THE CHARACTERISATION OF A TWO STAGE FLOTATION CIRCUIT TREATING A COMPLEX PLATINUM AND CHROMITE ORE

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by

Francis de Beer

B Eng (Chemical Engineering) (Potchefstroom)

Department of Chemical Engineering

University of Cape Town

Rondebosch

7700

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SYNOPSIS

A methodology was developed to characterise the complex MF2 circuit at Impala Platinum Ltd. The circuit was divided into individual units and each unit was evaluated over a period of time. The performance of each unit in terms of platinum, palladium, rhodium and chromite recoveries was determined by sampling around the unit. The changes in ore floatability characteristics were monitored by carrying out batch flotation tests on samples taken from the feed to each unit. The mixing in the units was characterised by performing residence time distribution studies on the units.

The batch flotation results were modelled using a distributed rate approach. The model yielded parameters which were associated with three floatability classes, viz. a fast floating, a medium floating and, a slow floating class and a class for valuable minerals which did not float. The model was not able to discriminate between mineralogical and particle size effects on flotation rates. This would be an important prospect of a future investigation.

The rate constants associated with each floatability fraction obtained from the batch flotation together with residence time distribution results were used in association with the actual recoveries from each unit. These factors were used to simulate the performance of the circuit. The same model was used for all the units and the differences in cell design as indicated by hydrodynamics, aeration rate, etc. were accounted for in the mixing parameters and flotation rates.

This procedure to simulate plant performance was successfully applied to three different circuits thus confirming its validity. The difference in the three simulated circuits was in the cleaning stages. Among other things, the simulation shows that the best platinum, palladium and rhodium recoveries could be obtained if the greatest cleaning capacity was situated in the secondary stage of the circuit.

Although the changes in the circuit configuration could be simulated to establish the effect on the platinum, palladium and rhodium recoveries, the chromite recovery could not be simulated. This is due to the fact that chromite is mainly floated by entrainment which is not incorporated into the model. Since the chromite recovery is crucial for the design of

optimum cleaning circuits in the flotation of UG-2 ore, research needs to be done to quantify the entrainment phenomenon.

ACKNOWLEDGEMENTS

TABLE OF CONTENTS

CHAPTER 1

INTRODUCTION

CHAPTER 2

THE CIRCUIT

CHAPTER 3

EXPERIMENTAL DETAILS

CHAPTER 4

RESULTS

CHAPTER 5

MODELLING

CHAPTER 6

DISCUSSION

CHAPTER 7

CONCLUSIONS

LIST OF TABLES

LIST OF FIGURES

NOMENCLATURE

a   Precious Mineral Grade

C   Concentration

C   Sampling Constant

C_θ Dimensionless Concentration

D   Axial Dispersion Coefficient

d   Particle diameter

d_b Bubble Diameter

d_c Column Diameter

d_l Liberation size

E   Residence Time Distribution

E_k Particle Collection Efficiency

E_θ Residence Time Distribution in terms of Dimensionless Time

f   Shape factor

f   Scale-up Factor for Flotation Rate Constants

F   Number of Feed Streams

F   Froth Stability Factor

F_c Mass Flow Rate of Concentrate Stream

F_f Mass Flow Rate of Feed Stream

g   Particle Size Range Factor

H_c Column Collection Zone Height

H_f Column Froth Zone Height

J   Number of Simple Junctions

J_g Superficial Gas Velocity

k   Flotation Rate Constant

k_f Flotation Rate Constant for Fast Floating Class

k_i Flotation Rate Constant for Particle Size i

k_m Flotation Rate Constant for Medium Floating Class

k_s Flotation Rate Constant for Slow Floating Class

l   Liberation Factor

L   Length of a Vessel

M   Mass of Representative Sample

M_f	Fraction of Fast Floating Minerals
M_i	Fraction Associated with Floatability Class i
M_m	Fraction of Medium Floating Minerals
M_o	Fraction of Valuable Minerals not Floatable
M_s	Fraction of Slow Floating Minerals
n	Number of Floatability Classes
N	Number of Tanks in Series
N_d	Vessel Dispersion Number
P_a	Probability of Particle-Bubble Adhesion
P_c	Probability of Particle-Bubble Collision
P_e	Probability of a Particle being retained on the Gas Bubble and lifted through the Liquid
P_f	Probability of a Particle being retained in the Froth
P_x	Probability of Success
Q	Volumetric Flow Rate
R	Recovery
$R_{0,i}$	Maximum Theoretical Recovery for Particle Size i
R_∞	Maximum Theoretical Recovery
s	Statistical Error committed by Sampling
S	Number of Simple Separators
t	Flotation Time
$\bar{t}$	Nominal Residence Time
u	Interstitial Velocity of Fluid
V	Cell Volume
W_C	Work of Cohesion of Liquid
W_d	Work of Adhesion from Dispersion Forces
W_h	Work of Adhesion from Hydrogen Bonding of Water to Solid Surface Groups
W_i	Work of Adhesion from Interaction with Ionic Sites
W_{SL}	Work of Adhesion of Liquid to Solid
x	Average Particle Size
δ	Dirac Delta Function
Θ	Correction Factor for fixing Time Zero in Batch Flotation Experiments
θ	Static Contact Angle at the Liquid/Solid Interface
θ	Dimensionless Time in Residence Time Distributions

σ^2	Variance of Residence Time Distribution Curve
σ_{LG}	Surface Tension between Liquid and Gas Phases
σ_{SG}	Surface Tension between Solid and Gas Phases
σ_{SL}	Surface Tension between Solid and Liquid Phases
ρ	Relative Density of Precious Mineral
τ	Mean Residence Time
ϕ	Fruitfulness Factor

Chapter One

Introduction

The platinum-group metals (PGMs) are a group of six chemically similar elements, namely platinum (Pt), palladium (Pd), rhodium (Rh), ruthenium (Ru), iridium (Ir) and osmium (Os). These elements, together with gold and silver, comprise the precious metals.

The PGMs have only received recognition as separate and unique metals in the last 200 years, although platinum was used by the Egyptians around 800 BC (Buchanan, 1988). McDonald and Hunt (1982) reviewed the history of scientific experimentation and investigation of platinum. Platinum was first formally described in 1750 in a letter by Dr William Brownrigg in which he provided an account of preliminary experiments carried out by himself and Charles Wood. In 1804 William Hyde Wollaston recognised that the native platinum did not consist of just one element, but also of palladium, rhodium, iridium and osmium. Once these had been identified it was possible to separate the different metals and obtain high purity platinum. Ruthenium was only discovered some 40 years later.

Sir Humphry Davy, designer of the miner's lamp, informed the Royal Society in 1817 that platinum was an ideal catalyst, being capable of promoting chemical reaction without undergoing any changes itself. Michael Faraday, in 1830, found that platinum was ideal for use in the making of glass, as it does not contaminate the final product.

The history of the refining of platinum from a laboratory scale operation to an industrial process can be associated with the career of George Matthey (1825 - 1913). Matthey started his career as an apprentice in 1838 by joining PN Johnson. He later became the chairman of Johnson Matthey and Co. Ltd, a name which has become synonymous with the PGM industry.

PGMs have excellent catalytic qualities and are resistant to chemical corrosion. They have a high melting point and are chemically inert over a wide range of temperatures. This makes them ideal for a number of specialised industries.

The automobile industry is the largest consumer of PGMs. According to a report released by the Minerals Bureau (Barry and Odendaal, 1993), in 1992, autocatalytic converters accounted for 34, 10 and 87 % of the total demand for platinum, palladium and rhodium, respectively. The report also states that 40% of the 1992 demand for platinum was consumed by the jewellery industry; of this amount, 85% was consumed by Japan.

Other important uses for PGMs are laboratory equipment, catalysts for petroleum and fertiliser industries, thermocouples, electric contacts, anti-cancer drugs, dental alloys, fuel cells and in high temperature vessels for the manufacture of glass.

The platinum mines of South Africa and the Stillwater Mine in the USA are the only major primary producers of platinum in the world (Barry and Odendaal, 1993). Other large producers of PGMs recover the metals as a by-product of base metal production.

In South Africa the main output is derived from the well known Merensky Reef, with the remainder being mined from the UG-2 chromium seam and the Platreef occurring near Potgietersrus.

1.1 Merensky and UG-2 ores

1.1.1 Geology

The largest reserves of PGMs in the world are situated in the North West region of South Africa and are part of the well known Bushveld complex (Figure 1.1). The mines in the area have exploited the Merensky Reef for the past sixty years or so. The Merensky Reef is estimated to contain some 17 million kg of PGMs (Corrans et al., 1982). Below the Merensky reef lies a series of chromium seams. One of these, the Upper Ground 2 Reef (UG-2) contains reserves of at least 31 million kg of PGMs.

The Merensky Reef is a pyroxentic layer and lies between footwalls and hanging walls of spotted and mottled anorthosite. It ranges in thickness from between 0.6 to 15 metres. At the narrow parts the noble-metal values occur throughout and where it is thicker, they are mostly found close to the hanging wall (Liddell et al., 1986). Two chromitite stringers,
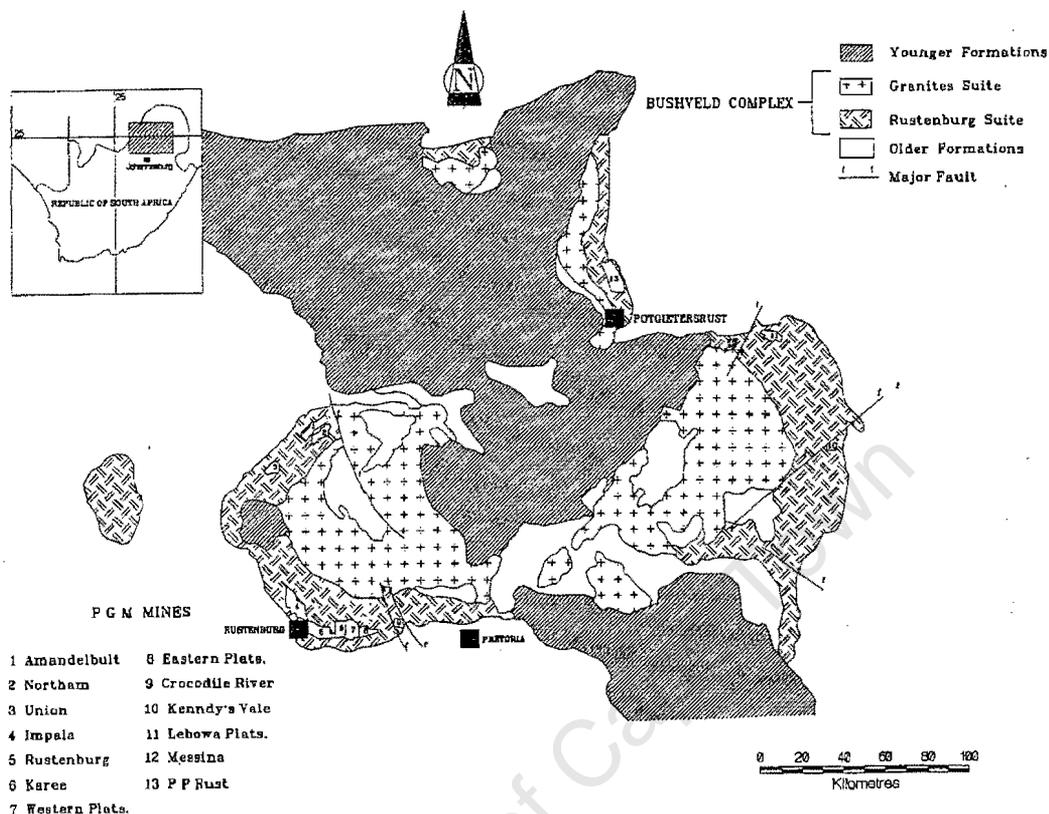
Figure 1.1: Geological map of the Bushveld Complex showing the locations of the PGM producers (Barry and Odendaal, 1993).

which vary in thickness from a few millimetres to approximately 60 millimetres, are present in the Reef. The sulphide mineralisation tends to be associated with these stringers. The upper stringer usually contains economic noble-metal mineralisation, and mining operations therefore include this chromitite (Liddell *et al.*, 1986). Secondary talc is widespread, occurring especially at the grain boundaries between sulphides and pyroxenes.

The UG-2 seam is comparable with the Merensky Reef but lies approximately 150 metres below it in the western Bushveld and 330 metres below it in the east. The UG-2 seam is between 15 and 255 cm thick (Corrans *et al.*, 1982).

1.1.2 Mineralogy

1.1.2.1 Merensky Reef

Mineralogical studies done on the Merensky Reef by J.C.I (Vermaak and Hendriks, 1976) revealed that the major silicate minerals are orthopyroxene (50 to 70%), plagioclase (11 to 25%) and clinopyroxene (2 to 27%). The main oxide minerals present were reported to be chromite, magnetite, ilmenite and rutile.

The base metal sulphide (BMS) content of the Reef was found to be 2.9% by volume. Wagner (1929) reported a BMS content of between 2 and 3 volume percent. The main sulphide minerals present in the Merensky Reef are pyrrhotite (45%), pentlandite (32%), chalcopyrite (16%), pyrite (2 to 4%) and minor amounts of cubanite and mackinawite (Vermaak and Hendriks, 1976). According to Liddell *et al.* (1986), the sulphides occur mostly in the interstices of the silicate grains, while sulphide inclusions in silicates and oxides are also present to some extent.

The content or grade of the PGMs present in the Merensky Reef varies from 3 to 11 g/t (Liddell *et al.*, 1986). The PGM mineralogy in the Reef varies with location and is complex. The PGM minerals are mostly associated with the base-metal sulphides. The minerals occur as occlusions, intergrowths and in solid solutions in the sulphides. In the Merensky Reef, PGM values are absent where no sulphides occur, but the converse is not true (Liddell *et al.*, 1986). A mineralogical study by Impala Platinum (Louwrens and McLaren, 1990) identified the 18 PGM minerals listed in Table 1.1. According to Liddell *et al.* (1986) the majority of the PGM minerals in the Reef are associated with pentlandite, occurring either in the pentlandite grains or, more often, at the grain boundary between the pentlandite and the gangue. Vermaak and Hendriks (1976) reported that the average grain size of the precious minerals is 236 ± 45μm.

1.1.2.2 UG-2 Reef

The UG-2 Reef consists mainly of chromite (60 to 90%), gangue silicates, orthopyroxene (5 to 25%), plagioclase (5 to 15%) and base-metal sulphides (1%). The BMS occur mainly

interstitially to the chromite grains (Corrans *et al.*, 1982) or are enclosed in silicate or chromite grains (McLaren, 1985).

Table 1.1: PGM minerals present in the Merensky 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-Fe Alloy	Pt-Fe
Moncheite	PtTe$_2$
Merenskyite	PdTe$_2$
Os-Ir alloy	
Potarite	PdHg
Genkinite	Pd-Pt-Sb
Vysotskite	PdS
Pt-Pd alloys	
Plumbopalladinite	Pd$_3$Pb$_2$
Majakite	PdNiAs
Stibiopallandinite	Pd-Sb
Pt-Ag-Se Alloy	
Pt-Rh sulphide	

The main sulphides observed in UG-2 ore are pentlandite, chalcopyrite and pyrrhotite. Corrans *et al.* (1982) also report cobalt-pentlandite and millerite as main sulphides observed in the UG-2 samples from Western Platinum Pty (Ltd) mines. McLaren (1985) evaluated samples from the Impala mines and does not mention the observations of Corrans and co-workers but mentions the presence of pyrite. Louwrens (1992), who mainly analysed for

BMS, confirmed that pentlandite, pyrrhotite, chalcopyrite and pyrite are the main BMS, and also mentioned traces of millerite, galena, spharelite, barite, digenite and bornite.

According to Liddell *et al.* (1986) only 70 to 80% of the PGM minerals in UG-2 ore are associated with the BMS, 10 to 20% are disseminated in the silicate gangue and 5% in the chromite. PGM minerals may also occur as single grains and aggregates along the boundaries of two or more chromite grains in contact with gangue particles. The PGM minerals listed in Table 1.1 were also found in UG-2 ore by Louwrens and McLaren (1990). The average grain sizes of PGM minerals have been reported to be smaller than $15\mu m$ (Liddell *et al.*, 1986; Corrans *et al.*, 1982; McLaren, 1985).

1.1.3 Mining

The mining costs for UG-2 ore are lower than for Merensky ore since UG-2 ore has a higher density than Merensky ore (Liddell *et al.*, 1986).

In South African platinum mines, four distinct underground methods are in use (Chadwick, 1980; Newman, 1973). The methods are known as the two scattered mining methods (mechanised and Bilby's methods), longwall mining method and the up-dip mining method.

1.1.3.1 Scattered mining methods

The scattered mining method is used where geological disturbances (e.g. potholes) prohibit the mining of extensive areas. From the original raise, which becomes the centre gully of the stope, a series of panels are advanced in the direction of the strike. Advance strike gullies (A.S.G.'s) form the top and bottom limit of each panel. A.S.G.'s are always kept slightly ahead of the advancing panels to establish a free face for blasting and for purposes of exploration, and to facilitate the removal of ore (Hochreiter *et al.*, 1985). The broken ore can be removed via a system of scrapers (hence the term mechanised) or by hand lashing (Bilby's method), with two persons per cutting. The ore is lashed into Bilby cars operating on monorail strike tracks along which the Bilby cars are pushed towards the centre gully.

1.1.3.2 Longwall mining method

The longwall mining method is used in the exploration of large, geologically undisturbed areas of reef (Hochreiter *et al.*, 1985). It requires numerous panels on dip, advancing together along the direction of strike. The ore is removed by means of face and A.S.G. scrapers.

1.1.3.3 Up-dip mining method

The up-dip mining method is so called because the panels are advanced in an up-dip direction. Very little cleaning of the blasted area is needed since the blasting is directed so that the blasted rock is concentrated in the dip gully. A dip gully scraper then scrapes the ore down to boxholes.

1.1.3.4 Run-of-mine ore

Liddell et al. (1986) reported that there is a difference in the noble-metal distribution in the run-of-mine (ROM) ore size fractions for the two Reefs. In Merensky ore the noble-metal values are evenly spread since the noble-metal distribution follows the mass distribution. UG-2 ROM however shows a finer particle size and in the fine sizes (smaller than 75μm) the noble-metal distribution is greater than the mass distribution. This indicates that the minerals containing the noble metals are liberated during mining, owing to the mineralisation and friability of the ore.

As discussed in section 1.1.2, the majority of the sulphides are contained in the interstices of the silicate and chromite grain. Blasting and scraping liberates these minerals. The data published by Liddell *et al.* (1986) shows that the UG-2 ROM ore does not only contain more fines than Merensky ROM ore, but that the fines are also enriched in noble-metals.

The floatability of the ROM ores from both Reefs decreases with particle size. As Merensky ROM ore has little fines this poses only problems for UG-2 ROM ore. Since the fines in UG-2 ROM ore are richer in PGM minerals than the bigger sizes, it will imply a decrease in the overall recovery of PGMs associated with UG-2 ore. It is obvious, as Liddell *et al.* (1986) suggest, that consideration be given to the use of underground blasting and handling

1.1.4 Extractive metallurgy of UG-2 and Merensky ore

The treatment process for PGMs can be divided into four different stages:

(1) comminution of the ore and concentration by physical techniques such as flotation,

(2) pyrometallurgical concentration,

(3) hydrometallurgical separation of the base metals and concentration of the PGM product and,

(4) final refining to produce the individual PGM.

The range of minerals present in Merensky and UG-2 ores, their relative densities, shape, discrete particle size, and association, all therefore present a challenge to the metallurgist in optimising the primary extraction process. This challenge is heightened by the fact that the deposits currently being exploited are not constant in mineralisation. Since Merensky ore has been treated for over 50 years the main challenge lies with UG-2 ore which has only been exploited since the beginning of the 1980's.

1.1.4.1 Comminution

The comminution of Merensky ore and UG-2 differ since UG-2 ore is more friable in nature and the PGMs are contained in smaller sulphide particles, and are much finer than in Merensky ore.

The PGM mines use crushing, rod milling, ball milling and autogenous milling in the comminution of Merensky ore (Hochreiter *et al.*, 1985). The grinding circuits vary in configuration. Earlier plants tend to use two stage grinding consisting of ball mills in an open circuit followed by closed-circuit ball mills. Newer plants tend to use single stage closed-circuit ball milling.

Until 1985, Impala Platinum Limited was the only known platinum venture not using conventional crushing: semi-autogenous mills alone were in use (Hochreiter *et al.*, 1985).

From a grinding point of view, the ore from the Merensky Reef is extremely hard. Good recovery requires the ore to be ground to a size range adequate to ensure the liberation of the valuable minerals. Corrans *et al.* (1982) report a grind of 60% passing -75 μm in use for Merensky ore in single stage grinding.

Extensive test work was done by the Council for Mineral Technology referred to as Mintek in the late seventies and early eighties in comminution, flotation and smelting of UG-2 ore. They recommended a 80% passing -75μm grind to enable liberation of the PGM minerals in the ore (Corrans *et al.*, 1982). Liddell *et al.* (1986) made a recommendation for regrinding of the coarse particles in the final tailings to enable an increase in the final recovery.

There are two types of comminution circuits practised in the processing of UG-2 ore today. Some mines use crushing and grinding 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 draw back 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 UG-2 plants tend to mill run-of-mine ore because of cost factors. According to Goodall (1995), three types of grinding practices are used in grinding run-of-mine ore. Fully autogenous (FAG) mills, semi-autogenous (SAG) mills with steel loads of 4 to 12% and run-of-mine (ROM) mills with steel loads similar to ball mills are in use.

A grind of 80% passing -75μm, necessary to liberate the PGM minerals, can be achieved by using one or more grinding stages. Goodall (1995) reports that single stage, two stage and even three stage grinding are in use in the UG-2 comminution plants. In more than one stage grinding, tailing streams from primary flotation circuits are usually reground.

1.1.4.2 Flotation

In order to recover PGM bearing minerals from Merensky and UG-2 ore flotation is the process that are used. In this process as applied in the case of Merensky and UG-2 ore bodies alloys of noble metals and other metals are not recovered. A detailed review of the circuits used in this process as well as the flotation chemistry will be presented in Section

1.3.8.3.

The concentrate obtained from the flotation process is fed to the smelter.

1.1.4.3 Smelting

The objective of smelting is to convert solids into liquids which then separate by density. The sulphide minerals form a matte, which is then treated further; the gangue, which has been fluxed with limestone, is discarded as slag. The flotation concentrate has to be dried to a moisture content of less than 5% before it is fed to the furnace. This can be achieved by filtering and partial drying. The more common practise is to agglomerate the concentrate by pelletizing or spray drying, which reduces dust losses and provides a porous charge through which reaction gases arising from decomposing of the sulphides can escape.

The Merensky concentrate is smelted in a rectangular electric furnace that has six Soderberg electrodes in line (Ireland, 1980; Mostert *et al.*, 1973; Newman, 1973). As the concentrate and added flux melt, two liquid phases form: a slag with a relative density of 2.7 and a matte with a relative density of 5.3. The flux is added to the charge to reduce the melting point of the slag and, hence, its viscosity (Liddell *et al.*, 1986).

The matte produced from the furnace is transferred in the molten state to a Pierce-Smith converter. The iron sulphide is oxidized to FeO, and the sulphur is removed as SO_2. The converter slag has a high PGM content and is usually returned to the smelter for the recovery of the PGMs.

The smelting of UG-2 is different to Merensky concentrates. The first difference lies in the metal oxide content of UG-2 concentrate which is higher in MgO, SiO and Al_2O_3 (Liddell *et al.*, 1986). This indicates that higher smelting temperatures and possibly larger addition of flux would be necessary for the production of a slag with a viscosity low enough for an acceptable separation of the matte and the slag to take place. The second important difference lies in the chromite content. The chromite content for UG-2 concentrate is almost ten times that of the Merensky concentrate. If this chromite is allowed to deposit in the furnace hearth it would build up and reduce the working volume of the furnace.

Mintek initiated intensive mineralogical studies to try and understand the build up of chromite during the smelting process (Corrans *et al.*, 1982). The following conclusions were made:

(1) The accretion normally found in the hearth of a furnace has an entirely different composition from that resulting from the concentrate, indicating that the chromium in the latter has crystallised from a matte containing chromium.

(2) The partitioning of chromium between the matte and the slag is strongly dependent on the partial pressure of oxygen and,

(3) under reducing conditions, the formation of chromite in the matte and at the interface between the matte and the slag is restrained.

(4) The temperature has a significant effect on the crystallisation of chromite crystals from both the matte and the slag.

(5) The chromite that dissolves from the matte is largely found in the pyrrhotite lattice.

Laboratory- and pilot-scale tests conducted by Mintek (Liddell *et al.*, 1986), revealed that UG-2 concentrate should be smelted in a circular electric furnace with three graphite electrodes, as practised at Western Platinum mine (Hochreiter *et al.*, 1985). A circular furnace will better withstand the operating condition of high power densities and higher temperatures. The mentioned operating conditions create more turbulence in the bath. The higher turbulence helps to dissolve the chromite in the matte and slag and therefore enables the removal of the chromite with tapping.

The UG-2 matte is blown in a manner similar to that used for Merensky matte. The matte is then treated for the removal of the base metals and the residues are send to the PGM refinery for the separation of the individual noble metals.

1.2 Flotation

Since the main emphasis of the research was on the characterisation of the flotation process, this section will review various aspects of flotation relative to this study.

In a flotation process, hydrophobic particles attach to gas bubbles which remove them from the aqueous environment while hydrophilic particles remain suspended in the system.

This phenomenon will be explained in more detail by discussing the following aspects that are most important for the understanding of the flotation process, from a scientific point of view.

i.   Thermodynamic conditions

ii.   Particle bubble attachment

iii.   Particle floatability

1.3.1 Thermodynamic conditions

The application of the principles of thermodynamics to flotation systems has contributed significantly to the understanding of some of the underlying concepts of the flotation process (Fuerstenau, 1976).

Thermodynamics predicts whether or not a change will tend to occur and says nothing about the rate at which the change takes place. Though thermodynamics cannot tell a flotation engineer what the mineral recovery will be at a given temperature or under given solution conditions, it may help to make some predictions as to how the flotation response may change with temperature, type of collector, type of mineral etc. The one criticism against the practical need for studying the thermodynamics of flotation is that thermodynamics is concerned mainly with equilibrium processes and, in the time span during which flotation takes place, the system may not be in equilibrium.

1.2.1.1 Thermodynamics of wetting

Bubble-particle contact is one of the key factors controlling the process of flotation. To enable the mineral particle to become floatable the air bubble must adhere to the mineral surface. At the stage where the bubble adheres to the surface of the particle a three phase contact of solids, air and liquid phases occurs.

The general thermodynamic condition for three phase contact is defined by Young's equation,

$$\sigma_{SG} = \sigma_{SL} + \sigma_{LG}\cos(\theta)$$

where σ_{SG}, σ_{SL} and σ_{LG} are the surface tensions of the solid/gas, solid/liquid and liquid/gas interfaces, respectively, and Θ is the contact angle.

The work required to separate liquid water from the solid surface and replace it by an
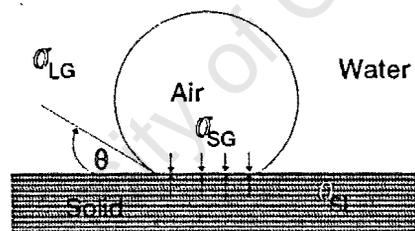
Figure 1.2: Contact angle between bubble and solid phase in a aqueous medium.

adsorbed water layer in equilibrium with saturated water vapour is equal to the work of adhesion, and is given by Dupre's equation (Dupre, 1867),

$$W_{SL} = \sigma_{SG} + \sigma_{LG} - \sigma_{SL}$$

The combination of Young's and Dupre's equations yields an expression for the work of adhesion,

$$W_{SL} = \sigma_{LG}(1 + \cos\theta)$$

Therefore, for air to displace water so that a finite contact angle is obtained with an air bubble,

$$W_{SL} < 2\sigma_{LG}$$

and the work of cohesion of liquid, W_C (Leja, 1982),

$$W_C = 2\sigma_{LG}$$

Thus,

$$W_{SL} < W_C$$

Thus a thermodynamic condition for flotation is that the work of adhesion of the water to the solid must be less than the work of cohesion for water.

The work of cohesion of water to the solid is made up of dispersion forces (W_d), hydrogen bonding of water to polar sites (W_h) and interaction with ionic sites (W_i). When most minerals are crushed, ionic or covalent bonds are broken, giving rise to highly reactive, charged sites which readily hydrate. The resulting surface has substantial $W_h + W_i$ components. For all solids W_d is less than W_C so that adhesion to an air bubble can only be obtained if the $W_h + W_i$ component are diminished appreciably.

In flotation this is achieved by adsorbing surfactants, referred to as collectors, at the solid-water interface to eliminate or shield polar sites or replace them with sites which do not ionise or take part in hydrogen bonding.

Flotation collectors are characterised by

i)   a polar group, usually ionic, which interacts with the mineral surface and gives the reagent solubility, and

ii)   a non-polar group, typically a hydrocarbon chain, of sufficient size to impart hydrophobicity to the mineral surface.

If the $W_h + W_i$ component are small, flotation will be possible without adding collector. That means that all solids would be hydrophobic if they did not carry polar or ionic groups

(Laskowski, 1974).

1.2.2 Particle bubble attachment

The capture of a particle by a bubble can be thought of as occurring in a series of stages. In the collision stage each particle that collides with a bubble stays in contact with the bubble surface for a certain time, called the contact time. When the particle and bubble move closer together the liquid between them forms a film and the next stage of film thinning starts. If the contact time is sufficiently long the film will thin to a point where rupture will occur spontaneously forming a stable three phase aggregate.

The movement of the particle around the bubble was studied by Jameson *et al.* (1977), Brown (1965) and Sutherland and Wark (1955). In the process of particle bubble collision, coalescence will almost certainly occur over the front of the bubble (Jameson *et al.*,1977), and particles adhering to the surface will then be swept to the rear by the relative motion between bubble and liquid. Evidence of the above was given by Sutherland and Wark (1955) when they photographed galena particles suspended from the rear of a bubble.

1.2.3 Particle floatability

A mineral particle is classified as floatable when it can successfully attach to an air bubble and be removed with it from the flotation slurry. For a particle to be floatable, not only must dewetting of the particle be thermodynamically favourable (i.e. the particle must be hydrophobic), but the following other criteria have also to be satisfied (Laskowski, 1974):

i) The particle must collide with a bubble;

ii) The disjoining film separating the particle and bubble must thin, rupture and recede within the contact time; and

iii) The particle-bubble aggregate formed must be of sufficient strength to withstand shearing forces in the flotation cell.

1.2.4 Flotation reagents

Flotation reagents ensure that the flotation process is highly selective and efficient. Flotation reagents vary very widely in composition; they include organic and inorganic compounds, acids and alkalis, salts of various compositions, substances both soluble and insoluble in water. Reagents are classified in three groups, according to the purpose that they serve nl., collectors, regulators or modifiers and frothers.

1.2.4.1 Collectors

Collectors are organic compounds which act selectively on the surfaces of certain mineral particles rendering them hydrophobic and thus aiding their attachment to air-bubbles.

Flotation collectors are characterised by

i)   a polar group, usually ionic, which interacts with the mineral surface and gives the reagent solubility, and

ii)   a non-polar group, typically a hydrocarbon chain, of sufficient size to impart hydrophobicity to the mineral surface.

Collectors may fall into three main groups:

i)   non-polar oils;

ii)   organic acids and their metal salts; and

iii)   organic bases and their acid salts.

These three classes are not mutually exclusive and a specific collector may contain some the characteristics of the other classes.

1.2.4.2 Frothers

Frothers are reagents which increase the dispersion of bubbles and the stability of the froth, and decrease the surface tension of the water. This organic substance concentrates at

liquid/gas interfaces, changing their surface energy and degree of hydration. Frothers ensure that the floated particles will not fall back into the flotation pulp before they can be removed.

1.2.4.3 Regulators or modifiers

Frequently, in flotation, the collector does not attach with adequate strength to the valuable mineral, or it attaches to the gangue minerals as well. Regulators are reagents that ensure that only the wanted particles float. Regulators can be divided into three types:

i)   activators, which enhance collector attachment to the valuable mineral;

ii)   depressants, which ensure that unwanted minerals such as gangue do not float by enhancing their hydrophilic character.

iii) pH modifiers which regulate the pH of the pulp so that the optimum conditions for collection, activation or depression, are achieved.

1.2.5 Flotation kinetics

Flotation kinetics is the study of the variation in amount of froth overflow product with flotation time, and the quantitative identification of all rate controlling variables (Arbiter and Harris, 1962). With such variables maintained constant, the algebraic relationship between the proportion of mineral floated and flotation time is a flotation rate equation. This contains the constant values of all the rate determining variables implicit in one or more rate constants which must be evaluated from experimental data.

The earliest contribution on flotation kinetics came in the mid-Thirties from Beloglozov (1939), Garcia-Zuniga (1935) and Gründer and Kadur (1940), with interest increasing in recent years. The broad goal of these studies is the reduction of the total variability of a flotation system to equations among individual variables, or among a lesser number of dimensionless groups (Johnstone and Thring, 1957). The attainment of this goal will yield many practical advantages. The quantitave understanding of the process should lead to improvements both in metallurgical results and flotation cell design.

1.2.5.1 Flotation rate

Schuhmann (1942) pointed out that the specific flotation rate has the same dimensional significance as the specific reaction rate for a first order chemical reaction, and that the two parameters are similarly related to the kinetics of the respective process they describe. Thus chemical kinetics is concerned with interactions between atoms, molecules, or ions, while flotation kinetics deals with interactions between bubbles and particles.

From the assumption that flotation is a first order reaction the rate equation is,

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

where C is the concentration of the valuable mineral and k the rate constant.

1.2.5.2 The optimum flotation time

The optimum flotation time is the time at which the flotation rates of the two components to be separated are equal (Agar, 1983). Obviously there are several restrictions that must be observed for this statement to hold true. Mechanical factors such as the particle size, pulp density, air flow and agitation must be constant as well as the chemical conditions. In short, all the factors that affect the rate constant must be fixed in order to focus attention on the flotation time.

1.2.6 The effect of variables on flotation kinetics

1.2.6.1 Effect of mixing

Mixing or the degree of turbulence in the flotation cell affects the rate of flotation. If good mixing occurs in the pulp the contact between the particle and the bubble will increase and therefore promote particle-bubble adhesion. However, if too much mixing is applied, detachment of the particles from the bubbles will occur. If the degree of mixing is too low, the particles will not have the required kinetic energy to be attached to the air bubbles. It is clear that the degree of mixing is very important in a flotation cell.

The degree of mixing in a cell will depend on feed rate and the hydrodynamics in a flotation cell, i.e. cell and impeller design and degree of agitation. Jameson *et al.* (1977) stated that an increase in the feed rate reduces the time required for flotation. This has been explained by the fact that, with an increase in the flow rate, a reduction of pulp impoverishment, caused by mixing, takes place. Mackenzie and Matheson (1963) found that the rate of flotation for a given particle size range increased with increased impeller speed. Studies (Kirchberg and Topfer, 1965) on the effect of turbulence, by varying impeller speed, showed that an optimum impeller speed exists. The conclusion was made that turbulence should be kept low to prevent detachment of coarse particles. Factors that raise the turbulence and lead to sharp deflexion of the flow, should be avoided. Li *et al.* (1993) found that recovery rises steadily with the increase in agitator speed. However, they stressed the fact that the intensity of agitation does not exert uniform influence on the recovery level of different size classes.

1.2.6.2 Effect of air flow rate

An increase in the air flow rate will result in an increase in the rate constant value (Dobby and Finch, 1990). Higher air flow rates into a flotation cell will increase the bubble size. Various authors (Brown, 1965; Reay and Ratcliff, 1973; Dobby and Finch, 1990) found that the rate of removal of fine particles will increase with a decrease in bubble size. Therefore, an optimum air flow rate is needed.

1.2.6.3 Effect of particle size

The decrease in flotation rate with particle size is an expected result and has been shown to be the case in a conventional batch flotation cell (Jameson *et al.*, 1977). It was concluded from a study performed on a pilot flotation column (Mills, 1992) that the effect of particle size on the rate constant had an optimum value. Dobby and Finch (1990) obtained the same result in the flotation of galena in a pilot column. Imaizumi and Inoue (1965) report that the optimum in conventional flotation is between 50 and 150μm.

The effect of particle size on collision efficiencies and induction time has been discussed by various authors (Jowett, 1980; Dobby and Finch, 1990; Mills, 1992). The collision efficiencies were determined using the experimental rate constants. Mills (1992) reports that

the collision efficiencies increased dramatically with increasing particle size while the attachment efficiencies decreased with increasing particle size. This is in agreement with the proposal of Dobby and Finch (1990) that the opposing changes in the collision and attachment efficiencies result in an optimum particle size for collection. Jowett (1980) found that induction time decreases with an increase in particle size. This trend was confirmed by Mills (1992).

1.2.7 Flotation machines

Froth flotation has been an important minerals separation process for well over 50 years. The design of flotation machines was often a matter of experience combined with judgement. The machines were not always selected on the basis of criteria developed from basic principles (Young, 1982). A wide range of flotation machines are now available for different applications. Through the years new machines had been developed, especially over the last two decades. The designs of the new machines are moving away from the conventional type design although the conventional machines are still the most widely used in the world of mineral processing.

1.2.7.1 Conventional Mechanical Flotation cells

The conventional mechanical cell is the most common type of flotation machine used in the industry (Young, 1982). The dominance of the mechanical machines is more a question of commercial realities than of design excellence: the major Western manufacturers make only this type, and most flotation engineers are familiar with no other (Young, 1982).

The conventional mechanical cell can be characterised by a mechanically driven impeller which agitates the pulp and disperses air into it. It is normal to have a number of impellers in series in one vessel. This cell 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 is now the most widely used (Wills, 1988). 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 machines receive air from a blower.

The first mechanical machines were of the "cell-to-cell" type. They were complicated in comparison with designs such as the Forrester pneumatic machines, and this led to the introduction of the "open flow" concept. Over a period of 30 years these two types were cautiously developed to a unit capacity of about 2.8 m³ (Young, 1982). Young (1982) states that, at that point, the "open flow" machines were in the ascendancy, for several reasons:

-       simpler cheaper construction
-       simpler maintenance
-       better suspension of coarse solids
-       higher unit capacity

However, "cell-to-cell" machines were particularly used in smaller plants and multi-stage cleaner floats where the pumping action of the impellers permitted the transfer of intermediate flows without external pumps. A period of relative rapid development then occurred in which "open flow" machines were introduced in sizes up to 14 m³ whilst conventional "cell-to-cell" machines developed more slowly to 5.7 m³ and, eventually, also to 14 m³. The most important application of the large "cell-to-cell" machines was in fine coal cleaning where the users reported a significant improvement of selectivity over "open flow" designs (Young, 1982). In the late seventies and early eighties the "open flow" machines become available in sizes up to 42.5 m³. Co-current with this expansion of the more conventional mechanical units, the Maxwell cell was introduced in sizes up to 57 m³.

1.2.7.2 Mechanical flotation machine design

The new designs for this type of machine tend to have a specific technical advantage (better air dispersion, better flotation of fine particles, reduced reagent consumption, etc.). Modifications of older designs also occur and are intended to reduce overall costs, use new materials, or rectify a design defect.

Flotation cells from any given manufacturer come in a series of sizes, which have geometrical similarity. They can be characterised by (Barbery, 1984):

-       a geometric structure (e.g. parallelepiped, cylindrical) which defines the overall shape

of the cell (Figure 1.3),

- impeller design (impellers are often surrounded by a stator and a diffuser) and
- means of introducing air into the cell (diaphragm, diffuser, hollow impeller shaft, pipe located beneath the impeller, etc...)

The basic functions of a mechanically agitated cell are as follows:

- to maintain particles in suspension,
- to disperse air flow into bubbles in the slurry and
- to create relevant conditions for contact between air bubbles and hydrophobic particles.

Barbery (1984) gives the following criteria that cell design has to fulfill:

- the transfer of particles in the feed slurry from the feed inlet to the two discharge points (concentrate launder for hydrophobic particles and next cell or outlet for hydrophilic particles). It is important that the particles entering the cell be removed relative rapidly without having to be ground to fine sizes by attrition in the impeller area.
- low turbulence in the top part of the cell so that the froth layer is not disturbed and detachment of hydrophobic particles is minimised.
- easy start up after stoppage (due to a power failure for example) and sedimentation.

Barbery (1984) gave a detailed description of cell and impeller design (different impeller designs are presented in Figure 1.4) where dimensionless numbers are used to quantify the hydrodynamics of mechanical flotation cells. The usual numbers considered as well as typical values are (Harris, 1976):

Power number	0.5 - 5
Reynolds' number	$10^5 - 10^6$
Froude number	0.1 - 5
Air Flow number	0.01 - 0.1
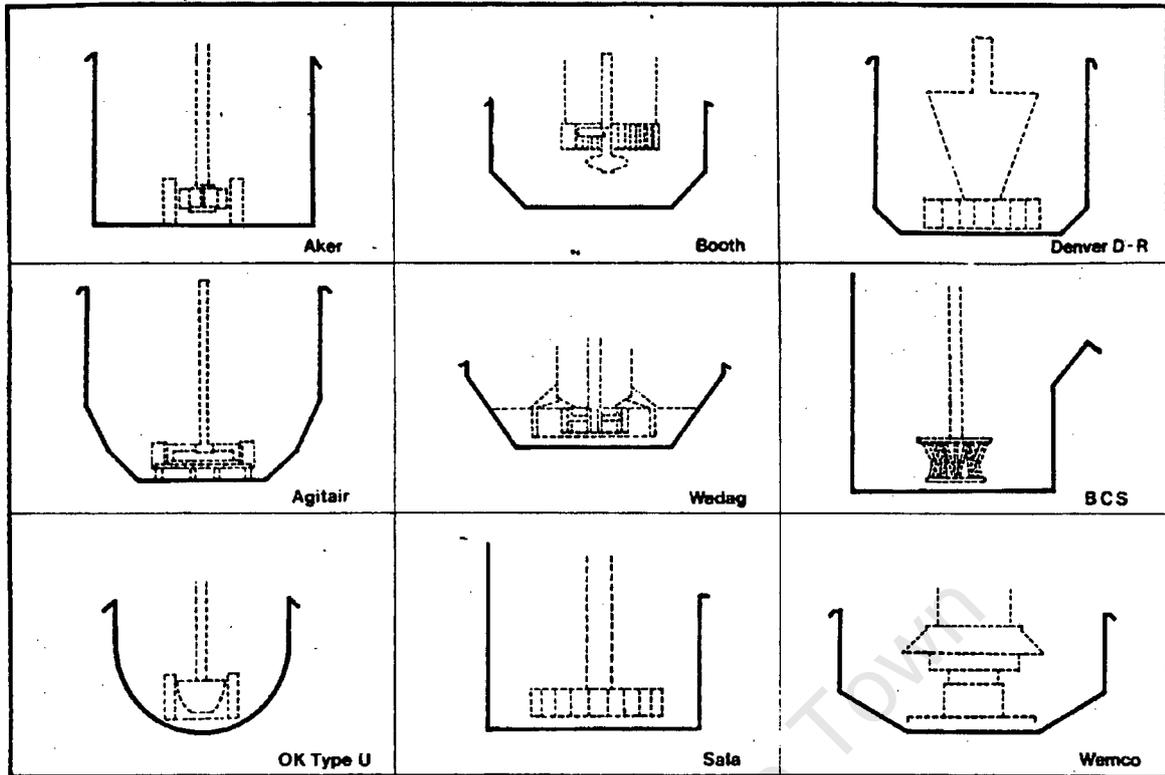
Figure 1.3: Cell tank profiles for "open flow" machines (Young, 1982).
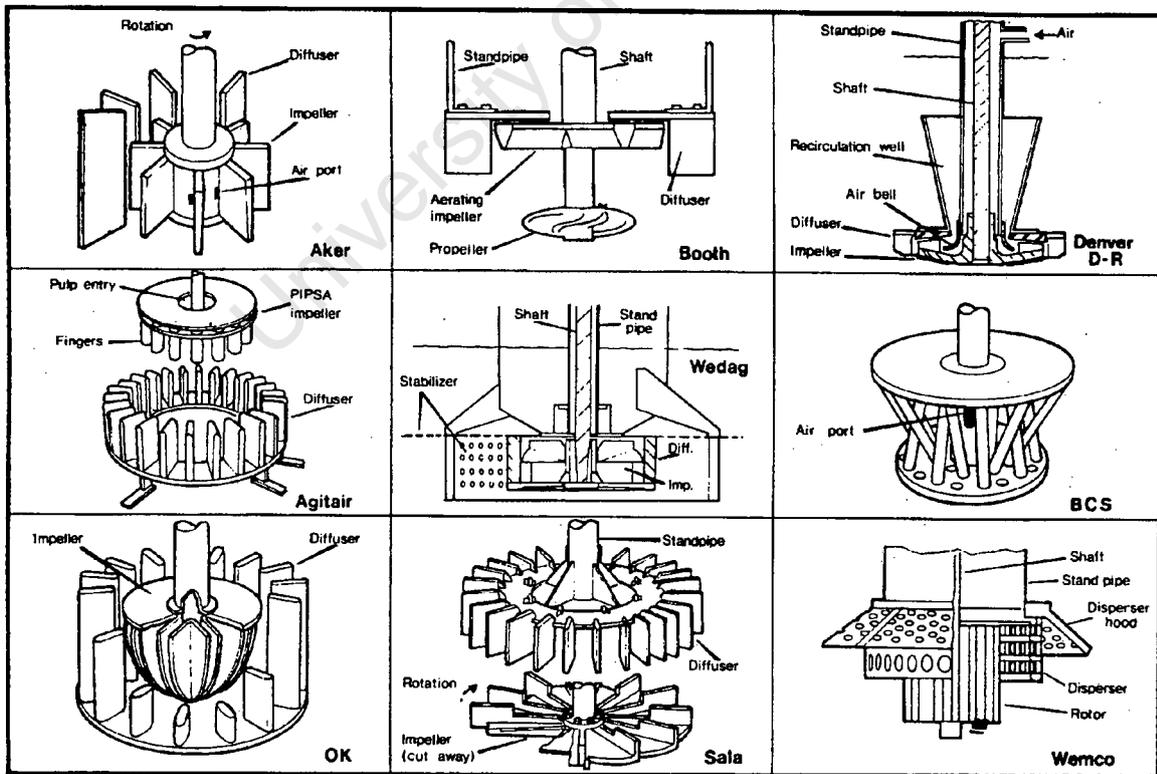
Figure 1.4: Impeller mechanism details of "Open flow" machines (Young, 1982).

The Weber number as well as the impeller peripheral speed are also important parameters.

1.2.7.3 Column flotation cell

Flotation columns are froth flotation devices that essentially eliminate gangue entrainment in the concentrate by counter current washing. Metallurgical performance along with low capital and operating costs have made column flotation popular in mineral processing.

The "conventional" column cell as used today was patented in the 1960s by Boutin and Trembley (Boutin and Trembley, 1964). Their column device was a complete break from the conventional mechanical cell still in use today. The column cell is schematically shown in Figure 1.5.
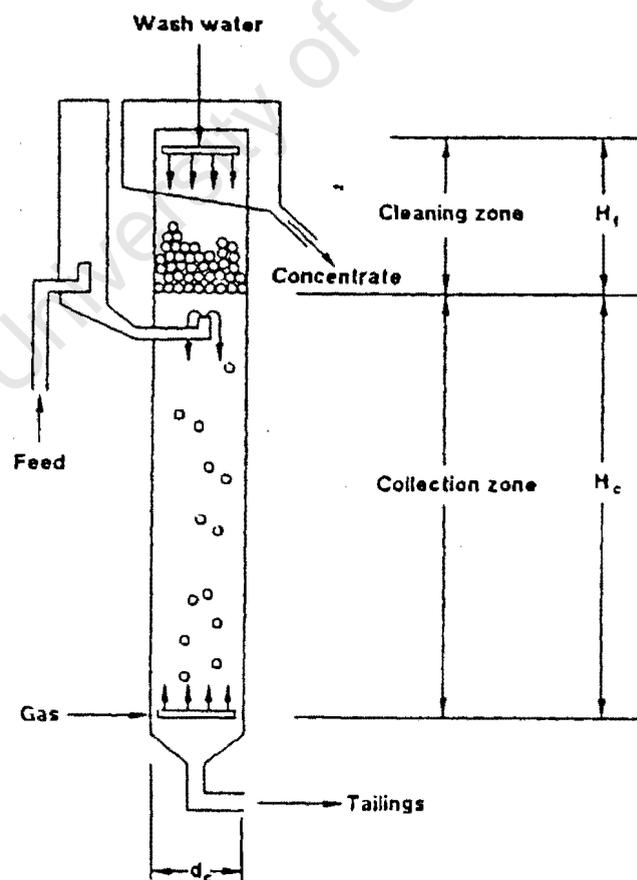
Figure 1.5: Column flotation cell (Dobby and Finch, 1990).

Commercial units are usually 9 to 15 meters high and 0.3 to 3 meters in diameter (Dobby and Finch, 1990). The cross section may be square (the design supplied by Column Flotation Company of Canada) or circular. Feed enters approximately one third of the way down the column and is distributed evenly. The slurry then travels down the column over the cross sectional area of the cell to meet a rising stream of bubbles which are produced in a sparging device. Particles which collide and attach to the bubbles are carried up the column, eventually reaching the interface between the pulp or collection zone and the froth zone. The froth zone in primarily gas with solids and liquid in the lamella between the bubbles. Washwater, which is added at the top of the column, filters down through this bed and washes the froth of entrained particles.

Apart from the shape, two features which distinguish the column from other flotation machines are the bubble generation system and the use of wash water.

1.2.7.4 Column flotation cell design

The lack of precise technical data for the design of industrial scale column flotation cells forced the industry to build columns on the experience and intuition of the constructor and the plant personnel (Brzezina and Sablik, 1991). Dobby and Finch (1990) proposed a scale-up method where parameters (rate constants and carrying capacity) are obtained from laboratory/pilot-scale columns. By making use of these parameters a industrial column could be designed with the appropriate collection zone size, bias rate, gas holdup and gas rate.

The columns designed in industry are very similar in shape, the point of input and of take-off of concentrate and tails. The main difference distinguishing them relates to the method of air dispersion employed (Brzezina and Sablik, 1991).

The objective of bubble generation in column flotation, as in any froth flotation system, is to produce relative small bubbles at a moderate air rate (typically, superficial gas velocity J_g = 0.5 to 2 cm/s). The size of the bubble produced is determined by the type of generation system and frother type and quantity. Thus, the optimum performance of a flotation column depends to a large extent on the design of the air sparger.

Spargers can be divided into two different types; internal and external spargers. Internal spargers are most commonly used (Dobby and Finch, 1990) and are situated near the base of the column. There are two broad categories of internal spargers: porous spargers (such as sintered glass) and single and multinozzle spargers (such as punched plate). The pre-mixing of air and liquid before introduced into the column is the main feature of an external sparger. An example of an external sparger is the recently developed Cominco system where the premixed water-air solution is injected through small orifices. This system seems to give an improved performance over porous media internal type spargers with greater durability and control.

1.2.7.5 Flotation machine selection

Flotation machines should ideally be selected to optimise metallurgical performance at minimum costs - both capital and operating. The selection of these machines is complicated by the fact that the same flotation machines are supposedly designed to separate valuable minerals from ores with widely different characteristics.

Poling (1980) proposed the following criteria for the general selection of flotation machines:

1. Machine selection should depend primarily on the type of flotation to be performed.
For example, for the handling of coarse solids, particularly accidental tramp oversize from plugged cyclones, mechanical cells of a high power number are preferred (Harris, 1976). High power number signifies high rate of displacement of pulp by the impeller. Another example is the flotation of low grade ore, when a low ratio of available froth area to cell volume can give improved froth stability and improved recovery (Poling, 1980).

2. Capital and operating costs are important.
While low bids often win supply contracts for flotation machines, this practise should not always be followed. Optimizing metallurgical performance is far more important than the life of a plant (Poling, 1980). Low maintenance, low down-time, low supervision, ease of automation are also critically important and must be learned through experience.

3. Other factors - such as the competence and quality of service available locally, availability of spare parts, previous experience of the design engineer and metallurgist with similar equipment and corporate ties with equipment manufacturers, can also affect machine selection. Too often operating personnel are not involved in these equipment selections.

1.2.8 Flotation circuits

Flotation plants always consist of several stages which are usually called roughing, scavenging or cleaning depending on their position in the plant. Each stage usually consists of several flotation cells, referred to as a flotation bank, connected so that the tailings pass sequentially through the cells of each stage. The first few cells in a flotation bank where high grade concentrate is collected, is known as a rougher. Rougher concentrates are usually sent for another concentration step. The cells handling the second concentration step are the cleaners. The last few cells in a flotation bank where low grade froths are present are scavengers. The scavenger concentrate is usely recycled to the head of the rougher or sent to a cleaner. The collection of stages with their interconnecting feed, concentrate and tailing streams is called the plant or circuit configuration. The plant or circuit configuration depends on the type of ore to be treated. A plant that must extract and separate two different minerals will have a different configuration from one that must separate one mineral from the gangue.

1.2.8.1 Circuit layout and 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. Bulatovic and Wyslouzil (1988) indicate that flowsheet configuration affects metallurgical performance significantly.

Traditionally, flowsheet selection is based solely on the sequential flotation of two or more valuable minerals, where complete circulation of the intermediate products within the circuit is practised (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 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; and

(3) past experience and trial and error techniques.

1.2.8.2 Circuit control variables

The stability of a circuit will depend on the operation and control of the roughers, scavengers and cleaners. The control variables which enable the operator to control the performance of the circuit are:

(1) the addition rates of the different reagents,

(2) the aeration rate to each cell and

(3) the level of the pulp in the cell.

The change in any variable at any bank in a circuit causes a change in the composition and flow rate of each stream leaving that bank, and changes in all later streams in the circuit; also in earlier streams if there is a recycle stream going back to a previous bank.

1.2.8.3 Flotation circuits treating UG-2 ore

The flotation circuits employed for Merensky and UG-2 ore are designed essentially for the flotation of base-metal sulphides (BMS). Alloys of noble metals and other metals are not recovered. Flotation takes place at 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 in ratios up to 7:3 (Liddell *et al.*, 1986). Copper sulphate is added as an activator to the

milling circuit or 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 of more consistent composition. Owing to the talcose nature of the ore, use is made of a gangue depressant, which is normally an organic colloid.

The flotation circuit configurations differ for the treatment of Merensky and UG-2 ore. This can be expected since there is a difference in the overall as well as PGM mineralogy of the two ores. Usually for Merensky ore a single flotation stage, with a rougher stage and sometimes multi-stage cleaning, is used. The main sulphide minerals floated in a plant treating Merensky ore are pentlandite, chalcopyrite and pyrrhotite. Corrans *et al.* (1982) report a 87% PGM recovery in Merensky ore flotation in industry.

The flotation circuit layout for the treatment of UG-2 ore is much more complex than for Merensky ore. This is due to more slow floating minerals in UG-2 as well as a constraint on the composition of the final concentrate. The chromite (Cr_2O_3) content of the final concentrate must be below 3%. A higher chrome content will cause severe problems in the smelting process. It is believed that the chromite is recovered through physical entrainment. Therefore, a rougher and cleaning circuit with a long residence time to enable flotation of the slow floating minerals, are used and multi stage cleaning with a brittle froth, to reduce the amount of physical entrained particles, and cater for the Cr_2O_3 problem.

There are several different milling and flotation configurations used on the platinum concentrators today. Goodall (1995) describes four configurations (Figure 1.6), viz.:

(1) Mill-Float (MF1),
(2) Mill-Mill-Float (M2F),
(3) Mill-Float-Mill-Float (MF2) and
(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 40 - 50% passing -75μm to enable the flotation of already liberated particles before overgrinding occurs. A rougher and scavenger with a two or three stage
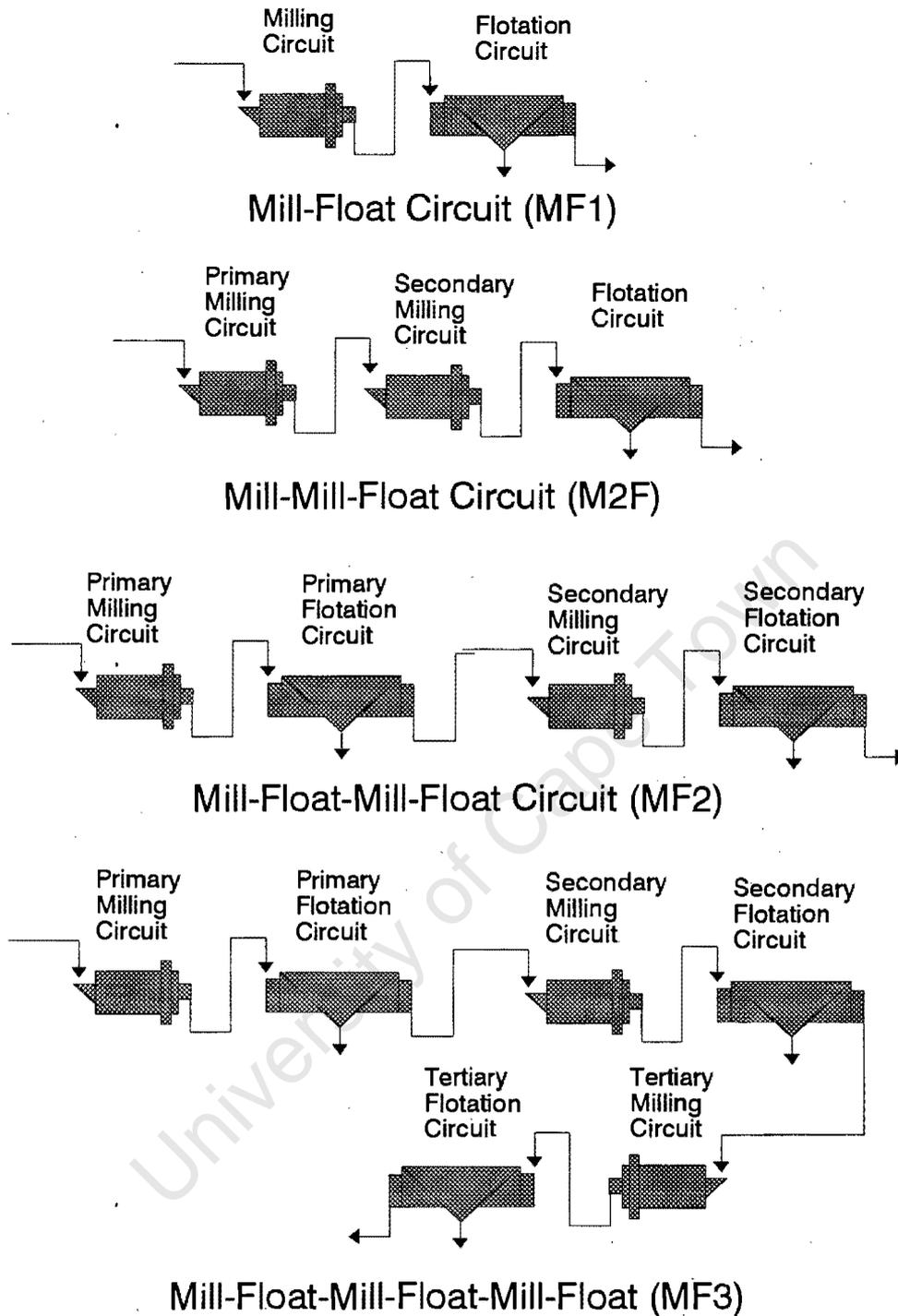
Figure 1.6: Schematic diagram of flotation circuits treating UG-2 ore (Goodall, 1995).

cleaning circuit is common practise for the primary float. The tails from the primary section are usually reground to 80% passing -75 μm to liberate the PGMs in the fine sulphide grains (< 15 μm). The secondary float usually has more cleaner capacity than the primary since the

presence of slow floating minerals will be higher than in the primary float.

1.3 Residence time distribution

1.3.1 General theory

The residence-time distribution (RTD) of a reactor is a characteristic of the mixing that occurs in a chemical reactor. The RTD is determined experimentally by the injection of an inert tracer, at time zero, and measuring the tracer concentration in the effluent stream as a function of time. The system is therefore disturbed and the response to this stimulus will give the desired information about the system. Any type of input signal may be used to create a response.

The two most commonly used input signals are a step signal and a pulse signal. The drawbacks in the step injection of a tracer are that it is sometimes difficult to maintain a constant tracer concentration in the feed. Obtaining the RTD from this test also involves differentiation of the data that could lead to large errors. A third problem lies with the large amount of tracer required for the test. The principal difficulties with the pulse technique lie in the problems connected with obtaining a reasonable pulse at the reactor entrance. The injection must take place over a very short period compared with residence time in various segments of the reactor or reactor system (Fogler, 1992). There must also be a negligible amount of dispersion between the point of injection and the entrance to the reactor.

The elements of fluid, in a vessel, will take different routes to the exit and therefore require different lengths of time to do so. The distribution of these times is called the residence time distribution E of fluid. The RTD is presented in such a way that the area under the curve is unity and is called normalised.

$$\int_0^\infty E dt = 1$$

The RTD curve is characterised by its first two moments, τ, the mean residence time, and σ^2, the variance. The mean residence time reflects the mass transport phenomena in that it

defines the relationship between the vessel volume and the volumetric feed rate. The variance in turn reflects the degree of mixing which occurs within the vessel. The mean residence time can be determined as follows,

$$\tau = \int_0^\infty tE\,dt$$

and the variance,

$$\sigma^2 = \int_0^\infty (t - \tau)^2 E\,dt$$

In treating models it is more convenient to measure time in units of mean residence time. This then gives a dimensionless measure.

$$\theta = \frac{t}{\tau}$$

There are two types of ideal flow; plug flow in which the fluid stream passes through the vessel without mixing with the vessel contents, and fully mixed flow in which the fluid elements are immediately dispersed throughout the vessel. Actual or nonideal flow has components of both types of ideal flow.

Many models can be used to characterise nonideal flow within vessels. Dispersion models draw on analogy between mixing in actual flow and a diffusional process while other models build a chain or network of ideal mixers.

1.3.2 Tanks-in-series model

The tanks-in-series model is well documented in the literature (Levenspiel, 1972). The E curve and moments of this model are easy to obtain since problems of boundary conditions (methods of tracer injection and measurement) do not intrude. For N tanks in series

$$E_\theta = \frac{N(N\theta)^{N-1}}{(N-1)!} e^{-N\theta}$$

If $N = 1$ the system will behave as a continuous stirred tank reactor (CSTR) and if N becomes very large, the behaviour of the system approaches that of a plug-flow reactor.

With the tanks-in-series model the recovery of a flotation vessel can be predicted as a function of

$$R = f(k, \tau, N)$$

where k is the first order rate constant. The equation for recovery is,

$$R = 1 - \frac{1}{(1 + k\frac{\tau}{N})^N}$$

1.3.3 Axial dispersion model

In this model there is an axial dispersion of material governed by Fick's law of diffusion. All the contribution to backmixing of fluid flowing in the x direction can be described by the following expression,

$$\frac{\partial C}{\partial t} = D \frac{\partial^2 C}{\partial x^2}$$

where D is the axial dispersion coefficient, which uniquely characterises the degree of backmixing during flow. Levenspiel (1972) gives this dispersion model in dimensionless form,

$$\frac{\partial C}{\partial \theta} = (\frac{D}{uL}) \frac{\partial^2 C}{\partial z^2} - \frac{\partial C}{\partial z}$$

with the dimensionless numbers $\theta = \frac{t}{\tau} = \frac{tu}{L}$ and $z = \frac{x}{L}$ where L is the length of the vessel and u is the interstitial velocity of the fluid. The dimensionless group, $(\frac{D}{uL})$ called the vessel dispersion number, is the parameter that measures the extent of axial dispersion. Thus

$$(\frac{D}{uL}) \rightarrow 0 \quad \text{negligible dispersion, hence plug flow}$$

$$(\frac{D}{uL}) \rightarrow \infty \quad \text{large dispersion, hence mixed flow}$$

The solution to this dispersion model depends on the boundary conditions. The boundary

conditions are the points where the tracer is injected into the vessel and where the response is measured. The two conditions treated by most workers in this field are the closed vessel and open vessel (Levenspiel, 1972). For the closed vessel there is a change in the flow pattern at the boundaries and for the open vessel there is no change in the flow pattern at the boundaries. The closed vessel case has to be solved numerically while an analytical solution exists for the open vessel (Levenspiel, 1972).

1.3.3.1 Closed vessel

Manqiu Xu et al. (1991) used the finite difference method to get a numerical solution for a closed vessel. The following finite difference equation is derived by the authors,

$$E_{i,j+1} = E_{i,j} + \frac{N_d \Delta \theta}{\Delta x^2} (E_{i+1,j} - 2E_{i,j} + E_{i-1,j}) - \frac{\Delta \theta}{2 \Delta x} (E_{i+1,j} - E_{i-1,j}) \qquad (1.9)$$

with $i = 0...M$ and $j = 0...N$, where $\Delta \theta = \dfrac{\tau}{N}$, $\Delta x = \dfrac{1}{M}$ and N_d is the vessel dispersion number.

The initial and boundary conditions are transformed into the finite difference equation as follows,

$$j = 0, \quad E_{i,0} = 0 \text{ for all } i, 0 < i < M$$

$$i = 0, \quad E_{0,j+1} = \frac{N_d}{\Delta x + N_d} (E_{i,j} + \delta)$$

$$i = M, \quad E_{M,j+1} = E_{M-1,j}$$

where δ is the Dirac delta function (tracer input):

$$\delta = M, \text{ at } t = 0, j = 0$$
$$\delta = 0, \text{ elsewhere.}$$

The results computed by equation (1.9) must be examined for stability and accuracy. Instability can be avoided if the following criteria are met,

$$\frac{N_d \Delta \theta}{\Delta x^2} \leq \frac{1}{2}$$

$$\frac{N_d}{\Delta x} \geq \frac{1}{2}$$

The finite difference equation converges to the exact solution of the partial differential equation as Δx and $\Delta\theta$ tend to zero (Manqui Xu *et al.*, 1991).

1.3.3.2 Open vessel

With the open vessel case another complication is introduced, in that the way in which the concentration of the tracer is measured will determine the shape of the RTD curve. If the concentration of the tracer is measured in a discrete fashion then the boundary conditions will closely represent the closed vessel boundary conditions. Continuous measurement, where the flow of the fluid is not disturbed, is more in accord with the open vessel. The following analytical solution for continuous measurement, without disturbing flow, is given (Levenspiel, 1972),

$$C_\theta = \frac{1}{2\sqrt{\pi\theta(\frac{D}{uL})}} \exp[-\frac{(1-\theta)^2}{4\theta(\frac{D}{uL})}]$$

where C_θ is the dimensionless concentration.

1.3.4 Application of residence time distribution studies to flotation

Residence time distribution studies have been widely used in the flotation industry, especially column flotation. Various authors reported the use of RTD studies to model the collection zone of a flotation column (Dobby and Finch, 1990; Mills, 1992).

Dobby and Finch (1985) performed RTD studies on full scale flotation columns with typical operating conditions. The 0.46m and 0.91m diameter columns were each 13m high. Fluorescein and non-floatable manganese dioxide tracers were used to monitor the liquid and solids flow. Both the tracers prove to be problematic in their application, the fluoroscein was absorbed on the particles in the system and the feed material contained a small amount of

MnO_2. Their results show, that for large columns, the axial dispersion coefficient follows the dependence upon column diameter. The experiments have also verified the observation by other authors that the axial dispersion coefficient of solids is equal to that of liquid.

The application of RTD studies in the evaluation of conventional flotation machines is not well documented in the literature. Nesset (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 the various process and economical trade-offs when selecting the number of cells and the size. The paper also deals with the problem of internal recirculation in hog-trough cells. The detailed discussion on tracer characteristics and selection is probably the most valuable information in the paper.

1.4 The modelling of the flotation process

"Notwithstanding the importance of this process to the mineral industry, the development of quantitative engineering tools needed for the evaluation of the flotation process is still largely in the evolutionary stage." (Dowling *et al.*, 1985)

In the literature, the construction of mathematical models of flotation has been approached in various ways. This depends on the philosophy of the researcher as well as the expected usage of the model and the allowable investment in personnel and time (Lynch *et al.*, 1981). Published models range from the type based on fundamental equations of science and engineering to the empirical type. It must be emphasised that models of complex processes as flotation can never represent the actual process completely and it is necessary to recognise the weaknesses of any model.

The importance of describing the flotation process has been realised by numerous workers in the mineral processing industry (Woodburn and Loveday, 1965; King, 1974; Huber-Panu *et al.*, 1976; Klimpel, 1980; Lynch *et al.*, 1981). In the last 30 years there have been sporadic efforts to use kinetics of flotation models to describe laboratory and industrial data

quantatively (Huber-Panu *et al.*, 1976; Klimpel, 1980; Lynch *et al.*, 1981, Dowling *et al.*, 1985; Loveday and Raghubir, 1995) and more recently the pace has quickened. Models present the worker with a means to assess and predict the performance of a flotation cell, hence the opportunity to improve the operation of the cell.

Woodburn (1970) divided mathematical modelling into three categories: characterisation, control and optimisation. In many cases it is desirable to be able to describe in terms of a number or set of numbers the flotation characteristics of an ore. It must be recognised that characterisation implies at least a minimum predictive power for the model, e.g. prediction of recovery for a constant flotation time (Woodburn, 1970). Mehrotra (1988) and Yingling (1993) reviewed in detail the design and optimisation of circuit networks splitting the work done into two groups on the basis of problem formulation and solution strategies. Both groups employ a structural parameters approach to represent circuit configuration alternatives, which, without care, may result in solutions involving complicated circuit configurations (Yingling, 1993).

The present research work focused on the characterisation of a flotation circuit and this section will therefore focus on some models used for this category.

1.4.1 Empirical models

Empirical models are obtained from statistical correlations between dependent and independent variables of the flotation process. These models usually apply to sections and entire flotation plants. The statistical correlations for these models can be obtained by normal multiple linear regression methods or spline regression methods (Mular, 1972). Early empirical models were provided by Faulker (1966).

The main advantage of an empirical model is that they are often less expensive in personnel and time than other types and if the expected use of the model is well defined, the resulting model can be a satisfactory engineering tool (Lynch *et al.*, 1981). However, predictions outside the model's data base should be treated warily while examination of the performance of alternate circuits for a particular ore is not possible. Further, the model can vary greatly for different ores and circuit arrangements (Lynch *et al.*, 1981).

1.4.2 Probability models

The first step towards an analytical, as opposed to purely empirical, investigation of flotation kinetics was made by Schuhmann (1942) who considered the rate determined step to be particle-bubble collision. Gaudin (1932) had earlier stated this was the predominant mechanism for particle-bubble attachment in flotation cells. Schuhmann proposed that the recovery rate of particles is related to the probability of success, P_x, of a sequence of events which must occur in a flotation compartment before a particle can reach the concentrate launder. He developed the following equation for particles of average size x:

$$P_x = P_c\, P_a\, F\, x\, V$$

where P_c and P_a are the probabilities of particle-bubble collision and adhesion, F is the froth stability factor and V is the cell volume. Schuhmann assumed that flotation is a first order rate process and defined the flotation rate constant k as:

$$k = P_c\, P_a\, \phi$$

where ϕ is the fruitfulness factor which compensates for the fact that some particles which are attached to the bubble can later become detached. It was reported by Lynch *et al.* (1981) that Tomlinson and Fleming suggested that

$$F = P_e\, P_f$$

where P_e is the probability of a particle being retained on the gas bubble and lifted through the liquid; and P_f is the probability of the particle being retained in the froth. Once a relationship for the joint probability is established, it is relative simple, using population balance models, to compute the mass rates of flotation. This approach has been followed by Kelsall (1960) who postulated a probability of flotation , which referred to the probability of a solid particle in the feed to the cell leaving in the concentrate.

Lynch *et al.* (1981) concluded that to define the probabilities of successive events in a particular mechanism, approximations of doubtful validity are necessary in order to apply fundamental laws to the complex gas-solid-liquid system. Futher the resultant equations can contain parameters which are difficult to evaluate.

1.4.3 Kinetic models

1.4.3.1 Simple first order model

In the early 1940's, Schuhmann (1942) pointed out that the specific flotation rate has the same dimensional significance as the specific reaction rate for a first order chemical reaction. Also that the two parameters are similarly related to the kinetics of the respective processes they describe. Thus chemical kinetics is concerned with interactions between atoms, molecules, or ions, while flotation kinetics deals with interactions between bubbles and particles. This approach, assuming that flotation is a first order process, has been widely accepted and used in the modelling of the flotation process (Huber-Panu et al., 1976; Klimpel, 1980; Dowling et al., 1985, Lynch et al., 1981; Loveday and Raghubir, 1995).

From the assumption that flotation is a first order reaction the rate equation is,

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

where C is the concentration and k the rate constant. Integration leads to,

$$C = C_o e^{-kt}$$

for the conditions $C_i = C_o$ when t=0, and $C_i = C$ when t=t.

Morris (1952) suggested that all the mineral particles capable of floating usually do not float. Thus, it may be preferable for the value chosen for C_o to refer only to that portion of the feed which will float in prolonged time. The concentration of the material remaining in the tailings C_∞ subtracted from materials in the feed C_o gives a corrected value, $C_o = C_\infty$, to replace C_o and C is replaced by $C - C_\infty$.

$$C - C_\infty = (C_o - C_\infty) e^{-kt}$$

Garcia (1935) and Morris (1952) obtained C_∞ by trail and error; it is the value that makes the plot ln $(C_o - C_\infty)/(C - C_\infty)$ versus t linear.

In terms of recovery, the equation thus becomes,

$$R = R_\infty(1 - e^{-kt})$$

where R_∞ is the maximum theoretical flotation recovery.

Because of the experimental uncertainty in fixing time zero for a batch flotation experiment, it is necessary to introduce an additional correction factor (Θ) which is peculiar to batch data (Agar and Barrett, 1983); thus the equation is modified to:

$$R = R_\infty(1 - e^{-k(t + \Theta)})$$

A plot of $\ln((R_\infty - R)/R_\infty)$ versus $(t + \Theta)$ should produce a linear line of slope k. However, R_∞ and Θ are both unknown.

The scheme used for simultaneously fitting the three parameters begins with an arbitrary estimate of R_∞ (usually 1.0); calculation of k and Θ with a least squares routine then calculation of a residual sum of squares. R_∞ is then decreased by a previously set but adjustable amount and the calculation is repeated until values of k, Θ and R_∞ are found which minimize the residual sum of squares.

1.4.3.2 The distributed size parameter model

Earlier work (Gaudin, 1932) had pointed to the effect of different size particles on flotation recovery. Trahar (1981) reviewed the role of particle size on flotation and emphasised how crucial it is to recognise its effects in flotation processes. The effect of particle size on the flotation rate was first studied by Gaudin et al. (1942) but their findings were not brought to a logical conclusion. Sutherland (1948) did subsequently and concluded that the total flotation rate was the sum of the rates of each size class. According to Fichera and Chudacek (1992), Tomlinson and Fleming were the first to test experimentally and confirm the validity of the size distribution model,

$$R = \sum_{i=1}^{n} R_{0,i} (1 - e^{-k_i t})$$

with i denoting a particular particle size and $R_{0,i}$ being the maximum theoretical recovery for

particle size i. Their results showed that with use of a size distributed model for flotation, there existed a potential for its use as a predictive model.

1.4.3.3 The floatability parameter model

Fichera and Chudacek (1992) reviewed batch flotation models for conventional flotation cells and they recognised another distributed parameter model that has been investigated in the literature as the distributed floatability model. This model lumps together the effect of particle size as well as all other factors influencing flotation such as particle shape, surface properties etc. Imaizumi and Inoue (1965) realised that the non-linearity of curves of unfloated material plotted against time on a log-linear axes was not because of a non-first order rate but because of a distribution of different floatabilities of the material in the feed sample.

The distributed floatability model can be written as

$$R = \sum_{i=1}^{m} M_i (1 - e^{-k_i t})$$

where R is the mineral recovery, M_i is the fraction associated with a floatability class and k_i the rate constant of floatability class i.

Other authors (Woodburn and Loveday, 1965) also accepted the concept of the distributed first order rate constant because of the heterogeneous nature of mineral feeds, in which exist a distribution of particle sizes, mineralogical compositions, particle shape etc. Woodburn (1970) made it clear that particle size is not the sole source of variation in the rate constants. He also stated that it is a natural consequence to free the variation of k from a direct dependence on particle size, and to postulate instead a one-dimensional spectrum of k values distributed through the entire solid mass. He used a model that implies a continuous distribution, $f(k)$,

$$R = R_\infty \left(1 - \int_0^\infty f(k) \, e^{-kt} dk\right)$$

The evaluation of the precise form of $f(k)$ explicitly proved to be a matter of some difficulty.

Huber-Panu *et al.* (1976) assumed that if f(k) follows a rectangular distribution then $f(k) = \frac{1}{k}$. The equation therefore became,

$$R = R_\infty(1 - \frac{1}{Kt}(1 - e^{-Kt}))$$

This well known equation was later popularized by Klimpel (1980) for reagent testing.

It was recently concluded by Fichera and Chudacek (1992) that the particle size effect is one of the most predominant factors influencing the rate of flotation and the easiest to analyse. Other factors influencing particle floatability are not so easy to isolate and therefore cannot be accounted for individually. However, these factors can be accounted for collectively in the form of a floatability spectra either for mineral feed or its various sizes.

The difference between the batch or semi-batch and continuous process is that for the continuous case the residence time of all the elements in the system is distributed. Woodburn and Loveday (1965) incorporated the residence time distribution function, $E(t)dt$, in their distributed rate model used in batch flotation to make it applicable to plant cells. The distributed rate model therefore becomes,

$$R = R_\infty (1 - \int_0^\infty \int_0^\infty f(k) \, E(t) \, e^{-kt} \, dk \, dt)$$

It is now well established that, for all practical purposes, a conventional flotation cell behaves like a perfectly mixed reactor (Woodburn and Loveday, 1965; Loveday, 1966; Lynch *et al.*, 1981), but the mean residence time is a function of operating condition, e.g., particle size, density, rpm, etc. (Mehrotra, 1988). If the cell is a perfect mixer with mean residence time τ, it was shown that the mass flow rates of particular species with rate constant k in the concentrate stream, F_c, at steady state is described by (Mehrotra and Kapur, 1975)

$$F_c(k) = F_f(k) (\frac{k\tau}{1 + k\tau})$$

where F_f is the feed flow of the particular species.

Lynch *et al.* (1981) showed that for N perfectly mixed, isolated compartments in series, with constant retention time per compartment, the cumulative fractional recovery of a floatable component from N compartments may be calculated as,

$$R = 1 - \frac{1}{(1 + k\tau)^N}$$

This equation can therefore predict the recovery of a floatable component in a flotation bank if the specific rate constant, k, the residence time, τ, and the number of cells, N, are known.

1.4.4 Two-phase models

Arbiter and Harris (1962) were among the first to suggest a two-phase model of the flotation machine attempting to account for kinetic behaviour, both transient and steady state. The model distinguishes between the froth phase and the pulp phase and assumes that ideal mixing occurs in both phases and that there is two-way mass transfer between the phases according to simple kinetics. This model postulated by Arbiter and Harris was extended by Harris and Rimmer (1966) and Harris (1978), who developed the two-phase model into multiphase models with several layered phases in the froth to account for concentration gradients. It was concluded by Harris (1978) that the two-phase model describes the steady state satisfactorily but it is only marginally successful in describing the transient state. He also stated that the model does not scale-up well and the model does not take water flow or air flow between the phases into account although flow in and out of the system is described.

1.4.5 Water recovery models

Ideally, a simulation model should be able to predict the water losses in the concentrate and tailings stream, since an estimate of water flow rates around the cell is necessary because of its direct bearing of the residence time calculations. Schrimgeour *et al.* (1970) assumed a flotation rate for water and calculated water balance on this basis. The limitation of this approach is that water behaviour along a entire bank cannot be simulated (Smith, 1984). Loveday and Marchant (1972) assumed a value for the residence time in the cell and calculated the rate of water flow from the knowlegde of cell volume and pulp density. King (1973) assumed a constant value for pulp density for the concentrate stream, thus enabling

a water balance to be made and residence time to be calculated. Sutherland (1977) observed that for the lead flotation circuit, the assumption of fixed pulp density in the tailing stream gave good results. This assumption, according to Mehrotra (1988), is not likely to be valid since it implies that the feed and concentrate material have the same pulp density. This would introduce significant errors in the cleaning section since here the pulp density decreases dramatically while travelling down the bank.

A two component model assuming fast and slow rate coefficients can be used to reproduce the experimental data (Lynch *et al.*, 1981), however the approach has no physical significance and only serves to highlight the lack of knowledge in this area (Smith, 1984). Smith (1984) in an attempt to model water behaviour made the following observations:

(1) Water is recovered to the concentrate when no solids are present,

(2) the percent solids in a flotation concentrate tends towards a maximum value,

(3) water recovery down a set of rougher scavenger banks can be modelled by a two component model and,

(4) a relationship exists between water recovery and mineral recovery.

Smith (1984) considered water recovery to be due to lifting in the wake of ascending air bubbles and therefore the water flow to the froth due to air. Two proportional constants were used; one for the transfer of water from the pulp to the froth due to the action of air and another for the transfer of water from the pulp to the froth due to the action of solids.

1.5 Objectives of the research

The largest reserves of Platinum-group metals (PGM) in the world are situated in the North West Province of South Africa and are part of the well known Bushveld Complex. The Platinum mines in the area have exploited the Merensky Reef for the past sixty years or so. Below the Merensky Reef lies a series of Chromium seams, one of these being the Upper Ground 2 Reef (UG-2) which contains significant reserves of PGM minerals. This Reef was not exploited until the beginning of the 1980's, but since then the proportion of this ore relative to the total ore mined has increased steadily. The design of efficient, high capacity concentrators to treat UG-2 ore become increasingly important.

There are a number of important differences in the concentration of UG-2 ore in comparison to Merensky ore, owing to the high chromite content, and the predominance of slow floating minerals. In particular, the chromite content of the product from the concentrator must be maintained below specified limits in order to prevent a rapid build-up of chromite in the hearth of the furnace during smelting, as this reduces the working volume of the furnace and makes tapping difficult. This constraint has had a significant impact on the design of the milling and flotation circuits treating UG-2 ore.

At the Impala Platinum concentrator, near Rustenburg, a complex two stage milling and flotation circuit has been developed to treat the UG-2 ore. The optimisation of this type of circuit with respect to PGM recovery and chromite rejection represents a challenging problem, owing to the high variability of many important parameters in the circuit, such as ore characteristics, flow rates, and pulp densities. This has made highly impractical, if not impossible, the use of classical mass balance techniques to adequately characterise the performance of the circuit. A different approach to circuit characterisation and optimisation was therefore required, to accommodate a high degree of variability in a complex circuit.

The main objectives of this research were:

1)   to develop a methodology to characterise the complex MF2 circuit by careful analysis of the performance of various key integral sections/units of the circuit. This was done by carrying out batch flotation tests on key streams;

2)   to use the batch flotation results to obtain kinetic parameters;

3)   to characterise the hydrodynamic behaviour of each unit using classical RTD experimental procedures;

4)   to use batch flotation kinetic parameters, the RTD results and actual recoveries in mass balance equations to determine scale-up parameters and

5)   to use scale-up factors to simulate the performance of the circuit and to evaluate this procedure using other circuits.

Chapter Two

The circuit

The flotation circuits used to float UG-2 ore are designed essentially for the flotation of base-metal sulphides (Liddell *et al.*, 1986). Circuit configurations used in the flotation of UG-2 ore are more complex than the circuits used to float other ores containing PGM minerals, on account of the different composition of UG-2 ore which consists of slow floating minerals as well as a high chromite content.

The chromite places a constraint on the design of circuits treating UG-2 ore since the final concentrate should only contain between 2 - 3% of chromite to prevent problems in the smelting process. The circuits are therefore designed with more than one cleaning stage, usually two to three (Liddell *et al.*, 1986).

In the processing of UG-2 ore the major costs are in getting the ore to the surface and therefore maximum recovery is essential in the flotation process. To obtain maximum PGM recovery the particles containing the PGM minerals should be floated as soon as they are liberated, but the PGM grains in the ore tend to be less than 12 - 15 μm and hence a fine grind (80% -75μm) is desirable (Goodall, 1995). In order to meet these requirements the use of two stage milling and flotation (MF2) is common practice.

At the Impala Platinum concentrator a MF2 circuit has been developed to treat UG-2 ore. The objective of the present research project was to characterise this type of circuit to enable the optimisation of the circuit configuration.

2.2 The different circuit layouts

A trial and error method was used in the optimisation process of the MF2 type circuit operated at the Impala Platinum concentrator treating UG-2 ore. This created several flowsheets of the configuration or layouts of the circuit. During the research period five different configurations were operated but only two were investigated. In this section the two relevant circuit configurations will be discussed in detail.

The research project was divided into two phases. The first phase investigation was done on a different circuit configuration than the second phase. The first circuit is referred to as the Phase 1 circuit and the second as the Phase 2 circuit.

2.1.1 The Phase 1 circuit

The Phase 1 circuit layout is presented in Figure 2.1 and consisted of two milling stages.

Figure 2.1: The Phase 1 circuit layout

The circuit layout was in a MF2 configuration which is a mill followed by a float followed by a mill and another float. The primary and secondary milling circuits were in closed circuit configuration. In each case, the mill product was sent to a cyclone and the cyclone underflow was sent back to mill since it consisted of coarse material while the overflow was sent to the float. The primary flotation section consisted of a rougher/scavenger and three cleaning stages. The scavenger tails as well as the primary cleaner tails were sent to the secondary mill discharge sump. The material sent to the secondary milling circuit first went

through the cyclone to separate the fines from the coarse material to prevent over-grinding.

The primary rougher concentrate was sent to the final concentrate sump while the scavenger concentrate was sent to the primary cleaner. The concentrate of the first few cells on the primary cleaner was treated as final concentrate while the middlings were sent to the primary recleaner. The concentrate coming from the first few cells of the primary recleaner was also treated as final concentrate while the middlings were sent to the primary column which acted as a re-recleaner. The primary recleaner tails as well as the column tails were sent to the primary cleaner feed sump.

The secondary rougher concentrate were sent to the final concentrate sump while the scavenger concentrate went to the secondary cleaner. The secondary scavenger tails were treated as final tails. The concentrate from the first few cells of the secondary cleaner went to the final concentrate sump while the middlings were sent to the secondary column. The secondary cleaner tails were sent to the secondary mill discharge sump. The secondary column tails were directed to the secondary cleaner feed sump.

2.1.2 The Phase 2 circuit

The Phase 2 circuit layout was in a MF2 configuration and is presented on Figure 2.2. The primary and secondary milling circuits were both closed circuits where the cyclone underflow was returned to the mill and the overflow to the flotation section.

To maximise PGM recoveries in the circuit a philosophy that high grade liberated PGM minerals should be recovered to final concentrate as soon as possible was implemented. The concentrate of the first few cells of the conventional cells in the circuit were therefore sent to the final concentrate sump (Figure 2.2). The primary scavenger tails were directed to the secondary mill discharge sump while the scavenger concentrate went to the primary cleaner. The primary cleaner tails were directed to the primary surge tank to give the PGM minerals in the tails another opportunity to report to the final concentrate sump. The primary cleaner middlings were the primary column's feed material and the column tails were recycled to the primary cleaner feed box.
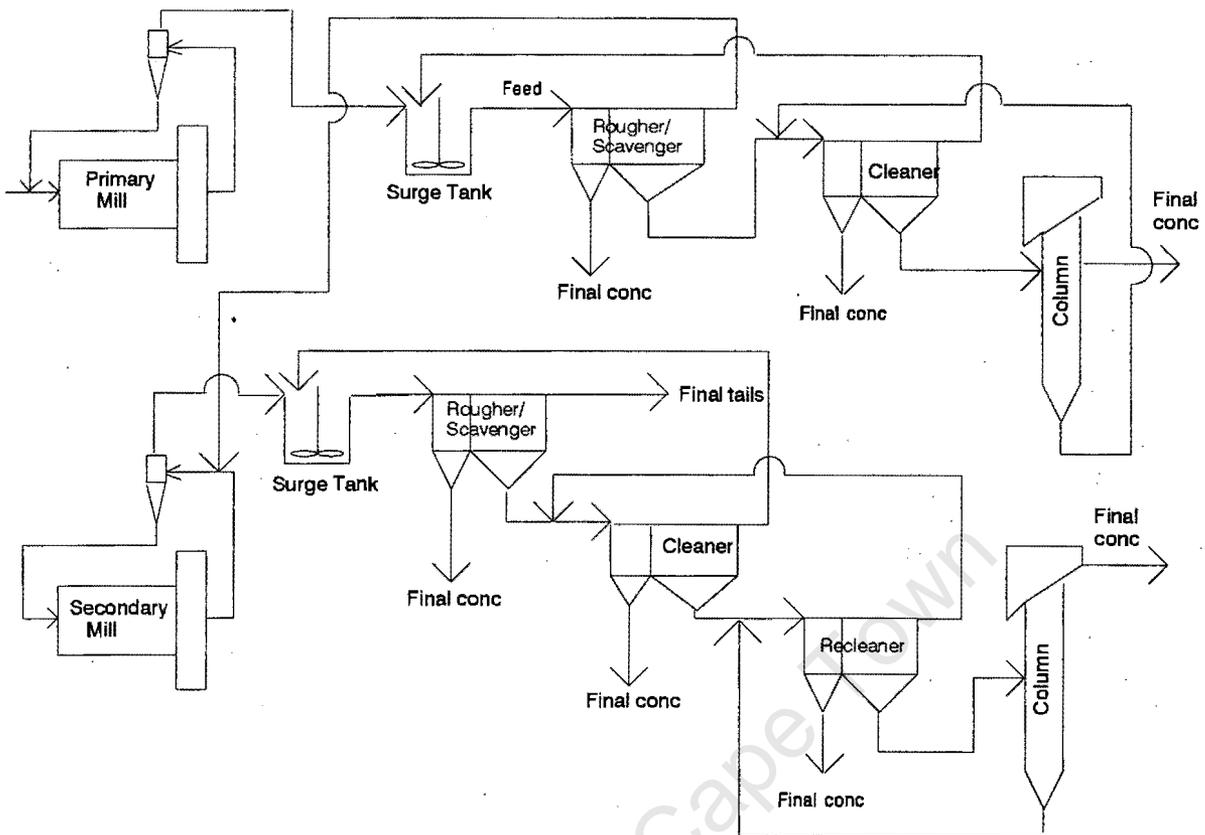
Figure 2.2: The Phase 2 circuit layout.

The secondary scavenger tails were treated as final tails while the scavenger concentrate was sent to the secondary cleaner. The secondary cleaning circuit consisted of three cleaning stages compared to the two of the Phase 1 circuit. The secondary cleaner tails were directed to the secondary surge tank to increase the probability of the PGM minerals in the tails reporting to the final concentrate. The secondary cleaner middlings were sent to the secondary recleaner. The secondary recleaner tails were recycled to the feed sump of the cleaner. The secondary recleaner middlings were the feed to the secondary column which acted as a re-recleaner. The column tails were directed to the secondary recleaner feed sump.

2.2 Description of the equipment in use

The equipment used in the Phase 1 and Phase 2 circuits was identical with only the location of certain flotation cells and streams being different. This section describes the equipment

as used in the Phase 2 circuit (see Table 2.1).

Table 2.1: Equipment of the Phase 2 circuit

Unit	Cell type	Installed power of cell mechanism (kW)	Capacity per cell (m³)	Number of cells in series
Primary section Surge tank		75	26	
Rougher/scavenger	Wemco	18.7 - 22.5	8.4	12
Cleaner	Wemco	3.7	0.6	12
Column	Cominco		15.2	
Secondary section Surge tank		75	26	
Rougher/scavenger	Wemco	18.7 - 22.5	8.4	12
Cleaner	Dorr Oliver	15	8.4	4
Recleaner	Wemco	3.7	0.6	12
Column	Cominco		8.5	

2.3 The operating conditions

The operating conditions, i.e., grind, pH, SG and reagents will be discussed in this section.

2.3.1 Grind

The primary mill was operated to produce a grind of 40% -75μm. The coarse grind limits the entrainment of chromite into the primary flotation concentrates and prevents the over-grinding of particles already liberated through milling or mining. The secondary mill which receives the primary section tailings operated at a grind of 80% -75μm. The regrinding of the tails liberates the minerals not yet liberated in the primary milling and since the PGM grains in UG-2 ore tend to be < 12μm a fine grind is desirable.

2.3.2 pH

The effect of pH on PGM grade and recovery had been investigated on Wildebeest North 2 shaft UG-2 ore and Bafokeng North 7A shaft UG-2 ore (Van Rensburg, 1991). The pH was varied between 9 and 12.5. Low pH values were not investigated since the investigator reported that most collectors are more stable in alkaline conditions and the corrosion of cells and pipework is minimized under these conditions (Van Rensburg, 1991). The best results were obtained in the pH range 9 - 10 for both shafts. Since the natural pH is around 9 it was recommended that no modifications be made to the current pH.

2.3.3 % Solids

The primary cyclone overflow which was the feed stream of the primary float had between 49% and 52% solids. The feed to the secondary float had 39% to 43% solids.

2.3.4 Reagents

A 50:50 mix of isobutyl xanthate (SIBX) and dithiophosphate (DTP) was used as a collector in the flotation of UG-2 ore. $CuSO_4$ was used as an activator while Acrol, which is an organic colloid used to depress talcaceous minerals, was used as a depressant. Dowfroth which is an alcohol-based frother was used during the research period.

Table 2.2: Reagent addition used in the Phase 1 circuit.

Reagent	Dosage points	Dosages (g/t)	
		Primary float	Secondary float
Collector	Conditioner	60	30
Activator	Conditioner	90	45
Depressant	Rougher feed box	230	130
	Cleaner feed box	90	60
Frother	Rougher feed box	5 - 10	–

The dosage points and dosages of the reagents are presented in Table 2.2 and Table 2.3.

Table 2.3: Reagent addition for the Phase 2 circuit

Reagent	Dosage points	Dosages (g/t)	
		Primary float	Secondary float
Collector	Conditioner	60	30
Activator	Conditioner	90	45
Depressant	Rougher feed box	80	40
Frother	Rougher feed box	5 - 10	-

2.4 The different operating philosophies

The circuit was operated to maximise PGM recovery and minimise chromite recovery. PGM recovery was maximised by sending the fast floating minerals in the first few cells of each flotation unit to final concentrate. The flow rates of these final concentrate streams depended on the chromite content of the concentrate. The chromite content in the final concentrate was monitored using the Amdel in-stream analyzer to maintain the chromite to below 2%.

The first operating philosophy that was followed was implemented in the Phase 1 and Phase 2 circuit. The 2% chromite content was maintained by altering the flow rates of the final concentrate streams for each unit. If the chromite content was above the 2% limit then the final concentrate flow rates were decreased by first starting at the secondary rougher then the cleaners and if the chromite was still not below the limit the primary rougher final concentrate flow rate was altered followed by the cleaners. If the chromite content of the final concentrate was well below the 2% limit an increase in the flow rates of the final concentrate streams was implemented starting at the primary cleaner followed by the rougher. If the chromite content was still below 2% the flow rates of the secondary cleaners followed by the rougher were steadily increased.

The positive point about this approach was that the chromite content of the final concentrate

was always on target. The main drawback was that the flow rates of the final concentrate streams varied significantly and it was difficult to maximise the PGM recovery since the concentration ratio (CR) was not constant at an optimum value.

The second operating philosophy was implemented on the Phase 2 circuit and the objective was to maintain a constant optimum CR where a maximum PGM recovery was obtained and the chromite content of the final concentrate was on target. If the chromite content of the final concentrate was higher than instructions permitted then the flow rates of the final concentrate streams were only decreased when a high chromite content was maintained over a 2 to 3 hour period. Since mass balances showed that the PGM content of the first few rougher cells were very high and the chromite content compared well to the cleaner concentrates, their final concentrate flow rates were the last ones to be decreased while the cleaners' final concentrate flow rates were adjusted first.

The second operating philosophy gave the best PGM recoveries for the Phase 2 circuit configuration. The chromite content of the final concentrate tended to be higher over longer periods (2 - 2.3%) as well as lower (1.8 - 2%) over longer periods than before.

Chapter Three

Experimental details

Experimental design is always the most important aspect of any research project. In this project a methodology for the evaluation of complex circuits was developed, since the classical mass balance approach for the characterisation of a flowsheet was not applicable for the circuit under investigation. The main reason for this was the high variability of important parameters in the circuit, such as ore characteristics, flow rates, and pulp densities.

Since a new approach had to be followed and no guidelines were available, the methodology was developed during the research. New techniques were used as the need for certain data arose. This methodology consisted of unit performance evaluation using sampling and batch flotation tests, residence time distribution studies, "down-the-bank" sampling of certain unit operations and mass balance campaigns.

The techniques used to collect the data for the characterisation of the circuit are described in this chapter. Since sampling and sample preparation is the basis of all the applied methods, it is the first subject under discussion in this chapter. The assay techniques and procedures used to analyse the samples are presented and are followed by the discussion of the evaluation of the circuit which consists of the evaluation of unit performances, residence time distribution studies, "down-the-bank" sampling and mass balance campaigns.

3.1 Sampling and sample preparation procedures

3.1.1 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. If this is not so bias is easily introduced and the sample is not representative.

The samples were collected by intersecting a stream at regular intervals. For effective

representability a complete cross-section of the ore stream was taken with each sample cut.

3.1.2 The sample size

The size of the representative sample (M) to give the required accuracy was calculated by using Gy's method (Gy, 1979). Gy's basic sampling equation can be written as:

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

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

Assuming normal distribution, 67 out of 100 assays of samples would lie within $\mp s$ of the true assay; 95 out of 100 assays would be within $\mp 2s$ of the true assay, and 99 out of 100 would be within $\mp 3s$ 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 an acceptable probability level (Wills, 1988).

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

$$C = f\, g\, l(\frac{\rho}{a})$$

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:

$$d_z/d_x > 4, \qquad g = 0.25$$
$$d_z/d_x \text{ is } 2 - 4, \qquad g = 0.5$$
$$d_z/d_x < 2, \qquad g = 0.75$$
$$d_z/d_x = 1 \qquad g = 1$$

where ρ 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:

$$I = (\frac{d_l}{d})$$

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

The sample size needed for a 95% confidence level for PGM and $\%Cr_2O_3$ analysis was roughly calculated for each type of stream that was sampled during the evaluation period and is listed in Table 3.1.

Table 3.1: Calculated sample sizes

Stream	PGM grade (g/t)	C (g/cm)	M (g)
Primary rougher feed	3 - 5	0.002	270.26
Primary rougher concentrate	300 - 600	1.00×10^{-5}	0.02
Primary scavenger concentrate	50 - 70	7.14×10^{-5}	0.43
Primary scavenger tails	0.7 - 1	0.009	1158.27
Primary cleaner feed	50 - 90	7.14×10^{-5}	0.43
Primary cleaner concentrate	150 - 250	3.00×10^{-5}	0.06
Primary cleaner middlings	50 - 90	3.75×10^{-5}	0.08
Primary cleaner tails	15 - 35	3.57×10^{-4}	2.13
Secondary rougher feed	1 - 2	0.006	810.79
Secondary rougher concentrate	40 - 80	4.29×10^{-5}	0.09
Secondary scavenger middlings	25 - 40	1.43×10^{-4}	0.85
Final tails	0.6 - 0.8	0.01	1351.32
Secondary cleaner feed	25 - 55	1.19×10^{-4}	0.71
Secondary cleaner concentrate	80 - 100	4.29×10^{-5}	0.09
Secondary cleaner middlings	50 - 60	6.01×10^{-5}	0.13
Secondary cleaner tails	10 - 20	3.57×10^{-5}	2.13

The relative density (ρ) of the precious mineral was calculated by combining the densities of pentlandite, pyrrhotite and chalcopyrite. As describe in section 1.1, the PGMs in UG-2 ore

are primarily associated with these three minerals.

Since the samples were taken in ten-litre buckets, the mass of the dry solids in the sample will differ for each sample and will depend on the % solids in the stream sampled. Ten litres of sample proved to be more than adequate to meet the confidence levels required.

3.1.3 Sampling points

The feed streams of the rougher/scavenger units were sampled at the overflow of the surge tanks. In the preliminary investigation (Phase 1) the rougher feed was sampled at the discharge of the feed stream into the feed box of the rougher. The collected samples proved to be non-representative since the high flow rate of the feed stream made it impossible to obtain a cut of the complete cross-section of the stream. Therefore, an alternative sampling point was located and the overflow of the surge tank proved to be the ideal location. The rougher concentrate and scavenger concentrate (middlings) streams were sampled at the final concentrate sump and the feed sump of the cleaner respectively. The tailing stream was sampled at the dead box of the tailing sump where the flow rate of the stream was reduced to enable representative sampling.

The cleaners' feed streams were sampled at the feed boxes of the different units. The concentrate and middling streams were sampled at the discharge points into the different sumps. The primary and secondary cleaner tailing streams were sampled at the point of discharge into the primary and secondary surge tanks respectively. The tailing stream of the secondary recleaner was sampled at the secondary cleaner feed sump. The feeds to the columns were sampled at the columns' feed sumps while the concentrate streams were sampled at the final concentrate sump and the primary and secondary column tailing streams at the primary cleaner feed and secondary recleaner sump respectively.

3.1.4 Sample preparation

The samples collected from the sampling campaigns were prepared for analysis in the metallurgical laboratory. The samples were collected in ten litre buckets. The buckets filled with sample as well as the empty buckets were weighed for future calculations of liquid to

solid ratios. Dewatering of the samples was done in two stages. Firstly pressure filtering was applied followed, secondly with drying in a laboratory oven. The mass of the dried sample was recorded before it was sent to the laboratory for analysis.

3.2 Assay procedures

In the preliminary investigation the samples were assayed for PGM and %chromite. The PGM results were obtained by using the Fire Assay technique while the %chromite was determined with UV/VIS spectroscopy. In Phase 2 of the evaluation the samples were assayed for the individual metals, platinum, palladium and rhodium using the SAFT technique. Details of all the assay procedures are given in Appendix I.

3.2.1 Fire Assay

In a melt, the precious metals are collected in molten lead, which forms a separate phase, by solvent extraction. After solidifying, the phases are separated, and the lead and silver, which is added as an auxiliary collector, removed by a combination of evaporation, oxidation and absorption in ceramic vessels. The precious metal bead remaining is weighed on a sensitive balance. The Fire Assay method is presented in more detail in Appendix I.

In an alternative modification to this standard method, there is a procedure similar to the above, but without the addition of silver. This method is referred to as the Twenty Minute Method.

3.2.2 The SAFT technique

The samples are weighed into a litharge based flux. The mixture is fused so that the platinum group metals are collected in the reduced lead. Once the melt has been poured and cooled the slag is detached from the solid lead button. These buttons are then subsequently specially prepared to that they may be presented to a spark emission source which is calibrated to determine platinum, palladium and rhodium. The SAFT technique is presented in more detail in Appendix I.

3.2.3 UV/VIS spectroscopy

The sample is fused, leached in water and iron removed by filtration. Excess peroxide is destroyed by the addition of manganese sulphate. After making to volume the solutions are analysed colorimetrically. The determination of chromite by UV/VIS spectroscopy is presented in more detail in Appendix I.

3.3 Circuit evaluation

" *When evaluating the performance of a commercial flowsheet, a method consisting of laboratory studies and plant surveys of individual unit operations was found to be effective.*" (Bulatovic and Wyslouzil, 1993).

The circuit was divided into individual units and sections (see Figure 3.1 and 3.2) and each evaluated individually over a period of a few days. A bank of flotation cells is referred to as a unit and more than one flotation bank is referred to as a section. The sampling of all the process streams around the unit enabled the determination of the unit's performance in terms of mineral or metal recovery. Each unit was sampled for a period three times longer than the calculated residence time. In addition, feed and tailing samples from each unit were subjected to batch flotation in the laboratory. The objective was to float the samples under the same conditions as on the plant. The samples were therefore floated as quickly as possible after collection. The batch flotation results were used as a diagnostic indicator in the evaluation of the unit performances. Collection of all the data occurred over a period of three to four months.

Two different circuit configurations were evaluated by dividing the circuit into units and sections. The work done on the first configuration (Figure 3.1) is referred to as Phase 1. Phase 1 represents the preliminary work and includes sampling around the units and batch flotation test work on the feed and tailing streams of each unit. The investigation done on the second configuration (Figure 3.2) is referred to as Phase 2. The Phase 2 evaluation includes sampling around the units, batch flotation test work on the feed streams to each unit and on the final tails, residence time distribution studies, "down-the-bank" sampling and circuit mass balance campaigns.
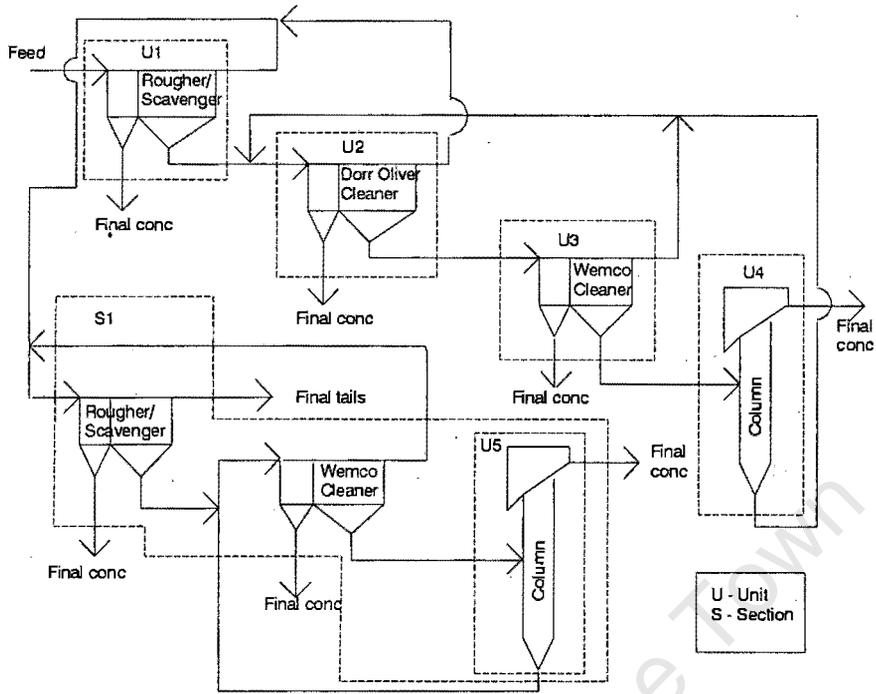
Figure 3.1: Units and section for Phase 1 (Preliminary work).
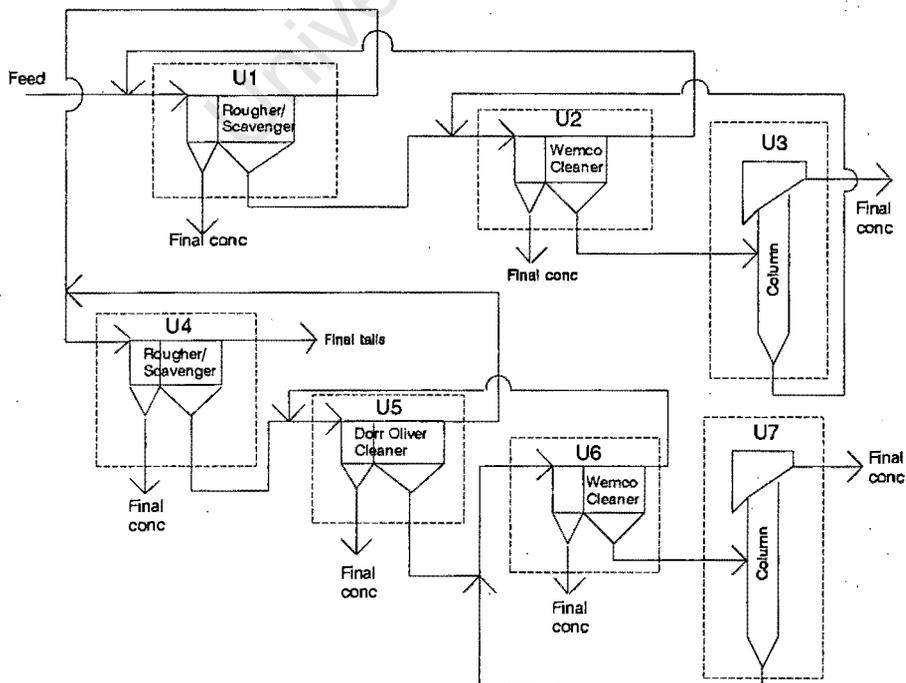
Figure 3.2: Units for Phase 2

The batch flotation procedure as well as the equipment and operating parameters are discussed in the following sections. After the batch flotation, the experimental procedure used in the residence time distribution studies is discussed. The "down-the-bank" sampling follows and Section 3.3 is ended with the discussion of the mass balance campaigns.

3.3.1 The batch flotation procedure and equipment

3.3.1.1 Equipment and operating parameters

A D12 Denver batch flotation machine, with a two or eight litre cell depending on the size of the sample, was the standard batch flotation equipment. The fixed flotation parameters are listed in Table 3.2.

3.3.1.2 Flotation time

Preliminary batch flotation tests were conducted to establish the suitable flotation time for further test work. Figure 3.3 shows that twenty-five minutes is a suitable flotation time for this type of ore.

Table 3.2: Fixed flotation parameters

Parameters	Conditions
Ore type	UG2
Pulp density	same as plant
Impeller speed (8 l)	1500 rpm
(2 l)	1200 rpm
Temperature	30 - 35 °C
Froth depth	2 - 3 cm
pH	8 - 9
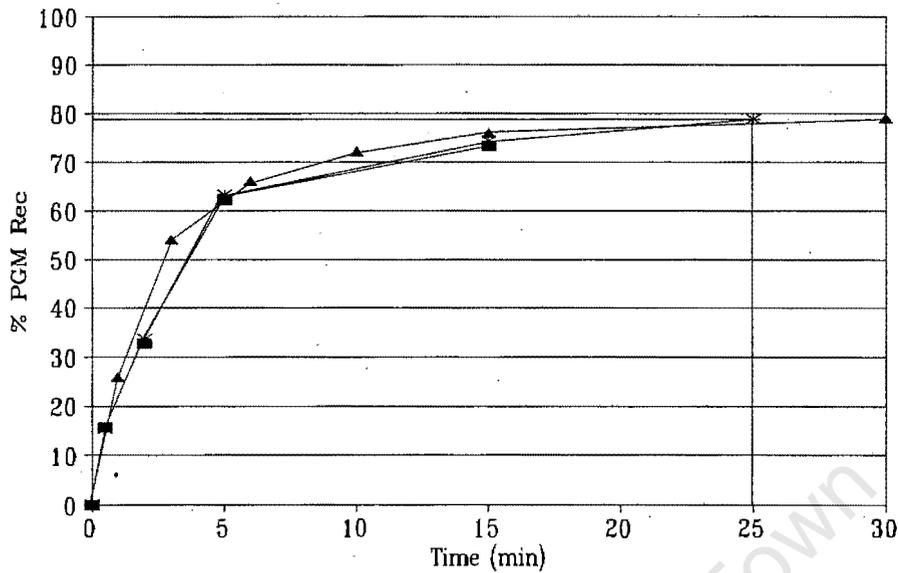
Figure 3.3: PGM recovery versus time curve for primary
rougher feed.

3.3.1.3 Standard procedure

The sample collected on the plant was poured into an eight litre batch flotation cell and stirred to keep the solids in suspension. Before the flotation started a feed sample was taken. Since the samples were collected on the plant no reagents were added except for the samples taken at the head of the primary and secondary rougher/scavengers. Only frother and depressant were added in these cases since the collector and activator were added into the conditioners and were already present in the samples. The conditioning time for the frother was five minutes and one minute for the depressant.

The air valve was opened at time zero and concentrates were sampled by scraping in 10 second intervals. The concentrate was taken over a 25 minute period. Five concentrates were taken as described in Table 3.3. The concentrate intervals differ for the batch tests of Phase 1 and 2. It was found in the Phase 1 batch tests that not enough concentrate was collected in some cases even after 30 seconds. Therefore, the first interval was extended to 1 minute.

After the air valve was closed, the concentrate samples were weighed, filtered in a Buchner

type filter and dried in the laboratory oven. The remaining sample or tails was pressure filtered and dried. Each sample's mass was recorded before the filtering process and after it was dried. This data enabled the calculation of the water recoveries.

Table 3.3: Concentrate intervals

Concentrate	Intervals for Phase 1 (min)	Intervals for Phase 2 (min)
1	0 - ½	0 - 1
2	½ - 2	1 - 2
3	2 - 5	2 - 5
4	5 - 15	5 - 15
5	15 - 25	15 - 25

3.3.2 Residence time distribution studies

Residence time distribution measurements involve the introduction of a tracer to the vessel feed and either continuous monitoring or periodic sampling (discrete method) of the tracer in the vessel discharge.

3.3.2.1 Liquid phase assumption

In a flotation plant one is most concerned with the behaviour of the solids. The use of solid tracers, by no means impossible, does tend to add to both the complexity and cost of the test. Therefore the assumption that the liquid phase of the slurry approximates the behaviour of the solid phase had been made. Dobby and Finch (1985) have shown that the validity of this approximation increases with the decrease in particle size. Since the particle size in the flotation circuit under investigation is very fine, viz. between 50% and 80% passing -75 μ for the feed to the roughers and even finer for the feed to the cleaners, the agreement between the liquid phase and solid phase residence time distributions should be very close.

3.3.2.1 Tracer selection

The guidelines used by Nesset (1988) for selecting a suitable tracer was followed. Preliminary background "noise" measurements were taken to establish a constant baseline. Because a constant baseline could be established as well as the ease of detectability, low cost and the non-interference with the process made sodium chloride the obvious choice.

3.3.2.3 Test procedure

A saturated solution of sodium chloride (NaCl) was prepared in a 25 litre bucket and introduced into the feed box of the vessel under investigation. The objective was to create a pulse input and the bucket's contents was therefore emptied into the feed box as quickly as possible (within 1 to 2 seconds). Discrete conductivity measurements were taken, in three minute intervals, at the vessel discharge. The samples were taken in a 1 litre bucket. The sampling period was extended until the conductivity reading advanced to the closest base line reading. It was found that the most consistent readings were obtained by stirring the sample for 15 seconds, leaving it for a minute to enable the solids to settle, and keeping the probe above the settling solids.

3.3.3 "Down-the-bank" sampling

"Down-the-bank" sampling was conducted on the primary rougher unit, since changes in the configuration of the circuit will have the least effect on this unit. The main objective of the "down-the-bank" sampling was to obtain data from the plant for the scale-up of kinetics. Therefore, a recovery versus time profile had to be constructed.

The most important aspect of the "down-the-bank" sampling campaign was the accurate sampling of the froth that comes over the lip of the cell. This was done by cutting across the lip of the cell with the sampler. The sampling was done over a three hour period.

3.3.4 Mass balance campaigns

A unique solution for a steady-state mass balance requires a minimum number of sampling

points. Wills (1988) describes the simple nodal analysis that can be used for calculating the number of sampling points needed for a unique solution.

3.3.4.1 Nodal analysis

The flowsheet can be reduced to a series of nodes where process streams either join or separate (Figure 3.4). The circuit under investigation could be divided into seven nodes where each flotation unit represents a node.

A connection matrix C_{ij} was generated where each element in the matrix is, +1 for stream j flowing into the ith node, -1 for stream j flowing out of the ith node and 0 for stream j not appearing at the ith node. There are twenty flow streams and seven nodes.
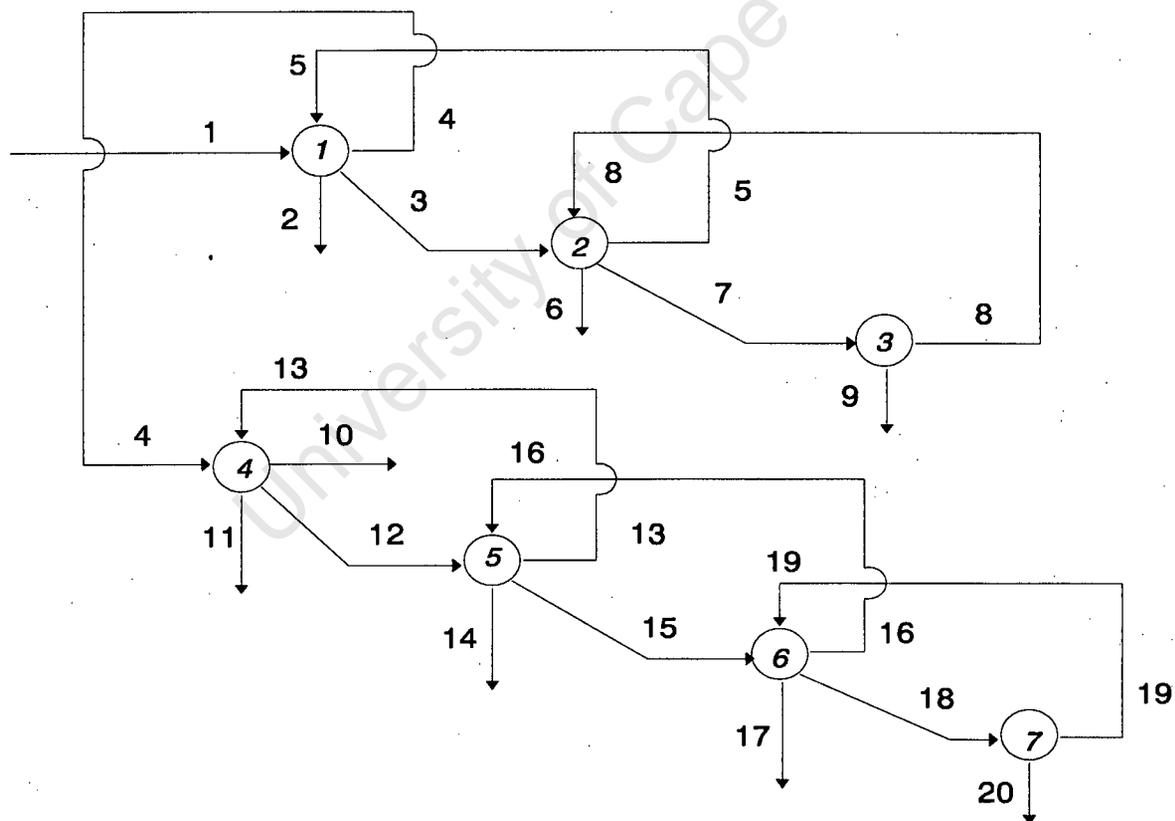
Figure 3.4: Flowsheet in node form

The connection matrix thus has seven rows and twenty columns as shown below:

$$
\begin{bmatrix}
+1 & -1 & -1 & -1 & +1 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 \\
0 & 0 & +1 & 0 & -1 & -1 & -1 & +1 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 \\
0 & 0 & 0 & 0 & 0 & 0 & +1 & -1 & -1 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 \\
0 & 0 & 0 & +1 & 0 & 0 & 0 & 0 & 0 & -1 & -1 & -1 & +1 & 0 & 0 & 0 & 0 & 0 & 0 & 0 \\
0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & +1 & -1 & -1 & -1 & +1 & 0 & 0 & 0 & 0 \\
0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & +1 & -1 & -1 & -1 & +1 & 0 \\
0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & +1 & -1 & -1 \\
\end{bmatrix}
$$

The contents of each column represent the individual streams, and when summed must equal +1, -1 or 0, any other result indicating an error in the input of the data, +1 indicates a feed stream, -1 a product stream and 0 an internal stream. The summation of the columns show that there is one feed stream and eight product streams. The elements of each row represent the individual nodes and if the number of "+1" entries (n_p) and the number of "-1" entries (n_n) are counted, then n_p and n_n can be used to assess the number of simple nodes:

Number of simple junctions (J) = n_p - 1
Number of simple separators (S) = n_n - 1.

There are thus 5 simple junctions and 12 simple separators (Table 3.4), and the minimum number of streams that must be sampled is:

$N = 2(F + S) - 1$
$\quad = 2(1 + 12) - 1$
$\quad = 25$

Since their are only twenty available streams, five additional mass flows were required to supplement the reference stream in order to calculate a balance.

Table 3.4: Nodes classification

Nodes	n_p	n_n	J	S
1	2	3	1	2
2	2	3	1	2
3	1	2	0	1
4	2	3	1	2
5	2	3	1	2
6	2	3	1	2
7	1	2	0	1
			5	12

3.3.4.2 Sampling campaign

The streams were sampled over a six to eight hour period in half hour increments. The locations of the sampling points are shown in Figure 3.5.
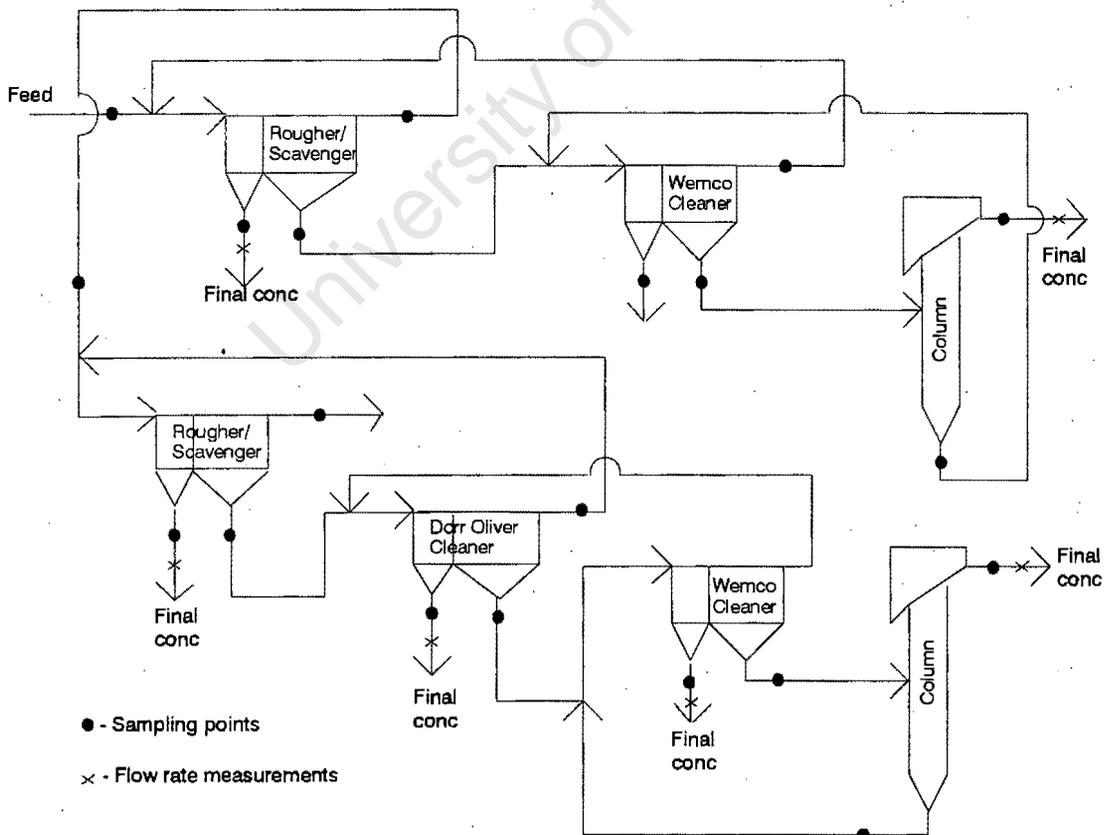
Figure 3.5: Sampling points and locations where flow rates were measured.

· The mass flows were taken by measuring the time to fill a ten-litre bucket. Mass flows of the primary rougher concentrate, primary column concentrate, secondary rougher concentrate, secondary cleaner concentrate, secondary recleaner concentrate and secondary column concentrate were measured.

The samples were prepared as described in section 3.1 and sent for platinum, rhodium, palladium and chromite analysis.

Chapter Four

Results

The circuit under investigation was evaluated by dividing it into different units, and each unit was evaluated separately. The experimental techniques described in Chapter 3 were used to evaluate the different units. This chapter presents the data obtained during the evaluation.

The chapter starts with the reproducibility data obtained from the different experimental techniques. The circuit evaluation was done in two phases. The first phase, which was the preliminary work, follows the reproducibility section. The second phase is presented after the preliminary work.

4.1 Reproducibility of experimental techniques

The objective of this section was to show that the experimental techniques and procedures used in the investigation were reproducible. The reproducibility of the sampling, assay, batch flotation, residence time distribution studies and mass balance techniques will be discussed.

4.1.1 Sampling

As explained in Section 3.1, sampling is the basis for most of the techniques used to evaluate the performance of the circuit. It was therefore crucial that the sampling technique produced representative samples reproducibly.

The sampling technique that was used is described in Section 3.1. The streams were sampled by intersecting the cross section of the pipe with a 1 litre sampling bucket. The primary cleaner tails were sampled three times in succession to test for reproducibility. Each sample was collected over a period three times longer than the cleaners' calculated residence time. The samples were screened and the results are presented in Figure 4.1. The sample standard deviation (STDS) and the standard error (STE) are listed in Table 4.1. The standard error for the size fractions is less than 1.5% and the average standard error for all

the size fractions is 1.1%.
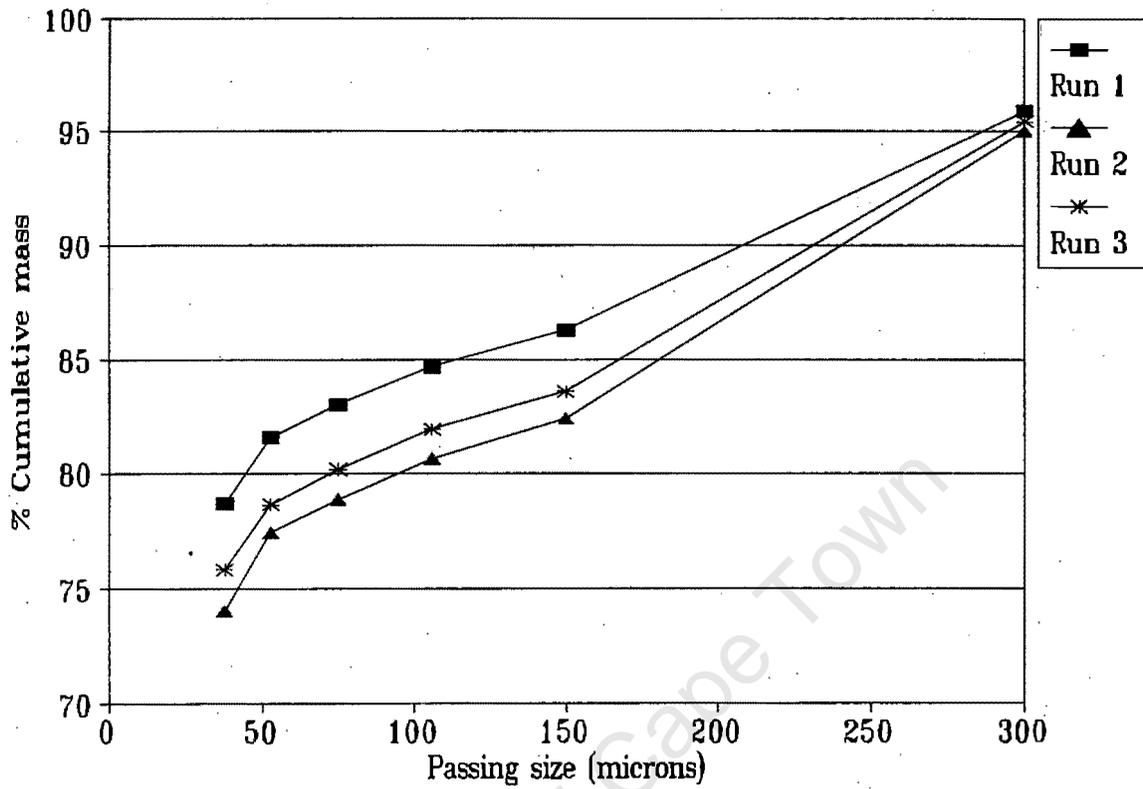
Figure 4.1: Size distribution curve for Primary cleaner tails

Table 4.1: Sample standard deviation and sample error.

Size interval (µm)	Sample standard deviation (STDS) (%)	Standard error (STE) (%)
-300	0.45	0.26
-150	2.00	1.15
-106	2.06	1.19
-75	2.12	1.22
-53	2.12	1.22
-38	2.36	1.36

Since these samples were taken on a full scale plant where a high degree of variability existed in parameters like flow rates and pulp densities, this sampling technique gave an acceptable degree of reproducibility.

4.1.2 Assaying

The samples taken during Phase 1 were analysed for PGM content as well as for % chromite. The Phase 2 samples were analysed for platinum, palladium, rhodium and % chromite. The PGM content of a sample was analyed using the 6 hour Fire Assay method while the platinum, palladium and rhodium content of a sample was determined using the SAFT technique. % Chromite was determined using UV/VIS spectroscopy. The three assay methods are presented in detail in Appendix I.

The precision of the assay techniques varied for different types of sample. Standard deviations for each assay technique were calculated from standard ranges[1] and are presented in Table 4.2.

Table 4.2: Standard deviations calculated for the different assay methods.

Type of Sample	Standard	deviations	(%)	Fire Assay	UV/VIS Spectroscopy
	SAFT				
	Pt	Pd	Rh	PGM	Chromite
Final concentrate	4.00	3.94	4.15	5	1.7
Flotation tails	5.08	5.88	6.67	10	1.4
Cyclone overflow	4.71	4.93	4.00	5	1.4

4.1.3 Batch flotation

The batch flotation technique used in this investigation is described in Section 3.3. The

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[1]The standard ranges for the different assay methods were obtained from the Impala Platinum Limited Mineral Processes Laboratory.

reproducibility of the batch flotation test procedure used in this investigation was tested using a 30 litre sample taken at the primary rougher tails. Three batch flotation tests were conducted on the sample. The results are shown in Figure 4.2.
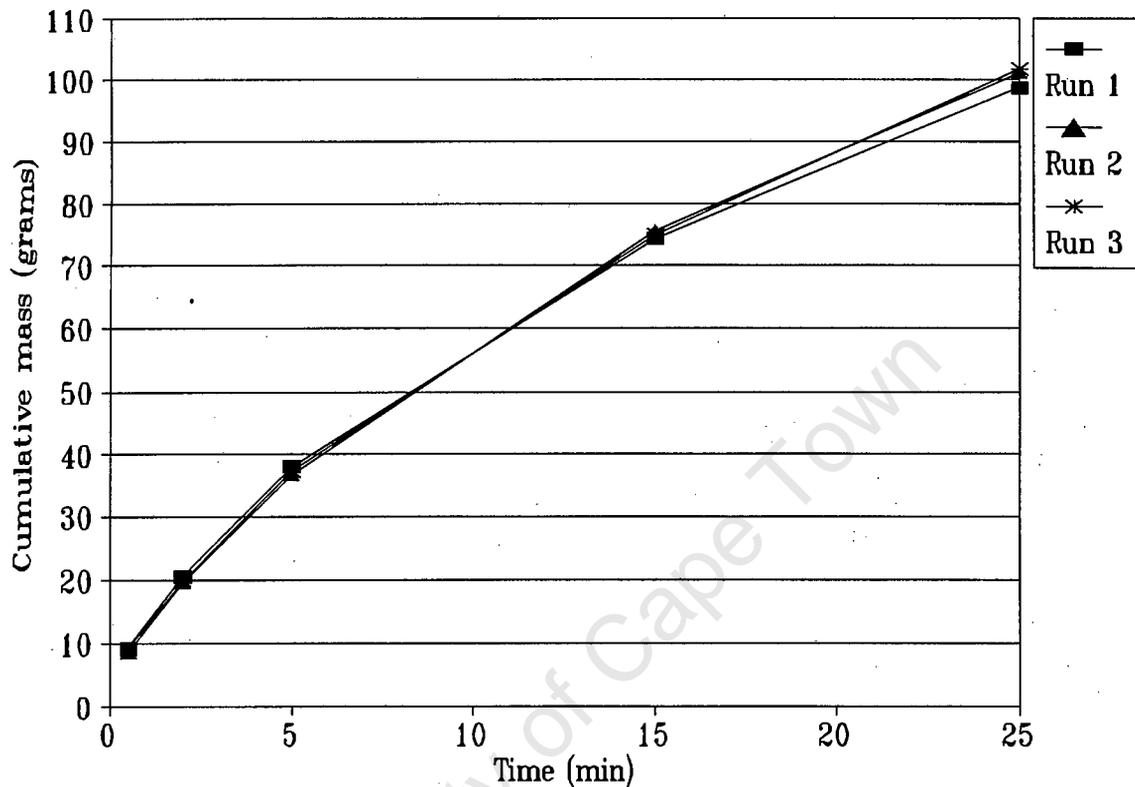
Figure 4.2: The mass of solids floated in the batch flotation tests done on Primary cleaner tails.

The sample standard deviation (STDS) and the standard error for each test are presented in Table 4.3.

The average standard error is 1.3% for the three batch flotation tests. Therefore, the reproducibility that was obtained using this batch flotation procedure is acceptable.

The recoveries of the batch flotation tests were calculated using a calculated head grade, obtained from the tailings assay result (Appendix F). If the calculated head grade and the assayed head grade resulted in a final recovery difference of 10% or more, the test was rejected.

Table 4.3: Sample standard deviations and standard errors

Time (min)	Sample standard deviation (STDS) (%)	Standard error (STE) (%)
0.5	3.8	2.2
2	3.1	1.8
5	1.7	1.0
15	0.7	0.4
25	1.7	1.0

A sensitivity analysis was carried out to determine the effect of a 10% change in the tailings assay results on the final recovery. The analysis was carried out on tailings results from floating primary rougher feed (0.6 - 0.8 g/t for platinum) and primary recleaner (column) feed (4 - 5 g/t for platinum). The 10% change in the tailing assay from floating primary rougher feed resulted in a 2% difference in the final recovery. The change in the tailings assay from floating primary recleaner feed resulted in 1% difference in the final recovery.

4.1.4 Residence time distribution studies

The test procedure discussed in Section 3.3 was used in all the residence time distribution tests. The test was done in triplicate on the primary rougher and primary column and in duplicate on the primary cleaner and secondary cleaner. Table 4.4 shows the calculated mean residence times.

Table 4.4: Mean residence times (τ) calculated from residence time distribution data.

Unit	Run 1 τ (min)	Run 2 τ (min)	Run 3 τ (min)
Primary rougher	34.63	33.97	33.15
Primary cleaner	8.55	8.65	
Primary column	44.23	37.97	42.84
Secondary cleaner	30.10	30.42	

The mean residence times for the primary rougher, primary cleaner and secondary cleaner

were almost the same for the different test runs. The difference between mean residence times of the first and second test runs for the primary column was more than 6 minutes. The performance of a column is very sensitive to changes in the feed flow rate which will affect the mixing in the unit. Since the primary column feed flow was very unstable during the investigation it was expected that a variation in the mean residence times would occur.

4.1.5 Mass balance campaigns

Three mass balance campaigns were conducted during a period when the operating philosophy, as explained in Chapter 2, resulted in flow rates that were more stable than those measured in the past. This operating philosophy could only be practised for a short period owing to too high a chromite level in the concentrate which created problems in the smelter. Thus the mass balances are "snap shot" type data and represent the circuit for only a short period of 6 hours. The flow rates around the primary rougher/scavenger cells are presented in Table 4.5.

Although reasonable reproducibility could be obtained for the three mass balances, this was not generally the case, owing to variations in the operating philosophies as a result of the excessive chromite levels in the final product going to the smelter.

Table 4.5: Mass flow rates (solids) around the primary rougher

Sampling location	Mass balance I Flow rates (t/hr)	Mass balance II Flow rates (t/hr)	Mass balance III Flow rates (t/hr)
Primary rougher feed	133.20	130.12	126.93
Primary rougher concentrate	1.80	2.16	1.66
Primary scavenger concentrate	3.60	3.52	3.98
Primary scavenger tails	128.43	124.42	121.07

4.2 Circuit evaluation: Phase 1 (Preliminary investigation)

The results obtained in this phase of the evaluation consist mainly of batch flotation results and unit recoveries. The calculations and assay results are presented in Appendices A and B. The circuit configuration and operation during the preliminary work were described in Section 2.2.

As explained in Section 3.3 the circuit configuration was different in this phase of the evaluation. The quest for optimal recoveries was the main reason for the changes that occurred in the circuit.

The PGM unit and batch flotation recoveries presented in this section were normalised for confidentiality reasons. A random value in each table was selected to normalise the data presented in that table. In certian cases where recoveries were compared between two tables the same normalisation value was used in both tables to enable comparison.

4.2.1 Unit recoveries

The recoveries were calculated using the two and three product formulae as described in Appendix B. The unit recoveries consist of PGM and chromite recoveries.

4.2.1.1 PGM recoveries

The unit performance or recovery was evaluated on three different days for each unit. The tests were not done in triplicate for reproducibility reasons but rather with the intention of evaluating the unit performance over a few days.

It is clear from Table 4.6 that the PGM recoveries obtained for the units are very consistent in some cases while variations occur for other cases. The variation in the PGM recoveries of the columns is a result of the high variability of the feed flow rates going into the columns.

Table 4.6: Normalised unit PGM recoveries

Unit	Run 1	Run 2	Run 3
Primary Section			
Rougher	1.49	1.49	1.49
Cleaner	1.76	1.75	1.78
Recleaner	1.73	1.72	1.80
Column	1.00	1.15	1.30
Secondary Section	1.15	0.94	0.79
Column		0.11	0.40

4.2.1.2 Chromite recoveries

The chromite recoveries are not consistent for the three sampling campaigns done on a unit (Table 4.7). The question of chromite being recovered through entrainment will be discussed in Chapter 6. The recovery depends on certain froth zone characteristics like water content, stability of the bubbles and froth depth. Variation of these characteristics will affect the consistency of the chromite recovery.

Table 4.7: % Unit chromite recoveries

Unit	Run 1	Run 2	Run 3
Primary Section			
Rougher	0.7	0.3	0.3
Cleaner	16.5	22.1	15.7
Recleaner	49.7	41.9	48.9
Column	23.3	18.1	43.4
Secondary Section	0.2	0.2	0.1
Column	6.9	2.4	

4.2.2 Batch flotation results

The batch flotation tests on a feed or a tailing stream were done in triplicate. Although the samples were taken at the same location for the three runs, they should not be the same. This is due to variations in flow rates, pulp densities and reagent dosages as well as ore characteristics. It must be re-iterated that the tests were not repeated for reproducibility reasons but rather to evaluate the variation in the floatability of a stream over a few days. The calculations and raw data collected in the batch flotation tests are listed in Appendix A. The results are discussed in the sections below.

4.2.2.1 PGM recoveries

The PGM recoveries listed in Table 4.8 are very consistent if the high variability of

Table 4.8: Summary of normalised PGM recoveries (batch flotation test work)

Sampling points for batch flotation test work	PGM recovery Run 1	PGM recovery Run 2	PGM recovery Run 3
Primary Section			
Rougher feed	1.30	1.41	1.36
Rougher tails	0.31	0.29	0.30
Cleaner feed	1.83	1.72	1.82
Cleaner tails	1.23	1.14	1.25
Recleaner feed	1.85	1.88	1.81
Recleaner tails	1.79	1.79	1.60
Column feed	1.72	1.79	1.81
Column tails	1.78	1.80	1.72
Secondary Section			
Rougher feed	1.24	1.02	1.09
Rougher tails	0.44	0.40	0.52
Column feed	1.69	1.80	
Column tails	1.66	1.57	

parameters in the circuit, which will have an effect on the floatability of the samples, are taken into account.

The batch flotation of the primary rougher feed gave PGM recoveries of around 1.35% while the plant unit performed at a 1.49% PGM recovery. In common practice it is usually found that the batch flotation results are higher than those obtained for the plant units. The feed samples for the batch flotation tests were therefore screened and the result was a 20% -75μ particle size distribution. The plant is operated at a 50% -75μ size distribution for the stream under investigation. It is obvious that the samples used in the batch flotation were not representative of the plant. Bias therefore occurred in the sampling process. The very high flow rates of the feed stream made it difficult to intersect the total cross section of the stream and resulted in biaser samples. Subsequent rougher feed samples were taken at the surge tank over flow where high flow rates were not a problem.

4.2.2.2 Chromite recoveries

The chromite recoveries obtained in the batch flotation tests are listed in Table 4.9. It is interesting to note that the recoveries increased as the sampling location moved to the later cleaning stages.

4.3 Circuit evaluation: Phase 2

The method applied in the evaluation of the circuit is described in Section 3.3. The difference between the second phase of the circuit evaluation and the first phase did not only involve the changes in the circuit configuration, but also included more specific assay results and water recovery data for the batch flotation tests as well as residence time distribution studies, "down-the-bank" sampling and mass balance campaigns.

4.3.1 Unit recoveries

The unit performances or recoveries were calculated by using the two and three product formulae as explained in Appendix H. The calculation of unit recoveries enabled an evaluation of the performance of the different unit operations in a certain configuration. The

Table 4.9: Summary of % chromite recoveries (batch flotation test work)

Sampling points for batch flotation test work	Run 1	Run 2	Run 3
Primary Section			
Rougher feed	0.5	0.7	0.5
Rougher tails	0.6	0.5	0.6
Cleaner feed	18.4	12.3	14.3
Cleaner tails	7.9	5.4	12.5
Recleaner feed	44.3	52.0	31.8
Recleaner tails	37.5	36.8	20.3
Column feed	48.5	48.1	58.1
Column tails	43.8	46.1	45.6
Secondary Section			
Rougher feed	2.5	3.3	1.8
Rougher tails	2.3	2.6	1.8
Column feed	39.1	57.1	
Column tails	34.2	32.5	

identification of problem areas in the circuit was also possible. Platinum, palladium, rhodium and chromite recoveries were determined and are presented in the same sequence.

The platinum, palladium and rhodium unit and batch flotation recoveries presented in this section were normalised for confidentiality reasons. A random value in each table was selected to normalise the data presented in that table. In certian cases where recoveries were compared between two tables the same normalisation value was used in both tables to enable comparison.

4.3.1.1 Platinum recoveries

The platinum recoveries of the units obtained during Phase 2 of the circuit evaluation are presented in Table 4.10. It is important to note that the primary and secondary columns gave

low platinum recoveries. Large variations in the platinum recoveries were observed for the different runs of the primary cleaner and secondary rougher.

Table 4.10: Normalised platinum recoveries

Unit	Platinum recovery Run 1	Platinum recovery Run 2	Platinum recovery Run 3
Primary section			
Rougher	3.73	3.69	3.66
Cleaner	4.19	3.39	4.12
Column	1.95	3.05	1.92
Secondary section			
Rougher	2.46	1.88	1.61
Cleaner	3.92	3.85	3.72
Recleaner	4.03	4.17	3.92
Column	1.00	1.32	1.15

4.3.1.2 Palladium recoveries

The palladium recoveries of the different units as obtained in Phase 2 of the evaluation of the circuit are presented in Table 4.11. The same low column recoveries, as observed for platinum, were observed for palladium and the recoveries were in the same order as the platinum recoveries. Large variations in the palladium recoveries for the different runs of the primary cleaner and the secondary rougher were observed.

4.3.1.3 Rhodium recoveries

The rhodium recoveries, as calculated during Phase 2 of the circuit evaluation, for the different unit operations are presented in Table 4.12. The observation about the low column recoveries that was made for platinum and palladium was also observed for rhodium. The

Table 4.11: Normalised palladium recoveries

Unit	Palladium recovery Run 1	Palladium recovery Run 2	Palladium recovery Run 3
Primary section			
Rougher	1.84	1.78	1.81
Cleaner	2.18	1.81	2.18
Column	0.95	1.44	1.00
Secondary section			
Rougher	1.38	0.88	0.77
Cleaner	1.95	2.04	1.84
Recleaner	2.01	2.08	2.09
Column	0.57	0.61	0.61

Table 4.12: Normalised rhodium recoveries

Unit	Rhodium recovery Run 1	Rhodium recovery Run 2	Rhodium recovery Run 3
Primary section			
Rougher	2.07	2.02	2.08
Cleaner	2.63	2.09	2.52
Column	1.10	1.68	1.13
Secondary section			
Rougher	1.56	1.00	1.19
Cleaner	2.44	2.23	2.55
Recleaner	2.37	2.56	2.44
Column	0.67	0.58	0.80

large deviations in rhodium recoveries for the different runs of the primary cleaner and secondary rougher were also observed. The unit recoveries for platinum, palladium and rhodium for each unit were all of the same order.

4.3.1.4 Chromite recoveries

The performance of each unit not only depends on the recovery of the different platinum group metals but also on the unit's ability to give low chromite recoveries. The % chromite recoveries, as calculated for Phase 2 of the circuit evaluation, are presented in Table 4.13. The chromite recoveries of the cleaners and recleaners were substantially higher than for the rougher units. This was expected since the chromite content of the roughers' feed (15 - 18 %) was much higher than for the feed to the cleaners (2 - 6 %). The common expectation was that the chromite recoveries would increase while the chromite grade in the concentrate would decrease with each cleaning stage. This was not the case for the columns since the chromite recoveries of the columns were lower than for the cleaners before them. This was due to the washing of the froth in the columns which is a main feature of these units.

Table 4.13: Chromite recoveries

Unit	%Chromite recovery Run 1	%Chromite recovery Run 2	%Chromite recovery Run 3
Primary section			
Rougher	1.5	1.7	2.3
Cleaner	47.9	22.9	42.3
Column	36.2	18.2	35.7
Secondary section			
Rougher	1.8	1.3	1.5
Cleaner	47.3	26.2	30.7
Recleaner	48.7	50.5	70.4
Column	23.5	12.0	65.9

4.3.2 Batch flotation data

The recoveries listed in section 4.3.2.1 to 4.3.2.4 were calculated by using a calculated feed grade as explained in Appendix F. The grades and mass distributions of the different batch flotation tests are listed in Appendix F. The batch flotation procedure used in the test work is described in Section 3.3.1. The main differences between the batch flotation results obtained from the Phase 2 test work compared to Phase 1 lie in the PGM assay results. The Phase 2 tests were assayed for specific platinum-group elements, platinum, palladium and rhodium while the Phase 1 results only gave a PGM result.

4.3.2.1 Platinum recoveries

The platinum recoveries for each batch flotation run carried out during Phase 2 of the evaluation are presented in Table 4.14. It is very interesting to note that the recoveries of the batch flotation tests carried out on the different feed samples of the units were almost the

Table 4.14: Batch flotation results: Normalised platinum recoveries

Sampling location	Platinum recovery Run 1	Platinum recovery Run 2	Platinum recovery Run 3
Primary section			
Rougher feed	3.75	3.73	3.65
Cleaner feed	4.27	3.77	4.27
Column feed	4.31	4.05	4.11
Secondary section			
Rougher feed	1.80	2.02	1.82
Rougher tails	0.42	0.71	0.51
Cleaner feed	3.60	3.59	3.59
Recleaner feed	3.95	4.23	3.99
Column feed	3.78	4.02	3.88

same as the platinum recoveries obtained for themselves (Table 4.10), except for the batch

flotation test results of the samples taken at the feed to the columns. The platinum recoveries for the batch flotation tests carried out on the feed samples of the columns were much higher than the platinum recoveries obtained from the columns.

4.3.2.2 Palladium recoveries

The palladium recoveries for each batch flotation run carried out during Phase 2 of the evaluation are presented in Table 4.15. The same observation made about the platinum recoveries of the batch flotation tests carried out on the different feed samples of the units that were almost the same as the platinum recoveries obtained for the units, was also made for palladium. The batch flotation tests results of the samples taken at the feed to the columns were again an exception. The palladium recoveries for the batch flotation tests carried out on the feed samples of the columns were much higher than the palladium recoveries obtained from the columns themselves.

Table 4.15: Batch flotation: Normalised palladium recoveries

Sampling location	Palladium recovery Run 1	Palladium recovery Run 2	Palladium recovery Run 3
Primary section			
Rougher feed	1.84	1.85	1.78
Cleaner feed	2.19	1.95	2.24
Column feed	2.10	2.00	2.10
Secondary section			
Rougher feed	0.86	0.88	0.81
Rougher tails	0.32	0.29	0.31
Cleaner feed	1.70	1.80	1.74
Recleaner feed	1.97	2.22	1.99
Column feed	1.98	2.18	2.11

4.3.2.3 Rhodium recoveries

The rhodium recoveries for each batch flotation run carried out during Phase 2 of the evaluation are presented in Table 4.16. The same observation made about the platinum and palladium recoveries of the batch flotation tests carried out on the different feed samples of the units that were almost the same as the platinum and palladium recoveries obtained for the units, was also made for rhodium. Again the columns were the exception and the rhodium recoveries for the batch flotation tests carried out on the feed samples of the columns were much higher than the palladium recoveries obtained from the columns.

Table 4.16: Batch flotation: Normalised rhodium recoveries

Sampling location	Rhodium recovery Run 1	Rhodium recovery Run 2	Rhodium recovery Run 3
Primary section			
Rougher feed	2.09	2.07	2.06
Cleaner feed	2.61	2.27	2.64
Column feed	2.54	2.40	2.43
Secondary section			
Rougher feed	0.94	0.88	1.06
Rougher tails	0.35	0.27	0.36
Cleaner feed	2.05	2.20	2.23
Recleaner feed	2.38	2.58	2.34
Column feed	2.27	2.21	2.27

4.3.2.4 Chromite recoveries

The chromite recoveries for each batch flotation run carried out during Phase 2 of the evaluation are presented in Table 4.17. The chromite recoveries obtained in the batch flotation test work differ depending on the sampling location. As expected the chromite recoveries became higher for samples taken at cleaning stages that followed each other since the amount of chromite to be rejected became less and less.

Table 4.17: Batch flotation: Chromite recoveries

Sampling location	%Chromite recovery Run 1	%Chromite recovery Run 2	%Chromite recovery Run 3
Primary section			
Rougher feed	1.2	1.0	1.4
Cleaner feed	39.8	17.2	35.1
Column feed	35.3	24.2	38.5
Secondary section			
Rougher feed	1.6	1.2	1.7
Rougher tails	1.4	1.2	1.5
Cleaner feed	17.3	20.5	22.3
Recleaner feed	36.7	42.6	38.4
Column feed	33.7	52.2	50.2

4.3.2.5 Water recoveries

Table 4.18: Batch flotation: Water recoveries

Sampling location	%Water recovery Run 1	%Water recovery Run 2	%Water recovery Run 3
Primary section			
Rougher feed	32.9	29.1	28.6
Cleaner feed	26.5	29.4	28.6
Column feed	32.2	30.2	28.7
Secondary section			
Rougher feed	32.7	26.1	31.3
Rougher tails	27.3	32.1	30.5
Cleaner feed	28.5	29.9	26.7
Recleaner feed	25.8	27.2	26.3
Column feed	20.3	24.9	24.6

The water recoveries measured for each batch flotation run carried out during Phase 2 of the evaluation are presented in Table 4.18. Every wet and dry sample mass was recorded. Therefore, the mass of the water present in each sample was calculated and the recoveries were determined. Although the amount of water in the samples differs for each sampling location, the water recoveries in the batch flotation cell were always in the range of 20 - 30%.

4.3.3 Residence time distribution studies

Residence time distribution studies have been performed on each type of flotation machine in the circuit. Studies were done on a Wemco rougher/scavenger, a Wemco cleaner, a Dorr Oliver cleaner and a flotation column. The data in its raw form is presented in Appendix D. The main objectives of the RTD studies were to determine the real residence times in the different vessels and to aid in the evaluation and characterisation of the units.

4.3.3.1 The Primary rougher/scavenger

The tracer tests for this unit were done in triplicate to test for reproducibility. From Figure 4.3 it is evident that the tracer tests were reproducible. The mean residence time, τ, which is also the first moment of the RTD curve was calculated to be 32 minutes.

The nominal residence time, which is 35 minutes, was calculated from the mass balances presented in section 4.3.5. The test procedure used, are therefore shown to be acceptable in that τ_{RTD} must be $\leq \tau_{nominal}$ since effective vessel volume must be $\leq$ actual volume at constant throughput.
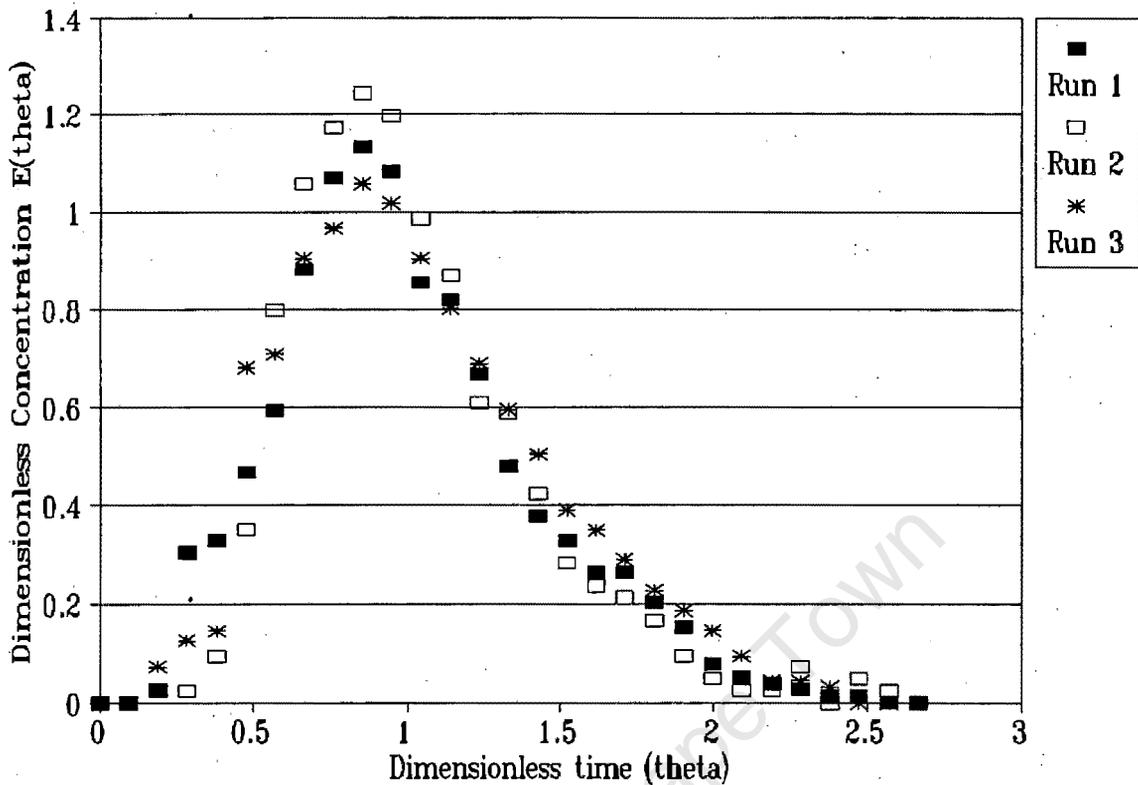
Figure 4.3: The RTD curve of the primary rougher/scavenger (Wemco).

4.3.3.2 Primary cleaner (Wemco)

The short residence time in the cleaner made it difficult to obtain enough points for a proper curve since discrete monitoring of the tracer was applied. However, two runs were conducted and reasonable reproducibility was obtained (Figure 4.4). The mean residence time was 8½ minutes while the nominal residence time was 11½ minutes.

4.3.3.3 Primary column

Three tracer tests were conducted on the Primary column and the results are presented on Figure 4.5.
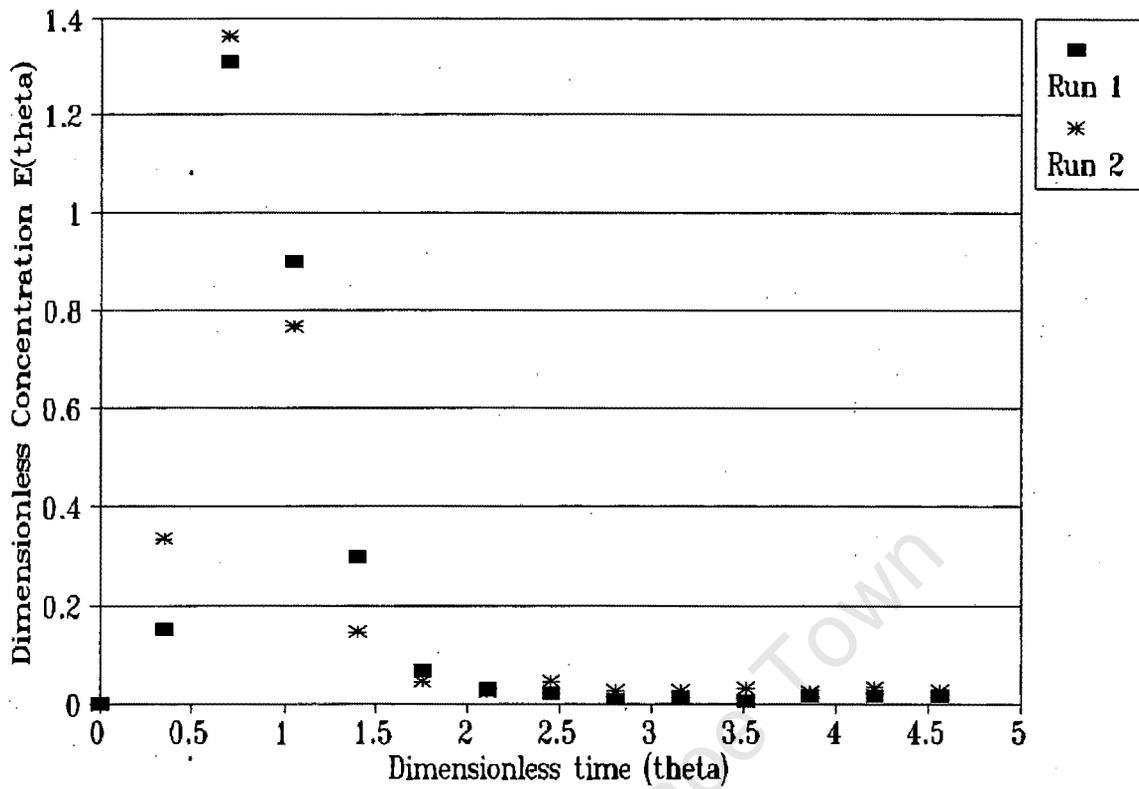
Figure 4.4: The RTD data of the primary cleaner (Wemco).
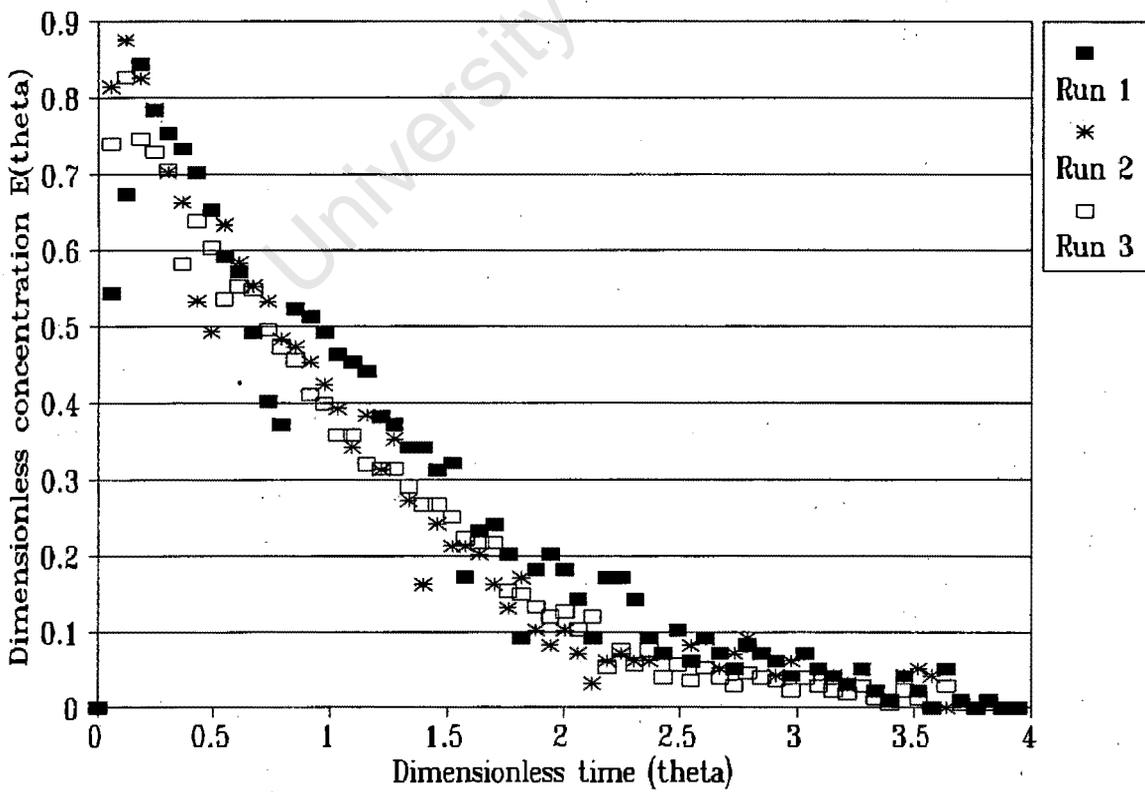
Figure 4.5: The RTD data for the primary column

Due to very unstable pulp flow rates into the column for run 1 and run 2, during the period of the RTD studies, it was observed that notable differences in the calculated mean residence times and variances (Table 4.19) occur. The effect of the unstable operation can also be seen on Figure 4.5 for run 1 and run 2 where the data follows an oscillation pattern which indicates internal recirculation (Levenspiel, 1972).

Table 4.19: First two moments of RTD curve

Run	Mean residence time (min)	Variance (min^2)
1	46	1330
2	44	1337
3	47	1475

The nominal residence time of ±48 minutes was calculated from the mass balances in Section 4.3.5. A gas holdup of 10% was assumed and a tailings pulp flow rate of 2.85×10^{-1} m^3/min was used for the calculations. The tailings flow rate was used since wash water was also added and had to be taken into account.

4.3.3.4 Secondary cleaner (Dorr Oliver)

Since this unit proves to have a stable operation only two runs were conducted in the RTD studies. Figure 4.6 shows that the reproducibility was excellent. The mean residence time was calculated as 30 minutes while the nominal residence time calculated from the mass balances was also 30 minutes.

4.3.4 "Down-the-bank" sampling

The "down-the-bank" sampling results, conducted on the primary rougher, are presented in Figure 4.7. The calculation of the results as well as the data collected during the sampling campaign can be found in Appendix E. The recoveries presented in Figure 4.7 were normalised for confidentiality reasons. The same value that was used in the normalisation of the unit recoveries were used to normalise the recoveries presented in Figure 4.7 to enable
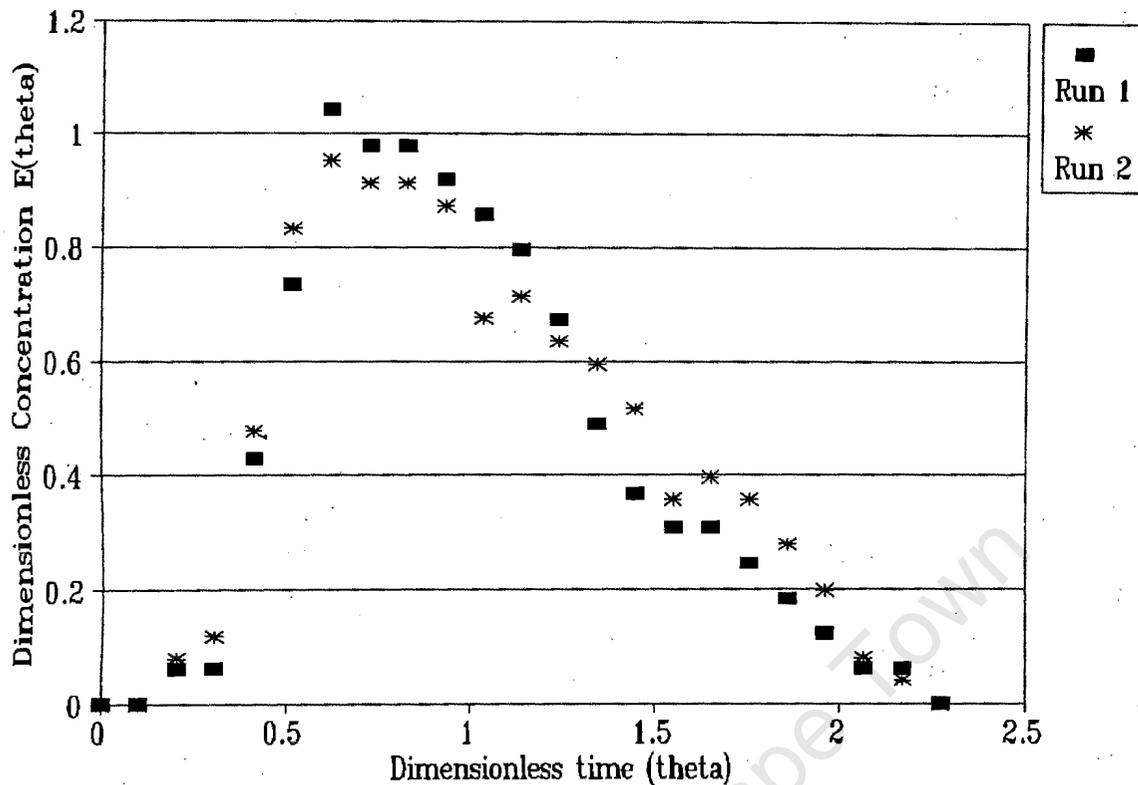
Figure 4.6: The RTD data for the secondary cleaner (Dorr Oliver)

a comparison between the data.

The generation of a recovery versus time curve for the plant unit was only possible with the assumption that the residence times in the different cells are all equal.

Figure 4.7 shows the profile obtained for platinum, palladium and rhodium. The platinum recovery for this unit was calculated as 3.84% while the unit evaluation gave unit recoveries of around 3.7%. The 3% difference can be associated with sampling errors since difficulties, such as inconsistent pulp levels, were experienced during the sampling of this unit. The same differences were calculated for palladium and rhodium.

The other important results obtained from this campaign are the chromite and water recoveries. The linear relationship between chromite recovery and water recovery (Figure 4.8) would seem to imply that chromite was recovered mainly through entrainment (Lynch *et al.*, 1981).

4.3.5 Mass balance data

The mass balance campaigns were preceded by a planning stage where the number of sampling points were calculated by doing a nodal analysis on the circuit layout as explained in section 3.3.4. From this analysis it was clear that 21 sampling locations and five measured flow rates are needed to calculate a unique solution for a mass balance of the circuit.

Three mass balance campaigns were conducted each a week apart. The mass balance data was used to evaluate and model the circuit. The mass balances were conducted in a period when the operating philosophy, as discussed in section 2.6, enabled more constant flow rates.

A computer program, Micbal version 1.0, was used for the mass balance calculations. This simplified the task of finding solutions for the unknown variables in the mass balance problem. Smoothing of some measured flow rates and assay values did occur. Sampling errors as well as errors in assay values made it impossible to solve the unknown variables without smoothing certain results. The program enables the smoothing of the results through selected standard deviations on the assay values. If contamination of the sample occurred during the sampling, or any other problems usually associated with sampling that would reduce the confidence in the result, a standard deviation could be associated with the assay value and make smoothing possible. The plant experience of the user enables the smoothing of the values to realistic numbers in order to gain a solution for the mass balance.

The results of the first mass balance campaign are presented in Figure C.1 and are referred to as Mass Balance I. Mass Balance II is presented in Figure C.2 and Mass balance III in Figure C.3 (Figures C1, C2 and C3 are presented in Appendix C). The mass flow rates, platinum (Pt) flow rates and grades and the % chomite (Cr_2O_3) are shown in the mass balance figures. The palladium and rhodium balances as well as the detailed chromite balances are given in Appendix C.

Nominal residence times, viz. unit volume/volumetric flow rate, were also calculated from the mass balance data (Table 4.20). The residence times for the columns were calculated by assuming a 10% gas holdup. Since the column feed flow rate was very inconsistent, the

tailings flow rate was used for the calculations.

Table 4.20: Nominal residence times

Unit	Nominal residence time (min)
Primary section	
Rougher	35
Cleaner	12
Column	48
Secondary section	
Rougher	32
Cleaner	30
Recleaner	15
Column	50

Chapter five

Modelling

In the previous chapter the results obtained in the evaluation of the circuit were presented. This chapter deals with the modelling of some of the results. The aim of the model is to give a steady state indication of the performance of the circuit. In order to reach this goal a methodology was developed to enable the modelling of the unit performances and the simulation of the circuit performance.

The methodology is presented in Figure 5.1 which shows that the results from the circuit evaluation can be divided into unit recoveries, batch flotation results and RTD studies. The modelling process started off with the kinetic modelling of the batch flotation results from which rate constants and ore floatability characteristics were calculated. The RTD studies were modelled using the tanks-in-series and axial dispersion models from which mixing parameters were derived.
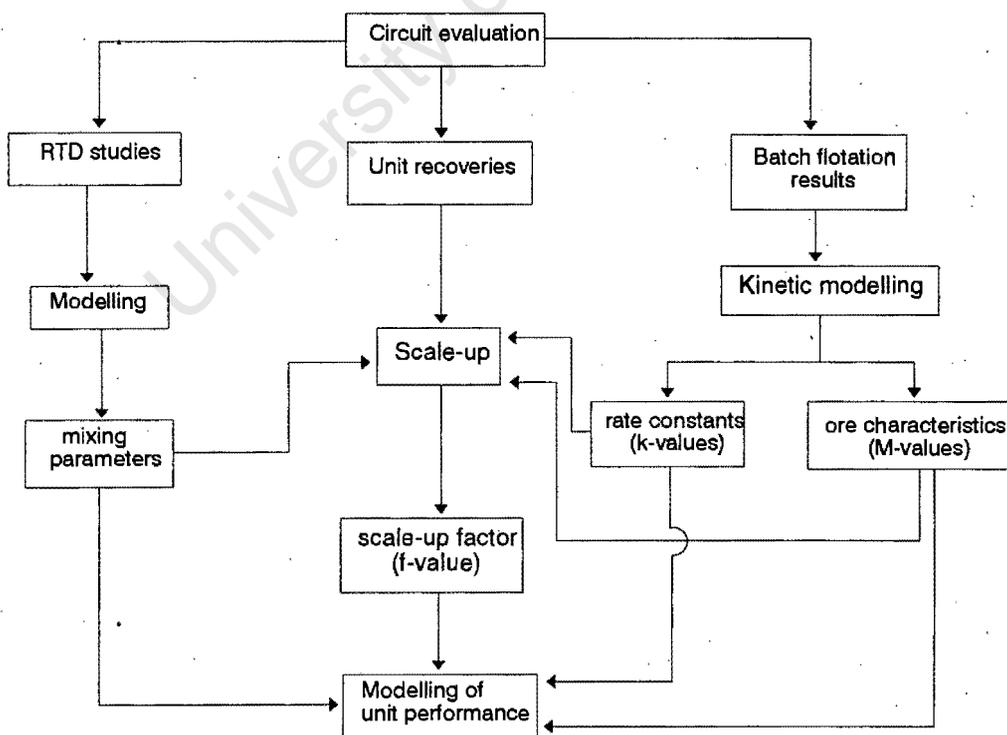
Figure 5.1: The methodology for the modelling of the unit performances.

The ratio of the plant rate constants to the batch flotation rate constants could then be determined, using the measured plant unit recoveries, and the ore characteristics determined in the batch tests. This scale-up method was tested using the data from the "down-the-bank sampling done on the Primary rougher. Applying the scale-up factor, the batch rate constants, the ore characteristics and the mixing parameters made it possible to model the performance of the unit operations.

Since the performance of each of the units could be modelled, the simulation of the steady state performance of the circuit was possible. The results of the circuit simulation were then compared to the circuit mass balance data, as well as historical plant data. Previous configurations of the circuit were also modelled, and the results compared to historical plant data.

5.1 Modelling of the batch flotation data

In the literature it has been assumed by many researchers (Woodburn and Loveday, 1965; Klimpel, 1980; Lynch, 1981) that flotation is a first order process, provided that the heterogeneous characteristics of the mineral particles in the ore is accounted for. While particle size is generally sufficient to adequately account for differences in particle floatability when modelling flotation as a first order process, even this simplification can nonetheless represent considerable difficulties in the analysis of complex, multi-mineral ores, such as platinum ores.

The batch flotation data was therefore modelled using the floatability parameter model. This model takes all the factors that influence floatability into account in a floatability spectra. The general form of the model is as follows:

$$R = \sum_{i=1}^{n} M_i (1 - e^{-k_i t})$$

where R is the mineral recovery, n is the number of floatability classes, M_i is the fraction associated with floatability class i and k_i is the characteristic rate constant of floatability class i.

The number of floatability classes (n) was calculated for the different batch flotation tests and it was found that all the batch flotation tests could be modelled using three floatability classes ($n = 3$). The general equation therefore becomes,

$$R = M_f(1 - e^{-k_f t}) + M_m(1 - e^{-k_m t}) + M_s(1 - e^{-k_s t})$$

where the three floatability classes are referred to as a fast floating, medium floating and a slow floating class. The equation has the mass balance constraint that

$$M_f + M_m + M_s + M_o = 1$$

where M_f - fraction of fast floating minerals

   M_m - fraction of medium floating minerals

   M_s - fraction of slow floating minerals

   M_o - fraction of valuable minerals not floatable (us. mineral "locked" in gangue)

   k_f, k_m, k_s - rate constants of the different floatability classes.

The application of this model to the flotation data in this study assumes that the rate constants associated with each floatability class remain unchanged in all the batch flotation tests. Only the proportion of each floatability class (M-values) varies depending on the sampling location in the circuit. The rate constants are therefore not a function of time or sampling location.

The UG-2 ore body consists of numerous different kinds of minerals of which chalcopyrite, pyrrhotite and pentlandite are the three main minerals present. This suggests that the use of 3 floatability classes to model the flotation data is reasonable.

The equation parameters were calculated using a Fortran program which solves nonlinear equations with linear variables. The program uses a powerful optimisation procedure called NPSOL. The program is listed in Appendix G. Parameters for the Phase 1 and Phase 2 batch flotation tests were calculated and the rate constants are presented in Table 5.1. The M-values are listed in Appendix G. The Phase 1 k-values and M-values were calculated from PGM assay results while the Phase 2 parameters were calculated from Pt, Pd and Rh results. The Phase 1 parameters were therefore calculated from 30 data sets and the Phase 2 parameters from 72 data sets. Each data set consists of 5 data points.

Table 5.1: Rate constants for batch flotation tests.

Floatability class	Phase 1 k-values (min^{-1})	Phase 2 k-values (min^{-1})
fast floating	8.2	1.90
medium floating	0.76	0.55
slow floating	0.078	0.072

The biggest criticism facing this modelling process will probably be the number of variables used to fit the data. However, for $n < 3$ the fit obtained with the observed values was very poor, and completely inadequate, while for $n = 3$, an excellent fit to all the data was obtained. The same three rate constants could be used to describe batch tests conducted on circuit streams as diverse as plant feed and scavenger tails.
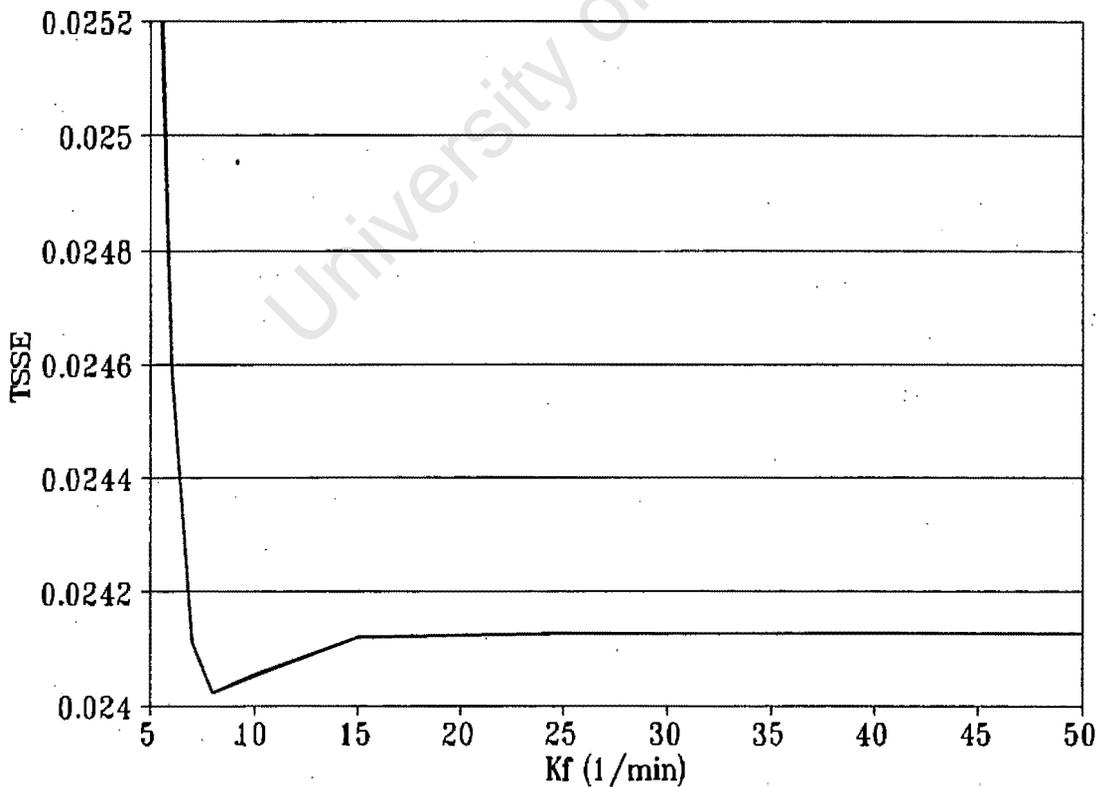
Figure 5.2: The effect of k_f on the total sum of the squares of the errors (TSSE).

The nonlinearity of the equation used as well as the number of variables, made it essential to test if the determined rate constant values were unique. The rate constants obtained in the modelling of the Phase 1 batch flotation data will be used as an example.

It was observed that the k_f could assume any value above 8 with almost no effect on the final result. Figure 5.2 shows clearly that an optimum value for k_f exists although the value of k_f does not have a big influence on the final result over a fairly wide range. However, it was found that if k_f was left out of the equation, the first few points on the recovery time profile of many of the tests could not be fitted. It was therefore necessary to have a fast floating component.

It was found that the k_m and k_s values had the biggest effect on the final result obtained from the model. It is clear from Figure 5.3 that optimum values for k_m and k_s do exist and the values are unique since the surface plot is convex. The minimum TSSE value was calculated
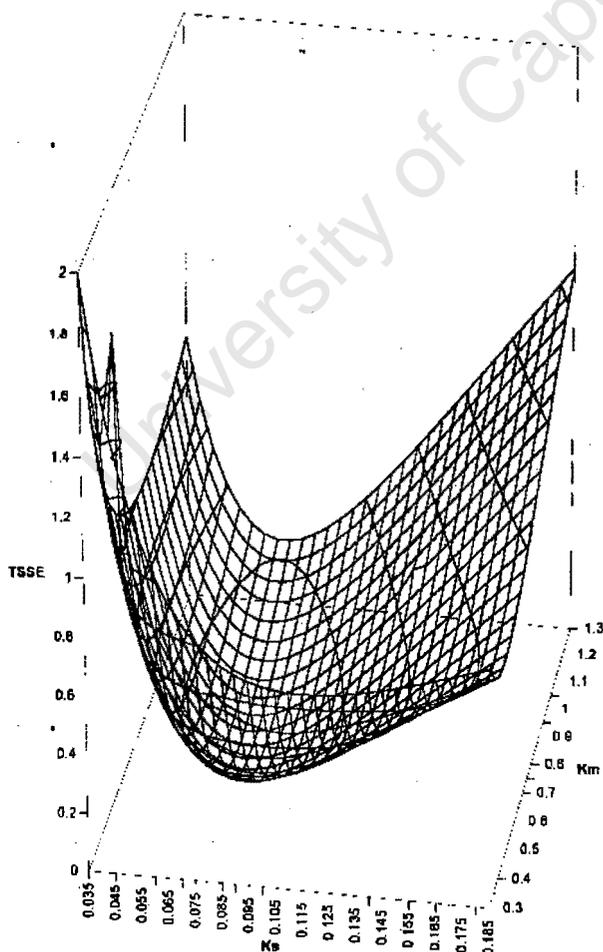
Figure 5.3: The effect of k_m and k_s on the total sum of the squares of the errors (TSSE).

as 0.024 and this represents an average "deviation" per point of 0.68%.

If the error associated with a batch flotation test is taken as 2.5% then the value of k_m would be in a range of 0.76±0.26 and k_s in a range of 0.078±0.02 to fall within an average deviation per point of 2.5%.

5.2 Modelling of RTD data

Two methods were used to model the RTD data of each of the units on which tracer tests were performed, namely the tanks-in-series model and the axial dispersion model. The theory underlying the two models was discussed in Section 1.4.

The tanks-in-series model was used to describe the mixing behaviour of the flotation banks since the number of tanks in series (N) obtained from the model could be physically related to the number of cells in series in a flotation bank. The axial dispersion model was used for comparison reasons to verify the results obtained from the tanks-in-series model.

5.2.1 Primary rougher

The tanks-in-series model was fitted to the data by using a least squares method. The best fit, as shown in Figure 5.4, was obtained with 9 tanks in series. The Wemco rougher/scavenger bank consists of 12 cells. The reason why the bank acts only as 9 tanks in series can be accounted for by the design of the flotation bank. Each set of four cells has an "open flow" configuration, so that there are three "tanks" with four agitators in each tank.

The solution to the axial dispersion model was numerically solved using the method of Manqiu Xu *et al.* (1991). The boundary conditions were those of a closed vessel. A least squares method was used to fit the model to the experimental data (Figure 5.4). The vessel dispersion number (N_d) was calculated as 0.07.
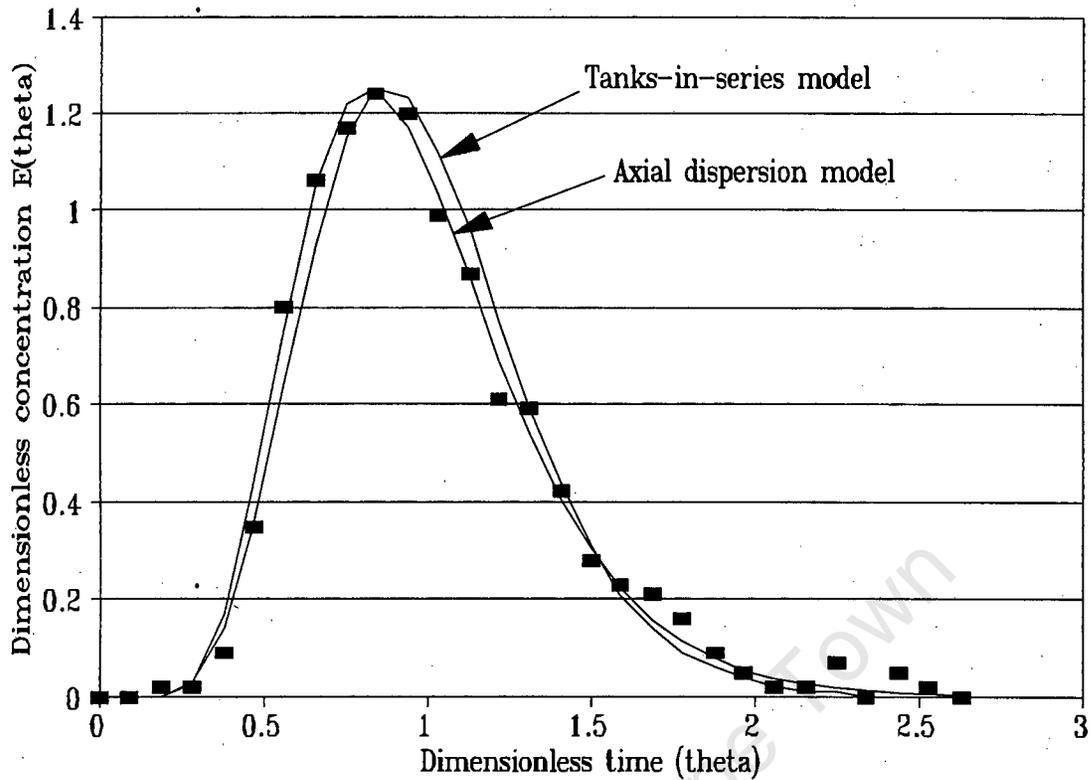
Figure 5.4: The fitting of the axial dispersion model and the tanks-in-series model to the RTD data of the primary rougher/scavenger.

5.2.2 Primary cleaner

The best fit obtained with the tanks-in-series model is presented in Figure 5.5 where N was found to be 13. Since the unit consists of 12 cells one would expect plug flow condition as the tanks-in-series model indicates.

The plug flow characteristics are confirmed by the axial dispersion model where $N_d = 0.06$.

5.2.3 Primary column

The tanks-in-series model was fitted to the RTD data and the best fit was obtained if $N = 1.08$ (Figure 5.6). Therefore, the column acts as a perfectly mixed tank reactor. Mills (1991) observed this for large diameter ($>$ 1 metre) columns.
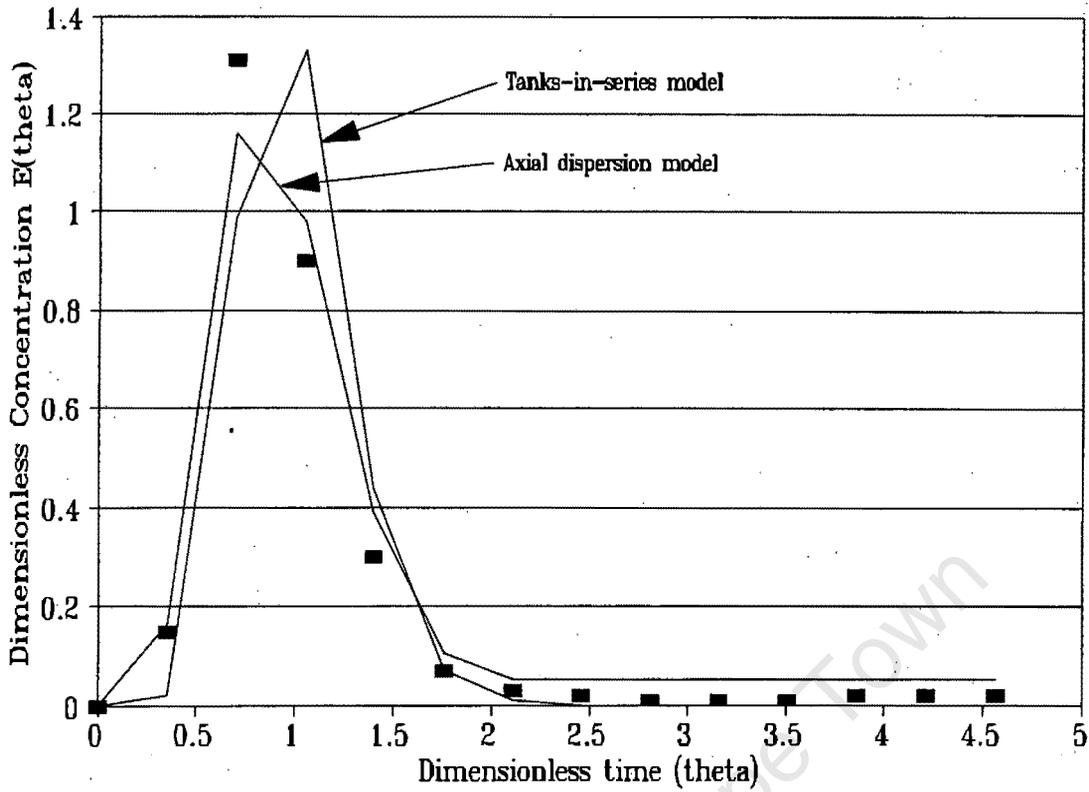
Figure 5.5: The axial dispersion model and tanks-in-series model fits to the primary cleaner RTD data.
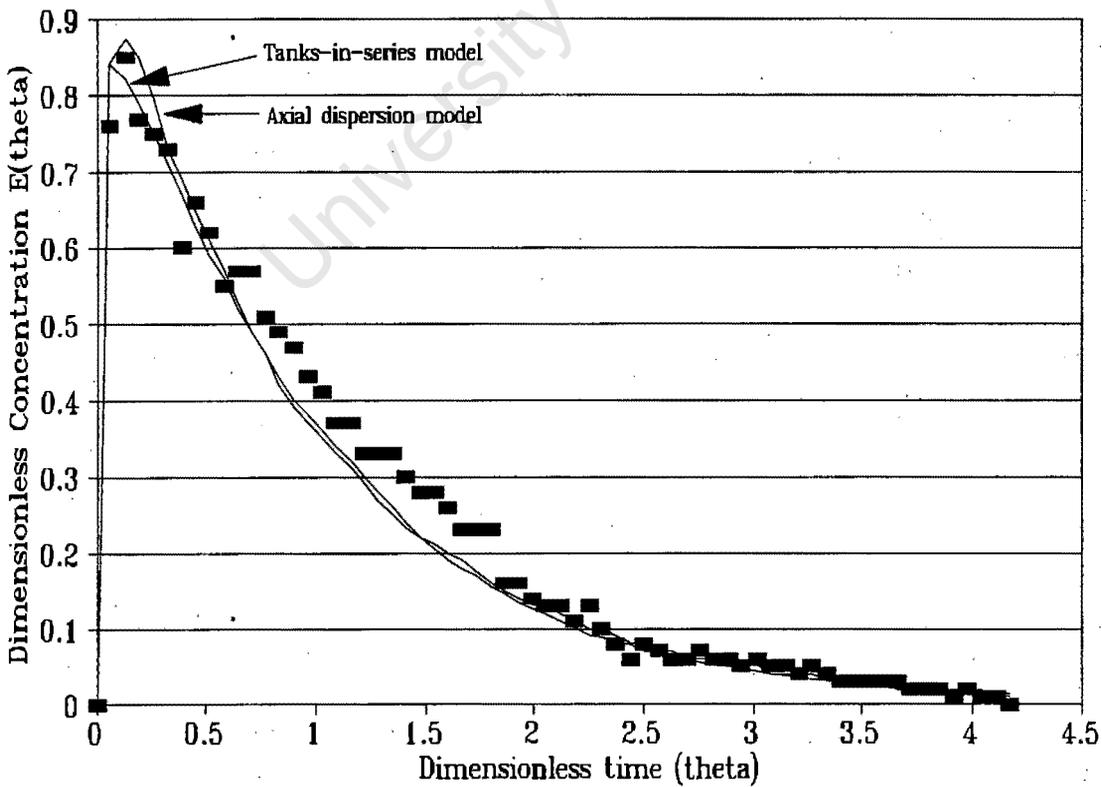
Figure 5.6: The axial dispersion model and tanks-in-series model fits to the RTD data of the primary column

The axial dispersion model produced a N_d of 1.8 which is a high number for the dispersion model (Levenspiel, 1972). Although the fit is not perfect (Figure 5.6) the result still confirms that the fluid in the column can be characterised by mixed flow.

5.2.4 Secondary cleaner

The best tanks-in-series model fit was obtained using a $N=4$ (Figure 5.7). This corresponds with the hydrodynamics of the unit, which consists of 4 cells.

The axial dispersion model was fitted to the data, as presented in Figure 5.7, and a $N_d =0.12$ was obtained. The low value of N_d indicates plug flow rather than mixed flow.
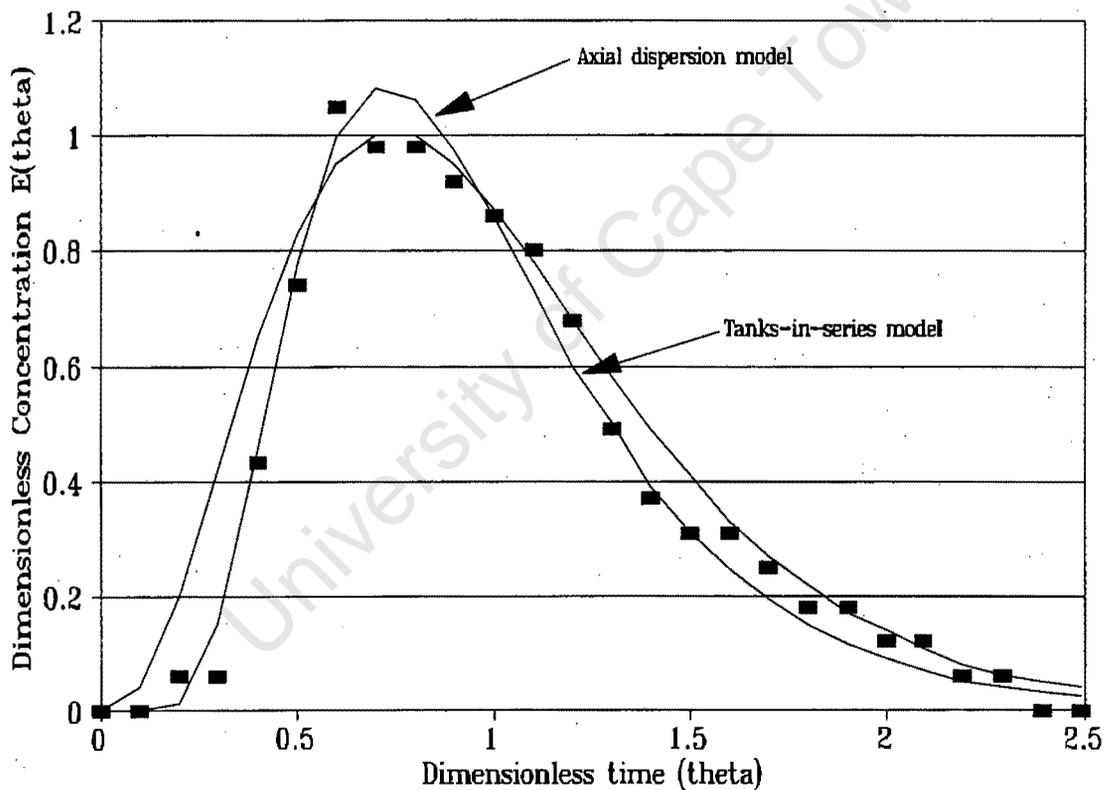
**Figure 5.7**: The axial dispersion model and tanks-in-series model fitted to the RTD data obtained from the secondary cleaner (Dorr Oliver).

5.3 Modelling of the unit recoveries

The difference between a batch and a continuous process is that for the continuous case the

residence time of all the elements in the system is distributed. The model used for the batch process was extended by incorporating the residence time distribution of the particles. The recovery is therefore a function of

$$R = f(M_i, k_i, \tau, N)$$

where M_i is the fraction associated with a floatability class, k_i is the flotation rate constant of floatability class i, τ is the residence time of the unit and N is number of tanks in series.

A general equation for mineral recovery in the flotation bank can be written as,

$$R = \sum_{i=1}^{n} M_i \int_0^{\infty} (1 - e^{-k_i t}) E(t) \, dt$$

Since it was established that $n=3$ and using the tanks-in-series model for N tanks in series, the equation becomes,

$$R = M_f \left(1 - \frac{1}{(1 + k_f \frac{\tau}{N})^N}\right) + M_m \left(1 - \frac{1}{(1 + k_m \frac{\tau}{N})^N}\right) + M_s \left(1 - \frac{1}{(1 + k_s \frac{\tau}{N})^N}\right)$$

5.3.1 Scale-up of the rate constants

The assumption that only residence time distribution is the difference between batch flotation and industrial units an over-simplification. In practice bubble loading, froth depth, frother concentration, bubble size, impeller efficiency, etc. have an impact. For normal operating conditions it may be assumed that these variables influence the rate of recovery of all particles to the same extent (Loveday *et al.*,1972; Woodburn, *et al.*, 1976).

Therefore, the rate constants used in the batch flotation modelling had to be scaled up for industrial flotation machines. For the scale-up process the assumption was made that the ratios between k_f and k_m, and k_f and k_s will stay the same in an industrial flotation cell. Therefore a scale-up factor (f) had to be found for each different flotation machine. Since the unit recoveries were known as well as the ore characteristics (M-values) of the feed stream to each unit a scale-up factor could be calculated.

The three batch flotation rate constants were divided by a scale-up factor (f) and the resulting rate constants were used in the model, with the mixing parameters (τ and N) (Table 5.2) and ore characteristics obtained from the modelling of the batch data.

Table 5.2: Mixing parameters used in the modelling

Unit	τ (min)	N
Primary rougher	32.00	9
Primary cleaner	8.65	13
Primary column	46.00	1
Secondary cleaner	30.00	4

The scale-up factor, f, was derived using the iterative procedure illustrated in Figure 5.8.
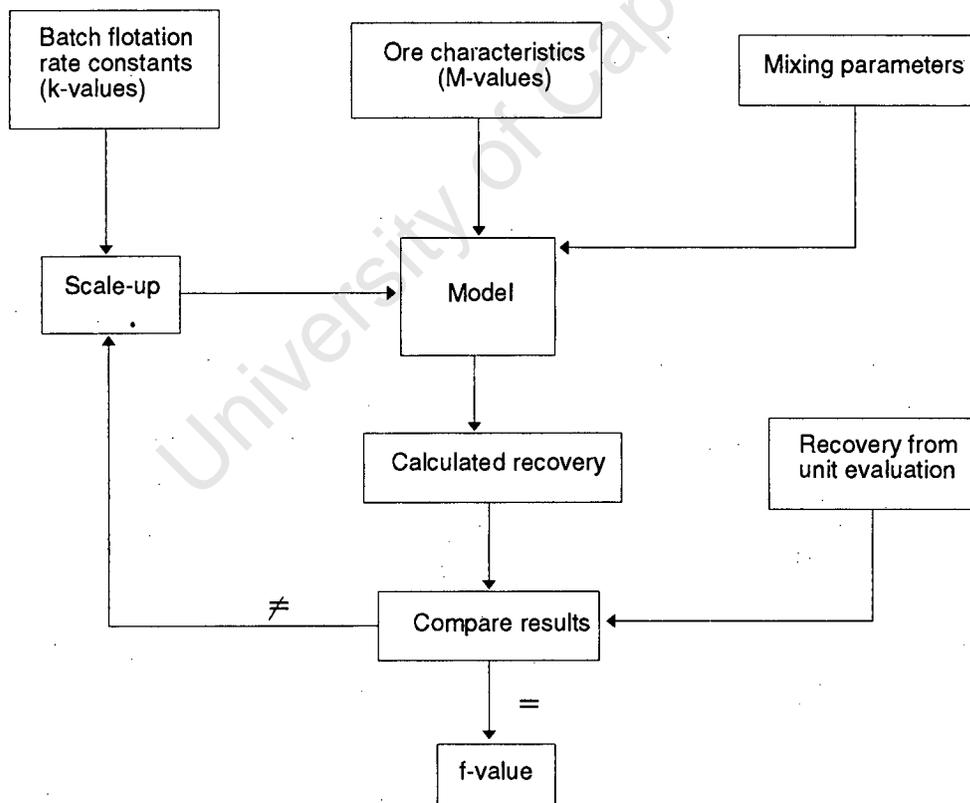
Figure 5.8: Methodology used for the scale-up of the batch flotation rate constants.

The scale-up factors (f) calculated for each unit are presented in Table 5.3.

It was assumed that the residence time in the different cells in a flotation bank stays constant. This assumption made it possible to create a recovery versus time profile for each unit. The scale-up of the rate constants was done with only the one point on the recovery versus time curve for the plant unit. After scale-up the model could be used to predict the performance of the plant unit at different residence times.

Table 5.3: Scale-up factors

Unit	Unit type	Scale-up factor (f)
Primary rougher	Wemco	1.35
Primary cleaner	Wemco	0.36
Primary column	Cominco	6.39
Secondary rougher	Wemco	1.35
Secondary cleaner	Dorr Oliver	0.61
Secondary recleaner	Wemco	0.36
Secondary column	Cominco	30.18

A "down-the-bank" sampling campaign (Appendix E) was conducted on the primary rougher, as this was the unit with the most stable performance in the circuit. The aim of the "down-the-bank" sampling campaign was to create a recovery versus time profile for the plant unit and compare it with the curve obtained from the modelling. Thus, it was a way of testing the validity of the scale-up method. Figure 5.9 show the "down-the-bank" sampling data for Platinum compared to the model.

The fractional palladium and rhodium recoveries obtained from the "down-the-bank" sampling campaign done on the primary rougher, shown in Figure 5.10 and Figure 5.11 respectively, were also compared to the model. The scale-up factors calculated for palladium and rhodium were the same as for platinum.

The same rate constants were used in the platinum, palladium and rhodium modelling. The

different results are due to the difference in distribution of the metals in the different floating classes, which is presented as ore floatability characteristics or M-values.
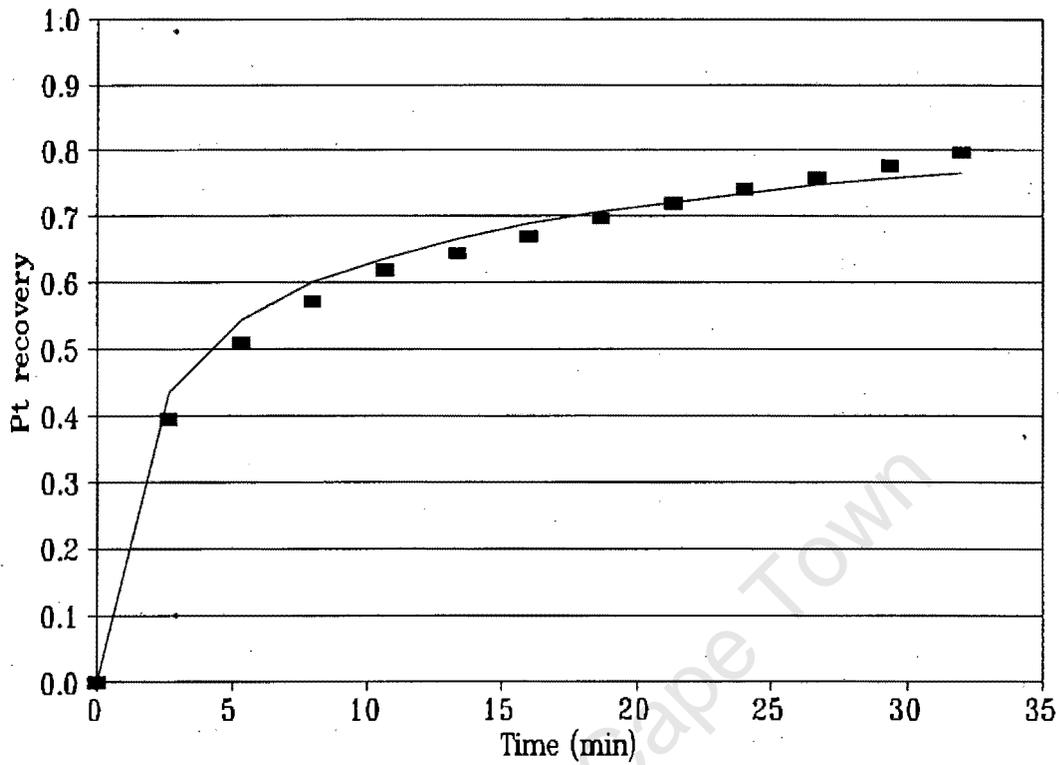
Figure 5.9: The "down-the-bank" sampling data compared to the model.
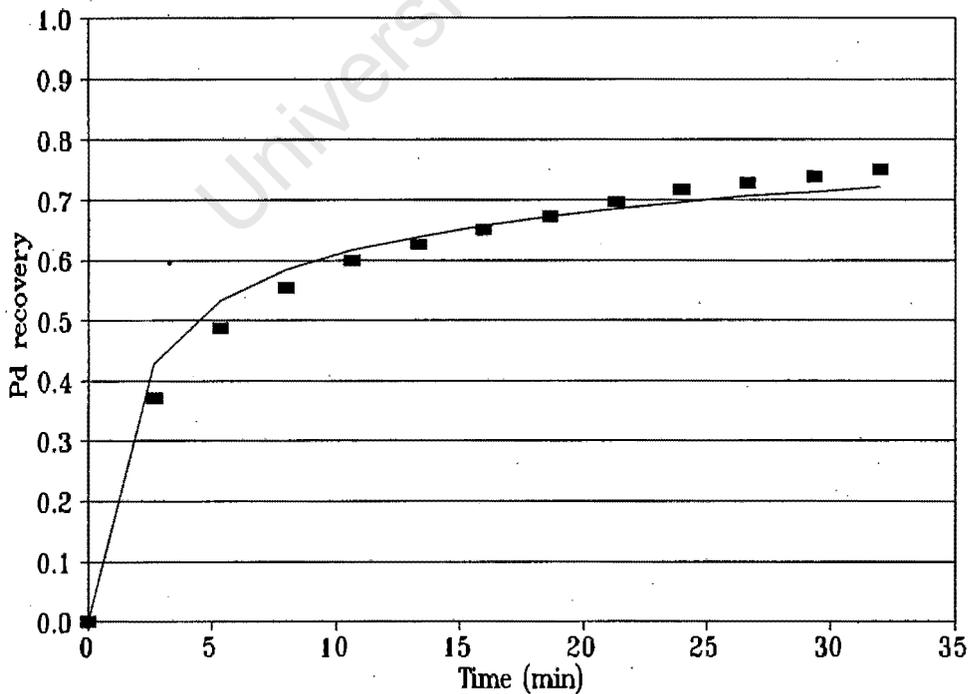
Figure 5.10: Palladium results obtained from the "down-the-bank" sampling campaign compared to the model results.
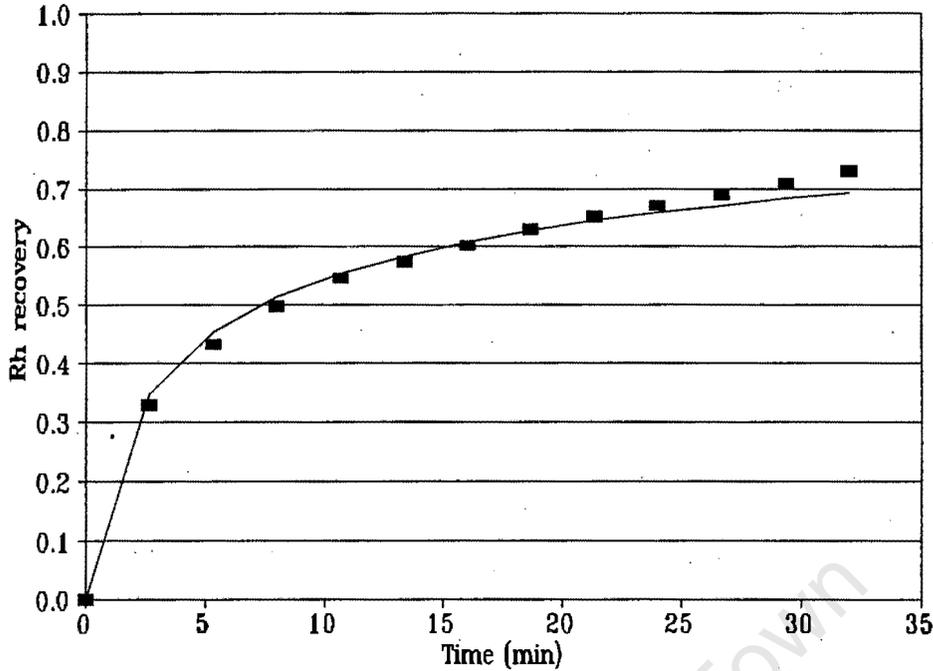
Figure 5.11: Rhodium results from the "down-the-bank" sampling campaign compared to the model results.

5.4 Simulation of the circuit recoveries

The recoveries of the circuit were simulated using the following information:

- ore floatability characteristics (M-values) obtained from the batch flotation tests,

- mixing parameters (τ, N) obtained from the RTD studies,

- the model and rate constants used in the modelling of the unit recoveries,

- the layout of the circuit,

- number of cells in a unit or flotation bank and

- the location and number of cells where final concentrate is drawn.

The ore floatability characteristics of the feed to the primary as well as secondary float section were used in the simulation (Table 5.4) since the characteristics of the feed to the secondary section differed from the primary section feed owing to the regrinding of the primary section tails which was the feed to the secondary section.

Table 5.4: Ore floatability characteristics associated with platinum.

Floatability fractions	Feed of primary float section	Feed of secondary float section
M_f	0.22	0.10
M_m	0.28	0.05
M_s	0.33	0.31
M_o	0.17	0.54

Table 5.4 gives the distribution of platinum in the different floatability classes. The effect of the regrind mill, where the primary rougher/scavenger tails is reground, was simulated by multiplying the platinum distribution (M-values) of the feed to the secondary section, by the total platinum content of the primary rougher/scavenger tails. For example, 10% of the platinum in the primary rougher/scavenger tails will be liberated to float as fast floating material (Table 5.4).

The assumption was made that the residence times of all the cells in a flotation bank is the same. Therefore, dividing the residence time of a bank by the number of cells in the bank will give the residence time of each cell in that bank. Since final concentrate is drawn from the first few cells of each flotation bank, the flotation banks could only be modelled if divided into two separate units (Figure 5.12).

The assumption was made that the rate constants in a particular unit do not change in any of the cells within the unit enabled the simulation of the platinum distribution in the final concentrate stream.

The results obtained from the simulation were compared to the mass balance data (Figure G.1 in Appendix G). The mass balance data represents an average of three mass balances conducted on different days. The normalised final platinum recovery predicted by the simulation for this circuit layout is 4.00%. The normalised average platinum recovery of the three mass balances is 3.96%. Historical data of the plant performance gave a normalised average of around 4.00%.
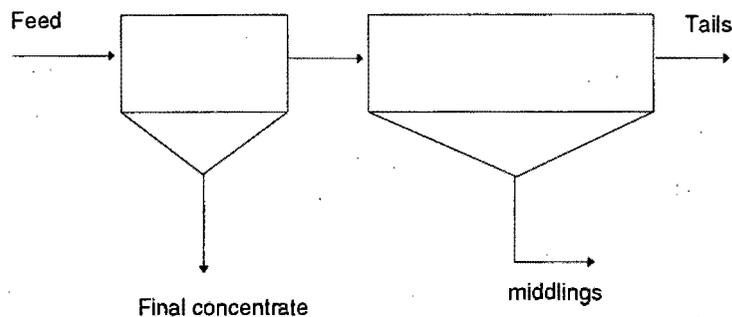
Figure 5.12: Approach followed in the modelling of a flotation bank with two concentrate streams.

5.5 Simulation of other configurations of the circuit

The simulation of the final platinum recovery of the circuit configuration as shown in Figure 5.13 was a success since the platinum recovery predicted by the simulation was within $\pm 0.3\%$ of the average platinum recovery calculated from historical plant data. This suggested that changes in the configuration of the circuit could be simulated, if adequate estimates of changes in the residence times of the units could be made.

An alternative circuit layout, shown in Figure 5.14, was therefore modelled using the same simulation procedure. This circuit differs from the circuit discussed above with respect to the secondary cleaner circuit. The cleaner, a Dorr Oliver unit, was removed. This decreased the residence time in the secondary cleaning circuit. The unit that acted as a recleaner in the circuit shown in Figure 5.13 acts as a cleaner in the circuit shown in Figure 5.14. The volume of the "new" cleaner is less than the removed or "old" cleaner which means a change in the residence time. The effect of this change can be estimated as follows: The nominal residence time is defined as the volume over the volumetric flow rate, $\bar{t} = \dfrac{V}{Q}$. If it is assumed that Q, the volumetric flow rate of the scavenger concentrate which was the stream that fed the "old" cleaner and in the circuit shown in Figure 5.14 feeds the "new" cleaner, remains constant, the change in t can be calculated by the volume difference between
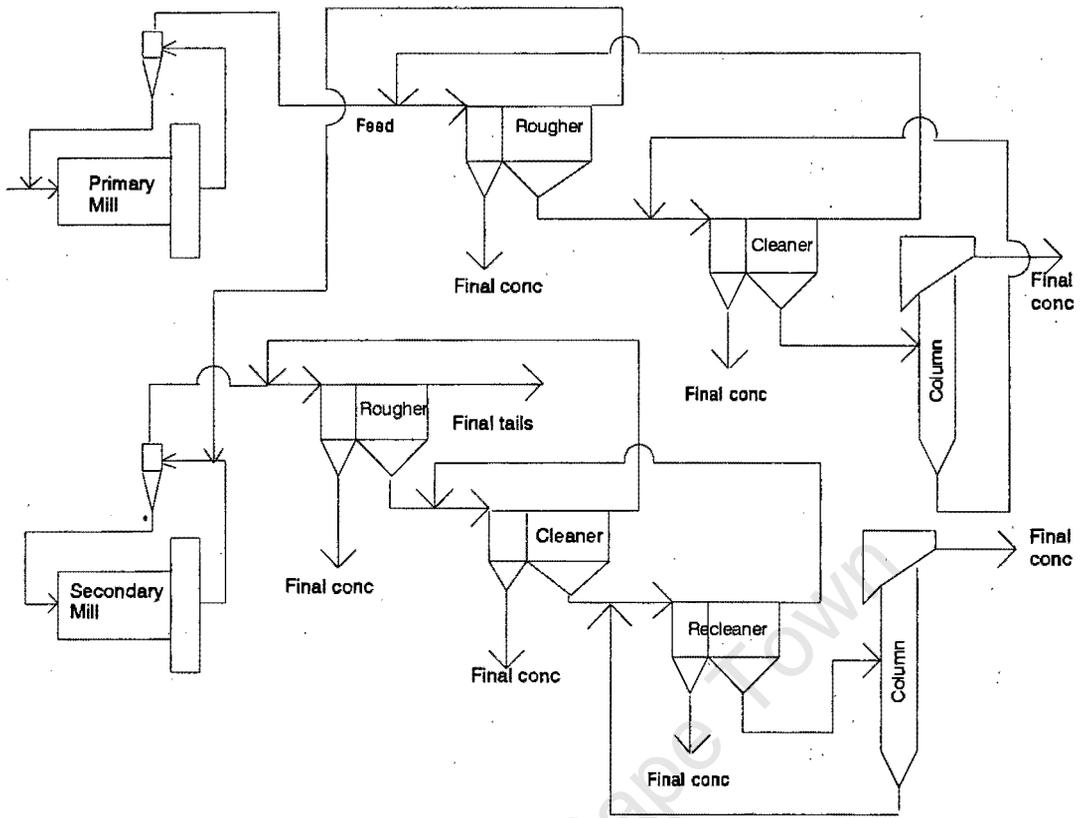
Figure 5.13: The circuit configuration during Phase 2.
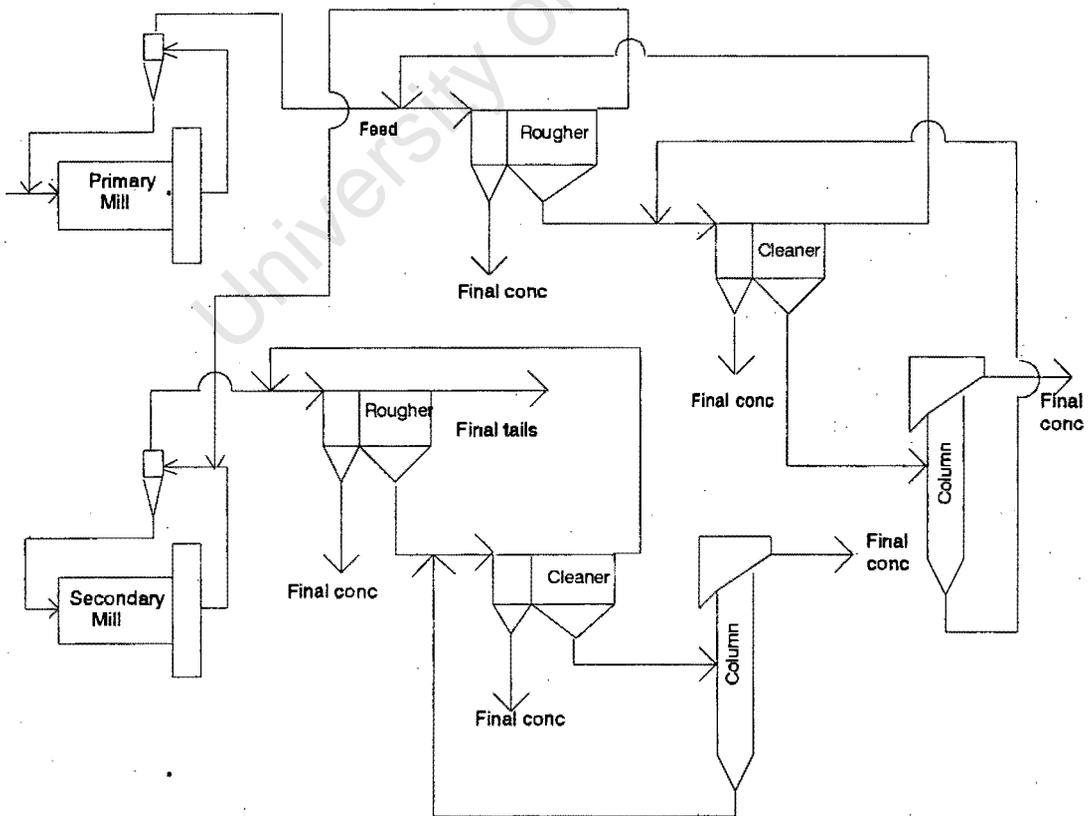
Figure 5.14: Circuit configuration used in the second simulation.

the two units, since τ for the Dorr Oliver cleaner is known (30 min). The cleaner bank has a volume 4.7 times smaller than the Dorr Oliver unit and therefore a τ of 6.35 min.

This simulation predicted a platinum recovery of around 3.84% for the circuit in Figure 5.14. Historical plant data for the circuit operated in this configuration indicated an average platinum recovery of 3.80%.

Figure 5.15 show another circuit which was operated during Phase 1 of the evaluation. The difference between this circuit and the original circuit lies in the cleaning sections. The Phase 2 circuit had the Dorr Oliver cleaner as a cleaner in the secondary section while in this circuit the Dorr Oliver cleaner acts as a cleaner in the primary section.

The residence times will change for the Dorr Oliver unit as well as for the primary recleaner as well as the secondary cleaner. The new nominal residence times were calculated using the method described above.
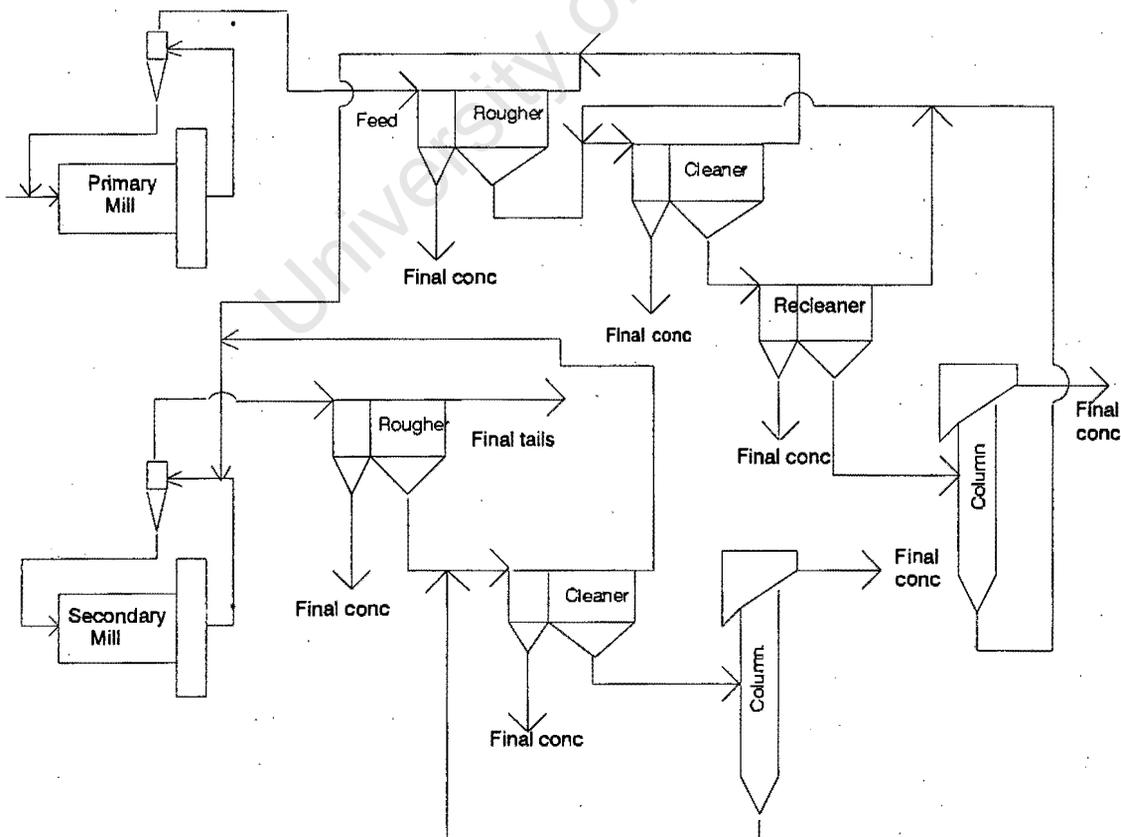
Figure 5.15: Circuit configuration during Phase 1.

This simulation predicted an 3.91% platinum recovery while the average platinum recovery based on historical plant data was 3.84%.

The success of the circuit simulation for 3 very different circuit configurations strongly supports the validity of the modelling methodology used in the evaluation of the circuit.

Chapter Six

Discussion

The main objective of this research was to characterise the complex MF2 circuit at the Impala Platinum concentrator, so as to lay the foundation for the future optimisation of the flotation process used for UG-2 ore at Impala. The high variability of parameters such as flow rates, pulp densities and ore characteristics made it impractical to use standard mass balance techniques to characterise the circuit. A mass balance only yields "snap shot" type indications of circuit performance. While such information is certainly useful, it contributes very little to an understanding of the dynamic behaviour of the circuit under conditions of high variability, and even less to the prediction of the performance of the circuit in different configurations. The latter is particularly important if the information gathered in the investigation is to be of value in circuit optimisation. The characterisation of the circuit was therefore approached using a methodology designed to provide information which could elucidate the mechanisms which influence the performance of the circuit.

The circuit was divided into separate units and each unit was evaluated separately over a period of a few days. All the streams around a unit were sampled to determine the units' performance and the feed and tailing streams were sampled to enable batch flotation test work to be carried out in the laboratory. A preliminary investigation, referred to as Phase 1, was conducted to test the suitability of the methodology for the characterisation of this type of circuit.

The results of the Phase 1 testing were used to refine the characterisation procedure. This methodology was then reapplied to the MF2 circuit, which had undergone a change in configuration at the end of the Phase 1. This subsequent test programme is referred to as Phase 2.

The PGM recoveries calculated for Phase 1 and the platinum, palladium and rhodium recoveries calculated for Phase 2 were normalised for confidentiality reasons. All the PGM recoveries were normalised using one random value to enable comparison between unit and

batch flotation recoveries. The same normalisation technique was applied to platinum, palladium and rhodium.

6.1 Phase 1: Preliminary work

The approach followed to characterise the MF2 circuit was initially started by evaluating the circuit. The method of dividing the circuit into units and evaluating each unit separately over a few days was tested during the preliminary phase. The performance of each unit, in terms of PGM recovery and chromite rejection, was determined. Since three performances on three different days was calculated for each unit, an indication of the stability of the units' performance in the specific circuit configuration was obtained. Table 4.6 shows that the units in the primary section gave stable results during the period of their evaluation. The only exception was the primary column where the PGM recoveries varied between 1.00 and 1.30%. It is perhaps not unexpected that the greatest variability would occur on a unit located at the "end" of the process.

In order to evaluate the performances of the units, an indication of the floatability of the material in the feed stream was needed. Batch flotation tests were assumed to provide a diagnostic indication of the floatability of the material in various process streams. If the PGM recovery of a unit was poor and the batch flotation results also showed a poor PGM recovery then it could be concluded that the unit recovery was low as a result of the nature of the feed material and not as a result of unstable or poor operation of the unit.

During Phase 1, a problem with the sampling of the primary rougher feed was found to have occurred. The batch flotation tests achieved a PGM recovery of approximately 1.36% while the plant unit performed at a 1.49% PGM recovery. In practice, the batch flotation recoveries should be higher or the same as those obtained in full-scale units, as the hydrodynamic conditions in the batch flotation cell are more ideal, and the energy input per volume of pulp is considerably greater. When the anomalous results obtained on the primary rougher feed was observed, the rougher feed batch test samples were screened and the size distribution was found to be 20% passing 75μm. The plant operates at a 50% passing 75μm size distribution for the primary rougher feed. It was therefore concluded that the sample was not representative of the primary rougher feed and that bias occurred during sampling. This was

owing to the fact that the stream was sampled at the feed box of the unit and it was difficult to get a complete cross cut of the stream due to its very high flow rate. It was therefore decided to sample the stream at an alternative location (the overflow of the surge tank), where high flow rates were not a problem.

The preliminary investigation proved that the circuit could be characterised by evaluating the performances of the individual units separately. An indication of how the units performed in the circuit and the stability of their performances was obtained. The feed stream of each unit was characterised in terms of a recovery versus time curve obtained from the batch flotation test work. The batch flotation results gave an indication of what PGM recovery could be achieved under laboratory or "ideal" conditions. Sampling techniques were refined, and the best sampling locations in the circuit were identified.

6.2 Phase 2: Characterisation of the circuit

The methodology which was tested and refined in Phase 1 was now used to carry out a comprehensive investigation on the circuit. In this study, known as Phase 2, the circuit was divided into units and each unit was evaluated separately over a few days. The evaluation of a unit, viz. sampling the streams and batch flotation tests, was repeated on three different days. The evaluation was not repeated for reproducibility purposes but to evaluate the change in the performance of each unit over a few days as well as the change in floatability of the material in the feed stream. The mixing in the different units was also characterised by RTD measurements using a salt tracer technique. The kinetic modelling of the batch flotation data obtained in the evaluation process enabled the ore and its flotation response in a particular unit, to be characterised.

6.2.1 Unit evaluation

6.2.1.1 Platinum recoveries

The 1% deviation in the platinum recoveries obtained for the primary rougher/scavenger (Table 4.10) showed that the operation of the unit was stable during the period of the evaluation. The platinum recoveries obtained from the batch flotation of the primary rougher

feed samples for runs 1, 2 and 3 were between 3.65% and 3.75% (Table 4.14) which were the same as the recoveries obtained for the primary rougher unit.

The platinum recoveries for the primary cleaner show a deviation of about 15% between run 2 and the other test runs (Table 4.10). This deviation could be as a result of unstable operation or changes in the floatability characteristics of the feed material treated by the unit. In order to resolve this issue, data that was obtained from batch flotation tests conducted on the feed samples collected during each test run, was studied (Table 4.14). The platinum recovery obtained from the batch flotation tests could therefore be used as a diagnostic indicator by comparing the batch flotation results to the unit platinum recoveries. The batch flotation test conducted for run 2, on primary cleaner feed, was 10% lower in platinum recovery when compared to the batch flotation results obtained for the other runs. It could therefore be concluded that a change occurred in the floatability characteristics of the ore treated by this unit during run 2 compared to the ore treated during the other two test runs.

Low platinum recoveries were observed for the primary and secondary columns (Table 4.10). A 23% difference in the platinum recoveries existed between run 1 and 3 and run 2 on the primary column. Again, the high deviation could be due to unstable operation of the columns or changes in the ore floatability characteristics. This issue could be resolved by studying the data obtained in the batch flotation tests conducted on the feed samples of each test run (Table 4.14). The batch flotation test work done on the primary column feed samples gave platinum recoveries of between 4.05 and 4.31%. The batch flotation results show that more than 4.00% of the platinum-bearing minerals in the feed stream were floatable while the plant unit could only recover a maximum of 3.05% of the platinum in the feed stream. Therefore in this case, the significant changes in the platinum recovery of the unit for the three campaigns are a result of unstable operating conditions and not a result of changes in the ore floatability characteristics for the different runs, since the batch flotation results show no significant changes in the floatability of the platinum minerals that were present in the feed stream.

The secondary rougher/scavenger gave platinum recoveries that varied between 2.46 and 1.61% (Table 4.10). Once again, this large variation could be ascribed to either unstable operation, or to changes in the ore floatability characteristics. The batch flotation tests

conducted on the feed samples of each test run for the secondary rougher feed samples gave platinum recoveries of between 1.80% and 2.02% (Table 4.14). The platinum recoveries of the rougher unit gave similar platinum recoveries except for run 1 where the unit gave a 2.46% platinum recovery while the batch flotation test on the feed sample for this specific run resulted in a 1.80% platinum recovery. The batch flotation results therefore gave no indication of changes that could have occurred in the ore floatability characteristics for this test run. The 18% difference in the platinum recoveries of run 1 and the other runs as well as the 13% difference between the unit platinum recovery and the platinum recovery obtained from the batch flotation test done on the feed sample during run 1 are unexplainable in terms of operating conditions or ore floatability characteristics. The deviations in the platinum recoveries could therefore be as a result of sampling or assaying errors.

The secondary cleaner gave very stable results, varying by no more than 4% (Table 4.10). This was also reflected in the platinum recoveries obtained from the batch flotation test work performed on the secondary cleaner feed samples, which were almost identical for all three test runs (Table 4.14). The 3.59% platinum recovery that was obtained in the three batch tests was about 5% lower than the platinum recoveries obtained for the cleaner unit. As stated previously, the expectation is that the batch flotation recoveries would generally be higher than the recoveries obtained for the plant unit. Since the opposite was true for all three runs in this case, the results suggested that the hydrodynamic conditions in the secondary cleaner were surprisingly superior to those in the batch flotation system used in this study. This phenomenon will be addressed in the discussion about the kinetic modelling of this unit.

The platinum recoveries calculated for the secondary recleaner were almost the same for the three different runs, which is indicative of stable operation during the period of evaluation (Table 4.10). The batch flotation test work gave platinum recoveries that compared well with unit recoveries for each different run (Table 4.14).

The batch flotation work performed on the secondary column feed samples gave recoveries of between 3.78% and 4.02% (Table 4.14) while the secondary column unit gave platinum recoveries of between 1.00% and 1.32% (Table 4.10). This mirrors to the same extent the results obtained in the primary column, although the difference is considerably greater. While

the considerable difference in the collection zone environment in the column compared to that of a batch cell makes a direct comparison rather unfair, if the batch flotation results are used as an indication of the floatability of the material in the feed stream of the unit the conclusion can be made that the columns were performing very poorly in terms of platinum recovery.

6.2.1.2 Palladium recoveries

Table 4.11 confirms the observations made for the platinum recoveries for the units. The palladium recoveries for each batch flotation run carried out during Phase 2 of the evaluation are presented in Table 4.15. The palladium batch results obtained for the primary rougher feed samples gave recoveries that are lower than the results obtained for platinum. The palladium recoveries obtained from the batch flotation tests performed on the feed samples of the unit were almost identical to the palladium recoveries obtained for the unit.

The palladium recoveries obtained for the primary cleaner show a large deviation between run 2 and the other two runs (Table 4.11). As explained previously for platinum, this could be as a result of unstable operation or changes that occurred in the floatability characteristics of the ore. As in the case of platinum, the batch flotation results for palladium (Table 4.15) were studied and it was found that a significant change occurred in the floatability characteristics of the ore floated during run 2, since the palladium recoveries obtained for the batch flotation test work performed on the primary cleaner feed samples showed a 10% difference between the other runs and run 2.

The primary column gave the same low recoveries for palladium as it did for platinum (Table 4.11). A large deviation in the palladium recoveries was observed. The batch flotation test work carried out on the feed samples of the primary column gave palladium recoveries of between 2.00% and 2.10% (Table 4.15). Although the batch flotation results showed that the minerals associated with palladium were highly floatable, the column unit could only recover between 0.95% and 1.44% of the palladium in the feed stream.

The units in the secondary section gave the same occurrences for palladium recovery as observed for platinum. The secondary rougher gave palladium recoveries of between 1.38

and 0.77% (Table 4.11). Again, the large deviation could be as a result of unstable operation or changes in the ore floatability characteristics of the feed stream. The batch flotation results for palladium obtained for the secondary rougher feed samples gave recoveries of between 0.81% and 0.88% (Table 4.15). The batch flotation results showed that no changes occurred in the ore floatability characteristics since the batch flotation recoveries were almost similar. The higher palladium recovery for this unit that was observed for run 1 could not explained in terms of operating conditions or changes in floatability characteristics of the ore. The deviation in the palladium recoveries for this unit could therefore be as a result of assay or sampling errors.

The palladium recoveries obtained for secondary cleaner and secondary recleaner gave very stable results and indicates a stable operation during the evaluation period (Table 4.11). The batch flotation test work carried out on the feed samples of the secondary cleaner gave palladium recoveries of between 1.70% and 1.80% (Table 4.15). The palladium recoveries for this unit were higher than for the batch flotation tests done on the feed samples. As stated previously, the expectation is that the batch flotation recoveries should generally be higher than the recoveries obtained for the plant unit. It therefore seemed that the hydrodynamic conditions in the secondary cleaner was surprisingly superior to that of the batch flotation cell in this study. The same occurrence was true for the platinum results for this unit. This phenomenon will be addressed in the discussion about the kinetic modelling of this unit.

The batch flotation results for palladium obtained for the secondary recleaner feed samples gave recoveries of 1.97% except for run 2 where the recovery was 2.22% (Table 4.15). The significant difference could be as a result of changes in the ore floatability characteristics. The unit palladium recovery for run 2 did not show a significant difference compared to the other runs and the difference in the batch flotation results could be as a result of sampling or assaying errors.

The secondary column palladium recoveries were very low with respect to the other cleaners (Table 4.11). The same observation was made for the platinum recoveries of this unit (Table 4.10). The batch flotation test work carried out on the feed samples of the secondary column gave palladium recoveries of between 1.98% and 2.18% (Table 4.15). Although the batch flotation results show that the palladium minerals in the feed were highly floatable, the

palladium recoveries for the column unit were only between 0.57% and 0.61%. As for platinum, the conclusion can be made that the columns were performing very poorly in terms of palladium recovery.

6.2.1.3 Rhodium recoveries

,The rhodium recoveries obtained for the primary rougher were lower than those obtained for platinum. The rhodium recoveries were also lower than those obtained for palladium. The constant recoveries obtained during the evaluation period show a stable operation for this unit. The rhodium recoveries for the primary rougher unit were almost identical to the batch flotation recoveries (Table 4.16).

A large deviation in the rhodium recoveries for the primary cleaner (Table 4.12) indicates an unstable operation or a change in the ore floatability characteristics of the feed material, as observed for platinum and palladium. The batch flotation test work carried out on the primary cleaner feed samples gave rhodium recoveries of approximately 2.60% except for run 2 where a 2.27% rhodium recovery was calculated (Table 4.16). As discussed for platinum and palladium, the 12% difference was as a result of a change in the ore floatability characteristics since the results of the batch flotation tests done on the primary cleaner feed showed a significant deviation between run 2 and the other two runs.

The batch flotation results for rhodium obtained for the test work done on the primary column feed samples gave recoveries of between 2.40% and 2.54% (Table 4.16). The column unit recovered only between 1.10% and 1.68% of the rhodium in the feed stream (Table 4.12). The primary column gave the same low recoveries for platinum and palladium. The large deviation in the unit recoveries were as a result of unstable operation since the batch flotation results showed no significant change in the floatability of the ore between the different runs.

The secondary cleaner which gave constant recoveries for platinum and palladium gave a 11% difference in the rhodium recoveries of run 2 and run 3 (Table 4.12). This observation is difficult to explain since the same samples gave constant platinum and palladium results and constant rhodium results were expected. The difference could be as a result of errors during assaying. The batch flotation test work done on the secondary cleaner feed samples gave

rhodium recoveries of between 2.05% and 2.23% (Table 4.16). These values were significantly lower than the rhodium recoveries of the cleaner unit. The general expectation is that the batch flotation recoveries should be higher than the plant values. The same trend was experienced for the platinum and palladium results which let one to believe that the hydrodynamic conditions in the secondary cleaner were superior to the batch flotation cell in this study. This phenomenon will be addressed in the discussion about the kinetic modelling of this unit.

The secondary recleaner gave constant results which indicates a stable operation during the evaluation period (Table 4.12). The secondary column show a poor ability to recover rhodium under the operating conditions experienced during the evaluation period (Table 4.12). The batch flotation test work carried out on the secondary column feed samples gave rhodium recoveries of between 2.21% and 2.27% (Table 4.16). Although the batch flotation results showed that the material in the column's feed was highly floatable the column unit could only recover between 0.58% and 0.80% of the rhodium present in the feed. The same trend was experienced for platinum and palladium. The conclusion could therefore be made that the secondary column was performing very poorly in terms of PGM recovery.

6.2.1.4 Summary of unit performances

The platinum, palladium and rhodium recoveries for the primary rougher showed that the operation of this unit was stable during the period of evaluation. A large difference in the platinum, palladium and rhodium recoveries for the primary cleaner occurred between run 1 and 3 and run 2. This was a result of changes that occurred in the floatability characteristics of the ore which was clearly showed in the batch flotation results. The batch flotation results were therefore used as a diagnostic indicator by comparing the batch flotation results to the unit recoveries.

Low platinum, palladium and rhodium recoveries were observed for the primary column as well as significant differences in the recoveries between run 1 and 3 and run 2. The significant changes in the platinum, palladium and rhodium recoveries of the unit for the three campaigns were a result of unstable operating conditions and not a result of changes in the ore floatability characteristics for the different runs, since the batch flotation results showed

no significant changes in the floatability of the platinum, palladium and rhodium minerals that were present in the feed stream.

The secondary cleaner gave unit platinum, palladium and rhodium recoveries that were higher than the batch flotation results. The expectation is that the unit recoveries would generally be lower than the batch flotation recoveries. Since the opposite was true for all three runs in this case, the results suggested that the hydrodynamic conditions in the secondary cleaner were surprisingly superior to those in the batch flotation system used in this study. This phenomenon will be addressed in the discussion about kinetic modelling of this unit.

The platinum, palladium and rhodium recoveries calculated for the secondary recleaner was indicative of a stable operation during the period of evaluation. The secondary column gave poor platinum, palladium and rhodium recoveries although the batch flotation results showed that the material in the feed stream to the unit was highly floatable.

The performance of each unit was not only measured in its ability to recover platinum, palladium and rhodium but also to minimise chromite recovery into the concentrate since chromite prove to be problematic in the smelting stages.

6.2.1.4 Chromite recoveries

Since chromite is a major problem in the smelting stages in the recovering of PGMs from UG-2 ore, the performance of a unit was also measured in terms of its, ability to reject chromite and not only its ability to recover PGM minerals.

According to Liddell et al. (1986) chromite in UG-2 ore is recovered through physical entrainment. The results from the "down-the-bank" sampling enabled the construction of a chromite recovery versus water recovery graph (Figure 4.8) for the primary rougher/scavenger unit. A linear relationship would suggest that chromite was recovered through entrainment (Lynch et al., 1981). Figure 4.8 clearly shows that the observation of Liddell et al.(1986) about the recovery mechanism for chromite is also true in this circuit treating UG-2 ore. The batch flotation test results also show a linear relationship between chromite recovery and water recovery (Figure 6.1).

The units were operated to minimize entrainment and maximize PGM recovery. The entrainment of chromite will depend on the water recovery, particle size, pulp density or solids concentration, mineral density, drainage of the solids in the froth as well as the
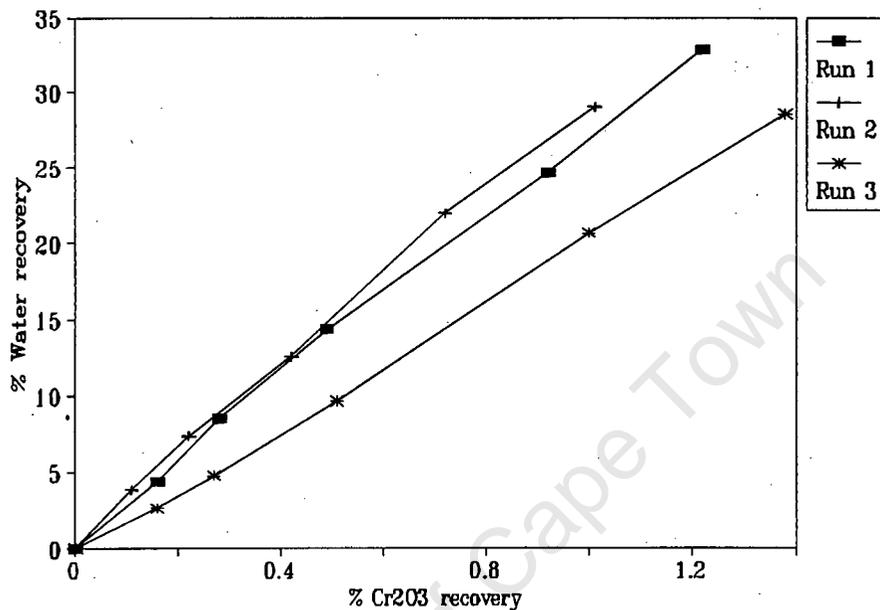
Figure 6.1: Water and chromite (Cr_2O_3) recoveries obtained from batch flotation tests done on primary rougher feed.

drainage of the water in the froth which in turn depends on the residence time of the particles and water in the froth (Bishop and White, 1976; Lynch *et al.*, 1981). The entrainment of the hydrophilic particle or in this case chromite can be minimised by controlling the pulp density of the feed material, frother addition, air flow rate through the cell and the froth depth in the cell.

The primary rougher/scavenger recovered between 1.5% and 2.4% of the chromite present in the feed material. This is significantly lower than the 23% to 48% recovered by the primary cleaners. The same occurred for the chromite recoveries for the batch flotation tests done on the feed material going to the primary rougher/scavenger and the primary cleaner. The major

differences in the condition for the two units were the pulp density which means a difference in solids concentration as well as the froth depth. Since the batch results gave chromite recoveries which were almost the same as the unit recoveries, the froth depth or condition could not be the major reason for the difference in chromite recovery between the two units since the froth depth and froth characteristics (water content, residence time of pulp, drainage of pulp, etc.) in a batch cell is not the same as in the plant units.

The chromite recovered through entrainment in the rougher/scavenger is likely to be recovered in the cleaner as well. Since the cleaner feed was the scavenger concentrate the rejection of the chromite will be more difficult than in the rougher unit especially since the pulp density in this cleaner is much lower viz., 1.6 in the rougher and 1.2 in the cleaner, and the recovery of water in the cleaner would be expected to be higher. The cleaner was therefore operated with a deep froth depth to increase the drainage of the water and so the entrained particles. Although the cleaner reduced the chromite content of the concentrate to 2 - 3% the relative chromite recovery is still higher than the rougher/scavenger.

The washing of the froth in the primary column proved to be effective since the chromite recovery of this unit was less than for the primary cleaner. The chromite recovery of the column was between 36 and 18% during the evaluation period.

The secondary recleaner recovered between 48% and 71% of the chromite that entered the unit. These values are remarkably higher than the cleaner chromite recoveries. After each stage it becomes more difficult to clean the chromite out of the concentrate and the increasing water content helps the recovery of chromite through entrainment. The chromite recoveries produced by the secondary column, which acted as a re-recleaner, were lower than those obtained for the recleaner. The washing of the froth in a column was the main reason for the lower chromite recoveries. The 65% chromite recovery obtained in test run 3 for the column was due to unstable addition of wash water which was caused by the presence of particulate matter in the wash water tank which plugged the holes through which the water escapes to wash the froth. The column therefore succeeded in rejecting the chromite by washing the froth.

The comparable batch flotation chromite recoveries are shown in Table 4.17. When these

results are compared to the plant unit performance it is clear that in all cases, except for the column, the batch flotation recoveries of chromite are lower than the plant indicating a lower extent of entrainment in the batch cell. The column, as would be expected, showed much lower chromite recoveries than the batch flotation cell.

6.2.2 RTD studies

Residence time distribution (RTD) studies were performed on each type of flotation machine in the circuit. The main objective of the RTD studies was to obtain the mean residence times (τ) and the mixing characteristics of each type of flotation machine in the circuit.

The mixing in the units was quantified by fitting the tanks-in-series model since the number of tanks in series (N) obtained from the model could be physically related to the number of cells in series in a flotation bank (Section 5.2). The axial dispersion model was also fitted (Section 5.2) and the dispersion number (N_d) obtained was used to verify the results of the Tanks-in-series model. Both these models are one parameter models and the parameters that were obtained from modelling the data gave a measure of the degree of mixing in the units.

The results obtained from the RTD studies (Section 4.3.3 and Section 5.2) are summarised in Table 6.1.

Table 6.1: Summary of RTD results

Unit	τ (min)	$\bar{t}$ (min)	N	N_d
Primary rougher	32	35	9	0.07
Primary cleaner	8½	11½	13	0.06
Primary column	44 - 47	48	1	1.8
Secondary cleaner	30	30	4	0.12

The primary rougher/scavenger unit could be modelled as 9 tanks in series which indicates plug flow characteristics rather than mixed flow. The low value of N_d, which according to Levenspiel (1972) indicates an intermediate amount of dispersion, is consistent with the

mixing characteristics obtained from the tanks-in-series model. The fact that the mean residence time (τ) was almost the same as the nominal residence time ($\bar{t}$) indicates that no channelling of fluid, stagnant regions or internal recirculation occurred in the primary rougher/scavenger unit (Levenspiel, 1972).

The primary cleaner unit was modelled as 13 tanks in series which indicates plug flow rather than mixed flow conditions. The N_d obtained from the axial dispersion model was 0.06 and according to Levenspiel (1972) indicates an intermediate amount of dispersion. The large deviation between the mean residence time (τ) and the nominal residence time ($\bar{t}$) was probably as a result of the experimental procedure rather than poor flow (channelling etc.) in the unit since the experimental procedure could not generate a sufficient number of points to give an accurate RTD curve because of the short residence time of the fluid in the unit.

The primary column was modelled as a perfectly mixed tank reactor ($N = 1$). Mills (1991) observed the same occurrence for columns with a diameter of more than 1 metre. The large N_d value of 1.8 confirms that the fluid in the column can be characterised close to mixed flow. The mean residence times for the three test runs varied between 44 - 47 minutes. This was due to unstable operation as a result of variations that occurred in the feed flow rate to the unit. The RTD curve obtained for run 1 (Figure 4.5) shows an oscillation pattern which indicates internal recirculation (Levenspiel, 1972). This is to be expected since the column was not equipped with baffles which could improve mixing in the unit. Baffles are recommended for columns with a diameter of more than 1 metre (Dobby and Finch, 1990).

The secondary cleaner was modelled as 4 tanks-in-series. The N_d value of 0.12 indicates a large amount of dispersion(Levenspiel, 1972) and confirms the tanks-in-series model result that the fluid could be described with mixed flow rather than plug flow. The τ and the nominal residence times ($\bar{t}$) were identical. This indicates a normal behaviour for the fluid in the unit with no channelling, stagnant regions or internal recirculation.

6.2.3 Modelling of the batch flotation results

As discussed in Section 5.1 the floatability parameter model was used to model the batch flotation results. The model derived to fit the data consisted of three floatability classes

namely a fast floating class, a medium floating class, a slow floating class and a class for valuable minerals which did not float. Batch flotation tests were performed on samples from the feed stream to each unit. The modelling of the batch flotation results therefore characterised the feed to each unit. The model provided a flotation rate constant for each floatability class and distributed the valuable minerals into the four classes.

The application of the floatability parameter model in this study assumes that the rate constants associated with each floatability class remain unchanged in all the batch flotation tests and that each class has a constant mean particle size and mineralogy or hydrophobicity. Only the proportion of each floatability class (M-values) varies depending on the sampling location in the circuit. This implies that a valuable mineral will float with the same rate wherever it is in the circuit if floated under the same conditions with the same procedure and equipment.

The M-values of the Phase 1 and Phase 2 batch flotation results are presented in Appendix G. The significance of the sampling location on the M-values is evident. The expectation is that the primary rougher feed should have more fast floating, medium floating and slow floating minerals than the primary rougher tails while the tails should have no fast floating and medium floating minerals, a small fraction of slow floating minerals and a large fraction of valuable minerals which do not float. The calculated M-values reflect this situation (Table 6.2).

The primary rougher tails were sent for regrinding and the effect of the regrinding is evident in the M-values calculated for the secondary rougher feed (Table 6.2). The large fraction of valuable minerals not floatable decreased significantly in the secondary rougher feed and these liberated minerals were distributed to the floatable fractions depending on the flotation rate.

In summary, the results of modelling the batch flotation data are corresponding with what would be expected from the ore in the feed stream to each unit. The ore is thus well characterised by treating it as material with four discrete fractions, viz. fast floating, medium floating, slow floating and non-floating material. The model, however, was not discriminating between mineralogical and particle size effects on flotation rates. This would be an important prospect of a future investigation.

Table 6.2: M-values for Phase 1 batch flotation data (PGM results)

Sampling location	Run	M_f	M_m	M_s	M_o
Primary rougher feed	1	0.08	0.19	0.48	0.25
	2	0.04	0.37	0.39	0.21
	3	0.05	0.25	0.48	0.22
Primary rougher tails	1	0.01	0.02	0.15	0.82
	2	0.03	0.00	0.14	0.83
	3	0.01	0.02	0.15	0.83
Secondary rougher feed	1	0.07	0.24	0.40	0.74
(Reground primary	2	0.05	0.16	0.37	0.76
rougher tails)	3	0.03	0.19	0.41	0.69

6.3 Simulation of the circuit performance

The approach through out this study was to investigate each unit individually. The modelling of the circuit performance was approached in the same way. The performance of each unit was modelled individually, by using the unit recoveries, the floatability fractions from the modelling of the batch flotation data (viz. M-values), the batch flotation rate constants and the mixing parameters. This has been shown diagrammatically in Figure 5.8. The fit between actual recoveries and predicted recoveries (Section 5.3) yielded unit scale-up constants. Thereafter the performance of the circuit was simulated. The same model was used for all the units and the differences in cell design as indicated by hydrodynamics, aeration rate, etc. were accounted for in the mixing parameters and flotation rates. A methodology was developed for the scale-up of the rate constants (Section 5.3.1) from which a scale-up factor (f) was calculated. Since each type of unit will create a different flotation environment, each machine will have a different scale-up factor (f) and therefore different flotation rates.

6.3.1 Validation of the methodology

The methodology for the scale-up of the rate constants used the unit recoveries obtained in the unit evaluation process to calculate f. Three recoveries existed for each unit and in most cases the recoveries differed since the unit evaluation was spread over a few days. Since the feed stream to each unit was characterised during each evaluation run, M-values or ore floatability characteristics could be generated for each run. The scale-up factor of each unit was calculated using three unit recoveries and three sets of M-values. For each unit three scale-up factors could be calculated and the three factors proved to be almost identical although different ore floatability characteristics and unit recoveries were used to generate them. This implies that the scale-up method generated unique scale-up factors for each unit.

The scale-up methodology used recoveries at one residence time only and after scale-up the models could predict recoveries of each unit at different residence times. "Down-the-bank" sampling was conducted on the primary rougher/scavenger unit to obtain recoveries at different residence times for the plant unit. The data obtained from the "down-the-bank" sampling campaign was compared to the recoveries predicted by the model at different residence times (Figure 5.9, Figure 5.10 and Figure 5.11). The average deviation per point between the predicted platinum recoveries at different residence times and platinum recoveries obtained from the "down-the-bank" sampling was 0.8%. The average deviation per point for the comparison of the Palladium and Rhodium results was also less than 1% (Figure 5.10 and Figure 5.11).

Since unique scale-up factors, and therefore rate constants, for each unit could be generated and the modelling of platinum, palladium and rhodium recoveries at different residence times for the primary rougher/scavenger could be confirmed by plant data proved that the methodology used to obtain the scale-up factors was valid.

6.3.2 The scale-up of the rate constants

In the rougher/scavenger units for the primary as well as secondary stages a scale-up factor (f) of 1.35 was obtained. Although the two rougher/scavenger units are identical, different operating conditions were used in each unit. The feed pulp density to the secondary

rougher/scavenger was 1.4 - 1.45 compared to a 1.55 - 1.6 pulp density for the primary rougher/scavenger. This would result in a difference in the % solids and the viscosity of the pulp which in theory will have an effect on the mixing and bubble size in the units and therefore on the flotation rate. Since the same flotation rate constants were obtained for the two units, the difference in feed pulp density did not change the rate of flotation. Since the cells in the rougher/scavenger units are 8.4 m^3 the expectation is that for this ore the pulp density effect on the flotation rate will be more evident in larger cells like 14m^3 cells than in the 8.4 m^3 cells.

In the cleaner and recleaner units, flotation rates faster than the batch flotation rates were obtained except for the columns. The primary cleaner and secondary recleaner, which were Wemco cleaners, produced the same scale-up factor of 0.36 while the secondary cleaner which was an Dorr Oliver unit, produced a 0.61 scale-up factor.

It was observed that the platinum, palladium and rhodium recoveries for the secondary cleaner were higher than the recoveries obtained in the batch flotation cell (Table 4.10 to Table 4.16). The comment was made that the hydrodynamic conditions in this unit must be superior to the batch flotation cell in this study. The rate constants calculated for this unit indicated that the hydrodynamic condition in the plant unit was indeed superior to the batch cell since the rate constants were significantly faster in the secondary cleaner than in the batch flotation cell. The rate constants obtained for the primary cleaner and secondary recleaner were 3.8 times faster than those used for the Wemco rougher/scavenger units. It was calculated that the power to volume ratio of the Wemco cleaners were ±3.5 times more than for the Wemco rougher/scavenger units. Since the Wemco units use self aeration, a difference in power will result in different aeration rates and the volume difference together with the aeration difference would result in a difference in the superficial gas velocity (J_g) of the unit. The flotation rate constant is approximately directly proportional to superficial gas velocity over bubble size, viz. $k \propto \dfrac{J_g}{d_b}$ (Dobby and Finch, 1990) and the higher J_g value would explain the faster flotation rates in the Wemco cleaners, assuming a constant bubble size.

The rate constants used for the primary column were 6.4 times slower than the batch flotation rate constants. If compared to the rate constants used for the modelling of the primary rougher/scavenger then the primary column produced rate constants 4.7 times slower than the

rougher/scavenger. In the relationship of $k = 1.5 \ E_k \dfrac{J_g}{d_b}$ (Dobby and Finch, 1990) the only parameter that was possible to calculate was J_g. It was found that the J_g of the column was the same as the J_g in each cell of the rougher/scavenger unit (±2 cm/s) and therefore this could not be the reason for the difference in the rate constants. A hypothesis concerning the difference in bubble size (d_b) and particle collection efficiency (E_k) in the two units was offered as an explanation for the slower flotation rates in the primary column. Jameson and Allum (1984) did a survey of bubble sizes in industrial flotation cells and found that in a Wemco rougher/scavenger bank with 8.4 m³ cells an arithmetic mean bubble size of 0.5 mm was produced with a J_g of ±1.7 cm/s and a gas hold up of $\pm14\%$. Flotation columns are usually operated with bubble sizes of 0.8 - 1.5 mm (Dobby and Finch, 1990). The particle collection efficiency of fine particles is higher in an agitated environment (Breytenbach, 1995) and therefore higher in the Wemco roughers than in the column cell. The larger bubble size and the lower particle collection efficiency of fine particles in the column compared to the Wemco rougher/scavenger may account for the lower flotation rate constants calculated for the column cells.

6.3.3 Simulation of the Phase 2 circuit

The simulation of the Phase 2 circuit (Figure 5.13) proved to be successful in terms of overall platinum, palladium and rhodium recovery. The distribution of these platinum group metals in the circuit was also obtained from the simulation. The overall platinum, palladium and rhodium recoveries obtained from the simulation could be compared to historical data gathered daily for metallurgical accounting. The distribution of these precious metals in the different streams in the circuit predicted by the simulation could be compared to mass balance data.

The good comparison between the distribution of platinum in the final concentrate streams calculated for the mass balance and predicted by the simulation (Figure G.1 in Appendix G) was evident in the 1% difference between the overall platinum recovery calculated from the mass balance and predicted by the simulation. The platinum recovery of the primary column calculated from the mass balance differed significantly (20%) from the recovery predicted by the simulation. The raw data of the mass balance were smoothed to balance the effect that the units will have on each other, and since the column was at the "end" of the process the

effect of smoothing will be more evident in the grades around the column. On the other hand, the simulation used kinetics and mixing characteristics which were determined by the individual evaluation of each unit over a few days. The main advantage of the simulation is that it can predict the change in performance of the circuit in terms of platinum, palladium and rhodium recovery if the floatability characteristics of the ore should change. This predictive power could be of immense benefit in the testing of various parameters such as grind which would alter the particle size of the ore and therefore the floatability in the batch cell. These newly obtained floatability characteristics can then be used in the simulation and the performance of the circuit could be predicted before testing the new conditions on the plant.

Although the simulation gives the performance of each unit in terms of platinum, palladium and rhodium recoveries it does not give an indication of chromite recovery. Since it was proved that chromite was recovered through physical entrainment in the froth (Section 6.2.1.4), the modelling of chromite recovery will only be possible if a model for entrainment could be obtained which means the froth zone needs to be modelled. Future work must therefore be focused on entrainment.

6.3.4 Simulation of different circuit configurations

The simulation of other configurations of the circuit (Figure 5.14, Figure 5.15) proved to be successful (Section 5.5) since the overall platinum, palladium and rhodium recoveries of both circuits deviated within 1% of historical plant data. The success of the simulation of three different circuit configurations strongly supports the validity of the modelling methodology used in the evaluation of the circuit.

The difference in the simulated circuits was in the cleaning stages. It was shown that the best platinum, palladium and rhodium recoveries could be obtained if the greatest cleaning capacity was situated in the secondary stage of the circuit. Since the secondary cleaner tails were recycled to the surge tank which means that the material will again go through the rougher and a high probability exists that these valuable minerals will get lost in the final tails, it was therefore a significant improvement by putting a highly effective cleaner in the secondary circuit which increased the cleaner capacity and decreased the recycling of

valuable minerals to the rougher. It was evident from the simulation of the three circuit that the Dorr Oliver cleaner had a significant impact on the recovery in the circuit. When the Dorr Oliver cleaner was removed from the circuit in Figure 5.14 the platinum recovery dropped by 1.5% and when the Dorr Oliver cleaner was placed in the secondary cleaning circuit as a cleaner (Figure 5.13) the platinum recovery increased approximately 3%.

Although the changes in the circuit configuration could be simulated to see the effect on the platinum, palladium and rhodium recoveries, the effect of the changes on the chromite recovery could not be simulated. Since the chromite recovery is crucial for the design of optimum cleaning circuits in the flotation of UG-2 ore it is strongly recommended that research be done in the field of froth characteristics and the modelling thereof.

The simulation of the three circuits was possible since the same primary and secondary grind and feed pulp density values were applicable to all three circuits. The effect of grind on the platinum, palladium and rhodium recoveries could be investigated by changing the grind and doing batch flotation tests on the mill product. The batch flotation results will enable the calculation of new M-values or ore floatability characteristics and the effect of the new grind could be simulated before implemented on the plant. The effect that finer or coarser grind will have on the different float fractions and the fraction of valuable minerals not floatable can also be investigated in the laboratory. The optimum grind for the secondary mill could be estimated by determining the grind where the fraction of valuable minerals not floatable can no longer be decreased as a result of milling.

Chapter Seven

Conclusions

A methodology was developed to characterise the complex MF2 circuit at Impala Platinum Ltd. The circuit was divided into individual unit operations and each unit was evaluated over a period of a few days. The unit performance in terms of platinum, palladium, rhodium and chromite recoveries was established by sampling around the unit and the changes in ore floatability characteristics were monitored by subjecting the feed samples taken at each unit to batch flotation tests. The mixing in the units was characterised by performing residence time distribution studies on the units.

7.1 Characterisation of the circuit

7.1.1 Unit performance

The platinum, palladium and rhodium recoveries for the primary rougher showed that the operation of this unit was stable during the period of evaluation. A large difference in the platinum, palladium and rhodium recoveries for the primary cleaner occurred between run 1 and 3 and run 2. This was a result of changes that occurred in the floatability characteristics of the ore which was clearly shown in the batch flotation results. The batch flotation results were therefore used as a diagnostic indicator by comparing the batch flotation results to the unit recoveries.

Low platinum, palladium and rhodium recoveries were observed for the primary column as well as significant differences in the recoveries between run 1 and 3 and run 2. The significant changes in the platinum, palladium and rhodium recoveries of the unit for the three campaigns were a result of unstable operating conditions and not a result of changes in the ore floatability characteristics for the different runs, since the batch flotation results showed no significant changes in the floatability of the platinum, palladium and rhodium minerals that were present in the feed stream.

The secondary cleaner gave unit platinum, palladium and rhodium recoveries that were higher

than the batch flotation results. The expectation is that the unit recoveries would generally be lower than the batch flotation recoveries. Since the opposite was true for all three runs in this case, the results suggested that the hydrodynamic conditions in the secondary cleaner were surprisingly superior to those in the batch flotation system used in this study (Section 7.4.1.).

The platinum, palladium and rhodium recoveries calculated for the secondary recleaner was indicative of a stable operation during the period of evaluation. The secondary column gave poor platinum, palladium and rhodium recoveries although the batch flotation results showed that the material in the feed stream to the unit was highly floatable.

7.1.2 Chromite recoveries

The performance of each unit was not only measured in its ability to recover platinum, palladium and rhodium but also to minimise chromite recovery into the concentrate since chromite prove to be problematic in the smelting stages.

According to Liddell *et al.* (1986) chromite in UG-2 ore is recovered through physical entrainment. The results from the "down-the-bank" sampling enabled the construction of a chromite recovery versus water recovery graph which clearly shows that the observation of Liddell *et al.*(1986) about the recovery mechanism for chromite is also true in this circuit treating UG-2 ore. The batch flotation test results also show a linear relationship between chromite recovery and water recovery indicating chromite recovery occurred by entrainment.

When the batch flotation chromite recoveries were compared to the plant unit recoveries it was clear that in all cases, except for the column, the batch flotation recoveries of chromite were lower than the plant, indicating a lower extent of entrainment in the batch cell. The column, as would be expected, showed much lower chromite recoveries than the batch flotation cell. The washing of the froth in the columns proved to be effective since the chromite recoveries in these units were also less than that of other cleaners.

Since it was shown that chromite was recovered through physical entrainment in the froth , the modelling of chromite recovery will only be possible if a model for entrainment could be

developed. This was not attempted in this presented study. Future work must therefore be focused on the modelling of the froth zone and its characteristics.

7.1.3 RTD studies

Residence time distribution (RTD) studies were performed on each type of flotation machine in the circuit. The main objective of the RTD studies was to obtain the mean residence times (τ) and the mixing characteristics of each type of flotation machine in the circuit.

The primary rougher/scavenger unit could be modelled as 9 tanks in series which indicates plug flow characteristics rather than mixed flow. The low value of N_d, which according to Levenspiel (1972) indicates an intermediate amount of dispersion, is consistent with the mixing characteristics obtained from the tanks-in-series model. The fact that the mean residence time (τ) was almost the same as the nominal residence time (t) indicates that no channelling of fluid, stagnant regions or internal recirculation occurred in the primary rougher/scavenger unit (Levenspiel, 1972).

The primary cleaner unit was modelled as 13 tanks in series which indicates plug flow rather than mixed flow conditions. The N_d obtained from the axial dispersion model was 0.06 and according to Levenspiel (1972) indicates an intermediate amount of dispersion.

The primary column was modelled as a perfectly mixed tank reactor ($N = 1$). Mills (1991) observed the same occurrence for columns with a diameter of more than 1 metre. The large N_d value of 1.8 confirms that the fluid in the column can be characterised close to mixed flow. Large variations in the mean residence times were due to unstable operation as a result of variations that occurred in the feed flow rate to the unit. The RTD curve obtained for run 1 (Figure 4.5) shows an oscillation pattern which indicates internal recirculation (Levenspiel, 1972). This is to be expected since the column was not equipped with baffles which could improve mixing in the unit. Baffles are recommended for columns with a diameter of more than 1 metre (Dobby and Finch, 1990).

The secondary cleaner was modelled as 4 tanks-in-series. The N_d value of 0.12 indicates a large amount of dispersion(Levenspiel, 1972) and confirms the tanks-in-series model result

that the fluid could be described with mixed flow rather than plug flow. The τ and the nominal residence times (t̄) were identical. This indicates a normal behaviour for the fluid in the unit with no channelling, stagnant regions or internal recirculation.

7.2 Characterisation of the ore

The floatability parameter model was used to model the batch flotation results. The model derived to fit the data consisted of three floatability classes namely a fast floating, a medium floating and a slow floating and a class for valuable minerals which did not float. Batch flotation tests were performed on samples from the feed stream to each unit. The modelling of the batch flotation results therefore characterised the ore that was fed to each unit. The model provided a flotation rate constant for each floatability class and distributed the valuable minerals into the four classes.

The application of the floatability parameter model in this study assumes that the rate constants associated with each floatability class remain unchanged in all the batch flotation tests and that each class has a constant mean particle size and mineralogy or hydrophobicity. Only the proportion of each floatability class varies depending on the sampling location in the circuit. This implies that the model assumes that a valuable mineral will float with the same rate wherever it is in the circuit if floated under the same conditions with the same procedure and equipment.

The results of modelling the batch flotation data correspond to what would be expected from the ore in the feed stream to each unit. The ore is thus well characterised by treating it as material with four discrete fractions, viz. fast, medium, slow floating and non-floating material. The model, however, did not discriminate between mineralogical and particle size effects on flotation rates. This would be an important aspects of a future investigation.

7.3 Simulation of the circuit

The performance of each unit was modelled individually, by using the unit recoveries, the floatability fractions from the modelling of the batch flotation data (viz. M-values), the batch flotation rate constants and the mixing parameters. This has been shown diagrammatically

The rate constants used for the primary column were 6.4 times slower than the batch flotation rate constants. If compared to the rate constants used for the modelling of the primary rougher/scavenger then the primary column produced rate constants 4.7 times slower than the rougher/scavenger. A hypothesis concerning the difference in bubble size (d_b) and particle collection efficiency (E_k) in the two units was offered as an explanation for the slower flotation rates in the primary column. The larger bubble size and the lower particle collection efficiency of fine particles in the column compared to the Wemco rougher/scavenger may account for the lower flotation rate constants calculated for the column cells.

7.3.2 Simulation of the circuit performance

The simulation of the Phase 2 circuit (Figure 2.2) proved to be successful in terms of overall platinum, palladium and rhodium recovery. The distribution of these platinum group metals in the circuit was also obtained from the simulation.

The main advantage of the simulation is that it can predict the change in performance of the circuit in terms of platinum, palladium and rhodium recovery if the floatability characteristics of the ore should change. This predictive power could be of immense benefit in the testing of various parameters such as grind which would alter the particle size of the ore and therefore the floatability of the ore in the batch cell. These newly obtained floatability characteristics can then be used in the simulation and the performance of the circuit could be predicted before testing the new conditions on the plant.

The difference in the three simulated circuits was in the cleaning stages. It was shown that the best platinum, palladium and rhodium recoveries could be obtained if the greatest cleaning capacity was situated in the secondary stage of the circuit. It was evident from the simulation of the three circuit that the Dorr Oliver cleaner had a significant impact on the recovery in the circuit. When the Dorr Oliver cleaner was removed from the circuit in Figure 5.14 the platinum recovery dropped by 1.5% and when the Dorr Oliver cleaner was placed in the secondary cleaning circuit as a cleaner (Figure 5.13) the platinum recovery increased approximately 3%.

Although the changes in the circuit configuration could be simulated to establish the effect

The rate constants used for the primary column were 6.4 times slower than the batch flotation rate constants. If compared to the rate constants used for the modelling of the primary rougher/scavenger then the primary column produced rate constants 4.7 times slower than the rougher/scavenger. A hypothesis concerning the difference in bubble size (d_b) and particle collection efficiency (E_k) in the two units was offered as an explanation for the slower flotation rates in the primary column. The larger bubble size and the lower particle collection efficiency of fine particles in the column compared to the Wemco rougher/scavenger may account for the lower flotation rate constants calculated for the column cells.

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Although the changes in the circuit configuration could be simulated to establish the effect

on the platinum, palladium and rhodium recoveries, the effect of the changes on the chromite recovery could not be simulated. Since the chromite recovery is crucial for the design of optimum cleaning circuits in the flotation of UG-2 ore it is strongly recommended that research be done in the field of entrainment.

The simulation of the three different circuits (Figures 2.1, 2.2, 5.14) was possible since the same primary and secondary grind, feed pulp density and reagent addition rates were applicable to all three circuits. The effect of grind on the platinum, palladium and rhodium recoveries could be investigated by changing the grind and doing batch flotation tests on the mill product. The batch flotation results will enable the calculation of new M-values or ore floatability characteristics and the effect of the new grind could be simulated before implemented on the plant. The effect that finer or coarser grind will have on the different float fractions and the fraction of valuable minerals not floatable can also be investigated in the laboratory. The optimum grind for the secondary mill could be estimated by determining the grind where the fraction of valuable minerals not floatable can no longer be decreased as a result of milling.

8. References

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APPENDIX A

Batch flotation results: Preliminary work

This appendix consists of all the batch flotation data generated during the preliminary work. The appendix is divided into two sections. The first section presents the PGM recoveries while the second section lists all the chromite recoveries.

The tables list the mass of each sample, the mass pull or mass distribution, the assay results, the cumulative grade and the cumulative recovery. The tables also list a calculated head which is the calculated feed grade of the test. The calculation was done by using the following equation,

$$f_{calc} = \frac{\sum_{j=1}^{n} (c_j m_j) + m_t t}{100}$$

where n is the number of concentrate samples, c_j is the assayed result of the concentrate sample, m_j is the mass fraction or distribution of the concentrate sample, t is assayed result for the tailing sample and m_t is the mass fraction of the tailing sample.

The equation that was used to calculate the cumulative grades is as follow,

$$Cum\ c_i = \frac{\sum_{j=1}^{i} (c_j m_j)}{cum\ m_j}$$

where i is the sample number. The recoveries was calculated as follow,

$$R_j = \frac{c_j m_j}{f_{calc}}$$

APPENDIX B

Unit performances: Preliminary work

This appendix consists of the unit recoveries for PGM and chromite. A positive unit performance is a high PGM recovery and a low chromite (Cr_2O_3) recovery. The unit recoveries were calculated by making use of the two or three product formula. The two product formula is as follows,

$$R = \frac{c}{f} \frac{(f - t)}{(c - t)} \times 100$$

where R is the recovery, c is the assay value of the concentrate, f is the assay value of the feed and t is the assay value of the tailings. The three product formula is used if there are two concentrate streams. For a flotation bank with concentrate and middlings the recovery of mineral a will be,

$$R_a = 100 \times \frac{a_c}{a_f} \frac{(a_f - a_m)(b_m - b_t) - (b_f - b_m)(a_m - a_t)}{(a_c - a_m)(b_m - b_t) - (b_c - b_m)(a_m - a_t)}$$

where the subscripts f, c, m and t are the assay values of the feed, concentrate, middlings and the tailings respectively.