the +25-38 μm and +106-150 μm size fractions. In the sample (combined size fractions), approximately eighty-five percent by number (~40% by volume) of the PGM's were smaller than 4 μm in diameter.

PGM mode of occurrence

Approximately 50 percent by volume of the PGM's in the -25 μ m size fractions were liberated (table 4.7). These PGM's had a maximum and average diameter of 5 and 3 μ m (-10 μ m fraction) and 20 and 7 μ m (-25 μ m fraction) respectively. Approximately 60 percent of these PGM's consist of PGM-sulphides. There is a gradual increase in the amount of PGM's associated with gangue (silicates and oxides) with increasing size from 15 percent by number in the -10 μ m fraction to approximately 70 percent in the +212 μ m fraction (figure 4.13). The oxide hosts consist mainly of chromite, although Fe-oxide (possibly magnetite) was also detected. The majority of the PGM's associated with silicates were locked in/on grain edges of pyroxene and, to a lesser extent, feldspar. A minor amount of PGM's were detected locked in Mg-silicate (talc) and biotite.

Table 4.6: PGM size distributions of the feed material to the pilot plant during phase I

SIZE CLASS (μ m)	COMBINED SIZE FRACTION		+212 μ m (12 PGM grains)		-212 μ m (14 PGM grains)		-150 μ m (26 PGM grains)		-106 μ m (25 PGM grains)		-75 μ m (25 PGM grains)		-53 μ m (29 PGM grains)		-38 μ m (33 PGM grains)		-25 μ m (75 PGM grains)		-10 μ m (71 PGM grains)	
	By		By		By		By		By		By		By		By		By		By	
	N	V	N	V	N	V	N	V	N	V	N	V	N	V	N	V	N	V	N	V
<2	37	11	32	3	46	13	45	13	56	7	56	12	50	7	36	12	50	17	46	10
2-4	52	56	42	14	33	27	45	47	28	13	24	20	31	15	57	68	33	24	39	32
4-6	10	26	12	12	18	36	3	8	-	-	8	17	11	20	7	20	17	59	8	18
6-8	1	7	8	14	-	-	7	32	-	-	8	27	4	7	-	-	-	-	3	11
8-10	-	-	1	5	3	24	-	-	8	34	4	24	-	-	-	-	-	-	2	11
10-12	-	-	1	7	-	-	-	-	8	46	-	-	-	-	-	-	-	-	1	7
12-14	-	-	1	9	-	-	-	-	-	-	-	-	-	-	-	-	-	-	<1	1
14-16	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
16-18	-	-	1	16	-	-	-	-	-	-	-	-	4	51	-	-	-	-	1	8
18+	-	-	1	20	-	-	-	-	-	-	-	-	-	-	-	-	-	-	<1	2

(By N = By Number (%), By V = By Volume (%))

The BMS hosts to PGM's were mostly pentlandite. PGM's also occurred on grain edges or locked in chalcopyrite, pyrite, pyrrhotite and millerite in decreasing order of abundance. In the +53 μ m size fractions, PGM-containing BMS were mostly locked in silicates or silicate/chromite chats while those larger than +106 μ m were all locked. In the -53 to +10 μ m size fractions, some liberated PGM-containing BMS occurred, whereas all of these BMS in the -10 μ m fraction were liberated. The BMS containing PGM's average between 6 μ m (-10 μ m fraction) and 24 μ m in diameter (-75 μ m fraction).

Table 4.7: PGM mode of occurrence of the feed material to the pilot plant during phase I

PGM MODE OF OCCURRENCE	-10 µm		-25 µm		-38 µm		-53 µm		-75 µm		-106 µm		-150 µm		-212 µm		+212 µm		COMBINED SIZE FRACTION	
	By		By		By		By		By		By		By		By		By		By	
	N	V	N	V	N	V	N	V	N	V	N	V	N	V	N	V	N	V	N	V
1	-	-	-	-	-	-	-	-	-	-	4	1	-	-	-	-	17	9	2	1
2	5	4	9	4	9	10	10	6	-	-	8	24	4	3	-	-	9	3	6	6
3	3	2	7	2	-	-	-	-	4	1	8	5	8	3	8	6	-	-	4	2
4	7	6	11	20	18	8	28	19	40	55	24	24	34	15	64	77	33	68	28	31
5	-	-	-	-	3	1	-	-	-	-	-	-	4	<1	7	7	8	4	2	1
6	-	-	-	-	12	10	14	16	8	3	8	1	4	1	7	2	-	-	6	3
7	-	-	-	-	3	3	-	-	-	-	-	-	-	-	-	-	-	-	<1	<1
8	-	-	1	<1	-	-	-	-	-	-	-	-	-	-	-	-	-	-	<1	<1
9	14	18	13	5	9	26	21	45	8	17	4	1	-	-	-	-	-	-	8	13
10	13	10	43	22	46	42	24	13	40	24	44	44	46	78	14	8	33	16	33	28
11	58	60	16	47	-	-	3	1	-	-	-	-	-	-	-	-	-	-	11	15

Legend		
1	On oxide and silicate grain boundaries	
2	On silicate grain edges	
3	On oxide grain edges	
4	Locked in silicate	
5	Locked in oxide	
6	On BMS and silicate grain boundaries	
7	On BMS and oxide grain boundaries	
8	On triple point between BMS, oxide and silicate	
9	On BMS grain edges	
10	Locked in BMS	
11	Liberated	

(By N =By Number (%), By V = By Volume (%), BMS = base metal sulphides)

("boundary" refers to the occurrence of PGM grains between two mineral grains, whereas "edge" refers to the occurrence at the intersection of resin and mineral, as observed in polished section.)

PGM-species

Table 4.8 shows the distribution of the PGM-species in the various size fractions. Approximately 40 to 65 per cent by number are sulphur-containing PGM's. The remainder consists mainly of elemental Pt, and Pt and/or Pd alloys with Fe, Pb, Cu or Sn and a relatively small percentage (10-25% by number) of PGE-tellurides, antimonides, arsenides, bismuthinides and compounds with Hg.

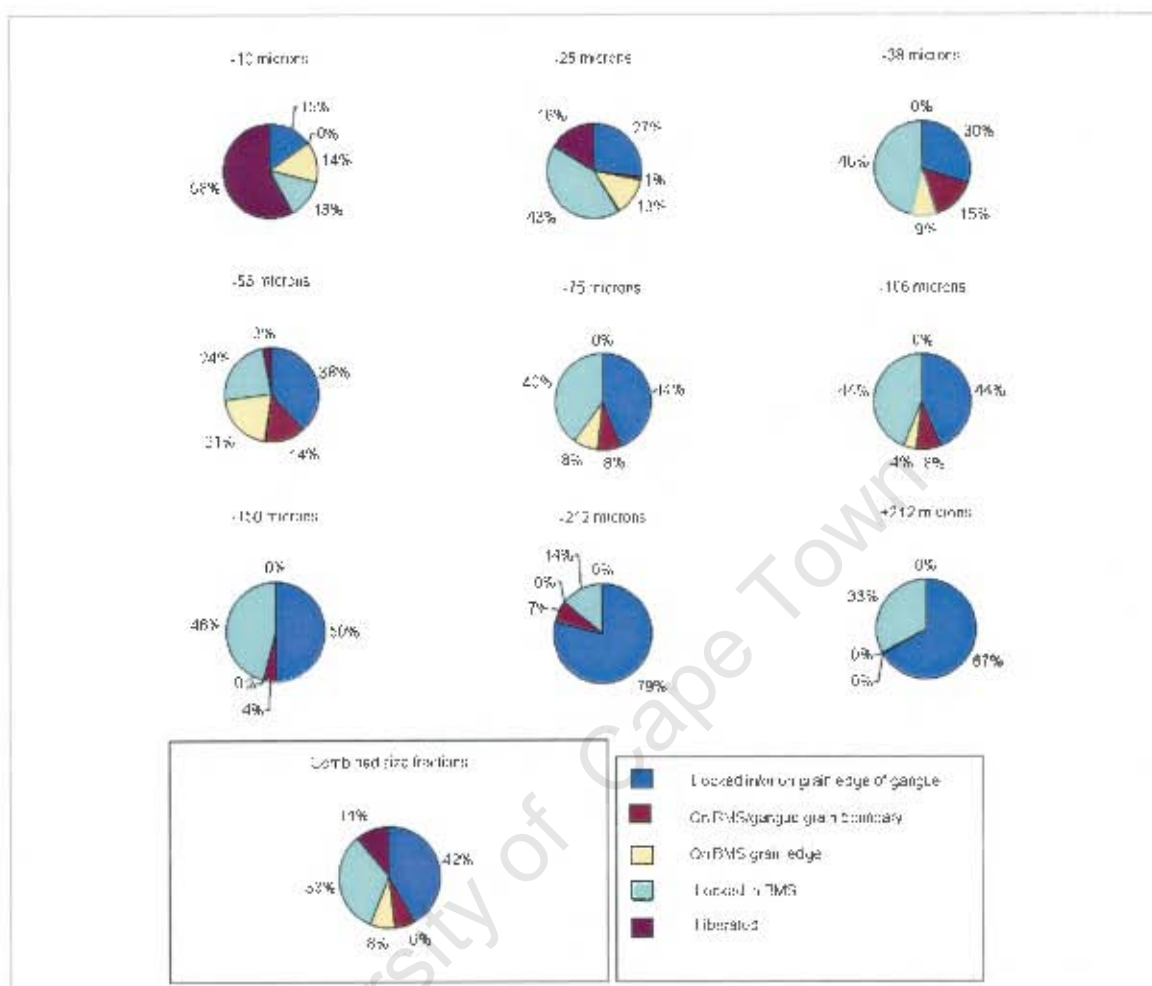


Figure 4.13: Graphical representation of the mode of occurrence of the PGM's in the size fractions of the feed material to the pilot plant during phase I

Float rank

The float rank of the PGM-containing particles as shown in table 4.9 can be calculated from the image analysis measurement. This is achieved by dividing the total area of potentially floatable component (i.e. the PGM and BMS areas) by the total area of the particle. The float rank is a measure typically used by mineralogists to determine whether sufficient liberation of the sample has been obtained and provides an indication as to how "well" the sample will respond to flotation. The PGM-containing particles progressively decrease in float rank towards the larger size classes due to an increasing proportion of unliberated PGM's and PGM-containing BMS (figure 4.14). There is a relatively large number (66%) of PGM-containing particles in the combined sample with a low float rank (<0.4). This indicates that the potentially floatable surface area of the majority of the particles in the sample had less than 40% of the surface available for collector adsorption and attachment to bubble surfaces, indicating that this sample should have a large fraction of relatively slow flotation rates. It should be noted

that the float rank of a particle should not be confused with the floatability of a particle whereby the latter takes into account all aspects relating to the flotation rate of a particle (including particle size and density particularly).

Table 4.8: PGM-species of the feed material to the pilot plant during phase I

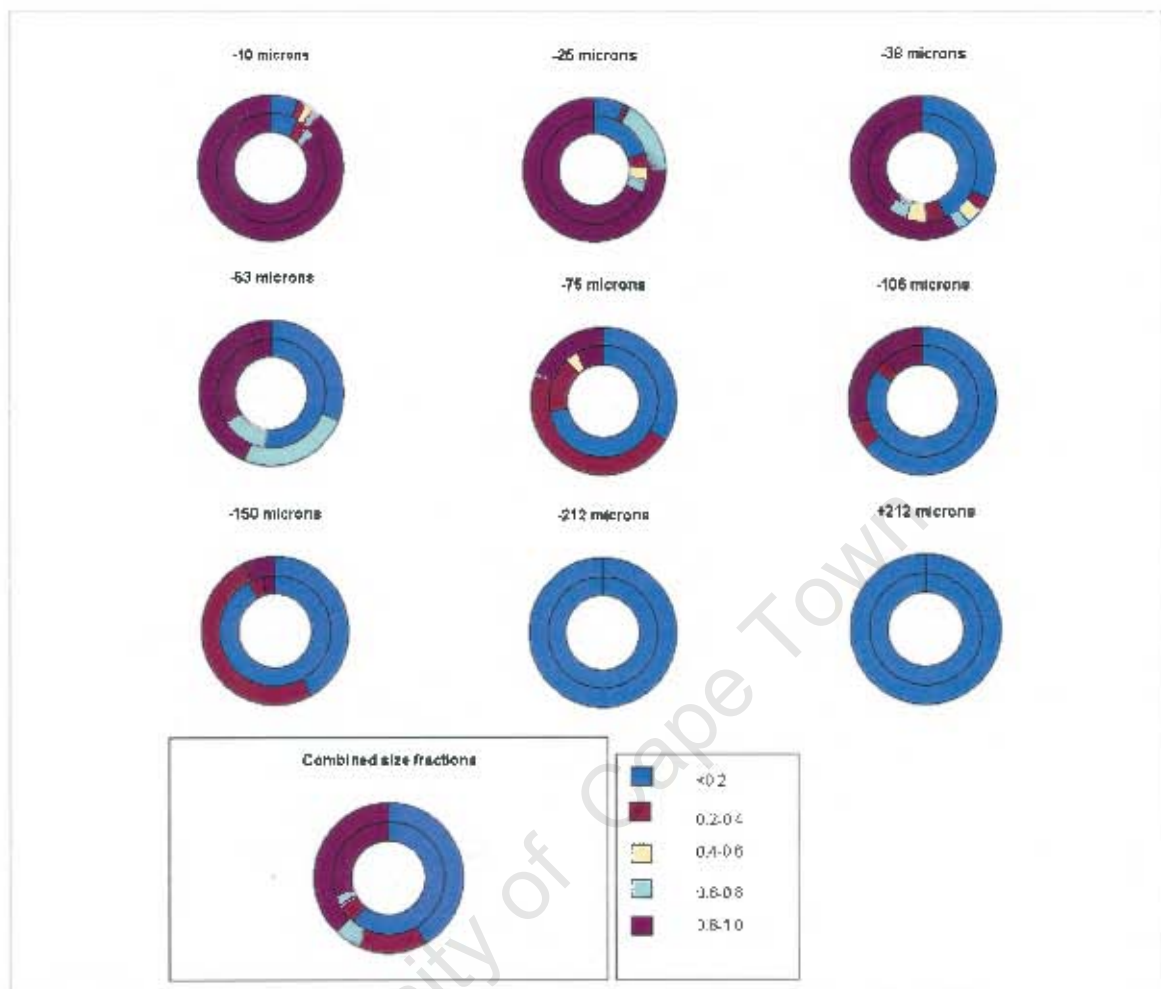
PGM SPECIES	COMBINED SIZE FRACTION																				
	-10 µm		-25 µm		-38 µm		-53 µm		-75 µm		-106 µm		-150 µm		-212 µm		+212 µm				
	By		By		By		By		By		By		By		By		By		By		
	N	V	N	V	N	V	N	V	N	V	N	V	N	V	N	V	N	V	N	V	
1	1	<1	1	2	6	2	-	-	-	-	4	1	4	1	7	6	8	1	3	1	
2	24	23	21	24	30	22	45	43	16	29	28	35	23	68	15	10	17	14	24	31	
3	15	14	15	29	12	10	14	4	24	6	12	3	23	8	14	10	25	14	17	10	
4	54	60	53	38	34	41	31	44	36	41	52	60	38	15	50	54	33	30	44	45	
5	3	2	8	2	-	-	3	1	16	2	4	1	8	7	-	-	8	36	6	4	
6	3	1	2	5	18	25	7	8	8	22	-	-	4	1	14	20	8	5	6	9	
Legend																					
1	AuAg										2	Elemental Pt. and Pt and/or Pd alloyed with Fe, Pb, Cu, Sn									
3	Pd, Pt or Rh compounded with Te, Sb, As, Bi, Hg										4	Pt and/or Pd-sulphide									
5	RuAsS, PtRhCoCuS										6	PGE-sulpharsenides									

(By N = By Number (%), By V = By Volume (%))

Table 4.9: Float rank of the feed material to the pilot plant during phase I

COMBINED SIZE FRACTION	+212 µm		-212 µm		-150 µm		-106 µm		-75 µm		-53 µm		-38 µm		-25 µm		-10 µm		FLOAT RANK
	By		By		By		By		By		By		By		By		By		
	N	V	N	V	N	V	N	V	N	V	N	V	N	V	N	V	N	V	
	61	42	100	100	100	100	92	42	84	64	72	34	52	31	43	32	20	7	<0.2
	5	15	.	.	.	.	51	4	6	4	16	47	.	.	6	3	4	1	0.2-0.4
	1	1	-	-	-	-	-	-	-	-	4	<1	-	-	6	4	4	<1	0.4-0.6
	3	5	-	-	-	-	-	-	-	-	.	.	14	25	6	3	4	17	0.6-0.8
	30	38	-	-	-	-	7	4	12	30	8	19	34	44	39	58	68	88	0.8-1.0

(By N = By Number (%), By V = By Volume (%))



(inner circle – Number percentage, Outer circle – Volume percentage)

Figure 4.14: Graphical representation of the shift in the float rank with increasing size of the feed material to the pilot plant during phase I

The following summary can be drawn from the results of the QEMSEM analysis and are listed below:

- Approximately 85 percent by number (40% by volume) of the PGM's in the sample are smaller than $4\mu\text{m}$ in diameter. The largest PGM grain ($18\mu\text{m}$ in diameter) was detected in the $-25+10\mu\text{m}$ size fraction.
- Liberated PGM's constituted approximately 10 percent of the total PGM content in the sample. The small size fractions ($<25\mu\text{m}$) contained 50-60 percent liberated PGM's by volume. Thirty percent of the PGM's in the combined sample were locked in BMS, which are mostly unliberated. Liberated PGM-containing BMS occurred mostly in the $-10\mu\text{m}$ fraction.
- The PGM's occurred mostly ($\sim 60\%$) as PGE-sulphides. The remainder consisted of elemental Pt, PGE-alloys (Fe, Pb, Cu or Sn) and Pt, Pd, or Rh compounds with Te, Sb, As, Bi, or Hg.
- The majority ($\sim 60\%$) of the PGM-containing particles in the sample had a low float rank of <0.4 . Ninety percent by number of the PGM-containing particles in the size

fractions larger than 106 μm had a very low float rank (<0.2). In the small size fractions ($<25\mu\text{m}$) a large proportion of the PGM-containing particles had a high float rank (0.8-1.0). These are the liberated PGM's and liberated PGM-containing BMS.

This chapter has been used to present a summary of the QEMSEM data obtained for the feed to the pilot plant. A method of using this data as a means to validate the regressed PGM distribution of this ore is presented in chapter 6 which is devoted to the validation of the models in this work.

4.5.1.3 Froth characterisation test work (Test2)

The froth zone is known to have a dominating effect on the overall kinetics of a flotation cell. Besides the chemical and physical pulp characteristics, the volumetric air flow rate and froth depth are the two operating parameters affecting the froth performance. These two parameters were varied in the first cell in every bank (test 1 circuit, figure 4.8) in order to model the effects of these parameters on the kinetics of the flotation cells in various parts of the plant. Only the first cell to each bank was characterised as it was assumed that only small differences in froth performance would occur down a bank, while large differences in froth stability and appearances occur in rougher, scavenger, cleaner and recleaner froths.

The rougher bank was operated in open circuit, on its own, starting with a very shallow froth depth. The air flow rate was set so that the froth just started flowing over the concentrate launder. Twenty minutes was allowed for the cell to reach steady state. Three timed samples of the rougher concentrate was taken straight after one another. This was to determine the standard deviation of the measurements caused as a result of possible operational variances, sampling errors and analytical measurement errors. Three samples of the rougher feed stream were then taken. The air flow rate to the cell was then increased, producing a slightly higher mass pull. Another twenty minutes was allowed for the cell to reach steady state, after which another set of samples were taken. The above was repeated for four different air flow rates, spanning the entire spectrum of feasible operating concentrate flow rates at the set froth depth, from very slow to very fast. Next, the froth depth was increased and the air flow rate re-adjusted repeating the same procedure described above. In total, four different froth depths were sampled, spanning the entire spectrum of feasible operating set points from very shallow to very deep froth depths. In total, sixteen combinations of froth depths and volumetric air flow rates were tested, sampling the feed and concentrate stream three times during each combination. Identical tests were performed for the first cell in the scavenger, high grade cleaner and low grade cleaner cells, except that in these cases only three air flow rates (high, medium and low mass pulls) were tested at four different froth depths and only two samples of the concentrate and feed samples were taken at each combination of operating set points. Table 4.10 lists the combination of froth heights and air flow rates used as well as the measured PGM, chromite, gangue and water flows with their standard deviations at each set point. During all the above tests,

the digilevel resistivity profile was logged in order to obtain accurate froth depth measurements. Bubble size distribution measurements were taken at each set of air flow rate set points used for every test. Four separate measurements were taken at each air flow rate in order to obtain a standard deviation of the average bubble size measured.

Table 4.10: Averaged experimental results from the froth characterisation tests of phase I

Froth Depth (cm)	Air Flow Rate (l/min)	PGM mg/hr	Stdev mg/hr	Chromite g/min	Stdev g/min	Gangue g/min	Stdev g/min	Water ml/sec	Stdev ml/sec	Dry Mass kg/hr	Stdev kg/hr	Vol Flow l/min	Stdev l/min
2.7	16.32	456.11	25.23	4.67	0.02	99.15	0.73	6.57	0.05	5.95	0.05	0.42	0.00
2.9	23.46	560.90	20.33	11.32	0.23	152.92	1.48	15.26	0.50	9.47	0.10	0.96	0.03
2.9	32.23	692.27	18.66	40.00	1.56	330.23	9.12	45.91	1.48	23.02	0.64	2.86	0.09
2.8	38.56	793.37	25.00					77.93	1.48	38.33	0.64	4.85	0.09
5.1	32.44	573.81	23.14	7.36	0.16	143.53	0.61	8.97	0.68	9.27	0.04	0.58	0.04
5	39.98	653.09	23.25	17.75	1.04	224.00	10.29	27.44	2.50	16.67	0.88	1.72	0.15
5.1	47.94	701.38	45.79	36.74	1.34	380.33	15.32	49.91	2.27	23.65	0.99	3.10	0.14
5.1	55.08	745.55	40.00	90.99	1.34	573.80	15.32	92.09	2.27	41.88	0.99	5.72	0.14
8	39.78	612.77	39.42	7.30	0.12	146.80	2.18	7.72	0.32	9.95	0.14	0.51	0.02
8.2	50.59	702.53	23.61	15.88	0.76	232.59	1.61	22.44	0.53	14.34	0.14	1.41	0.03
8	59.36	730.01	9.58	32.00	1.63	327.63	18.90	44.86	3.11	20.86	1.23	2.78	0.19
8	79.56	822.89	10.00	133.95	1.63	795.49	18.90	151.17	3.11	88.57	1.23	9.39	0.19
11.5	48.96	629.73	26.05	4.72	1.07	163.60	1.85	6.56	0.21	10.12	0.16	0.44	0.01
11.7	62.83	700.50	15.70	20.65	0.38	249.62	2.39	28.02	0.29	16.12	0.15	1.76	0.02
11.7	74.46	761.57	19.70	43.70	1.50	422.06	7.44	61.75	1.01	26.62	0.53	3.83	0.06
11.7	92	996.98	20.00	179.36	1.50	886.06	7.44	156.77	1.01	78.50	0.53	9.77	0.06

Froth Depth (cm)	Air Flow Rate (l/min)	PGM mg/hr	Stdev mg/hr	Chromite g/min	Stdev g/min	Gangue g/min	Stdev g/min	Water ml/sec	Stdev ml/sec	Dry Mass kg/hr	Stdev kg/hr	Vol Flow l/min	Stdev l/min
1.3	20.4	46.77	4.40	3.89	0.26	28.91	2.13	7.25	0.68	2.12	0.16	0.44	0.04
1.3	24.46	127.93	13.24	30.27	0.13	188.99	3.68	39.33	0.65	12.42	0.25	2.42	0.04
2.6	28.56	139.05	24.78	45.64	7.70	242.15	40.96	53.86	7.40	17.77	3.01	3.30	0.46
7.1	27.54	56.62	1.80	4.91	0.18	41.10	1.28	9.45	0.47	2.96	0.09	0.58	0.03
6.7	36.72	78.48	10.94	15.13	0.90	99.88	5.85	29.03	1.05	7.18	0.44	1.77	0.06
7.8	49.16	119.55	39.66	66.12	9.11	341.70	51.88	80.06	10.95	24.90	3.80	4.92	0.67
12	42.64	52.02	0.01	7.56	0.03	55.01	0.71	14.51	0.04	3.94	0.04	0.89	0.00
12.1	48.96	80.01	10.20	20.86	1.32	130.87	15.18	38.84	3.29	9.39	1.03	2.37	0.20
12.1	55.49	94.19	5.68	47.19	2.85	248.09	11.28	65.87	2.31	18.06	0.87	4.04	0.14
14	48.96	67.96	2.19	11.05	0.35	79.40	3.42	25.22	1.15	5.67	0.22	1.54	0.07
14	56.1	111.74	9.36	35.90	1.33	205.19	4.89	58.74	1.13	14.87	0.41	3.59	0.07

Froth Depth (cm)	Air Flow Rate (l/min)	PGM mg/hr	Stdev mg/hr	Chromite g/min	Stdev g/min	Gangue g/min	Stdev g/min	Water ml/sec	Stdev ml/sec	Dry Mass kg/hr	Stdev kg/hr	Vol Flow l/min	Stdev l/min
2.3	8.1	499.85	13.07	2.99	0.10	47.50	0.67	4.22	0.08	4.83	0.09	0.28	0.01
2.3	11.61	804.04	66.59	7.98	0.07	119.52	0.39	12.74	0.25	10.54	0.27	0.81	0.02
2.1	16.74	1079.55	244.84	21.26	4.61	256.28	54.73	37.40	8.03	20.54	4.44	2.34	0.50
5.1	12.42	504.90	39.08	3.89	0.15	71.65	1.56	3.30	0.13	6.34	0.24	0.23	0.01
4.6	21.92	782.71	36.42	11.44	0.31	174.07	4.22	18.17	0.80	13.95	0.40	1.15	0.05
5.8	29.27	836.18	15.76	23.07	0.13	269.62	7.17	43.40	0.07	20.57	0.50	2.70	0.01
7.2	21.33	697.86	43.26	6.92	0.17	127.68	2.06	7.66	0.14	10.59	0.04	0.51	0.01
7.9	26.46	630.37	60.71	10.17	1.51	166.92	11.52	17.26	0.49	12.89	1.00	1.10	0.03
8.2	34.56	966.88	235.10	28.51	6.67	344.58	101.28	60.88	17.80	12.89	1.00	3.77	1.10
10	27.81	689.82	3.71	8.76	0.38	155.34	1.80	10.61	0.04	12.33	0.11	0.69	0.00
10	32.4	721.97	37.79	11.94	0.09	156.32	12.06	21.11	0.60	12.69	0.85	1.33	0.03

Froth Depth (cm)	Air Flow Rate (l/min)	PGM mg/hr	Stdev mg/hr	Chromite g/min	Stdev g/min	Gangue g/min	Stdev g/min	Water ml/sec	Stdev ml/sec	Dry Mass kg/hr	Stdev kg/hr	Vol Flow l/min	Stdev l/min
4.3	13.5	81.32	6.44	1.84	0.08	26.24	0.49	3.33	0.02	1.96	0.05	0.21	0.00
4.4	18.9	167.20	11.84	8.42	0.01	90.13	3.77	16.78	0.43	6.51	0.18	1.04	0.03
4.2	23.22	164.48	14.66	14.72	0.05	134.86	2.36	26.34	0.24	9.57	0.09	1.62	0.01
7	22.41	141.45	4.12	3.83	0.01	48.93	0.94	5.57	0.14	3.67	0.04	0.35	0.01
7.2	25.87	188.79	2.96	5.96	0.13	75.27	1.64	10.15	0.18	5.48	0.08	0.63	0.01
7.1	31.97	199.03	10.34	14.63	1.55	150.33	6.35	28.43	0.26	10.61	0.51	1.75	0.02
9.7	29.05	144.16	15.88	4.47	0.52	64.17	0.89	6.64	0.30	4.64	0.14	0.42	0.02
10.2	33.21	154.52	2.79	7.36	0.30	88.52	1.62	13.44	0.63	6.31	0.10	0.84	0.04
9.7	37.21	164.75	19.94	12.74	0.80	122.78	6.36	24.57	1.54	8.72	0.50	1.51	0.09
13.6	35.69	106.66	1.11	4.34	0.18	54.37	2.80	7.50	0.12	3.90	0.16	0.47	0.01
13.6	46.66	172.29	1.11	20.71	0.73	188.06	13.41	39.46	0.76	12.27	0.31		

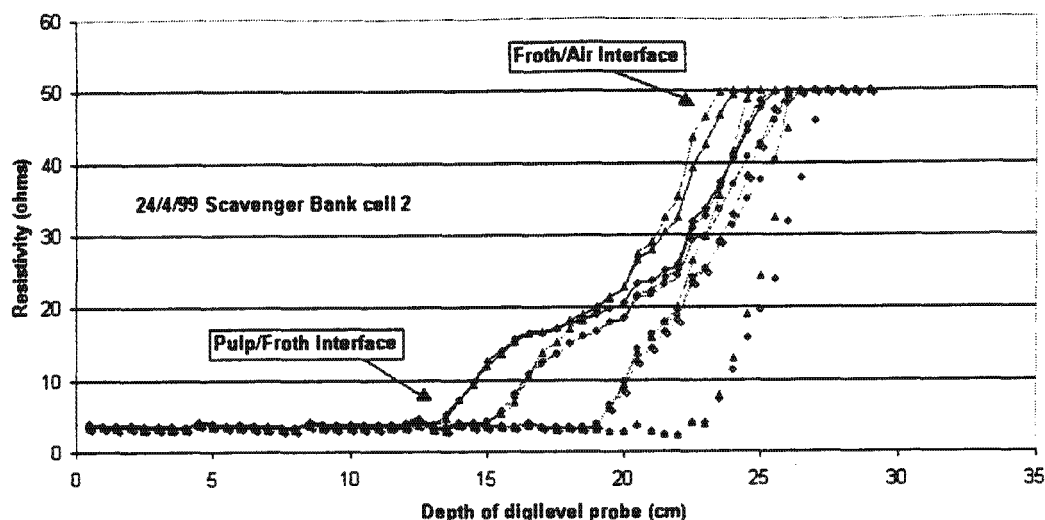


Figure 4.15: The resistivity profile for the scavenger froth characterisation test

Figure 4.15 shows the logged digilevel resistivity profiles obtained for the froth characterisation test work for the scavenger cell of the pilot plant test work of phase I. This figure was included to show the sensitivity of the digilevel profile in measuring the change in resistivity through the froth phase. A change in the slope of an individual resistivity profile can be interpreted as a change in the concentration of particles in the froth. This change in concentration with froth depth is clearly evident for the deeper froth depths in figure 4.15 highlighting the occurrence of increased drainage in the froth for the deeper froth depths. Another interesting observation from figure 4.15 is the change in the froth/air interface for increasing air flow rates. The froth/air interface rises as the gas hold-up in the cell is increased due to the increased upward buoyancy force of the rising air bubbles. Figure 4.16 shows the same trend observed for the rougher froth characterisation test, the interface of both phases are plotted during the time of day that the test was carried out.

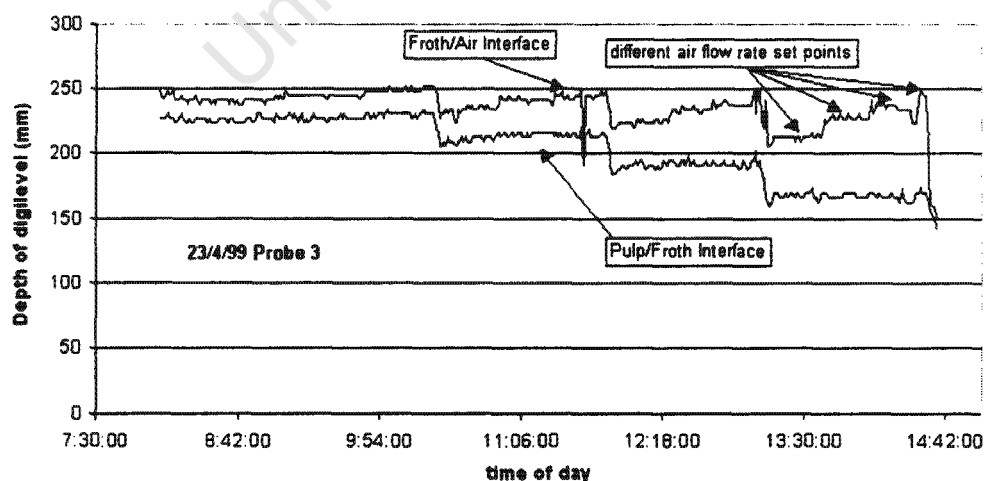


Figure 4.16: The increase in the froth/air interface with increasing air flow rates for the rougher froth characterisation test.

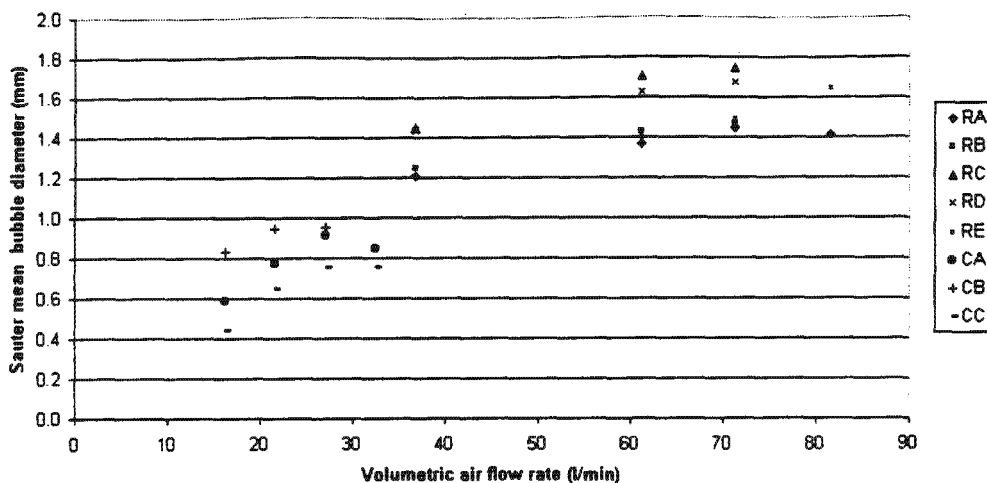


Figure 4.17: Sauter mean bubble diameters as a function of volumetric air flow rate for the rougher and cleaner banks

Figure 4.17 shows that the measured bubble diameters for the rougher and cleaner banks reached a maximum diameter and then seemed to plateau off at higher air flow rates. Smaller bubble diameters were measured in the cleaner banks.

4.5.1.4 Alternative circuit configurations (Test3)

The two circuit configurations shown in figures 4.18 and 4.19 were run in order to test whether a model derived from the data obtained in the first two tests (test 1 and test 2) could accurately predict the flotation kinetics of these circuits. The feed characteristics to both these circuit configurations were identical to the feed used during the other two tests described above. These two circuits were chosen as they were expected to give substantially different overall PGM grades and recoveries compared to the feed characterisation test. Exactly the same procedure was followed in obtaining flow rate and mass balance data as was described in test 1. Three steady state mass balances of every stream were taken for each configuration. The bubble sizes were measured in each flotation cell at the air flow rates used for the test.

Table 4.11: Averaged mass balance data for circuit configuration B. (test3)

CB	PGM mg/hr	Stdev	Chromite g/min	Stdev	Gangue g/min	Stdev	Water ml/sec	Stdev	Dry Mass kg/hr	Stdev	Vol Flow l/min	Stdev	Air Flow l/min	db mm	FH
Feed	1533.36	79.30	1362.81	121.81	5470.24	450.20	273.62	7.32	409.98	3.39	18.32	0.99			
RA	602.78	8.03	9.96	0.86	205.69	11.07	14.88	2.38	12.84	0.07	0.95	0.15	32.64	0.85	5.87
RB	363.97	32.98	9.74	1.85	178.69	19.25	17.17	4.07	11.31	0.13	1.08	0.25	32.37	0.89	6.10
RC	182.11	22.55	12.75	1.89	165.64	17.43	30.40	6.22	10.70	0.11	1.87	0.38	40.19	1.00	6.37
RD	136.22	13.15	10.23	2.23	127.10	20.19	22.92	6.05	8.24	0.13	1.41	0.37	36.86	0.84	6.50
RE	106.22	34.11	10.74	4.56	121.25	40.36	27.87	5.85	7.92	0.27	1.71	0.36	42.64	0.98	6.63
RT	606.36	77.38	1396.03	132.88	5657.44	444.40	370.50	13.47	423.21	3.46	24.19	0.97			
SA	81.72	26.61	8.16	0.76	77.31	6.30	18.27	2.30	5.13	0.04	1.12	0.14	26.52	1.03	5.80
SB	63.61	10.71	12.55	1.09	106.84	5.21	28.87	2.02	7.16	0.04	1.77	0.12	24.89	1.13	5.80
SC	42.19	2.42	9.43	0.75	73.02	4.96	20.98	1.97	4.95	0.03	1.28	0.12	24.89	1.05	5.57
SD	43.26	1.93	10.65	0.64	81.66	2.57	24.36	1.89	5.54	0.02	1.49	0.10	24.14	1.06	5.47
SE	37.84	8.72	11.25	4.00	79.35	22.96	28.56	2.13	5.44	0.16	1.74	0.13	27.27	1.13	5.37
ST	311.56	82.23	1322.54	72.32	5211.49	494.33	250.17	22.84	392.04	3.27	16.83	1.52			
HGCA	761.74	56.07	9.74	0.79	229.86	12.73	14.67	1.58	14.37	0.08	0.95	0.10	16.42	0.53	5.50
HGCB	365.39	16.89	6.01	0.08	130.14	5.57	12.04	0.35	8.17	0.03	0.76	0.02	11.93	0.78	5.30
HGCC	203.45	14.29	5.80	0.92	110.90	8.08	14.25	0.59	7.00	0.05	0.89	0.04	14.96	0.55	5.23
HGCT	310.19	28.56	39.20	5.33	415.46	50.24	82.62	11.67	27.28	0.33	5.08	0.71			
LGCA	668.43	145.79	4.03	0.54	144.92	13.21	4.05	0.40	8.94	0.08	0.28	0.03	13.00	0.80	7.70
LGCB	358.13	6.58	4.22	0.16	141.19	5.13	5.01	0.19	8.72	0.03	0.34	0.01	18.45	0.68	10.17
HGCC															
LGCT	158.83	15.96	13.70	4.44	170.44	34.70	30.12	3.70	11.05	0.23	1.86	0.23			

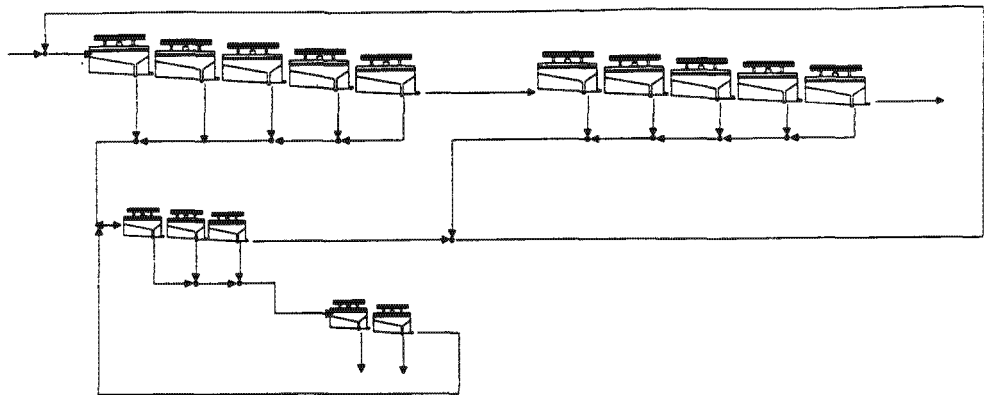


Figure 4.18: Circuit configuration B, standard rougher-scavenger-cleaner-recleaner circuit

Table 4.12: Averaged mass balance data for circuit configuration C, (test3)

CC	PGM mg/hr	Stddev	Chromite g/min	Stddev	Gangue g/min	Stddev	Water ml/sec	Stddev	Dry Mass kg/h	Stddev	Vol Flow l/min	Stddev	Air Flow l/min	db mm	FH
Feed	1473.33	82.20	1309.96	96.68	5413.24	257.89	272.71	9.57	403.39	19.39	18.23	0.64			
RA	629.53	38.24	8.00	0.24	166.48	7.49	9.00	1.06	10.47	0.44	0.59	0.07	32.64	0.85	5.60
RB	277.56	8.05	7.22	0.41	124.78	5.00	9.56	1.30	7.92	0.31	0.61	0.08	32.64	0.89	5.70
RC	97.14	23.51	7.73	0.47	90.66	6.26	14.86	0.57	5.90	0.40	0.92	0.04	38.76	1.00	6.23
RD	81.82	6.82	7.37	0.20	78.17	3.02	14.04	0.80	5.13	0.19	0.87	0.05	37.06	0.94	6.33
RE	80.44	8.60	5.41	0.67	58.65	7.88	10.17	1.74	3.84	0.51	0.63	0.11	37.74	0.98	6.47
RT	442.95	39.92	1527.95	77.75	5100.30	186.62	282.41	18.57	397.70	8.23	18.79	1.15			
SA	33.15	1.99	4.49	0.19	36.56	0.42	9.64	1.35	2.40	0.03	0.59	0.08	26.52	1.03	6.00
SB	44.33	3.26	8.35	0.24	61.08	1.33	16.46	0.51	4.17	0.08	1.01	0.03	25.30	1.13	5.40
SC	27.97	0.96	6.53	0.28	44.83	4.25	12.70	1.58	3.08	0.27	0.78	0.10	24.48	1.05	5.57
SD	23.60	0.44	6.58	0.30	42.75	2.96	12.43	0.99	2.96	0.19	0.76	0.06	22.85	1.06	5.30
SE	27.15	2.09	10.24	2.48	67.62	9.82	24.16	2.96	4.69	0.74	1.47	0.18	28.02	1.13	5.50
ST	316.04	31.88	1322.18	113.12	5008.39	336.50	213.39	20.45	379.83	26.15	14.56	1.30			
HGCA	645.65	89.75	7.31	0.44	163.01	10.86	10.09	0.25	10.22	0.68	0.65	0.02	16.25	0.53	6.37
HGCB	281.30	25.19	5.14	0.60	94.15	4.98	9.83	1.28	5.96	0.33	0.62	0.08	11.97	0.78	5.80
HGCC	102.28	13.46	2.45	0.29	46.01	5.89	4.79	0.79	2.91	0.37	0.30	0.05	11.47	0.55	5.70
HGCT	220.16	90.57	55.80	7.95	410.11	34.30	92.81	4.82	27.96	2.12	5.70	0.30			
LGCA	112.38	45.09	3.67	0.89	68.06	14.99	8.19	0.81	4.18	0.94	0.51	0.05	24.91	0.79	7.00
LGCB	74.55	20.27	3.49	0.51	57.82	8.38	10.26	0.60	3.68	0.53	0.63	0.04	23.89	0.61	8.33
LGCC	49.44	6.17	2.98	0.41	46.07	5.58	9.01	0.86	2.94	0.36	0.55	0.05	26.28	0.62	8.23
LGCT	123.32	59.84	41.56	5.64	275.20	35.05	67.34	11.01	19.01	2.23	4.13	0.67			

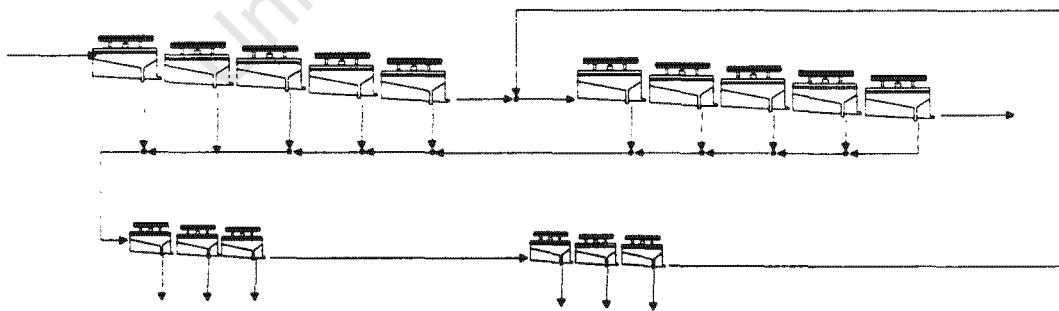


Figure 4.19: Circuit configuration C, standard rougher-cleaner circuit

4.5.1.5 Scale-up test (Test4)

In order to establish whether the flotation model could effectively account for the effect of residence time and cell volume on the flotation kinetics, a simple test was devised to evaluate both these factors simultaneously. A single eighty litre and thirty litre flotation bank were run in parallel, being fed by the same feed stream that was split in two to give approximately equal residence times across both banks. Figure 4.20 shows the experimental setup.

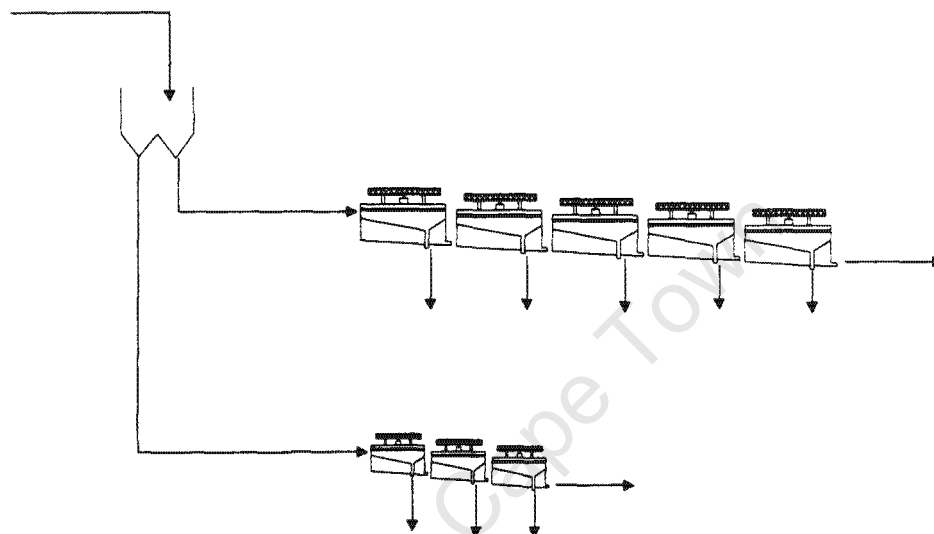


Figure 4.20: Scale-up test, 5*80l and 3*30l flotation banks run parallel to one another

Exactly the same feed characteristics were produced during this test as for test 1 described previously. The volumetric flow rate of pulp feeding the 5*80 l rougher bank was smaller than the flow rate feeding the rougher bank during test 1 as some of the feed was feeding the 3*30 l bank. The objective was to determine whether the model could accurately describe the kinetics with a change in residence time between both banks treating exactly the same feed material. In addition to this, the model could be tested to see whether it could describe the flotation kinetics between cells of different volumes treating the same feed material. Again, three sets of samples, separated by an hour residence time, were taken from every stream during this test. The froth depths and volumetric air flow rates were recorded after each set of samples were taken.

Table 4.13: Averaged mass balance data for the scale-up test, (test4)

	PGM	Stdev	Chromite	Stdev	Gangue	Stdev	Water	Stdev	Dry Mass	Stdev	Vol Flow	Stdev	Air Flow	db	FH
	mg/hr		g/min		g/min		ml/sec		kg/h		l/min		l/min	mm	cm
80 L															
Feed	1119.57	63.81	967.56	89.67	3953.12	77.48	191.74	1.92	295.24	8.59	12.87	0.12			
RA	496.57	36.25	7.18	0.71	137.29	6.87	7.24	0.75	8.67	0.39	0.47	0.04	36.52	1.20	6.20
RB	235.21	11.58	11.50	1.38	135.76	6.96	14.51	6.28	8.84	0.49	0.91	0.38	43.45	1.18	6.13
RC	62.06	4.43	5.97	0.39	56.58	1.78	9.54	0.55	3.75	0.12	0.59	0.03	43.04	1.45	6.43
RD	40.80	4.18	5.26	0.31	45.67	3.14	7.97	0.47	3.05	0.21	0.49	0.03	42.77	1.43	6.43
RE	39.29	2.85	5.83	0.27	46.75	4.35	9.86	0.75	3.16	0.28	0.61	0.04	46.85	1.18	6.47
RT	245.68	19.92	908.50	48.61	3554.36	122.93	142.61	5.12	267.77	8.90	9.80	0.27			
30 L															
Feed	315.12	32.52	211.97	45.21	925.59	119.73	52.39	0.89	68.25	9.85	3.46	0.09			
RA	129.28	20.24	3.32	0.11	47.51	0.82	4.84	0.22	3.05	0.05	0.30	0.01	24.57	1.02	6.50
RB	51.37	1.42	2.52	0.11	30.55	1.03	6.67	3.34	1.98	0.06	0.41	0.20	21.40	0.86	6.33
RC	26.25	1.89	2.56	0.34	24.88	0.75	6.14	0.28	1.65	0.05	0.38	0.02	23.42	1.09	6.50
RT	63.68	9.48	204.06	44.44	841.25	93.45	35.64	2.99	52.72	6.26	2.43	0.16			

4.5.2 Phase II test on the Mintek pilot plant

4.5.2.1 Interpretation of different feed floatability distributions (Test5)

Mass balance data together with corresponding laboratory rate floats were used to regress the feed distribution of the floatable PGM particles in chapter 3. The fractions together with the rate of each fraction determine the characteristics of the feed material entering the plant. If three floatable fractions are deemed sufficient to simulate the ore being tested (fast, slow, and non-floating), it is required to regress thirteen parameters by minimising the sum of errors between the measured and calculated recoveries from the pilot plant mass balance as well as the batch rate floats. Obviously a large combination of parameters are possible to achieve a satisfactory fit to the experimental data. It is desirable that the regressed parameters can be related in a meaningful way to the characteristics of the feed material being fed to the plant. In order to test the regression procedure, feed material obtained from two distinctly different milling conditions was characterised. Figures 4.21 and 4.22 show the experimental setup used. A sample of LONMIN Karee UG2 ore was used for this test work. Feed material A (coarse grind) was produced using a closed circuit peripheral discharge ball mill, fed at 900 kg/hr, with a 600 μ m vibrating screen producing a grind of +/-50% - 75 μ m. Feed material B (fine grind) was produced using a two stage milling circuit. The same primary mill used to produce feed material A was used except using an 800 μ m screen to lower the circulating load on the primary mill. The secondary mill was identical to the primary mill, except it was operated in closed circuit with a cyclone. The dry feed rate to the primary mill was 1200 kg/hr and milling was conducted at 70% solids. A final grind of +/- 85% - 75 μ m was produced. The exact details of the mill sizes and ball charges will not be elaborated on as this serves no purpose for the presentation of this work. What is important is to note the large difference in grind between the two materials produced for both characterisation tests.

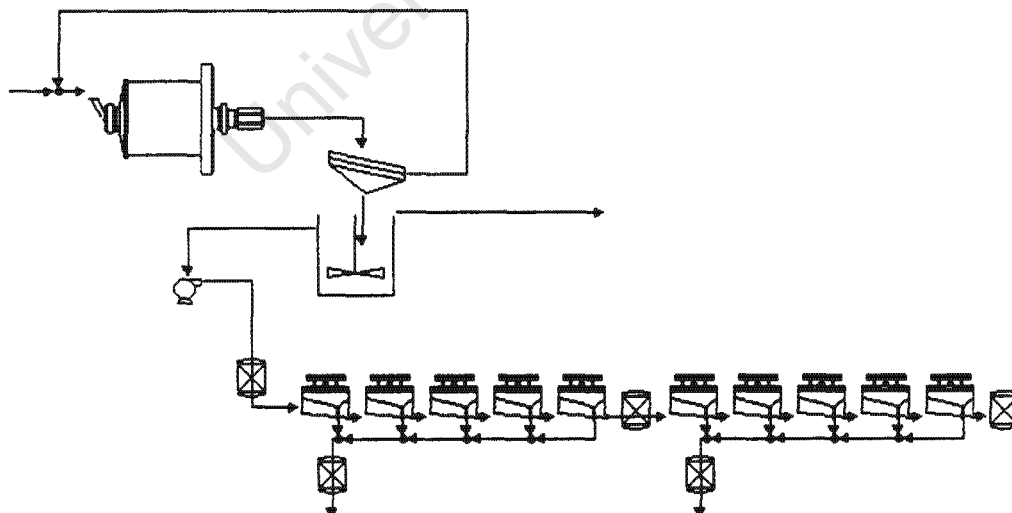


Figure 4.21: Closed circuit milling with a 600 μ m screen (coarse grind), showing streams sampled for batch rate floats

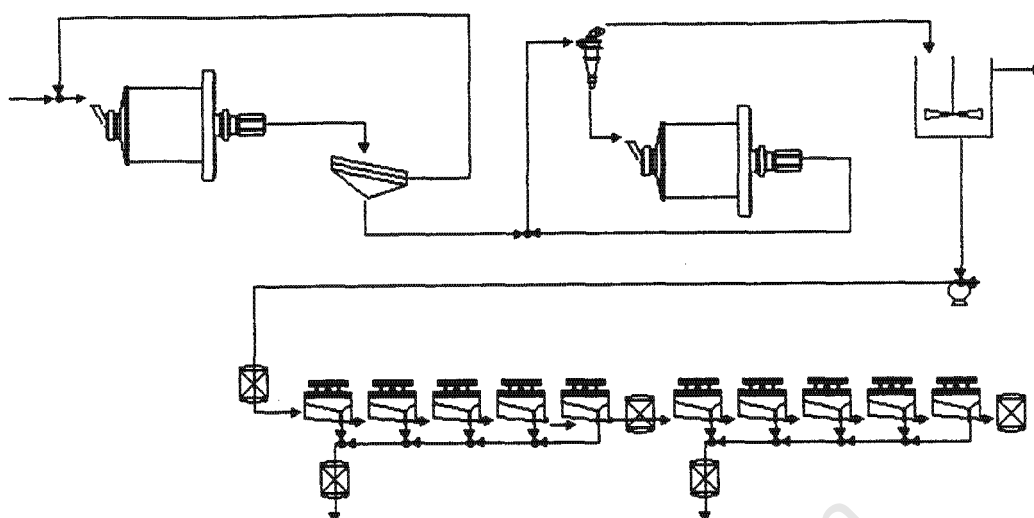


Figure 4.22: Closed circuit primary mill with an 800 µm screen followed by a closed circuit secondary mill with a cyclone (fine grind), showing streams sampled for batch rate floats

Both feeds were floated using two banks of 6x80 litre flotation cells placed in series to one another, fed at exactly the same pulp volumetric flow rates providing approximately 55 minutes of residence time across both banks. Three mass balances, each an hour apart were taken. Samples of the feed, combined rougher concentrate, combined secondary rougher concentrate as well as the tailings streams from both banks were taken for batch rate floatability tests. Exactly the same methods of collecting, preparing and analysing the samples were used as discussed in the previous sections of this chapter. The froth heights and volumetric air flow rates were measured in every cell after every mass balance. Enough sample was taken from every stream in order to measure the size distribution of every stream down the bank. Composite samples from each of the three mass balances were sized at $+106\mu\text{m}$, $-106\mu\text{m}+38\mu\text{m}$, $-38\mu\text{m}+25\mu\text{m}$, $-25\mu\text{m}+10\mu\text{m}$ and $-10\mu\text{m}$. This data was used to quantify the rates of flotation of each particle size class. This information was used to determine whether the regressed distributions (obtained from the mass balances and batch rate floats) could be correlated to the differences in the feed characteristics of the coarse and fine grind feed streams. Particular attention was given to the effect of particle size on each floatable fraction. Only the experimental results pertinent to the objectives of this work will be presented here. The detailed mass balance data is included in appendix E. The recovery of the different size fractions, the resulting batch tests as well as the modelling of the data will be presented in chapter 6 where this data is used to validate the overall model.

4.6 Model fit to new pilot plant data

This section presents the fit of the pilot plant campaign experimental data described above to the individual component models. It must be noted that the model predictions were made possible via the use of the experimentally determined water recovery and pulp and froth residence times. Chapter 5 describes an empirical water model to predict the water recovery and then presents a simulation algorithm that is used to calculate the pulp and froth residence times, providing a purely predictive simulation rather than a model fit as presented in the next few sections. The work presented here was necessary to obtain starting estimates of the model parameters in the simulation algorithm of chapter 5.

4.6.1 PGM model fit to pilot plant tests

The same analysis as described in chapter 3, section 3.1.2 was used in order to regress the parameters describing the PGM flotation kinetics from the pilot plant test work. The contribution of entrainment to the overall PGM recovery was included based on the conclusions drawn from the preliminary work on the mini plant. It was assumed that all PGM floatability components were subjected to the same degree of entrainment with water. Figure 4.23 is a schematic showing the separation of the PGM particles in the feed into floatability classes. The fraction of each floatability class that is not recovered by flotation is subjected to entrainment, proportional to a single degree of entrainment with water. All floatability classes having the same degree of entrainment with water, including the non-floating PGM fraction. Obviously this is a simplification of the true extent at which the PGM particles are entrained into the concentrate. The approach used in this work was adopted to limit the number of parameters describing the recovery of PGM particles and does however reflect a more realistic model by allowing a degree of entrainment contribution. This is particularly relevant to the non-floating fraction of PGM particles that could not be recovered in the previous PGM model of chapter 3. A more realistic approach would have been to use a size by liberation PGM model having a fraction of each size by liberation class subjected to entrainment which is a function of an entrainability distribution with particle size class. The approach adopted in this work was used in an attempt to avoid needing such a complex and detailed analysis and as can be seen from the results which follow, the need for such detail is probably not warranted. Equation 4.1 depicts the overall recovery of PGM particles as a function of a floatability distribution, including entrainment.

$$R_{PGM,k}^u = \sum_{i=0}^n m_{i,PGM,k} \left[\frac{P_{i,PGM} S_{b,k} \tau_{c,k} R_{f,PGM,k}}{1 + P_{i,PGM} S_{b,k} \tau_{c,k} R_{f,PGM,k}} + \left[1 - \frac{P_{i,PGM} S_{b,k} \tau_{c,k} R_{f,PGM,k}}{1 + P_{i,PGM} S_{b,k} \tau_{c,k} R_{f,PGM,k}} \right] * R_{w,k} * E_{PGM} \right] \quad \text{.....4.1}$$

where

$R_{w,k}$ is the water recovery in unit k

E_{PGM} is the degree of PGM entrainment for all PGM particles

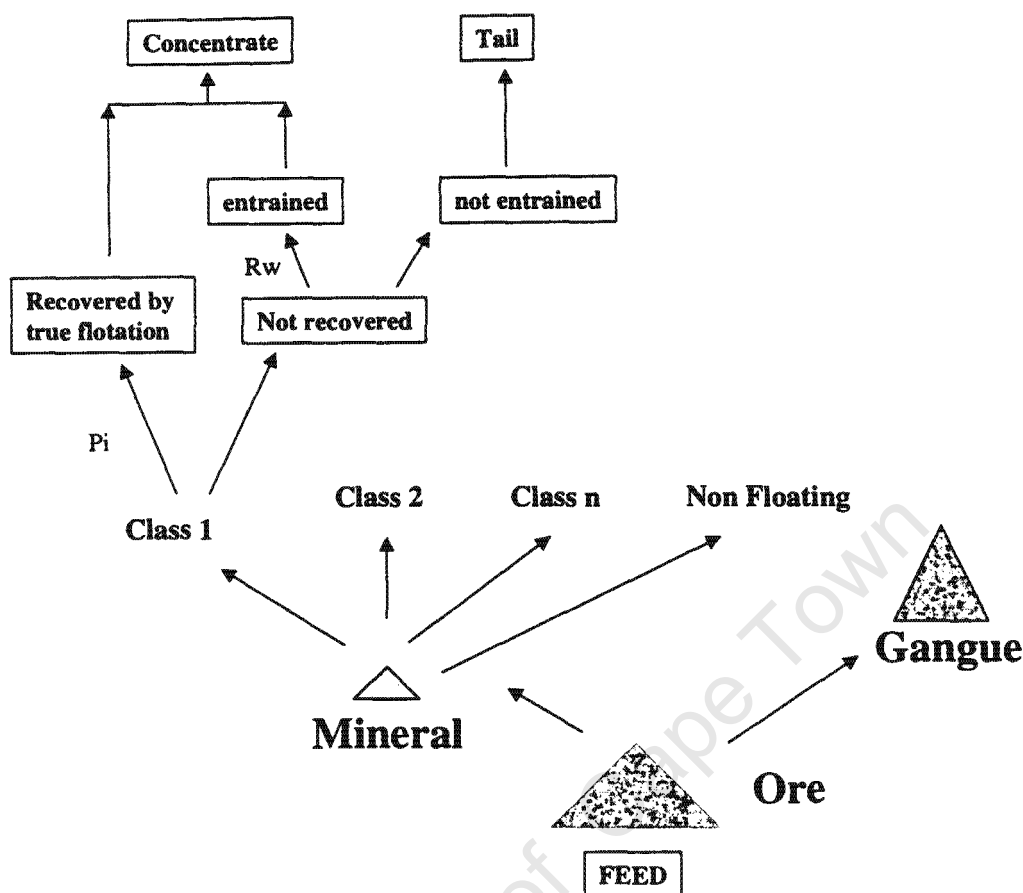


Figure 4.23: The separation of PGM components in the feed into floatability classes, depicting the mechanism adopted to include entrainment into the PGM model

Only three floatability classes (fast, slow and non) were used in the PGM model described by equation 4.1 for the pilot plant test work. Using four did not provide a much better fit to the experimental data, having virtually the same TSSE as when three were used. When analysing the parameters regressed for the mini plant (table 3.5) it becomes evident that the slow floating fraction was essentially a non floating fraction due to the extremely small floatability rate. This rate was still sufficient to act as an entrainment fraction, explaining why only three floatability classes were needed for the pilot plant regression analysis. The entrainment and the slow floatability parameters essentially had the same influence on the overall model fit. However, the basis for including an entrainment contribution puts the model onto a more realistic basis. Figures 4.24 to 4.28 show the PGM model fit to the experimental data from the pilot plant test work. Table 4.14 shows the regressed PGM model parameters for the pilot plant.

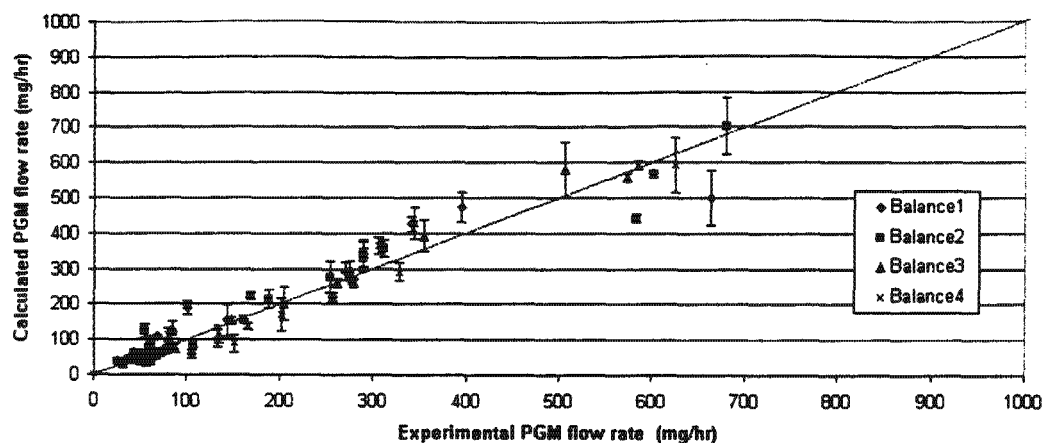


Figure 4.24: PGM model fit to experimental data for test 1, including measured standard deviations

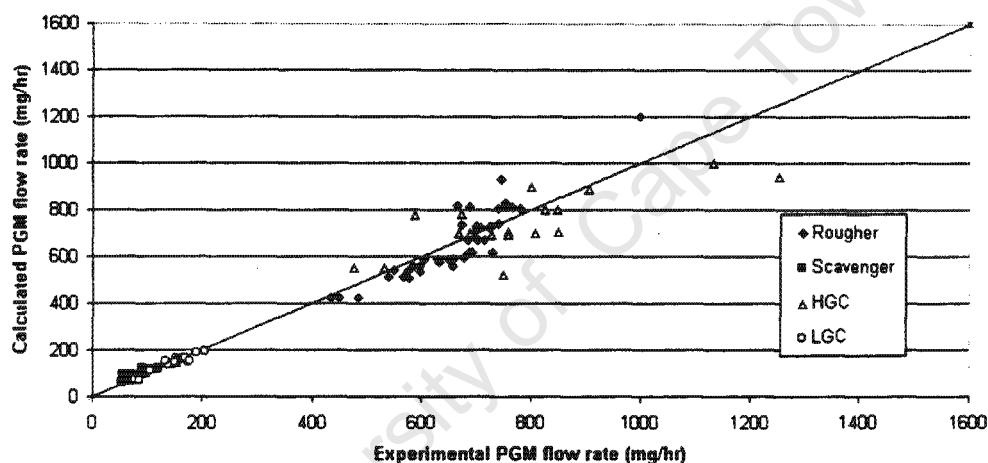


Figure 4.25: PGM model fit to experimental data for test 2

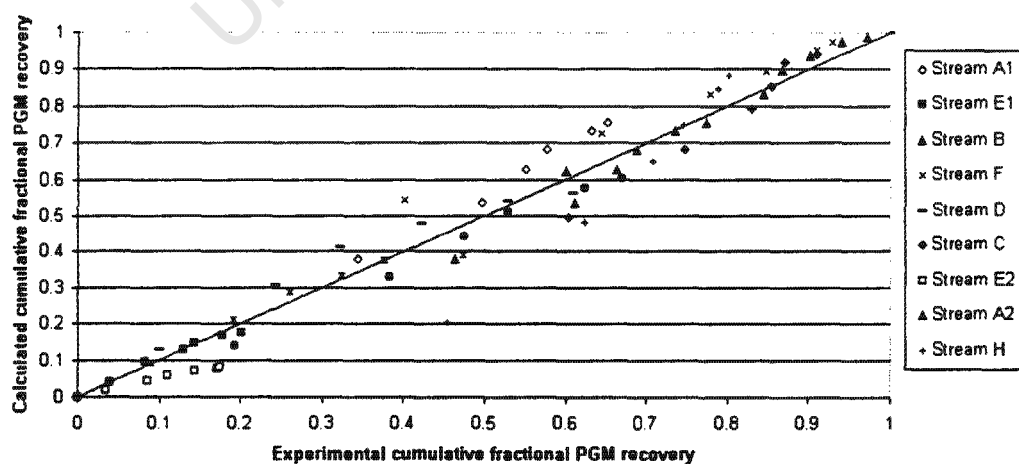


Figure 4.26: PGM model fit to experimental batch recovery time profiles (test 1)

Figures 4.24 to 4.26 show that a good fit to the experimental data was obtained. Figure 4.25 shows that the models fit for the rougher and HGC cells was not as accurate as for the scavenger and LGC cells. It would be expected that the experimental error associated with the faster floating fractions in these cells is greater than the slower fractions reporting to the scavenger and low grade cleaner concentrates. This is evident in figure 4.24 where the standard deviations associated with the high PGM flow rates are larger than the small PGM flow rates.

The PGM model parameters derived from the regression analysis in table 4.14 all seemed to take on reasonable values, as far as could be ascertained. A wider distribution of PGM floatability fractions were regressed in comparison to the ore studied in chapter 3, as was expected owing to the nature of the oxidised ore used for the pilot plant campaign. The QEMSEM analysis of the feed material is used to validate the regressed fractions of each PGM floatability class (m_i 's) in chapter 6. The parameters that were regressed from the feed and froth characterisation tests were used to predict the PGM flow rates for the two alternative circuit configurations of test 3.

Figure 4.27 is a comparison of the model fit to the experimental data for test 3. All the PGM flow rates were reasonably predicted for circuit configuration C, however the PGM predictions around the recleaner banks of circuit configuration B were not so good. Figure 4.27 has highlighted the model error for these streams. It must be noted that the model parameter set regressed for circuit configuration A was used (table 4.14). The high grade cleaner bank (HGC) froth stability parameter was used to predict the froth recovery in the recleaner bank in this circuit. It was quite apparent during the experiments that the froth characteristics in the recleaner bank were substantially different from that in the cleaner bank. The recleaner bank froth appeared much darker (more mineralised) as well as more stable with larger bubbles. It is evident that in order to obtain a good prediction of the PGM flow rates for the two alternative circuit configurations one needs to allow for the model to include the different froth zone performances by allowing for alternative froth recovery parameters.

The one major limitation of the PGM model was the unsatisfactory prediction of the first flotation cells concentrate PGM flow rate in the cleaner banks. This same phenomenon was found in a previous study modelling a UG2 ore on an industrial flotation circuit, as discussed in section 2.3.4 in the literature review, (Theron, 1998). A stable froth containing extremely floatable PGM and gangue particles made the flotation kinetics in these first flotation cells a little irregular compared to the rest of the plant. Assigning an independent froth recovery model for these cells in the circuit could be a means of obtaining a better model fit. It is possible that the flotation kinetics were inhibited in these cells, as described in section 2.2.3.1.5, and possibly a mechanism for accounting for this phenomena as proposed by Whiten (1972) should be incorporated into the PGM model.

Table 4.14: PGM model parameters derived from the regression of the pilot plant data

PARAMETER	VALUE
$P_{\text{fast,PGM}}$	2.47e-3
$P_{\text{slow,PGM}}$	2.55e-4
$m_{\text{fast,PGM}}$	0.28
$m_{\text{slow,PGM}}$	0.49
$\beta_{\text{rougher,PGM}}$	0.07
$\beta_{\text{scavenger,PGM}}$	0.09
$\beta_{\text{HGC,PGM}}$	0.07
$\beta_{\text{LGC,PGM}}$	0.01
α_{PGM}	0.00
ξ_{PGM}	0.58
$k_{\text{batch fast}}$	3.68
$k_{\text{batch slow}}$	0.25

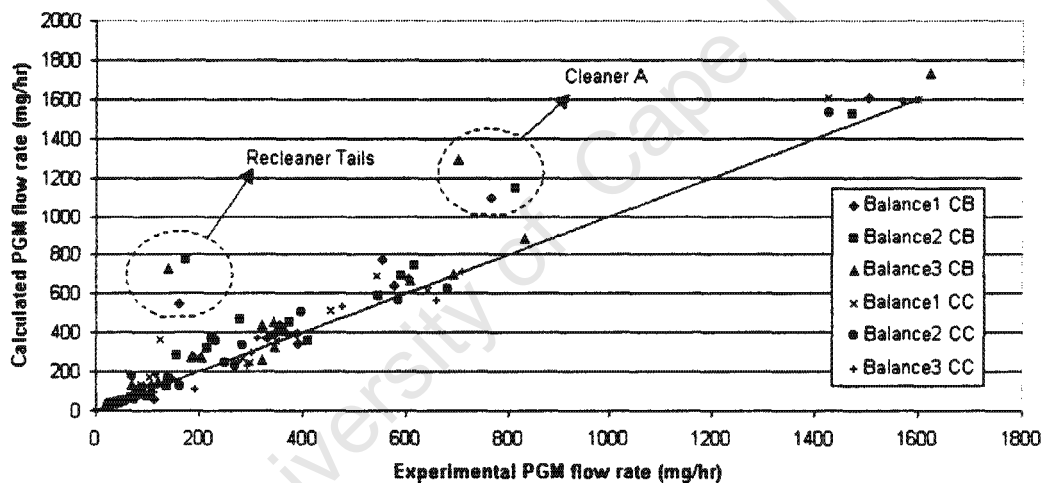
**Figure 4.27: PGM model fit to experimental data for test 3**

Figure 4.28 shows the model fit to the scale up test (test 4) using the same set of PGM parameters that were regressed from the feed and froth characterisation tests. All the PGM flow rates were predicted reasonably accurately except for the PGM flow rate of the first rougher cell in the eighty litre bank. This error is also attributed to the different flotation kinetics in the first flotation cells as discussed above. A need exists to investigate what is actually happening in these cells as they contain over 50% of all the valuable concentrate flowing through the circuit. A thorough understanding of this slightly erratic kinetic behavior of these cells would be extremely useful and beneficial to the enhanced control of UG2 flotation circuits. It is possible that bubble or froth overloading occurs (inhibited kinetics) in these first flotation cells treating most of the PGM particles as discussed previously, necessitating the need to include a model like the one described in section 2.2.3.1.5 of the literature review. It should be noted

however, that this additional complexity would be hard to justify, as the overall circuit performance is not significantly affected, as this is largely determined by the response of the slower floating minerals.

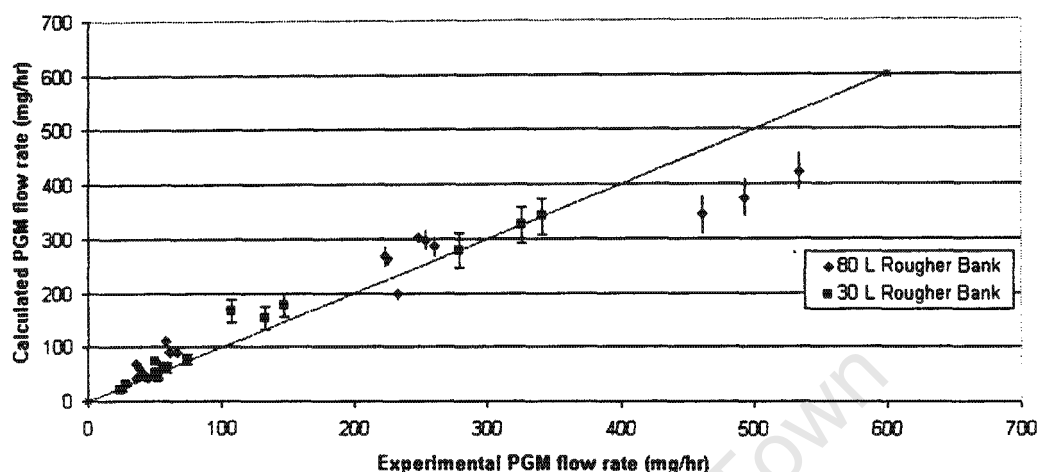


Figure 4.28: PGM model fit to experimental data for test 4

4.6.2 Chromite model fit to pilot plant tests

Figure 4.29 below has been included to show the consistency of the experimental data compared to the preliminary model development on the mini plant. A linear relationship was obtained (test 2) between the chromite and water recovery with a different slope for each bank in the circuit. This figure shows a decrease in the slope (entrainability) for the HGC, LGC, rougher and scavenger banks respectively. The highest slope in the HGC cell can be attributed to a higher fraction of small, entrainable chromite particles in this cell and therefore a higher recovery in this cell. The chromite particles not recovered in the HGC bank are sent to the LGC bank, these particles are less "entrainable", as otherwise they would have been entrained in the HGC. Following the same reasoning the difference in the slope between the rougher and scavenger banks can be explained. Figure 4.30 shows the model prediction for the chromite flow rate versus the experimental data from the feed characterisation test. Only the concentrate streams are depicted in this figure to depict the model fit for the smaller chromite flow rates as the feed and tailings chromite flow rates were obviously very much larger. Their predictions were however also within the measured standard deviations. The majority of the data points were predicted within their measured standard deviations in figure 4.30. Even though extreme care was taken to run the pilot plant as smoothly as possible, figure 4.30 highlights the dynamic nature of a flotation plant when looking at the error bars of the experimental data points. The extremely large standard deviations in figure 4.30 were for the scavenger concentrate cells which were pulled harder than the rougher cells to increase the recovery, explaining the higher chromite flow rates as well as the associated higher standard deviations.

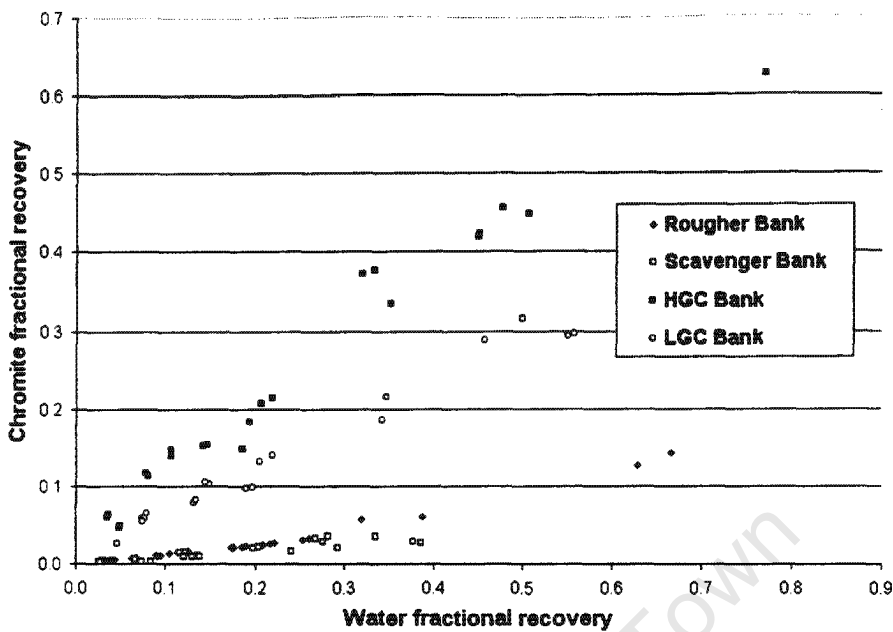


Figure 4.29: Linear relationship between the chromite and water recovery from the froth characterisation tests on the pilot plant

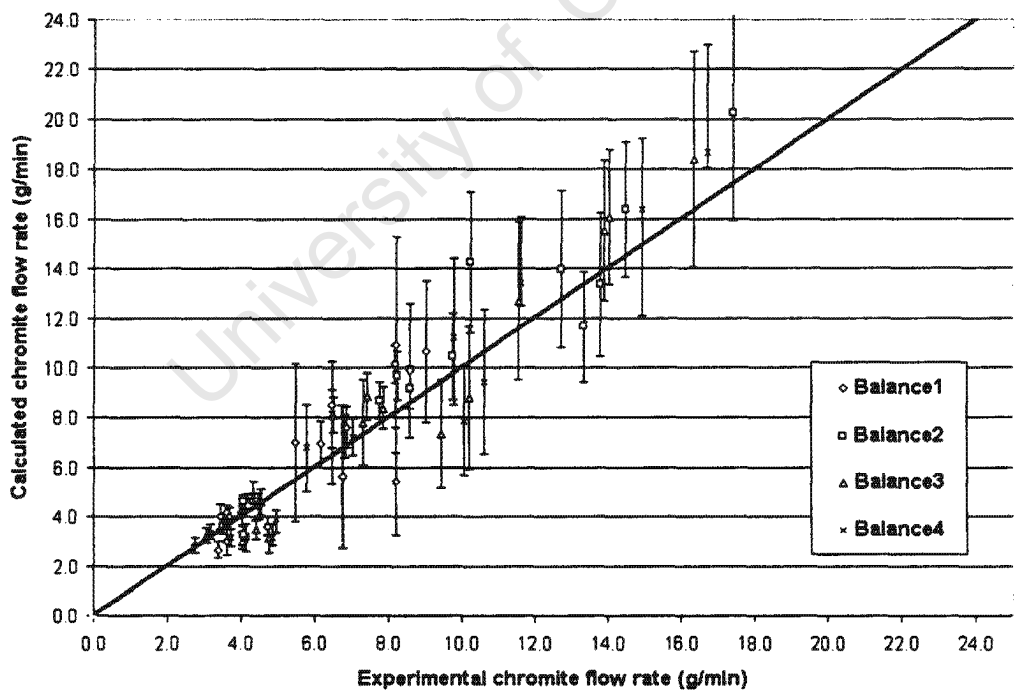


Figure 4.30: Chromite model fit to experimental data for test 1, including measured standard deviations

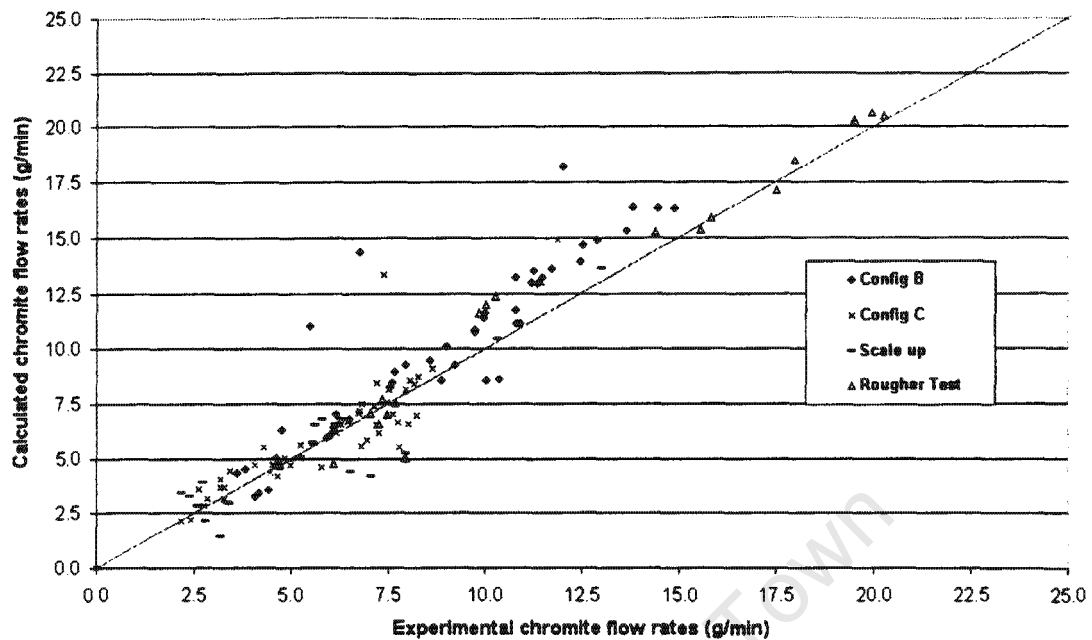


Figure 4.31 : Chromite model fit to experimental data for test 2, 3, 4

Figure 4.31 shows that the chromite flow rates could be adequately predicted for the alternative circuit configurations, scale up test as well as the froth characterisation tests using the same model parameters. This is provided the water recovery, froth and cell residence times can be predicted or measured from experiment. A total of 446 data points were predicted from the nine parameters in table 4.15 below.

Table 4.15: Chromite model parameters derived from the regression of the pilot plant data

PARAMETER	REGRESSED VALUES
$\phi_{act,Cr,feed}$	0.00675
ξ_{Cr}	18.33
δ_{Cr}	0.635
P_{Cr}	1.86e-4
$\beta_{rougher,Cr}$	0.1
$\beta_{scavenger,Cr}$	0.1
$\beta_{HGC,Cr}$	0.0038
$\beta_{LGC,Cr}$	0.002
α_{Cr}	0.02

The regressed parameters in table 4.15 correspond well with the parameters obtained from the mini plant test work. Figure 4.32 shows the cumulative particle size distribution and the entrainability function that describes the chromite entrainment kinetics on the pilot plant.

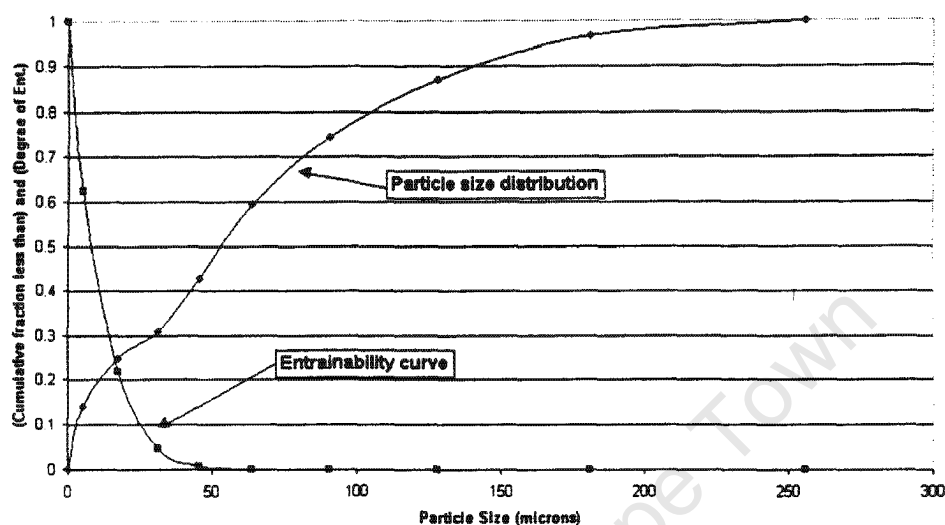


Figure 4.32: The pilot plant chromite entrainability function as well as the experimentally measured particle size distribution of the feed stream

Figure 4.32 shows the particle size distribution of the chromite particles in the feed to the pilot plant. The high proportion of fine chromite particles is attributed to the nature of the ball mill used during this work. The preliminary test work on the mini plant used a rod mill that produces a narrower particle size distribution preventing an over-grinding of the finer chromite particles which is obviously beneficial as this limits the fraction of entrainable chromite. Unfortunately a trade off is made on plants, as rod mills are more expensive and more difficult to operate than ball mills and therefore plants tend to opt for the latter. The entrainability curve in figure 4.32 shows that particles greater than 50 μ m will not be entrained with water into a concentrate.

4.6.3 Gangue model fit to pilot plant tests

Equation 4.2 shows an improved version for the gangue recovery equation based on the findings in chapter 3. The first part of the equation defines the recovery due to entrainment as a function of particle size and the second term defines the recovery due to the floating gangue particles based on the floatability distribution of the feed gangue particles to the plant. This equation was used to predict the gangue flotation kinetics using the pilot plant data.

$$R^o_{Ga,k} = (1 - \phi_{act,Ga}) * \sum_{i=1}^d (E_{Ga,i} * R_{w,k}) +$$

$$\phi_{act,Ga} \sum_{i=1}^n m_{i,Ga,k} \left(1 - \frac{1}{1 + P_{Ga,i} S_{b,k} \tau_{c,k} * ((1 - \alpha_{Ga,k}) * \exp(-\beta_{Ga,k} \tau_{f,k}) + \alpha_{Ga,k})} \right)$$

.....4.2

Figure 4.33 shows the gangue flow rate predictions of the model compared to the experimental data from test 1. The standard deviations are included showing that most of the data points were predicted within the experimentally determined variance. A total of 164 data points were generated from test 1 and the rougher froth characterisation test 2. These were used to regress the 11 parameters to describe the gangue kinetics. A model fit of 1.73 was obtained showing a relatively good fit to the experimental data.

Figure 4.34 shows the gangue flow rate predictions of the model compared to the experimental data from test 2, 3 and 4. The data points are evenly distributed around the model fit line ($y=x$) showing no bias in the model predictions (ie: no consistent over or under prediction of the data). It can be concluded that the gangue model described by equation 4.2 using a distribution of two gangue floatability classes as opposed to a single floating fraction (preliminary model) describes the gangue flotation kinetics of the pilot plant reasonably well.

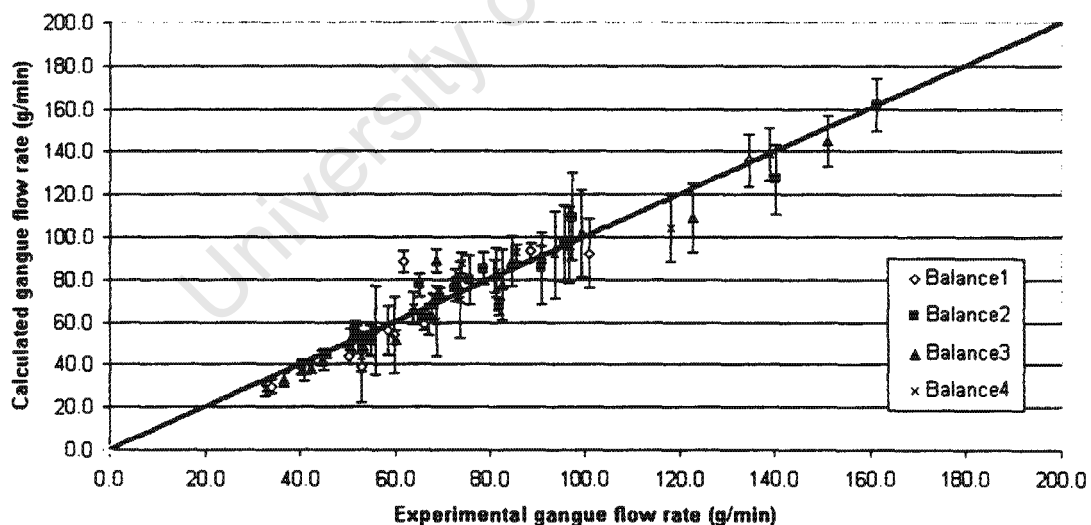


Figure 4.33: Gangue model fit to experimental data for test 1, including measured standard deviations

Table 4.16 shows the regressed gangue model parameters for the pilot plant. A lower entrainability value (ξ_{Ga}) was obtained for the gangue compared to the chromite. This is due to the higher specific gravity of the chromite particles causing a higher degree of entrainment throughout the size classes. The fraction of floatable gangue particles was calculated as 5% of the total dry mass in the feed. In a paper by Liddell et al (1986) it was found that on average, 4.85 % of the material reporting to the concentrates of UG2 circuits can be associated with non-PGM bearing sulphides and talc which correlates well with the regressed floatable gangue fraction.

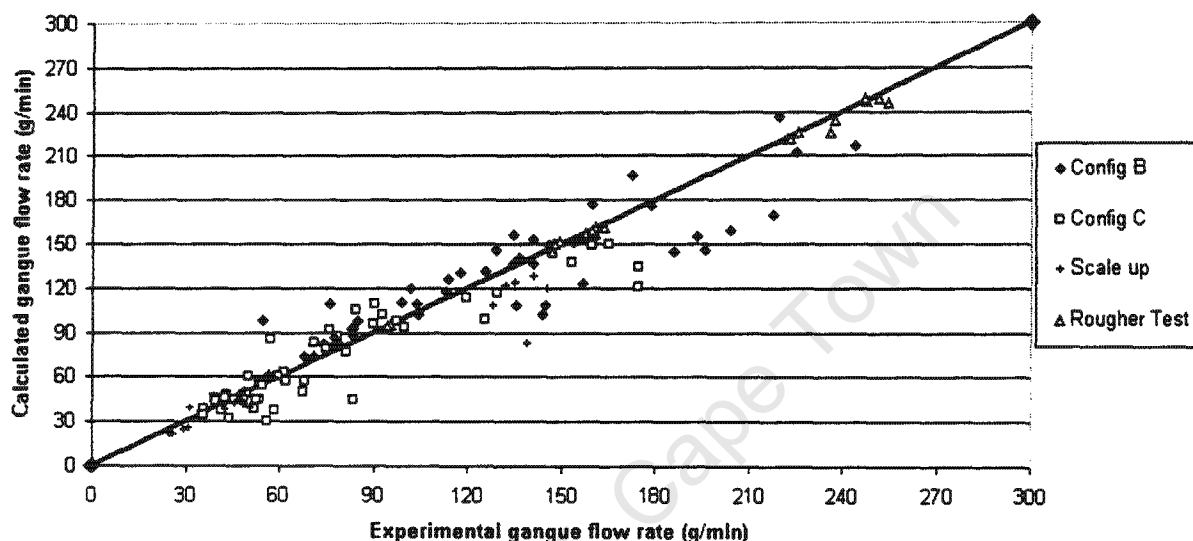


Figure 4.34 : Gangue model fit to experimental data for test 2, 3, 4

Table 4.16: Gangue model parameters derived from the regression of the pilot plant data

PARAMETER	VALUE
$\phi_{act,Ga,feed}$	0.05
ξ_{Ga}	17.42
δ_{Ga}	1.26
$P_{fast,Ga}$	1.77e-4
$P_{slow,Ga}$	1.42e-4
$m_{fast,Ga,feed}$	0.05
$\beta_{rougher,Ga}$	0.05
$\beta_{scavenger,Ga}$	0.07
$\beta_{HGC,Ga}$	0.03
$\beta_{LGC,Ga}$	0.01
α_{Ga}	0.03

Figure 4.35 shows the particle size distribution of the gangue particles in the feed to the pilot plant. Again, a high proportion of the gangue particles are in the $-38\mu\text{m}$ fraction owing to the ball milling as discussed in section 4.6.2. Gangue particles larger than $50\mu\text{m}$ do not get entrained by water into the concentrate according to the entrainability curve in figure 4.35. A higher proportion of the $-10\mu\text{m}$ gangue particles are entrained compared to that for chromite, owing to the larger area under the curve in figure 4.35 compared to figure 4.32. This is as a result of the smaller average specific gravity of the gangue particles compared to the chromite particles.

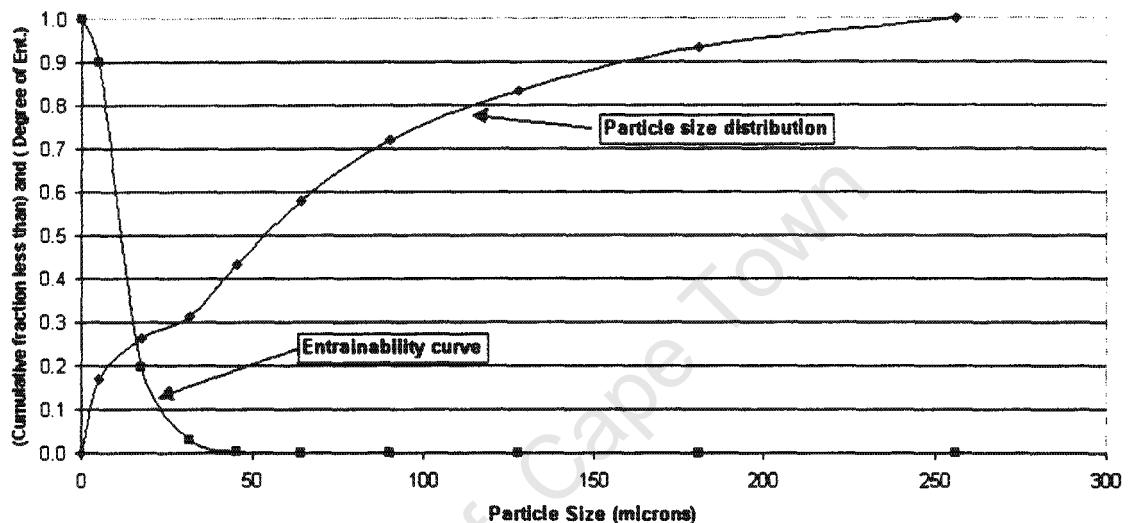


Figure 4.35: The pilot plant gangue entrainability function as well as the experimentally measured particle size distribution of the feed

4.7 Validation of Gorain's semi empirical bubble surface area flux (S_b) equation

Gorain (1997) developed a method of predicting the bubble surface area flux (S_b) in a flotation cell as a function of cell operating parameters. The bubble surface area flux is a function of the volumetric flow rate of air into the flotation cell as well as the sauter mean bubble diameter in the flotation cell. In order to carry out simulations on the pilot plant it was necessary to be able to predict the change in bubble diameter as a function of the air flow rate in the cell. The basis of this model has been discussed in the literature review (section 2.2.5.2.4).

As discussed in the literature review, in a mechanical flotation cell the production of S_b may be assumed to be dependent on the impeller peripheral speed (N_s), air flow rate per unit cell cross-sectional area (Q/A), impeller aspect ratio (As) and the 80 % passing feed size. (P_{80}). This can be expressed mathematically as:

$$S_{b,k} = a N_s^b \left(\frac{Q}{A} \right)^c A s^d P_{80}^e \dots \dots \dots 4.3$$

An Excel spreadsheet was used to regress the five empirical parameters in equation 4.3 to the results obtained from all the froth characterisation tests of phase 1, section 4.5.1.3. This test produced a wide variation of experimental results as a function of the volumetric air flow rates in all the flotation banks of the pilot plant circuit A, providing adequate experimental data to test the validity of this model. Figure 4.36 shows the model fit of equation 4.3 to the froth characterisation tests of the pilot plant.

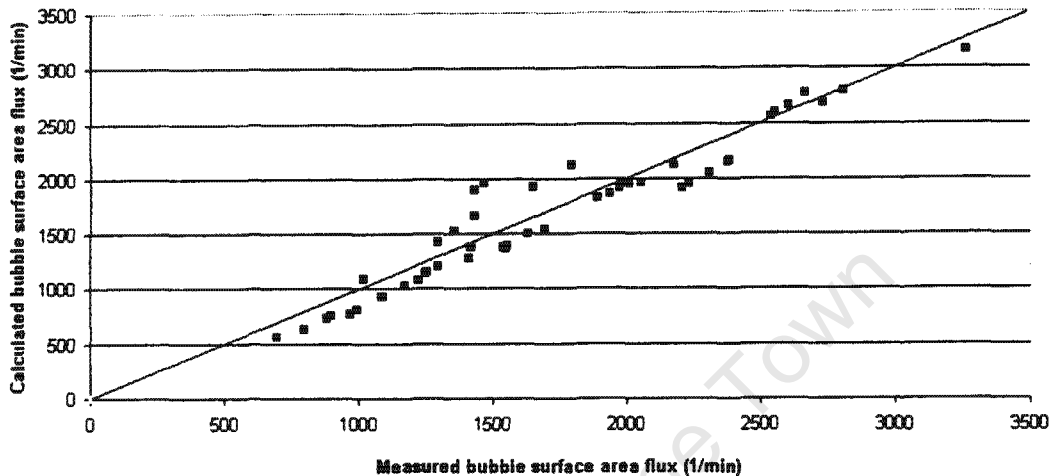


Figure 4.36: The prediction of the bubble surface area flux for the froth characterisation test work using Gorain's S_b empirical model

The fit to the experimental data was very good as is seen by the correlation in figure 4.36. The regressed parameters from the pilot plant using this model is shown in table 4.17, a comparison of these parameters to the parameters obtained by Gorain (obtained from 100 sets of experimental values obtained for a variety of impellers) is also shown.

Table 4.17: The regressed S_b parameters from the pilot plant compared to Gorain's parameters

PARAMETER	PILOT PLANT	GORAIN
a	125.00	123
b	0.514	0.44
c	0.662	0.75
d	0.00	-0.10
e	0.456	0.42

A close comparison of parameters a, b, c were obtained. As stated by Gorain (1997) the value for the parameter c is significantly higher than the other parameters b, d, e. Therefore the variable 'Q/A' has the most dominating effect on S_b compared to the other variables. The results from the pilot plant show that the impeller aspect ratio had no

effect on the prediction of S_b . This could possibly be due to the fact that only a small difference in this variable was presented by this data set (between the 30 litre cells and the 80 litre cells) and that this difference was not significant enough to present itself as an important variable for this work.

It is concluded that the use of equation 4.3 to predict the bubble surface area flux as a function of the cell operating parameters for the pilot plant is sufficiently accurate for simulation purposes.

University of Cape Town

Chapter 5

DEVELOPMENT OF THE SIMULATION ALGORITHM

In previous chapters, a methodology has been identified that can be used to describe the performance of a flotation circuit, based on sub-process models of mineral kinetics, froth recovery and entrainment. In addition, an algorithm for fitting these models to data to extract the model parameters has been presented, and it has been shown that the models were able to describe the performance of both the Mintek mini plant and pilot plant very well. This chapter looks at the application of these models for circuit simulation.

In order to apply the circuit modelling methodology to simulation, an additional sub-process model is required to describe the behaviour of water. All the preliminary work in the previous chapters made use of the experimentally determined pulp and froth residence times as well as the water recoveries to determine the model recoveries. In order to develop a comprehensive simulation algorithm the flow rate of water to the concentrate, as a function of cell operating parameters needs to be predicted. This chapter presents the development of an empirical water model based on the data from the froth characterisation tests presented in chapter 4. Following this, the simulation algorithm for the pilot plant treating UG2 ores is presented. This algorithm combines each of the four models describing the PGM, chromite, gangue and water kinetics into a mathematical form that is solved via a nested iteration sequence of successive approximations of the cell and froth residence times for each unit in the simulated circuit. The nested iteration sequence is discussed with reference to its stability and rates of convergence.

The output of the pilot plant simulation algorithm is then discussed using the preliminary parameters obtained in chapter 4 as well as the water model parameters obtained in this chapter. The output of the simulator is presented for a unit rougher cell, showing the resulting grade and recovery predictions as a function of the froth height and volumetric air flow rates to this cell. A comparison between the experimental grade recovery curve obtained from the rougher froth characterisation test and the simulation output is presented. The simulator output is discussed with respect to its boundary conditions as well as its form.

5.1 The development of an empirical water model

The importance of accurately describing the water kinetics in flotation circuit simulation cannot be underestimated. The volume of water in every concentrate stream of a plant typically constitutes more than sixty percent of the total volume and therefore accounts for over half the volume flows in the circuit. The volume flows control the residence times in the flotation cells of the circuit. These residence times are used in the mineral

recovery models. Obviously, an error in the water model results in errors in the floatable mineral kinetics as well as the entrainment kinetics.

The above clearly highlights the central role of the water recovery model in accurately simulating the performance of flotation circuits. Without an accurate water model to “drive” the other models in the simulator, the results obtained from any simulation exercise could be highly suspect. The next section describes an empirical approach to modelling the water kinetics from a bank of flotation cells in a circuit. This model incorporates the froth height and the volumetric air flow rate as variables. These are important control variables of a plant that are typically not included in other models presented in the literature.

5.1.1 Introduction to previous water modelling attempts

The purpose of this paragraph is to briefly describe the water models used previously in the literature and to show the lack of an appropriate model that can be used in the simulation algorithm of the pilot plant. Ideally, a simulation model should be able to predict the water losses in the concentrate and tailing streams, as well as the amount and location of water additions. However, no satisfactory model is yet available that allows this. Harris and Rimmer (1966) stressed the importance of the dependence of the hydrophilic gangue recovery rate on the water recovery rate in a flotation cell. The partial dependence of the recovery rate on hydrophobic gangue and valuables on the water recovery rate was outlined in this work. The complicated nature of this interdependence means that either a detailed mechanistic model for the water recovery is necessary or that a consistent methodology is required to overcome the need for a complicated water recovery prediction. This second approach was followed in the development of the work presented in this thesis, as the development of a comprehensive water recovery model was not the objective of this thesis. The objective was to develop a simulation algorithm of the pilot plant using existing models as far as possible.

Previous attempts at describing the water kinetics from a flotation cell have typically been very crude. Scrimgeour et al (1970) assigned a flotation rate equation for water as an individual species and calculated the water balance on this basis. This does not facilitate the prediction of water flow rates as a function of cell operating variables however. Smith (1984) pointed out that this approach is limited as the water behaviour can only be simulated for a bank of flotation cells and an entire circuit cannot be simulated. Loveday and Marchant, (1972) assumed a residence time in a flotation cell and calculated the water flow rate knowing the cell volume and pulp density. King (1973) assumed a constant value of the pulp density for the concentrate stream allowing a water balance to be calculated. This is known not to be true for all cells in a circuit, especially between the roughers and cleaners where thicker more stable froths occur in the latter. Woodburn et al (1965) took water as the reference species throughout the circuit and defined enhancement ratios for all other species based on what happens to water across a cell. Sutherland (1981), using data from a lead flotation

circuit, assumed a fixed pulp density in the tailings stream. This assumption was made because the pulp density of the concentrate was much less than the tails pulp density. However, this assumption lead to large errors in the water recovery prediction in the cleaning and recleaning stages of a circuit as the pulp density decreased dramatically down the bank. Figure 5.1 shows the large decrease in the percentage solids of the concentrate streams obtained from the pilot plant work (test1). This highlights the difficulties in the use of models such as those described by King and Sutherland.

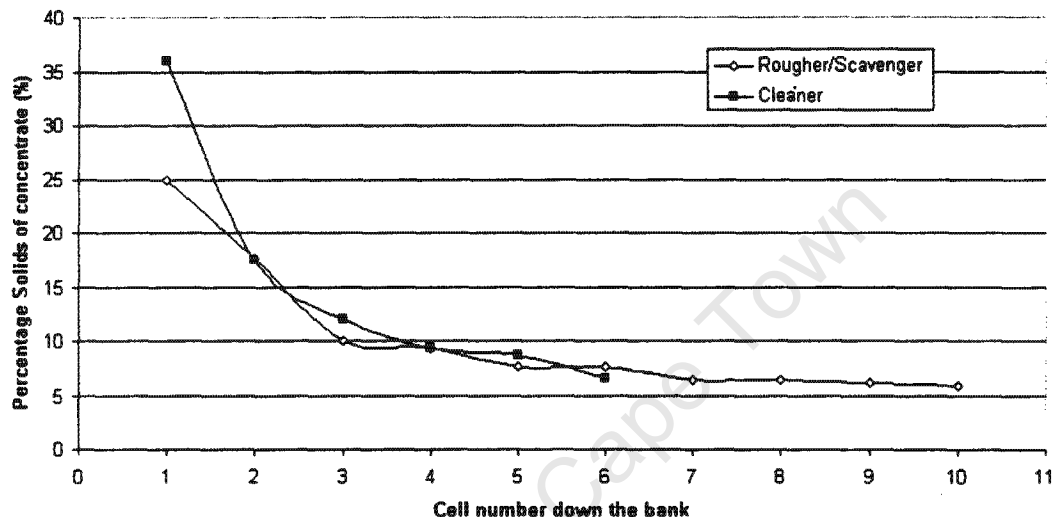


Figure 5.1: The decrease in the concentrate percentage solids down the rougher and cleaner banks of the pilot plant, (test 1, chapter 4)

King (1973) related the water flow rate to mass flow rate described by the equation below, provided the specific gravity of the pulp in the concentrate can be predicted or measured. This fixes the water flow rates for the whole plant. This approach is quite popular owing to its ease of use. Predictions can be reasonable if circuit changes explored in a simulator are not too radical.

$$Q_w = a F_s^b \dots\dots\dots 5.1$$

where

- Q_w concentrate volume flow of water (l/hr)
- F_s total concentrate mass flow rate (kg/hr)
- a water recovery slope parameter
- b water recovery exponent parameter.

In an attempt to use this model for the development of the simulator in this thesis, figure 5.2 was plotted from the experimental results obtained from the pilot plant froth characterisation tests (test 2, chapter 4). It must be remembered that this data was obtained from the first flotation cell to each bank of the feed characterisation circuit (figure 4.8, chapter 4) operating each cell at a number of different froth heights and volumetric air flow rates. The model seemed to work quite well predicting the water flow rate provided the dry mass flow rate is known and different slope and recovery parameters are used for each bank.

The curves in figure 5.2 show an increase in the water flow rate for a constant dry mass flow rate for the HGC, Rougher, LGC and Scavenger cells respectively. This makes sense, indicating the progression of increasing entrainment between the different banks as would be expected due to lower floatable minerals in the order described above.

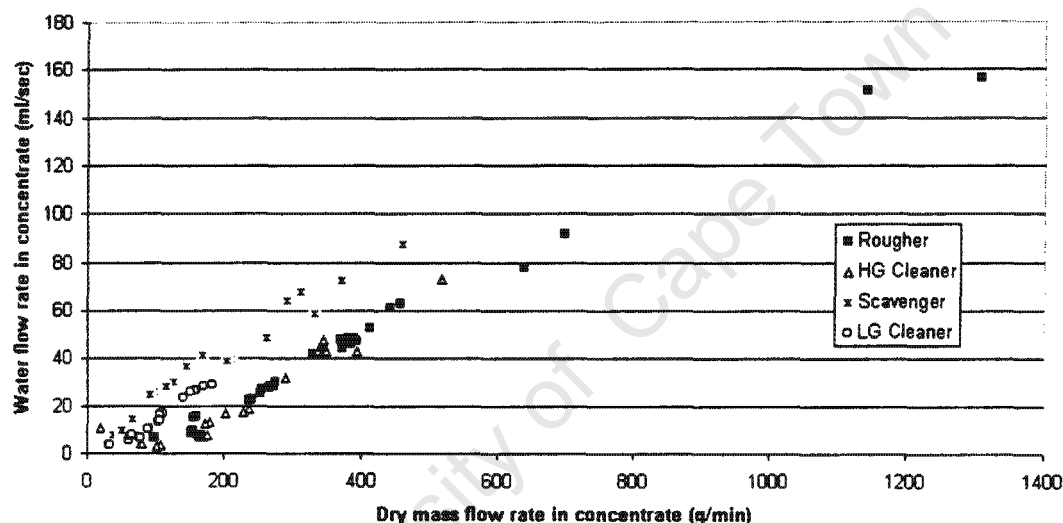


Figure 5.2: The relationship between the concentrate water flow rate and dry mass flow rate from the pilot plant froth characterisation tests (test 2, chapter 4)

One problem with using the model presented in equation 5.1 is that it is specific to an individual flotation cell in the circuit because it does not relate this relationship to any meaningful parameter such as the froth stability (which is a function of the flotation cell chemistry, solids type and concentration etc.). Using it to predict the water recoveries of all the flotation cells in the circuit would require either conducting froth characterisation tests in every flotation cell of the circuit (16 for the pilot plant, 32 parameters) or otherwise regressing these parameters to the experimental data of the pilot plant assuming the relationship (defined by the first flotation cell) does not change considerably down a bank. This is feasible as can be seen in figure 5.3, which shows how this relationship changes down the rougher bank. However, the primary drawback of this approach is that it does not provide a basis to account for the simulation of changes in cell operating conditions during simulation, such as gas flow rate and froth height in particular.

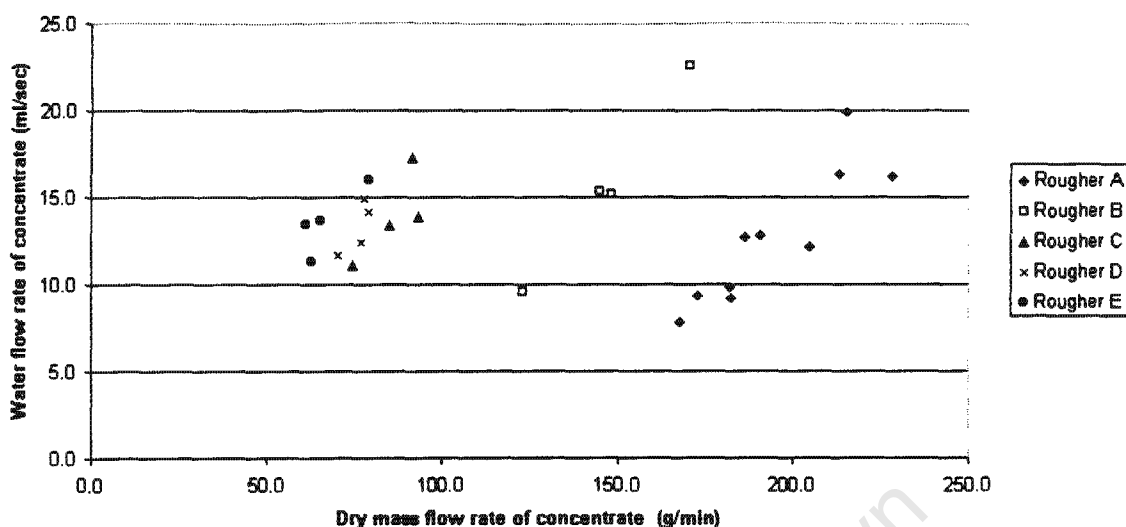


Figure 5.3: The change in the dry mass versus water flow rate trend down the rougher bank (test 1, chapter 4)

Most water models in the literature ignore the importance of the froth phase characteristics on the water recovery from a cell, (Moys, 1978), (Mehrotra, 1988). The importance of the froth height and aeration rate has been realised and utilised on flotation plants for many years, but the quantification of the effects of these variables in modelling studies has only recently received attention, (Engelbrecht and Woodburn, 1975). The froth height and air flow rate are the major controlling parameters used to vary the mass pull and therefore the water recovery from a flotation cell, (Langberg, 1988). Also ignored is the contribution of chemical effects on the water recovery from a flotation cell. Langberg, (1988) has suggested that the rate at which water is carried into the froth depends primarily upon the rate at which air bubbles pass through the froth/pulp interface and the size of the air bubbles. The rate at which water drains out of the froth depends upon the viscosity, the residence time of the bubbles, and their stability. The residence time depends upon the dimensions of the froth and the aeration rate. The residence time of bubbles in the froth phase is normally sufficient to allow much of the interstitial water entrained with the bubbles as they enter the froth to drain out again. As the liquid films separating the bubbles become thinner, coalescence of the bubbles becomes more likely. The stability of the froth depends upon the type and concentration of surfactant (frother) used to stabilise the froth, the concentration and surface properties of particles in the froth, as well as the stresses induced as the froth moves, (Gaudin, 1957). Klassen and Mokrowsov (1963) recognised that the particle size, shape and hydrophobicity as well as the concentration of particles (mineralisation) are important variables affecting the stability, and water content of flotation froths. Szatkowski (1987) studied the factors influencing the behaviour of flotation froths and pointed out that the frother concentration affects the rate of coalescence in the froth in two ways, influencing the initial bubble size and the froth stability. Unfortunately no

detailed model accurately incorporates frother concentrations around a flotation plant. Frother molecules are surface active and undergo separation throughout a flotation plant, with a concentration in the cleaner banks. This can be expected to affect the water recovery kinetics in these different banks.

5.1.2 Preliminary water model development

This section briefly describes an attempt to model the water recovery kinetics of the pilot plant using the approach of Scrimgeour et al (1970). He assigned a first order rate equation to water but did not explicitly include a froth model into this model that could account for the effects of the froth zone as described in section 5.1.1. An assumption was made that the froth model developed for all the other mineral kinetic models (as a function of concentrate froth residence time) could be used to include the froth kinetics and can be included into a first order equation to describe the water recovery in exactly the same way it was defined for the other mineral models defined previously. In this way, the water model would include froth stability effects as well as relating the water froth recovery to an exponential decay of some sort that is a function of the residence times in the froth phase depending on the froth stability parameter. Equations 5.2 to 5.4 show the proposed form of this model.^{5.1}

$$k_{w,k}^o = P_w S_{b,k} R_{f,w,k} \dots\dots\dots 5.2$$

where $k_{w,k}^o$ is the overall rate constant for water in flotation cell k that can be calculated directly from experimental data by manipulating equation 5.3, (first order recovery expression for water) knowing the experimentally determined overall water recovery ($R_{w,k}^o$) and the measured cell residence time in flotation cell k, ($\tau_{c,k}$).

$$R_{w,k}^o = \frac{k_{w,k}^o \tau_{c,k}}{1 + k_{w,k}^o \tau_{c,k}} \dots\dots\dots 5.3$$

where $R_{f,w,k}$ defining the froth recovery of water in flotation cell k can be calculated from

$$R_{f,w,k} = (1 - \alpha_w) \exp(-\beta_{k,w} \tau_{f,k}) + \alpha_w \dots\dots\dots 5.4$$

Manipulating equation 5.2 it is possible to solve for $P_w R_{f,w,k}$. Plotting $k_{w,k}^o / S_{b,k}$ versus froth residence time of the concentrate flow rate (volume froth/concentrate volume flow rate) the form of equation 5.4 can be determined and the parameters can be regressed. Figure 5.4 shows the exponential decrease in the froth recovery with froth residence time for the froth characterisation tests (test 2, chapter 4). Figures 5.5 and 5.6 show the resulting model fit of the water flow rates to the feed and froth characterisation tests

^{5.1} the parameters in this model are exactly as defined previously, equation 3.9

(test 1 and 2, chapter 4). The fit was very good and shows that the form of this model is capable of predicting the water flow rates for the pilot plant cells.

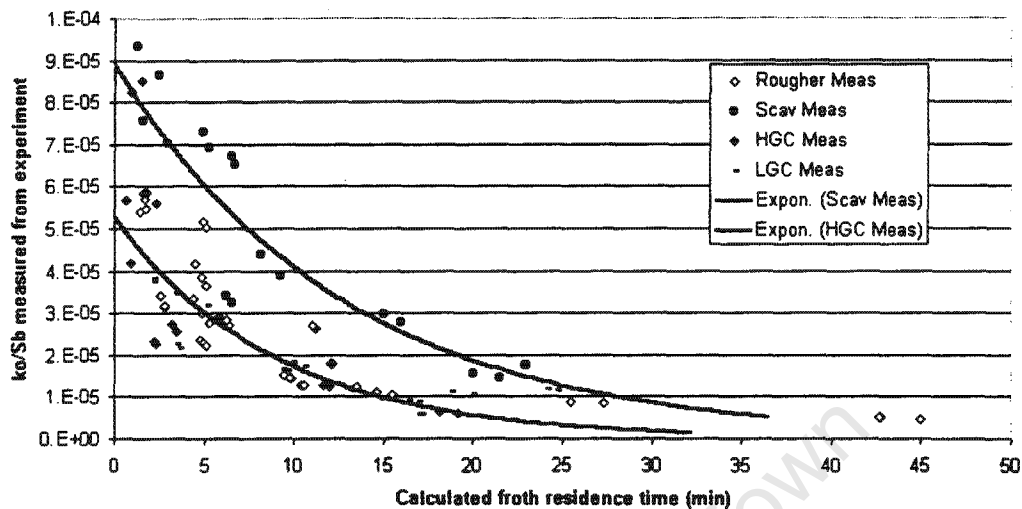


Figure 5.4: The exponential decay of the $k_w/S_{b,k}$ or $(P_w R_f)$ with concentrate froth residence time for the froth characterisation tests (test 2, chapter 4)

The problem with this approach arises when it is included into a simulation algorithm. The basic objective of a water model is the prediction of the flow rate of water in the concentrate, and when this is included in a simulation algorithm the stability of the algorithm becomes highly unpredictable. This is possibly because of the dependencies of convergence of the froth and cell residence times that rely heavily on the accuracy of the water model. A water model based on the froth residence time of air would be less problematic for simulation purposes as the water recovery would then be directly related to cell operating conditions, (Savassi, 1998).

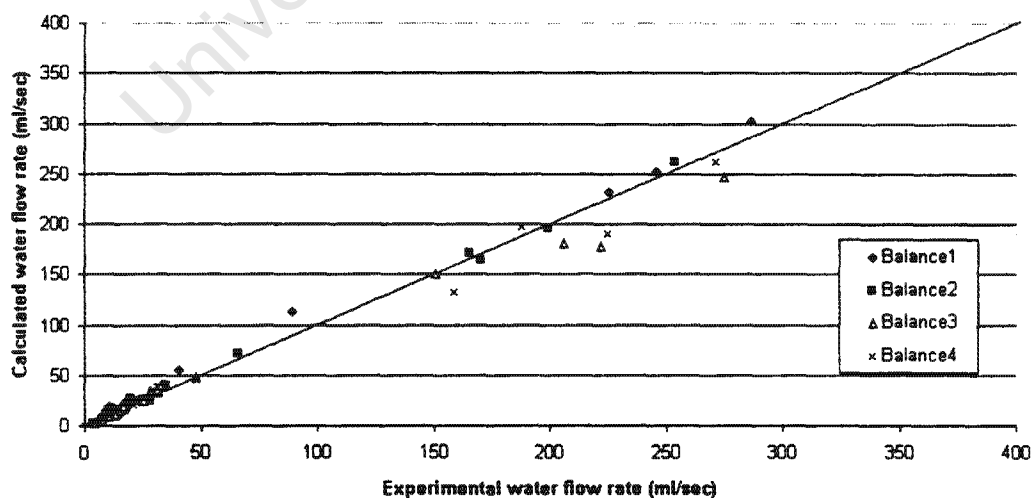


Figure 5.5: Preliminary water model fit to the feed characterisation test (test 1, chapter 4)

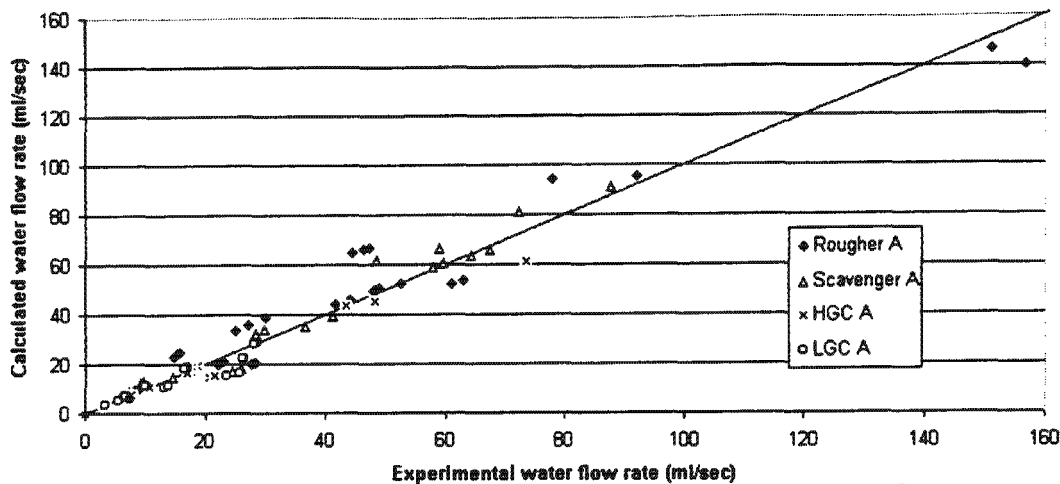


Figure 5.6: Preliminary water model fit to the froth characterisation test (test 2, chapter 4)

Using this froth model in a simulation algorithm requires two variables to converge (froth residence time as well as the cell pulp residence time). Having a water froth model that does not rely on an iteration sequence to obtain a solution (froth residence time), and is based exclusively on cell operating parameters increases the stability of the algorithm and converges a lot more rapidly to a solution. The next section describes such a model.

5.1.3 Water model based exclusively on cell operating parameters

The basis of this model is identical to the preliminary model except an alternative froth model based on the froth residence time of air is used as this variable is based exclusively on cell operating parameters which facilitates a more stable simulation algorithm. The assumptions in this model are:

- Water recovery from the pulp zone (excluding froth) is a first order process, affected by the cell residence time of the pulp and the bubble surface area flux in a flotation cell, assuming a dependency of bubble size with aeration rate, (Engelbrecht and Woodburn, 1975).
- Only the froth phase is affected by changes in chemical conditions between different banks of a circuit. This is not strictly true as the pulp phase water recovery is also affected by changes in the pulp chemistry, but these changes are incorporated into the change of the bubble size in the S_b model. This assumption considers the froth phase stability and mobility to be the dominant reason for changes in water recoveries induced by chemical effects, (Engelbrecht and Woodburn, 1975), and that these effects override the small change in pulp water recovery due to pulp chemistry.

- The froth recovery of water is affected by the stability of the froth in different banks of a flotation circuit as well as by the froth residence time of air in an individual flotation cell in the circuit. It is assumed that the stability of the froth is constant for a bank of flotation cells but is different in different banks within a circuit. This assumption is also not strictly true as the concentration of frother and floatable particles is depleted down a bank of flotation cells influencing the stability of the froth. The differences down a bank are small and it is assumed that the same trends are however still maintained allowing the simulator to at least approximate the effects of changing the air flow rates and froth heights in the different cells constituting a bank. If enough experimental data is available to define the froth stability effects for each individual cell across a bank then the accuracy of the model would be increased, but it is believed that this level of accuracy is not necessary to still obtain relatively reliable simulations.

Equation 5.5 describes the true gas residence time of air in the froth. Langberg (1988) pointed out that the value of the gas hold-up varies with froth height and that the composition of the froth varies from a bubble bed to a cellular foam with height. Steiner et al (1977) accounted for the effects of foam removal on the liquid hold-up profile in the froth. According to this model, most of the film drainage takes place within a few bubble diameters of the froth/pulp interface. Thereafter, it is argued that the liquid hold-up does not change rapidly with height, and the contribution of the films to the liquid overflow rate can be safely included to be constant. Therefore, the assumption is made in the development of the water model that the interstitial water in the plateau borders of the froth is sufficiently drained throughout the entire froth depth. This assumption in effect amounts to a constant gas hold-up with depth in the froth, (Shih and Lemlich, 1971), (Haas and Johnson, 1964).

$$\tau_{air} = \frac{\epsilon_g V_f}{Q_{air}} \dots\dots\dots 5.5$$

where

- τ_{air} is the residence time of air in the froth (min)
 ϵ_g is the gas hold-up in the froth zone of the flotation cell
 V_f is the effective volume of the froth (l)
 Q_{air} is the volumetric air flow rate into the flotation cell (l/min)

The data from the froth characterisation test was again used to calculate k_w°/S_b , and the corresponding residence time of air was calculated from the experimental data (assuming ϵ_g was unity). Figure 5.7 shows the resulting trends obtained for the froth characterisation test of each bank when plotting k_w°/S_b versus τ_{air} , as a function of the froth volume used during the tests. An exponential decay relationship is observed as a function of froth volume in each cell. Equation 5.6 is proposed to describe the variation in k_w°/S_b as a function of τ_{air} for different froth volumes.

$$R_{f,w,k} = \Omega_k \exp(\sigma_k * V_{f,k} - \chi_k * \tau_{air,k}) \dots\dots\dots 5.6$$

where χ_k , Ω_k and σ_k are empirically determined constants for a bank in the flotation circuit.

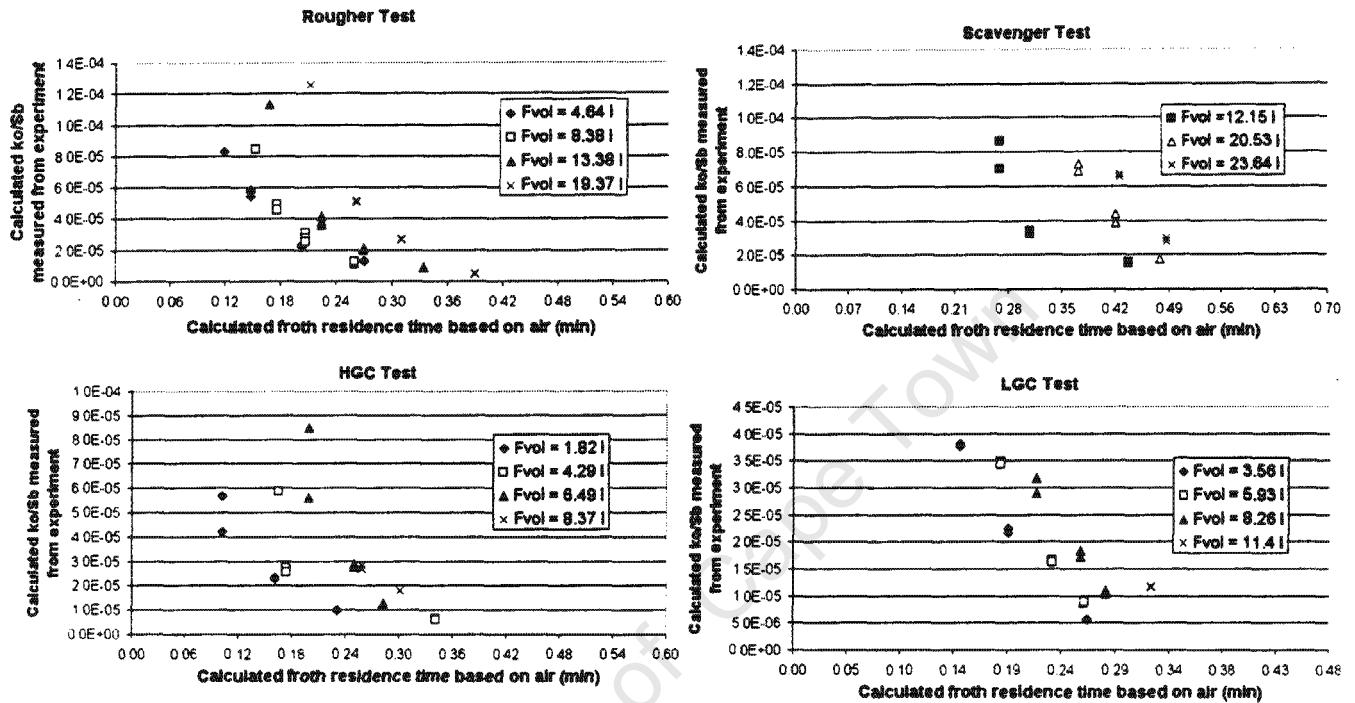


Figure 5.7: The trends observed when plotting k^o_w/S_b with τ_{air} for the froth characterisation tests, at different froth volumes

The form of this model (equation 5.6) was obtained from analysing the parameter variation with froth volume after fitting an exponential function to the above curves in figure 5.7. Moys (1978) studied the plug flow behaviour of flotation froths and used the bubble residence time in a plug flow model describing the froth drainage mechanism. Cooper (1966) as well as Cutting and Devenish (1975) have presented distributed models for the froth phase. Each of these models predicts an approximately exponential decrease in the recovery rate with increasing froth height. Langberg (1988) points out that the complex equations resulting from this work were abandoned in favour of much simpler linear transfer functions. In order to avoid the need to obtain complex information such as liquid hold-up, surface tension, particle size, viscosity, bubble lamella thickness, density (just to mention a few) it was believed that the model presented in equation 5.6, although purely empirical would suffice for the purpose of this investigation. A boundary condition placed on the model was that $R_{fw} < 1$ for all froth residence times of air. Table 5.1 shows the regressed model parameters for the data presented in figure 5.7.

Table 5.1: The regressed empirical water parameters for the pilot plant

PARAMETER	ROUGHER	SCAVENGER	HGC	LGC
Ω_k	3.00	2.55	1.67	1.86
σ_k	0.08	0.16	0.32	0.21
χ_k	14.03	13.03	17.06	16.92

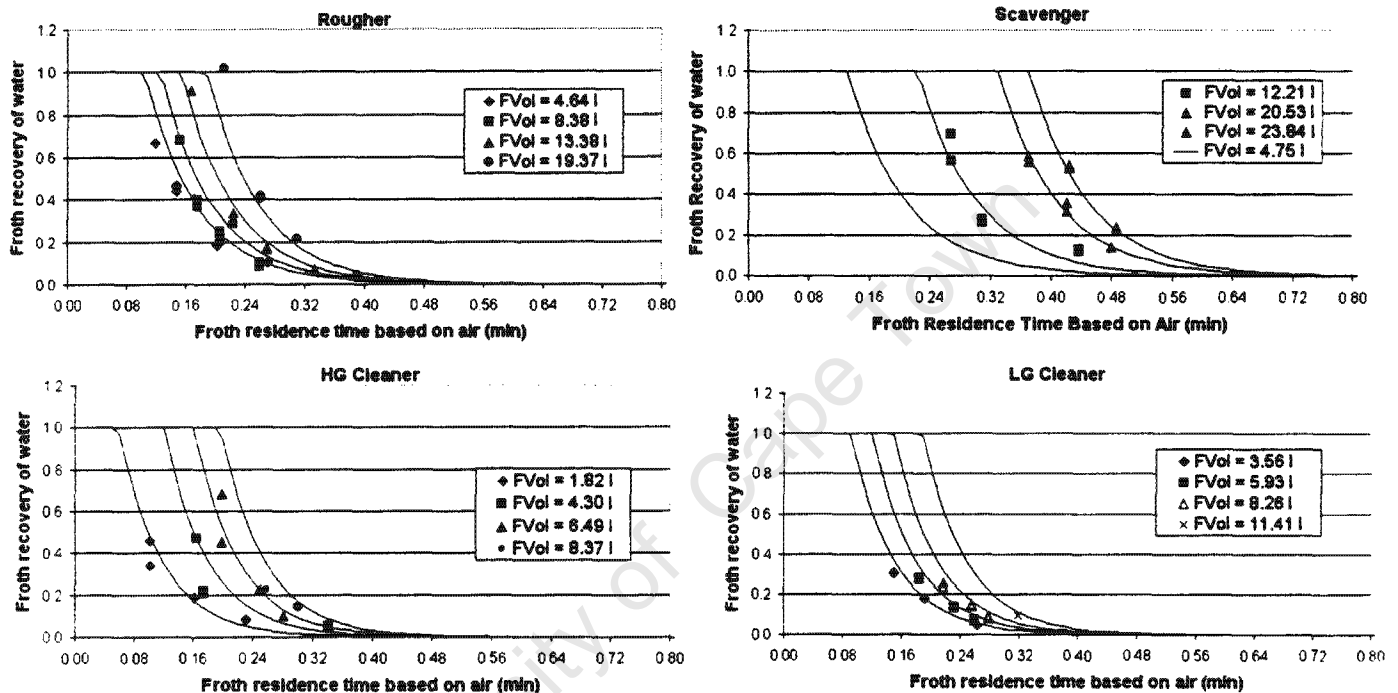
**Figure 5.8: The water froth recovery model fit to the experimental data of the pilot plant (test 2)**

Figure 5.8 shows the form of this model and figure 5.9 the fit to the experimental data obtained from the pilot plant tests. The correlation between measured and calculated values for all the data points obtained from every bank was reasonably good ($R^2 = 0.85$). The parameters in this model are believed to incorporate froth mobility and stability effects for different banks in the circuit. Even though this model is largely empirical it appears to be well able to describe the water recovery in the different banks for various combinations of froth heights and volumetric flow rates of air. It is relatively simple and is easy to implement as measuring water flows is quick and cheap (no assays are required) and the model provides a means to account for the effects of the operating variables on cell performance.

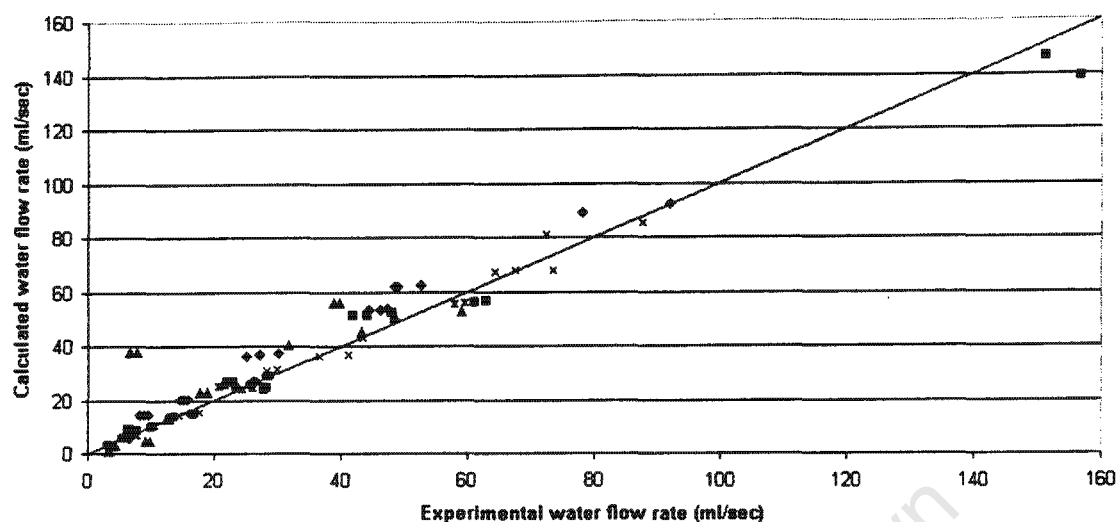


Figure 5.9: The water flow rate fit to experimental data from the pilot plant tests.

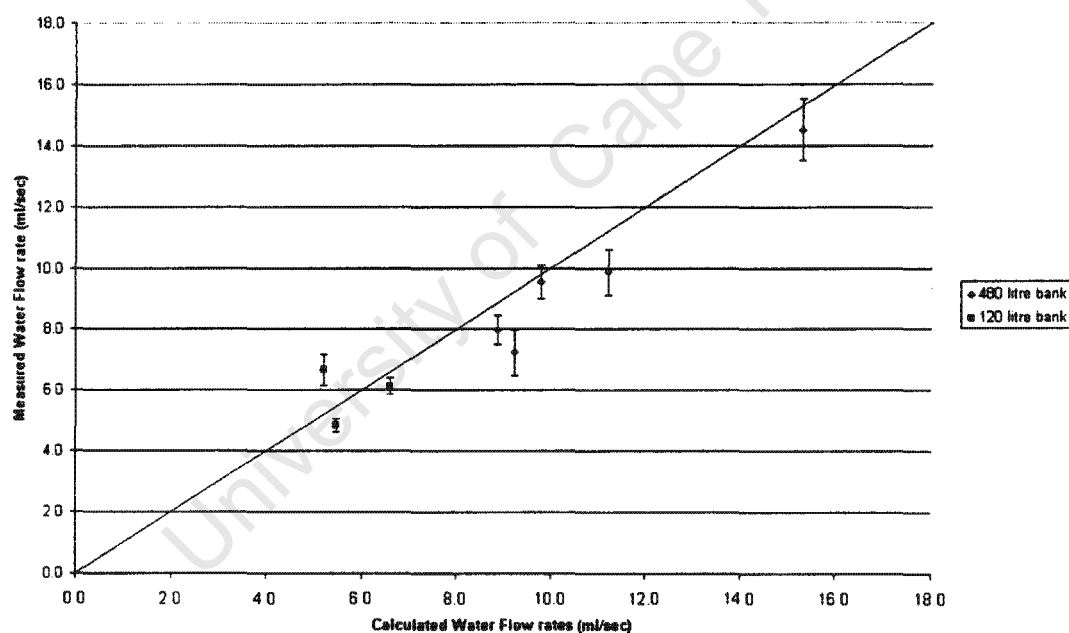


Figure 5.10: Water model fit for banks of two different sizes using the same model parameters (test 4, chapter 4)

The variables in this model ($\tau_{air,k}$ and $V_{t,k}$) were chosen in order to incorporate scale-up effects as they are not related to a specific dimension of a cell (such as froth height). In order to test the scale-up assumption of this model the data obtained from the scale-up test described in chapter 4 (test 4) was used to test the prediction of this model. A constant set of parameters were used to predict the water recoveries in the 120 litre and 480 litre banks. Figure 5.10 shows the resulting water flow rate fit to the experimental data of test 4. A reasonable prediction could be made for the cells of

different sizes. Obviously there exists a limit as to how far this is true as froth transportation distances to the cell lip as well as froth cell lip cross sectional areas start becoming important variables as the flotation cell size gets larger. However, it is apparent that varying the size of a flotation cell within certain limits is possible with this model. However, extrapolating results from pilot plant cells to industrial cells could represent a problem. This was considered beyond the scope of this study.

5.1.4. Conclusions drawn from the empirical water model development

There is considerable scope for further work in the area of water recovery modelling in flotation, particularly from a more mechanistic perspective. It must be stressed that the purpose of the model presented above was to obtain a relatively simple, robust method of predicting the water kinetics from the pilot plant cells as a function of the major control variables such as froth height and volumetric flow rates of air. While this model is purely empirical, it appears capable of providing reasonably accurate prediction of the water kinetics for a cell in a fixed position in a flotation circuit. It should be noted that attempting to simulate alternative flotation circuits with the derived parameters could well give less reliable water flow rate predictions.

5.2 Simulation algorithm development

An empirical water model has been presented and its importance to the entire simulation model has been highlighted. This section presents an algorithm that was developed for simulating the pilot plant facilities. This algorithm has been developed specifically for UG2 ores but the form of this algorithm should be generic to any ore treated by flotation. The section starts by presenting a detailed step by step simulation algorithm for the pilot plant. Followed by a brief summary.

5.2.1 Step by step simulation algorithm

This section presents a complete 'recipe' for simulating the flotation cells of the pilot plant. Included are all the model parameters, inputs and calculated variables in the model. Figure 5.11 is a schematic of the simulation algorithm that can be used to understand the steps in the following description.

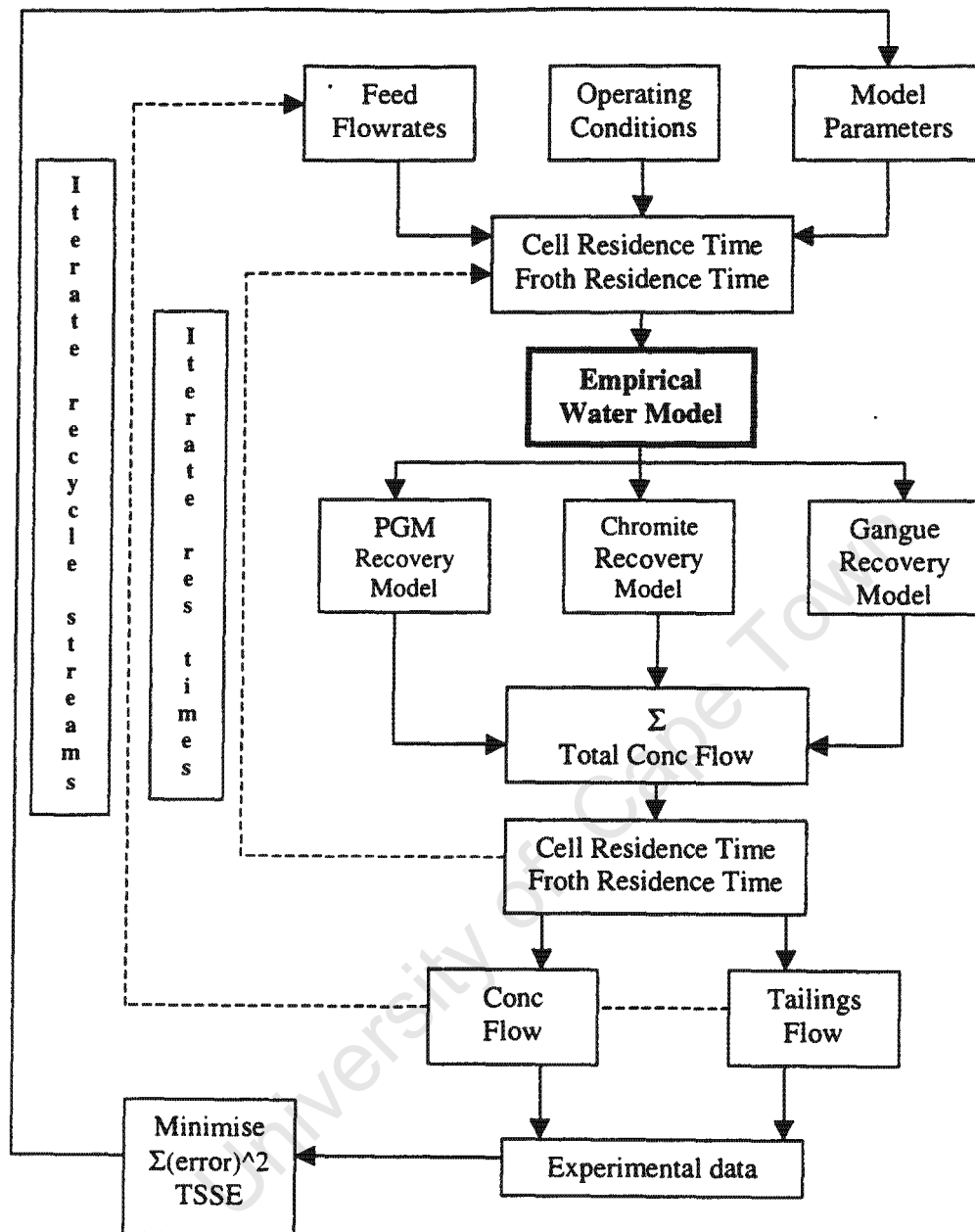


Figure 5.11: Schematic of the simulation algorithm

5.2.1.1. SIMULATOR INPUTS

The input values in this section give information about the flotation cells being simulated. These are usually design variables that are measured on the circuit and need to be determined before simulation can proceed. This information is used to calculate the sub-variables in the models.

A. OPERATING CONDITIONS

- Total volumetric air flow rate (l/min) (Q_{air})
- Impeller Speed (1/sec) (N)
- Cell dimensions
 - Height (cm) (H)
 - Depth (cm) (D)
 - Length (cm) (L)
- Stator Shaft Radius (cm) (S_r)
- Impeller/Stator Height (cm) (I_H)
- Impeller/Stator Radius (cm) (I_R)
- Froth Height (cm) (FH)
- Gas hold-up in the pulp (%) (ϵ_g)
- Volume Stator (l) (VS)
- Height of Imp/Stator above cell floor (cm) (IT)

B. FEED INPUTS

The feed inputs to the flotation cell include the feed flow rates of all the individual minerals and gangue species, the size distribution of these minerals and the quantity of water (% solids).

- Total dry feed mass flow rate (kg/hr) ($F^{o DM, feed}$)
- Percentage Solids (%) (S_{feed})
- PGM feed grade (g/t) ($G_{PGM, feed}$)
- Chromite feed grade (%) ($G_{Cr, feed}$)
- Specific gravity of combined feed (kg/l) (SG_{feed})
 - of PGM (kg/l) (SG_{PGM})
 - of Chromite (kg/l) (SG_{Cr2O3})
 - of Gangue (kg/l) (SG_{gangue})
- Size distribution of combined feed material, chromite and gangue (cumulative fraction weight undersize in size interval (d) of mineral (j)) for the following size fractions:

		<u>Combined</u>	<u>Chromite</u>	<u>Gangue</u>
+212 μ m	=	Y _{+212,DM}	Y _{+212,Cr}	Y _{+212,ga}
-212 μ m +150 μ m	=	Y _{-212,DM}	Y _{-212,Cr}	Y _{-212,ga}
-150 μ m +106 μ m	=	Y _{-150,DM}	Y _{-150,Cr}	Y _{-150,ga}
-106 μ m + 75 μ m	=	Y _{-106,DM}	Y _{-106,Cr}	Y _{-106,ga}
-75 μ m +53 μ m	=	Y _{-75,DM}	Y _{-75,Cr}	Y _{-75,ga}
-53 μ m + 38 μ m	=	Y _{-53,DM}	Y _{-53,Cr}	Y _{-53,ga}
-38 μ m + 25 μ m	=	Y _{-38,DM}	Y _{-38,Cr}	Y _{-38,ga}
-25 μ m + 10 μ m	=	Y _{-25,DM}	Y _{-25,Cr}	Y _{-25,ga}
-10 μ m	=	Y _{-10,DM}	Y _{-10,Cr}	Y _{-10,ga}

C. FLOTATION AND ENTRAINMENT PARAMETERS

The parameters describing the rates of flotation (floatability), the rates of entrainment as a function of particle size (entrainability), as well as the froth effects are included below.

PGM Parameters

- Fraction fast floating PGM in the feed $(m_{f,PGM})$
- Fraction slow floating PGM in the feed $(m_{s,PGM})$
- Fraction non floating PGM in the feed $(m_{o,PGM})$
- PGM fast floatability $(P_{f,PGM})$
- PGM slow floatability $(P_{s,PGM})$
- PGM entrainment parameter (E_{PGM})
- PGM froth stability parameter in bank or cell k $(\beta_{PGM,k})$
- PGM froth drainage parameter (α_{PGM})

Chromite Parameters

- Chromite entrainment parameter (E_{cr})
- Chromite drainage parameter (δ_{cr})
- Chromite floatability (P_{cr})
- Fraction activated chromite in feed (ϕ_{cr})
- Chromite froth stability parameter in unit or bank k $(\beta_{cr,k})$
- Chromite froth drainage parameter (α_{cr})

Gangue Parameters

- Gangue entrainment parameter (E_{Ga})
- Gangue drainage parameter (δ_{Ga})
- Gangue fast floatability $(P_{f,Ga})$
- Gangue slow floatability $(P_{s,Ga})$
- Fraction activated gangue in feed (ϕ_{Ga})
- Fraction fast float gangue in feed $(m_{f,Ga})$
- Gangue froth stability parameter in unit or bank k $(\beta_{Ga,k})$
- Gangue froth drainage parameter (α_{Ga})

Water Parameters

- Water floatability (P_w)
- Water empirical froth parameters (Ω_k)
 (σ_k)
 (χ_k)

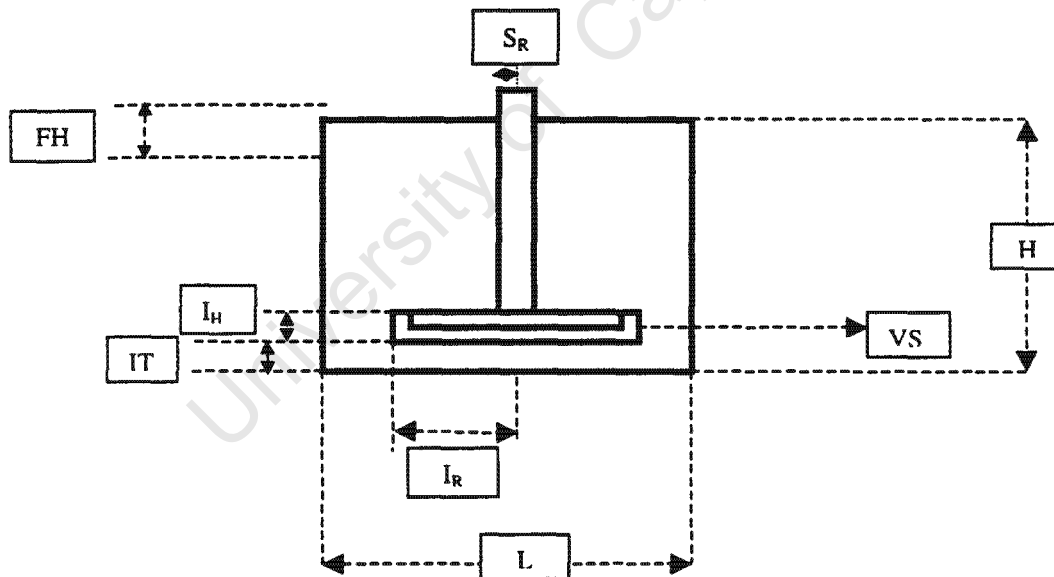
Where "k" refers to an individual flotation cell or bank of flotation cells in the circuit.

Bubble surface area flux prediction parameters

- *Bubble flux slope parameter* (a)
- *Impeller peripheral speed exponent parameter* (b)
- *Superficial gas velocity exponent parameter* (c)
- *Impeller aspect ratio exponent parameter* (d)
- *80% passing size exponent parameter* (e)

5.2.1.2. Derived model parameters

Using the operating conditions of section 5.2.1.1A the following cell characteristics are calculated to account for the volume occupied by the impeller/stator for every flotation cell in the simulated circuit. To obtain the effective operating volume of the pulp and froth phases, the dimensions of the impeller/stator have to be included. The froth depth could not be varied independently of the pulp volume and therefore the effective pulp volume is a function of the froth height used during the simulation. The gas hold-up (ϵ_g) in equation 5.12 refers to the fraction of the total pulp volume (excluding impeller/stator) occupied by the air bubbles in the flotation pulp. This parameter was not measured during the experiments due to the small cross sectional area of the flotation cells on the pilot plant and was therefore approximated as being 10% in the rougher banks and 15% in all the cleaner banks.



Legend	Description	80 litre flotation cell	30 litre flotation cell
I_H	Height of impeller/stator	8 cm	6 cm
I_T	Total height of imp/stat above cell floor	13 cm	11.5 cm
I_R	Radius of impeller	11.8 cm	8.5 cm
L	Flotation cell length	42 cm	32 cm
V_S	Volume stator	840 cm ³	270 cm ³
H	Flotation cell depth	42 cm	32 cm
FH	Froth depth	variable	variable
S_R	Stator shaft radius	4.62 cm	3.74 cm

Figure 5.12: Flotation cell and impeller stator dimensions for the pilot plant cells

- Total water flow in feed (ml/sec)

$$Q_{\text{water feed}} = \left(\frac{F^{\circ}_{DM, \text{feed}}}{S_{\text{feed}}} * 100 - F^{\circ}_{DM, \text{feed}} \right) * \frac{1000}{3600} \dots\dots\dots 5.7$$

- Feed volumetric flow rate (l/min)

$$Q_{\text{total feed}} = \frac{F^{\circ}_{DM, \text{feed}}}{60 * SG_{\text{feed}}} + Q_{\text{water feed}} * \frac{60}{1000} \dots\dots\dots 5.8$$

- Pulp shaft volume (l)

$$VS_{\text{pulp}} = \frac{\pi * S_R^2 * (H - IT - I_H - FH)}{1000} \dots\dots\dots 5.9$$

- Froth shaft volume (l)

$$VS_{\text{froth}} = \frac{\pi * S_R^2 * (FH)}{1000} \dots\dots\dots 5.10$$

- Impeller/Stator volume (l)

$$V_{\text{imp / stat}} = \frac{\pi * I_R^2 * I_H + VS}{1000} \dots\dots\dots 5.11$$

- Effective pulp cell volume (excluding volume occupied by impeller/stator) (l)

$$EPV = \left[\frac{D * L * (H - FH)}{1000} - VS_{\text{pulp}} - V_{\text{imp / stat}} \right] * (1 - \epsilon_g) \dots\dots\dots 5.12$$

- Effective froth volume (excluding volume occupied by impeller shaft) (l)

$$EFV = \frac{D * L * FH}{1000} - VS_{\text{froth}} \dots\dots\dots 5.13$$

- Bubble surface area flux predictor (1/min)

$$S_b = a * (N_s)^b * \left(\frac{Q_{\text{air}}}{L * D} \right)^c * (A_s)^{-d} * (P_{80})^{-e} \dots\dots\dots 5.14$$

where: P_{80} = derived from overall feed size distribution

$A_s = 2l_R/l_H$ (impeller aspect ratio)

$N_s = 2\pi l_R N$ (impeller peripheral speed)

5.2.1.3. Mineral entrainment characterisation

The entrainment of gangue and chromite (mineral j) with water is size dependent. The following equations are used to calculate the entrainability parameters for the different size fractions (d') to infer the difference in entrainment with particle size class.

$$\bullet \quad \overline{s}_{j,d'} = \frac{d'_{j,d'.\max} + d'_{j,d'.\min}}{2} \dots\dots\dots 5.15$$

where above is the arithmetic mean size of mineral (j) in size class (d') (μm)

(max) refers to upper size in size class (d') (μm)

(min) refers to lower size in size class (d') (μm)

$$\bullet \quad ADJ_{j,d'} = 1 - \frac{\ln\left(\frac{1}{\delta_j}\right)}{\exp\left(\frac{s_{j,d'}}{\xi_j}\right)} \dots\dots\dots 5.16$$

$$\bullet \quad E_{j,d'} = \frac{2}{\exp\left(2.292 * \left(\frac{s_{j,d'}}{\xi_j}\right)^{ADJ_{j,d'}}\right) + \exp\left(-2.292 * \left(\frac{s_{j,d'}}{\xi_j}\right)^{ADJ_{j,d'}}\right)} \dots\dots\dots 5.17$$

The output of the above calculation is a matrix with the entrainability values of the nine size fractions given in the feed input section (1B), for both chromite and gangue particles in the feed.

5.2.1.4. Mineral recovery calculations and iteration sequence

This section describes in detail the iteration algorithm used in calculating the PGM, chromite and gangue unit cell recoveries from an individual flotation cell in a circuit. Once these are obtained it describes the iteration of the flotation cell residence times.

A Initial approximation of flotation cell and froth residence times (based on concentrate flow rate)

An initial approximation of the cell residence time and froth residence time (based on concentrate flow) is calculated based on the feed flow rates to the flotation cell. These are used to calculate an initial approximation of the recovery of PGM, chromite and gangue. The initial approximation of the froth residence time (based on concentrate flow rate) used in the PGM, chromite and gangue models are calculated assuming a 1% volume pull based on feed.

- Initial approximation of cell residence time (min)

$$\tau_{c_{initial}} = \frac{EPV}{Q_{total_{feed}}} \dots\dots\dots 5.18$$

- Initial approximation of froth residence time based on concentrate flow rate (min)

$$\tau_{f_{initial}} = \frac{EFV}{Q_{total_{feed}} * 0.01} \dots\dots\dots 5.19$$

B Initial approximation of water flow rate in concentrate

Once all the feed inputs are known, it is possible to calculate the water recovery independently for a particular flotation cell in a particular bank as a function of the operating conditions. The calculated water recovery is used in the mineral and gangue recovery equations taking into account the entrainment of these particles with water.

- Calculate the froth residence time of air (min)

$$\tau_{air} = \frac{EFV}{Q_{air}} \dots\dots\dots 5.20$$

- Calculate the water froth recovery

$$R_{f_{water}} = \Omega \exp(\sigma * EFV - \chi * \tau_{air}) \dots\dots\dots 5.21$$

- Calculate the initial approximation of the water recovery

$$R_{w_{initial}} = \frac{P_w S_b \tau_{c_{initial}} R_{f_{water}}}{1 + P_w S_b \tau_{c_{initial}} R_{f_{water}}} \dots\dots\dots 5.22$$

- Calculate the initial approximation of the water flow rate in the concentrate

$$Q_{water,conc_{initial}} = R_{w_{initial}} * Q_{water,feed} \dots\dots\dots 5.23$$

C Initial approximation of the PGM flow rate in the concentrate

- Calculate the total PGM flow rate in the feed (mg/hr)

$$F_{PGM, feed}^o = G_{PGM, feed} * F_{DM, feed}^o \dots\dots\dots 5.24$$

- Calculate each PGM floatability component flow rate feeding the flotation cell (mg/hr)

$$F_{i, PGM, feed} = F_{PGM, feed}^o * m_{i, PGM, feed} \dots\dots\dots 5.25$$

where floatability components = i

- Calculate the initial approximation of the PGM froth recovery

$$R_{f, PGM, initial} = (1 - \alpha_{PGM}) \exp(-\beta_{PGM, k} \tau_{f, k, initial}) + \alpha_{PGM} \dots\dots\dots 5.26$$

where k = particular flotation cell or bank

- Calculate the initial approximation of the recovery of each PGM floatability component (i)

$$R_{i, PGM, k, initial} = \frac{P_{i, PGM} S_{b, k} \tau_{c, k, initial} R_{f, PGM, k, initial}}{1 + P_{i, PGM} S_{b, k} \tau_{c, k, initial} R_{f, PGM, k, initial}} + [1 - \frac{P_{i, PGM} S_{b, k} \tau_{c, k, initial} R_{f, PGM, k, initial}}{1 + P_{i, PGM} S_{b, k} \tau_{c, k, initial} R_{f, PGM, k, initial}}] * R_{w, k, initial} * E_{PGM} \dots\dots\dots 5.27$$

- Calculate the initial approximation of each PGM floatability component flow rate in the concentrate (mg/hr)

$$F_{i, PGM, conc, initial} = R_{i, PGM, initial} * F_{i, PGM, feed} \dots\dots\dots 5.28$$

- Calculate the initial approximation of the total PGM flow rate in the concentrate (mg/hr)

$$F_{PGM, conc, initial}^o = \sum_{i=1}^n F_{i, PGM, conc, initial} \dots\dots\dots 5.29$$

where n is the number of floatability components

D Initial approximation of the chromite flow rate in the concentrate

- Calculate the total chromite flow rate in the feed (g/min)

$$F_{Cr,feed}^{o} = F_{DM,feed}^{o} * \frac{G_{Cr,feed}}{100} * \frac{1000}{60} \dots\dots\dots 5.30$$

- Calculate the flow rate of each entrainable fraction of chromite in each size fraction (g/min)

$$F_{Cr,d',feed} = Y_{d',Cr,max} * F_{Cr,feed}^{o} * (1 - \phi_{Cr,feed}) - Y_{d',Cr,min} * F_{Cr,feed}^{o} * (1 - \phi_{Cr,feed}) \dots\dots\dots 5.31$$

where $Y_{d',Cr,max}$ = fraction smaller than top size of size class (d')

$Y_{d',Cr,min}$ = fraction smaller than bottom size of size class (d')

- Calculate the flow rate of the floatable chromite fraction in the feed (g/min)

$$F_{Cr,act,feed} = F_{Cr,feed}^{o} * \phi_{Cr,feed} \dots\dots\dots 5.32$$

- Calculate the initial approximation of the froth recovery of the floatable fraction of chromite

$$R_{f,Cr,k,initial} = (1 - \alpha_{Cr}) \exp(-\beta_{Cr,k} * \tau_{f,k,initial}) + \alpha_{Cr} \dots\dots\dots 5.33$$

where k = particular flotation cell or bank

- Calculate the initial approximation of the recovery of each entrainable chromite size fraction (d')

$$R_{Ent,Cr,d',initial} = R_{w,initial} * E_{Cr,d'} \dots\dots\dots 5.34$$

- Calculate the initial approximation of the recovery of the floatable chromite fraction

$$R_{act,Cr,initial} = \frac{P_{Cr} S_b \tau_c R_{f,Cr,k,initial}}{1 + P_{Cr} S_b \tau_c R_{f,Cr,k,initial}} \dots\dots\dots 5.35$$

- Calculate the initial approximation of the flow rate of entrainable and floating fractions of chromite in the concentrate (g/min)

$$F_{d',conc,initial} = R_{EntCr,d',initial} * F_{Cr,d',feed} \dots\dots\dots 5.36$$

$$F_{act,Cr,conc,initial} = R_{act,Cr,initial} * F_{Cr,act,feed} \dots\dots\dots 5.37$$

- Calculate the initial approximation of the total chromite flow rate in the concentrate (g/min)

$$F_{Cr,conc}^{o,initial} = \sum_{i=1}^{d'} F_{Cr,d',conc}^{o,initial} + F_{act,Cr,conc}^{o,initial} \dots\dots\dots 5.38$$

E Initial approximation of the gangue flow rate in the concentrate

- Calculate the total gangue flow rate in the feed (g/min)

$$F_{Ga,total,feed}^{o} = F_{DM,total,feed}^{o} * \frac{1000}{60} - F_{Cr,total,feed}^{o} - \frac{F_{PGM,total,feed}^{o}}{1000 * 60} \dots\dots\dots 5.39$$

- Calculate the flow rate of each entrainable fraction of gangue in each size fraction (d') in the feed (g/min)

$$F_{Ga,d',feed} = Y_{d',Ga,max} * F_{Ga,feed,total}^{o} * (1 - \phi_{Ga,feed}) - Y_{d',Ga,min} * F_{Ga,feed,total}^{o} * (1 - \phi_{Ga,feed}) \dots\dots\dots 5.40$$

where $Y_{d',Ga,max}$ = fraction smaller than top size of size class (d')

$Y_{d',Ga,min}$ = fraction smaller than bottom size of size class (d')

- Calculate the flow rate of the floatable gangue fraction in the feed (g/min)

$$F_{act,Ga,feed} = F_{Ga,total,feed}^{o} * \phi_{act,Ga} \dots\dots\dots 5.41$$

- Calculate each gangue floatability component flow rate feeding the flotation cell (g/min)

$$F_{i,Ga,feed} = F_{act,Ga,feed} * m_{i,Ga} \dots\dots\dots 5.42$$

where floatability components = i

- Calculate the initial approximation of the froth recovery of the floatable fraction of gangue in unit k

$$R_{f,Ga,initial} = (1 - \alpha_{Ga}) \exp(-\beta_{Ga,k} * \tau_{f,k,initial}) + \alpha_{Ga} \dots\dots\dots 5.43$$

- Calculate the initial approximation of the recovery of each entrainable gangue size fraction (d')

$$R_{Ent,Ga,d',initial} = R_{w,initial} * E_{Ga,d'} \dots\dots\dots 5.44$$

- Calculate the initial approximation of the recovery of the floatable gangue fractions

$$R_{i,Ga,initial} = \frac{P_{Ga,i} S_b \tau_c R_{f,Ga,initial}}{1 + P_{Ga,i} S_b \tau_c R_{f,Ga,initial}} \dots\dots\dots 5.45$$

- Calculate the initial approximation of the flow rate of entrainable and floating fractions of gangue in the concentrate (g/min)

$$F_{Ga,d',conc,initial} = R_{Ent,Ga,d',initial} * F_{Ga,d',feed} \dots\dots\dots 5.46$$

$$F_{i,Ga,conc,initial} = R_{i,Ga,initial} * F_{act,Ga,feed} \dots\dots\dots 5.47$$

- Calculate the initial approximation of the total gangue flow rate in the concentrate (g/min)

$$F_{Ga,conc,total,initial}^o = \sum_{i=1}^{d'} F_{Ga,d',conc,initial} + \sum_{i=1}^n F_{i,Ga,conc,initial} \dots\dots\dots 5.48$$

F Recalculation of flotation cell and froth residence times

Knowing the feed volumetric flow rates and the recoveries of the individual floatability and entrainability classes it is possible to calculate the total concentrate and tailings volumetric flow rates by summing the individual volume fractions of all the recovered floatability and entrainability classes. The calculated concentrate volume flow rate is used to calculate the froth residence time (τ_f) and the tailing volume flow rate is used to calculate the cell residence time (τ_c). The method of successive approximations is used iteratively until the guessed residence times equal the calculated residence times. The final solution is the concentrate and tailings volumetric flow rates that are either products from the circuit or are recycled to other cells in the circuit.

- Calculate the initial approximation of the total solids flow rate in the concentrate (kg/hr)

$$F_{DM,conc,total,initial}^o = \frac{F_{PGM,conc,total,initial}^o}{1 * 10^6} + \frac{F_{Cr,conc,total,initial}^o * 60}{1000} + \frac{F_{Ga,conc,total,initial}^o * 60}{1000} \dots\dots\dots 5.49$$

- Calculate the initial approximation of the total volumetric flow rate of the concentrate stream (l/min)

$$Q_{conc,initial} = \frac{F_{DM,conc,total,initial}^o}{60 * SG_{feed}} + Q_{water,conc,initial} * \frac{60}{1000} \dots\dots\dots 5.50$$

- Recalculate the actual cell residence time (min)

$$\tau_{c \text{ calc}} = \frac{EPV}{(Q_{\text{total feed}} - Q_{\text{conc initial}})} \dots\dots\dots 5.51$$

- Recalculate the actual froth residence time based on concentrate flow rate (min)

$$\tau_{f \text{ calc}} = \frac{EFV}{Q_{\text{conc initial}}} \dots\dots\dots 5.52$$

- Substitute the calculated flotation cell and froth residence times into the initial approximated ones (equations 5.18 and 5.19) and iterate until both parameters converge

$$\tau_{\text{cell initial}} = \tau_{\text{cell calc}} \dots\dots\dots 5.53$$

$$\tau_{f \text{ initial}} = \tau_{f \text{ calc}} \dots\dots\dots 5.54$$

This iteration sequence was calculated in an Excel spreadsheet using a circular reference calculation sequence. The convergence criteria could be stipulated and was set to a maximum of 1000 iterations with convergence being achieved when the difference in successive approximations was less than 0.001 for both the cell and froth residence times.

5.2.2. Flotation circuit simulation and iteration sequence

When simulating a flotation circuit the recirculated flow rates change the feed flow rates to the cell that has just been iterated (if the cell is not only receiving feed to the circuit) and therefore a nested iteration of successive approximations is used to get a final solution of the mass balance around each individual flotation cell as well as for the recirculating streams in the circuit. Exactly the same steps as described in section 5.2.1 are used in calculating the flotation kinetics for all the other flotation cells in the circuit. The feed to the other cells in the circuit are calculated by summing the material received from the other flotation cells in the circuit either via a calculated tailings flow rate or concentrate flow rate, linking all the flow rates to the external feed to the circuit.

$$Feed_k = \text{External feed} + \sum_{i=1}^k \text{Concentrates}_k + \sum_{i=1}^k \text{Tailings}_k \dots\dots\dots 5.55$$

where k = flotation cell in the circuit being simulated

A global iteration sequence is employed in obtaining a solution to the circuit mass balance. The recirculating streams are re-substituted until they converge to a desired level of accuracy.

5.2.3. Obtaining best fit model parameters

There are obviously a large number of parameters in a simulation model describing a complicated system such as a flotation cell. Therefore good starting estimates of the model parameters are initially required in order to obtain a good fit to the experimental data. Once realistic estimates of the parameters are obtained from individual component modelling (as described in chapters 3 and 4), the solver optimisation subroutine in Excel can be used to improve the fit to the experimental data. Sensible parameter constraints can be included in the optimisation to improve the rate of convergence and the final solution of parameters obtained. In order to obtain the best fit model parameters, the above iteration algorithm is followed allowing the parameters of all the models to vary while minimising the sum of errors between the experimental and calculated species flows for all components in the circuit (PGM, chromite, gangue and water).

$$\text{Minimize} = \sum \left(\frac{\text{calculated} - \text{measured}}{\text{StdDev}} \right)^2 \dots\dots\dots 5.56$$

The measured experimental data that was used in driving the optimisation subroutine for the simulation of the pilot plant included:

- Total PGM flow rate (mg/hr)
- Total chromite flow rate (g/min)
- Total gangue flow rate (g/min)
- Total water flow rate (ml/sec)
- Total volumetric flow rate (l/min)

The standard deviations of all the above flow rates were measured from multiple samples obtained from the pilot plant campaign. It must be appreciated that flotation circuits are generally very complicated and unstable systems. Some of the measured operating variables such as the froth depths, air flow rates, bubble diameters and gas hold-ups also had a measured standard deviation associated with them. The optimisation subroutine was allowed to vary these operating variables about their own measured standard deviations (over and above varying the model parameters). This allowed for a much more realistic optimisation (and obviously better model fit) as in the real world, these measured operating variables were varying about their standard deviations and are not fixed to exact values.

5.2.4 Summary of the simulation algorithm

The feed inputs to the flotation cell include the feed flow rates of all the individual mineral floatability classes and gangue species, the size distribution of these minerals and the quantity of water (% solids). The operating conditions in figure 5.11 include the dimensions of the flotation cell, the power input, froth height and volumetric flow rates of air. The parameters describing the rates of flotation (floatability), the rates of entrainment as a function of particle size (entrainability), as well as the froth effects are included and are usually obtained from realistic estimates of the parameters that are obtained from individual component modelling (as described in chapters 3 and 4).

Once all the feed inputs are known, it is possible to calculate the water recovery from a particular flotation cell in a particular bank as a function of the operating conditions. The calculated water recovery is used in the mineral and gangue recovery equations taking into account the entrainment of these particles with water. An initial approximation of the cell residence time and froth residence time is calculated based on the feed flow rates to the flotation cell. These are used to calculate an initial approximation of the recovery of PGM, chromite and gangue. Knowing the feed volumetric flow rates and the recoveries of the individual floatability classes it is possible to calculate the total concentrate and tailings volumetric flow rates by summing the individual volume fractions of all the recovered floatability and entrainability classes. The calculated concentrate volume flow rate is used to calculate the froth residence time (τ_{fc}) and the tailings volume flow rate is used to calculate the cell residence time (τ_c). The method of successive approximations is used iteratively until the guessed residence times equal the calculated residence times. The final solution is the concentrate and tailings volumetric flow rates, which are either products from the circuit or are recycled to other cells in the circuit. The recirculated flow rates change the feed flow rates to the cell that has just been iterated (if the cell is not only receiving feed to the circuit) and therefore a nested iteration of successive approximations is used to get a final solution of the mass balance around each individual flotation cell as well as for the recirculating streams in the circuit.

In order to obtain the model parameters, the above iteration algorithm is followed allowing the parameters of all the models to vary, minimising the sum of errors between the measured and calculated species flows for the entire circuit until a final set of parameters are obtained.

5.3 Unit cell simulations

This section presents the simulated fit, using the algorithm described in section 5.2 for the rougher froth characterisation test (test 2). The parameters obtained from the individual component modelling were used as starting parameters. These parameters were then optimised (using Excel solver) to obtain the final parameters from the pilot plant data. The simulation output for the unit rougher cell is presented to show the form of the model as a function of the grade and recovery while varying the froth depth and air flow rate to the flotation cell. The overall model fit for the feed characterisation test (test1) and the rest of test 2 is presented at the end of this section together with the final parameter set describing the kinetics of the pilot plant.

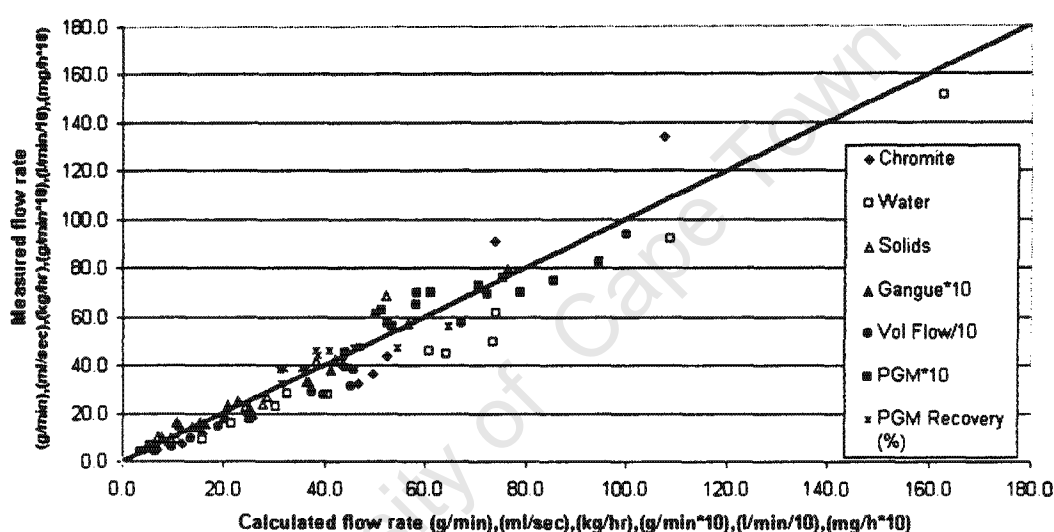


Figure 5.13: A comparison of the simulated output to the experimental data of the rougher froth characterisation test

Figure 5.13 compares the simulation output to the experimental data for total water, chromite, solids, gangue, volume, PGM flow and PGM recovery. The fit was relatively good showing that the interaction of all the component models in the simulator were working well in combination to predict the overall response of the flotation cell as a function of froth height and air flow rate. Figure 5.14 shows the corresponding PGM grade versus recovery curve from the experiments compared to the model prediction. Again the model predicted the shape of this curve within the bounds of experimental error. However, figure 5.14 shows that infinite grades are predicted at low recoveries from the flotation cell. This peculiarity will be addressed next where the sensitivity of the froth recovery model parameters are discussed. Figure 5.15 shows the simulated decrease in the froth recovery of attached particles with froth height for each component at a constant air flow rate of 30 l/min.

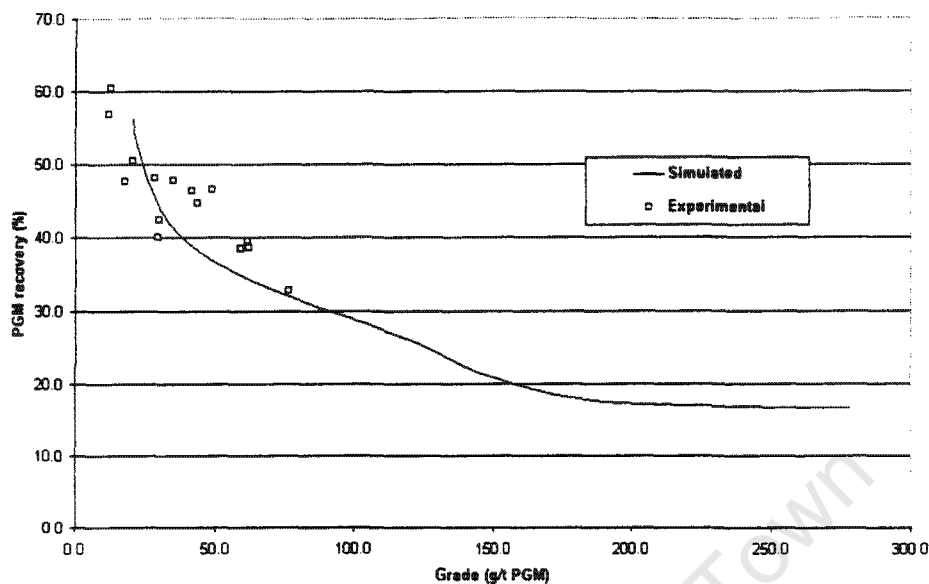


Figure 5.14: A comparison of the experimental and simulated grade versus recovery curve for the rougher froth characterisation test

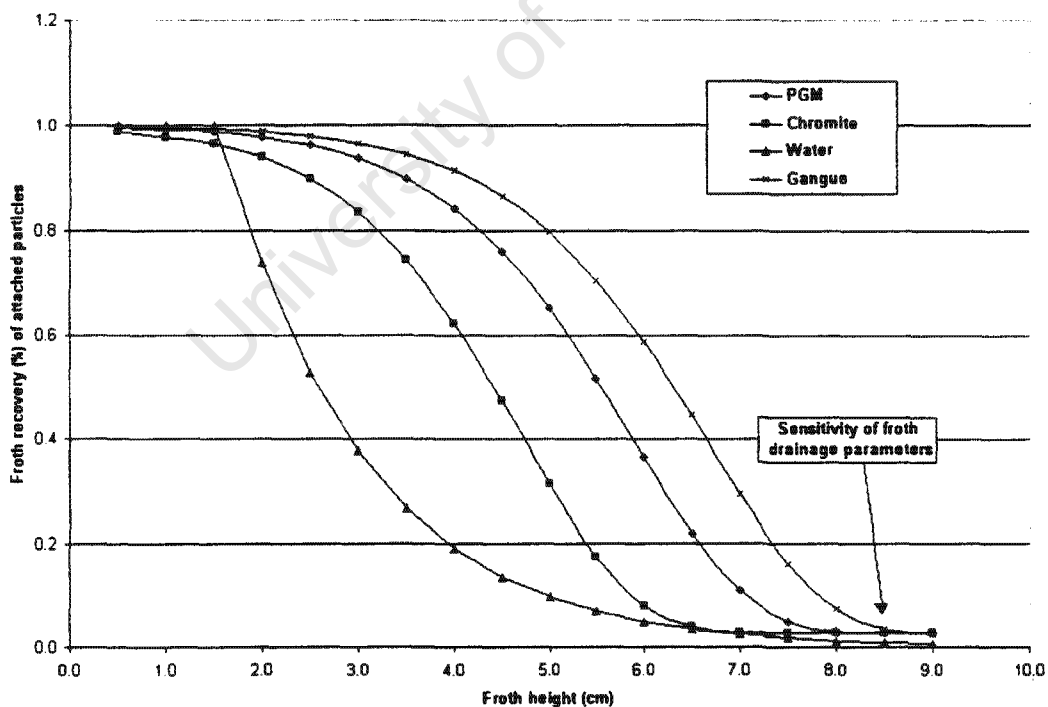


Figure 5.15: The decrease in the froth recovery of attached particles with increasing froth height, showing the sensitivity of the froth drainage parameters to the simulated boundary condition of the grade recovery curve

Figure 5.15 shows that for a fixed froth height the froth recovery of particles attached to bubbles is mineral specific decreasing in order of the average specific gravity of each component in the froth i.e: gangue, PGM and chromite. This appears to be reasonable; if attached hydrophobic particles in the froth are detached for some reason (via bubble coalescence perhaps) it would be expected that the particles with a high density should settle faster than the particles with a low density. The froth recovery of attached particles should not be confused with the froth recovery of entrained particles in the froth. These particles follow the water froth recovery curve in figure 5.15. This curve drops off a lot quicker with froth height carrying a large portion of hydrophilic entrained PGM, chromite and gangue particles with it. The higher proportion of entrained versus floatable classes for the chromite and gangue provides a logical basis for the observed increase in mineral grade with froth height in the flotation cell.

The sensitivity of the froth drainage parameter (α_i) becomes evident when analysing figure 5.15 at deep froth depths. If $\alpha_{\text{PGM}} > \alpha_{\text{Ga,Cr}}$ this would cause the simulator to predict infinite PGM grades at infinite froth depths. If $\alpha_{\text{PGM}} < \alpha_{\text{Ga,Cr}}$ then the boundary condition of the grade-recovery curve in figure 5.14 would predict zero PGM grade at infinite froth depths. Neither of these two boundary conditions would be expected to describe the real system, however. To avoid the simulator predicting an incorrect boundary condition such as the one described above, a constraint was imposed in the simulation algorithm that prevented the simulator from working in this region. A maximum grade constraint was imposed, above which the recovery from the cell was set to zero. This approach appears to be reasonable as figure 5.14 shows that grades of more than 100 g/t were unobtainable during the rougher froth characterisation test. Similar constraints were imposed in some of the work of Cutting et al (1981).

At shallow froth depths the simulator decreases the froth recovery of attached particles before the water starts draining out of the froth. This is due to the constraint imposed on the water froth recovery model ($R_{i,w} < 1$ for all froth residence times) causing a discontinuity in this curve. This clearly represents a spurious phenomena and a slightly more complicated water model that plateaus off to 100 percent $R_{i,w}$ at lower froth heights would certainly represent a better approximation of reality. This situation would occur when the buoyancy force caused by the rising air bubbles exceeds the downward gravitational force exerted on each component in the froth. Due to the higher specific gravity of the particles, it is expected that they should start draining before the interstitial water in the plateau borders starts draining. An attempt to apply such a model was not made as it is evident that this region of operation is not within a feasible operating range, making this additional complication not worth while.

Figure 5.16 shows the simulated effect of the operating variables (froth height and air flow rate) on the resulting PGM recovery from the rougher cell. Each curve represents the overall recovery of PGM as a function of froth depth at a constant air flow rate into the cell. At zero froth depth, the recovery increases with increasing air flow rate due to a higher bubble surface area flux levitating the floatable particles out of the pulp. The relative velocity at which particles fall out of the froth for any particle liquid system is determined by the spatial distribution of the bubbles, the hold-up of slurry in the froth

and the size, shape and density of the particles. This relative velocity together with the net upward velocity of the slurry liquid in the froth determines whether particles in the froth liquid have a net upward or downward velocity, (Moys, 1989). Increasing the froth depth initially has no effect on the recovery from the cell as the wake of ascending air bubbles creates a greater upward force than the downward gravitational force exerted on the particles in the froth. This results in virtually all the floated particles being recovered in the froth zone. As the froth depth is increased further a greater downward gravitational force results due to an increase in the head of froth above the pulp-froth interface. According to the froth recovery model this takes on the form of an exponential decay function with froth residence time. This results in a decrease in recovery with froth depth.

At higher air flow rates, the buoyancy force of ascending air bubbles is greater than at lower air flow rates, (Szatkowski, 1987) resulting in higher PGM recoveries at the same froth depths. At very deep froth depths, the overall PGM recovery drops to zero as the grade constraint is encountered. This point is reached at deeper froth depths as the volumetric air flow rate is increased. The higher the air flow the greater the upward buoyancy force and the lower the froth residence time of air resulting in a higher water froth recovery as well as mineral froth recovery. Obviously there is a limit to how much air can be put into a flotation cell before it starts flooding (where the impeller/stator is not able to break up all the air into bubbles causing excessively large bubbles in the centre of the flotation cell –commonly known as boiling). At this point the recovery and grade would drop off extremely rapidly as a result of an unstable pulp froth interface. This is not a feasible operating regime and should be avoided.

In order to avoid unfeasible operating regimes such as the one previously described, an empirical polynomial was imposed on the simulator whereby a maximum air flow rate was permissible for a particular froth depth. The data used to develop this constraint equation was obtained from the froth characterisation tests where the froth was pulled the hardest at each froth depth. These operating points were definitely on the point of becoming unfeasible as in reality a flotation cell would not be pulled as hard as occurred in these tests. Figure 5.17 shows this boundary condition from the four froth characterisation tests, indicating the maximum feasible air flow rate at the specified froth depth as well as the polynomial equations used to impose the boundary conditions.

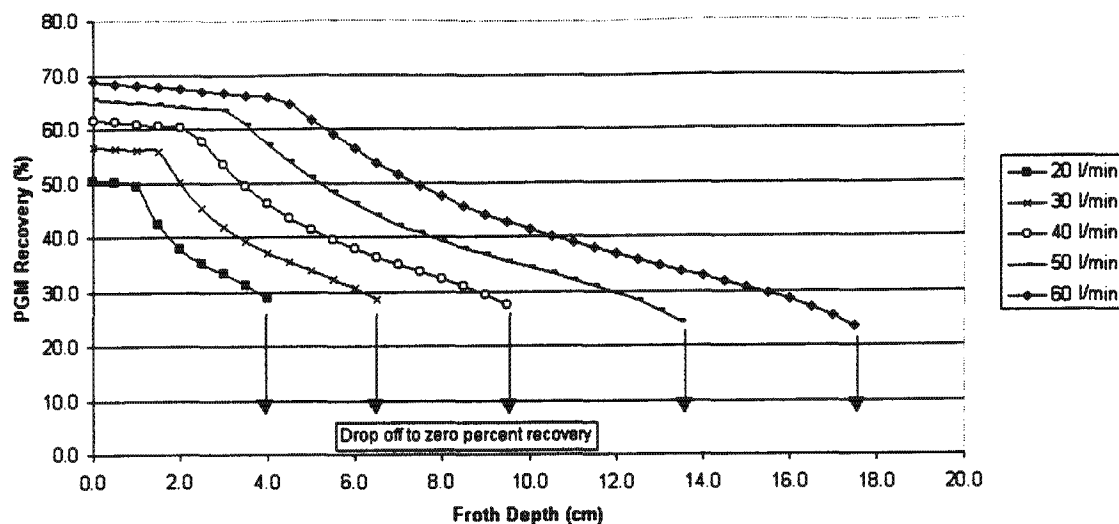


Figure 5.16: The simulated variation of PGM recovery with froth depth and air flow rate

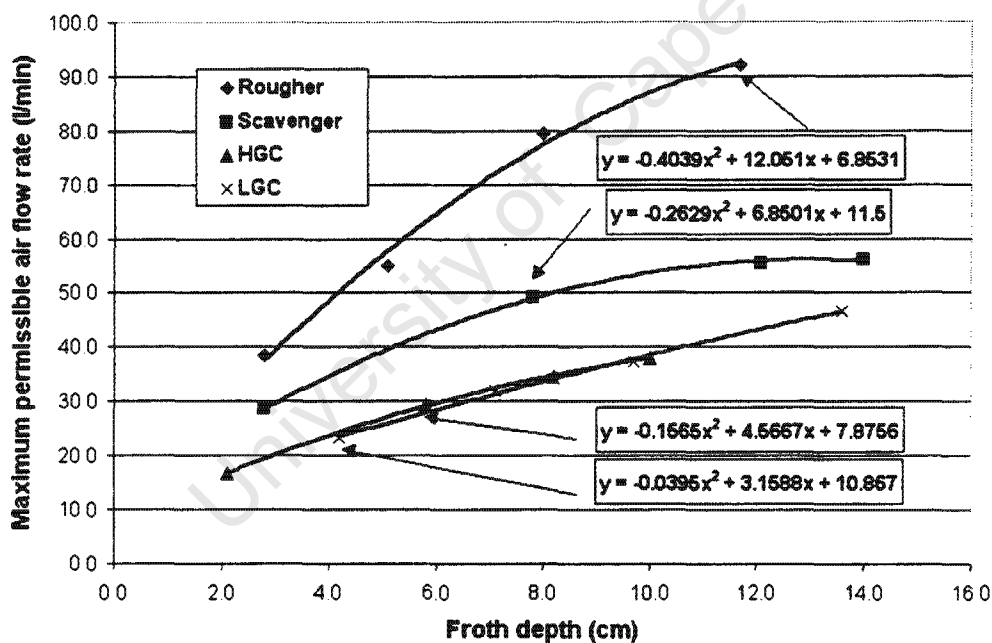


Figure 5.17: The boundary condition of maximum permissible air flow rates as a function of froth depth imposed on the simulation algorithm

Figure 5.18 shows the grade versus froth depth relationship for constant volumetric air flow rates. At very shallow froth depths the grade of the concentrate remains constant with increasing froth volume due to limited classification effects, the grade depending entirely on the flotation kinetics of the pulp. Only beyond a certain froth depth (critical

enrichment depth) does the gravitational force of the froth head above the pulp/froth interface exceed the buoyancy force of ascending air bubbles causing particles to return to the pulp. Froth drainage removes particles not strongly attached to bubble surfaces (generally gangue particles). If preferential attachment of valuables occurs, then the concentrate grade would be expected to improve, (Moys, 1989). The critical enrichment line in figure 5.18 is longer the higher the volumetric flow rate of air. This is again due to the increase in buoyancy force with an increase in air flow rate, (Szatkowski, 1987). As the air flow rate is increased it becomes possible to float with deeper froth depths. There is a limit to this, however, before gravitational forces destabilise the froth causing it to collapse resulting in no concentrate flow rate. This condition once again represents an unfeasible operating regime, which can be avoided by ensuring that reasonable froth depths are used which normally should not exceed 1/3 of the total cell depth. Engelbrecht and Woodburn (1975) point out that an optimum froth height exists to maximise selectivity in the froth at a specified recovery, this height being dependent on the aeration rate.

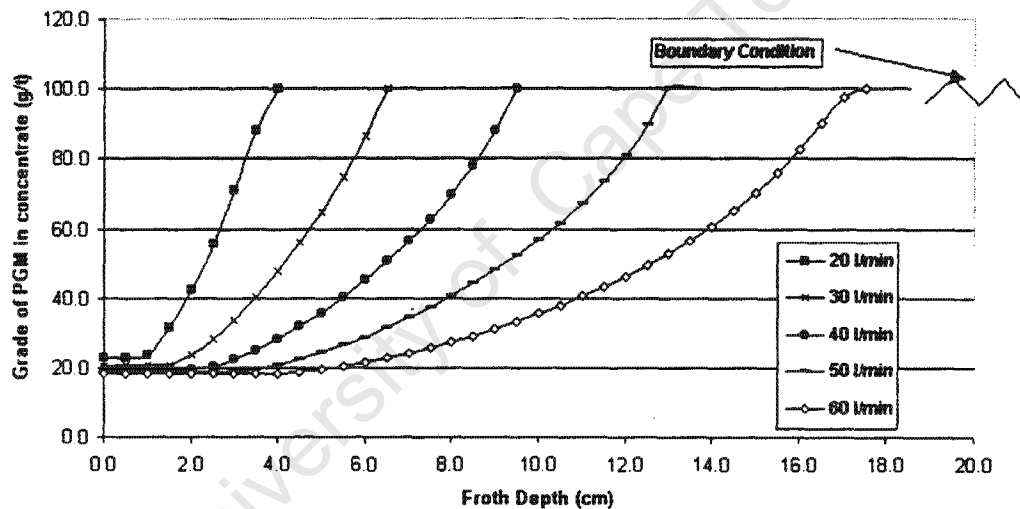


Figure 5.18: The simulated variation of PGM grade with froth depth and air flow rate

Figure 5.19 shows the grade-recovery relationship obtained from the simulation. This is essentially a combined plot of figure 5.16 and figure 5.18. Each line represents a constant volumetric air flow rate with a change in froth depth. For a constant air flow rate, gradually increasing the froth depth results in the expected decrease in recovery with an increase in grade. As already discussed, this is due to classification within the froth as a result of the drop back of material from the froth. Interestingly, the lines of constant air flow rate all tend towards a common ("thermodynamic") grade recovery curve at feasible operating froth heights and air flow rates. This indicates that the froth is essentially a hydraulic classifier as stated by Moys (1989) and a point on the grade-recovery curve is determined by the combined effects of the froth height and air flow rate due to a balance of upward and downward forces. The same force balance can be

achieved at different operating conditions ie: high air flow rate at a deep froth depth or a lower air flow rate at a shallower froth depth, each able to achieve the same point on the grade recovery curve. This grade recovery curve tends towards the feed grade of the cell at very high recoveries, as expected.

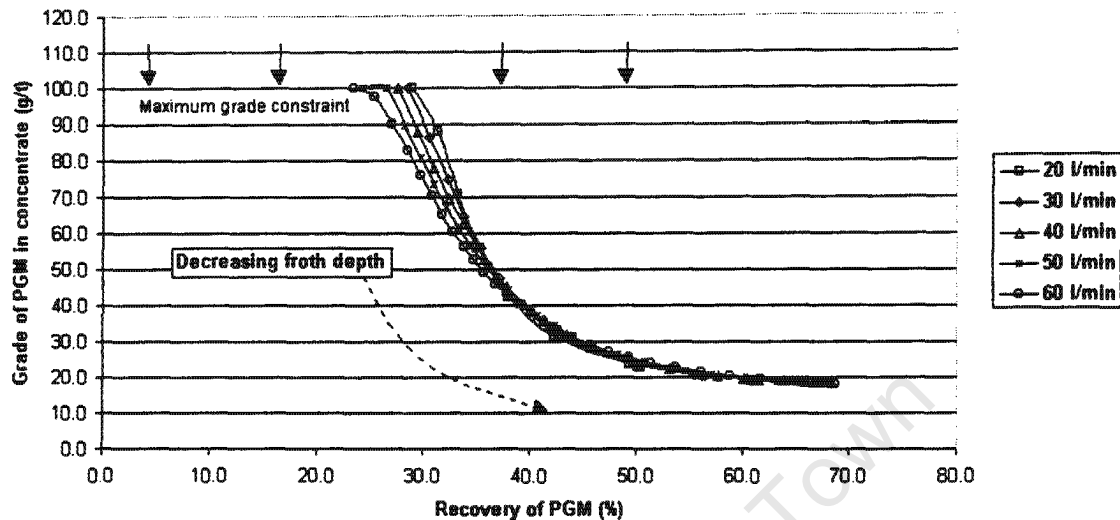


Figure 5.19: Simulated PGM grade versus recovery as a function of froth height for different volumetric air flow rates

Figures 5.20 – 5.23 show the comparison of the simulated output to the experimental data for tests 1 and 2. Table 5.2 shows the final parameter set describing the simulated kinetics for all the components of the pilot plant. A comparison of this parameter set to the preliminary parameter set obtained in chapter 4 shows that the optimisation did not need to vary the parameters by very much in order to fit the data with reasonable accuracy. In total, 1740 data points were used to optimise the 48 parameters in table 5.2.

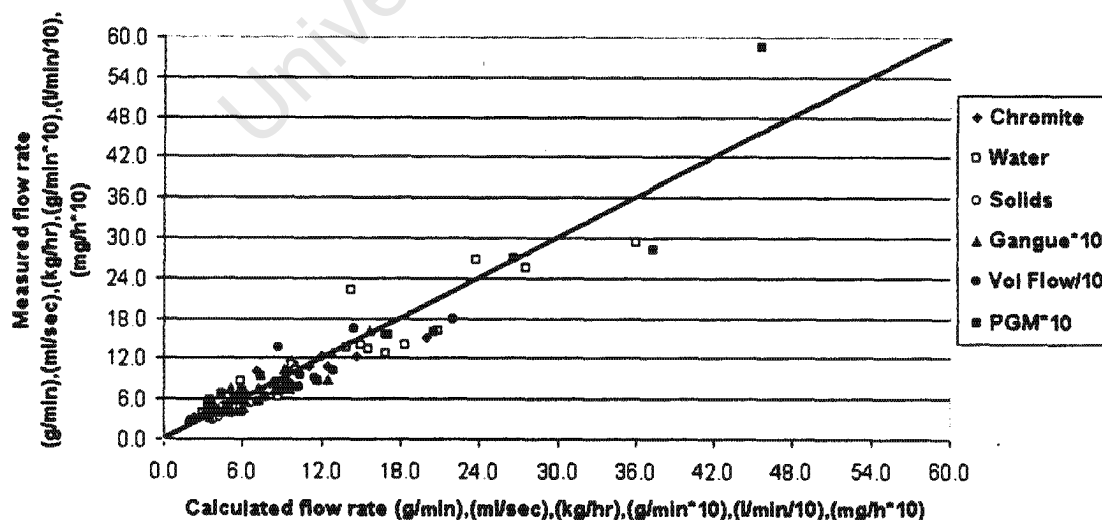


Figure 5.20: A comparison of the simulated output to the experimental data of the feed characterisation test

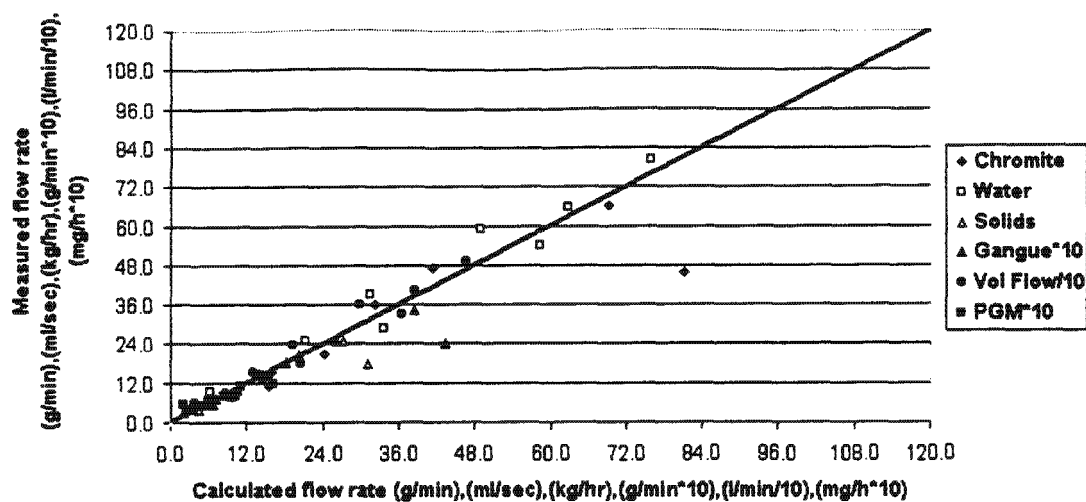


Figure 5.21: Simulated fit of the scavenger froth characterisation test

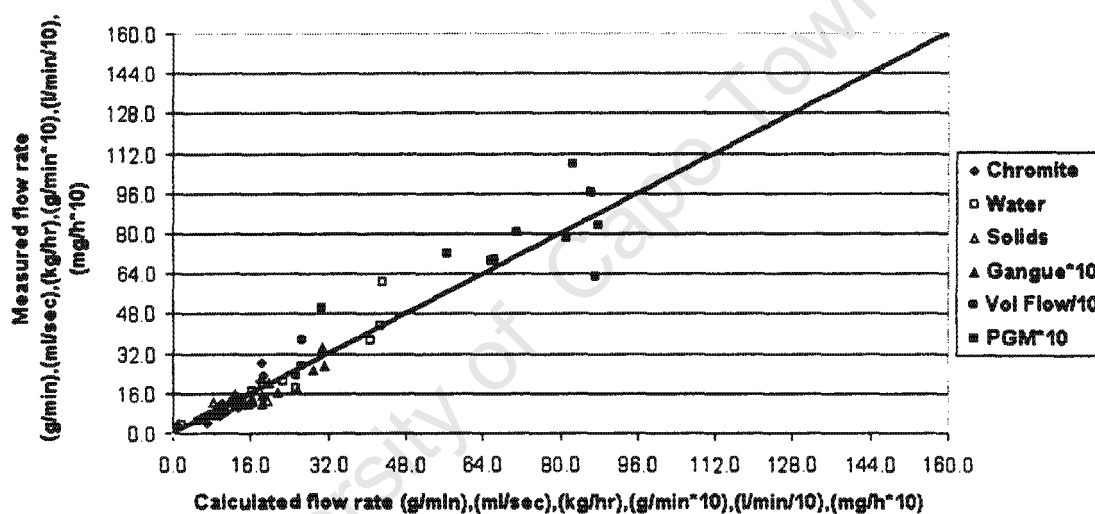


Figure 5.22: Simulated fit of the HGC froth characterisation test

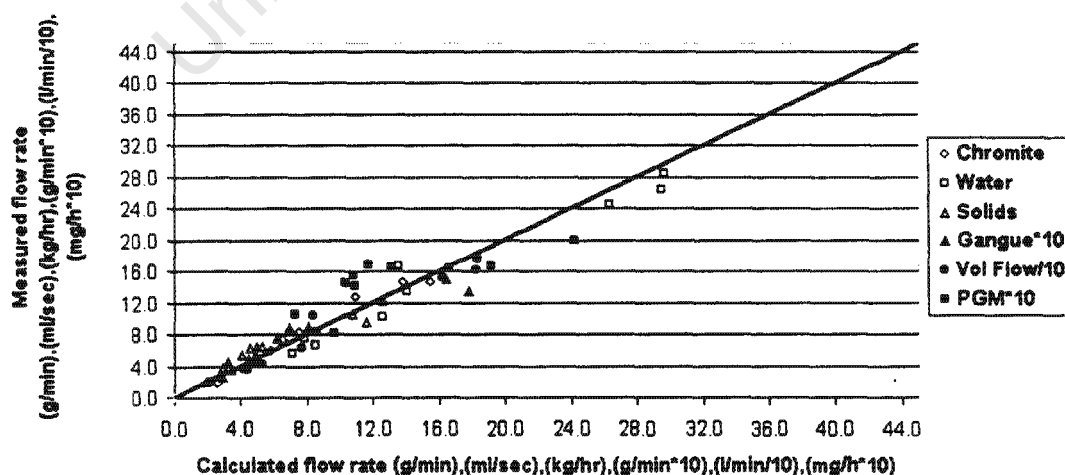


Figure 5.23: Simulated fit of the LGC froth characterisation test

Table 5.2: The final regressed parameter set for the pilot plant

PGM		WATER		CHROMITE		GANGUE		S _B	
Parameter	Value	Parameter	Value	Parameter	Value	Parameter	Value	Parameter	Value
$P_{fast,PGM}$	2.54e-3	P_w	1.24e-4	P_{Cr}	1.85e-4	$P_{f,Ga}$	6.51e-2	a	125.0
$P_{slow,PGM}$	2.43e-4	$\Omega_{rougher}$	2.77	ϵ_{Cr}	17.83	$P_{s,Ga}$	1.71e-4	b	0.51
$m_{fast,PGM}$	0.28	Ω_{scav}	2.23	δ_{Cr}	0.63	ϵ_{Ga}	17.35	c	0.66
$m_{slow,PGM}$	0.45	Ω_{hcc}	1.65	$\phi_{Cr,feed}$	0.0068	δ_{Ga}	1.25	d	0.011
E_{PGM}	0.54	Ω_{hcc}	1.87	$\beta_{r,Cr}$	0.100	$\phi_{Ga,feed}$	0.049	e	0.046
$\beta_{r,PGM}$	0.041	$\sigma_{rougher}$	0.12	$\beta_{s,Cr}$	0.100	$m_{fast,Ga}$	0.048		
$\beta_{s,PGM}$	0.052	σ_{scav}	0.15	$\beta_{hcc,Cr}$	0.014	$\beta_{r,Ga}$	0.02		
$\beta_{hcc,PGM}$	0.069	σ_{hcc}	0.24	$\beta_{hcc,Cr}$	0.002	$\beta_{s,Ga}$	0.07		
$\beta_{hcc,PGM}$	0.095	σ_{hcc}	0.19	α_{Cr}	0.028	$\beta_{hcc,Ga}$	0.03		
α_{PGM}	0.020	$\chi_{rougher}$	15.75			$\beta_{hcc,Ga}$	0.01		
		χ_{scav}	12.36			α_{Ga}	0.02		
		χ_{hcc}	14.67						
		χ_{hcc}	15.74						

5.4 Conclusions drawn from this chapter

This chapter has presented a water model that has been used in the development of the simulation algorithm of the pilot plant. This model is based on cell operating parameters allowing for a more stable convergence of the model solution. Alternative empirical water models have been discussed and some tested to determine the usefulness of these models for inclusion into the simulator. These models were found to be unsatisfactory for the purposes of this study.

A step by step simulation algorithm of the pilot plant has been presented describing in detail how the individual component models can be combined into a structure that is relatively simple, robust, stable and converges rapidly to a solution. The output of the simulation algorithm has been compared to the experimental results of the rougher froth characterisation tests. The comparison was good showing that the component models were able to work well in combination to predict the overall response of the flotation cell as a function of froth height and air flow rate. The sensitivity and interaction of the individual component froth recovery model parameters was discussed with particular reference to the boundary conditions of the simulated grade-recovery curve. Finally, the form of the simulator (with respect to the grade and recovery as a function of froth height and air flow rate) has been discussed in terms of the identification of suitable boundary conditions for the models during simulation.

Chapter 6

OVERALL MODEL VALIDATION

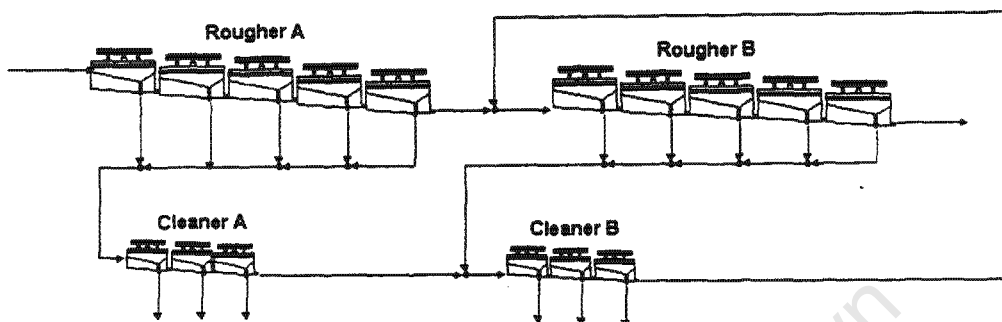
The aim of this chapter is to evaluate and test the simulation and modelling methodology presented in this thesis. The chapter is divided into three sections, the first of which is a comparison of the simulators prediction to the two alternative circuit configurations of test 3 as well as the scale-up test, test 4, in chapter 4. The next section is an attempt to validate the parameters of the PGM regression methodology using the experimental data obtained from the work of phase II in chapter 4 where two feed streams of differing grinds were subjected to the PGM regression analysis. The objective of this was to evaluate the correlation of the regressed parameters to measured characteristics of the ore. The final section makes use of the QEMSEM data obtained from the feed stream to the pilot plant during the feed characterisation test (test 1). This data is used to infer some significance on the PGM floatability mass fractions regressed for the Impala UG2 ore used during the phase I tests.

6.1 Predicting alternative circuit configurations with the simulator

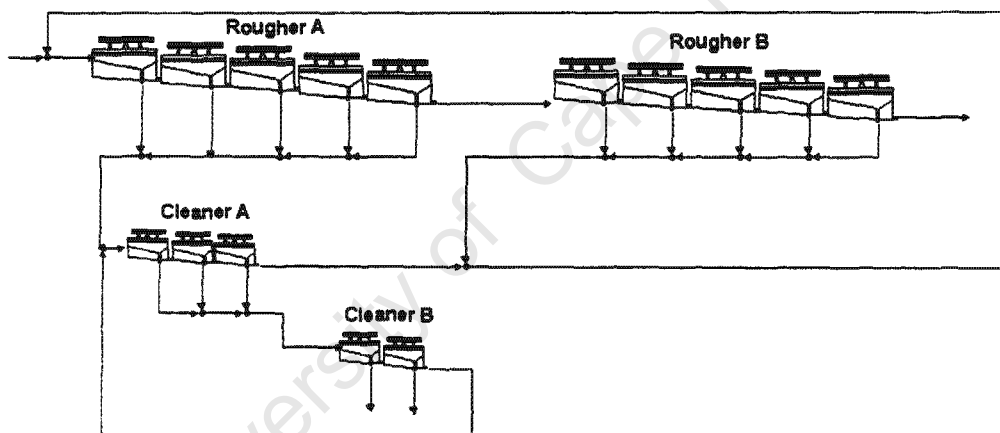
The pilot plant simulation algorithm presented in chapter 5 stressed the importance of the accuracy of the water model in being able to make reliable predictions of the flotation kinetics of the pilot plant. Unfortunately, the water model that has been developed for this thesis was not based on detailed fundamentals required to make accurate predictions of the water kinetics in any flotation bank or in any circuit configuration. This limits the usefulness of the simulation algorithm, particularly for the design of new "green field" flotation circuits. The ability of the simulation algorithm to predict the flotation kinetics of each component for the alternative circuit configurations was made possible by allowing the optimisation subroutine to vary the parameters that describe the water froth model (empirical parameters) as well as the froth models for PGM, chromite and gangue (stability and drainage parameters). These parameters were varied in order to minimise the TSSE between the experimental and predicted flow rates for each component in each circuit configuration. The parameters that described the floatability distribution of the feed, as well as the entrainability functions for gangue and chromite were kept constant. The basis for varying the froth model parameters is that the froth stability and therefore froth recovery is not expected to be the same for the various flotation banks in each circuit configuration. This is a logical assumption and highlights the necessity of a froth model that can allocate new froth recovery parameters for different flotation cells in a circuit. It would be useful to understand the significance of the parameters in the froth models so that this could be used when attempting to simulate alternative circuit configurations from a design perspective where no experimental data is available to fit the empirical parameters in the froth models. The variation of the regressed froth model parameters are presented for all three circuit

configurations for each component in order to evaluate their significance. The alternative circuit configurations are presented again in figure 6.1 for ease of reference, and the flotation banks are labelled according to their reference in the discussion.

CONFIGURATION A



CONFIGURATION B



CONFIGURATION C

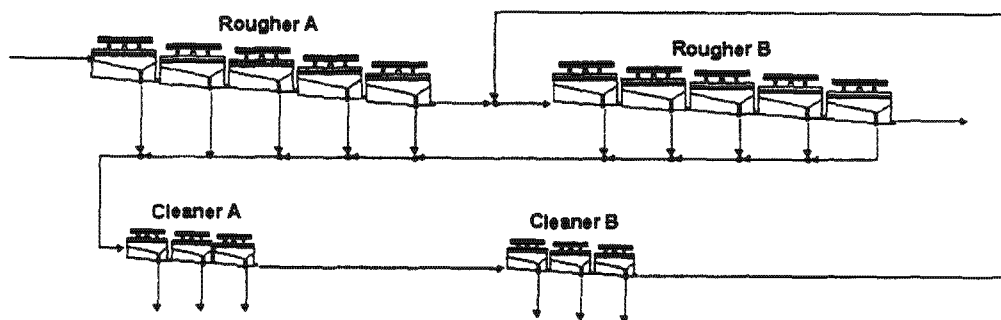


Figure 6.1: The alternative circuit configurations simulated for the pilot plant

Figure 6.2 shows the comparison of the simulated flotation kinetics and the experimental data for circuit configuration B. The froth model parameters (water, PGM, chromite and gangue) accounted for 27 parameters while 288 experimental data points were used in the regression to get a fit of the model parameters. A model fit of 4.07 was obtained, which is considered relatively high, owing to the relatively large change in the circuit configuration. The major cause for the high value of the model fit coefficient was due to the extremely small measured standard deviations associated with the experimentally determined dry mass flow rates. Another significant contribution to the model error was the PGM flow rate predictions for the first cells on cleaner A and B. The high standard deviations of these PGM flow rates is attributed to the extremely high kinetics of the fast floatable fraction of PGM particles reporting to these concentrates. This phenomenon was detected in the individual component PGM modelling of chapter 4 section 4.6.1, and is believed to be a result of inhibited flotation kinetics in these two flotation cells. A model fit of 2.05 was obtained when ignoring the contribution of these errors to the TSSE showing that this accounts for half of the error associated with the model fit. The model fits for water, chromite and gangue were relatively good.

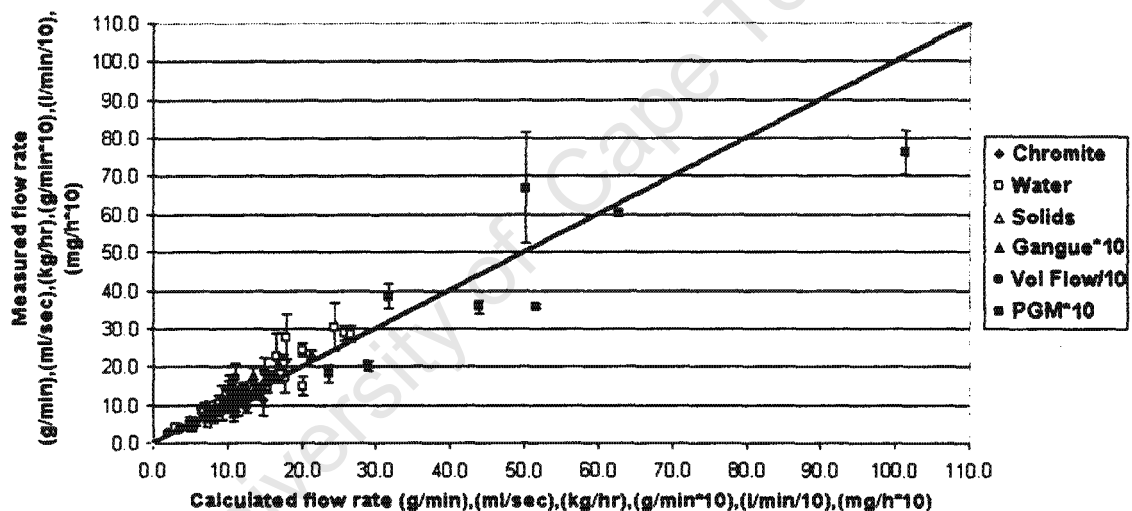


Figure 6.2: The simulated flotation kinetics of circuit configuration B, test 3

Figure 6.3 shows the comparison of the simulated flotation kinetics and the experimental data for circuit configuration C. The froth model parameters (water, PGM, chromite and gangue) accounted for 27 parameters while 288 experimental data points were used in the regression to get a fit of the model parameters. A model fit of 2.04 was obtained showing very good agreement between the experimental flow rates and the calculated flow rates for all the components. This is attributed to the small difference between this circuit configuration and circuit configuration A, which was used to regress the floatability and entrainability parameters. It can be concluded that the more the simulated circuit deviates from the original circuit (where the floatability and entrainability parameters were regressed from) the less unreliable the output of the simulation, owing to limitations in the water recovery model.

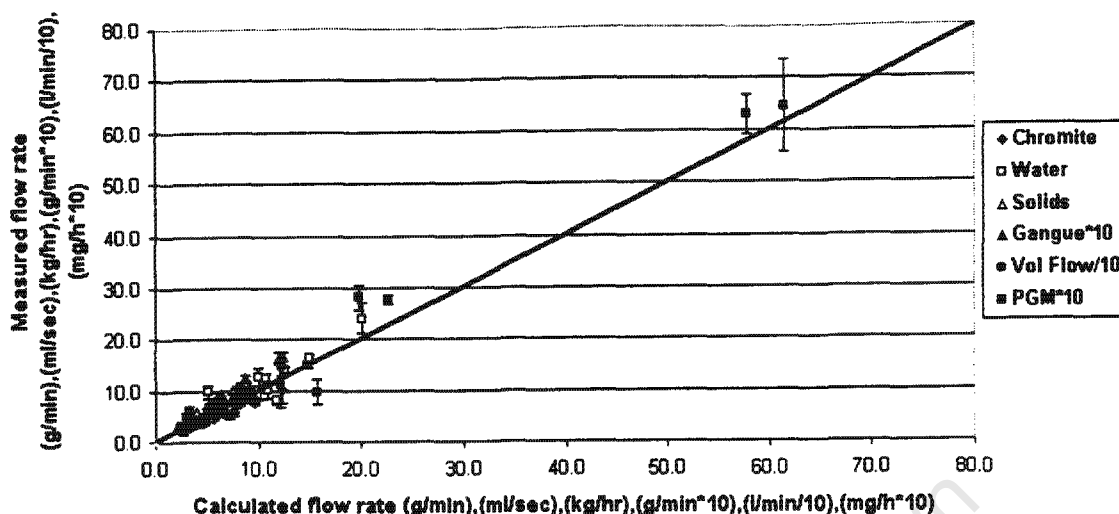


Figure 6.3: The simulated flotation kinetics of circuit configuration C, test 3

Figure 6.4 shows the simulated flotation kinetics for the scale-up test. Exactly the same parameters (including the water and mineral froth parameters) as those used for the feed characterisation test (circuit configuration A) were used to predict the flow rates for both banks of flotation cells. A satisfactory comparison was obtained for all component flow rates. This seems to indicate that the simulation algorithm is capable of predicting the flotation kinetics for flotation cells of different sizes but within the same order of magnitude. It is believed that the water model would be inaccurate when scaling up to industrial flotation cells however as a result of the empirical nature of this model.

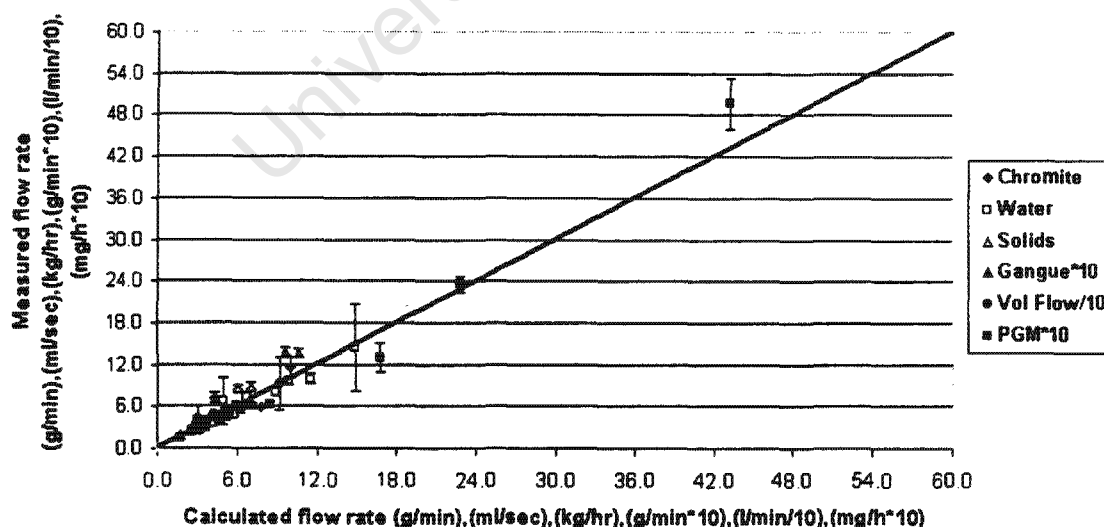


Figure 6.4: The simulated flotation kinetics of the scale-up test, test 4

Table 6.1: The variation of the water froth model parameters for each bank in each circuit configuration

	CONFIG A	CONFIG B	CONFIG C
$\Omega_{\text{rougher A}}$	2.77	3.86	2.67
$\Omega_{\text{rougher B}}$	2.23	2.31	3.62
$\Omega_{\text{cleaner A}}$	1.65	1.48	2.18
$\Omega_{\text{cleaner B}}$	1.87	2.68	2.03
$\sigma_{\text{rougher A}}$	0.12	0.13	0.07
$\sigma_{\text{rougher B}}$	0.15	0.12	0.15
$\sigma_{\text{cleaner A}}$	0.24	0.09	0.54
$\sigma_{\text{cleaner B}}$	0.19	0.18	0.30
$\chi_{\text{rougher A}}$	15.75	16.49	14.04
$\chi_{\text{rougher B}}$	12.36	9.70	12.84
$\chi_{\text{cleaner A}}$	14.67	6.72	17.02
$\chi_{\text{cleaner B}}$	15.74	8.67	16.92

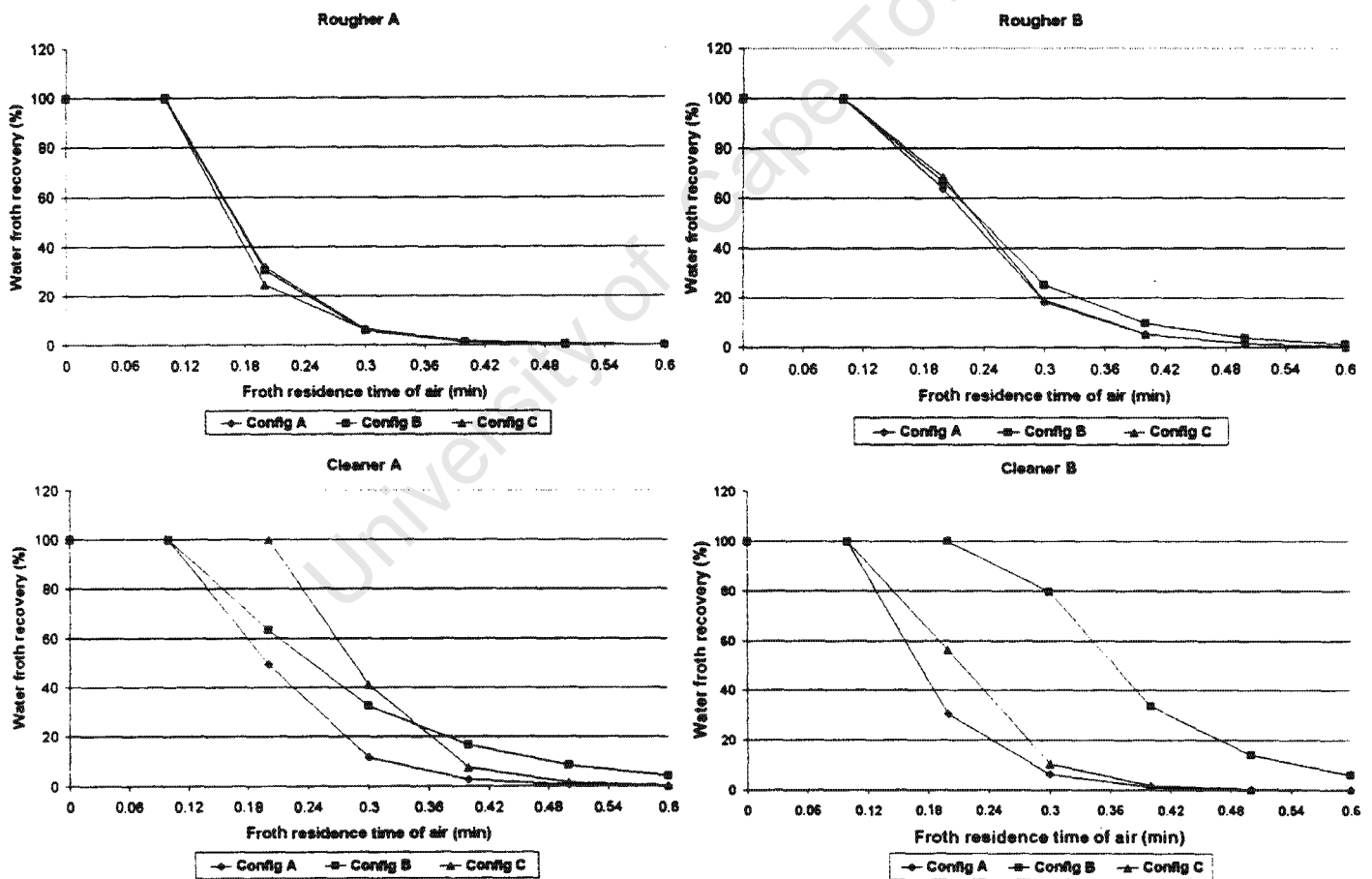
**Figure 6.5: The change in the water froth recovery in different banks for the different circuit configurations**

Table 6.1 shows the change in the regressed water model parameters for each circuit configuration. The section that follows shows the significance of these different parameters showing that if a simulation of a completely different circuit structure is required, it is possible to some extent to obtain relatively accurate predictions knowing how the parameters vary according to the location of the flotation cells in the circuit. Figure 6.5 shows the water froth recovery as a function of froth residence time of air in the different banks for the three different circuit configurations. Figure 6.6 shows how the parameters in the water froth recovery (R_{fw}) model influences the shape of the curve.

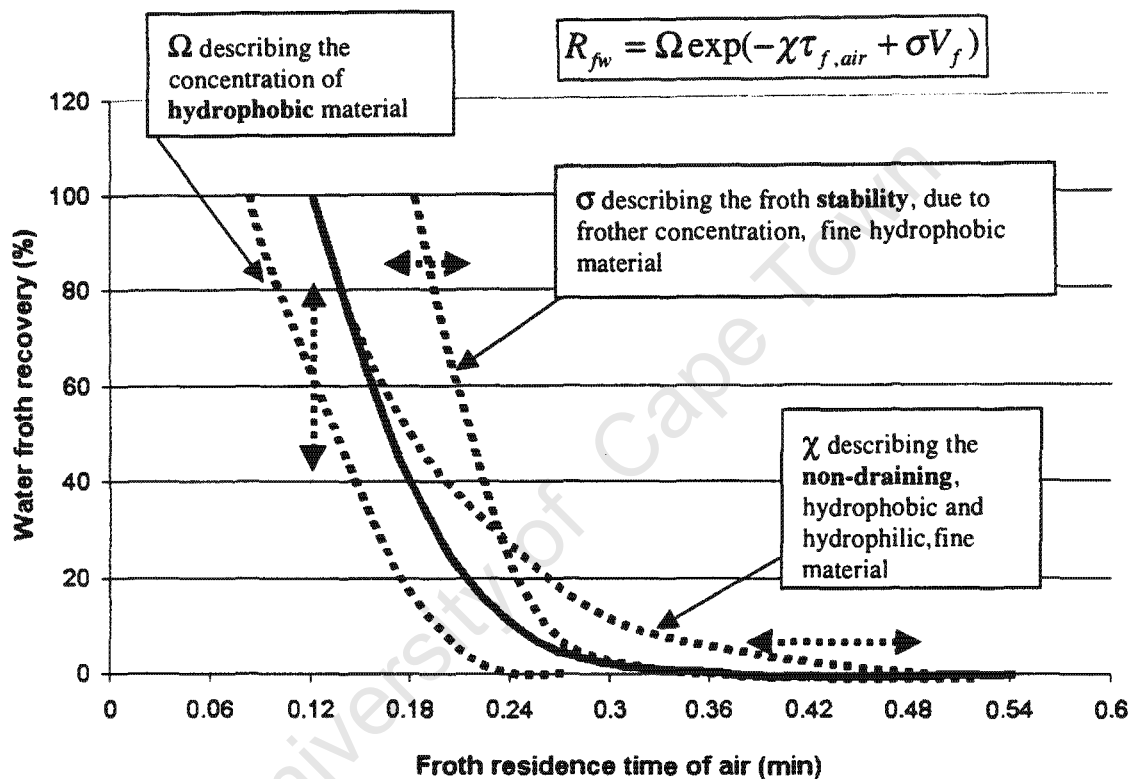


Figure 6.6: The influence of the water froth recovery parameters on the water froth model

From analysing the change in the water froth recovery model parameters that were regressed for the alternative circuit configurations in table 6.1 a number of conclusions have been drawn. Firstly, the Ω parameter moves the entire froth recovery curve of water up or down the y-axis. It is hypothesised that a variation in this parameter is influenced by the amount of floatable particles feeding a bank of flotation cells, representing the contribution of the hydrophobic floatable particles on the recovery of water. The largest $\Omega_{rougherA}$ parameter in table 6.1 was regressed for circuit configuration B (CB), depicting the higher proportion of floatable material due to the circulating load. The $\Omega_{rougherA}$ parameters are the same for circuit configuration A (CA) and circuit configuration C (CC) as they were receiving the same amount of hydrophobic material. The largest $\Omega_{rougherB}$ parameter was regressed for CC, corresponding to the higher

proportion of floatable material from cleaner B tailings stream. The larger Ω_{rougherB} parameter in CB compared to CA can be ascribed to the smaller overall PGM recovery from CB (smaller concentrate production rate) resulting in a higher proportion of floatable material to rougher B bank in CB, even though CA has a circulating load from cleaner B tails to the rougher B bank. The largest Ω_{cleanerA} parameter was regressed for CC; in this circuit all the floatable material from both rougher A and B banks are feeding cleaner A bank. In CA and CB, the hydrophobic fraction is split, a proportion going to cleaner B bank (CA) and another proportion being recycled (CB). The largest Ω_{cleanerB} parameter is expected for CB if this reasoning is valid. Table 6.1 shows that it does have the highest value. The second largest Ω_{cleanerB} parameter was regressed for CC as a result of the reduced residence time in cleaner A bank. The smallest Ω_{cleanerB} parameter was regressed for CA indicating that the cleaner B bank was only treating the low grade material from the cleaner A tailings and rougher B concentrate.

The σ parameter in the water froth recovery model determines the slope of the water froth recovery (R_{fw}) exponential decay as depicted in figure 6.6. The larger σ the steeper the decrease in R_{fw} with froth residence time of air (τ_{air}). This parameter is affected by the stability of the froth as a result of the concentration of frother as well as the amount of fine hydrophobic and hydrophilic material in the stream feeding the bank. The larger the σ parameter the steeper the R_{fw} curve, which could be considered to indicate a more brittle froth that drains quickly. This is clearly evident when looking at the σ_{cleanerB} values in table 6.1. The largest σ_{cleanerB} parameter was regressed for CC; the froth in this bank was definitely more brittle than the froths in cleaner B banks of CA and CB. This was due to the fact that the majority of the floatable fine material was recovered in the cleaner A bank in CC. The smallest σ_{cleanerB} parameter was regressed for CB, this "recleaner" banks' froth was very stable as a result of the higher frother concentration and high fraction of floatable, fine material feeding this bank. The σ_{rougherB} parameters were all essentially the same, although a slightly more stable froth was observed for CB. This could indicate that the recycle streams of CA and CC decreased the stability of the froth in these banks.

The χ parameter in the water froth recovery model determines how long the R_{fw} curve takes to approach zero water froth recovery. The smaller χ , the longer the curve takes to reach zero R_{fw} , this is associated with the amount of slow draining material in the froth as a small fraction of water is still typically recovered at high froth residence times of air. This is clearly evident when comparing the χ parameters for the three circuit configurations in table 6.1. It was observed however that this parameter was not very sensitive, and needed to vary over a wide range to substantially change the shape of the curve. The biggest variation of this parameter between the alternative circuit configurations was noticed for the cleaner B banks. The smallest value regressed was in CB. This bank was receiving very fine and very hydrophobic material (recleaner) that would not be expected to drain from the froth. This type of froth is observed in these cells even in the absence of air addition. The effect of a small χ_{cleanerB} value for CB can be seen in figure 6.5, where the persistent nature of this hydrophobic material, contributing to the higher water recoveries at high froth residence times of air is apparent. Again, the smallest χ_{cleanerA} parameter was regressed for CB as a result of the

recleaner tailings from cleaner B in this circuit carrying along with it a high proportion of non-draining material in comparison to CA and CC. The difference in χ values for the rougher A and B banks are relatively small and this can also be seen in figure 6.5. The major fraction of hydrophobic, fine material reports to the final concentrate and does not influence these banks to any great extent.

Table 6.2: The variation of the PGM froth model parameters for each bank in each circuit configuration

	CONFIG A	CONFIG B	CONFIG C
$\beta_{\text{rougher A.PGM}}$	0.0414	0.0411	0.0405
$\beta_{\text{rougher B.PGM}}$	0.0518	0.0521	0.0506
$\beta_{\text{cleaner A.PGM}}$	0.0691	0.0728	0.0715
$\beta_{\text{cleaner B.PGM}}$	0.0947	0.0705	0.1065
α_{PGM}	0.0200	0.0211	0.0207

Table 6.2 depicts the different PGM froth stability (β) and drainage parameters (α) regressed for the different circuit configurations. The beta parameters describe the stability of the froth phase, the smaller beta the higher the froth recovery for the same froth residence time based on concentrate flow rate. The alpha parameter describes the amount of non-draining material from the froth. No major differences in the PGM froth recovery parameters were required to simulate the different circuit configurations. The only major difference that can be seen in table 6.2 is the large froth stability parameter for circuit configuration C in the cleaner B bank. The low froth recovery of PGM's in this bank is attributed to the brittle froth in this bank as described in the previous discussion of the water model parameters. The PGM particles feeding this bank would be expected to consist of composite PGM grains with a slow floatability. The very small change in the PGM froth recovery model parameters indicates why a satisfactory fit was obtained in chapter 4 for the individual PGM component modelling in section 4.6.1.

Table 6.3: The variation of the chromite froth model parameters for each bank in each circuit configuration

	CONFIG A	CONFIG B	CONFIG C
$\beta_{\text{rougher A.Cr}}$	0.100	0.081	0.101
$\beta_{\text{rougher B.Cr}}$	0.100	0.087	0.100
$\beta_{\text{cleaner A.Cr}}$	0.014	0.012	0.0038
$\beta_{\text{cleaner B.Cr}}$	0.002	0.014	0.0024
α_{Cr}	0.0280	0.0282	0.0202

The parameters in table 6.3 and 6.4 describe the froth recovery of the hydrophobic chromite and gangue particles attached to bubbles. The froth recovery of the entrained hydrophilic particles follow the water froth recovery curve. Owing to the small fraction of floatable chromite in the feed to all three circuit configurations, the parameters in table

6.3 were not found to be very sensitive during the modelling exercise. Varying these parameters did not contribute significantly to a change in the overall model fit. The very small froth stability parameters in the cleaner A and B banks show that the fine, floatable chromite particles in the cleaner banks have virtually 100 % froth recovery. This is a major problem on most industrial UG2 flotation plants, making it extremely difficult to get less than 3% chromite in the final concentrate and it is believed that this is due to the contribution of these particles.

Table 6.4: The variation of the gangue froth model parameters for each bank in each circuit configuration

	CONFIG A	CONFIG B	CONFIG C
$\beta_{\text{rougher A.Ga}}$	0.0200	0.0205	0.0199
$\beta_{\text{rougher B.Ga}}$	0.0695	0.0592	0.0681
$\beta_{\text{cleaner A.Ga}}$	0.0272	0.0207	0.0258
$\beta_{\text{cleaner B.Ga}}$	0.0065	0.0364	0.0068
α_{Ga}	0.0200	0.0282	0.0199

Even though the froth recovery parameters in table 6.4 are associated with the gangue particles attached to bubbles, they were much more sensitive to the model fit for the alternative circuit configurations. Five percent of the gangue in the feed stream was floatable and these particles had a relatively high froth recovery rate compared to the entrained gangue particles. As a result, the stream feeding the cleaner banks had over 50% floatable gangue and therefore the gangue froth stability parameters in the cleaner banks played a crucial role in the overall gangue recovery predictions in the simulations. The very small froth stability parameters for the cleaner B bank in configurations A and C is due to the fact that most of the gangue particles feeding these two banks had been entrained with water by the rougher B bank and therefore the contribution of the floatable gangue fractions was relatively smaller. Most of the floatable gangue was recovered in rougher A and cleaner A banks. The larger froth stability parameter in the cleaner B bank of CB is as a result of the high fraction of floatable particles coming directly through rougher A and cleaner A banks. The final grade predictions of the CB simulation were very sensitive to the value of this parameter. The larger value of this parameter shows the effectiveness of this bank in increasing the grade of PGM's (decreasing the recovery of gangue) with concentrate recleaning by effectively controlling the concentrate froth residence time in this bank.

In conclusion, the ability of the simulator to predict the alternative circuit configurations was relatively good provided the water froth recovery parameters as well as the mineral froth recovery parameters were changed appropriately. The analysis of the froth model parameters in this section has highlighted the correlation of the regressed parameters with the physical conditions in the cells and banks in the various circuits. Accurate determination of the empirical water model plays a very important role in defining the feasibility and accuracy of any simulation results. The mineral froth recovery model parameters were found to be less sensitive in this regard. The Ω empirical parameter in the water froth recovery model was considered to be associated with the contribution of

the floatable hydrophobic particles to the overall water recovery. The higher the concentration of floatable particles the larger the associated Ω parameter. The σ parameter in the water froth recovery model was found to determine the slope of the water froth recovery (R_{fw}) function which can be considered to be affected by the stability of the froth. The χ parameter in the water froth recovery model determines how long the R_{fw} curve takes to approach zero water froth recovery. The smaller χ , the longer the curve takes to reach zero R_{fw} , and this was considered to be associated with the amount of very slow draining material in the froth. Understanding the significance of the parameters in the empirical water recovery model offers some potential to use this type of approach to simulate "green field" designs where no data is available to regress the different empirical constants in this model. It must be noted however, that obtaining the necessary experimental data to regress the parameters in the water froth recovery model is very simple. Only water flow rate measurements are required at a number of different froth depths and air flow rates. This is quick and easy to measure and no assays are required. Finally, it was observed that the simulation of the scale-up tests was very good without any parameter adjustment in any of the models, including the froth models. This shows the potential of the simulator to predict the overall response of a flotation circuit with a change in residence times in various banks as well as the potential for scale-up. The ability of the water model to scale-up directly to industrial flotation cells from pilot plant flotation tests is not expected to work sufficiently well as a result of the empirical nature of this model. This requires further investigation.

6.2 Regression of two differing feed floatability distributions

The modelling methodology presented in chapter three and four assumes that the floatability components are conserved around units and nodes in a flotation circuit. This is provided there are no regrind stages or chemical addition points in the circuit that could change the nature of the floatable particles. Not much work has been done to assess, quantitatively, what a regrind mill does to the floatability distribution of an ore. One attempt to describe this phenomenon is the work of Runge (1997). An assumption was made in this work that a regrind mill does not influence the rate (or floatabilities, P 's) of the particles but instead redistributes the mass fractions (m 's) by liberating more particles, thereby decreasing the non floating fraction and possibly grinding composite slow floating particles to render them fast floating. This approach is based on no scientific evidence, but it is logical as there are still two fractions commonly observed by plant operators after a regrind mill (fast or slow) and there is still a proportion of material that could be locked and therefore non-floating. In order to test the PGM regression methodology presented in this thesis, two feed streams with different feed grinds were evaluated, using a sample of Lonmin Karee UG2 ore. The experimental set-up was presented in chapter 4, section 4.5.2. Three mass balances as well as batch rate floats on the combined rougher and scavenger concentrates as well as all tailings streams were sampled (at a fixed set of operating conditions) for both feed streams. Figure 6.7 shows the particle size distributions of the coarse and fine grind streams fed to the pilot plant. The grind of the coarse and fine streams was approximately 50% -75 μ m and 85% -75 μ m respectively.

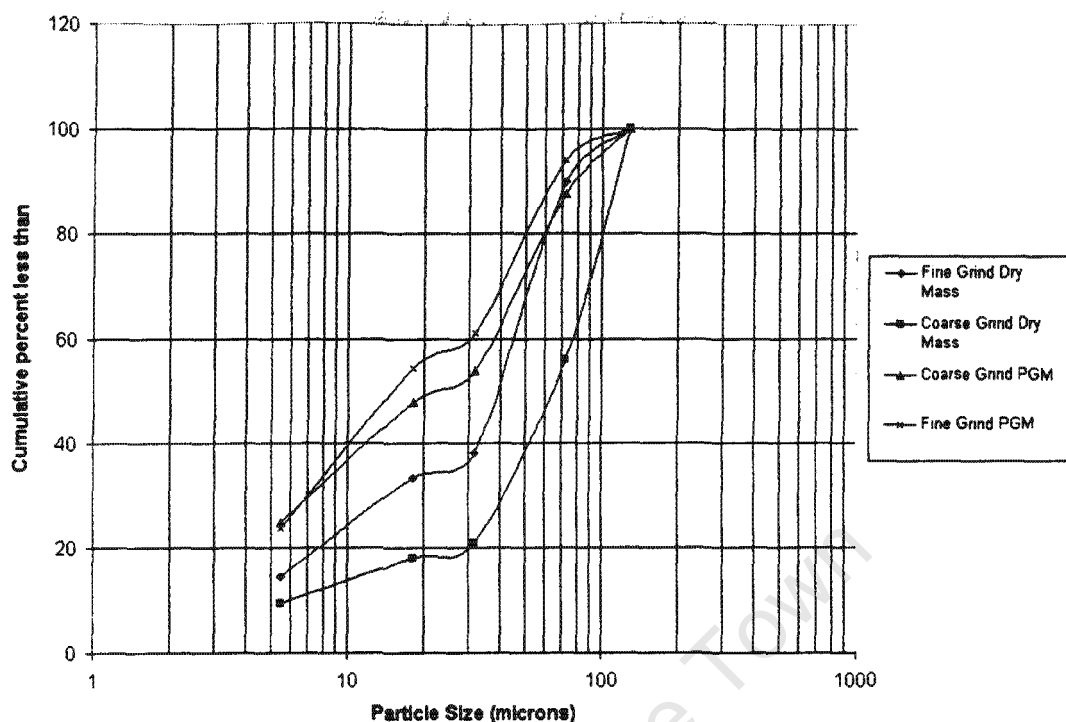


Figure 6.7: Particle size distributions of the two different feed streams fed to the pilot plant

The fine grind stream curve is above the coarse grind stream throughout the size distribution showing that very different streams were being introduced as feed to the pilot plant. Figure 6.7 also shows the size distributions of the PGM particles highlighting the difference in the PGM particle distributions in both streams. Figures 6.8 and 6.9 show the cumulative PGM recovery time curves as a function of particle size class for both the coarse and fine grind streams.

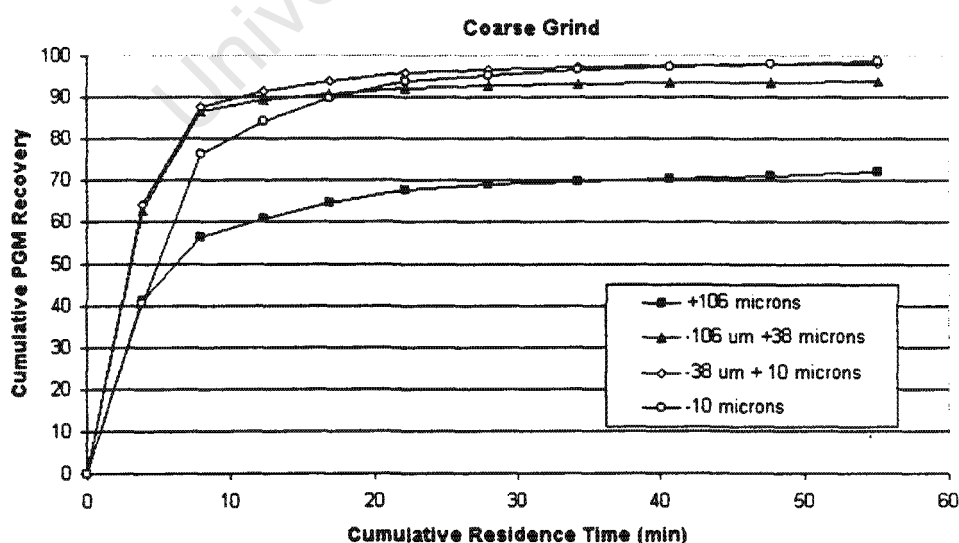


Figure 6.8: Cumulative PGM recovery versus time for the coarse grind stream

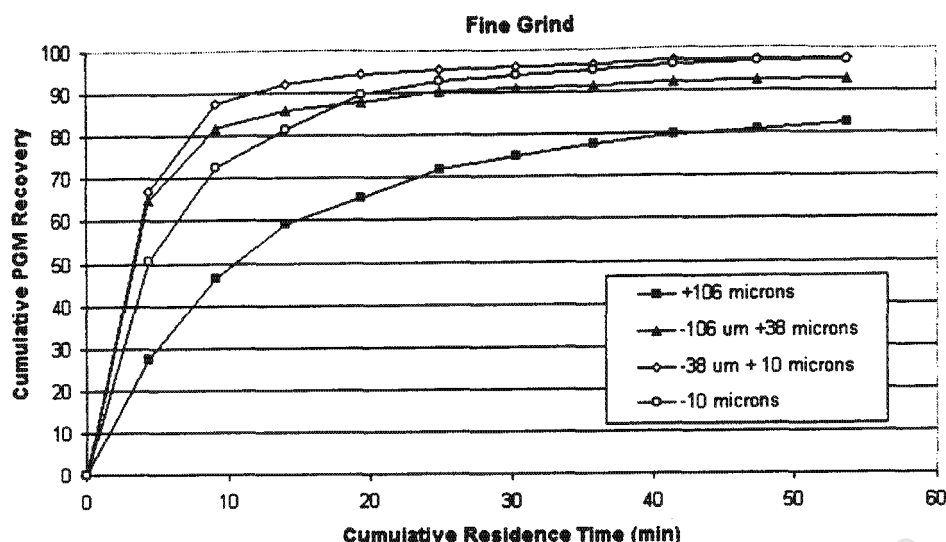


Figure 6.9: Cumulative PGM recovery versus time for the fine grind stream

What is clearly evident in figures 6.8 and 6.9 is the increase in the overall recovery of the +106 μ m fraction for the fine grind stream. This shows that some extremely coarse locked PGM particles were liberated (but still remained larger than 106 μ m) resulting in a ten percent increase in the overall recovery for this size fraction. The difference in the flotation rates between the different size fractions, particularly for the -10 μ m and the +106 μ m fractions is evident. The -10 μ m size fraction floated much slower than the +10 μ m-106 μ m size classes for both the coarse and fine grinds. This phenomena has been observed in many flotation studies, and is discussed in section 2.2.2.2 in the literature review. Figure 6.10 shows the recovery-size response for the coarse grind feed material. This figure shows the smaller overall recoveries of the non-floating +106 μ m particles as well as the slower flotation rates for the -10 μ m and +106 μ m particles. The optimum size for the best recovery was found for the +25-106 μ m fraction.

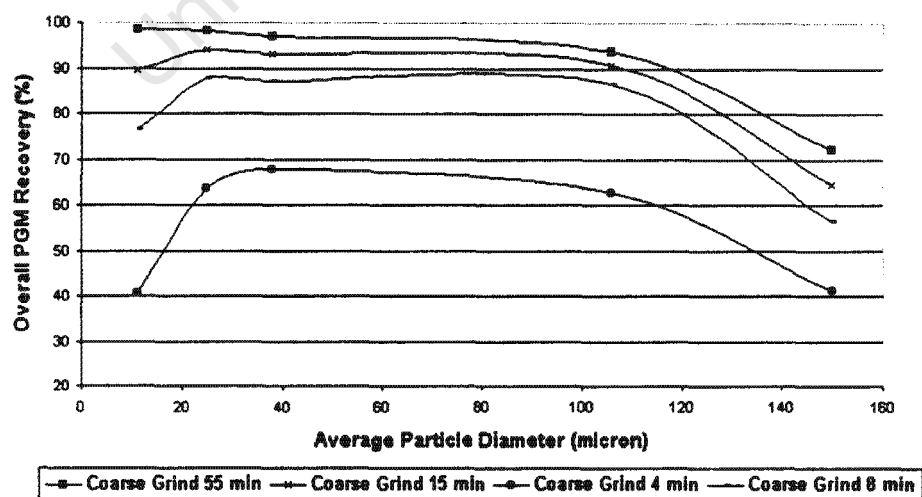


Figure 6.10: The overall recovery with time as a function of size class for the coarse grind material

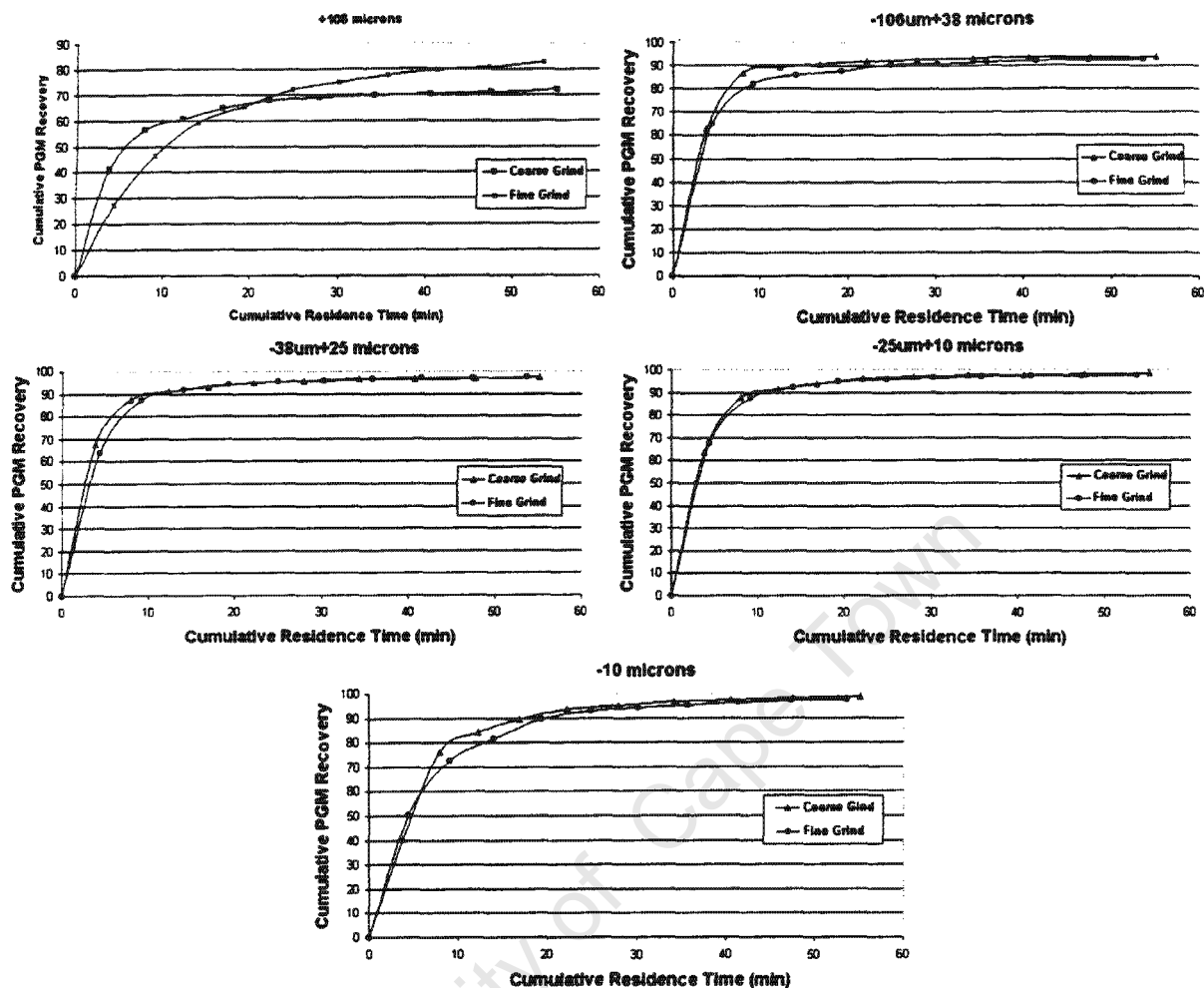


Figure 6.11: A comparison of the flotation rates between the coarse and fine grind feeds for each particle size class.

Figure 6.11 compares the flotation rates between the coarse and fine grind feeds for each particle size class. The initial flotation rates were always higher for the coarse grind, showing a higher recovery in the first flotation stage for all size classes. The rate of flotation of the $-10\mu\text{m}$ fraction was higher for the coarse grind for all residence times. It is possible that the fine grind feed produced extremely fine PGM grains in this fraction, over-grinding them which slowed down the flotation rates of this fraction even further. The $+106\mu\text{m}$ recovery time response was very different for the two grinds. The final recovery difference (at high residence times) is attributed to increased liberation. All the recovery time profiles in figure 6.11 were essentially the same except for the $+106\mu\text{m}$ fraction. A simple first order recovery expression for a perfectly mixed vessel ($R = k\tau / (1 + k\tau)$) was fitted to the recovery time profiles in figure 6.11 and the regressed rate constants were plotted as a function of particle size class in figure 6.12. The rate constants regressed from this equation ignore the influence of entrainment as well as the effect of the froth on the overall recovery. Figure 6.12 shows that the rate of flotation is smaller for all size fractions for the finer feed material, indicating that a regrind stage could well affect the rates of flotation of particles, and not just the relative mass in a

particular rate class, casting some doubt on the assumption that floatability parameters can be considered to remain constant through a regrind stage. However, this analysis is based purely on the effect of particle size class on the flotation kinetics. Floatability includes all aspects relating to the flotation rate of the particles, including liberation, and using figure 6.12 alone to evaluate this assumption is insufficient.

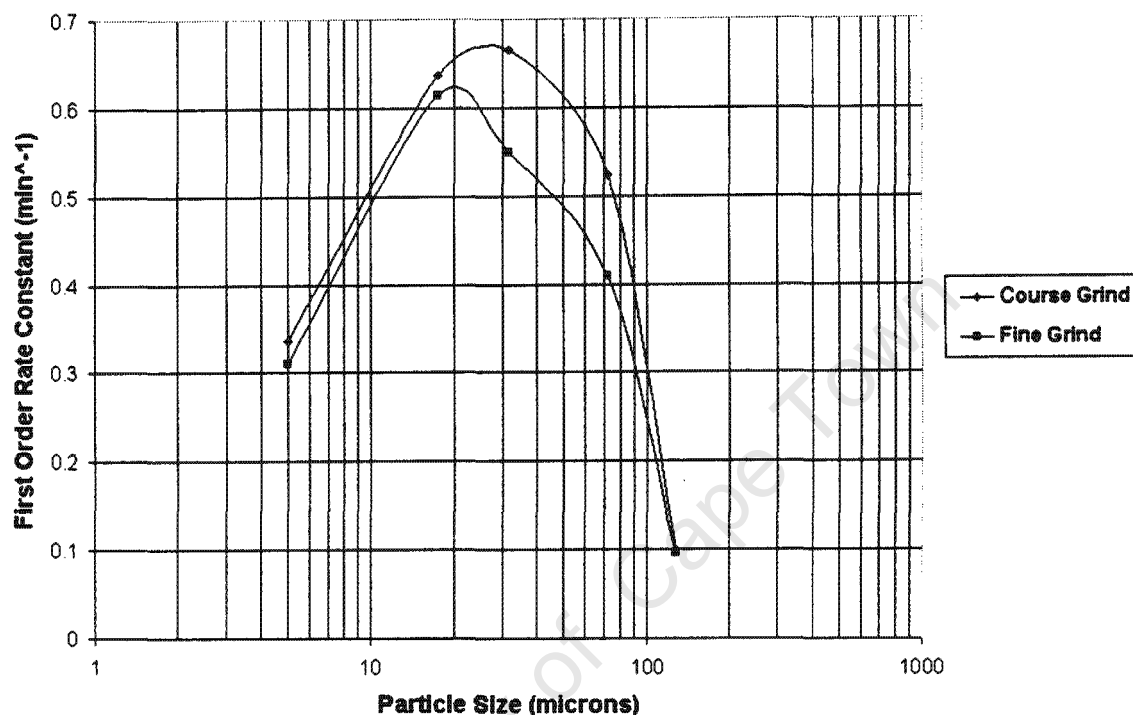


Figure 6.12: A comparison of the first order flotation rate constants with particle size for the coarse and fine grind streams

The objective of this work was to evaluate the correlation of the parameters regressed using the PGM modelling methodology presented in chapters 3 and 4 to the measured mineralogical characteristics of the ore. The preliminary discussion was presented to help provide some qualitative information on the difference between the flotation characteristics of the two feed materials to the pilot plant, as a basis for comparison with a more quantitative analysis.

The same analysis as described in chapter 3, section 3.4 was used in order to regress the parameters describing the PGM flotation kinetics. The assumption used by Runge (1997) that the floatability parameters (P 's) of the fast and slow floating fractions were the same for both feed streams was used. The effect of the different grind was assumed to shift the proportions of fast, slow and non floating fractions between the floatability classes. This was necessary to provide a basis for comparison between the regressed mass fractions. Different froth stability parameters were fitted to the rougher and scavenger banks as in the analysis of phase I tests. The parameter sets that produced the smallest error (TSSE) between the calculated and experimental data

obtained for the coarse and fine grind streams are shown in table 6.5. This table also includes the PGM model parameters obtained from phase I using the Impala UG2 ore. Figure 6.13 shows the overall PGM flow rates predicted by the model compared to the experimental data for both feed streams. The comparison of the predicted batch floatability profiles compared to the experimental batch floatability data is shown in figures 6.14 and 6.15.

Table 6.5: The PGM model parameters obtained from Phase I and Phase II tests on the pilot plant

PARAMETERS	LONMIN KAREE UG2 (PHASE II)		IMPALA UG2 (PHASE I)
	COARSE GRIND	FINE GRIND	
$P_{f,PGM}$	1.22E-3	1.22e-3	2.54e-3
$P_{s,PGM}$	1.06e-4	1.06e-4	2.43e-4
$m_{f,PGM}$	0.75	0.74	0.28
$m_{s,PGM}$	0.16	0.21	0.45
$m_{o,PGM}$	0.09	0.05	0.27
$\beta_{rougher,PGM}$	0.04	0.04	0.04
$\beta_{scavenger,PGM}$	0.06	0.02	0.05
α_{PGM}	0.02	0.02	0.02
E_{PGM}	0.33	0.59	0.54

The fit to the experimental data was excellent for both the pilot plant PGM flow rates as well as for the batch floatability tests even though the same floatability parameters were used to model the PGM kinetics for both streams. In total 136 data points (plant PGM flow rate data and batch tests) from both feed streams were used to regress 16 parameters. A model fit of 1.69 was obtained from the regression, and most of the predicted data points were within the experimentally determined standard deviations. A qualitative evaluation of the difference between the coarse and fine grind parameters suggested that they are sensible. Firstly, the PGM entrainment parameter for the coarse grind stream was smaller than that of the fine grind stream. This would indicate that a larger proportion of PGM's were entrained with water for the finer grind. This is expected, as entrainment is considered to be purely size dependent and therefore the higher fraction of finer PGM particles resulted in a higher regressed degree of entrainment for this feed stream. The entrainment parameter was assumed constant for all particle size classes in the development of the entrainment PGM model. The interpretation given here would indicate that the entrainment parameter is slightly weighted towards the finer particle size classes however. The entrainment parameter regressed from the work of phase I using the Impala UG2 ore lay in between the entrainment parameters regressed for the two feed streams in phase II. This correlates well with the conclusions drawn with respect to the phase II entrainment parameters, as the grind used for the Impala work was 60% -75 μ m which is in between the two grinds used in phase II. This suggests that the entrainment parameter varies directly with grind.

The next obvious difference in the parameters in table 6.5 is the smaller fraction of non floating material ($m_{o,PGM}$) for the fine grind stream. This can be ascribed to the liberation of some of the locked PGM particles in the larger size fractions, supported by the higher recovery of the +106 μ m fraction at high residence times in figure 6.11. The effect of finer grinding seemed to be detrimental to the amount of fast floating material in the feed stream as the fast floating fraction decreased slightly for the fine grind stream. This can also be supported from the recovery time response of the -10 μ m particles in figure 6.11. The fine grind recovery time profile lay below the coarse grind recovery time profile for all residence times. Figure 6.12 shows that a smaller first order rate constant was regressed from the fine grind recovery time profile compared to the coarse grind. This correlates well with the regressed decrease in the fraction of fast floating component in the feed. It should be noted that another possible reason for the decreased fraction of fast floating components could be due to increased surface area with fine grinding resulting in an increase in reagent adsorption on newly liberated hydrophobic sites and subsequently less reagents per surface area in the feed stream for the fine grind stream compared to the coarse grind stream.

The froth recovery parameters showed an increase in the froth stability parameter (β) between the rougher and scavenger bank indicating a decrease in the froth recovery of attached particles down the bank of flotation cells. Savassi (1998) developed a technique of measuring the froth recovery of attached particles and observed that froth recovery dropped from 70% at the head of a rougher to 10% at the tailings end of the plant on which he conducted the investigation (Mount Isa Copper Concentrator). A smaller froth stability parameter was obtained for the scavenger bank for the fine grind stream indicating that a higher froth recovery was achieved in this bank. It would be expected that a higher froth recovery would occur with respect to the finer ore.

Finally, the difference in the floatability distributions between the Impala UG2 ore and the LONMIN UG2 ore is highlighted in table 6.5. The LONMIN UG2 ore had a much higher fraction of faster floating components compared to the Impala UG2 ore. This can possibly be ascribed to the fact that the Impala ore had been exposed to the surface a lot longer than the LONMIN sample and could well have undergone some oxidation. The LONMIN sample was freshly mined and showed no signs of degradation. It must be noted that oxidation would be expected to affect the finer size fractions the most as these particles have the highest surface area. It should be noted that the Impala UG2 ore had a high proportion of -10 μ m particles. The large difference in the floatability distributions between these two UG2 ores is not an unfamiliar occurrence, and it is well documented that the mineralogy of UG2 ores is extremely variable and can change dramatically from location to location, (Liddell, 1986).

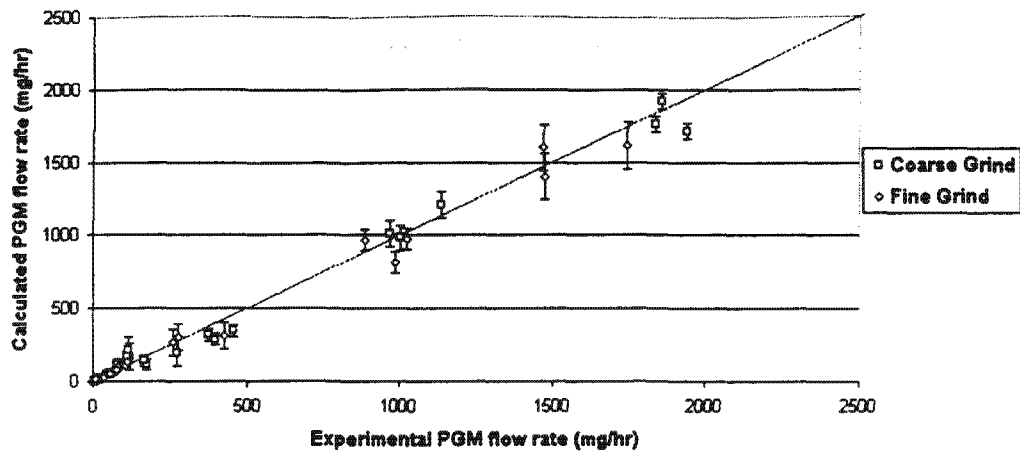


Figure 6.13: The comparison of the PGM flow rate predictions of the model compared to the experimental flow rates for the coarse and fine grind streams.

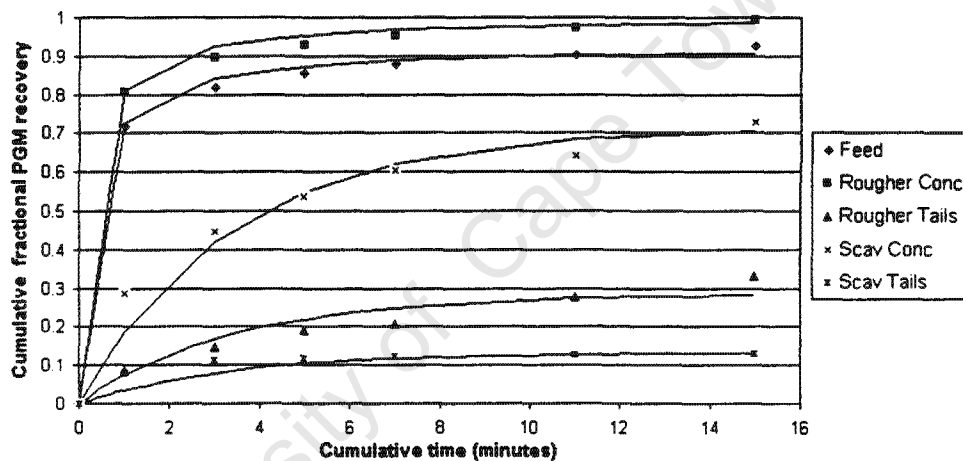


Figure 6.14: A comparison of the predicted and experimental batch PGM recovery time profiles for the coarse grind feed material

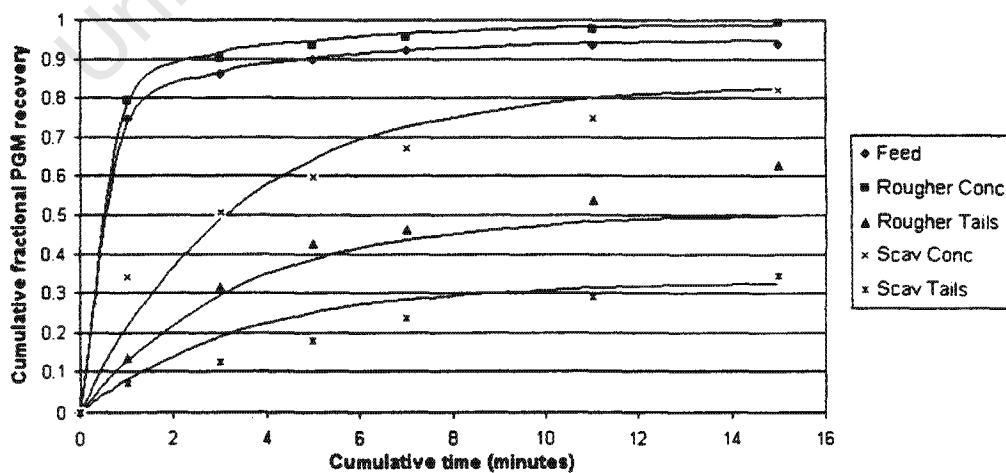


Figure 6.15: A comparison of the predicted and experimental batch PGM recovery time profiles for the fine grind feed material

In conclusion, a very good agreement was obtained from the PGM model compared to the experimental results from both streams, suggesting that the use of a constant set of floatability parameters can be used to model different grinds of the same feed material. This provides a basis for including regrind stages into the simulation of flotation circuits. The significance of the regressed parameter sets for the two streams has been discussed in conjunction with the analysis of the particle size recovery time profiles. The results suggest that the PGM entrainment parameter varies linearly with the percentage $-75\mu\text{m}$. The different floatability distributions regressed for phase I and phase II work highlights the extremely variable nature of UG2 ores.

6.3 Correlation of a derived floatability distribution with mineralogical data

At present, there is no direct method of measuring a floatability distribution. As such there is no direct method that can be used to verify the regressed parameters that describe the distribution of the various PGM floatability classes in the feed to a circuit. The regression of a floatability distribution is highly non-linear and it is possible that a number of different parameter sets could be obtained that could give a similar, satisfactory fit to an experimental data set. The final set of parameters obtained frequently relies heavily on good initial estimates of the parameters. The Monte Carlo Boot Strap technique has been used earlier in this thesis to address this issue. This section offers another approach to quantify the mass fractions of the floatability components obtained from the regression of the pilot plant data set. The method makes use of information obtained from the QEMSEM analysis of the feed material to the pilot plant during phase I, presented in chapter 4. The QEMSEM analyser makes use of computer based image analysis techniques, coupled with a scanning electron microscope to identify the mode of occurrence of the PGM particles in a sample. It was hypothesised that if enough information regarding the PGM particles in the feed to the pilot plant was known and if this data was used statistically it should be able to approximate the regressed PGM floatability distribution. This is the first attempt, as far as the author knows, that QEMSEM data has been used to quantify the regressed floatability distribution of a feed to a platinum flotation circuit.

The method presented here relies very much on the fact that the data obtained from the QEMSEM search is truly representative of all the particles in the feed to the pilot plant. Obviously there are millions of particles with differing characteristics feeding a flotation plant but statistically, there should be an overall "blue-print" of what the characteristics of the PGM particles are like, such as the average particle size distribution, extents of liberation, PGM association with various minerals, to name a few. The separation of particles into floatability components is based on information obtained from the QEMSEM analysis, such as whether a PGM particle is liberated, locked in gangue or base metal sulphides (BMS) as well as the expected rates of flotation based on particle size effects, which are backed-up from size-recovery data obtained from section 6.2.

The raw data from the QEMSEM analysis is presented in appendix F and is summarised in tables 4.5 to 4.9 in chapter 4. This data depicts every PGM particle that was detected from an average of four searches in nine size fractions. The total average number of grains (4 sections) detected in all size fractions was 310 and is listed in the top row of table 4.6. This is used as a basis for quantifying the PGM floatability distribution regressed in chapter 5, table 5.2. Table 6.6 below shows the number of particles detected in each size class. The information summarised in table 4.7 is used to proportionate these particles into different modes of occurrences. As an example, 58 percent by number of PGM particles in the $-10\mu\text{m}$ size class were found to be fully liberated. Therefore, of the 71 particles found in the $-10\mu\text{m}$ size class 41 (71×0.58) were considered to be fully liberated. The same calculation has been performed for all the particle size classes and for each PGM mode of occurrence in table 4.7.

Further information obtained from the QEMSEM data is used to separate these particles into fast, slow and non floatability classes. As a starting point, it was found that 66 percent of PGM-containing particles in the combined sample had a low float rank of less than 0.4 (section 4.5.1.2, float rank). Note that the float rank represents an indication of potentially floatable component area (PGM+BMS) divided by the total area of the particle. Therefore it has been assumed that less than 34 percent of the particles should be fast floating. This is not strictly true as the float rank does not take into account the inertial effects associated with particle size and density on the recovery. However, it does relate the floatability to the extent of liberation which would be expected to directly influence the rate of flotation.

Another assumption that was adopted in this analysis was that all PGM particles locked in silicates and oxides would be non floating PGM grains due to the hydrophilic nature of these gangue components. These are highlighted in dark grey in table 6.6 for all size fractions. Further, the QEMSEM analysis indicated that all PGM-containing BMS's larger than $106\mu\text{m}$ were locked in silicates or silicate/chromite chats and these were therefore regarded as non floating particles. Composite PGM/oxide and PGM/silicate grains larger than $106\mu\text{m}$ with the PGM exposed on the grain edges or boundaries were also considered non-floating. This is assumed to be as a result of the extremely small nature of the PGM grains in this ore ($85\% < 4\mu\text{m}$, $93\% < 6\mu\text{m}$) as well as the size of this hydrophobic PGM grain in comparison to the hydrophilic gangue constituent on a PGM/gangue particle. To substantiate this assumption, figure 6.16 shows the resulting rate of flotation as a function of particle size obtained from the LONMIN work (phase II). Particles in this ore greater than $106\mu\text{m}$ had an extremely slow rate of flotation and together with the fact that the ore from phase I was found to be somewhat weathered (oxidised) it was considered reasonable to assume that these particles had a very small probability of flotation. These assumptions allow all the expected non floating particles to be highlighted in dark grey in table 6.6.

Table 6.6: The proportionment of PGM particles into their floatability components

Total # Particles	310								
Particle size class (microns)	+0-10	+10-25	+25-38	+38-53	+53-75	+75-106	+106-150	+150-212	+212
Arithmetic mean particle size in size class (microns)	5	17.5	31.5	45.5	64	90.5	128	181	256
Total # Particles in Size Class	71	75	33	29	25	25	26	14	12
On oxide and silicate grain boundaries	0	0	0	0	0	1			
On silicate grain edges	1	7	3	3	0	2			
On oxide grain edges	2	5	0	0	1	2			
Locked in silicate									
Locked in oxide									
On BMS and silicate grain boundaries	0	0	4	4	2	2			
On BMS and oxide grain boundaries	0	0	1	0	0	0			
On triple point between BMS, oxide and silicate	0	1	0	0	0	0			
On BMS grain edges	10	18	3	6	2	1			
Locked in BMS	9	32	15	7	10	11			
Liberated	41	12	8	1	8	8	8	0	8
	71	75	33	29	25	25	26	14	12

The next, more complicated step was to separate the remaining grains into fast and slow floating fractions. The first assumption was that all PGM particles less than 10 μ m (except those locked in gangue) were slow floating. These are highlighted in light grey in table 6.6. It is well known that the major mineral recovery losses in many industrial UG2 flotation circuits are due to the low flotation rates of the particles less than 10 μ m, (Trahar, 1981), (Liddell et al, 1986). Figure 6.16 shows the slow flotation rates for the LONMIN UG2 ore in this size fraction. It must be emphasised that the particles in this size fraction are believed to float due to the high float ranks associated with increased liberation of smaller particles, but these particles require more energy per unit flotation cell volume to overcome the inertial forces required to attach to the surface of bubbles.

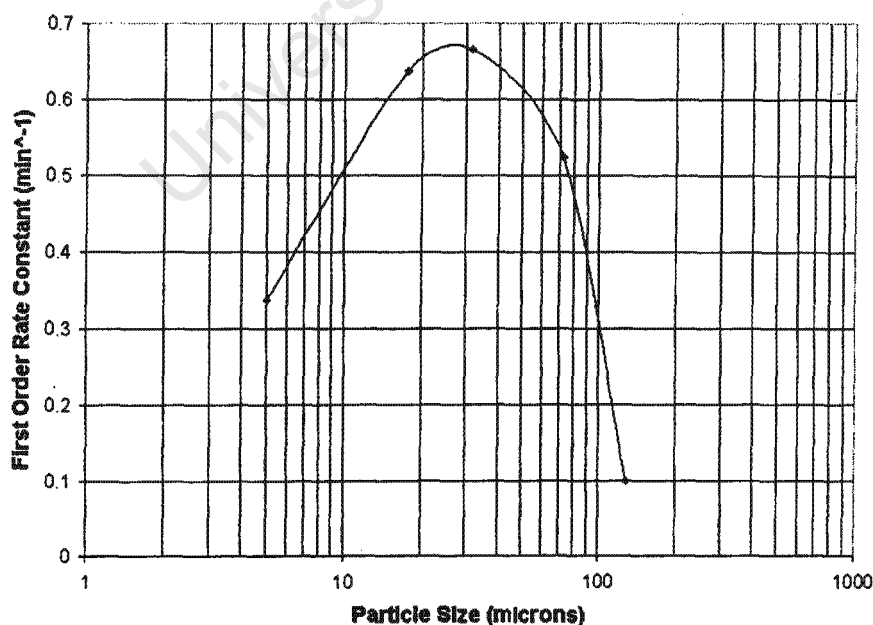


Figure 6.16: The overall flotation rate constant as a function of particle size, LONMIN UG2, phase II.

All composite PGM/gangue particles with the PGM exposed on the edge or boundary smaller than $106\mu\text{m}$ were considered to be slow floating due to the smaller size of the hydrophilic gangue mineral in association with the hydrophobic PGM particle. The gangue component is sufficient to restrict the flotation rate of these particles due to the reduced exposed surface available for attachment of collector as well as to bubbles. The composite PGM grains associated with gangue and BMS's on the boundary or triple point are also considered slow floating. Boundary refers to the occurrence of PGM grains between two mineral grains. The added complexity of these composite particles as well as their reduced float rank suggest that it is reasonable to assume that these particles would have slow recovery rates.

Fully liberated PGM particles larger than $10\mu\text{m}$ were found to have a high float rank in the mineralogical analysis, and these particles were therefore all assumed to be fast floating. The fast floating particles are represented by the unshaded region in table 6.6. The BMS particles are also hydrophobic, and therefore PGM grains associated with these particles in size classes coarser than $10\mu\text{m}$ have also been assigned as fast floating material. These particles are in the perfect size range to have sufficient inertial energy to overcome the film boundary surrounding bubbles and also have a sufficiently high float rank to render them fast floating. The final piece of evidence needed to complete the table in table 6.6 is the note made in section 4.5.1.2 that in the $+53\mu\text{m}$ size fractions, PGM-containing BMS's are mostly locked in silicates or silicate/chromite chats restricting virtually all the exposed hydrophobic surface. Therefore particles greater than $53\mu\text{m}$ but smaller than $106\mu\text{m}$ (where the size of the gangue particle associated with the composite grain is not overriding) have been assigned into the slow floating fraction.

Having proportioned all the particles into their respective floatability classes, the final thing to do is to add up all the particles in each shaded zone and calculate the fraction of each floatability class by dividing this number by the total number of particles detected. Table 6.7 presents the calculated floatability fractions compared to the regressed parameters obtained in chapter 5, table 5.2.

Table 6.7: A comparison of the calculated and regressed floatability components for the feed to the pilot plant, phase I

PARAMETER	CALCULATED	REGRESSED
Fast ($m_{\text{fast,PGM}}$) fraction	28	28
Slow ($m_{\text{slow,PGM}}$) fraction	40	45
Non ($m_{\text{non,PGM}}$) fraction	31	27

The agreement between the calculated and regressed parameters in table 6.7 can be seen to be very good. However, some care needs to be taken in judging the significance of the correspondence indicated by the above, rather qualitative, analysis. A more quantitative understanding of the role of mineralogical and liberation information in terms of a priori prediction of ore floatability is still some way off, particularly for ores

as complex as used in this study. Nonetheless, this analysis does offer an indication of the extent to which it may be possible to use such information to provide estimates of ore floatability in the design of new circuits.

6.4 Conclusions derived from this chapter

The purpose of this chapter was to evaluate the ability of the simulator to predict the flotation kinetics of two alternative circuit configurations as well as to predict the flotation kinetics with a change in residence time and size of a flotation bank. It was found that the water froth model parameters were the most sensitive to the change in circuit configuration. The significance of the empirical parameters in the water froth recovery model was evaluated with reference to its values regressed for each circuit configuration. The froth recovery parameters for PGM, chromite and gangue were found to be relatively robust during the simulation of the various circuits. However the gangue froth model parameters showed more variation. Overall, the results of both simulation exercises matched the experimental data very well.

Comparing the regressed PGM parameter sets for different feed grind sizes showed a good correlation between the fitted model parameters and measurable stream characteristics. It has been shown that this could provide a basis for including regrind in a simulation algorithm. It was observed that the PGM entrainment parameter varied linearly with the grind of the feed material, expressed on a % passing 75 μm basis.

The regressed PGM floatability mass fractions of the UG2 Impala ore of phase I were evaluated in conjunction with the information obtained from the QEMSEM analysis of the feed material to the pilot plant, with encouraging results. More work is required in this area.

AN OPTIMISATION STUDY USING THE SIMULATOR

The development of this thesis has progressed from describing the models that can be used to predict the flotation kinetics of the components in UG2 platinum ores, to the presentation of an algorithm that combines each of these component models so that they can be used in a simulator. The output of the simulator has been assessed with respect to its form and boundary conditions. The ability of the algorithm to simulate two different circuit configurations has been assessed by comparing the simulator results to experimental data. The simulator was shown to be capable of predicting the flotation kinetics for material feeding two banks of rougher cells of different sizes and at a reduced feed rate without needing to change the froth parameters describing the flotation kinetics in the different banks. However, it was shown that simulation of alternative circuit configurations was more problematic, owing to difficulties in assigning appropriate froth parameters during simulation. The simulations presented in this chapter are restricted to the limitations of the simulator as described above. Despite this, the work presented in this chapter gives an example how the simulator can be usefully employed in its present form.

7.1 Introduction to the optimisation test using the simulator

This chapter presents a case study in which the simulator is used to evaluate the effective use of the recycle mechanism in re-circulating separation systems. The application of the multiple product recycle mechanism is discussed with reference to common platinum flotation circuit designs.

There is very little discussion in the literature as to why the practice of recycling middling streams within a circuit, without fundamentally changing the properties of the recycled particles, often proves to be successful. Many rule of thumb attempts at describing this practice have been based on simple qualitative concepts such as; 'return the stream to the point where its assay value and water composition are nearest those of the stream that it joins', (Taggart, 1945), (Jowett and Sutherland, 1985). In the same paper, Jowett and Sutherland (1985) recognised that circuit configurations or interconnections, especially the pattern of the recycle streams is an important control issue in the design and operation of complex circuits. It has been observed that the greater the number of stages and the higher the circulating loads, the better the separation, (Imaizumi and Inoue, 1965), (Loveday and Marchant, 1972). Lauder (1991) made an attempt to highlight the principle behind the recycle mechanism using a relatively simple mathematical explanation. The conclusions from this work were that a network of flotation cells with recycling streams aids in the selectivity of valuable components from gangue components, providing combinations of recovered valuable and gangue components that are unobtainable with the same number of stages in an open network. He concluded that; "The recycle mechanism therefore operates by effectively enhancing the

selectivity of the species with the higher flotation probability at the expense of the species with the lower flotation probability”.

It is the purpose of this chapter to explore the ‘recycle mechanism’ in terms of the separation of a complex ore. It is proposed to consider a complex ore as an ore that contains a large range of particles that float at different rates. This is particularly true for the UG2 platinum ore used as the basis for this study. In previous chapters it has been shown that the ore can be considered to consist of a number of distinct floatability classes, with relatively large proportions of fast, slow and non floating (locked or slimed) components present in the feed. As a result, a recent trend in the design of new UG2 concentrators is to produce multiple concentrate streams where, in effect, the floatability component classes can be considered to be concentrated in separate cleaning stages in the circuit. The results and conclusions drawn in this chapter are based on trends and observations made using the simulator developed in the previous chapters. The results from this evaluation clearly highlight the validity of this trend in the design of UG2 circuits, despite its largely “evolutionary” nature.

7.2 The assumptions made and restrictions imposed on the simulation algorithm

The performance of two flotation circuits is evaluated, the first circuit in figure 7.1 was the same as circuit configuration A as described in figure 4.8 and will be called the ‘Mintek’ circuit for the purpose of this chapter. This circuit produces two separate concentrates, a high grade concentrate which can be considered to comprise of mostly the fast floating components, and a second concentrate comprising the slower (middlings) floating components. The second circuit is a standard rougher/cleaner circuit, with recycled cleaner tails going to the head of the rougher bank. This circuit cleans all the rougher concentrate in a single stage as opposed to the Mintek circuit that cleans some of the rougher concentrate in a high grade cleaner (HGC) bank and some of the ‘scavenger’ concentrate and HGC tails in a combined low grade cleaner (LGC) bank. The other major difference between these two circuits is the position of recycled material to the rougher bank. The Mintek circuit recycles material to the head of the scavenger bank while the standard circuit recycles material to the head of the rougher bank. The importance of these differences will become apparent from the results of the evaluation.

The parameters describing the Mintek circuit have been regressed previously in this thesis. The standard rougher/cleaner circuit was virtually identical to circuit configuration C as presented in figure 4.19, the only difference being that the recycle stream was sent to the rougher B bank and not rougher A. It was assumed that this change was sufficiently small to simulate this circuit configuration relatively accurately using the parameters regressed for circuit configuration C in chapter 6. In order to simplify the calculation, only single flotation cells were used in the optimisation study and not a bank of flotation cells as indicated in figures 4.18 and 4.19.

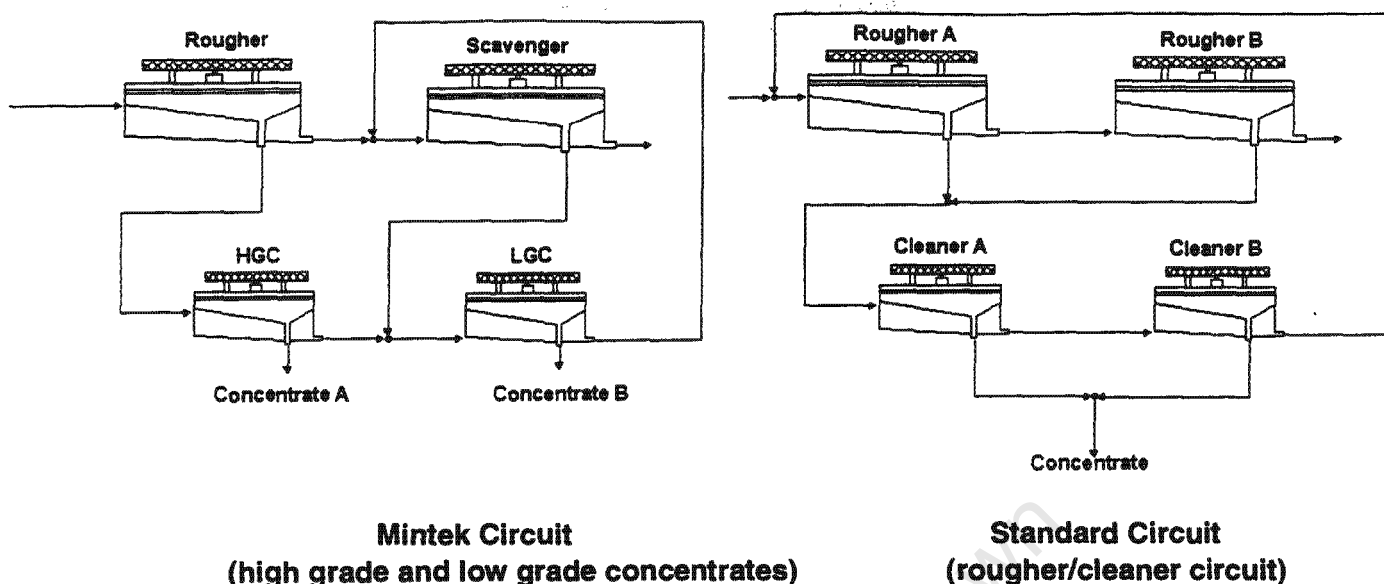


Figure 7.1: The two circuit configurations compared in the optimisation study

The performances of these two circuits were compared using solver to search for the maximum overall recovery from each circuit for a specified overall grade. To do this the solver subroutine was allowed to vary the volumes of the individual flotation cells in figure 7.1 for both circuits while keeping the total circuit volume constant (180 litres). This procedure was repeated for a number of specified grades in order to determine the grade versus recovery curves for the two configurations in figure 7.1. A few constraints were required to perform this analysis, as indicated below:

- The flotation cell dimensions were always perfectly square.
- The froth height for any calculated flotation cell size was fixed at 20 percent of the cell depth.
- The flux of bubble surface area per cross sectional area of the flotation cell (S_b) was kept constant for any volume of flotation cell.
- The volume occupied by the impeller stator was calculated as a constant percentage of the total cell volume, having the same proportions as the pilot plant cells.
- Based on the constant S_b and froth dimensions, it was considered reasonable to assume that the froth recovery would be constant for any flotation cell size for each flotation cell in the two circuits.
- The feed volumetric flow rate and composition was kept constant for all the calculations, for both circuits.

7.3 Results from the optimisation study

Figure 7.2 shows the resulting grade-recovery curves obtained from the evaluation of both circuit configurations. This figure clearly shows that the Mintek circuit was able to outperform the standard circuit, especially at higher grades.

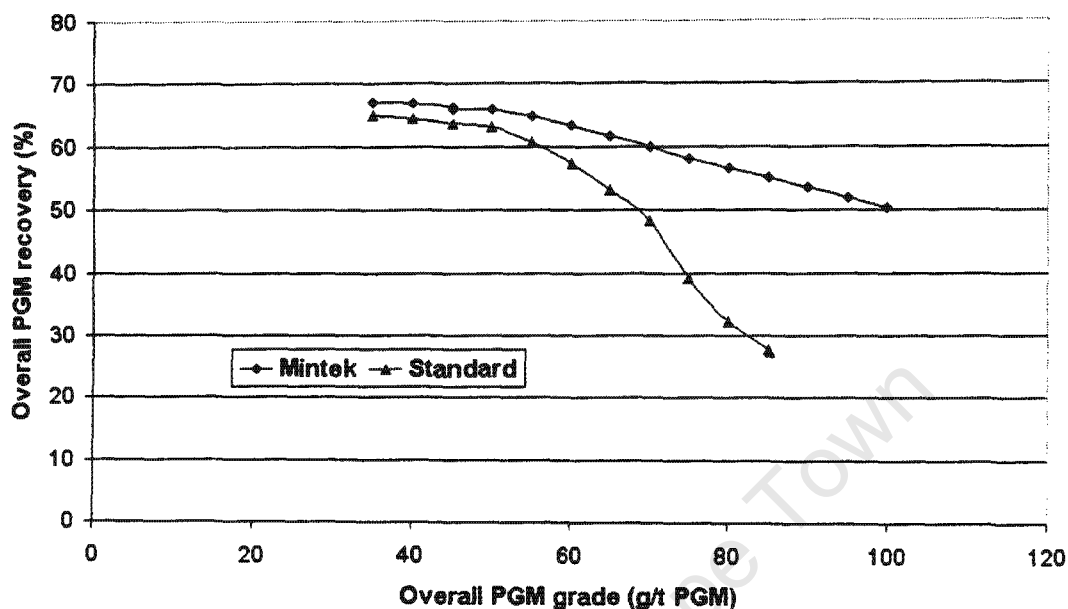


Figure 7.2: A comparison of the grade recovery curves for the Mintek and Standard circuit configurations

The two grade recovery curves in figure 7.2 approach one another at low grades which tends to indicate that the performance of the two circuits converge at higher recoveries and lower grades. Figure 7.3 shows the optimum distribution of flotation cell volumes at high recoveries for both circuits. The dashed lines indicate very small volumetric flow rates from both circuits; highlighted are the major volumetric flows. This clearly shows why the two grade recovery curves in figure 7.2 approach one another at higher recoveries.

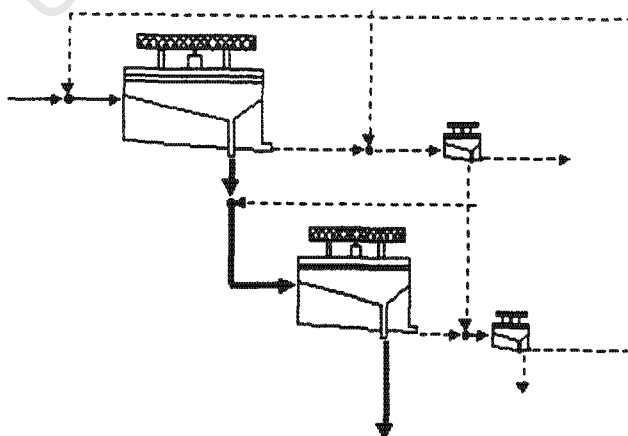


Figure 7.3: The distribution of flotation cell volume for both circuit configurations at maximum recovery

At high recoveries both configurations approach a single large rougher bank (for recovery) and a large cleaner bank that virtually recovers all the rougher concentrate. A very small circulating load exists for this extreme case, which adversely affects the grade. The Mintek and Standard circuits are identical at lower grades producing the same maximum recoveries. Figure 7.4 shows the distribution of flotation cell volumes for the Mintek circuit at the other extreme end of the grade recovery curve (maximum grade).

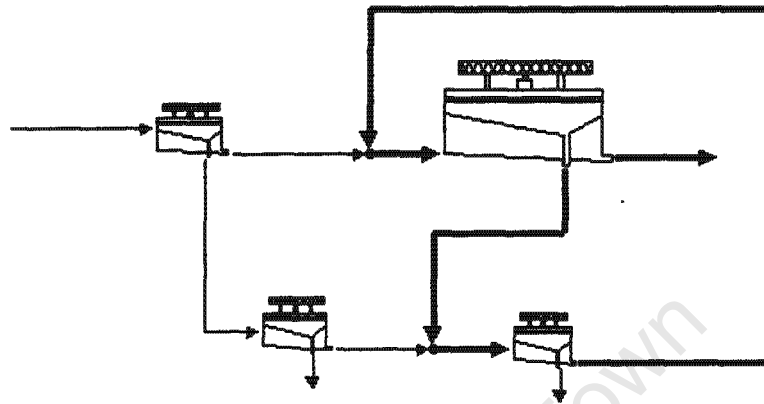


Figure 7.4: The distribution of flotation cell volumes for the Mintek circuit at the maximum grade

With the Mintek circuit running at maximum grade, figure 7.4 shows the optimum distribution of flotation cell volumes. The rougher and HGC banks are just large enough to pull off the fast floating material while a very large scavenger bank creates a high circulating load in order to enrich the slower floating “middling” components. The LGC bank is just large enough to obtain the maximum recovery at the set grade. The magnitude of the circulating load (based on feed volumetric flow) in the two circuit configurations was the only variable that was changing throughout the grade recovery curves in figure 7.2. In order to understand why the grade-recovery curve of the standard circuit dropped below the Mintek circuit, the composition of the circulating load (percent of each PGM floatability component in the circulating load) was plotted along the grade-recovery curve for the two circuits in figure 7.5.

Figure 7.5 shows that at high recoveries and low grades the composition of the recirculating stream consisted primarily of non-floating particles. As grade increases, the circulating load increases. With this, the percentage slow floating components start increasing, displacing the amount of non-floating material. As the grade increases, a point is reached where the circulating load becomes so large that the fast floating components are swept completely through the cleaner cells, resulting in the breakthrough curve of fast floating components in figure 7.5. At this point, the final concentrate consists only of fast floating material that is pulled off slow enough (at the expense of recovery) in order to obtain the maximum grades. The most important thing to note here is the point at which the breakthrough of fast floating components occurs between the Mintek and Standard circuit. The Mintek circuit can accommodate much higher circulating loads before the fast floating components are ‘washed through’ both cleaner banks and are recirculated.

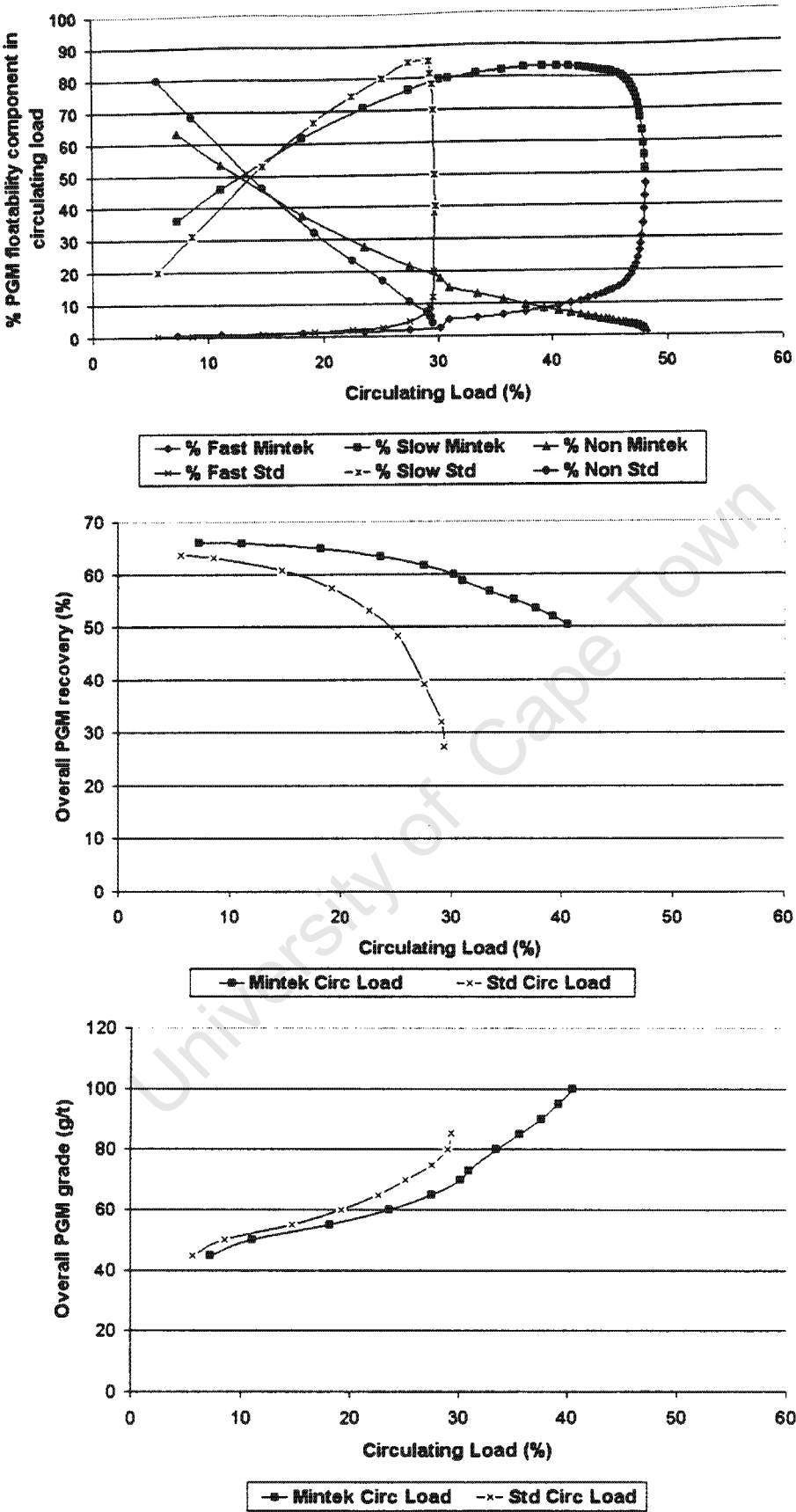


Figure 7.5: The variation of the percentage floatability component, PGM recovery and grade with circulating load for both circuit configurations

Figure 7.5 also shows the grade and recovery as a function of the circulating load for the two circuits. The reason for the faster decline in recovery with grade (figure 7.2) in the standard circuit is as a result of the smaller circulating load allowable before the breakthrough of the faster floating components. If a maximum limit was not imposed on the total circuit volume, figure 7.5 shows that infinite grades would be obtained at infinite circulating loads for both circuits.

It is very important to characterise the chromite recovery kinetics in UG2 ore flotation, as chromite causes problems during the later smelting stages of the concentration process. In fact, the overall design of UG2 flotation circuits are aimed specifically at minimising the recovery of chromite into the final concentrate. It was therefore important to determine how the overall percentage chromite in the concentrate varied while the circulating load was changed during the optimisation tests described previously. Figure 7.6 shows the percentage chromite in the final concentrate as a function of the circulating load. The percentage chromite decreases with an increase in circulating load for both circuit configurations. The conclusion from the work of Lauder (1991) was that a network of flotation cells with recycling streams aids in the selectivity of valuable components from gangue components, providing combinations of recovered valuable and gangue components that are unobtainable with the same number of stages in an open network. Therefore the recycle mechanism was operating by effectively enhancing the selectivity of the species with the higher flotation probability (PGM particles) at the expense of the species with the lower flotation probability (floatable fraction of chromite).

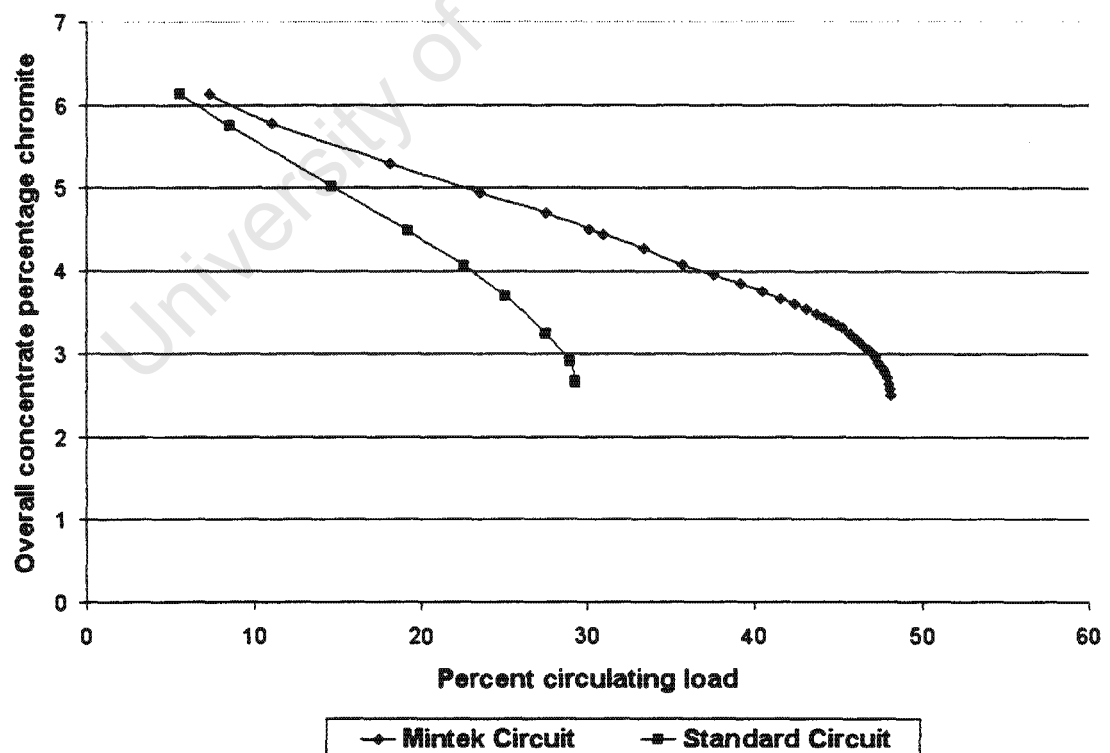


Figure 7.6: The decrease in the concentrate percentage chromite with an increasing circulating load

7.4 Discussion of the optimisation study results

The results obtained from this simulation exercise support the findings of Lauder (1991) that the recycle mechanism operates by enhancing the selectivity of the species with the higher flotation probability. This is depicted in figure 7.5 whereby the non-floating material is displaced by the medium floating material and finally this being displaced by the faster floating material as the recirculating load is increased to achieve higher grades.

The comparison of these two circuits highlights the extra degree of freedom obtained in controlling the grade and recovery from the Mintek circuit. The faster floating material is allowed to by-pass the circulating load while the middling floating components are allowed to gradually enrich themselves in the circulating load, without passing into the tailings stream due to the large scavenger bank. This material is then recovered by the low grade cleaner bank. In the case of the standard circuit, this extra degree of freedom is lost as the fast floating components in the circulating load effectively 'suppress' the middlings floating material (via the recycle mechanism), resulting in lower maximum recoveries compared to the Mintek circuit at the same grades. The breakthrough of fast floating PGM particles occurs at lower recirculating loads because the recycle stream affects all the flotation cells in this circuit. The lower the circulating load, the lower the final PGM grade of the concentrate. In addition, figure 7.6 shows that the PGM floatability components 'squeeze out' the amount of chromite in the final concentrate according to the recycle mechanism principle which is favorable for UG2 concentrations. It is interesting to note that in the work on the optimisation of flotation networks by Green (1984), every optimised circuit configuration had multiple concentrate product streams as that described by the Mintek circuit. This work did not investigate the possible reasons as to why this result was obtained, however.

It is apparent from this work that the Mintek circuit only really outperforms the standard circuit at higher grades and lower recoveries. This is not really useful as most plants, especially platinum concentrators, are highly recovery 'driven', owing to the high value of the concentrate produced. However, it is possible to extend this principle to two stage cleaning circuits to address this issue.

Most platinum flotation plants are designed with very high cleaner residence times with a cleaner to rougher residence time ratio of 3 to 3.5. However, the recleaner to rougher residence time ratio is typically only around one. This practice has become a rule of thumb for the design of these plants as a result of past experience and online optimisations on a variety of plants. Figure 7.7 shows a typical flowsheet for the majority of UG2 cleaning circuit configurations found in industry. Included in this figure are the typical pulp residence times for the various flotation banks in the circuit.

Initially it is not apparent as to the logic behind the high cleaner residence time in figure 7.7. Using the simulator and the results obtained from the Mintek circuit, it becomes apparent that the high volume of cleaner capacity is to increase the recovery of the high fraction of slower floating material commonly found in this ore. This is done by allowing a high circulating load to develop between the cleaner and

recleaner banks. The rougher bank is run purely for recovery with a small circulating load from the cleaner bank. This circuit can easily be adapted to include the Mintek process as shown in figure 7.8. Liddell (1986) stated that the fine sulphides in industrial UG2 flotation circuits have slow flotation rates, and the cleaner circuit must have a large volume to ensure that the residence time of these particles is sufficiently long, otherwise they recirculate to the rougher stage and are essentially lost to the rougher tailings stream.

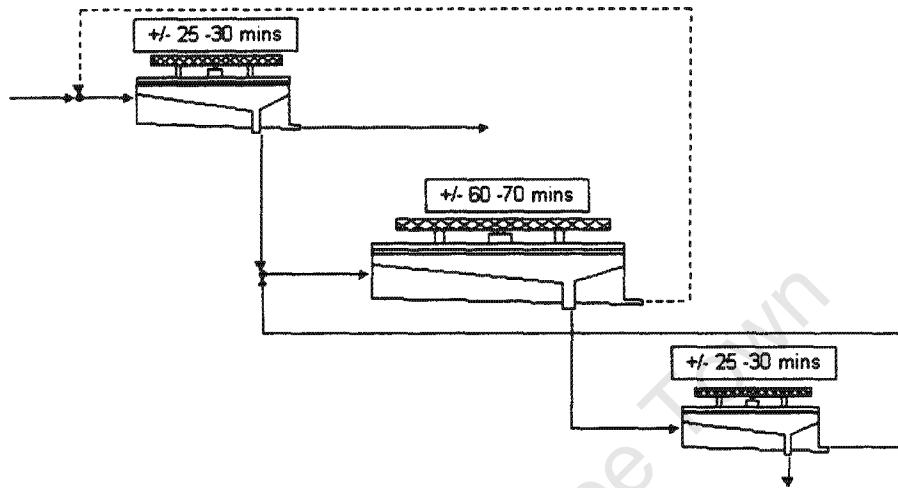


Figure 7.7: A common UG2 cleaner circuit configuration depicting typical pulp residence times in each bank

If the first cell to the cleaner bank is allowed to produce a final high grade concentrate and the recycled stream sent to the second cell on the cleaner bank, the fast floating material will be effectively separated from the circulating load. A high circulating load of only slower floating material will develop between the two cleaners, while a small circulating load to the rougher bank will ensure high recoveries.

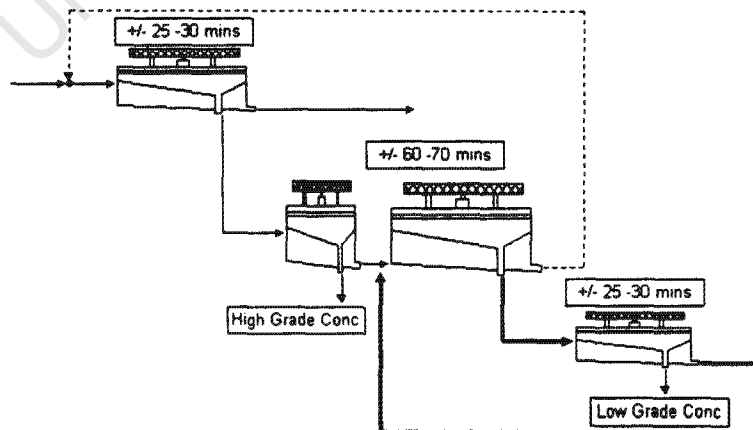


Figure 7.8: Depicting how the principle behind the Mintek circuit can be applied to existing UG2 circuit configurations

It is worth noting that there are some disadvantages to operating a plant with very high circulating loads. A general belief by many operators is that a certain amount of degradation (usually ascribed to oxidation) occurs to recycling material through every pass. It is widely held that the longer material stays in the recirculating load the lower its flotation rate. It should be noted that this phenomena was not apparent in any of the data generated in this study, where floatability was shown to be conserved with respect to all the UG2 ores that were used. However, many UG2 concentrators either discard the cleaner tailings stream or send it to a regrind mill for further liberation. The belief is that if the material in the cleaner tailings stream did not float even after a long residence time, it will not float at all. A common phrase exists, 'retreat rather than recycle'. This study suggests that this may not be the case. Another disadvantage of high circulating loads is that larger equipment is required, and hence capital costs are higher. In addition, control becomes more difficult. Obviously a trade off exists between the size of an allowable circulating load and the capital costs involved. The simulator presented in this thesis could provide a useful basis for such an evaluation.

7.5 Conclusions

This chapter has highlighted the advantages of the recycle mechanism in continuously operated recirculating flotation circuits. For ores with a high fraction of slow floating particles the separation of the fast floating and slow floating components during cleaning allows a high circulating load of only slow floating PGM particles to develop without the faster floating material being recycled. This allows higher grades to be achieved at the same recoveries. The separation of the floatability components gives an extra degree of freedom in controlling the grades and recoveries from a plant. The implications from this work suggest that it would be possible to achieve higher grades at the same recoveries from existing plant operations treating UG2 platinum ores.

Even though the simulator that has been developed in this thesis is restricted to small changes in circuit configurations as a result of a lack of fundamental water recovery models in the literature, it has been shown that valuable information can still be obtained from semi-empirical correlations of the flotation process.

Chapter 8

CONCLUDING REMARKS

The primary objective of this thesis was to investigate whether a kinetic simulation methodology could be developed to simulate the kinetics of the flotation pilot plant facilities at the Council for Mineral Technology (Mintek). The emphasis of the work was to determine whether existing models in the literature could be used to simulate the flotation kinetics of a platinum ore obtained from the Upper Group 2 Chromitite Seam (UG2) reef of the Bushveld Igneous Complex in the North Western Province of South Africa. The thesis was initiated in order to gain a greater understanding as to the reasons behind the recent circuit design trends of many new industrial platinum concentrators. Over a period of time the typical circuits have evolved from a conventional rougher/cleaner/recleaner circuit to what has been called the "Mintek" circuit for the development of this thesis. It was necessary to look at this problem from a simulation point of view to understand the mechanism behind this design trend in order to understand why experience tends to point in this direction. In addition, a simulation model could be used to exploit this principle to its full capability and could possibly be used to design even better circuits by providing a deeper understanding of the principles guiding the operation of these circuits.

In order to investigate this, the need for a relatively simple and practical model of the flotation pilot plant facilities was required. A robust and stable circuit simulation methodology that could converge rapidly to a solution and could be able to predict the response of a flotation circuit with acceptable accuracy was required. The development of this thesis progressed from describing the individual models that were used to predict the flotation kinetics of each major component found in UG2 platinum ores. An algorithm was then presented that combined each of these component models so that they could be used in a simulator. The output of the simulator was assessed with respect to its form and boundary conditions. The ability of the algorithm to simulate two different circuit configurations was determined by comparing the simulator's output to experimental data. It was found that a limitation of simulating alternative circuit configurations was that the kinetics of the froth phase needs to be known for different banks in a circuit, particularly with respect to the water froth model, before reliable results could be obtained from the simulator. The simulator was capable of predicting the flotation kinetics for material feeding two banks of rougher cells of different sizes and at a reduced feed rate without needing to change the froth parameters describing the flotation kinetics in the different banks. This provided the tool to investigate the advantages of the Mintek circuit configuration by using the simulation algorithm developed in this thesis in an optimisation study.

The conclusions developed through the progression of this work start in section 8.1 where the preliminary component models are presented. Section 8.2 discusses the conclusions of the water model developed in this thesis as well as the conclusions drawn from the development of the simulation algorithm. Section 8.3 and 8.4 briefly

describe the conclusions obtained from the work that was done to validate the simulation model as well as the implication and insights obtained from the optimisation study conducted using the simulator.

8.1 Conclusions from the individual component modelling work

The underlying assumptions in the simulation methodology had to be initially tested and the conclusions from this work are discussed in this section. The models used to predict the recoveries of each component in the UG2 ore are described and the conclusions from the application of the individual component models are discussed. All the individual component modelling work made use of the measured water recoveries as well as pulp and froth residence times to obtain a model fit. This modelling exercise was a preliminary step to obtain initial estimates of the model parameters in each component model. These were used in the simulation methodology whereby these variables were refined using an overall iterative solution algorithm.

8.1.1 Conservation of floatability

Conservation of the PGM floatability components was observed with respect to all the ore samples used in this study, as determined using the batch technique proposed by Runge et al (1997). It is concluded that the floating PGM particles maintained their rates of flotation throughout the circuit, and that the effect of possible surface oxidation did not influence the flotation rates of the PGM particles. A point worth noting is that it has been assumed that the various floatability components do not interfere with one another in various parts of the plant as well as within every batch floatability test. This assumption has not been validated but it is concluded in this work that due to the very dilute nature of the floatable minerals in UG2 ores, this influence should not be a major concern. The assumption that there is no interaction between each component seems to be justified.

8.1.2 The model used to simulate the PGM flotation kinetics

The preliminary PGM model developed for the mini plant in chapter 3 did not include an entrainment contribution in the overall recovery of the PGM minerals. Analysing the fit to the data for the mini plant tests indicated the need to include an entrainment contribution as a result of the progressive under prediction of the PGM flow rates down both the rougher and cleaner banks. An entrainment model was subsequently included in the analysis of the pilot plant work, using the assumption that all PGM floatability components were subjected to the same degree of entrainment with water. This represented a simplification of the true extent to which the PGM particles would be entrained into the concentrate. The approach used was adopted to limit the number of parameters describing the recovery of PGM particles. This nonetheless represented a more realistic model by allowing a degree of entrainment contribution particularly for the

non-floating fraction of PGM particles that could not be recovered in the preliminary mini plant PGM model. A more realistic approach would have been to use a size by liberation PGM model having a fraction of each size by liberation class subjected to entrainment which is a function of an entrainability distribution with particle size class. The approach adopted in this thesis was used to attempt to avoid the need for such a complex and detailed analysis. The equation that follows depicts the overall recovery of PGM particles that was used to simulate the PGM flotation kinetics.

$$R_{k\text{ PGM}}^o = \sum_{i=0}^n m_{i,\text{PGM},k} \left[\frac{P_{i,\text{PGM}} S_{b,k} \tau_{c,k} R_{f,\text{PGM},k}}{1 + P_{i,\text{PGM}} S_{b,k} \tau_{c,k} R_{f,\text{PGM},k}} + \left[1 - \frac{P_{i,\text{PGM}} S_{b,k} \tau_{c,k} R_{f,\text{PGM},k}}{1 + P_{i,\text{PGM}} S_{b,k} \tau_{c,k} R_{f,\text{PGM},k}} \right] * R_{w,k} * E_{\text{PGM}} \right]$$

The methodology developed by Harris (1997) in which batch flotation tests are used in combination with plant mass balance data to regress the parameters describing the PGM floatability distribution was found to work very well. This method provided a highly constrained, non-linear optimisation problem, that was found to be amenable to solution in a spreadsheet. A three step algorithm was developed whereby the parameters describing the PGM flotation kinetics could be effectively regressed. This comprised using only the batch floatability tests (obtained directly from streams in the steady state flotation circuit) initially as mass balance tools in order to get a first approximation of the feed stream floatability mass fractions. These values were then used as initial parameter estimates while using the plant mass balance data to regress the plant floatability and froth characterisation parameters. The third step was then to regress all the parameters together to obtain the parameter set providing the best overall model fit. This approach proved computationally robust despite the highly non-linear nature of this problem. A Monte Carlo Boot Strap analysis was conducted on the regressed PGM parameters. The froth stability parameters were shown to be the most influential to the resulting model fit. Only three floatability classes (fast, slow and non-floating) were required in the PGM model. From analysing the PGM parameters regressed for the mini plant it was evident that the slow floating fraction regressed during this phase of the study was essentially a non-floating fraction that was recovered largely by entrainment. The model fit for the PGM flow rates for the pilot plant tests can be considered satisfactory. The experimental error associated with the faster floating PGM fractions was found to be greater than the slower PGM fractions recovered in the scavenger and low grade cleaner banks. The PGM froth recovery model proved to be capable of providing a satisfactory fit to the pilot plant froth characterisation experimental data.

The one major limitation of the PGM model was the poor prediction of the concentrate PGM flow rate in the first flotation cell in the cleaner bank. A highly stable froth containing a high concentration of floatable PGM and gangue particles made the flotation kinetics in these first flotation cells inconsistent with that observed in the rest of the plant. It is likely that the kinetics in these cells was inhibited due to overloading of the froth zone (and/or the pulp zone) which could not be accounted for in the current models. This requires more work.

Very different PGM floatability distributions were regressed for the three different UG2 ores investigated in this thesis. The large difference in the floatability distributions was not unexpected, as it is well documented that the mineralogy of UG2 ores is extremely variable and can change dramatically from location to location, (Corrans et al, 1982), (Liddell, 1986).

8.1.3 The model used to simulate the chromite flotation kinetics

A linear relationship between the chromite and water recovery was found for each flotation bank for both the mini and pilot plant campaigns. This entrainment phenomenon has been widely observed in the literature. However, it was also found necessary to allow for the recovery of a small fraction of the chromite by true flotation, either owing to activation, or locking with sulphides and/or PGM particles. A hyperbolic secant equation developed by Savassi (1995) was found to well describe the degree of entrainment as a function of particle size for chromite. It can be concluded that realistic model parameters were regressed based on the correlation between the regressed and the experimentally determined chromite feed particle size distributions on the mini plant. The equation that follows presents the chromite recovery model used in the simulation model. The two separate contributions to the overall recovery of chromite are represented, the first being the entrainment contribution and the second the floatable fraction contribution.

$$R_{Cr,k}^o = (1 - \phi_{act,Cr}) * \sum_{i=1}^{d'} (E_{Cr,d'} * R_{w,k}) + \phi_{act,Cr} * \left(1 - \frac{1}{1 + P_{Cr} S_{b,k} \tau_{c,k} * ((1 - \alpha_{Cr,k}) * \exp(-\beta_{Cr,k} \tau_{f,k}) + \alpha_{Cr,k})} \right)$$

An extremely good model fit for all the tests on both the mini and pilot plant, including the alternative circuit configurations was obtained using this equation. It must be emphasised that the accuracy of this model depends on the accuracy of any water recovery predictions. The entrainability curve indicated that chromite particles greater than 50µm would not be entrained by water on the pilot plant. The disadvantage of using a ball mill as a grinding media has been highlighted from analysing the feed particle size distribution of chromite between the mini plant and pilot plant tests. The mechanism of grinding in a ball mill tends to produce a higher proportion of fine grained chromite. The use of a rod mill on the mini plant substantially reduced this problem.

8.1.4 The model used to simulate the gangue flotation kinetics

The preliminary model describing the recovery of gangue on the mini plant was not satisfactory giving a relatively poor model fit to the experimental data. This can be ascribed to the complicated mechanism by which gangue is recovered into a flotation concentrate. Despite this, the regression indicated a higher degree of entrainment for

the gangue compared to chromite, as would be expected given the lower specific gravity of the gangue compared to chromite. A more elaborate model was presented in chapter 4 for modelling the gangue flotation kinetics of the pilot plant. In this model the floatable fraction of gangue in the feed was assigned a distribution of flotation rates. It was assumed that even though a very large range of different floatable gangue minerals exist in the feed, having different properties (particle sizes, surface properties, extents of liberation, particle densities, shapes etc.) the response of these minerals to flotation can still be described using a distribution of floatabilities, such as was used for the PGM particles. The equation that follows shows the gangue recovery equation used in the simulation algorithm. The first part of the equation defines the recovery due to entrainment as a function of particle size and the second term defines the recovery due to the floating gangue particles based on the gangue floatability distribution of the feed material to the plant.

$$R_{Ga,k}^o = (1 - \phi_{act,Ga}) * \sum_{i=1}^{d'} (E_{Ga,d'} * R_{w,k}) + \phi_{act,Ga} \sum_{i=1}^n m_{i,Ga,k} \left(1 - \frac{1}{1 + P_{Ga,i} S_{b,k} \tau_{c,k} * ((1 - \alpha_{Ga,k}) * \exp(-\beta_{Ga,k} \tau_{f,k}) + \alpha_{Ga,k})} \right)$$

A good model fit was observed using this equation, allowing for two floatable gangue fractions (fast and slow). The slower flotation rates (P 's) of the floatable gangue fraction compared to the floatable PGM components shows the potential of separation for these two components. The added contribution of the gangue recovery by entrainment lowers the enrichment of PGM concentrates substantially however. It would be expected that the mechanism by which gangue is recovered into a flotation concentrate should be more complex than both the recovery mechanisms of the PGM and chromite particles. This is because the gangue particles (defined in the analysis of this thesis as all the dry mass excluding the PGM and chromite particles) have a range of densities and surface characteristics over and above a particle size distribution. The good correlation of the model to the experimental data seems to indicate that all the entrained gangue particles have sufficiently similar properties in UG2 ores (ie: no widely varying particle densities, shapes etc.) and that the only main variable affecting the entrainment of the gangue is the size distribution of the particles.

The fraction of floatable gangue particles in the feed to the pilot plant work of phase I tests correlated well with the work of Liddell et al (1985) who found that on average 4.85% of the concentrate of UG2 ores is associated with non-PGM bearing sulphides whereas a value of 5.0% was calculated from the parameter regression of the pilot plant data. According to the model parameters, gangue particles larger than 50 μ m do not get entrained by water into the concentrate. Particles larger than this are recovered as composite gangue-sulphide particles by true flotation. A higher proportion of the <50 μ m gangue particles are entrained into UG2 concentrates as opposed to chromite as a result of the lower average specific gravities of the gangue constituents.

8.2 Conclusions obtained from the water model and simulation algorithm work

8.2.1 The empirical water model

The importance of accurately describing the water kinetics cannot be underestimated. This thesis has stressed the central role of the water recovery model in accurately simulating the performance of flotation circuits. A number of empirical water models were evaluated in this thesis for inclusion into a simulator. None of these models were found to be suitable for use in this study. As a result, an empirical water model that could account for the effect of cell operating conditions was developed in this thesis, based on the froth residence time of air. The equation that follows describes the overall water recovery model developed for inclusion into the simulation algorithm. The pulp phase water recovery can be adequately represented by assuming a first order recovery rate of water that is affected by the cell residence time of the pulp and the bubble surface area flux, similar to that proposed by Engelbrecht and Woodburn (1975).

$$R_{w,k}^o = 1 - \frac{1}{[1 + P_w S_{b,k} \tau_{c,k} (\Omega_k \exp(\sigma_k V_{f,k} - \chi_k \tau_{air,k}))]}$$

The basic assumption in the development of this model was that a constant gas hold-up with froth depth can be approximated and this is supported by the work by Shih and Lemlich (1971) and Haas and Johnson (1965). They concluded that the water drains from the froth within a few bubble diameters of the pulp froth interface. Smith (1984) described the recovery of water to the flotation concentrate as a result of the contribution of two mechanisms, one for the transfer of water from the pulp to the froth due to air and the second for the transfer of water from the pulp to the froth due to the contribution of solids. The water model presented here essentially does the same thing, the air transporting water from the pulp to the froth, while the evaluation of the froth model parameters shows the respective contributions due to the action of solids. The analysis of the froth model parameters in this thesis has highlighted the significance associated with the regressed parameters. The Ω empirical parameter in the water froth recovery model is associated with the contribution of the floatable hydrophobic particles to the overall water recovery. The higher the concentration of floatable particles the larger the associated Ω parameter. The σ parameter in the water froth recovery model determines the slope of the water froth recovery (R_{fw}) exponential decay which is affected by the stability of the froth. The χ parameter in the water froth recovery model determines how long the R_{fw} curve takes to approach zero water froth recovery. The smaller χ the longer the curve takes to reach zero R_{fw} , this is associated with the amount of non-draining material in the froth causing a small fraction of water to be recovered at high froth residence times of air. Understanding the significance of the parameters in the empirical water recovery model has potential in being able to simulate "green field" designs where no data is available to regress the different empirical constants in this model.

The correlation between the measured and calculated water flow rates from the pilot plant tests was good using the water model presented above. However the model was found to be limited to fixed circuit configurations and specific flotation cells in a circuit.

8.2.2 Simulation algorithm

A step by step simulation algorithm of the pilot plant has been developed in this thesis describing in detail how the individual component models can be combined into a structure that is relatively simple, robust, stable and converges rapidly to a solution. The output of the simulation algorithm was compared to the experimental results of the rougher froth characterisation tests showing good correlation. The parameters in the froth models were found to be the most influential to the accuracy of the simulator. The froth drainage parameters affected the boundary conditions of the simulated grade-recovery curve, depending on the relative values of these parameters for the different mineral models. The overall form of the simulator's output with respect to the variation of air flow rate and froth height were found to be both feasible and sensible.

8.3 Conclusions regarding the validity of the simulation algorithm

8.3.1 Simulating alternative circuit configurations

The ability of the simulator to predict the overall flotation kinetics of the two alternative circuit configurations investigated in this thesis was relatively good provided the water froth recovery parameters as well as the mineral froth recovery parameters were changed appropriately. Accurate determination of the empirical water model parameters was found to play a very important role in determining the accuracy and feasibility of any simulation results. The froth recovery parameters for PGM, chromite and gangue were less influential on the simulator's predictive ability.

8.3.2 Validating the regressed PGM parameter set for the pilot plant

Comparing the regressed PGM parameter sets for different feed grind sizes showed a good correlation between the fitted model parameters and measurable stream characteristics. The PGM flotation kinetics as a function of particle size class has supported the differences observed between the regressed parameter values. The assumption that the floatability rates remain the same for differing feed grinds provides a basis for including regrind in a simulation algorithm. The satisfactory model fit of the experimental data for both feed grinds indicates that this assumption is feasible. It was observed that the PGM entrainment parameter varied linearly with the grind of the feed material, expressed on a % passing -75 μ m basis.

The regressed PGM floatability mass fractions of the UG2 Impala ore of phase I were evaluated in conjunction with the information obtained from the QEMSEM analysis of the feed material to the pilot plant. The basis of this work relied a lot on inference and logical assumptions regarding the influence of physical variables affecting the flotation rates of various types of particles presented to the flotation circuit. Even though this is the case, the correlation between the regressed parameters and calculated parameters was encouraging. This supports the feasibility and physical significance of regressing these parameters from the pilot plant data.

8.4 Conclusion and implications of the circuit optimisation study

This thesis was initiated in order to gain a greater understanding as to the reasons behind the recent circuit design trends of many new industrial platinum concentrators having multiple concentrate product streams from the circuits. The usefulness of the simulation algorithm in its current form has been presented in an optimisation study, even though there are a few limitations to its predictive ability. The simulation algorithm from the pilot plant experiments has highlighted the advantages of the recycle mechanism in continuously operated flotation circuits. For ores with a high fraction of slow floating particles the separation of the fast floating and slow floating components during cleaning is beneficial. In combination with this, allowing a high circulating load of only slow floating PGM particles to develop without the faster floating material being recycled allows higher grades to be achieved at the same recoveries. The separation of the floatability components gives an extra degree of freedom in controlling the grades and recoveries from a plant. The implications from this work suggest that it would be possible to achieve higher grades at the same recoveries from existing plant operations treating UG2 platinum ores.

Even though the simulator that has been developed in this thesis is restricted to small changes in circuit configurations as a result of a lack of fundamental water recovery models in the literature, it has been shown that valuable information can still be obtained from semi-empirical correlations of the flotation process.

Chapter 9

RECOMMENDATIONS

The most important step to developing a truly comprehensive simulator is the development of an independent water recovery model. This model needs to include the change in frother concentrations around a plant. This is one of the only variables that has not been assigned a "flotation rate", after-all it does get concentrated into the cleaner circuits. It is believed that an independent frother model that can be used in a water model would provide a good starting point to developing such a water model. An alternative to the above but less fundamental, would be to evaluate to what extent the empirical water model parameters developed in this thesis are capable of predicting alternative circuit configurations. Empirical correlations that relate the change in the variables affecting the parameters in the water model would provide a basis for simulating alternative circuit designs from a "green field" perspective.

The nature of the parameter regression analysis is highly non-linear and methods have been proposed in this thesis to evaluate whether the parameters in the models have physical significance. The major conclusions regarding the sensitivity of the parameters in the simulation algorithm was that the parameters describing the froth recovery models had the largest effect on the overall model fit. It is understandable that the pulp phase recovery model is trying to push material out of the flotation cell while the froth models are trying to push material back into the pulp of the flotation cell. This competition between the two is what causes the sensitivity in the resulting parameter sets, and this highlights the non-linear nature of the modelling. Savassi (1997) has developed a direct method of actually measuring the froth recovery of attached particles in flotation cells. This should be done for all experiments so that the froth models have a direct experimental measurement to be compared (and possibly fitted) in the regression analysis and no need would exist to derive these values. This would constrain the problem even further, and help decrease the error and sensitivity associated with the froth model parameters. The assumption made in this thesis is that froth recovery of attached particles is mineral specific but not floatability specific. This is a difficult problem to unravel as the floatability of a particle relates to all aspects influencing the flotation rate of the particle. Unfortunately detachment and froth drainage mechanisms involve variables such as particle size and mineral specific gravity. To link the two would be extremely difficult. If a truly comprehensive understanding is required to develop more accurate simulators, these kind of issues need to be understood on a more fundamental basis. Methods need to be developed to quantify the assumption in the mineral froth recovery models used in this thesis that the froth recovery of attached particles is not different for different floatability classes. Given that this was an assumption that was used in the development of the simulation algorithm in this thesis, the satisfactory model fit would tend to indicate that this assumption has a reasonable foundation.

The contribution of entrainment with water in the PGM model that was used in this thesis does not describe the true mechanism by which these particles are entrained. Savassi (1998) has developed an equation, through a detailed analysis of entrainment in industrial flotation cells that incorporates entrainment on a more realistic basis. This model should be compared to the one used in this thesis to determine the impact of assuming the unrecovered PGM particles are subjected to entrainment subsequent to flotation as applied in this thesis.

The one major limitation of the PGM model was the unsatisfactory prediction of the first flotation cells concentrate PGM flow rate in the cleaner banks. The same phenomenon was found previously modelling a UG2 ore on an industrial flotation circuit, (Theron, 1998). A stable froth containing extremely floatable PGM and gangue particles made the flotation kinetics in these first flotation cells a little irregular compared to the rest of the plant. Assigning an independent froth recovery model for these cells in the circuit could be a means of obtaining a better model fit. However, a need exists to investigate what is actually happening in these cells as they contain over 50% of all the valuable concentrate flowing through the circuit. A thorough understanding of this slightly erratic kinetic behaviour of these cells would be extremely useful and beneficial to the enhanced control of UG2 flotation circuits.

A method of quantifying the significance of the PGM model parameters was investigated in this thesis whereby two feed streams of differing grinds were subjected to the PGM regression analysis. The underlying assumption in this analysis was that the floatability rates remained constant with a change in grind, the only thing that is affected is the distribution of the floatability component mass fractions. This provides a basis for including regrind mills into the simulator which is extremely valuable for the simulation of platinum flotation circuits. Assessing the influence of a number of different grinds using the same technique used in the development of this thesis would be an extremely simple but valuable analysis of the effect of grind on a floatability distribution. Empirical models that describe the variation in the PGM model parameters with feed grind would provide a basis for overlapping the simulation of a milling circuit and flotation circuit. The analysis of this type of information would provide a basis to conclude whether the floatability rates are conserved with a change in feed grind, and probably be able to quantify to what extent this assumption remains valid as well as to identify the limitations using this assumption. The combination of these currently independent stages of the concentration process would definitely be beneficial to the overall understanding and optimisation of the flotation process.

A comprehensive analysis such as the one described in this thesis using QEMSEM data obtained from all streams in the circuit would provide a basis to assess what type of particles are associated with a particular floatability component. An analysis of this type of information would help assess to what extent the assumptions made in chapter 6 are relevant. This information could also be used to simulate the flotation kinetics of particles in a UG2 feed based on a mineral by size by liberation analysis. The large number of parameters inherent in such an analysis would not be useful for the direct

simulation of industrial flotation circuits, but would provide a means of quantifying the assumptions made by using the more simple floatability component model.

Finally, the purpose of developing the simulation algorithm in this thesis was to try understand whether multiple concentrate product flotation circuits are indeed better than the more standard bulk concentrate cleaning circuits used over the years. The conclusions obtained from the optimisation study presented in chapter 7 needs to be tested on an operating pilot plant or perhaps an industrial flotation circuit. The effective control of the residence times in the circuit (and therefore the circulating loads) needs to be comprehensively evaluated. If this principle does work on real plants, then it would be easy to apply an economic optimisation criteria in the simulation algorithm to determine the optimum circulating load based on the costs and implications associated with the operation of plants with high circulating loads.

University of Cape Town

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APPENDIX A – Analytical Techniques

- **The determination of PGM+Au**

The sample is mixed with a flux consisting of fused borax, sodium carbonate, silica, lead oxide, and maize meal. Silver nitrate solution is added to the mixture before being placed in a fusion furnace. The mixture of molten lead and slag is transferred to an iron mould and allowed to solidify. The lead is then separated from the slag and placed in a cupel in a muffle furnace, where the lead is absorbed by the material of the cupel, and a prill containing the PGM+Au and silver is left on the surface of the cupel. The prill is then subjected to a cupellation at high temperature for the removal of silver and the lead residue. After being cleaned the prill is weighed and the total PGM+Au is calculated. Individual PGM's for lead collection are read using ICP-MS or ICP-OES after preparation/dissolution of the prill. All the PGM samples analysed were done by the analytical department at Mintek. The measurement method is accredited. At least one reference sample similar to the samples being analysed is included with each batch of samples. The maximum variation between duplicate samples did not exceed 5%.

- **The Determination of Chromite**

An x-ray fluorescence analyser was used to analyse the percentage chromite (Cr_2O_3) in the samples collected from the tests. The analyser had to be initially calibrated using samples of known grades that were determined using chemical analysis. The chemical analysis technique is described below.

A direct titrimetric method for the determination of chromium ores was used. The sample is fused with sodium peroxide, leached in water and acidified with dilute sulphuric acid. The chromium is oxidized to the hexavalent state by boiling with an excess of potassium permanganate. The excess potassium permanganate is destroyed by the addition of hydrochloric acid and the chromate titrated with ferrous ammonium sulphate, using ferroin as indicator. The relative standard deviation of the method at the 50% Cr_2O_3 level is 0.016. The accuracy of the method based on the analysis of RMS is $\pm 0.5\%$ absolute at the 50% concentration level.

Once the analyser was calibrated with the samples obtained from the chemical analysis, it could measure samples over the entire spectrum of calibrated samples. The method of measurement is based on energy dispersive x-ray fluorescence technology and on the use of radioactive isotopes. The machine provides a method for quantitative and qualitative analysis of samples in powder, solid, liquid and briquette form. Radiation from the radioactive source strikes the sample and excites fluorescent radiation. The spectrum of the fluorescent radiation contains the information required for the analysis. The radiation is measured with a proportional counter, and the data is processed by a microcomputer in the electronic unit. The XRF analyser produces primary x-rays which illuminate the sample and cause it to emit characteristic x-rays which are associated with a particular wavelength. The x-ray detection process involves

the following steps: adsorption of the x-ray photon in the proportional counter gas filling, formation of an electron avalanche, acceleration of the electrons to the high voltage wire and conversion to a charge pulse proportional to the x-ray energy. Each pulse lasts a few seconds so that each x-ray photon is detected sequentially. The pulse count in each element channel divided by the measurement time is called the net intensity which is then converted into element concentration.

The Mintek Technical Memorandum TMNo 170921 concludes a slight systematic error of 0.01 % in the equipment. For individual grades, at 99% and 90 % confidence levels, no significant difference was found between paired values for 24 measurements each at high, medium and low concentrations. It was therefore concluded that using the XMET was sufficiently accurate enough to obtain reliable estimates of the percentage chromite within the samples tested.

University of Cape Town

APPENDIX B – Monte Carlo Boot Strap Technique

The term 'Monte Carlo method' is used to describe a wide range of techniques, all of which are based on the use of random numbers, (Kojovic, 1996). These methods are used to explore the behaviour of systems and models which are intrinsically random in nature, and/or which are too large or complex to be calculated "deterministically". Flotation plants are usually associated with high standard deviations in the measured experimental data as a result of the dynamic nature of the process. A method is required in order to quantify the error associated with the parameters from the floatability component modelling exercise. Many possible minimum solutions can be found as a result of the error associated with experimental data obtained from flotation plants, as well as the fact that the solution obtained from a non-linear search allows a high interaction between the different parameters. The parameter set that is obtained is a mean of a distribution of possible parameter sets, and it is important to be able to quantify the error associated with the distribution of the parameter set obtained. The next section describes a method of generating distributions of the parameters using the Monte Carlo boot strap technique.

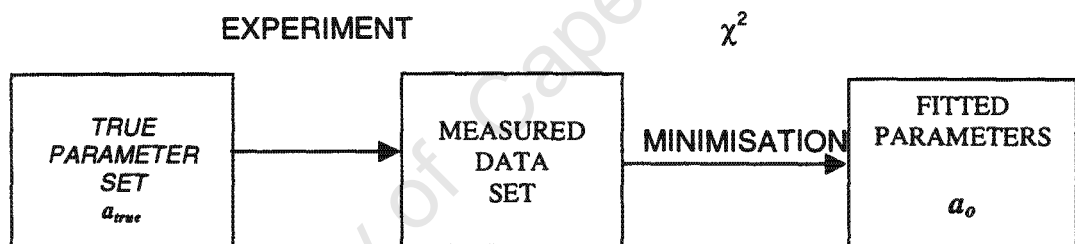


Figure B1: Relationship between "true" and "fitted" model parameters, (Kojovic JKMRC,1996)

Assume in an ideal world that there is a "true" set of model parameters a_{true} that would fit the kinetic experimental data perfectly. The relationship between the best fit model parameters defined as $a(0)$ and a_{true} is depicted in figure B1. Inherent in the data set obtained from a flotation plant are experimental errors due to a multitude of factors that propagate through from the measured values. Assuming four different mass balances are conducted. Each separate mass balance generates a data set D(1), D(2), D(3) and D(4). From each data set, a corresponding best fit parameter set can be obtained $a(1)$, $a(2)$, $a(3)$ and $a(4)$. Therefore, the best fit parameter set $a(0)$ is one member of a population of possible parameter sets. If infinite mass balances could be performed, a distribution of the parameters "a" could be obtained.

The basic idea of generating this probability distribution is to replace the unknown parameters a_{true} by the known "best-fit" parameters $a(0)$. Although we know that the parameters from the regression analysis are not the true parameters, we consider a fictitious world in which $a(0)$ are the true parameters and use them to generate synthetic data sets D(i) which simulate the errors in the experiment measuring the data set D(0).

These data sets are then used to obtain synthetic best-fit model parameters $a(i)$. Figure B2 demonstrates the principle behind the Monte Carlo boot strap technique.

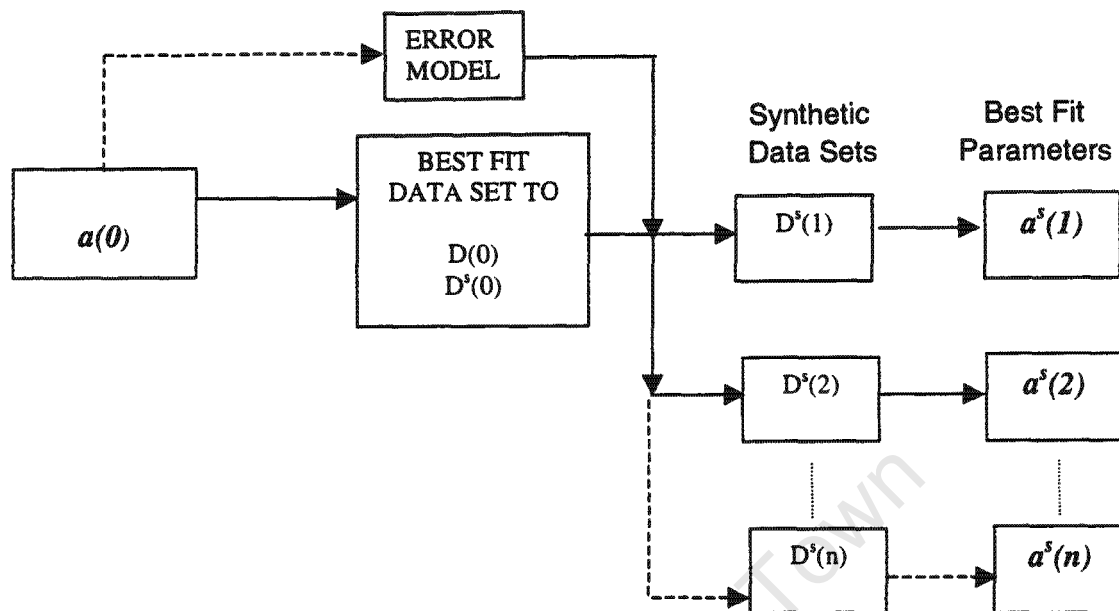


Figure B2: Monte Carlo simulations of the data gathering and parameter estimation process, Kojovic (1996)

The technique used for calculating a synthetic data set involves changing each measured experimental point by a randomly created, normally distributed measured deviation. The size of the deviation is a function of the standard deviation of the experimental data points obtained from the four mass balances.

$$D_{\text{new},i} = D_{\text{measured},i} + (N_{\text{random}})(\text{StDev}_{\text{measured},i}) \dots \dots \dots \text{B1}$$

where

- $D_{\text{new},i}$ is the new data point for stream (i) or batch recovery from stream (i)
 $D_{\text{measured},i}$ is the actual measured data point for stream (i) or batch recovery from stream (i)
 N_{random} is the randomly generated normally distributed number
 $\text{StDev}_{\text{measured},i}$ is the measured standard deviation of the measured data point, $D_{\text{measured},i}$

APPENDIX C – Raw Mini Plant Data

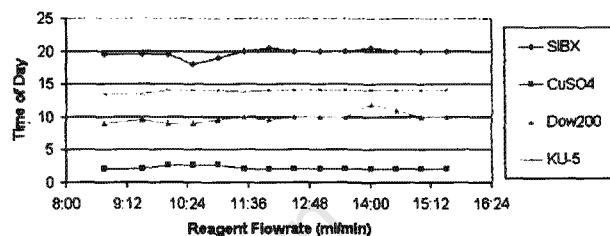
C.1 Feed and flotation cell operating set points

MEASURED REAGENT FLOW RATES

Date: 4/12/97

Time	SIBX	CuSO ₄	Dew200	KU-5
	ml/min	ml/min	ml/min	ml/min
8:45	19.5	2	9	13.5
9:30	19.5	2	9.5	13.5
10:00	19.5	2.5	9	14
10:30	19	2.5	9	14
11:00	19	2.5	9.5	14
11:30	20	2	10	13.8
12:00	20.5	2	9.5	14
12:30	20	2	10	14
13:00	20	2	10	14
13:30	20	2	10	14
14:00	20.5	2	12	14
14:30	20	2	11	14
15:00	20	2	10	14
15:30	20	2	10	14

Reagent Flowrates During Campaign

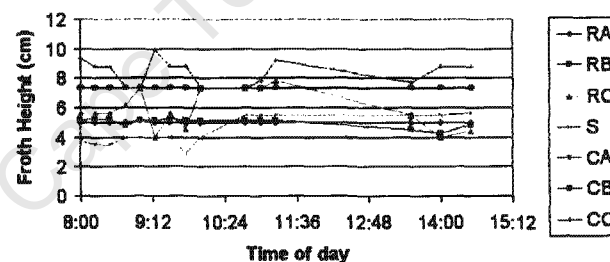


MEASURED FROTH HEIGHTS

Date: 4/12/97

Time	RA	RB	RC	S	CA	CB	CC
	cm	cm	cm	cm	cm	cm	cm
8:00	5	5.1	5.05	3.7	7.3	7.3	9.35
8:15	5	5.2	5.05	3.45	7.3	7.3	8.8
8:30	5	5.1	4	4	7.3	7.3	8.8
8:45	5	4.75	6.2	4	7.3	7.3	7.3
9:00	5.1	5.1	7.3	4	7.3	7.3	7.3
9:15	5	5.1	4	4	7.3	7.3	8.8
9:30	5	5.2	5.05	4	7.3	7.3	8.8
9:45	5	5.1	4.05	2.8	7.3	7.3	8.8
10:00	5	5.1	7.3	4	7.3	7.3	7.3
10:15	5	5.1	7.3	5.5	7.3	7.3	7.3
11:00	5	5.1	7.3	5.5	7.3	7.3	7.85
11:15	5	5.1	7.05	5.5	7.3	7.3	8.2
13:30	5	4.5	6.5	5.5	7.3	7.3	7.7
14:00	5	4.25	4	5.5	7.3	7.3	8.8
14:30	5	4.8	4.45	5.05	7.3	7.3	8.8
Average	5.00886667	4.98	6.98	4.43333333	7.3	7.3	8.4

Froth Heights for Campaign

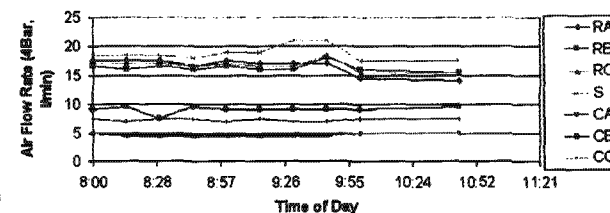


MEASURED AIR FLOW RATES

Date: 4/12/97

Time	RA	RB	RC	S	CA	CB	CC
	l/min	l/min	l/min	l/min	l/min	l/min	l/min
8:00	17.5	16.5	17	18.5	5	9	7.5
8:15	17.5	16	17	18.5	4.5	9.5	7
8:30	17.5	16.5	17	18.5	4.5	7.5	7.5
8:45	16.5	16	16.5	18	4.5	9.5	7.5
9:00	17.5	16.5	17	19	4.5	9	7
9:15	17	16	16.5	19	4.5	9	7.5
9:30	17	16	16.5	21	4.5	9	7
9:45	17	16	16.5	21	4.5	9	7
10:00	14.5	16	15.5	17.5	5	9	7.5
10:15	14	15.5	15.5	17.5	5	9.5	7.5

Air Flow Rates for Campaign

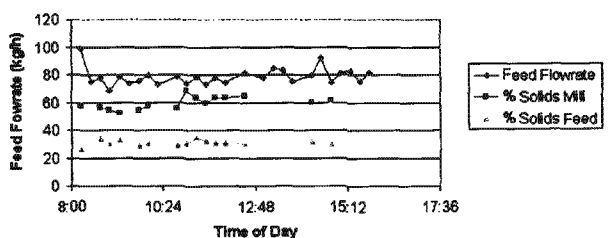


MEASURED FEED FLOW RATES

Date: 4/12/97

Time	Feed Flowrate kg/h	% Solids Mill Discharge	% Solids Mill Feed	Mill Speed (rpm)	% less 75 micron
8:15	98.6376	57	26	51.4	
8:30	74.484			51.4	
8:45	77.5632	56	34	51	
9:00	68.0256	54	30	51	
9:15	70.9848	52	33	51	
9:30	73.4112				51.77
9:45	75.7056	54	29		
10:00	79.3656	57	30	51.4	
10:15	72.482				
10:45	70.336	55	29		55.06
11:00	73.3008	68	30	51.4	
11:15	70.192	63	35		
11:30	72.984	59	32		
11:45	77.628	63	31		
12:00	74.2008	63	31		58.62
12:30	81.4272	64	30		
13:00	77.5392				
13:15	84.6744				
13:30	82.9704				
13:45	74.9944				
14:15	79.4232	60	32	51.4	
14:30	91.812				
14:45	74.184	61	30	51.4	62.39
15:00	80.3328				
15:15	82.5216				53.31
15:30	75.0408				
15:45	81.3848				
Average	78.4500000				

Feed Flowrate for Campaign



C.2 Raw mass balance data**MEASURED DATA FROM ALL TESTS****Mass Balance 1**

	Dry Mass Flow (g/min)	Water Flow (ml/sec)	PGM Grade (g/t)	Cr2O3 Grade (%)	Froth Height (cm)	Air Flow Rate (l/min)	Bubble Diameter (mm)
Feed	1044.80	53.75	4.52	25.08			
RA	29.97	10.49	120.00	7.87	5.64	16.60	0.74
RB	25.36	12.34	26.15	10.41	6.63	16.30	0.85
RC	29.55	16.13	13.60	12.47	7.48	16.70	0.91
RT	1144.00	44.41	0.33	26.11			
SC	12.35	8.67	13.50	15.43	5.10	18.85	1.11
Tails	992.56	33.23	0.30	24.85			
CA	21.78	5.47	126.00	5.52	7.30	4.65	0.74
CB	20.69	8.12	55.20	7.28	7.30	9.00	0.80
CC	14.19	5.03	39.50	6.83	8.40	7.30	0.62
CT	36.05	26.31	7.12	16.92			

Mass Balance 2

Feed	1206.34	49.39	4.52	25.61			
RA	39.87	14.59	95.85	10.22	5.64	16.60	0.74
RB	27.36	12.57	23.60	12.51	6.63	16.30	0.85
RC	26.22	13.42	12.80	13.45	7.48	16.70	0.91
RT	1162.51	32.13	0.46	26.08			
SC	9.66	4.95	10.90	17.02	5.10	18.85	1.11
Tails	1152.86	27.61	0.28	26.00			
CA	19.77	5.61	154.00	6.06	7.30	4.65	0.74
CB	20.88	9.17	73.30	8.00	7.30	9.00	0.80
CC	14.46	7.00	38.70	8.47	8.40	7.30	0.62
CT	39.96	20.73	4.91	18.75			

Mass Balance 4

Feed	1114.80	49.59	4.13	24.43			
RA	46.98	18.11	86.30	9.10	5.64	19.60	0.74
RB	24.48	10.65	21.65	12.38	6.63	19.30	0.85
RC	19.56	8.20	14.40	12.24	7.48	19.70	0.91
RT	1144.08	36.67	0.31	25.44			
SC	13.88	5.39	10.30	16.34	5.10	18.85	1.11
Tails	1122.24	32.20	0.28	26.43			
CA	17.72	6.77	150.00	4.92	7.30	4.65	0.74
CB	19.76	5.71	46.20	6.64	7.30	9.00	0.80
CC	11.30	4.91	36.70	5.46	8.40	7.30	0.62
CT	37.17	18.17	5.07	18.28			

Mass Balance 4+5

Feed	1145.24	49.34	4.13	25.87			
RA	46.98	18.11	86.30	11.01	5.64	19.60	0.74
RB	24.48	10.65	21.65	11.33	6.63	19.30	0.85
RC	19.56	8.20	14.40	11.55	7.48	19.70	0.91
RT	1105.28	35.94	0.31	26.01			
SC	13.88	5.39	10.30	17.00	5.10	18.85	1.11
Tails	998.74	29.22	0.29	24.85			
CA	23.81	8.36	103.00	6.20	7.30	4.65	0.74
CB	18.76	6.53	58.60	6.31	7.30	9.00	0.80
CC	14.31	5.23	36.30	6.77	8.40	7.30	0.62
CT	37.17	18.17	5.07	17.44			

Circuit Configuration A

Feed	1155.20	51.27	4.90	23.23			
RA	25.80	6.52	157.50	6.79	5.50	22.50	1.05
RB	23.55	8.74	31.80	9.79	6.60	27.25	0.97
RC	21.45	8.84	20.00	10.92	6.78	26.63	1.13
RT	1097.30	35.25	0.28	26.44			
SC	12.90	8.14	16.60	13.35	3.30	30.00	1.41
Tails	1111.40	38.39	0.47	25.02			
CB	12.67	1.93	58.20	4.72	7.00	14.50	0.91
CC	7.43	2.28	46.50	5.45	7.00	10.67	0.76
CT	27.00	15.05	8.26	14.46			

Circuit Configuration B

Feed	1104.30	51.32	3.68	24.59			
RA	28.65	10.39	123.00	8.33	5.50	27.00	1.02
RB	28.00	16.02	25.00	10.72	6.60	30.50	1.40
RC	29.63	15.73	13.10	13.25	6.78	30.25	1.03
Tails	949.13	36.70	0.38	25.01			
CA	31.00	11.65	101.00	5.66	4.40	21.00	0.67
CB	14.50	5.22	41.50	5.79	4.40	14.50	0.67
CT	51.20	33.02	10.20	16.07			
RCA	12.75	2.23	148.00	2.99	6.50	11.00	1.02
RCB	12.55	2.59	73.35	3.79	7.00	14.00	0.75
RCC	10.35	4.55	43.70	5.16	7.00	13.25	0.90
RCT	8.55	7.26	7.00	11.61			

Cleaner Water A

Feed	1342.44	51.28	4.15	25.01			
RA	30.36	9.12	104.00	7.32	5.13	23.333	1.0378
RB	27.75	12.39	34.50	10.63	5.13	25	0.8398
RC	23.15	10.80	20.90	11.19	6.60	23.5	0.8757
RT	1309.32	39.43	0.46	26.79			
SC	9.73	4.59	16.00	15.99	5.53	34.333	1.4122
Tails	1299.59	36.10	0.32	25.62			
Water				0.00			
CA	15.46	4.26	105.00	5.35	6.50	10.5	0.8455
CB	14.80	3.67	64.50	5.67	7.00	13.5	0.8955
CC	9.10	3.22	52.50	6.32	7.00	10.33	0.7325
CT	38.41	21.15	6.33	16.70			

Cleaner Water B

Feed	1073.04	52.50	4.24	25.41			
RA	27.13	6.29	147.00	5.00	5.13	23.33	1.0378
RB	19.45	8.04	25.90	7.92	5.13	25.50	0.8398
RC	17.21	7.42	17.50	8.84	6.60	23.50	0.8757
RT	1053.12	84.19	0.47	26.62			
SC	15.89	11.57	10.70	15.63	3.30	34.67	1.45
Tails	1037.23	60.56	0.46	26.69			
Water		25.08		0.00			
CA	14.50	4.94	107.00	4.39	6.50	14.50	0.9215
CB	13.06	3.49	54.10	4.59	7.00	16.50	0.98612
CC	8.96	3.12	36.50	5.49	7.00	16.00	0.8992
CT	27.98	35.27	10.10	14.81			

Cleaner Water C

Feed	1073.04	52.29	4.24	24.11			
RA	30.90	9.43	107.00	6.33	5.39	25.00	1.023
RB	24.25	12.13	28.30	8.30	6.05	25.50	0.8398
RC	20.97	11.05	16.90	9.05	6.60	23.83	0.8757
RT	1034.09	73.24	0.28	24.81			
SC	15.03	11.50	10.20	14.25	3.30	34.00	1.4122
Tails	1019.05	51.15	0.48	23.78			
Water		29.39		0.00			
CA	21.03	10.58	84.20	5.20	6.50	15.67	0.9919
CB	19.43	9.97	36.70	5.95	7.00	17.50	0.9202
CC	10.95	6.31	39.30	5.55	7.00	14.00	0.75658
CT	22.14	35.13	11.40	12.74			

Reduced Throughput

Feed	873.94	31.517	3.00	26.40			
RA	29.32	12.37083333	62.00	10.16	3.33	24.66	1.02602
RB	12.395	6.901833333	22.60	12.19	4.40	27.00	0.94588
RC	14.1	8.11125	15.20	13.05	5.78	26.33	1.1252
RT	840.997	20.15958333	0.29	26.46			
SC	3	1.22975	26.90	14.59	3.52	34.67	1.45
Tails	837.997	18.92983333	0.76	26.00			
CA	12.595	4.627583333	87.00	6.11	6.50	10.33	0.83086
CB	13.735	4.247083333	45.60	6.13	7.00	11.80	0.8474
CC	8.53	4.5125	29.90	6.82	7.00	10.83	0.7709
CT	19.872	14.803	5.05	17.53			

Rougher Tests Results

Test 1	Dry Mass Flow (g/min)	Water Flow (ml/sec)	PGM Grade (g/l)	C203 Grade (%)	Fresh Height (cm)	Air Flow Rate (l/min)	Bubble Diameter (mm)
Feed	984.90	59.19	5.25	25.13			
RA	26.17	12.34	143.00	13.80	2.5	10	0.7396
RB	21.14	7.74	35.70	12.15	3.2	10	0.7368
RC	6.37	2.75	62.30	9.19	3.6	10	0.9139
RT	931.22	36.35	0.30	26.15			
Test 2							
Feed	1079.43	58.76	3.77	25.14			
RA	10.21	2.32	181.00	4.67	3.5	10	0.7396
RB	19.11	8.74	50.00	12.00	3.4	10	0.7368
RC	24.45	11.91	32.20	10.52	3.5	10	0.9139
RT	1025.66	24.72	0.47	27.03			
Test 3							
Feed	1151.91	57.43	4.95	25.40			
RA	26.17	12.34	143.00	8.89	2.5	10	0.7396
RB	23.17	12.34	35.70	12.37	3	10	0.7368
RC	34.20	18.75	17.30	14.18	3.2	10	0.9139
RT	1068.36	14.00	0.51	25.88			
Test 4							
Feed	1175.28	57.57	5.40	25.72			
RA	81.53	30.00	60.10	14.15	3	15	1.0176
RB	21.91	10.33	25.30	12.20	3.8	15	0.7304
RC	26.20	12.75	17.70	14.09	4	15	0.8737
RT	1045.64	4.49	0.41	25.50			
Test 5							
Feed	1198.65	57.71	6.04	26.04			
RA	15.70	4.56		9.00	4	15	1.0176
RB	12.11	3.72	50.80	7.37	5	15	0.7304
RC	11.31	3.73		13.00	5.2	15	0.8737
RT	738.00	64.53	0.41	25.19			
Test 6							
Feed	1184.07	57.21	5.74	25.72			
RA	26.56	8.07	188.00	6.54	5.2	15	1.0176
RB	21.60	9.14	42.50	9.60	4.7	15	0.7304
RC	19.66	9.52	26.60	11.29	4.5	15	0.8737
RT	1116.05	24.48	0.32	27.12			
Test 7							
Feed	1189.49	56.71	5.09	25.41			
RA	67.57	26.30	66.20	13.66	5.5	20	0.8951
RB	27.44	13.19	26.80	13.30	6	20	0.8602
RC	31.17	3.35	14.50	14.91	6.4	20	0.7575
RT	1043.31	13.87	0.28	25.50			
Test 8							
Feed	1074.18	53.73	3.81	26.31			
RA	43.11	15.58	89.60	11.52	6.2	20	0.8951
RB	13.81	5.47	32.90	10.26	6.9	20	0.8602
RC	18.13	9.24	19.90	13.77	7.5	20	0.7575
RT	999.13	21.51	0.28	25.91			
Test 9							
Feed	1167.26	53.59	2.96	25.92			
RA	23.66	7.37	94.70	8.40	6.6	20	0.8951
RB	12.64	4.12	41.20	8.41	7.5	20	0.8602
RC	13.44	5.34	28.60	9.92	8	20	0.7575
RT	1117.52	38.47	0.28	25.72			
Test 10							
Feed	1260.33	53.45	4.35	25.53			
RA	56.92	20.10	75.4	12.89	7.5	25	1.023
RB	22.17	9.76	24.30	13.39	8.2	25	0.8398
RC	28.25	14.16	11.50	15.21	8.3	25	1.0863
RT	1152.99	11.43	0.28	27.24			
Test 11							
Feed	1189.64	54.27	3.73	25.12			
RA	49.48	17.05	88.10	11.95	7.9	25	1.023
RB	16.71	6.88	27.70	10.71	8.7	25	0.8398
RC	18.12	8.97	16.60	13.67	8.9	25	1.0863
RT	1105.34	17.15	0.28	25.86			
Test 12							
Feed	1118.94	55.09	3.72	24.72			
RA	35.21	10.84	89.35	8.99	8.2	25	1.023
RB	16.61	6.37	25.60	10.39	9.2	25	0.8398
RC	16.00	7.02	20.90	11.43	9	25	1.0863
RT	1051.12	27.10	0.28	25.35			
Test 13							
Feed	1118.94	55.09	3.26	24.72			
RA	30.90	9.29	82.40	8.85	8.7	25	1.023
RB	14.89	4.11	33.30	8.49	10	25	0.8398
RC	13.30	4.49	24.00	10.00	10.2	25	1.0863
RT	1060.06	35.25	0.28	25.47			
Test 14							
Feed	1215.12	55.05	5.62	26.02			
RA	32.74	10.88		10.00	10.5	30	0.8597
RB	19.54	9.03	23.40	11.96	11.3	30	1.1
RC	19.48	9.88		13.00	11.2	30	1.0339
RT	937.60	23.91	0.28	25.84			
Test 15							
Feed	1112.67	53.54	3.71	24.30			
RA	34.33	11.75	88.10	9.73	11	30	0.8597
RB	17.96	7.62	27.00	11.31	11.8	30	1.1
RC	17.55	8.08	18.40	12.41	11.7	30	1.0339
RT	1042.84	23.04	0.28	26.48			
Test 16							
Feed	1010.22	52.03	5.89	22.58			
RA	40.89	12.64	117.00	7.78	11.5	30	0.8597
RB	19.98	7.98	29.00	10.79	12.3	30	1.1
RC	19.55	9.02	18.00	10.90	12.2	30	1.0339
RT	930.01	22.39	0.28	25.50			

Cleaner Test Results

Test 1							
	Dry Mass Flow (g/min)	Water Flow (ml/sec)	PGM Grade (g/t)	Cr2O3 Grade (%)	Froth Height (cm)	Air Flow Rate (l/min)	Bubble Diameter (mm)
<i>Cleaner Feed</i>	68.60	25.15	69.80	11.11			
CA	12.54	0.69	187.00	3.64	8	12.5	0.9919
CB	5.62	0.20	134.00	3.21	8.5	12.5	0.879
CC	4.90	0.06	68.60	4.12	8.5	12.5	0.8992
CT	45.53	24.19	9.06	15.00			
Test 2							
<i>Cleaner Feed</i>	87.29	36.02	79.60	10.89			
CA	23.13	3.76	145.00	4.18	8	15	0.9039
CB	16.00	2.32	86.20	4.26	8.5	15	0.9202
CC	9.29	1.52	51.70	4.61	8.5	15	0.8992
CT	38.87	28.42	7.65	15.96			
Test 3							
<i>Cleaner Feed</i>	104.60	39.93	108.50	13.44			
CA	30.78	5.52	140.50	4.44	8	17.5	0.9514
CB	23.29	4.56	75.80	4.43	8.5	17.5	0.997
CC	15.37	2.51	48.20	4.49	8.5	17.5	0.997
CT	26.95	19.72	6.96	16.63			
Test 4							
<i>Cleaner Feed</i>	95.36	27.03	58.40	11.10			
CA	41.01	8.05	144.00	4.88	8	20	0.92
CB	23.57	7.66	49.80	5.87	8.5	20	0.94
CC	14.77	3.72	34.30	5.30	8.5	20	0.93
CT	16.01	11.30	5.27	18.10			
Test 5							
<i>Cleaner Feed</i>	82.89	36.01	49.90	10.53			
CA	19.56	5.20	104.00	5.47	6.5	10	0.8089
CB	16.98	3.92	101.00	6.23	7	12.5	0.7661
CC	9.65	3.06	55.30	5.99	7	10	0.7072
CT	48.49	29.79	8.07	17.56			
Test 6							
<i>Cleaner Feed</i>	123.91	49.81	47.50	12.19			
CA	47.67	14.08	159.00	7.16	6.5	12.5	0.9919
CB	26.73	5.08	74.90	5.38	7	12.5	0.879
CC	19.96	9.02	51.90	6.58	7	12.5	0.8992
CT	29.55	17.76	7.40	17.53			
Test 7							
<i>Cleaner Feed</i>	85.18	35.56	41.00	11.77			
CA	25.04	12.51	92.50	7.80	6.5	15	0.9039
CB	14.97	6.69	58.80	6.13	7	15	0.9202
CC	12.70	7.35	32.30	7.73	7	15	0.8992
CT	11.12	6.21	12.50	18.39			
Test 8							
<i>Cleaner Feed</i>	90.84	43.46	33.90	11.01			
CA	14.53	3.87	115.00	4.81	5	7.5	0.7363
CB	8.01	1.19	95.30	4.33	5.5	8.5	0.81272
CC	17.56	9.65	41.20	7.45	5.5	7.5	0.6216
CT	51.04	39.00	7.21	17.05			
Test 9							
<i>Cleaner Feed</i>	85.56	39.30	38.50	10.02			
CA	23.15	10.41	88.70	6.72	5	10	0.8089
CB	10.47	2.53	84.90	5.08	5.5	10	0.7661
CC	22.19	13.71	30.40	9.72	5.5	10	0.7072
CT	22.69	13.34	5.73	17.42			
Test 10							
<i>Cleaner Feed</i>	99.46	43.30	51.20	12.33			
CA	42.68	19.96	79.55	9.27	5	12.5	0.9919
CB	21.02	8.63	64.90	7.31	5.5	12.5	0.879
CC	23.06	13.70	26.90	10.26	5.5	12.5	0.8992
CT	12.99	1.69	5.98	18.46			

C.3 Batch floatability data set from mini plant feed characterisation test

Batch Floatability Tests

Experimental PGM Assay (g/t)																
Time	Fresh Feed	Fresh Feed	Fresh Feed	Tails	Tails	RA	RB	RC	Rougher	Scav	Time	Cleaner	CA	CB	CC	Cleaner
(min)	1	2	3	1	2	conct	conct	conct	Tails	conct		Feed	conct	conct	conct	Tails
1	218.00	293.00	171.00	2.72	1.13	360.00	75.40	47.30	21.00	31.60		147.00	277.50	162.00	79.10	25.34
3	42.40	47.90	30.60	2.72	5.24	122.00	41.10	27.40	9.75	44.90		44.30	127.00	89.50	49.00	10.90
5	14.90	13.80	14.80	3.08	3.29	37.40	26.80	18.90	8.58	15.50		24.00	45.30	48.80	27.50	10.50
7	12.60	9.91	10.60	2.78	8.80	25.60	19.60	17.90	7.78	9.47		13.80	26.00	34.10	24.80	8.88
10	8.37	4.92	5.61	2.03	2.88	22.40	18.70	16.80	12.00	11.40		12.40	21.20	25.80	18.40	7.36
15	7.64	4.55	5.97	6.97	4.04	15.20	15.80	14.10	5.81	6.74		7.71	18.90	18.10	14.40	11.30
Tails	0.45	0.61	0.32	0.35	0.30	6.70	9.38	9.51	0.41	3.44		5.52	10.30	13.90	10.50	3.42
Total Dry Masses (g)																
Time	Fresh Feed	Fresh Feed	Fresh Feed	Tails	Tails	RA	RB	RC	Rougher	Scav	Time	Cleaner	CA	CB	CC	Cleaner
(min)	1	2	3	1	2	conct	conct	conct	Tails	conct		Feed	conct	conct	conct	Tails
1	11.56	12.35	11.32	7.38	8.63	12.31	13.07	9.97	4.25	3.95		8.93	29.98	12.78	11.85	5.86
3	9.33	12.69	9.32	13.59	4.77	14.66	13.48	12.28	4.19	4.19		8.69	18.09	9.00	11.82	7.69
5	6.19	8.64	6.72	11.73	4.00	6.50	9.12	8.83	3.09	4.33		5.00	11.16	5.77	7.97	6.82
7	4.53	8.02	6.67	4.99	3.26	4.80	4.27	3.99	2.90	5.41		3.55	6.54	3.32	5.35	4.41
10	6.45	8.89	7.41	4.69	4.18	3.52	6.20	4.57	6.17	3.80		4.95	5.37	3.16	4.88	6.34
15	10.22	11.20	7.28	5.68	6.71	4.91	5.25	3.93	13.15	5.10		5.61	4.22	3.55	4.57	6.34
Tails	661.08	778.68	678.17	972.33	880.52	22.64	22.50	22.47	1316.47	51.99		28.12	11.33	11.20	15.65	67.86
Total	709.36	840.37	726.89	1020.39	912.07	69.34	73.89	66.04	1350.22	78.77		64.85	86.69	48.78	62.09	105.32

C.4 Bubble size data as a function of volumetric air flow rate

Rougher A				Rougher B				Rougher C				Scavenger			
Air Flow Rate (l/min)	Mean Bubble Diameter (mm)	Std Deviation (mm)	Bubble Flux (1/sec)	Air Flow Rate (l/min)	Mean Bubble Diameter (mm)	Std Deviation (mm)	Bubble Flux (1/sec)	Air Flow Rate (l/min)	Mean Bubble Diameter (mm)	Std Deviation (mm)	Bubble Flux (1/sec)	Air Flow Rate (l/min)	Mean Bubble Diameter (mm)	Std Deviation (mm)	Bubble Flux (1/sec)
10	0.615	0.4014	18.06685	10	0.7368	0.3907	15.08023	10	0.9139	0.4486	12.15791	15	0.5119	0.425	32.55844
12.5	0.7912	0.4201	17.55421	12.5	0.6708	0.3691	20.70496	12.5	0.6352	0.4175	21.86538	17.5	0.9076	0.5076	21.42402
15	1.0172	0.478	16.36485	15	0.7304	0.4119	22.81855	15	0.8737	0.5488	19.07596	20	1.2663	0.5046	17.27608
17.5	1.0699	0.4643	18.17408	17.5	0.9309	0.4376	20.88779	17.5	0.9319	0.5175	20.86538	22.5	1.2131	0.7524	20.60836
20	0.8951	0.58	24.82652	20	0.8602	0.4448	25.83379	20	0.7575	0.4651	29.33627	25	0.8175	0.6706	33.97893
22.5	1.0452	0.5811	23.91887	22.5	0.8325	0.5031	30.03003	22.5				27.5	0.6395	0.5659	48.11899
25	1.023	0.55	27.15325	25	0.8398	0.5694	33.07666	25	1.0863	0.5985	25.571	30	1.4122	0.6679	23.60383
27.5	1.0406	0.5331	29.3634	27.5	0.9724	0.5886	31.42283	27.5	1.1594	0.4893	26.35463				
30	0.8597	0.509	38.77322	30	1.4025	0.6782	23.76708	30	1.0339	0.6246	32.24038				
Cleaner A				Cleaner B				Cleaner C							
Air Flow Rate (l/min)	Mean Bubble Diameter (mm)	Std Deviation (mm)	Bubble Flux (1/sec)	Air Flow Rate (l/min)	Mean Bubble Diameter (mm)	Std Deviation (mm)	Bubble Flux (1/sec)	Air Flow Rate (l/min)	Mean Bubble Diameter (mm)	Std Deviation (mm)	Bubble Flux (1/sec)				
5	1.3861	0.6238	4.008048	5	0.711	0.5116	7.813721	5	0.6232	0.4246	8.914563				
7.5	0.7363	0.5464	11.31785	7.5	0.8438	0.462	9.875958	7.5	0.6216	0.3498	13.40626				
10	0.8089	0.4682	13.73608	10	0.7661	0.4809	14.50347	10	0.7072	0.4443	15.71141				
12.5	0.9919	0.4656	14.00231	12.5	0.879	0.4692	15.80078	12.5	0.8992	0.4881	15.44583				
15	0.9039	0.48	18.43862	15	0.9202	0.4784	18.112	15	0.6615	0.533	25.19526				

C.5 Chromite feed particle size distribution raw data

Chromite Particle Size Results

5 Feed						
Particle Size	Dry Mass +Paper	Dry Mass	% Cr2O3	Cr2O3 (g)	% Less than	
0	0				0.00	
38um	38	78.20	74.80	24.41	18.26	25.44
38-53um	53	39.40	36.00	24.08	8.67	37.52
53-75um	75	64.60	61.20	22.83	13.97	56.99
75-106um	106	89.30	85.90	19.78	16.92	80.56
106-150um	150	69.60	66.20	16.35	10.82	95.64
+150um	200	18.80	15.40	20.32	3.13	100.00
Total		359.90		71.77		
Concentrates						
0	0				0.00	
38um	38	105.80	102.40	12.46	12.76	88.16
38-53um	53	22.18	18.70	2.94	0.55	91.96
53-75um	75	25.18	21.70	2.74	0.59	96.06
75-106um	106	19.40	16.00	2.81	0.32	98.28
+106um	150	13.50	10.10	2.46	0.25	100.00
Total		185.90		14.48		

APPENDIX D – Raw Pilot Plant Data – IMPALA UG2

D.1 Raw mass balance data

CIRCUIT Measured	Feed Characterisation - Mass Balance 1											
	Duration (sec)	Mass Bag (g)	Wet Mass+Bag+Clip (g)	Dry Mass (g)	Water Mass (g)	Chromite (%)	PGM (g/t)	Dry Mass Flow (g/min)	Water Flow (ml/sec)	C2O3 Flow (g/min)	PGM Flow (mg/hr)	Gangue Flow (g/min)
Feed	5.3	51.1	2231.2	647.1	1531.5	19.8	3.7	7270.8	286.8	1439.3	1609.8	5831.5
Rougher A Conc	90.7	51.2	1162.3	276.0	833.6	4.5	60.5	182.5	9.2	8.2	662.6	174.3
Rougher B Conc	100.9	50.6	1224.6	207.0	965.5	5.5	34.7	123.1	9.6	6.7	256.3	116.4
Rougher C Conc	90.4	51.3	1453.1	140.5	1259.8	7.0	18.1	93.3	13.9	6.5	101.3	86.8
Rougher D Conc	90.9	51.1	1296.9	116.4	1127.9	9.0	14.9	76.8	12.4	6.9	68.7	70.0
Rougher E Conc	90.7	51.0	1623.1	119.4	1451.2	10.3	12.9	79.0	16.0	8.2	61.1	70.8
Rougher E Tails	5.7	49.8	2533.8	816.0	1655.5	21.1	1.0	6716.0	225.7	1414.3	394.9	5301.7
Scavenger A Conc	87.0	50.7	1228.2	88.8	1087.2	8.5	13.9	61.2	12.5	5.5	51.1	55.8
Scavenger B Conc	92.2	50.9	1754.5	109.8	1592.3	12.0	12.1	71.4	17.3	8.6	51.9	62.8
Scavenger C Conc	107.7	50.9	1427.1	88.0	1286.7	12.6	10.4	49.0	12.0	6.2	30.6	42.9
Scavenger D Conc	84.6	49.8	1656.9	97.9	1507.7	13.0	8.7	69.4	17.8	9.0	36.2	60.4
Scavenger E Conc	91.0	50.4	1759.9	99.3	1608.7	12.5	6.4	65.5	17.7	8.2	25.3	57.3
Scavenger E Tails	4.8	51.2	1775.2	546.9	1175.6	21.5	0.8	6854.9	245.9	1473.8	341.9	5381.0
HGCC Conc	120.0	51.0	764.3	256.0	455.8	2.8	78.2	128.0	3.8	3.6	800.6	124.4
HGCC Tails	120.0	51.0	982.0	162.7	766.8	4.2	55.2	81.4	6.4	3.4	269.4	78.0
HGCC Conc	90.0	50.7	1137.3	130.1	955.0	5.4	31.0	86.7	10.6	4.7	161.3	82.0
HGCC Tails	23.8	51.1	2002.5	201.9	1748.0	11.1	8.6	258.7	40.3	28.8	133.5	229.9
LGCC Conc	118.0	50.0	1256.9	62.1	1143.3	5.7	29.5	59.9	9.7	3.4	106.0	56.4
LGCC Tails	118.0	50.5	970.0	79.7	838.3	6.8	25.0	40.5	7.1	2.8	60.6	37.8
LGCC Conc	87.0	50.4	1119.7	69.8	998.0	7.6	18.2	48.1	11.5	3.6	52.6	44.5
LGCC Tails	44.0	51.4	1269.5	75.2	1141.4	13.8	5.6	426.8	69.3	59.0	144.2	367.8

CIRCUIT Measured	Feed Characterisation - Mass Balance 2											
	Duration (sec)	Mass Bag (g)	Wet Mass+Bag+Clip (g)	Dry Mass (g)	Water Mass (g)	Chromite (%)	PGM (g/t)	Dry Mass Flow (g/min)	Water Flow (ml/sec)	Chromite (g/min)	PGM Flow (mg/hr)	Gangue Flow (g/min)
Feed	4.5	49.9	1682.2	481.1	1149.7	20.8	3.9	6372.2	253.8	1328.3	1502.6	5043.8
Rougher A Conc	60.6	58.6	1483.1	217.6	1265.4	6.2	52.5	215.6	19.9	13.3	679.1	202.2
Rougher B Conc	60.6	58.5	1603.8	172.6	1371.2	8.1	27.1	170.8	22.6	13.8	277.7	157.0
Rougher C Conc	60.6	58.9	1199.6	92.3	1046.9	10.7	9.9	91.4	17.3	9.8	54.4	81.7
Rougher D Conc	90.8	58.0	1531.0	117.8	1353.7	11.0	12.8	77.9	14.9	8.6	59.8	69.3
Rougher E Conc	90.7	58.3	1398.9	96.5	1240.6	11.9	11.0	65.2	13.7	7.8	43.0	57.4
Rougher E Tails	7.3	51.3	2175.2	770.3	1352.1	21.0	0.8	5751.3	165.4	1209.4	289.9	4541.9
Scavenger A Conc	55.1	58.4	1727.8	99.4	1588.5	11.8	12.2	108.2	28.5	12.7	79.2	95.5
Scavenger B Conc	38.1	58.4	1369.5	72.8	1226.8	12.6	10.5	114.6	32.2	14.5	72.2	100.1
Scavenger C Conc	75.4	49.9	1501.9	77.5	1373.0	13.4	8.5	61.6	18.2	8.3	31.4	53.4
Scavenger D Conc	61.2	50.3	1727.0	90.0	1595.2	11.6	7.3	88.3	25.9	10.2	38.8	78.0
Scavenger E Conc	57.7	50.7	2181.4	112.7	2016.5	14.9	6.1	117.2	36.0	17.4	42.7	99.8
Scavenger E Tails	6.0	50.1	1822.7	579.2	1191.9	21.5	0.9	5821.1	199.6	1258.5	310.8	4562.6
HGCC Conc	120.3	51.7	754.1	242.6	458.3	3.4	80.4	121.0	3.8	4.2	583.5	116.8
HGCCB Conc	120.6	50.9	1104.4	174.1	877.9	4.7	55.7	86.6	7.3	4.1	289.5	82.5
HGCCC Conc	90.2	50.6	1195.9	131.2	1012.6	4.6	32.4	87.3	11.2	4.1	169.6	83.2
HGCC Tails	15.4	50.2	1805.9	137.8	1616.4	11.6	9.6	326.1	86.1	38.0	188.4	208.1
LGCC Conc	111.1	48.1	1463.7	116.4	1297.7	7.1	28.5	62.9	11.7	4.5	107.5	58.4
LGCCB Conc	115.0	48.5	1309.1	91.7	1167.4	7.1	28.5	47.9	10.2	4.2	26.8	43.6
LGCCC Conc	96.2	49.3	1468.4	85.9	1331.7	7.6	20.3	53.6	13.8	4.1	65.3	49.5
LGCC Tails	18.1	50.3	3701.3	241.3	3408.2	13.6	6.5	651.7	170.1	88.3	254.2	583.4

CIRCUIT Measured	Feed Characterisation - Mass Balance 3											
	Duration (sec)	Mass Bag (g)	Wet Mass+Bag+Clip (g)	Dry Mass (g)	Water Mass (g)	Chromite (%)	PGM (g/t)	Dry Mass Flow (g/min)	Water Flow (mL/sec)	Chromite (g/min)	PGM Flow (mg/hr)	Gangue Flow (g/min)
Feed	5.6	58.4	2226.7	626.1	1540.7	19.1	3.9	6708.2	275.1	1284.0	1553.6	5424.2
Rougher A Conc	90.7	59.3	1509.8	288.3	1160.7	5.0	44.2	190.8	12.8	9.4	506.0	181.3
Rougher B Conc	92.5	58.8	1694.5	229.2	1405.0	6.9	29.2	148.6	15.2	10.2	260.4	138.4
Rougher C Conc	90.6	59.2	1400.3	128.1	1211.5	8.6	16.6	84.9	13.4	7.3	84.5	77.6
Rougher D Conc	90.6	59.0	1464.6	119.6	1284.5	9.9	12.8	79.2	14.2	7.8	60.8	71.3
Rougher E Conc	90.7	59.2	1373.1	92.5	1219.9	10.7	11.7	61.2	13.5	6.5	43.0	54.7
Rougher E Tails	14.4	59.4	3604.9	1186.0	2359.0	20.1	1.0	6143.5	206.1	1235.8	353.9	4907.7
Scavenger A Conc	53.5	59.0	1521.2	87.7	1373.0	11.7	12.0	98.3	25.6	11.5	70.8	86.8
Scavenger B Conc	55.3	58.7	1914.9	105.2	1749.5	12.3	8.6	114.2	31.7	14.0	59.0	100.2
Scavenger C Conc	84.4	58.5	1557.8	84.3	1413.5	12.4	8.8	59.9	16.7	7.4	31.5	52.5
Scavenger D Conc	57.9	58.9	1816.4	97.1	1558.9	13.8	6.5	100.7	28.7	13.9	39.5	86.8
Scavenger E Conc	50.9	58.9	1814.1	100.4	1653.3	13.8	5.9	118.3	32.5	16.4	41.9	101.9
Scavenger E Tails	6.0	58.4	1800.8	563.0	1177.9	20.2	0.8	6251.6	222.1	1256.7	288.8	4995.9
HGCC Conc	120.5	58.5	797.8	261.0	476.8	3.6	73.5	130.0	4.0	4.7	573.1	125.2
HGCC Conc	120.6	58.2	1105.6	179.5	866.4	4.5	51.3	89.3	7.2	4.0	275.0	85.3
HGCC Conc	90.8	58.1	1110.4	124.9	925.9	29.9	8.6	82.6	10.2	5.4	148.1	77.1
HGCC Tails	22.3	58.8	1966.1	158.0	1747.8	10.3	8.6	262.9	47.7	27.1	135.0	236.7
LGCC Conc	113.5	58.1	1425.1	118.4	1247.1	7.3	21.7	62.6	11.0	4.6	91.5	58.0
LGCC Conc	118.8	58.8	1157.8	86.1	1022.4	7.1	21.7	43.0	8.6	3.1	56.0	39.9
LGCC Conc	104.6	59.1	1411.7	85.7	1265.4	7.5	17.9	49.2	12.1	3.7	52.8	45.5
LGCC Tails	12.4	58.8	2046.9	122.2	1864.4	13.9	5.7	599.5	151.1	83.1	204.3	516.4

CIRCUIT Measured	Feed Characterisation - Mass Balance 4											
	Duration (sec)	Mass Bag (g)	Wet Mass+Bag+Clip (g)	Dry Mass (g)	Water Mass (g)	Chromite (%)	PGM (g/t)	Dry Mass Flow (g/min)	Water Flow (ml/sec)	Chromite (g/min)	PGM Flow (mg/hr)	Gangue Flow (g/min)
Feed	9.0	58.7	3455.1	959.0	2435.9	20.7	4.0	6414.7	271.6	1328.7	1539.5	5086.0
Rougher A Conc	90.7	58.6	1502.5	282.0	1160.4	5.4	55.8	186.5	12.8	10.1	624.4	176.4
Rougher B Conc	90.5	58.5	1866.6	219.2	1367.4	7.3	32.0	145.4	15.3	10.6	279.1	134.7
Rougher C Conc	90.5	58.8	1172.2	112.1	999.8	7.8	18.0	74.3	11.0	5.8	80.2	68.5
Rougher D Conc	90.8	58.5	1224.4	106.4	1058.0	9.7	13.9	70.3	11.7	6.8	58.6	63.5
Rougher E Conc	90.5	58.7	1181.7	94.4	1027.1	11.2	12.6	62.6	11.4	7.0	47.3	55.6
Rougher E Tails	6.5	58.8	1906.0	618.6	1227.1	21.9	1.0	5683.9	187.9	1245.3	344.4	4438.6
Scavenger A Conc	60.9	58.9	1951.3	115.2	1785.8	11.9	13.2	82.4	21.3	9.8	65.3	72.6
Scavenger B Conc	63.8	58.7	1746.4	94.3	1591.9	13.1	10.4	88.6	24.9	11.6	55.3	77.0
Scavenger C Conc	94.6	58.9	1543.5	84.0	1399.1	12.2	9.7	53.3	14.8	6.5	30.9	46.8
Scavenger D Conc	65.4	59.0	2062.0	107.2	1894.3	15.2	6.2	96.4	29.0	14.9	36.6	83.4
Scavenger E Conc	62.1	58.7	2131.5	117.0	1954.3	14.8	6.1	113.0	31.5	16.7	41.0	96.3
Scavenger E Tails	4.8	59.1	2070.9	687.4	1322.9	21.8	0.9	5867.8	225.0	1274.7	306.3	4593.0
HGCC Conc	120.5	58.9	813.0	251.7	500.9	3.9	77.8	125.4	4.2	4.9	595.2	120.5
HGCC Conc	120.3	59.3	1171.0	195.4	914.8	3.8	56.0	97.4	7.6	3.7	327.3	93.7
HGCC Conc	120.5	59.3	1549.8	186.6	1322.4	6.0	33.3	82.9	11.0	4.9	165.7	78.0
HGCC Tails	28.3	59.7	1542.5	122.4	1358.9	11.7	9.8	259.7	48.1	30.5	152.1	229.2
LGCC Conc	120.7	59.1	1565.1	109.0	1395.5	8.0	27.1	54.2	11.6	4.4	88.1	49.8
LGCC Conc	119.9	59.2	1196.9	78.5	1057.7	8.1	23.2	39.3	8.8	3.2	54.7	36.1
LGCC Conc	99.3	59.3	1508.2	78.8	1368.6	8.9	18.3	47.6	13.8	4.2	52.3	43.4
LGCC Tails	9.8	58.6	1716.9	101.3	1555.5	14.4	5.4	619.6	158.6	89.0	201.9	530.6

CIRCUIT	Circuit Configuration B - Mass Balance 1											
Measured	Duration	Mass Bag	Wet Mass+Bag+Clip	Dry Mass	Water Mass	Chromite	PGM	Dry Mass Flow	Water Flow	Chromite	PGM Flow	Gangue Flow
	(sec)	(g)	(g)	(g)	(g)	(%)	(g/l)	(g/min)	(ml/sec)	(g/min)	(mg/hr)	(g/min)
Feed	5.8	41.9	2306.8	655.4	1635.8	20.6	3.7	6756.7	276.1	1369.9	1504.0	536.8
Rougher A Conc	90.7	41.7	1850.9	322.8	1484.9	4.6	47.3	213.6	16.4	9.7	606.1	203.8
Rougher B Conc	44.0	41.5	1035.7	144.4	848.3	5.5	33.3	196.8	19.3	10.8	393.2	185.9
Rougher C Conc	60.6	41.4	2191.6	195.4	1953.3	7.7	17.5	193.5	32.2	14.9	203.2	178.6
Rougher D Conc	44.9	41.5	1370.2	114.2	1213.0	7.7	15.9	152.7	27.0	11.7	145.7	141.0
Rougher E Conc	61.1	41.6	2027.3	149.9	1834.3	8.8	12.2	147.2	30.0	12.9	107.8	134.3
Rougher E Tails	6.6	41.7	3239.6	776.4	2420.0	19.9	1.4	7101.2	388.9	1412.4	579.5	5688.8
Scavenger A Conc	77.2	41.3	1546.8	109.8	1354.2	9.3	21.8	85.4	18.1	7.9	111.6	77.4
Scavenger B Conc	46.3	41.4	1570.8	97.6	1430.3	10.8	9.5	126.5	30.9	13.7	71.8	112.9
Scavenger C Conc	76.8	41.7	1885.2	112.0	1730.0	11.4	8.3	87.5	22.5	10.0	43.7	77.5
Scavenger D Conc	61.3	41.5	1689.0	95.8	1550.2	11.5	7.6	93.7	25.3	10.8	42.6	83.0
Scavenger E Conc	60.2	41.8	1770.9	97.5	1630.1	12.9	7.0	97.2	27.1	12.5	41.0	84.7
Scavenger E Tails	11.6	42.0	3556.8	1053.0	2470.3	21.0	1.0	6610.9	245.0	1365.4	388.7	5225.4
Cleaner A Conc	61.4	42.0	1259.2	240.8	974.9	4.4	54.4	235.5	15.9	10.4	769.7	225.1
Cleaner B Conc	90.5	41.8	1366.6	197.9	1125.4	4.5	44.0	131.2	12.4	5.9	346.4	125.3
Cleaner C Conc	60.7	42.0	976.2	107.9	824.8	4.5	29.4	106.6	13.6	4.8	188.1	101.9
Cleaner Tails	19.2	41.6	2034.7	159.2	1832.4	8.9	11.1	498.8	95.7	44.5	332.2	454.3
Re-Cleaner A Conc	95.3	42.0	642.2	219.2	379.5	2.8	57.1	138.0	4.0	3.8	555.6	134.2
Re-Cleaner B Conc	150.1	42.2	1199.3	372.9	782.7	3.0	39.6	149.1	5.2	4.4	354.2	144.7
Re-Cleaner Tails	76.1	41.9	2620.8	217.8	2359.6	7.2	15.6	171.8	31.0	12.4	160.8	159.4

CIRCUIT Measured	Circuit Configuration B - Mass Balance 3											
	Duration (sec)	Mass Bag (g)	Wet Mass+Bag+Clip (g)	Dry Mass (g)	Water Mass (g)	Chromite (%)	PGM (g/l)	Dry Mass Flow (g/min)	Water Flow (ml/sec)	Chromite (g/min)	PGM Flow (mg/hr)	Gangue Flow (g/min)
Feed	7.6	42.5	3120.9	945.1	2131.8	19.8	3.6	7432.0	279.4	1468.8	1523.1	5863.2
Rougher A Conc	90.5	42.9	1452.1	309.1	1089.6	4.5	49.5	204.9	12.1	9.2	608.6	195.7
Rougher B Conc	90.5	43.1	1421.6	248.0	1129.0	4.6	35.2	164.5	12.5	7.6	347.4	156.9
Rougher C Conc	90.7	43.2	2409.6	237.5	2127.4	7.2	19.6	157.2	23.5	11.4	184.8	145.8
Rougher D Conc	88.4	42.8	1621.0	164.4	1412.3	6.9	18.1	111.6	18.0	7.7	121.2	103.9
Rougher E Conc	95.8	42.6	2209.4	130.0	2039.3	6.7	14.6	81.5	21.3	5.5	71.4	76.0
Rougher E Tails	6.3	42.6	2911.5	698.4	2179.0	20.0	1.5	7605.3	384.7	1520.0	693.6	6085.3
Scavenger A Conc	112.6	42.7	2000.7	147.3	1809.2	9.6	12.9	78.5	16.1	7.5	60.8	70.9
Scavenger B Conc	69.4	42.6	2013.6	131.7	1837.8	10.0	9.7	115.5	26.9	11.5	67.5	104.0
Scavenger C Conc	109.2	42.8	2232.2	138.8	2049.1	11.3	8.6	76.3	18.8	8.6	39.4	67.7
Scavenger D Conc	90.8	42.8	2212.1	134.1	2033.7	11.2	7.9	88.7	22.4	10.0	41.8	78.7
Scavenger E Conc	90.4	42.7	2631.5	92.5	2494.8	11.0	7.6	61.4	27.6	6.8	28.0	54.5
Scavenger E Tails	6.2	43.7	2462.6	722.5	1694.9	19.0	0.8	7037.3	275.1	1338.6	320.9	5898.7
Cleaner A Conc	90.3	43.7	1553.3	344.1	1164.0	3.9	51.2	226.7	12.9	8.9	702.5	219.8
Cleaner B Conc	90.7	43.6	1318.1	204.1	1088.9	4.5	42.6	135.0	11.8	6.1	345.1	128.9
Cleaner C Conc	69.3	43.6	1187.9	138.2	1004.6	5.1	28.7	119.6	14.5	6.2	205.9	113.4
Cleaner Tails	36.3	43.5	3196.9	286.1	2865.8	8.3	11.3	472.6	78.9	39.3	320.4	433.4
Re-Cleaner A Conc	148.2	43.4	1113.9	405.7	663.3	2.8	84.5	154.3	4.5	4.6	833.0	159.7
Re-Cleaner B Conc	120.5	43.4	925.5	280.2	600.4	3.0	43.7	139.5	5.0	4.2	365.7	135.3
Re-Cleaner Tails	89.1	43.6	2693.9	226.7	2322.1	6.6	15.5	152.7	26.1	10.1	142.0	142.6

CIRCUIT Measured	Circuit Configuration B - Mass Balance 2											
	Duration (sec)	Mass Bag (g)	Wet Mass+Bag+Clip (g)	Dry Mass (g)	Water Mass (g)	Chromite (%)	PGM (g/l)	Dry Mass Flow (g/min)	Water Flow (ml/sec)	Chromite (g/min)	PGM Flow (mg/hr)	Gangue Flow (g/min)
Feed	6.0	41.8	2255.6	627.9	1554.4	19.5	3.9	6310.6	265.4	1225.7	1472.9	5080.8
Rougher A Conc	85.1	41.8	1362.0	251.8	1086.9	4.8	43.3	226.5	16.1	10.9	593.6	217.6
Rougher B Conc	64.1	41.8	1527.3	218.0	1286.0	5.3	33.6	204.1	19.8	10.8	411.4	193.3
Rougher C Conc	46.5	41.9	1836.9	143.0	1650.5	6.5	14.3	184.5	35.5	12.0	158.3	172.5
Rougher D Conc	60.4	41.6	1747.4	148.7	1555.6	7.6	16.0	147.7	25.8	11.3	141.8	136.4
Rougher E Conc	60.5	41.6	2168.2	168.7	1956.4	8.3	13.9	187.3	32.3	13.8	139.5	153.5
Rougher E Tails	5.8	43.7	2721.6	618.5	2057.9	19.5	1.4	6453.9	357.9	1256.7	546.0	5196.2
Scavenger A Conc	82.5	41.5	1875.7	127.3	1705.4	9.7	13.1	92.5	20.7	9.0	72.7	83.5
Scavenger B Conc	70.5	42.0	2214.1	136.5	2034.1	10.7	7.4	116.1	28.8	12.5	51.5	103.6
Scavenger C Conc	60.2	41.4	1431.1	83.9	1304.3	11.6	8.7	83.6	21.7	9.7	43.4	73.9
Scavenger D Conc	69.9	44.4	1930.8	110.1	1774.8	11.9	8.0	94.5	25.4	11.2	45.4	83.3
Scavenger E Conc	62.6	43.8	2103.0	118.1	1939.6	12.8	6.6	113.2	31.0	14.5	44.6	98.8
Scavenger E Tails	14.5	42.0	5453.6	1577.9	3832.2	20.9	0.6	5953.9	230.3	1243.5	225.1	4710.3
Cleaner A Conc	75.7	41.9	1517.3	320.5	1153.4	3.9	53.4	254.1	15.2	10.0	614.0	244.0
Cleaner B Conc	90.2	41.9	1329.9	213.8	1072.7	4.2	43.9	142.2	11.9	6.0	374.6	136.2
Cleaner C Conc	90.1	42.1	1553.4	186.1	1323.7	5.2	29.1	123.9	14.7	6.5	216.3	117.4
Cleaner Tails	30.3	41.9	2462.1	198.3	2220.4	8.6	11.8	382.5	73.3	33.8	277.9	358.7
Re-Cleaner A Conc	120.8	41.7	779.9	291.0	445.7	2.5	71.1	144.6	3.7	3.6	616.7	140.9
Re-Cleaner B Conc	120.5	42.2	923.5	295.5	583.3	2.8	40.0	147.7	4.8	4.1	354.4	143.6
Re-Cleaner Tails	90.1	41.9	2110.5	149.4	1917.7	8.2	12.7	228.0	33.3	18.6	173.7	209.3

CIRCUIT Measured	Circuit Configuration C - Mass Balance 1											
	Duration (sec)	Mass Bag (g)	Wet Mass+Bag+Clip (g)	Dry Mass (g)	Water Mass (g)	Chromite (%)	PGM (g/l)	Dry Mass Flow (g/min)	Water Flow (ml/sec)	Chromite (g/min)	PGM Flow (mg/hr)	Gangue Flow (g/min)
Feed	7.8	41.6	2951.3	825.1	2083.1	19.6	3.7	6363.2	267.8	1247.7	1424.1	5115.5
Rougher A Conc	93.6	42.2	1190.3	270.0	876.6	4.8	61.8	173.2	9.4	8.2	642.0	164.9
Rougher B Conc	85.7	41.8	1185.5	195.6	946.6	5.6	34.8	137.0	11.0	7.6	286.0	129.4
Rougher C Conc	102.2	41.7	1711.8	156.2	1512.4	7.8	19.2	91.7	14.8	7.2	105.7	84.5
Rougher D Conc	117.0	41.8	1746.4	160.4	1544.7	8.7	15.1	82.2	13.2	7.1	74.5	75.1
Rougher E Conc	123.5	41.2	1400.8	121.3	1236.8	8.9	14.7	58.9	10.0	5.2	52.0	53.7
Rougher E Tails	4.4	41.2	1864.9	472.9	1149.3	22.5	1.2	6522.8	264.2	1469.8	454.0	5063.0
Rougher F Conc	120.3	41.6	1187.1	79.4	1064.6	11.4	14.7	39.6	8.8	4.5	34.9	35.1
Rougher G Conc	90.4	41.5	1663.8	105.5	1515.3	12.3	11.2	70.0	16.8	8.6	47.0	61.4
Rougher H Conc	90.1	41.6	1171.4	72.6	1055.7	13.0	9.5	48.4	11.7	6.3	27.6	42.1
Rougher I Conc	90.4	41.5	1253.7	74.9	1135.8	13.6	7.8	49.7	12.6	6.8	23.1	42.9
Rougher J Conc	76.0	41.3	1851.0	81.3	1726.9	11.5	6.8	64.2	22.7	7.4	26.0	56.8
Rougher J Tails	8.7	42.1	3496.5	1177.7	2277.2	21.7	0.9	6250.9	191.6	1357.3	348.8	4893.6
Cleaner A Conc	88.7	42.7	1154.0	237.0	872.8	4.3	56.5	160.3	9.8	8.9	543.5	153.4
Cleaner B Conc	122.4	42.3	1307.1	193.1	1070.2	4.9	52.7	94.6	8.7	4.6	299.2	90.0
Cleaner C Conc	167.0	42.1	688.6	115.3	709.7	5.3	35.0	41.4	4.2	2.2	87.0	39.2
Cleaner C Tails	30.2	41.6	3023.5	228.9	2751.5	14.0	4.5	454.9	91.1	63.8	123.6	381.2
Cleaner D Conc	120.3	42.5	1124.1	124.2	955.9	5.3	22.4	81.9	7.9	3.3	83.2	98.7
Cleaner E Conc	120.5	42.1	1327.9	112.3	1170.0	5.6	18.9	58.9	9.7	3.2	64.5	53.7
Cleaner F Conc	120.4	42.3	1149.8	88.8	1017.2	6.0	17.1	44.2	8.4	2.6	45.4	41.6
Cleaner F Tails	31.8	41.5	1937.1	145.4	1748.7	14.4	5.3	274.3	55.0	39.5	87.2	234.7

Circuit Configuration C - Mass Balance 2													
CIRCUIT	Measured	Duration	Mass Bag	Wet Mass+Bag+Clip	Dry Mass	Water Mass	Chromite	PGM	Dry Mass Flow	Water Flow	Chromite	PGM Flow	Gangue Flow
		(sec)	(g)	(g)	(g)	(g)	(%)	(g)	(g/min)	(ml/sec)	(g/min)	(mg/hr)	(g/min)
Feed		7.6	41.9	2933.5	863.6	2026.5	18.5	3.5	8817.9	286.6	1280.8	1427.7	5567.0
Rougher A Conc		121.3	41.5	1602.6	388.6	1191.0	4.3	53.6	182.4	9.8	7.8	586.6	174.6
Rougher B Conc		117.9	41.3	1360.1	260.0	1057.3	5.1	34.0	132.3	9.0	6.8	270.0	125.5
Rougher C Conc		105.0	41.3	1849.7	183.7	1623.2	7.6	11.2	105.0	15.5	8.0	70.5	97.0
Rougher D Conc		104.6	41.4	1745.3	154.5	1547.9	8.4	15.6	88.6	14.8	7.5	82.9	81.1
Rougher E Conc		113.3	41.6	1541.0	139.6	1368.3	8.3	15.6	73.9	12.0	6.2	69.2	67.7
Rougher E Tails		3.9	41.3	1325.1	448.2	834.1	24.6	1.0	6579.7	281.7	1616.3	398.7	4562.5
Rougher F Conc		90.4	41.2	1055.0	106.0	1012.3	10.7	12.9	40.1	11.2	4.3	31.0	35.8
Rougher G Conc		90.3	41.5	1662.6	106.0	1513.6	11.6	9.6	70.4	16.8	8.2	40.7	62.2
Rougher H Conc		90.2	41.8	1437.9	85.0	1308.6	12.1	8.6	56.6	14.6	6.8	29.1	49.7
Rougher J Conc		90.4	41.6	1329.2	78.9	1207.2	12.9	7.6	52.4	13.4	6.7	23.9	45.6
Rougher J Tails		80.2	41.3	1790.0	87.9	1689.3	13.6	4.9	87.6	27.6	11.9	25.9	75.7
Cleaner A Conc		8.6	41.7	2755.8	851.4	1851.2	20.1	0.8	5940.0	216.4	1195.7	285.1	4744.3
Cleaner B Conc		121.1	59.3	1680.8	368.4	1251.6	4.3	82.2	182.6	10.3	7.8	881.4	174.8
Cleaner C Conc		129.5	58.9	1743.3	227.7	1455.2	5.5	39.9	105.5	11.2	5.8	252.5	98.7
Cleaner C Tails		141.1	58.9	806.2	120.4	624.4	4.7	35.0	51.2	4.4	2.4	107.5	46.8
Cleaner D Conc		29.1	41.6	2843.8	211.9	2589.8	10.9	8.9	437.4	89.1	47.9	233.5	389.5
Cleaner E Conc		120.4	59.1	1332.5	176.2	1086.7	5.1	31.2	87.8	9.1	4.5	164.3	83.3
Cleaner F Conc		120.6	59.0	1378.0	112.1	1206.4	6.2	16.9	55.8	10.0	3.4	56.5	52.3
Cleaner F Tails		26.0	41.6	2333.4	180.1	2130.2	14.0	4.4	343.1	76.1	47.9	90.4	295.1

Circuit Configuration C - Mass Balance 3													
CIRCUIT	Measured	Duration	Mass Bag	Wet Mass+Bag+Clip	Dry Mass	Water Mass	Chromite	PGM	Dry Mass Flow	Water Flow	Chromite	PGM Flow	Gangue Flow
		(sec)	(g)	(g)	(g)	(g)	(%)	(g)	(g/min)	(ml/sec)	(g/min)	(mg/hr)	(g/min)
Feed		9.3	58.6	5654.5	1077.4	4517.0	20.3	3.7	8986.5	283.7	1421.3	1558.2	5567.2
Rougher A Conc		120.6	59.0	1338.1	337.5	940.1	4.8	65.5	167.9	7.8	8.0	659.9	159.9
Rougher B Conc		102.6	59.4	1164.7	216.6	887.2	5.7	36.4	126.7	8.6	7.3	276.7	119.4
Rougher C Conc		107.6	58.8	1777.9	176.6	1541.0	8.2	19.5	98.5	14.3	8.0	115.2	90.4
Rougher D Conc		107.5	59.2	1733.8	153.7	1519.4	8.8	17.1	85.8	14.1	7.5	88.0	78.3
Rougher E Conc		108.5	59.4	1080.9	107.3	922.7	8.2	16.9	59.4	8.5	4.8	80.2	54.5
Rougher E Tails		4.2	58.6	1610.0	531.3	1018.6	22.1	1.2	6783.3	301.3	1497.8	476.2	5285.5
Rougher F Conc		95.5	58.6	971.5	64.5	946.9	11.5	13.8	40.5	8.9	4.7	33.5	35.8
Rougher G Conc		90.1	58.9	1591.5	101.9	1429.2	12.2	11.1	67.9	15.9	8.3	45.2	59.6
Rougher H Conc		91.1	58.9	1214.8	74.6	1079.8	13.2	9.2	49.1	11.9	6.5	27.2	42.7
Rougher I Conc		90.3	59.9	1157.8	69.1	1027.3	13.6	8.6	45.9	11.4	6.2	23.8	39.7
Rougher J Conc		80.2	58.9	1478.8	82.7	1336.7	13.9	6.0	82.4	22.2	11.5	29.6	70.9
Rougher J Tails		9.6	59.1	3374.0	1087.0	2226.4	20.8	0.8	6800.8	232.2	1413.6	314.2	5387.3
Cleaner A Conc		118.2	58.7	1584.3	331.1	1193.0	4.3	70.6	188.1	10.1	7.2	712.0	180.9
Cleaner B Conc		119.1	59.3	1368.1	194.1	1133.2	5.1	49.8	97.8	9.5	5.0	292.2	92.8
Cleaner C Conc		147.5	59.3	1031.7	129.7	841.2	5.2	35.5	52.7	5.7	2.8	112.4	50.0
Cleaner C Tails		36.0	59.1	3067.0	237.9	2788.5	11.0	10.0	505.5	98.2	55.8	308.3	449.7
Cleaner D Conc		120.3	59.2	1086.6	119.2	906.7	5.5	25.1	59.5	7.5	3.3	89.6	56.2
Cleaner E Conc		119.8	59.4	1389.1	110.8	1217.4	5.8	18.4	55.5	10.2	3.2	61.2	52.3
Cleaner F Conc		120.0	59.2	1184.6	94.3	1029.6	6.1	16.4	47.2	8.6	2.9	46.4	44.3
Cleaner F Tails		39.8	59.2	3109.6	221.1	2827.8	11.2	9.6	333.0	71.0	37.2	192.4	295.7

Scale-up Test - Mass Balance 1													
CIRCUIT	Measured	Duration	Mass Bag	Wet Mass+Bag+Clip	Dry Mass	Water Mass	Chromite	PGM	Dry Mass Flow	Water Flow	Chromite	PGM Flow	Gangue Flow
		(sec)	(g)	(g)	(g)	(g)	(%)	(g)	(g/min)	(ml/sec)	(g/min)	(mg/hr)	(g/min)
80 Feed		7.8	42.1	2167.6	620.8	1503.2	19.1	3.7	4787.7	153.2	916.4	1080.0	3671.2
80 Rougher A Conc		124.4	41.7	1186.9	314.2	829.5	4.3	50.8	151.6	6.7	6.5	462.1	145.1
80 Rougher B Conc		72.3	41.6	1304.8	186.4	1096.1	7.5	27.2	138.1	15.1	10.3	225.4	127.8
80 Rougher C Conc		95.2	41.5	1012.5	102.3	867.2	9.4	15.9	64.5	9.1	6.1	61.5	58.4
80 Rougher D Conc		103.3	41.9	990.5	93.6	863.5	10.3	12.3	54.4	8.3	5.6	40.1	48.8
80 Rougher E Conc		104.2	42.0	1114.3	98.7	972.1	10.8	12.4	56.9	9.3	6.1	42.3	50.7
80 Rougher Tails		24.8	42.0	3249.5	1121.7	2084.3	20.9	0.9	4322.3	144.7	903.5	223.0	3418.7
30 Feed		33.9	41.9	2331.5	538.1	1780.0	17.2	4.9	953.8	51.7	164.1	278.7	789.7
30 Rougher A Conc		179.9	41.9	1114.6	155.4	915.8	6.5	42.8	51.8	5.1	3.4	133.1	46.5
30 Rougher B Conc		151.8	41.6	861.2	85.0	733.1	7.8	26.3	33.6	4.8	2.6	53.0	31.0
30 Rougher C Conc		122.9	41.9	815.8	55.4	717.0	8.0	15.2	27.1	5.8	2.2	24.7	24.9
30 Rougher Tails		62.3	42.0	3343.7	927.4	2372.8	17.2	1.1	892.6	38.1	154.0	56.8	736.6

Scale-up Test - Mass Balance 2													
CIRCUIT	Measured	Duration	Mass Bag	Wet Mass+Bag+Clip	Dry Mass	Water Mass	Chromite	PGM	Dry Mass Flow	Water Flow	Chromite	PGM Flow	Gangue Flow
		(sec)	(g)	(g)	(g)	(g)	(%)	(g)	(g/min)	(ml/sec)	(g/min)	(mg/hr)	(g/min)
80 Feed		7.3	41.4	2017.7	594.8	1380.0	19.2	3.8	4502.2	189.6	933.4	1111.8	3562.8
80 Rougher A Conc		144.8	41.4	1386.6	336.6	1007.9	5.1	59.1	139.1	7.0	7.1	493.3	132.0
80 Rougher B Conc		79.8	40.6	874.9	199.5	633.3	7.4	25.8	150.0	7.9	11.2	232.3	138.9
80 Rougher C Conc		113.7	41.4	1221.6	114.4	1064.3	9.2	16.0	60.4	9.4	5.5	57.9	54.8
80 Rougher D Conc		161.8	40.9	1511.4	137.4	1331.6	10.3	12.0	50.9	8.2	5.2	36.7	45.7
80 Rougher E Conc		96.5	41.6	1163.1	85.6	1034.4	10.9	12.2	53.2	10.7	5.8	38.9	47.4
80 Rougher Tails		21.6	41.7	1771.7	551.7	1176.9	19.4	1.0	4448.5	146.4	862.6	253.6	3986.0
30 Feed		35.1	41.8	2580.5	696.4	1840.8		4.6	1189.4	52.4	0.0	325.4	1189.4
30 Rougher A Conc		161.8	41.4	931.5	135.2	753.4	6.4	35.7	50.1	4.7	3.2	107.4	46.9
30 Rougher B Conc		151.9	41.8	1727.6	85.3	1599.0	7.1	25.0	33.7	10.5	2.4	50.5	31.3
30 Rougher C Conc		180.8	41.3	1145.3	76.1	1026.4	9.6	15.1	28.4	6.4	2.8	25.7	25.6
30 Rougher Tails		99.5	41.6	5080.8	1796.7	3212.0	20.3	0.9	1083.2	32.3	219.5	59.8	863.7

CIRCUIT Measured	Scale-up Test - Mass Balance 3										PGM (g/min)	Water Flow (m/sec)	Chromite (g/min)	PGM Flow (mg/hr)	Gangue Flow (g/min)
	Duration (sec)	Mass Bag (g)	Wet Mass+Bag+Clip (g)	Dry Mass (g)	Water Mass (g)	Chromite (%)	PGM (g)	Dry Mass Flow (g/min)	Water Flow (m/sec)	Chromite (g/min)					
80 Feed	8.5	41.3	2405.4	721.1	1641.5	20.6	3.9	5072.2	192.4	1046.9	1186.9	4025.3			
80 Rougher A Conc	154.1	42.1	1666.9	366.5	1246.8	5.6	62.4	142.7	8.1	7.9	534.4	134.8			
80 Rougher B Conc	94.4	41.9	2214.5	241.6	1929.5	8.5	26.9	153.6	20.5	13.0	248.0	140.6			
80 Rougher C Conc	132.2	42.2	1524.6	138.5	1342.4	10.0	17.7	62.8	10.2	6.3	86.7	56.6			
80 Rougher D Conc	167.9	42.3	1424.5	132.8	1247.9	10.5	15.8	47.5	7.4	5.0	46.0	42.5			
80 Rougher E Conc	111.9	41.5	1198.5	89.0	1066.5	11.7	12.8	47.7	9.5	5.6	36.6	42.1			
80 Rougher Tails	10.4	42.1	1828.5	647.1	1137.8	20.8	0.9	4617.8	136.8	969.4	260.4	3658.4			
30 Feed	34.1	42.2	2575.1	721.5	1809.9	20.0	4.5	1269.5	53.1	254.0	341.2	1015.5			
30 Rougher A Conc	173.6	42.2	1020.3	146.2	830.4	6.7	48.6	50.5	4.8	3.4	147.4	47.1			
30 Rougher B Conc	180.1	42.2	874.4	86.2	746.5	8.0	26.4	31.9	4.7	2.6	50.6	29.4			
30 Rougher C Conc	153.5	42.0	1062.1	68.7	949.9	10.2	17.6	26.8	6.2	2.7	28.4	24.1			
30 Rougher Tails	58.8	42.0	3327.4	1136.0	2147.9	20.6	1.1	1180.2	36.6	238.7	74.5	921.4			

CIRCUIT Measured	Low Grade Cleaner Test										PGM (g/min)	Water Flow (m/sec)	Chromite (g/min)	PGM Flow (mg/hr)
	Duration (sec)	Mass Bag (g)	Wet Mass+Bag+Clip (g)	Dry Mass (g)	Water Mass (g)	Chromite (%)	PGM (g)	Dry Mass Flow (g/min)	Water Flow (m/sec)	Chromite (g/min)				
150LC25 1	180.2	60.6	757.4	97.3	598.0	5.8	39.5	32.4	3.3	1.9	76.8			
150LC25 2	179.7	43.1	747.2	100.4	602.2	5.3	42.7	39.5	3.4	1.8	85.9			
150LFeed25 1	27.9	43.2	2085.6	252.6	1788.3	11.0	10.3	543.8	64.2	59.9	336.1			
150LFeed25 2	16.8	43.4	1592.3	205.4	1342.0	11.0	11.3	734.4	80.0	80.5	498.0			
150LC35 1	71.1	43.2	1390.9	131.3	1214.9	7.6	23.9	110.8	17.1	8.4	158.8			
150LC35 2	69.5	43.1	1312.0	123.2	1144.2	7.9	27.5	106.4	16.5	8.4	175.6			
150LFeed35 1	17.6	57.8	2102.9	257.0	1786.6	12.9	10.1	877.6	101.7	112.8	531.8			
150LFeed35 2	20.9	59.0	1724.3	161.2	1502.6	12.8	10.8	463.4	72.0	59.5	277.8			
150LC43 1	45.0	65.5	1380.2	120.4	1192.8	9.2	16.0	180.5	26.5	14.8	154.1			
150LC43 2	46.2	65.2	1398.1	122.0	1209.4	9.3	18.4	158.4	26.2	14.7	174.8			
150LFeed43 1	53.3	65.7	2199.7	254.9	1877.6	12.5	9.4	286.9	35.2	35.8	162.0			
150LFeed43 2	37.6	65.8	2626.1	313.0	2245.8	12.8	8.9	500.4	59.8	63.9	268.1			
120LC41 5 1	163.4	58.1	1118.6	165.5	893.5	6.3	39.6	80.8	5.5	3.8	144.4			
120LC41 5 2	163.6	58.3	1155.9	168.3	927.8	6.2	37.4	61.7	5.7	3.8	139.5			
120LFeed41 5 1	25.4	58.4	2544.6	297.2	2187.5	11.3	9.9	703.4	86.3	79.6	418.3			
120LFeed41 5 2	24.1	57.7	1883.6	168.7	1455.7	12.6	9.3	420.7	60.5	53.0	233.5			
120LC47 9 1	101.6	58.2	1260.2	156.3	1044.2	6.4	30.1	92.3	10.3	5.9	166.7			
120LC47 9 2	120.8	57.9	1452.1	182.0	1210.7	6.7	31.5	90.4	10.0	6.0	170.9			
120LFeed47 9 1	24.5	57.9	2110.9	191.9	1869.6	13.0	9.1	470.0	75.9	61.0	256.9			
120LFeed47 9 2	24.3	65.8	1750.2	168.9	1514.0	12.9	8.9	417.4	62.4	53.9	223.1			
120LC59 2 1	52.9	58.4	1736.8	161.4	1515.5	8.6	18.8	182.9	28.6	15.7	206.3			
120LC59 2 2	52.0	58.0	1674.8	148.0	1467.3	7.9	18.7	170.9	28.2	13.5	191.7			
120LFeed59 2 1	20.6	58.5	2054.5	276.8	1717.7	11.2	9.7	806.6	83.4	90.2	471.4			
120LFeed59 2 2	21.6	58.4	1974.7	153.0	1761.8	13.0	11.2	425.8	81.7	56.3	286.1			
90LC53 8 1	160.7	58.6	1296.2	202.5	1033.6	5.4	29.3	75.6	6.4	4.1	132.9			
90LC53 8 2	150.3	58.6	1288.0	197.8	1030.1	6.1	32.8	79.0	6.9	4.8	155.4			
90LFeed53 8 1	19.5	58.3	2040.6	217.3	1763.5	12.2	10.2	688.6	90.4	81.6	409.2			
90LFeed53 8 2	25.7	58.1	2398.2	239.8	2098.8	12.4	9.1	560.3	81.7	69.3	305.6			
90LC61 5 1	109.3	58.1	1667.6	189.2	1418.8	6.9	25.1	103.9	13.0	7.1	156.5			
90LC61 5 2	100.3	58.2	1630.7	177.9	1383.1	7.1	23.9	106.4	13.9	7.6	152.5			
90LFeed61 5 1	21.8	58.2	1513.0	147.0	1306.3	13.0	8.7	405.0	60.0	55.5	211.1			
90LFeed61 5 2	25.0	58.5	1894.2	174.7	1669.5	13.3	8.8	419.8	66.5	52.7	222.4			
90LC68 9 1	50.3	58.2	1476.7	126.8	1290.2	8.8	19.7	151.3	25.7	13.3	178.9			
90LC68 9 2	50.4	58.1	1360.9	117.2	1184.1	8.7	18.0	139.5	23.5	12.2	150.7			
90LFeed68 9 1	41.8	57.9	2449.4	233.8	2156.2	13.7	9.3	335.7	51.6	46.0	186.5			
90LFeed68 9 2	33.5	57.7	1960.4	186.3	1714.9	11.5	9.0	333.4	51.1	39.2	180.8			
60LC66 1 1	160.0	58.4	1414.7	168.4	1186.4	6.7	28.1	63.1	7.4	4.2	106.5			
60LC66 1 2	160.3	58.4	1454.6	178.9	1215.8	6.7	26.1	67.0	7.6	4.5	104.9			
60LFeed66 1 1	24.8	57.7	1640.3	183.3	1397.8	11.9	9.3	442.8	56.3	52.8	246.8			
60LFeed66 1 2	29.8	58.3	1979.4	223.5	1686.1	12.2	11.2	449.7	56.9	55.1	302.2			
60LC66 4 1	30.2	58.1	1373.0	104.8	1208.6	9.7	13.8	208.1	40.0	20.2	172.3			
60LC66 4 2	30.6	58.2	1354.0	102.5	1191.8	10.6	29.1	200.8	39.9	21.2	350.7			
60LFeed66 4 1	45.3	57.5	1898.5	239.4	1600.1	13.0	9.7	317.2	35.3	41.3	183.7			
60LFeed66 4 2	49.2	58.5	1941.0	180.1	1700.9	14.7	8.2	219.8	34.6	32.4	108.5			

CIRCUIT Measured	Rougher Test										Gangue Flow (g/min)	
	Duration (sec)	Mass Bag (g)	Wet Mass-Bag+Chp (g)	Dry Mass (g)	Water Mass (g)	Chromite (%)	PGM (g/l)	Dry Mass Flow (g/min)	Water Flow (ml/sec)	Chromite (g/min)		PGM Flow (mg/hr)
200RA10 1	110.2	57.1	972.0	183.8	729.5	4.7	80.6	100.1	6.6	4.7	484.0	95.4
200RA10 2	101.3	44.0	876.0	166.6	653.9	4.7	75.9	98.7	6.6	4.7	449.5	94.0
200RA10 3	110.4	44.0	948.7	182.0	721.2	4.7	73.3	98.9	6.5	4.7	434.9	94.2
200Feed10 1	8.1	57.6	2710.2	836.0	1815.1	21.3	3.8	6215.6	224.9	1321.0	1402.2	4884.6
200Feed10 2	8.3	44.0	2764.2	853.0	1865.7	21.0	3.7	6151.4	224.2	1293.7	1354.5	4857.7
200Feed10 3	8.6	44.0	2579.3	926.4	2007.4	20.6	3.7	6448.3	232.9	1330.0	1416.0	5118.3
200RA11 5.1	73.2	42.0	1391.6	194.6	1153.5	6.4	66.2	159.6	15.8	10.3	536.2	149.3
200RA11 5.2	80.4	42.0	1480.8	211.1	1226.2	6.4	80.0	157.6	15.3	10.0	567.2	147.5
200RA11 5.3	71.1	42.0	1277.9	185.1	1049.3	6.3	61.6	156.2	14.8	9.6	577.3	146.4
200Feed11 5.1	6.4	42.0	2236.1	702.5	1469.1	20.5	3.7	6617.0	233.8	1365.9	1457.1	5261.0
200Feed11 5.2	5.8	42.0	2032.2	631.6	1367.1	20.9	3.8	6489.0	232.4	1368.5	1483.4	5130.5
200Feed11 5.3	6.7	57.6	2391.7	737.8	1694.8	20.7	3.6	6666.8	239.8	1375.8	1425.9	5281.0
200RA15 8.1	28.2	42.0	1526.2	180.7	1302.0	9.9	30.3	384.6	48.2	38.1	689.2	346.5
200RA15 8.2	28.4	42.0	1568.3	166.1	1338.7	9.9	28.4	393.9	47.2	38.0	671.1	354.9
200RA15 8.3	28.8	42.0	1500.5	179.1	1277.9	9.6	31.6	372.6	44.3	35.9	706.5	336.7
200Feed15 8.1	5.6	42.0	2061.2	647.6	1370.1	20.1	3.7	6801.6	243.4	1385.8	1540.4	5515.8
200Feed15 8.2	6.0	42.0	2213.5	692.5	1477.5	19.6	3.7	6880.5	245.0	1363.9	1517.3	5636.6
200Feed15 8.3	6.6	42.0	1957.7	925.3	988.9	19.8	3.6	8450.2	150.5	1672.2	1840.5	6778.0
200RA16 9.1	15.8	57.6	1456.9	188.0	1229.8	12.3	20.7	638.8	77.9	78.5	793.4	560.2
200Feed16 9.1	5.3	57.7	1947.6	603.6	1284.8	20.1	3.8	6898.3	244.7	1384.0	1568.7	5514.3
170RA14 8.1	102.7	57.4	1235.1	263.7	912.5	4.6	64.5	154.0	8.9	7.1	596.2	147.0
170RA14 8.2	102.4	58.2	1177.3	263.2	854.4	4.7	62.2	154.2	8.3	7.3	575.3	146.9
170RA14 8.3	90.2	57.1	1166.3	233.6	874.1	4.7	58.0	155.4	9.7	7.4	560.0	148.0
170Feed14 8.1	5.5	57.5	1862.2	595.6	1207.6	19.4	3.4	6487.5	219.6	1257.6	1325.5	5239.8
170Feed14 8.2	5.3	57.4	1911.5	619.5	1233.1	19.4	3.9	7039.8	233.5	1367.6	1534.6	5672.1
170Feed14 8.3	5.9	57.8	1961.3	629.5	1272.5	19.8	3.7	6445.4	217.2	1279.2	1438.6	5166.2
170RA19 3.1	45.1	59.5	1821.3	205.9	1354.4	7.1	41.2	274.1	30.1	19.5	677.6	254.6
170RA19 3.2	46.6	42.0	1508.8	198.5	1266.8	7.0	42.4	255.6	27.2	18.0	650.3	237.6
170RA19 3.3	43.2	59.2	1326.4	182.6	1083.1	6.9	41.5	253.6	25.1	17.5	631.3	236.0
170Feed19 3.1	6.3	57.2	2304.4	722.3	1523.4	17.4	3.3	6836.6	240.3	1187.6	1361.7	5648.1
170Feed19 3.2	5.3	60.0	1953.3	569.0	1322.8	19.7	3.6	6429.4	249.1	1267.6	1377.2	5161.8
170Feed19 3.3	5.9	58.0	2036.5	626.1	1348.9	19.9	3.8	6442.1	230.6	1279.9	1480.4	5162.1
170RA23 5.1	27.5	58.4	1567.4	178.3	1329.2	8.7	29.4	389.0	48.3	33.8	686.2	355.2
170RA23 5.2	24.4	57.9	1805.6	167.5	1278.7	8.8	30.4	412.7	52.5	36.3	752.8	376.4
170RA23 5.3	26.2	59.1	1508.4	166.4	1281.4	9.0	29.1	380.9	48.9	34.3	665.1	346.7
170Feed23 5.1	4.3	58.0	2235.4	700.5	1475.4	19.7	3.8	9889.4	347.2	1944.3	2237.0	7945.1
170Feed23 5.2	5.3	58.6	1988.7	619.7	1288.9	19.7	3.8	6989.1	242.3	1379.3	1583.6	5809.8
170Feed23 5.3	5.2	57.7	1936.4	617.6	1259.6	20.2	3.3	7139.9	242.7	1439.5	1430.8	5700.4
170RA27 1	14.6	58.0	1569.7	189.4	1340.8	12.0	17.8	698.1	92.1	83.8	745.5	614.3
170Feed27 1	5.8	57.4	2087.5	666.6	1362.0	20.1	3.7	6955.8	236.9	1394.8	1550.9	5561.0
140RA19 5.1	91.0	57.7	976.7	248.7	688.8	4.6	60.6	164.0	7.3	7.5	596.2	156.5
140RA19 5.2	92.5	56.2	1046.2	254.9	731.6	4.7	58.9	165.3	7.9	7.7	584.3	157.6
140RA19 5.3	90.3	57.6	1026.0	253.3	713.6	4.6	65.1	168.4	7.9	7.7	657.8	160.7
140Feed19 5.1	5.2	58.2	1938.5	630.6	1248.2	20.8	3.6	7332.6	241.9	1526.6	1993.8	6805.9
140Feed19 5.2	5.8	57.5	2009.1	679.6	1360.5	20.3	3.7	7018.2	234.2	1428.1	1537.0	5580.1
140Feed19 5.3	5.8	57.8	2063.7	664.2	1340.2	19.8	3.7	6859.2	230.7	1360.5	1522.7	5498.7
140RA24 8.1	60.0	58.3	1621.0	236.8	1324.4	6.1	48.3	236.7	22.1	14.4	686.9	222.3
140RA24 8.2	50.4	58.3	1378.7	200.5	1118.4	6.5	48.3	238.8	22.2	15.6	892.1	223.3
140RA24 8.3	60.2	58.1	1698.0	241.9	1386.5	6.6	50.4	241.3	23.0	15.8	729.6	225.4
140Feed24 8.1	5.3	57.8	1941.3	565.1	1316.9	19.5	3.3	6373.3	247.5	1241.5	1277.2	5131.8
140Feed24 8.2	5.5	58.3	2016.8	649.8	1307.2	21.0	3.9	7088.7	237.7	1490.6	1646.0	5598.1
140Feed24 8.3	5.5	58.1	2052.1	637.0	1365.5	21.3	3.8	6949.1	246.5	1479.0	1596.9	5470.1
140RA29 1.1	27.0	57.5	1336.1	148.1	1129.0	9.0	36.6	329.1	41.8	29.7	722.7	299.4
140RA29 1.2	27.3	58.0	1534.7	167.9	1307.3	8.9	33.4	369.7	48.0	33.0	740.9	326.7
140RA29 1.3	32.7	58.0	1891.2	187.4	1444.3	9.0	36.2	344.0	44.2	31.0	726.4	312.9
140Feed29 1.1	5.1	56.7	1845.7	586.4	1201.1	21.2	3.8	6958.5	234.1	1456.7	1551.4	5401.7
140Feed29 1.2	6.4	58.2	1958.2	622.3	1276.2	21.5	3.6	6940.1	237.2	1494.4	1490.7	5445.7
140Feed29 1.3	5.2	58.0	1948.6	625.1	1264.0	20.2	3.6	7198.8	242.6	1456.9	1537.7	5741.9
140RA39 1	7.3	58.4	1294.0	138.1	1086.0	15.4	12.0	1142.9	151.2	176.4	822.9	966.5
140Feed39 1	6.1	58.6	2142.9	677.8	1405.0	20.7	3.6	6710.9	231.8	1392.2	1445.5	5318.6
110RA24 1	121.5	57.9	1216.9	347.0	810.5	4.6	61.1	171.4	6.7	8.0	628.4	163.4
110RA24 2	120.4	58.7	1204.8	338.8	805.8	4.7	64.8	169.8	6.7	7.9	656.4	160.9
110RA24 3	120.5	58.0	1154.0	333.2	761.3	3.7	60.7	166.0	6.3	6.1	604.4	159.8
110Feed24 1	7.2	57.3	2880.9	934.8	1887.3	21.1	3.7	7779.2	263.1	1638.0	1708.3	6141.2
110Feed24 2	5.4	57.9	2024.1	644.8	1319.9	21.4	3.8	7204.5	245.8	1544.8	1621.0	5659.7
110Feed24 3	5.8	58.1	2034.6	643.5	1331.5	20.6	3.9	6679.9	230.4	1376.9	1559.1	5303.0
110RA30 8.1	58.7	58.0	1979.3	265.6	1654.2	7.3	43.9	266.9	27.7	19.5	703.0	247.3
110RA30 8.2	58.5	58.1	1962.9	260.8	1642.5	7.6	42.6	267.5	28.1	20.3	683.7	247.2
110RA30 8.3	58.6	58.1	1980.9	265.0	1666.3	7.4	43.9	271.4	28.3	20.0	714.8	251.4
110Feed30 8.1	5.7	58.5	2112.5	684.6	1367.9	21.2	3.5	7181.1	239.1	1521.8	1516.7	5659.3
110Feed30 8.2	5.8	58.2	2089.0	673.6	1365.7	20.9	4.1	6992.4	234.6	1462.1	1707.5	5630.3
110Feed30 8.3	6.3	58.6	2307.3	745.0	1502.2	20.7	3.5	7060.5	236.9	1459.2	1472.1	5591.3
110RA36 5.1	26.5	58.2	1926.7	201.7	1885.3	10.5	27.9	457.2	62.9	48.0	765.3	408.1
110RA36 5.2	22.3	58.3	1889.4	164.5	1365.1	10.2	27.9	442.2	61.2	45.1	740.3	397.1
110RA36 5.3	19.4	58.2	1392.0	143.1	1189.2	10.4	29.4	441.7	61.2	45.1	779.1	395.6
110Feed36 5.1	5.5	58.8	1999.7	645.2	1294.2	20.5	3.7	7000.4	234.0	1434.3	1570.9	5556.0
110Feed36 5.2	5.2	58.8	1956.1	632.9	1262.9	20.2	3.3	7373.6	245.2	1492.8	1477.7	5880.8
110Feed36 5.3	5.4	58.5	2046.1	660.4	1325.7	20.2	3.9	7283.8	243.7	1469.9	1700.0	5813.9
110RA45 1.1	6.1	58.4	1147.4	132.8	954.7	15.6	12.7	1308.4	156.8	204.3	997.0	1104.1
110Feed45 1.1	5.3	57.5	1936.8	627.0	1249.8	20.2	3.9	7071.4	234.9	1431.0	1546.2	5640.4

CIRCUIT Measured	Cleaner and Scavenger Test										PGM Flow (mg/hr)
	Duration (sec)	Mass Bag (g)	Wet Mass-Bag+Clip (g)	Dry Mass (g)	Water Mass (g)	Chromite (%)	PGM (g/l)	Dry Mass Flow (g/min)	Water Flow (ml/sec)	Chromite (g/min)	
190CA15.1	176.8	58.6	1029.6	233.9	736.8	3.7	103.0	79.4	4.2	2.9	450.6
190CA15.2	162.5	58.4	975.0	221.0	694.1	3.8	104.0	81.6	4.3	3.1	509.1
190CFeed15.1	17.2	57.9	1817.3	231.4	1526.5	8.4	49.7	807.7	88.8	67.8	2408.5
190CFeed15.2	14.1	58.6	1392.6	155.5	1177.0	8.8	41.4	653.6	63.7	58.5	1648.3
250SA115.1	176.8	58.0	1354.6	98.8	1196.3	10.5	21.7	33.5	6.8	3.5	43.7
250SA115.2	162.5	58.3	1416.8	101.0	1256.0	10.4	22.3	37.3	7.7	3.9	48.9
250SFeed115.1	11.0	57.8	2101.1	1036.3	1005.5	26.9	0.9	5888.0	91.7	1526.1	312.9
250SFeed115.2	8.6	58.0	3358.3	1517.9	1781.9	24.1	0.8	10553.2	206.5	2542.6	500.2
190CA215.1	80.1	58.5	1334.2	238.9	1036.3	4.5	79.3	178.9	12.9	8.0	651.1
190CA215.2	70.5	58.5	1148.3	202.7	895.6	4.6	73.1	172.6	12.6	7.9	757.0
190CFeed215.1	20.8	57.6	2017.8	214.8	1743.9	8.8	33.6	620.2	83.9	54.5	1250.3
190CFeed215.2	17.4	58.4	1864.7	192.0	1612.8	7.5	33.5	652.1	92.7	49.9	1330.8
250SA12.1	34.4	58.1	1550.4	120.5	1370.3	14.4	10.9	208.9	39.8	30.2	137.3
250SA12.2	41.3	58.0	1802.8	140.2	1603.1	14.9	9.7	203.9	36.9	30.4	118.6
250SFeed12.1	13.5	58.2	2243.5	943.6	1240.2	23.3	0.7	4184.5	91.7	974.7	175.7
250SFeed12.2	11.9	57.7	2471.3	1046.3	1365.8	23.0	1.0	5275.5	114.8	1211.2	310.2
190CA31.1	38.3	58.1	1958.9	251.6	1647.7	6.2	52.9	384.7	43.1	24.5	1252.7
190CA31.2	62.2	58.1	2332.1	300.4	1972.1	6.2	52.1	290.0	31.7	18.0	906.4
190CFeed31.1	21.8	57.7	2079.1	191.9	1828.0	8.7	29.0	529.4	84.0	46.0	921.1
190CFeed31.2	20.0	57.5	2222.2	233.3	1929.9	8.8	35.5	701.0	96.6	61.7	1493.0
250SA14.1	30.8	58.1	1886.1	133.9	1492.6	15.4	7.8	260.7	46.4	40.2	121.5
250SA14.2	32.1	58.6	2128.2	177.4	1880.7	15.4	7.9	331.6	58.9	51.1	156.6
250SFeed14.1	6.2	58.1	2545.4	1212.2	1273.6	24.9	0.8	11807.1	206.8	2942.9	695.1
250SFeed14.2	6.8	58.2	2348.4	988.7	1321.0	23.4	1.0	8610.7	195.7	2011.1	906.3
180CA23.1	175.7	59.4	974.4	317.7	696.8	3.5	81.8	108.5	3.4	3.8	532.5
180CA23.2	187.1	58.5	882.6	286.2	636.4	3.5	77.4	102.8	3.2	3.6	477.3
180CFeed23.1	20.5	58.8	2193.4	246.2	1886.9	8.7	37.8	720.6	92.0	62.9	1634.3
180CFeed23.2	19.8	58.1	2089.2	220.4	1868.2	8.6	30.1	688.6	91.4	57.2	1207.4
210SA135.1	156.0	58.8	1808.4	125.6	1422.5	9.9	19.1	48.3	9.1	4.8	56.3
210SA135.2	123.5	58.8	1372.1	104.0	1207.8	10.0	19.1	50.5	9.8	5.0	57.9
210SFeed135.1	3.4	58.8	1900.0	625.9	1213.8	23.8	0.4	10816.9	352.8	2594.4	281.7
210SFeed135.2	4.4	58.4	2271.1	638.6	1572.6	21.7	1.5	8629.7	354.2	1871.6	756.0
180CA40.1	86.7	58.4	2004.3	338.8	1605.6	4.9	56.8	237.2	18.7	11.7	808.5
180CA40.2	93.6	58.0	2061.7	365.1	1647.1	4.9	56.4	227.7	17.6	11.2	757.0
180CFeed40.1	16.0	58.5	1567.4	214.4	1293.0	8.3	29.3	804.0	80.8	66.9	1413.4
180CFeed40.2	20.2	58.5	2036.2	164.5	1813.7	8.4	35.8	688.8	90.0	41.3	1052.1
210SA18.1	67.6	58.6	2212.0	140.7	2011.2	12.6	11.5	125.0	28.8	15.8	86.2
210SA18.2	73.2	58.6	2270.9	139.7	2071.1	12.7	10.3	114.5	28.3	14.5	70.7
210SFeed18.1	5.8	58.4	2017.7	564.6	1207.2	21.0	1.7	6081.9	250.1	1279.4	613.1
210SFeed18.2	5.0	57.8	1566.9	302.9	1204.7	21.0	2.2	3613.1	239.5	758.8	470.4
180CA54.1	28.4	58.2	1448.4	159.3	1229.4	6.8	40.8	337.0	43.3	23.0	825.0
180CA54.2	32.9	58.5	1680.8	191.2	1429.6	6.6	40.5	348.7	43.5	23.2	847.3
180CFeed54.1	16.8	58.0	1788.5	183.8	1545.2	8.4	36.2	654.9	91.8	55.1	1383.1
180CFeed54.2	14.9	57.7	1866.7	160.6	1445.9	8.4	29.2	646.3	97.0	54.4	1132.3
210SA24.1	14.4	58.0	1432.3	110.2	1262.6	15.8	5.4	459.8	87.8	72.6	147.6
210SA24.2	18.7	57.7	1528.4	115.5	1363.7	16.1	4.1	370.2	72.3	59.7	91.6
210SFeed24.1	4.8	58.0	1290.9	481.4	750.0	24.3	1.0	7641.3	198.4	1859.0	453.9
210SFeed24.2	4.7	58.1	2406.9	815.6	1533.7	22.5	1.2	10389.8	325.6	2335.6	760.5
130CA39.1	114.3	58.8	1260.7	337.0	863.4	4.0	68.6	177.0	7.6	7.0	728.6
130CA39.2	121.9	58.2	1363.5	357.6	946.2	3.9	63.2	176.0	7.8	6.8	667.3
130CFeed39.1	16.7	58.4	1918.0	207.0	1661.1	8.7	35.8	742.8	98.8	64.5	1596.6
130CFeed39.2	16.0	58.6	1736.4	172.5	1503.8	8.6	34.1	648.1	94.2	55.9	1326.0
170SA20.1	114.3	58.5	1840.6	126.0	1654.6	11.4	13.1	66.2	14.5	7.5	52.0
170SA20.2	121.9	58.5	1965.9	132.5	1773.4	11.6	13.3	65.2	14.5	7.6	52.0
170SFeed20.1	8.0	58.1	2704.9	945.5	1699.8	20.9	2.1	7054.8	211.7	1475.7	873.2
170SFeed20.2	5.9	58.5	1654.7	380.1	1214.6	22.0	0.9	3872.0	206.2	852.7	213.7
130CA49.1	71.9	57.8	1597.1	271.7	1286.1	5.0	49.5	226.7	17.6	11.2	673.3
130CA49.2	61.4	58.6	1307.5	208.0	1039.4	4.5	48.2	203.1	16.9	9.1	587.4
130CFeed49.1	15.9	58.6	1610.7	182.0	1368.6	8.8	38.9	686.4	85.0	60.7	1602.0
130CFeed49.2	13.8	58.4	1545.0	159.3	1325.8	8.9	36.1	693.6	96.2	62.1	1502.4
170SA24.1	49.0	58.1	2215.5	137.8	2018.1	12.9	8.6	168.6	41.2	21.8	87.2
170SA24.2	47.2	58.2	1896.2	113.6	1722.9	13.8	8.4	144.4	36.5	19.9	72.8
170SFeed24.1	4.8	58.2	2404.1	788.1	1576.3	21.8	1.1	9541.4	329.8	2099.4	647.9
170SFeed24.2	5.1	58.6	2170.5	721.2	1389.2	23.2	1.2	8634.9	274.0	1977.3	604.3
130CA64.1	37.4	58.5	2082.7	215.1	1807.6	6.9	38.7	344.8	48.3	23.8	800.6
130CA64.2	33.8	58.1	2632.8	291.3	2481.9	6.4	36.5	517.4	73.5	33.2	1133.1
130CFeed64.1	13.4	58.5	1422.7	124.4	1238.3	8.4	29.5	559.1	92.8	46.9	989.6
130CFeed64.2	14.4	58.2	1641.5	170.0	1411.8	8.4	35.5	706.4	97.8	59.4	1504.6
170SA27.1	23.2	59.5	1665.2	112.5	1491.7	15.5	5.2	290.7	64.2	45.2	90.2
170SA27.2	22.3	57.7	1880.9	115.7	1506.0	15.8	5.3	311.2	67.5	49.2	98.2
170SFeed27.1	4.1	58.2	1769.8	522.4	1187.7	19.6	1.1	7853.6	290.4	1505.5	519.6
170SFeed27.2	5.1	58.6	1497.7	467.8	969.8	22.8	1.0	5482.0	189.4	1250.1	338.8
110CA51.1	89.7	58.6	1314.3	305.3	948.9	4.2	56.5	204.3	10.6	8.5	692.4
110CA51.2	101.8	58.2	1493.9	350.9	1083.3	4.4	55.4	206.7	10.6	9.0	687.2
110CFeed51.1	19.6	58.6	2378.8	260.7	2058.0	8.5	33.5	797.2	104.9	67.7	1802.5
110CFeed51.2	20.4	58.5	2211.5	226.4	1923.1	8.2	30.2	670.5	94.1	55.2	1218.9
150SA24.1	68.5	58.1	1954.4	110.9	1783.9	11.6	11.4	97.1	26.0	11.3	66.4
150SA24.2	75.5	57.7	2017.4	115.7	1842.5	11.7	12.6	91.9	24.4	10.8	89.5
150SFeed24.1	5.0	58.4	1482.2	502.8	919.5	21.9	1.0	6703.6	183.9	1318.7	365.6
150SFeed24.2	5.7	58.3	1639.8	484.4	1095.6	21.5	1.5	5081.1	191.5	1094.4	466.4
110CA60.1	88.9	58.0	1486.0	217.5	1218.0	5.4	56.3	221.6	20.7	11.9	748.7
110CA60.2	80.4	57.9	1563.4	202.9	1301.1	6.0	57.5	201.5	21.5	12.0	695.3
110CFeed60.1	22.8	58.3	1649.0	113.7	1475.5	10.9	46.8	298.9	64.7	32.7	839.4
110CFeed60.2	27.0	58.1	1932.8	128.3	1744.9	10.9	49.6	284.8	64.6	31.1	847.5
150SA27.1	25.3	58.1	1869.1	105.3	1503.2	14.6	7.8	252.6	59.5	36.8	118.4
150SA27.2	26.0	58.0	1671.2	105.3	1506.4	14.4	7.2	243.0	57.9	36.0	105.1
150SFeed27.1	4.2	57.5	2192.0	676.2	1466.8	21.7	1.4	9883.1	347.7	2099.8	795.9
150SFeed27.2	4.8	57.7	1701.2	496.0	1146.0	21.3	1.2	6187.1	238.3	1317.8	460.3

D.2 Bubble size data as a function of volumetric air flow rate

CIRCUIT: Measured	Feed Characterisation										
	db 1	Std Dev	db 2	Std Dev	db 3	Std Dev	db 4	Std Dev	db Ave	Std Dev	Std Dev
	(mm)	(mm)	(mm)	(mm)	(mm)	(mm)	(mm)	(mm)	(mm)	(mm)	(%)
Rougher A	0.883	0.284	0.947	0.280	0.883	0.262	0.859	0.260	0.893	0.038	4.2
Rougher B	0.984	0.286	1.154	0.381	1.097	0.364	0.951	0.262	1.046	0.095	9.1
Rougher C	1.179	0.443	1.289	0.506	1.201	0.448	1.210	0.443	1.220	0.048	3.9
Rougher D	1.283	0.491	1.246	0.442	1.181	0.421	1.192	0.430	1.225	0.048	3.9
Rougher E	1.184	0.409	1.202	0.427	1.157	0.389	1.188	0.412	1.183	0.019	1.6
Scavenger A	1.619	0.774	1.572	0.781	1.314	0.714	1.555	0.801	1.515	0.137	9.0
Scavenger B	1.184	0.528	1.197	0.559	1.086	0.490	1.091	0.468	1.139	0.059	5.2
Scavenger C	1.084	0.468	1.105	0.486	1.093	0.487	1.115	0.462	1.099	0.013	1.2
Scavenger D	1.017	0.421	1.000	0.406	0.980	0.378	0.986	0.397	0.996	0.017	1.7
Scavenger E	0.959	0.396	0.982	0.393	0.978	0.401	0.870	0.407	0.947	0.052	5.5
HGCA	0.639	0.282	0.561	0.243	0.534	0.250	0.616	0.277	0.587	0.048	8.2
HGCB	0.876	0.360	0.833	0.362	0.826	0.325	0.812	0.351	0.837	0.028	3.3
HGCC	0.592	0.266	0.590	0.298	0.712	0.292	0.691	0.260	0.646	0.064	10.0
LGCA	1.070	0.581	1.137	0.610	1.023	0.547	0.991	0.491	1.055	0.064	6.0
LGCB	0.704	0.333	0.561	0.278	0.716	0.272	0.533	0.225	0.629	0.095	15.1
LGCC	0.827	0.362	0.748	0.381	0.778	0.349	0.765	0.325	0.779	0.034	4.3

CIRCUIT: Measured	Configuration B										
	db 1	Std Dev	db 2	Std Dev	db 3	Std Dev	db 4	Std Dev	db Ave	Std Dev	Std Dev
	(mm)	(mm)	(mm)	(mm)	(mm)	(mm)	(mm)	(mm)	(mm)	(mm)	(%)
Rougher A	0.810	0.270	0.882	0.280	0.814	0.265	0.909	0.286	0.854	0.049	5.8
Rougher B	0.847	0.260	0.946	0.228	0.846	0.271	0.924	0.283	0.891	0.051	5.8
Rougher C	0.999	0.318	0.969	0.295	1.018	0.358			0.995	0.025	2.5
Rougher D	0.889	0.275	0.965	0.295	0.953	0.308	0.950	0.301	0.939	0.034	3.6
Rougher E	0.957	0.364	0.971	0.342	1.007	0.400	0.978	0.319	0.978	0.021	2.2
Scavenger A	1.043	0.355	1.010	0.337	1.024	0.333	1.035	0.348	1.028	0.014	1.4
Scavenger B	1.220	0.457	1.201	0.430	1.045	0.324	1.069	0.328	1.133	0.090	7.9
Scavenger C	1.010	0.268	1.075	0.299	1.043	0.277	1.054	0.276	1.045	0.027	2.6
Scavenger D	1.029	0.279	1.069	0.326	1.094	0.319	1.054	0.272	1.061	0.027	2.6
Scavenger E	1.136	0.322	1.105	0.305	1.146	0.344	1.150	0.341	1.134	0.021	1.8
HGCA							0.530	0.214	0.530	0.027	5.1
HGCB	0.854	0.279	0.783	0.285	0.759	0.299	0.724	0.301	0.780	0.055	7.0
HGCC	0.618	0.224	0.523	0.187	0.547	0.247	0.524	0.264	0.553	0.045	8.1
LGCA	0.484	0.184	0.633	0.243	0.593	0.256	0.693	0.202	0.601	0.088	14.7
LGCB	0.586	0.215	0.669	0.294	0.704	0.275	0.749	0.238	0.677	0.069	10.2

CIRCUIT: Measured	Configuration C		db 2	Std Dev	db 3	Std Dev	db 4	Std Dev	db Ave	Std Dev	Std Dev
	db 1	Std Dev									
	(mm)	(mm)	(mm)	(mm)	(mm)	(mm)	(mm)	(mm)	(mm)	(mm)	(%)
Cleaner D	0.780	0.340	0.833	0.223	0.714	0.291	0.827	0.230	0.788	0.055	7.0
Cleaner E	0.675	0.251	0.509	0.216	0.652	0.258			0.612	0.090	14.7
Cleaner F	0.709	0.273	0.605	0.270	0.551	0.248			0.625	0.076	12.2
CIRCUIT: Measured	Rougher Test 18			36.72	Vmin						
	db 1	Std Dev	db 2	Std Dev	db 3	Std Dev	db 4	Std Dev	db Ave	Std Dev	Std Dev
	(mm)	(mm)	(mm)	(mm)	(mm)	(mm)	(mm)	(mm)	(mm)	(mm)	(%)
Rougher A	1.367	0.362	1.087	0.301	1.216	0.323	1.147	0.305	1.204	0.121	10.0
Rougher B	1.465	0.455	1.100	0.396	1.195	0.314	1.224	0.310	1.246	0.155	12.5
Rougher C	1.520	0.452	1.449	0.416	1.467	0.432	1.355	0.389	1.448	0.069	4.7
Rougher D	1.500	0.444	1.423	0.428	1.392	0.382	1.421	0.395	1.434	0.046	3.2
Rougher E	1.298	0.298	1.244	0.288	1.203	0.263	1.044	0.428	1.197	0.109	9.1
CIRCUIT:	Rougher Test 23-27.5			46.92	-	56.1	Vmin				
Rougher A	0.883	0.284	0.947	0.280	0.883	0.262	0.859	0.260	0.893	0.038	4.2
Rougher B	0.984	0.286	1.154	0.381	1.097	0.364	0.951	0.262	1.046	0.095	9.1
Rougher C	1.179	0.443	1.289	0.506	1.201	0.448	1.210	0.443	1.220	0.048	3.9
Rougher D	1.283	0.491	1.246	0.442	1.181	0.421	1.192	0.430	1.225	0.048	3.9
Rougher E	1.184	0.409	1.202	0.427	1.157	0.389	1.188	0.412	1.183	0.019	1.6
CIRCUIT:	Rougher Test 30			61.2							
Rougher A	1.403	0.547	1.373	0.488	1.248	0.499	1.440	0.536	1.366	0.083	6.1
Rougher B	1.355	0.451	1.405	0.468	1.502	0.517	1.448	0.489	1.428	0.063	4.4
Rougher C	1.654	0.592	1.651	0.581	1.750	0.729	1.783	0.761	1.709	0.067	3.9
Rougher D	1.651	0.599	1.612	0.579	1.605	0.566	1.645	0.613	1.628	0.023	1.4
Rougher E	1.414	0.457	1.383	0.468	1.401	0.454	1.414	0.471	1.403	0.015	1.1
CIRCUIT:	Rougher Test 35			71.4							
Rougher A	1.605	0.663	1.494	0.658	1.357	0.656	1.344	0.600	1.450	0.124	8.5
Rougher B	1.488	0.536	1.463	0.612	1.477	0.541	1.445	0.516	1.468	0.019	1.3
Rougher C	1.763	0.669	1.738	0.689	1.781	0.774	1.714	0.636	1.749	0.029	1.7
Rougher D	1.822	0.634	1.671	0.588	1.534	0.663	1.675	0.701	1.676	0.118	7.0
Rougher E	1.543	0.574	1.542	0.577	1.400	0.592			1.495	0.082	5.5
CIRCUIT:	Rougher Test 40			81.6							
Rougher A	1.439	0.617	1.464	0.697	1.344	0.691			1.416	0.063	4.5
Rougher B											
Rougher C											
Rougher D											
Rougher E	1.661	0.668	1.704	0.646	1.576	0.652	1.140	0.484	1.647	0.065	4.0
CIRCUIT:	Cleaner Test 30			16.2							
HGCA	0.639	0.282	0.561	0.243	0.534	0.250	0.616	0.277	0.587	0.048	8.2
HGCB	0.876	0.360	0.833	0.362	0.826	0.325	0.812	0.351	0.837	0.028	3.3
HGCC	0.464	0.234	0.418	0.268					0.441	0.033	7.4
CIRCUIT:	Cleaner Test 40			21.6							
HGCA	0.785	0.332	0.777	0.346	0.748	0.332	0.792	0.313	0.775	0.019	2.5
HGCB	0.960	0.394	0.922	0.397	0.981	0.379	0.936	0.363	0.950	0.026	2.7
HGCC	0.592	0.266	0.590	0.298	0.712	0.292	0.691	0.260	0.646	0.064	10.0
CIRCUIT:	Cleaner Test 50			27							
HGCA	0.804	0.381	0.982	0.444	0.937	0.408	0.942	0.427	0.916	0.077	8.4
HGCB	1.014	0.483	0.948	0.413	0.954	0.423	0.904	0.430	0.955	0.045	4.8
HGCC	0.771	0.261	0.680	0.231	0.812	0.266			0.754	0.067	8.9
CIRCUIT:	Cleaner Test 60			32.4							
HGCA	0.854	0.345	0.881	0.370	0.799	0.343			0.845	0.042	4.9
HGCB											
HGCC	0.781	0.304	0.721	0.252	0.753	0.265	0.754	0.277	0.751	0.030	4.0

Experimental PGM Assay			(g/t)								
Time	Fresh Feed	Scav Tails	HGCA A	HGCC	LGCT	LGCC	HGCT	Scav Conc	Rougher Conc	Scav Tails 2	Fresh Feed 2
(min)	1	1									
1	96.50	4.21	138.00	60.50	11.90	76.80	24.90	11.90	75.80	7.13	219.00
3	37.70	6.14	71.00	35.60	9.67	36.20	18.20	16.90	29.00	5.77	66.00
5	22.40	5.26	51.00	29.40	9.65	35.00	14.60	14.90	25.30	4.79	24.80
7	21.50	3.05	64.70	23.40	9.39	23.00	13.10	14.80	20.40	4.53	18.50
10	19.00	4.13	52.90	22.20	9.10	20.60	11.10	13.70	18.30	3.66	16.90
15	11.00	2.98	54.30	16.20	9.25	17.30	9.24	12.00	18.50	0.33	10.00
Tails	1.28	0.72	14.30	11.20	3.89	9.54	4.16	5.15	6.64	0.66	1.18
Total Dry Masses			(g)								
Time	Fresh Feed	Scav Tails	HGCA A	HGCC	LGCT	LGCC	HGCT	Scav Conc	Rougher Conc	Scav Tails 2	Fresh Feed 2
(min)	1	1									
1	13.60	7.50	362.70	60.20	6.00	28.30	17.60	9.60	60.40	3.60	5.90
3	15.50	5.60	286.80	62.40	8.30	22.70	23.70	9.70	37.50	6.70	6.20
5	9.40	7.30	39.40	41.40	5.80	11.50	14.30	6.00	25.10	3.90	6.10
7	4.50	3.80	42.80	27.00	5.30	8.00	9.50	7.80	8.90	5.40	3.50
10	11.00	6.70	60.50	25.90	4.80	9.70	19.00	8.80	6.90	5.60	8.10
15	7.00	6.20	49.30	9.80	8.30	3.60	11.60	7.80	16.30	7.20	10.40
Tails	1038.20	891.50	160.80	58.00	107.50	100.00	179.80	86.90	101.90	931.90	534.90
Total	1099.20	928.60	1002.30	284.70	145.80	183.80	275.50	136.60	257.00	964.30	575.10

[illegible]

APPENDIX E – Raw Pilot Plant Data – LONMIN UG2

E.1 Raw mass balance data

1200kg/hr (Closed Circuit Secondary Mill)

Mass Balance 1 25/03/99											
CIRCUIT Measured	Duration	Mass Bag	Vet Mass+Bag+Cl	Dry Mass	Water Mass	Froth Height	Air Flow	PGM	Dry Mass Flow	Water Flow	PGM Flow
	(sec)	(g)	(g)	(g)	(g)	(cm)	(l/min)	(g/l)	(g/min)	(ml/sec)	(mg/hr)
Dry Feed	8.1	57.5	2644.8	2585.8	0.0			3.9	19225.3	0.0	4533.3
S/O		57.9	3397.8	2707.7	630.7			1.9			
1 Mill Discharge	10.9	58.2	5200.7	3902.8	1339.2			3.8	20866.3	122.3	4692.7
SAU	3.5	57.3	3427.6	1092.1	2276.7			3.9	18562.6	645.0	4321.4
F Feed	7.0	58.1	2069.5	594.4	1415.5			4.6	5102.1	202.5	1417.4
RA	150.8	57.8	1714.4	239.9	1415.2	4.9	33.7	179.0	95.4	9.4	1024.9
RB	150.8	58.4	2174.7	208.4	1906.4	5.0	35.1	86.4	82.9	12.6	429.7
RC	150.8	56.7	2134.0	114.1	1961.7	6.2	40.2	30.2	45.4	13.0	82.3
RD	150.8	57.1	2115.7	125.1	1932.0	6.5	40.6	21.4	49.8	12.8	63.9
RE	150.8	57.3	1800.0	85.0	1456.2	6.4	44.7	15.7	33.8	9.7	31.8
RT	17.3	57.6	3141.7	1118.8	1963.8			0.3	4794.9	145.0	80.6
SA	194.2	57.1	878.2	50.2	769.4	8.4	30.8	19.7	15.5	4.0	18.3
SB	194.2	57.4	1265.7	64.7	1142.1	8.6	30.6	13.8	20.0	5.9	16.6
SC	194.2	57.9	2345.8	89.0	2197.4	8.7	36.7	8.1	27.5	11.3	13.4
SD	194.2	57.7	1125.6	42.9	1023.5	9.0	31.6	8.7	13.3	5.3	7.0
SE	194.2	57.3	1280.5	49.4	1172.3	9.1	36.8	4.8	15.3	6.0	4.4
ST	16.7	57.1	4369.9	1928.8	2382.5			0.3	4703.3	112.5	79.0
C Feed	2.3	58.1	3735.2	1817.2	1868.4			2.4	46996.6	801.0	6711.1
C/O	3.9	57.6	3496.4	1103.0	2334.3			4.3	17189.6	606.3	4414.3
CAU	5.7	57.7	3943.5	3019.3	865.0			1.8	32063.4	153.1	3482.1
2 Mill Discharge	6.6	57.9	4510.5	3454.5	996.6			1.5	31262.4	150.3	2738.6

Mass Balance 2 25/03/99											
CIRCUIT Measured	Duration	Mass Bag	Vet Mass+Bag+Cl	Dry Mass	Water Mass	Froth Height	Air Flow	PGM	Dry Mass Flow	Water Flow	PGM Flow
	(sec)	(g)	(g)	(g)	(g)	(cm)	(l/min)	(g/l)	(g/min)	(ml/sec)	(mg/hr)
Dry Feed	6.6	58.8	2217.7	2157.4	0.0			3.8	19702.3	0.0	4515.8
S/O		59.1	4945.1	4003.9	880.6			1.7			
1 Mill Discharge	14.4	59.2	6466.5	4199.4	2226.4			3.8	17521.8	153.4	4016.0
SAU	3.9	59.3	3675.1	1185.3	2429.0			3.7	18329.4	626.0	4058.1
F Feed	8.8	58.6	2578.5	736.7	1781.7			3.6	5005.9	201.8	1069.3
RA	131.2	57.9	1447.2	193.3	1194.5	5.1	34.7	167.0	88.4	9.1	866.0
RB	131.2	58.5	1763.1	116.6	1586.5	5.4	34.5	87.1	53.3	12.1	278.8
RC	131.2	57.7	2151.9	144.8	1947.9	5.9	40.6	27.3	66.2	14.9	108.5
RD	131.2	57.4	2021.9	123.4	1839.6	6.3	38.4	14.6	56.5	14.0	49.5
RE	131.2	57.5	1708.5	90.3	1559.2	6.4	44.7	15.7	41.3	11.9	38.9
RT	9.4	57.9	2436.6	821.5	1555.7			0.3	4700.1	139.8	79.0
SA	174.0	58.3	1002.9	40.8	902.3	7.7	32.6	15.2	14.1	5.2	12.8
SB	174.0	57.7	814.6	32.9	722.5	7.9	28.8	13.2	11.3	4.2	9.0
SC	174.0	57.3	1704.5	78.2	1567.5	7.9	35.7	7.4	27.0	9.0	12.0
SD	174.0	58.2	1031.7	37.1	934.9	8.6	32.6	7.0	12.8	5.4	5.4
SE	174.0	58.2	1866.9	75.8	1731.4	9.0	45.1	4.9	26.1	10.0	7.7
ST	22.0	57.5	3622.5	576.6	3186.9			0.3	4608.8	106.1	77.4
C Feed	3.2	59.1	4808.6	2333.8	2414.2			2.4	43895.9	756.8	6215.7
C/O	4.1	58.6	4166.4	1351.5	2754.8			3.5	19826.4	673.5	4139.8
CAU	6.9	59.2	4931.0	3769.3	1101.0			1.5	33015.8	180.7	3030.8
2 Mill Discharge	7.5	59.0	4893.8	3703.2	1130.1			1.6	29507.6	150.1	2744.2

Mass Balance 3 25/03/99											
CIRCUIT Measured	Duration	Mass Bag	Vet Mass+Bag+Cl	Dry Mass	Water Mass	Froth Height	Air Flow	PGM	Dry Mass Flow	Water Flow	PGM Flow
	(sec)	(g)	(g)	(g)	(g)	(cm)	(l/min)	(g/l)	(g/min)	(ml/sec)	(mg/hr)
Dry Feed	8.0	57.2	2647.6	2632.4	-43.5			5.0	19842.2	0.0	5916.9
S/O		58.0	4059.1	3178.4	820.2			1.8			
1 Mill Discharge	10.3	58.5	4581.5	3199.3	1422.2			3.9	18654.8	138.2	4342.8
SAU	3.7	57.9	3280.3	1013.0	2207.9			5.1	16338.7	593.5	4989.8
F Feed	7.9	59.0	2365.1	864.6	1640.0			4.9	5022.2	206.5	1473.5
RA	138.3	43.0	1365.3	211.8	1109.0	5.1	34.3	179.0	91.9	8.0	986.7
RB	138.3	43.0	1554.7	116.6	1393.6	5.7	34.9		50.6	10.1	0.0
RC	138.3	43.0	1681.9	98.1	1539.3	5.7	39.6	31.3	42.6	11.1	79.9
RD	138.3	43.0	1640.0	89.7	1505.9	5.8	39.2	23.4	38.9	10.9	54.6
RE	138.3	43.0	1277.5	67.3	1165.7	6.0	44.5	18.4	29.2	8.4	32.2
RT	10.9	59.5	2593.6	944.6	1588.0			0.4	4769.1	158.0	111.6
SA	168.2	43.0	939.3	46.8	848.0	8.4	34.7	15.9	16.7	5.0	15.9
SB	168.2	44.0	683.3	34.7	603.1	8.5	30.4	13.6	12.4	3.6	10.1
SC	168.2	43.0	1252.6	79.5	1128.6	8.8	36.7	7.6	28.4	6.7	12.9
SD	168.2	44.0	832.4	38.3	748.6	8.9	34.5	7.9	13.7	4.5	6.5
SE	168.2	43.0	1690.0	68.9	1576.7	8.9	46.7	6.1	24.6	9.4	8.9
ST	29.7	43.0	2607.5	1018.7	1544.3			0.3	4673.4	128.9	78.5
C Feed	3.7	58.0	5846.1	2895.0	2891.6			3.4	46568.4	775.2	9388.2
C/O	5.1	58.5	5148.8	1452.4	3636.4			4.3	17188.2	717.2	4393.3
CAU	6.2	57.2	4145.7	3071.2	1015.8			1.6	29862.9	165.2	2912.4
2 Mill Discharge	7.0	58.9	4314.9	3281.2	973.3			1.0	28245.6	139.6	1745.6

900kg/hr (Closed Circuit Secondary Mill)

CIRCUIT Measured	Mass Balance 1 24/03/99		Vel Mass+Bag+Cl	Dry Mass	Water Mass	Froth Height	Air Flow	PGM	Dry Mass Flow	Water Flow	PGM Flow
	Duration	Mass Bag									
	(sec)	(g)	(g)	(g)	(g)	(cm)	(l/min)	(g)	(g/min)	(ml/sec)	(mg/hr)
Dry Feed	15.1	42.0	3745.6	3691.3	10.8			3.9	14677.1	0.7	3460.9
S/O		43.0	2985.4	2357.4	583.5			2.0			
1 Mill Discharge	18.7	42.0	5972.8	4050.2	1879.1			3.8	12995.3	100.5	2970.7
S/U	8.8	42.0	5129.3	2085.1	3000.7			4.1	14265.2	342.2	3526.4
F Feed	10.4	42.0	3500.4	1383.0	2073.9			4.1	7963.5	199.0	1939.9
RA	144.2	42.0	2030.4	224.6	1762.3	6.5	33.7	173.0	93.5	12.2	970.2
RAT		42.0	3117.3	1253.3	1820.5			1.1			
RB	144.2	42.0	2485.3	181.1	2260.7	6.7	36.1	83.7	75.4	15.7	378.4
RC	144.2	42.0	2225.0	100.3	2081.2	6.8	41.2	32.6	41.7	14.4	81.6
RCT		42.0	3225.8	1458.4	1713.9			0.6			
RD	144.2	42.0	1960.0	65.4	1831.1	6.9	42.8	24.0	35.5	12.7	51.2
RE	144.2	42.0	4673.3	202.8	4427.0	7.1	56.5	9.7	84.4	30.7	49.1
RT	18.5	42.0	4867.2	2212.9	2410.8			0.3	7192.5	130.6	120.8
SA	152.1	42.0	3112.1	133.3	2936.3	7.4	46.3	5.9	52.6	19.3	18.5
SAT		42.0	3470.1	1939.5	1487.1			0.8			
SB	152.1	42.0	2720.9	106.2	2571.2	7.4	45.3	8.7	41.9	16.9	21.9
SC	152.1	42.0	1291.9	52.3	1196.1	8.5	44.9	6.1	20.6	7.9	7.6
SCT		42.0	3503.5	1784.5	1675.5			0.9			
SD	152.1	42.0	1401.0	56.3	1301.2	8.7	44.9		22.2	8.6	0.0
SE	152.1	42.0	980.9	54.7	982.7	8.8	53.0	4.0	21.6	5.8	5.2
ST	86.6	42.0	4860.9	2680.5	2156.9			0.4	7033.6	72.2	177.2
C Feed	3.9	42.0	4698.3	2209.8	2445.0			1.6	33737.4	622.1	3157.8
C/O	6.5	42.0	3489.7	1041.8	2404.4			2.4	9801.8	369.3	1405.7
C/U	10.9	42.0	6490.7	4365.7	2081.5			1.2	24009.3	190.8	1728.7
2 Mill Discharge	7.8	42.0	3647.5	1995.8	1618.2			1.3	15236.3	206.9	1206.7

CIRCUIT Measured	Mass Balance 2 24/03/99		Vel Mass+Bag+Cl	Dry Mass	Water Mass	Froth Height	Air Flow	PGM	Dry Mass Flow	Water Flow	PGM Flow
	Duration	Mass Bag									
	(sec)	(g)	(g)	(g)	(g)	(cm)	(l/min)	(g)	(g/min)	(ml/sec)	(mg/hr)
Dry Feed	10.7	42.0	2558.8	2515.3	0.0			3.0	14117.7	0.0	2524.2
S/O		42.0	2674.3	2076.5	554.3			2.0			
1 Mill Discharge	11.3	42.0	3840.4	1985.8	1811.1			3.6	10581.5	160.8	2292.0
S/U	8.4	42.0	4092.2	1525.8	2522.9			3.5	14282.1	393.6	2956.4
F Feed	12.2	42.0	2632.6	776.5	2012.6			6.8	7117.9	226.5	2882.7
RA	144.2	42.0	1833.5	219.6	1570.4	6.5	33.7	183.0	91.4	10.9	1003.6
RAT		42.0	3606.5	1346.0	2217.0			0.9			
RB	144.2	42.0	2258.5	184.5	2030.5	6.7	36.3	99.2	76.8	14.1	456.9
RC	144.2	42.0	1954.3	96.4	1814.4	6.8	40.8	36.2	40.1	12.6	91.9
RCT		42.0	3841.1	1500.5	2297.1			0.4			
RD	144.2	42.0	1850.1	86.6	1720.0	6.9	42.8	27.3	36.0	11.9	69.0
RE	144.2	42.0	4729.6	202.0	4484.1	7.1	56.7	11.1	84.0	31.1	56.0
RT	11.1	42.0	2806.5	1250.4	1612.6			0.7	6789.5	145.9	272.9
SA	198.9	42.0	2753.9	112.8	2597.6	7.4	45.9		34.0		0.0
SAT		42.0	3820.4	2367.2	1509.7			0.4			
SB	198.9	42.0	2490.2	95.6	2351.1	7.4	44.9	8.7	28.8	11.8	15.1
SC	198.9	42.0	1262.8	59.8	1159.5	8.5	44.5		18.0	5.8	0.0
SCT		42.0	3509.7	2027.1	1439.1			0.6			
SD	198.9	42.0	1492.2	66.5	1382.2	8.7	44.9	5.5	20.1	6.9	6.6
SE	198.9	42.0	1588.5	64.0	1481.0	8.8	52.8	5.2	19.3	7.4	6.0
ST	333.7	42.0	6301.3	2989.8	3268.0			0.4	6669.2	100.8	169.1
C Feed	3.7	42.0	4277.7	2015.7	2218.5			1.3	32864.7	602.9	2524.0
C/O	7.3	42.0	6018.7	1775.0	4200.2			2.3	14629.1	577.0	1983.7
C/U	7.1	42.0	4032.3	2979.2	1009.6			0.9	25247.5	142.6	1333.1
2 Mill Discharge	6.9	42.0		2584.4				1.1	22408.1		1436.6

CIRCUIT Measured	Mass Balance 3 24/03/99		Vel Mass+Bag+Cl	Dry Mass	Water Mass	Froth Height	Air Flow	PGM	Dry Mass Flow	Water Flow	PGM Flow
	Duration	Mass Bag									
	(sec)	(g)	(g)	(g)	(g)	(cm)	(l/min)	(g)	(g/min)	(ml/sec)	(mg/hr)
Dry Feed	9.8	42.0	2417.4	2351.7	22.2			3.4	14442.4	0.0	2937.6
S/O		42.0	3222.5	2534.9	644.1			2.3			
1 Mill Discharge	12.1	43.0	4192.4	2874.4	1273.5			3.8	14241.5	105.2	3230.0
S/U	5.5	42.0	3463.4	1207.0	2212.9				13215.3	403.8	0.0
F Feed	12.3	43.0	3272.8	963.3	2266.0			5.6	7199.3	222.5	2427.6
RA	138.6	43.0	2356.8	235.3	2077.0	5.3	36.7	186.0	101.9	15.0	1136.9
RAT		43.0	3381.2	1291.2	2045.5			0.6			
RB	138.6	43.0	1732.1	155.2	1532.4	5.7	36.3	99.1	67.2	11.1	399.5
RC	138.6	43.0	1530.6	84.1	1402.0	6.4	41.8	36.6	36.4	10.1	80.0
RCT		42.0	3696.7	1526.3	2126.9			0.5			
RD	138.6	43.0	1403.7	76.3	1282.9	6.5	42.8	28.9	33.0	9.3	57.3
RE	138.6	43.0	3952.3	181.7	3726.1	6.6	57.1	11.8	78.7	26.9	55.7
RT	10.9	43.0	2929.2	1249.1	1635.6			0.3	6882.1	150.2	115.6
SA	187.9	43.0	1492.4	75.5	1372.4	7.9	45.3	14.2	24.1	7.3	20.5
SAT		42.0	4442.7	3321.5	1077.7			0.3			
SB	187.9	43.0	1543.3	74.0	1424.8	7.9	44.9	10.7	23.6	7.6	15.2
SC	187.9	43.0	1136.4	59.0	1032.9	8.4	44.7	9.2	18.8	5.5	10.4
SCT		42.0	3660.6	1817.1	1700.0			0.4			
SD	187.9	43.0	1316.3	63.2	1208.6	8.7	44.9	8.0	20.2	6.4	9.7
SE	187.9	43.0	1862.9	96.7	1719.7	8.4	52.6	3.9	31.5	9.2	7.4
ST	12.9	42.0	4054.6	2075.6	1935.5			0.3	6763.8	114.2	113.6
C Feed	3.8	42.0	4436.8	2064.2	2329.1			1.5	32337.7	608.1	2929.8
C/O	5.6	43.0	5195.1	1701.9	3448.7			2.1	16202.1	614.7	2249.8
C/U	5.7	42.0	3383.1	2345.5	994.1			0.8	24517.4	173.2	1118.0
2 Mill Discharge	8.0	43.0	4823.4	3380.9	1398.0			0.9	25293.5	174.3	1320.3

E.2 Batch floatability data

Batch Floatability Tests						Batch Floatability Tests					
Experimental PGM Assay (g/t)						Fine Grind					
Time	Fresh Feed	Rougher Conc	Rougher Tails	Scav Conc	Scav Tails	Time	Fresh Feed	Rougher Conc	Rougher Tails	Scav Conc	Scav Tails
(min)	1	1	1	1	1	(min)	2	2	2	2	2
1	148.00	117.00	9.97	21.10	16.20	1	137.00	188.00	2.08	114.00	2.75
3	35.50	75.80	5.89	12.40	5.11	3	31.70	60.80	3.13	35.70	1.94
5	19.20	44.40	4.74	8.41	0.75	5	14.40	38.60	2.40	22.80	2.04
7	13.80	34.40	1.93	5.93	0.84	7	9.38	30.40	0.91	22.70	2.39
11	12.10	27.30	6.97	3.11	0.73	11	6.35	24.60	1.47	19.30	1.83
15	8.60	19.80	4.52	5.82	0.61	15	1.06	18.50	2.66	15.90	1.75
Tails	0.58	7.29	0.40	3.09	0.55	Tails	0.39	5.08	0.08	4.36	0.17
Total Dry Masses (g)											
Time	Fresh Feed	Rougher Conc	Rougher Tails	Scav Conc	Scav Tails	Time	Fresh Feed	Rougher Conc	Rougher Tails	Scav Conc	Scav Tails
(min)						(min)					
1	32.90	27.90	7.40	9.00	5.70	1	30.20	86.50	14.80	17.20	9.20
3	19.40	18.80	8.50	8.70	7.00	3	20.20	35.90	13.00	14.90	10.00
5	13.20	12.40	7.30	7.40	7.00	5	15.00	19.50	10.40	13.10	9.80
7	11.40	10.10	7.30	7.40	6.30	7	13.40	13.60	9.60	10.30	8.70
11	12.60	11.30	8.50	8.30	7.20	11	13.00	16.80	11.20	12.30	10.20
15	18.40	15.20	10.00	10.00	8.20	15	15.80	17.50	14.00	14.90	11.70
Tails	870.10	54.00	1394.50	58.70	1782.60	Tails	857.40	60.60	1055.50	131.50	1386.20
Total	978.00	149.70	1443.50	109.50	1824.00	Total	965.00	250.40	1128.50	214.20	1445.80
PGM											
(g)											

E.3 Recovery versus size data for both coarse and fine streams**Coarse Grind**

Dry Mass						
Measured	Dry Mass Overall	Fractions +106um	Fractions -106um	Fractions -38um	Fractions -25um	Fractions -11um
F Feed	7426.9	0.44	0.35	0.03	0.08	0.09
RA	95.6	0.16	0.34	0.04	0.19	0.26
RB	73.1	0.16	0.30	0.04	0.19	0.31
RC	39.4	0.12	0.18	0.04	0.21	0.45
RD	34.9	0.13	0.16	0.04	0.20	0.46
RE	82.4	0.07	0.09	0.03	0.17	0.46
RT	6954.7	0.56	0.26	0.04	0.09	0.06
SA	36.9	0.07	0.05	0.03	0.23	0.62
SB	31.5	0.06	0.05	0.03	0.21	0.65
SC	19.2	0.12	0.04	0.02	0.22	0.60
SD	20.8	0.14	0.05	0.02	0.18	0.60
SE	24.1	0.39	0.03	0.02	0.14	0.42
ST	6822.2	0.63	0.21	0.03	0.07	0.06
PGM Grade						
	Grades Overall	Grades +106um	Grades -106um	Grades -38um	Grades -25um	Grades -11um
F Feed	5.5	0.6	5.9	10.0	13.1	12.3
RA	180.7	107.0	186.0	239.0	239.0	110.0
RB	94.0	72.1	94.3	125.0	127.0	72.2
RC	35.8	41.0	39.6	39.9	35.0	28.4
RD	26.7	35.9	29.2	27.2	26.1	23.3
RE	10.9	22.7	18.2	13.9	10.7	7.7
RT	0.4	0.7	0.4	0.4	0.5	0.8
SA	10.0	17.0	39.8	19.2	9.1	5.6
SB	9.4	15.6	33.1	13.0	7.0	5.6
SC	7.7	13.0	32.8	8.3	5.7	5.4
SD	6.7	8.7	26.0	13.6	5.4	4.2
SE	4.4	4.5	26.8	7.7	4.2	4.1
ST	0.4	0.5	0.3	0.3	0.3	0.3

Fine Grind

Dry Mass						
Measured	Dry Mass Overall	Fractions +106um	Fractions -106um	Fractions -38um	Fractions -25um	Fractions -11um
F Feed	5379.8	0.10	0.52	0.05	0.19	0.15
RA	91.9	0.09	0.40	0.04	0.20	0.27
RB	62.3	0.10	0.28	0.06	0.21	0.36
RC	51.4	0.11	0.21	0.05	0.19	0.44
RD	48.4	0.08	0.16	0.04	0.20	0.51
RE	34.8	0.15	0.29	0.04	0.14	0.39
RT	4784.0	0.14	0.51	0.04	0.18	0.13
SA	15.4	0.20	0.11	0.03	0.16	0.51
SB	14.6	0.23	0.11	0.02	0.16	0.48
SC	27.6	0.16	0.08	0.03	0.17	0.56
SD	13.2	0.21	0.08	0.02	0.14	0.55
SE	22.0	0.25	0.08	0.03	0.14	0.51
ST	4691.1	0.16	0.47	0.05	0.18	0.13

PGM Grade						
	Grades Overall	Grades +106um	Grades -106um	Grades -38um	Grades -25um	Grades -11um
F Feed	4.4	2.0	2.9	6.6	8.4	7.5
RA	175.0	52.5	145.0	271.0	266.0	128.0
RB	86.8	48.8	81.7	107.0	117.0	71.2
RC	29.6	33.5	30.0	32.8	34.9	25.1
RD	19.8	24.3	21.2	19.9	21.8	20.5
RE	16.6	19.9	19.1	13.3	15.6	14.7
RT	0.3	0.8	0.3	0.3	0.3	0.5
SA	16.9	14.3	35.3	26.3	20.3	10.1
SB	13.5	12.9	27.5	23.7	14.4	8.7
SC	7.7	6.8	23.0	11.1	9.0	4.8
SD	7.9	6.4	21.4	6.9	7.7	5.4
SE	5.3	5.0	18.9	7.3	6.6	2.6
ST	0.3	0.6	0.3	0.3	0.3	0.3

APPENDIX F – Raw QEMSEM data

(ECD = Equivalent circle diameter, GE = PGM's on grain edges, GB = PGM's on grain boundaries, Lib = Liberated PGM's, Lock = PGM's locked in particles, feld = feldspar, px = pyroxene, chr = chromite, pent = pentlandite, chalc = chalcopyrite, mill = millerite, py = pyrite, po = pyrrhotite)

-10 μm

PGM -TYPE	MODE OF OCCURRENCE	HOST	PGM ECD (μm)	BMS ECD (μm)	GANGUE ECD (μm)
Pt.Rh.Co.Cu.S	GE	feld	3.43		7.02
Pd.Hg	GE	px	2.17		1.77
Pd.Sb	GE	px	1.53		24.70
Pt.Fe	GE	px	1.77		4.77
Pt.Fe	GE	chr	1.25		2.50
Pt.S	GE	chr	2.80		2.17
Pt.Fe	GE	chalc	1.98	1.77	
Pd.Pb	GE	pent	4.24	5.01	
Pd.Sb	GE	pent	0.89	1.77	
Pt.Fe	GE	pent	2.34	2.34	
Pt.Fe	GE	pent	2.50	4.34	
Pt.Pd.S	GE	pent	1.53	8.89	
Pt.Pd.S	GE	pent	2.94	3.75	
Pt.Pd.S	GE	pent	3.43	2.50	
Pt.S	GE	pent	6.74	7.02	
Pt.Pd.S	GE	po	2.17	3.07	
Pd.Pb	Lib		1.98		
Pd.Pb	Lib		3.31		
Pd.Pb	Lib		4.85		
Pd.S	Lib		2.17		
Pd.S	Lib		3.19		
Pd.S	Lib		4.15		
Pd.Sb	Lib		2.50		
Pd.Sb	Lib		4.85		
Pd.Pb.Te	Lib		1.98		
Pt	Lib		3.31		
Pt.Fe	Lib		1.25		
Pt.Fe	Lib		2.66		
Pt.Fe	Lib		3.19		
Pt.Fe	Lib		3.31		
Pt.Pd.S	Lib		0.89		
Pt.Pd.S	Lib		1.77		
Pt.Pd.S	Lib		1.77		
Pt.Pd.S	Lib		2.94		
Pt.Pd.S	Lib		3.07		
Pt.Rh.Co.Cu.S	Lib		1.98		
Pt.S	Lib		0.89		
Pt.S	Lib		1.25		
Pt.S	Lib		1.25		
Pt.S	Lib		1.53		
Pt.S	Lib		1.77		
Pt.S	Lib		2.17		
Pt.S	Lib		2.34		

Pt.S	Lib		2.50		
Pt.S	Lib		3.07		
Pt.S	Lib		3.54		
Pt.S	Lib		4.15		
Pt.S	Lib		5.16		
Pt.Pd.Ni.S	Lib		2.17		
Pt.Pd.Ni.S	Lib		2.34		
Pt.Pd.Ni.S	Lib		3.65		
Pt.Pd.Ni.S	Lib		4.93		
Pt.Rh.Ru.As.S	Lib		1.53		
Pt.Rh.Ru.As.S	Lib		2.17		
Pt.Te	Lib		2.50		
Pt.Te	Lib		3.31		
Pt.Te	Lib		3.75		
Pt.S	Lock	chalc	3.07	14.08	
Au.Ag	Lock	pent	0.89	3.31	
Pd.S	Lock	pent	1.53	7.24	
Pt.Fe	Lock	pent	1.53	2.34	
Pt.Fe	Lock	pent	3.43	17.02	
Pt.Pd.S	Lock	pent	1.98	3.07	
Pt.S	Lock	pent	3.43	3.07	
Pt.S	Lock	pent	3.86	19.23	
Pt.Te	Lock	laurite	0.89	4.15	
Pt.Pd.S	Lock	px	3.19		7.40
Pt.S	Lock	px	1.98		18.24
Pt.Fe	Lock	biotite	1.53		26.81
Pt.Pd.S	Lock	px	3.65	9.53	13.04
Pd.Bi.Te	Lock	talc	2.94		2.66

+10-25 μ m

PGM -TYPE	MODE OF OCCURRENCE	HOST	PGM ECD (μ m)	BMS ECD (μ m)	GANGUE ECD (μ m)
Pt.S	GB	pent/chr/sil	1.77	2.50	21.80
Pt.Te	GB	pent/laurite	2.18	4.97	0.00
Pt.Fe	GE	feld	1.99		3.45
Pt.S	GE	feld	4.46		23.16
Pt.S	GE	feld	3.89		12.36
Pt.S	GE	feld	2.36		3.89
Pt.Te	GE	feld	3.34		13.82
Pt.S	GE	px	1.54		3.57
Pt.Fe	GE	px	3.89		6.55
Pd.Pb	GE	chr	2.68		19.34
Pt.Rh.Co.Cu.S	GE	chr	0.89		15.58
Pt.S	GE	chr	1.99		27.00
Pt.Fe	GE	magnetite	5.43		4.28
Pd.Hg	GE	rutile	1.78	1.54	17.41
PtFerm	GE	chalc	6.79	6.18	
Pd.Pb	GE	pent	2.82	9.40	4.09
Pt.Fe	GE	pent	3.22	3.57	
Pt.Fe	GE	pent	3.22	22.60	
Pt.Pd.S	GE	pent	1.98	5.01	
Pt.Pd.S	GE	pent	2.17	4.85	
Pt.Pd.S	GE	pent	2.18	8.18	
Pt.Pd.S	GE	pent	3.54	10.77	
Pt.S	GE	pent	1.78	14.82	15.06

Pd.Bi.Te	GE	po	1.77	13.42	
Au	Lib		6.55		
Pd.Hg	Lib		13.57		
Pt.As	Lib		5.53		
Pt.Bi.Te	Lib		1.99		
Pt.Fe	Lib		11.77		
Pt.S	Lib		1.25		
Pt.S	Lib		1.78		
Pt.S	Lib		5.60		
Pt.S	Lib		19.81		
Pt.Cu	Lib		2.52		
Pt.Pd.Ni.S	Lib		3.22		
Pt.Rh.As.S	Lib		9.37		
Pt.Fe	Lock	chalc	3.09	2.52	
Pt.Fe	Lock	chalc	7.88	13.44	
Pt.S	Lock	chalc	6.05	5.43	
Pt.Te	Lock	chalc	2.96	12.36	6.68
Pd.S	Lock	pent	5.80	13.68	6.56
Pd.S	Lock	pent	2.34	15.30	
Pd.Sb	Lock	pent	2.34	7.02	
Pt	Lock	pent	2.50	14.10	
Pt.Fe	Lock	pent	6.79	10.67	
Pt.Pd.S	Lock	pent	2.36	3.45	
Pt.Pd.S	Lock	pent	3.65	10.62	
Pt.Rh.Co.Cu.S	Lock	pent	3.99	14.30	8.03
Pt.Rh.Co.Cu.S	Lock	pent	2.94	10.55	
Pt.Rh.Co.Cu.S	Lock	pent	2.96	11.46	
Pt.S	Lock	pent	1.77	7.51	23.94
Pt.S	Lock	pent	1.78	9.93	19.40
Pt.S	Lock	pent	2.50	15.05	14.41
Pt.S	Lock	pent	0.89	22.05	9.49
Pt.S	Lock	pent	5.85	26.07	7.19
Pt.S	Lock	pent	1.78	14.44	2.68
Pt.S	Lock	pent	1.53	12.67	
Pt.S	Lock	pent	1.99	14.77	
Pt.S	Lock	pent	2.94	10.95	
Pt.S	Lock	pent	4.63	9.40	
Pt.Cu.S	Lock	pent	2.94	18.67	
Pt.Pd.Ni.S	Lock	pent	3.09	19.05	2.96
Pt.Pd.Ni.S	Lock	pent	4.77	10.05	
Pt.FE	Lock	py	4.51	13.45	
Pt.Rh.Co.Cu.S	Lock	py	1.54	3.34	
Pt.Pd.S	Lock	laurite	1.26	27.57	
Pd.Te	Lock	laurite/pent	1.54	14.05	5.28
Pt.Pd.S	Lock	feld	1.26		24.18
Pt.Rh.Co.Cu.S	Lock	px	2.36		11.00
Pt.S	Lock	px	2.68		15.19
Pt.S	Lock	px	2.36		10.05
Pt.S	Lock	px	1.78	2.36	10.13
Pt.S	Lock	px	0.89	4.24	4.24
Pt.As	Lock	px	17.66		12.52
Pt.Fe	Lock	talc	6.91		19.52

+25-38 μ m

PGM -TYPE	MODE OF OCCURRENCE	HOST	PGM ECD (μ m)	BMS ECD (μ m)	GANGUE ECD (μ m)
Pd.Pb	GB	pent/feld	2.35	8.09	22.46
Pd.Pb	GB	pent/px-rim	1.99	29.80	10.44
Pd.Pb	GB	pent/px	1.26	4.95	25.90
Pt.S	GB	laurite/feld	4.26	7.75	20.24
Pt.Rh.Co.Cu.S	GB	laurite/px	2.95	2.51	17.99
Pt.Rh.Co.Cu.S	GE	px	2.51		8.14
Pt.Pd.S	GE	px	1.54		23.22
Pt.Sb.As	GE	talca	4.35		24.72
Pt.Pd.S	GE	chalca	8.19	13.36	
Pd.Pb	GE	mill	1.99	20.55	17.77
Au.Ag	GE	py	1.54	25.38	8.24
Pt.Pd.S	Lock	chalca	2.18	23.49	7.75
Pt.Te	Lock	chalca	1.54	17.57	
Pd.Pb	Lock	pent	4.26	30.64	
Pt.Fe	Lock	pent	2.81	3.55	21.40
Pt.Fe	Lock	pent	1.26	4.53	13.85
Pt.Fe	Lock	pent	3.44	14.19	31.90
Pt.Fe	Lock	pent	2.35	25.43	
Pt.Pd.S	Lock	pent	1.26	1.54	1.99
Pt.Rh.Co.Cu.S	Lock	pent	2.51	6.59	16.69
Pt.Rh.Co.Cu.S	Lock	pent	4.01	53.70	
Pt.Rh.Co.Cu.S	Lock	pent	4.01	53.71	
Pt.Rh.Co.Cu.S	Lock	pent	4.01	53.72	
Pt.S	Lock	pent	0.89	26.30	14.57
Pt.S	Lock	pent	2.35	22.11	
Au.Ag	Lock	py	1.99	32.79	
Pt	Lock	chr	1.54		51.93
Pt.Pd.S	Lock	feld	1.26		32.10
Pt.S	Lock	feld	1.99	10.95	21.65
Pt.S	Lock	px	2.81	14.97	12.21
Pd.As	Lock	px	2.35	6.71	12.69
Pd.Hg	Lock	feld	1.54		32.62
Pt.S	Lock	pent	1.26	2.51	29.72

+38-53 μ m

PGM -TYPE	MODE OF OCCURRENCE	HOST	PGM ECD (μ m)	BMS ECD (μ m)	GANGUE ECD (μ m)
Pd.Pb	GB	chalca/talca	1.78	27.51	21.23
Pd.Pb	GB	pent/px-rim	3.66	21.42	9.81
Pd.Pb	GB	pent/px-rim	3.66	21.42	9.81
Pd.Pb	GB	pent/px-rim	3.66	21.42	9.81
Pt.Pd.S	GE	feld	2.67		31.46
Pt.Pd.S	GE	px/chr	2.35		39.21
Pt.Hg	GE	talca	1.78	4.44	37.46
Pd.Pb	GE	chalca	2.51	25.78	
Pd.Pb	GE	chalca	6.03	30.38	
Pt.Pd.S	GE	chalca	4.78	34.73	
Pd.Sb	GE	pent	1.99	5.10	16.24
Pt.Pd.S	GE	pent	7.22	33.67	21.21
Pt.S	GE	py/px	1.54	32.78	11.17
Pd.Hg	Lib		1.78		

Ir(Rh).As.S	Lock	chalc	3.20	20.40	13.96
Pt.Fe	Lock	chalc	3.32	6.16	22.95
Pt.Fe	Lock	pent	1.62	48.44	
Pt.Fe	Lock	pent	1.62	48.44	
Pt.Fe	Lock	pent	1.62	48.44	
Pt.Rh.Co.Cu.S	Lock	pent	1.78	24.15	13.88
Pt.S	Lock	pent	1.54	15.41	4.26
Pt.S	Lock	feld	3.87	5.40	40.73
Pt.Fe.Cu	Lock	feld	1.54		24.43
Pt.S	Lock	feld	1.54		24.43
Pt.Fe	Lock	px	2.18		35.77
Pt.Pd.S	Lock	px	3.20		35.67
Pt.Pd.Ir.As.S	Lock	px	3.44		20.86
Pd.Pb	Lock	talc/chr	2.35		35.82
Pt.As	Lock	px	0.89	6.88	53.46

+53-75 μ m

PGM -TYPE	MODE OF OCCURRENCE	HOST	PGM ECD (μ m)	BMS ECD (μ m)	GANGUE ECD (μ m)
Pd.Pb.Te	GB	chalc/feld	2.81	9.10	56.19
Pd.Sb	GB	pent/px	2.67	11.34	63.26
Pd.Sb.As.S	GE	chr	1.78		47.82
Pt.S	GE	pent	1.78	28.07	43.04
Pt.Pd.Ni.S	GE	pent	9.36	11.62	
Pt.S	Lock	chalc	0.89	2.35	51.36
Pd.S	Lock	pent	1.99	7.43	54.02
Pd.Pt.Sn	Lock	pent	1.54	10.81	40.60
Pd.Te	Lock	pent	1.99	14.76	40.91
Pt.Fe	Lock	pent	2.18	18.70	30.30
Pt.Rh.Co.Cu.S	Lock	pent	0.89	45.50	55.21
Pt.S	Lock	pent	1.78	3.32	63.39
Pt.S	Lock	pent	1.78	12.21	48.85
Pt.S	Lock	pent	9.53	20.47	76.05
Pt.S	Lock	pent	3.44	52.87	
Pd.Hg	Lock	feld	1.99		60.46
Pt.Rh.Co.Cu.S	Lock	px/chr	2.67		86.69
Pt.Rh.Co.Cu.S	Lock	px	1.78	9.15	48.63
Pt.Rh.Co.Cu.S	Lock	px	1.26	13.82	60.90
Pd.As.S	Lock	px	10.51	20.28	39.26
(Pt.Pd).(Te.Bi)	Lock	px	1.99		66.69
Pt.S	Lock	feld	3.55	25.24	69.45
Pt.Fe	Lock	px	11.41	20.49	39.44
Pt.Pd.Pb	Lock	talc	3.97		66.29
Pt.Te	Lock	talc	1.54	3.87	45.19

+75-106 μ m

PGM -TYPE	MODE OF OCCURRENCE	HOST	PGM ECD (μ m)	BMS ECD (μ m)	GANGUE ECD (μ m)
Pt.Pd.Cu	GB	chr/px	1.78		91.43
Pt.S	GB	pent/feld	0.89	5.96	89.64
Pt.Pd.Ni.S	GB	pent/feld	0.89	1.54	83.52
Pt.Fe	GE	feld	6.16		56.89
Pd.Pb	GE	px	5.69		13.94
Pd.Hg	GE	chr	1.54		102.17

Pd.Pb	GE	chr	3.66		47.77
Au.Ag	GE	pent	1.78	14.55	50.71
Pd.Pb	Lock	chalc	2.35	60.29	29.01
Pd.Bi.As	Lock	chalc	1.54	7.27	72.82
Pt.Pd.S	Lock	chalc	4.35	40.80	72.67
Pt.Fe	Lock	pent	0.89	6.59	108.43
Pt.Fe	Lock	pent	3.20	2.67	47.57
Pt.Pd.S	Lock	pent	1.26	5.83	66.22
Pt.Pd.S	Lock	pent	3.08	9.02	
Pt.Rh.Co.Cu.S	Lock	pent	1.78	11.17	78.50
Pt.S	Lock	pent	1.54	20.68	91.57
Pt.S	Lock	pent	1.99	15.13	62.47
Pt.S	Lock	pyrrh	8.52	98.01	0.00
Pt.S	Lock	feld	6.47	28.07	70340
Pt.Pd.Ni.S	Lock	feld	2.81	4.62	71.67
Pt.Cu.S	Lock	px	1.78	6.77	62.93
Pt.S	Lock	px	1.78	5.18	67.79
Pt.Pd.Ni.S	Lock	px	3.66	8.71	79.88
Pd	Lock	px	1.78		117.68

+106-150 μ m

PGM -TYPE	MODE OF OCCURRENCE	HOST	PGM ECD (μ m)	BMS ECD (μ m)	GANGUE ECD (μ m)
Pt.S	GB	pent/talc	1.99	8.29	99.97
Pt.S	GE	feld	3.87	2.81	69.68
Au.Ag	GE	chr	2.67		86.41
Pt.Rh.As	GE	chr	3.20		72.81
Pt.	Lock	chalc	6.09	6.71	50.94
Pt.Fe	Lock	chalc	1.54	2.51	153.22
Pd.Pb.As	Lock	laurite	2.95	6.35	96.29
Pd.Pb	Lock	pent	1.54	8.29	131.59
Pd.Pb	Lock	pent	2.18	15.69	120.66
Pt.Fe	Lock	pent	16.43	18.68	78.51
Pt.Rh.Co.Cu.S	Lock	pent	5.96	23.41	11.20
Pt.S	Lock	pent	1.78	5.55	79.51
Pt.S	Lock	pent	1.78	9.28	68.44
Pt.S	Lock	pent	5.96	17.02	127.94
Pt.Pd.Ni.S	Lock	pent	1.99	19.91	70.43
Pt.Pd.Ni.S	Lock	py	1.99	8.14	136.80
Pd.Pb	Lock	magnetite	1.26		82.79
Pd.Bi.Te	Lock	feld	3.77	8.00	72.27
Pt.Fe	Lock	feld	1.78	8.19	105.25
Pt.Rh.Co.Cu.S	Lock	feld	1.26	9.97	101.72
Pt.Pd.S	Lock	px	3.08	2.67	81.28
Pt.Fe	Lock	px	3.20		126.43
Pt.Fe	Lock	chalc	5.96	24.96	272.62
Pt.S	Lock	talc/chr	1.78		110.54
Pt.As.S	Lock	talc/px	1.78	2.95	120.25
Pt.S	Lock	feld	0.89	6.82	144.55

+150-212 μm

PGM -TYPE	MODE OF OCCURRENCE	HOST	PGM ECD (μm)	BMS ECD (μm)	GANGUE ECD (μm)
Pt.Te	GB	chalc/px	1.26	15.92	125.91
Pt.Pd.S	GE	chr	2.51		156.92
Pd.As.S	Lock	pent	2.51	17.46	39.65
Pt.S	Lock	pent	1.26	15.18	156.08
Pt	Lock	chr	2.67		155.57
Pt.Pd.Ni.S	Lock	feld	3.55	18.45	117.88
Pt.S	Lock	feld/chr	4.44		145.63
Au.Ag	Lock	px	2.35		15.47
Pd.Pt.Sb.As	Lock	px	2.95		175.73
Pt.Pd.S	Lock	biotite	1.54		135.36
Pt.Fe	Lock	talc	1.78		168.48
Pd.As.S	Lock	px	3.66		20.20
Pt.S	Lock	px	1.78	14.81	90.29
Pt.S	Lock	talc	2.67		143.65

+212 μm

PGM -TYPE	MODE OF OCCURRENCE	HOST	PGM ECD (μm)	BMS ECD (μm)	GANGUE ECD (μm)
Pd.Te	GB	px/chr	2.35		163.47
Pt.S	GB	px/chr	1.78	8.29	267.63
Pt.Pd.S	GE	px	1.54		186.79
Rh.Ir.As.S	Lock	chalc	2.18	22.76	251.89
Pt	Lock	pent	2.35	19.99	220.83
Pt.S	Lock	pent	1.26	6.47	215.06
Pt.Pd.Te	Lock	pent	1.99	22.95	226.83
Pt.Pd.Sb.As	Lock	chr	1.99		281.00
Pd.Pb	Lock	px/chr	2.81		242.46
Ru.Rh.S	Lock	px	5.89	8.33	224.57
Au.Ag	Lock	px	1.26		200.89
Pt.S	Lock	px/chr	4.70	8.71	258.77