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School of Chemical and Metallurgical Engineering

Beneficiation of Recycled Process Water at DRDGOLD's ERGO Brakpan Plant and Evaluating its Effect on Gold Recovery

MSc Thesis

by

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EXECUTIVE SUMMARY

In January 2019, DRDGOLD had implemented a 38 % decrease in externally sourced water, produced by Rand Water, resulting in the amplified utilisation of process water. ERGO Brakpan's Metallurgical Laboratory has however shown improved gold recoveries when utilising potable water over process water in laboratory test work. The overall impact of utilising process water is twofold; a reduction in gold recovery due to an accumulation of contaminants within the process water, as well as penalties imposed due to effluent water and tails discharge streams exceeding allowable contaminant concentrations. To improve the understanding of the effect of water quality on gold leaching and gold recovery, strategies for the beneficiation of the process water at the ERGO Brakpan gold mine operations are presented.

Laboratory scale leach test work showed a significant improvement in gold recoveries between the use of potable water, used as a benchmark, and untreated process water. A two-sample t-test undertaken on the respective leach results confirmed a significant difference in gold recoveries between the use of potable water and untreated process water.

Atomic absorption analysis of ERGO Brakpan's process water, confirmed the presence of appreciable concentrations of heavy metals identified as iron (Fe), nickel (Ni), zinc (Zn) and manganese (Mn). A study of the effect of these identified contaminants on gold recovery was undertaken by spiking potable water with a known concentration of the respective contaminant. Leach tests showed that iron, nickel and zinc have the largest negative effect on gold recovery, with iron and nickel having a greater detrimental effect on gold recovery than zinc.

Other contaminants present within ERGO Brakpan's process water were sulfate and calcium (Ca). Sulfates were shown to have a possible passivation effect, which negatively influenced gold recoveries; however, this was to a lesser extent than that caused by the heavy metals. Calcium, when found in excess, showed an increase in gold recovery indicating the possible formation of a calcium aurocyanide complex.

Lime softening was shown as efficient in the reduction of heavy metal concentration by metal hydroxide precipitation as well as showing a reduction in sulfate concentration by the precipitation of gypsum. Lime softened process water was shown to produce similar gold recoveries as potable water and was an attractive beneficiation option as it is readily available at the ERGO Brakpan plant.

The effect of lime treated process water at elevated temperatures showed improved gold recoveries; however, the heating requirement for the plants daily consumption of 60 Ml/day would not be economically viable.

An economic study for the potential full-scale installation of a lime treatment process and gypsum crystallization for treatment of 20 Ml/day of process water showed an estimated CAPEX of R49 085 809 and an estimated monthly OPEX of R1 521 408. A discounted cash flow analysis showed a Net Present Value of approximately R330 000 000 within the project life with an internal rate of return of 76 %, which far exceeded the project's discount rate of 10%.

DECLARATION

The work presented in this thesis was undertaken at the School of Chemical and Metallurgical Engineering, University of the Witwatersrand in Johannesburg, South Africa, from February 2019 until September 2020.

All work presented in this thesis is original unless otherwise stated. It has neither in whole nor part been submitted previously to any other University or Institute as part of a degree.

I, Arshir Narain, student number 705882, declare that:

1. The research reported in this thesis, except where otherwise indicated, is my original research.

2. This thesis has not been submitted for any degree or examination at any other university.

3. This thesis does not contain other persons' data, pictures, graphs or other information, unless specifically acknowledged as being sourced from other persons.

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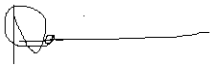

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- Lastly, I would like to thank the personnel from the MAED Analytical Lab for their assistance with procurement of required chemicals, assisting with the atomic adsorption analysis as well gold assays required.

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NOMENCLATURE

Symbol	Description	Unit
AAS	Atomic Absorption Spectrometer	
AMD	Acid Mine Drainage	
CAPEX	Capital Expenditure	
CIL	Carbon-in-leach	
CIP	Carbon-in-pulp	
DCF	Discounted Cash Flow	
g/t	Grams of gold per ton of solids	g/t
ICP-MS	Inductively coupled plasma mass spectrometry	
ICP-OES	Inductively coupled plasma optical emission spectroscopy	
LR	Laboratory Reagent grade (Purity ≥ 85%)	
M	Molarity	mol/l
MAED	Mechanical Analysis and Engineering Design Metallurgical Laboratory	
mg/l	Concentration	mg/l
MW	Molecular weight	g/mol
n	Number of moles of species	mol
NPV	Net Present Value	
P	System operating pressure	bar
OPEX	Operating Expenditure	
Q	Volumetric flow rate	m^3/s
RO	Reverse Osmosis	
Senfroth	1,1,3 Triethyloxybutane	
Senfroth	Alkyl Dithiophosphate	
SG	Specific Gravity	
SNPX	Sodium Normal Propyl Xanthate	
t	Time	seconds/minutes
TSF	Tailings Storage Facility	
UF	Ultrafiltration	
V	Molar Volume	m^3
XRD	X-Ray Diffraction	
% S	Percentage Solids	

GLOSSARY

Word	Description
Dissolution	The amount of gold leached from the solids into solution
DRDGOLD	DRDGOLD Limited is a mid-tier, unhedged gold producer and a world leader in surface gold tailings retreatment
Eluate	A solution of pregnant gold obtained during the process of elution
Eluent	A solution comprised of cyanide (CN) and sodium hydroxide (NaOH) used to remove adsorbed gold from activated carbon by means of washing
Elution	Elution is the process of extracting adsorbed gold from activated carbon by washing with a solvent (eluent)
ERGO	ERGO Mining (Proprietary) Limited – refers to the entire central and eastern mining operations
ERGO Brakpan Plant	DRDGOLD's flagship gold processing plant located in Brakpan
Flotation	Flotation is a process for selectively separating hydrophobic materials from hydrophilic
Fire Assay	Analysis technique used to determine the precious metal content within silver and/or gold ores.
Leach	The extraction of gold from the 4L50 ore into a cyanide rich solution
MAED	The metallurgical laboratory that undertook the solids, solution and carbon gold loading analysis
Process Water	Water that is recycled and reused within the gold leaching process
Potable Water	Municipal supplied water (supplied by Rand Water)
Washed Residue	Refers to the gold grade of the solids within tails which has been filtered, re-pulped with potable water and filtered again, prior to assay. This allows for free gold to be removed/washed off the solids thus ensuring accurate assay readings

CHAPTER 1: INTRODUCTION

1.1 BACKGROUND ON DRDGOLD ERGO MINING PROPRIETARY LIMITED (ERGO)

DRDGOLD is a South African gold producer and a world leader in the recovery of gold from the retreatment of surface tailings. DRDGOLD ERGO Mining (Proprietary) Limited (ERGO) specialises in retreatment of gold surface mine tailings, with numerous operations and tailings reclamation sites spanning 163 km from the western to the central and eastern regions of the Witwatersrand. The ERGO operations, illustrated in Figure 1.1, have two main plants, namely the Knights Plant located in Germiston and the flagship plant located in Brakpan, while City Deep, a former plant, has been re-commissioned as a pump station. In addition, DRDGOLD concluded its acquisition of Far West Gold Recoveries (FWGR), previously the West Rand Tailings Retreatment Project, from Sibanye-Stillwater at the end of July 2018. The following research project focuses on the ERGO Plant operations in Brakpan.

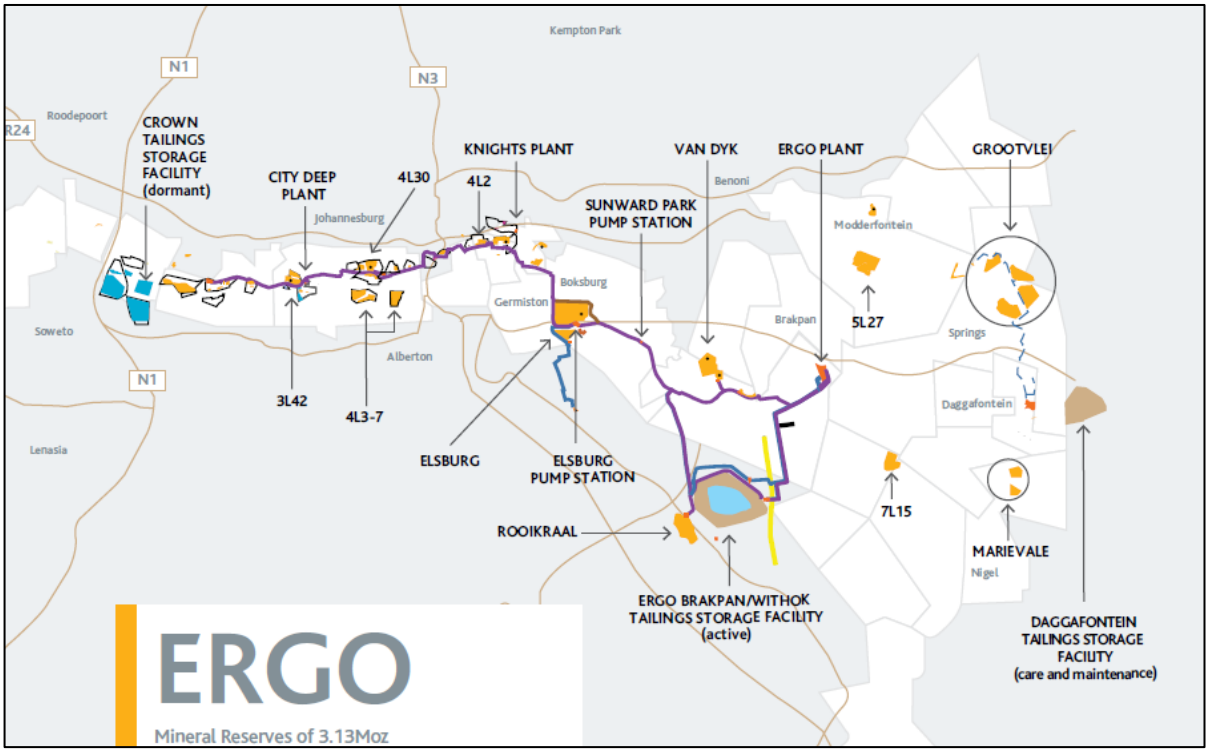

Figure 1.1: DRDGOLD operations spanning 163 km from the western to the central and eastern regions of the Witwatersrand (DRDGOLD Limited, 2019)

ERGO Mining Operations (Proprietary) Limited is 74% owned by DRDGOLD Limited and 26% by Broad Based Black Economic Empowerment (BBBEE) partners – Khumo Gold SPV Proprietary Limited (Khumo) and the DRDSA Empowerment Trust (DRDGOLD Limited, 2019). DRDGOLD is

one of the oldest continuously listed companies on the JSE, with a secondary listing on the NYSE (DRDGOLD Limited, 2019). A breakdown of the DRDGOLD structure is illustrated in Figure 1.2.

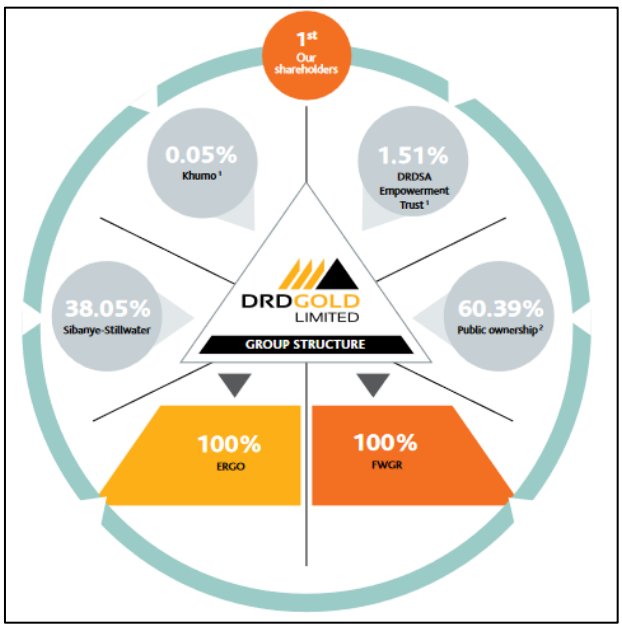

Figure 1.2: A breakdown of the DRDGOLD structure (DRDGOLD Limited, 2019)

ERGO's flagship plant is situated roughly 50 km east of Johannesburg, in Brakpan. The ERGO Brakpan plant aims to produce approximately 250 kg of gold by treating 1.8 to 2.0 Mt of material per month (reclaimed from several mine tailing sites). Figure 1.3 illustrates a simplified block flow diagram of the ERGO Brakpan operations.

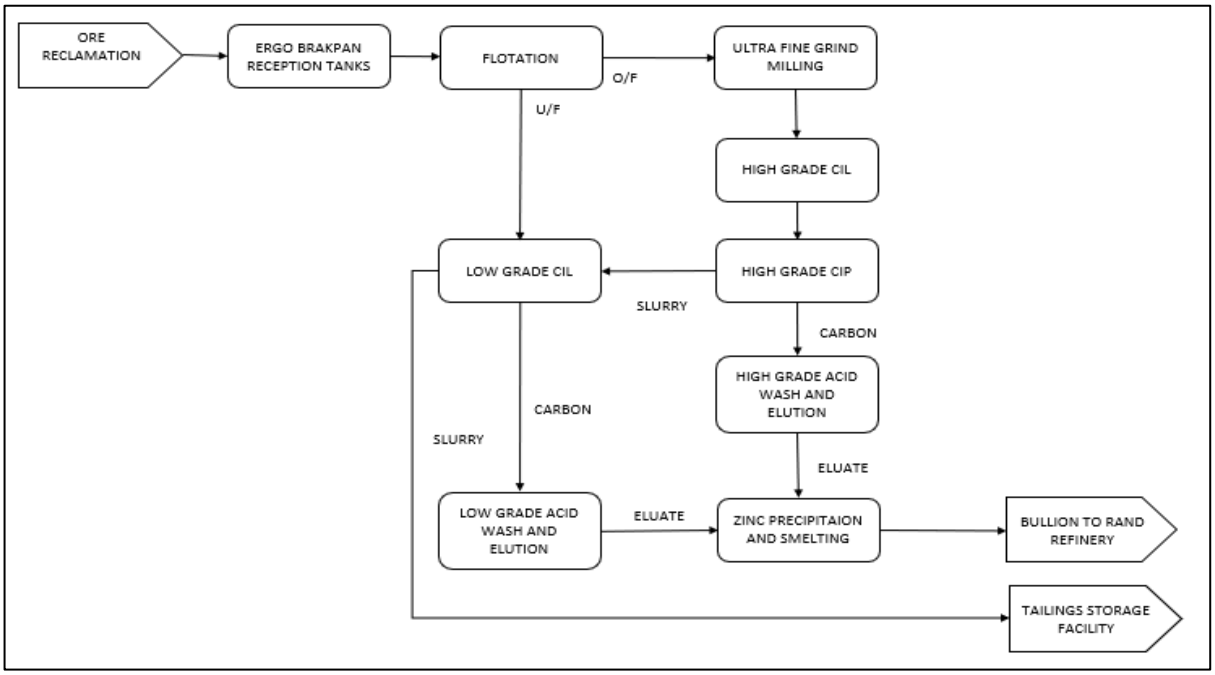

Figure 1.3: Block flow diagram of ERGO Brakpan operations

The gold recovery process proceeds by initially floating the feed material, thus separating the high-grade pyrite concentrate from the low-grade float tails (DRD Gold, n.d.). The concentrated pyrite proceeds to the high-grade plant, where the pyrite is milled in a series of fine-grind vertical mills prior to being leached in a unique combined carbon in leach – carbon in pulp (CIL-CIP) leach circuit. Simultaneously, the float tails are sent to the low-grade plant where they are treated in a carbon in leach (CIL) circuit. Utilising cyanide as a lixiviant, the gold loads onto activated carbon, from both the high-grade and low-grade plant leach circuits, respectively. The gold is recovered through elution of the loaded carbon, followed by zinc precipitation of the eluate before being smelted into bullions, which proceed to Rand Refinery.

Owing to the nature of the ERGO Brakpan operations, which focus on retreatment of surface mine tailings, vast quantities of material are treated daily (approximately 64 000 t/day), which in turn requires approximately 60 Ml/day of water. Since water is a strategic resource for our water scarce country, several mining operations are at risk due to water restrictions leading to limited supply of water. This has caused an amplified utilisation of "grey water" with the focus placed on recycling and reusing process water. Therefore, from a financial and an environmental standpoint, it would be beneficial for ERGO Brakpan to recycle its process water.

Approximately 60 % (36 Ml/day) of water utilised at ERGO Brakpan is recovered and re-used from the thickener overflows at the plant as well as the main tailings' deposition site. This figure includes harvested rainwater and run-off water. The remaining 24 Ml/day required by ERGO Brakpan are obtained from two important 'grey' water sources. Between 8 Ml to 12 Ml/day are sourced from the Trans-Caledon Tunnel Authority neutralisation plant, which treats acid mine drainage on behalf of the government, while the balance is treated sewage, which is drawn from the recently completed Rondebult treatment plant (DRD Gold, n.d.).

DRDGOLD's CEO Mr Niel Pretorius had implemented a 38 % decrease in externally sourced water produced by Rand Water from January 2019. This has resulted in a greater need to recycle and reuse process water. The current operation at the ERGO Brakpan plant recycles process water by means of thickeners. The flotation concentrate is sent to two 55 m diameter high grade thickeners and the flotation tails are sent to four 135 m diameter low grade thickeners. The thickeners operate by continuous gravity settling (sedimentation) of the solids in suspension. The thickened solids are removed together with a portion of the water as thickened underflow. The remaining water, ideally containing no solids, forms the "overflow" from the thickener. This overflow is pumped to ERGO Brakpan's settling pond for reuse in other plant processes. While this may allow larger suspended solids to settle, some inorganic contaminants (such as heavy metals) as well as organic contaminants (such as flotation reagents) are still present. This has resulted in numerous downstream problems such as chokes and blockages around the plant and an increased risk of these undesirable metals being precipitated during zinc precipitation.

The Metallurgical Laboratory at ERGO Brakpan performs weekly flotation and leach tests on each of the feed streams with the aim of monitoring plant performance. These tests simulate the plant operation in a smaller, more controlled environment, while attempting to keep all dependent variables such as leach residence time, reagent strength and addition rate, as well as carbon concentration, close to those of the plant operation. The only exception is that the laboratory utilises potable water, while the plant utilises process water due to the high purchase expense of potable water. The results obtained from the laboratory tests showed lower washed residues (gold loading on the tails solids) than those achievable by the plant. In addition, higher gold recoveries were observed from the Metallurgical Laboratory tests compared to plant recoveries.

Therefore, it is imperative to identify contaminants within the process water that negatively affect plant recoveries and to determine an effective purification technique to remove these contaminants. A comparison of the achievable gold recoveries in both the leach and float circuits while utilising potable water, process water and cleaned process water was also required. A clearly formulated action plan related to water management and treatment, the associated recovery losses as well as environmental liabilities would be of benefit to enable the continued, sustainable operation at ERGO Brakpan into the future.

1.2 PROBLEM STATEMENT

Although the results for weekly leach and flotation laboratory tests conducted by the Metallurgical Laboratory at ERGO Brakpan show improved gold recoveries which highlight the benefit of utilising potable water over process water, there has been no previous investigation into quality of the process water and possible purification options that might be considered at the ERGO Brakpan plant. Based on the contaminants identified in ERGO Brakpan's process water, it would also be important to understand and compare the effects that potable water, process water and cleaned process water may have on the performance of the leach and float circuits at ERGO Brakpan.

An understanding of the effects, of using the different waters will facilitate the implementation of optimisation processes at the ERGO Brakpan plant. Thus, the primary objective of this research is to analyse ERGO Brakpan's process water to identify key contaminants, with subsequent removal of contaminants that cause suppression of gold recovery. An investigation of the impact that different water sources (potable water, process water and cleaned process water) have on both the floatation and leach circuits respectively, will also be undertaken.

1.3 RESEARCH QUESTIONS

In order to achieve a better understanding of the effect of water quality on gold recovery, as well as to guide the lab-scale test-work presented within this thesis, the following research question were posed:

- What are the predominant contaminants within the process water at ERGO Brakpan?
- Are the contaminants organic and/or inorganic in nature?
- Which of the contaminants present in the process water cause suppression of gold recovery?
- What is the most efficient method to remove the contaminants identified above?
- What effect does the cleaned process water, compared to potable water have on gold recovery?
- Can the lab scaled test-work for the removal of contaminants from process water be scaled up for plant use?
- Will potential improved gold recoveries outweigh the economic costs of process water purification on a plant scale?

1.4 RESEARCH OBJECTIVES

The research questions were used to develop the research objectives.

To gain a better understanding of ERGO Brakpan's process water makeup; the following objectives have been defined for this project:

- To analyse potable water and various sources of process water to determine the contaminants within the ERGO Brakpan process water.
- To determine which potential contaminants in the process water, cause suppression of gold recovery.
- To undertake laboratory studies to develop a process(es) to selectively remove such contaminant(s).
- To compare the gold recoveries of the cleaned process water to potable water (which will be utilised as a benchmark).
- To design a treatment process(es) which will effectively remove gold suppressing contaminants that may be implemented on a plant scale.
- To evaluate the Net Present Value (NPV) of the project to determine if the potential improvement in gold recoveries would outweigh the economic costs of cleaning the process water on a plant scale.

1.5 SCOPE OF THE STUDY

The primary research method for this study was a quantitative scientific investigation resulting in the generation of statistically analysable data by systematic manipulation of independent variables. This research methodology allowed one to make inferences about the relationship between independent and dependent variables. The primary research method was supported by a literature review which assisted in guiding the selection of the experimental design employed.

In order to meet the objectives associated with the fulfilment of the project, the following tasks were carried out:

1. An extensive literature review was performed to understand the conventional technologies available for mine wastewater treatment and their performance. Inductively coupled plasma mass spectrometry (ICP-MS) was used to provide a comprehensive analysis of ERGO Brakpan's process water to determine the presence of contaminants with the highest concentrations. This provided insight into the effects that the identified contaminants had on gold leaching and flotation.
2. Once the predominant contaminants had been identified, the apparatus required for the laboratory scale test work was designed and commissioned. This included a preliminary control experiment where potable water and untreated process water were used for leach tests as well as flotation tests, followed by leaching of the flotation tails. This provided insight into the highest achievable gold recoveries using the different water sources.
3. Experimental runs were repeated in triplicate to determine whether significant differences in gold recoveries were evident when using potable water and untreated process water. Sufficient replication of experimental work was vital to facilitate a statistical evaluation of data, which would strengthen the validity of conclusions made in terms of beneficiation options.
4. Thereafter, potable water was spiked with increased concentrations of individual contaminant species to determine the effect on gold recovery of each contaminant species. This served to isolate an individual ion/group of ions and allowed for more comprehensive beneficiation approach rather than following a "blanket" overall treatment approach.
5. Thereafter, specialised reagents were added to ERGO Brakpan's untreated process water to selectively and chemically precipitate contaminants that were identified to have the largest effect on gold recovery. The concentrations of the contaminants were measured before and after beneficiation to determine the degree of contaminant removal. This was accomplished by use of an Atomic Absorption Spectrometer (AAS), which detected the metal contaminant concentrations. X-Ray Diffraction (XRD) was use for analysis of the crystalline compounds present in the precipitants formed.

6. The treated process water was used in subsequent flotation and leach tests to determine the efficiency of the proposed beneficiation options. The overall gold recovery was determined by fire assay analysis to determine the amount of gold available before and after flotation and leaching.
7. The Net Present Value (NPV) of the project was calculated to determine if the revenue from the increased potential gold recovery will outweigh the economic costs of implementing the proposed beneficiation options on a plant scale.

1.6 OUTLINE OF THESIS

This thesis comprises of six chapters. Chapter 1 provides an introduction into the thesis, highlights the pertinent research objectives as well as provides the scope of study. An overview of the five ensuing chapters are as follows:

Chapter 2: Theoretical Background and literature review

This chapter provides a plant overview of the operations at ERGO Brakpan as well as a theoretical background that highlights the conventional technologies that are available for mine wastewater treatment. An extensive literature review detailing the effect that the presence of various metal cations as well as inorganic contaminants, within sulfide ore slurries, has on gold recovery is also presented.

Chapter 3: Methodology and Equipment

This chapter describes the experimental design and procedures undertaken to remove contaminants from ERGO Brakpan's process water under laboratory conditions. The experimental procedures for both the initial leach tests as well as flotation tests, followed by leaching of the flotation tails tests are described under laboratory conditions. The purpose of the experiments was to confirm the removal of contaminants by chemical precipitation under conditions specified in literature as well as to determine the associated effects on gold recovery.

Chapter 4: Results and Discussions

This chapter provides a statistical analysis comparing the gold recoveries between potable water and process water. Laboratory scale leach test results are also presented where potable water was spiked with known concentrations of identified contaminants within ERGO Brakpan's process water, with the aim to determine ion(s) that have the largest detrimental effect on gold recovery. The use of lime and soda ash softening as viable beneficiation options, at various operating temperatures, are presented with the resultant effect on gold recovery.

Chapter 5: Plant scale viability and Economic Evaluation

This chapter summarises the potential installation of a full-scale lime treatment process followed by gypsum crystallization. The plant design, proposed plant layout, as well as the capital expenditure (CAPEX) and operating expenditure (OPEX) are presented for treatment of 20 Ml/day of process water. A Discounted Cash Flow (DCF) analysis as well as Net Present Value (NPV) evaluation of the proposed treatment are also presented.

Chapter 6: Conclusions and Recommendations

This chapter summarises the relevant findings of this thesis and provides recommendations for future work.

CHAPTER 2: THEORETICAL BACKGROUND AND LITERATURE REVIEW

2.1. FOCUS OF THE THEORETICAL BACKGROUND AND LITERATURE REVIEW

The use of cyanide as a lixiviant for gold leaching has been used globally since it was first patented over 100 years ago. There have been vast advances in gold recovery over the last century, most notably the development and implementation of carbon-in-pulp processing in the 1970's (Rees, 2000). However, the processing and extraction of gold has become more complex as simple free milling oxide ores have depleted over the past decade (Rees, 2000). Therefore, processing improvements are required so that more complex refractory ores become viable to mine. A growing concern in industry is the vast quantities of water required to maintain optimal slurry densities required for optimal gold leaching. Since water is a strategic resource for our water scarce country, several mining operations are at risk due to water restrictions leading to a limited supply of water. This has caused an amplified utilisation and focus placed on recycling process water. There is a vast array of literature in the field of gold extraction using cyanide, with a variety of diverse subject areas, however, limited literature is available on the effects of water quality on gold leaching.

This chapter reviews the literature on the gold cyanidation process and provides an overview of the gold recovery process at the ERGO Brakpan plant. The importance of water in both the mining and the processing of gold ores are highlighted.

An extensive review of the literature highlighting the effects on gold recovery by cyanidation leaching in the presence of contaminants such as heavy metals ions, as well as inorganic contaminants such as calcium and sodium is presented. In this thesis, it is essential to first review the literature on the effects that the presence of contaminants have on gold extraction as well as the effects of utilising process water in gold extraction, so that areas requiring further attention, or that have been neglected, can be identified.

An extensive review of the available water purification techniques predominantly utilised in industry is also presented. There is a large selection of literature in the field of water treatment techniques, therefore in this review, only the papers specifically relevant to this field of study have been analysed.

2.2. OVERVIEW OF THE GOLD RECOVERY PROCESS AT THE ERGO BRAKPAN PLANT

Figure 2.1 illustrates a simplified process flow diagram of the ERGO Brakpan gold processing plant. A detailed process flow diagram of the ERGO Brakpan gold plant is present in Appendix A, Figure A.1.

2.2.1. Feed reception and flotation

There are four main feed streams of reclaimed tailings that enter the flagship ERGO Brakpan plant. These are deposited into three reception tanks at a specific gravity (SG) of 1.45 (49 % solids). Each of the reception tanks acts as a dilution tank, where process water is added to dilute the plant feed to a SG of 1.32 (37 % solids), which is the required density for flotation. The diluted material proceeds into three flotation streams, each consisting of two banks of five primary, five secondary, four tertiary-1 and four tertiary-2 flotation cells. The core operating function of the flotation cells is to recover pyrite concentrate in the overflow launders while the tails from each cell flows to the next cell, creating a rougher-scavenger flotation circuit. The concentrate overflow is transported to the high grade unique CIL-CIP circuit, whereas the flotation tails are sent to the low grade CIL circuit.

2.2.2 ERGO Brakpan's low grade CIL circuit

The flotation tails proceed to a thickener distribution box, which feeds the tails to one of four 135 m thickeners. The purpose of the thickeners is to adjust the tails density to a desired SG of 1.45 (49 % solids) and the thickeners overflow (clear process water) is transported to a settling pond for reuse in plant operations. The underflow from the thickeners (at a SG of 1.45) is transported to a "de-sanding circuit", where large debris such as stones and sticks (and other organic material) is removed by a series of four linear screens. The de-sanding circuit underflow is fed to the low-grade carbon in leach (CIL) circuit, where the feed is split into two streams, A and B.

Each stream of ERGO Brakpan's low grade CIL circuit consists of eleven 2000 m^3 tanks, with the first tank of each stream acting as a preconditioning tank where slaked lime ($Ca(OH)_2$) is added to increase the pH from approximately 8 to 10.5. The lixiviant (cyanide) as well as liquid oxygen is added into the second tank in each leach stream. Regenerated and fresh carbon is introduced into the last tank of each leach stream and flows in a counter-current direction to the slurry until the third tank in each stream. The "loaded carbon" is then removed and sent to an acid wash and elution circuit to clean and strip the gold off the carbon respectively. The slurry tails from the last tank from each CIL leach stream proceed to the plant residue before being pumped to the main tailings depositing dam.

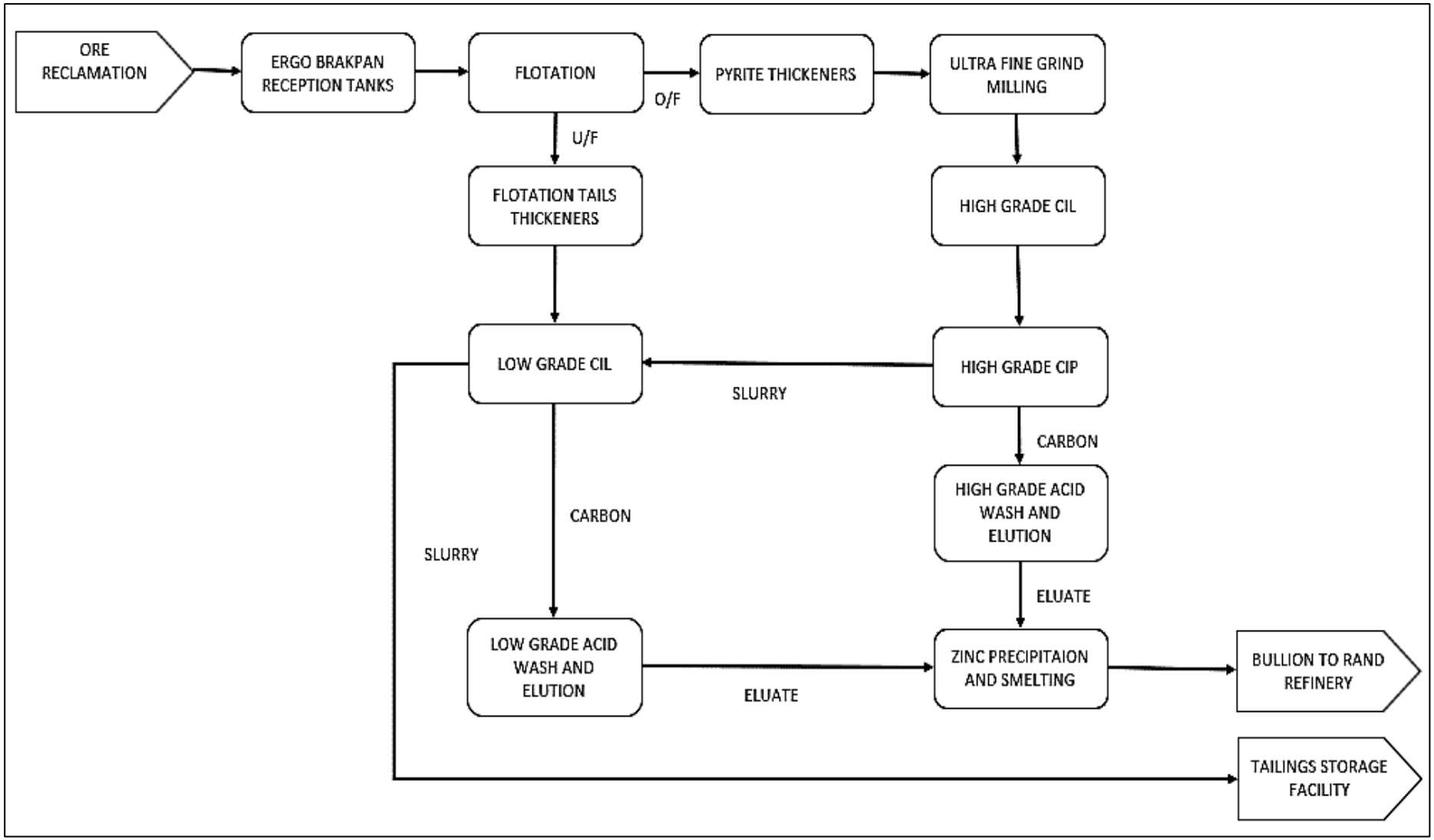

Figure 2.1: Summarised process flow diagram of the ERGO Brakpan gold plant

2.2.3. ERGO Brakpan's CIL-CIP high grade circuit

The concentrated pyrite that overflows the primary flotation cells is sent to a 55 m high-grade thickener, while the concentrated pyrite that overflows the secondary and tertiary flotation cells is sent to a 42 m low-grade thickener. The purpose of the thickeners is to adjust the pyrite density to a desired SG of 1.45 (49 % solids). Both thickener underflows are mixed in a pyrite stock tank before proceeding to the mill feed distribution box, where the mixed concentrate is distributed between four vertical ultra-fine-grind mills. The milled product is transported to the pyrite leach preconditioning tank, where the pH is increased to 10.5 by the addition of lime. The conditioned pyrite concentrate proceeds to the pyrite leach carbon-in-leach (CIL) tanks which comprises of six 1000 m^3 tanks in series. The lixiviant (cyanide) is added to the first two tanks while liquid oxygen is added to the first three tanks. Both the slurry and carbon from the tails tank (tank six) in the high grade CIL circuit proceeds to a catch tank in the carbon-in-pulp (CIP) circuit. Carbon is pumped from the CIP head tank to tank four of the pyrite high grade CIL circuit. The carbon proceeds in a counter current direction, from tank four to tank three in the high grade CIL circuit before proceeding to tank 6 and subsequently to the CIP catch tank.

The CIP circuit consists of six tanks in series. The carbon that reports to the CIP catch tank is therefore loaded with gold and proceeds to the high-grade elution circuit. The high-grade circuit targets one elution per day resulting in a new tank taking up the role as the head tank, while the previous head and catch tanks become the new catch and tails tank respectively. This rotation of head, tail and catch tanks occurs daily with high activity regenerated and fresh carbon added to the CIP tails tank. The slurry circulates through the six tanks in the CIP circuit before proceeding to the low-grade CIL stream B.

2.2.4. Acid washing, elution and zinc precipitation

Carbon from both the low and high-grade circuits undergoes an acid wash using 5 % hydrochloric acid (HCl) to remove both organic and inorganic foulants before being neutralised with 5 % sodium hydroxide (NaOH). The neutralised carbon undergoes elution with the pregnant eluate solution sent to eluate tanks. The spent carbon is regenerated in kilns before being quenched with process water and recycled back into the circuit. The pregnant eluate proceeds to the newly commissioned smelt house that utilises zinc precipitation to precipitate gold from the eluate. The resulting precipitate is filtered, and the filtered cake is calcined before being smelted to form gold bullion which is sent to the Rand Refinery.

2.3. WATER IN THE MINING INDUSTRY

2.3.1. Water intensive gold extraction processes

Water is a vital resource required for the following gold extraction processes (Sub-Committee of Consulting Metallurgists of the Minerals Council of South Africa, 1987):

- Metal extracted by acid or alkali attack (leaching)
- Solid-liquid separation (filtration/ thickening)
- Metal concentration (milling)
- Mining processes (reclamation and drilling)
- Tailings storage facility seepage

These processes are predominantly undertaken at ambient temperatures to limit corrosion. The most important constituents in recycled mine process water are the organic matter, calcium, sulfate as well as colloidal silica and suspended solids. Predominant contaminants within effluent tailing streams are typically aluminium, gold, uranium and zinc (Bester, 2012).

2.3.2. Water utilised for ore extraction – mining processes

The mining process starts with excavation of the reef plane (by drilling and blasting of horizontal tunnels) and is followed by establishing connecting tunnels. Mining of the reef requires widening of the winze-raise connections and is achieved by drilling and blasting to form slopes. Water is continuously fed down a central channel to achieve wet drilling. Conditions in stopes are almost always wet because of water ingress from rock fissures. Wet drilling is desired as the fines from broken rock are very quickly converted to mud. The ore, converted to slurry contains approximately 8 % water and is transported to the surface for further treatment. Excess water is collected at the base of the shaft and is pumped to the surface for re-use (Sub-Committee of Consulting Metallurgists of the Minerals Council of South Africa, 1987).

2.3.3. Water utilised for milling - metal concentration

The valuable constituent in gold ores normally occupies only an extremely minor proportion of the total ore volume, thus resulting in the predominant use of milling circuits. To ensure optimal gold recovery, the ore is milled so that 50 % of the feed material passes through 75 µm to ensure sufficient gold liberation. Dilution water is added to the mill feed to ensure a constant 2:1 water-to-solid ratio is maintained (Bester, 2012). The water dilution ratio influences the efficiency of the classification system. Generally, the basis of control is the adjustment of the cyclone feed water-to-solids ratio. The aim of

the classification system is to control the size fraction of the mill product by returning oversize particles to the mill, which mixes with incoming un-milled ore (Bester, 2012).

2.3.4. Recycling process water with thickeners – solid/liquid separation

While the milling and grinding process requires a very high ratio of liquid to solids, further downstream processes require that the liquid content of the pulp is low. Therefore, the solids must be dewatered by separating water from pulp, which is achieved using continuous thickeners. The feed pulp is introduced at the centre of the thickener and the settled solids are continuously removed as the underflow. The overflow waters from the thickeners is recycled and re-used in the milling process or other downstream processes (Bester, 2012).

2.3.5. Leaching - metal extracted by acid or alkali attack

The thickened underflow is leached to dissolve the gold from the solids into an aqueous solution. Leaching is achieved by contacting the gold ore with a lixiviant. A soluble cyanide salt such as sodium cyanide (NaCN) is the preferred and predominant lixiviant utilised in this industry, due to its high gold selectivity (Bester, 2012). Lime is added to the thickener underflow prior to the introduction of sodium cyanide to increase the slurry alkalinity and thus minimize the formation of poisonous hydrogen cyanide gas (HCN), which occurs under acidic conditions. The control of the pulp dilution is very important in the leaching vessels. If the dilution level is too low, the pulp viscosity inhibits mass transfer of both reagents and products with a resultant adverse effect on the dissolution process (Bester, 2012). Conversely, should the pulp be excessively diluted then the overall residence in the leaching plant is reduced with the same effect. In most plants the leaching section is designed to operate at a water-to-solids ratio in the range 1:1 to 2:1 (Bester, 2012). The thickener underflow pulp usually produces a water-to-solids ratio of approximately 3:5; therefore, additional dilution is required before entering the leaching circuit (Bester, 2012).

The carbon-in-pulp (CIP) process recovers gold in solution by contacting activated carbon with the pulp and separating the two by screening. The slurry/pulp is leached by use of several adsorption vessels in series where the pulp flows continuously from the first tank to the last, while carbon flows in a counter current direction from the last to the first tank. The loaded carbon is transferred to an external screen where the carbon is recovered and washed before entering the elution circuit. Water is used to transport carbon from the adsorption tanks to the acid wash and elution tanks (Bester, 2012).

2.3.6. Water utilised in acid wash process

Calcium, as an inorganic contaminant, has a detrimental effect on activated carbon activity, as high concentrations lead to low gold recoveries by inhibiting the elution process. Its removal is therefore an essential part of the process. Removal of the calcium deposit is achieved by addition of sufficient

hydrochloric acid (HCl) and adequate contact time with the carbon. The acid treatment usually occurs before elution; therefore, the acid needs to be fully neutralised before addition of the eluent (sodium hydroxide (NaOH) and cyanide (CN) solution). This is achieved by backflushing the column with water followed by circulation of a weak sodium hydroxide solution to ensure pH is above 10 (Bester, 2012).

2.3.7. Water utilised in the elution process and electrowinning/zinc precipitation

The eluate solution, usually a weak sodium hydroxide solution (approximately 2 % NaOH strength) is pumped to the elution column to strip the gold from the carbon. Gold is recovered from the eluate by either electrowinning or zinc precipitation. The spent eluent is recycled to the elution circuit. An alternative process to recover gold from the eluate is precipitation/cementation of the gold onto zinc dust. This process was utilised by the gold industry for almost a century before the advent of carbon technology and is a well proven technology that is still used in some operations today (Bester, 2012).

2.3.8. Water utilised for carbon regeneration

Throughout the adsorption process, where the carbon is in contact with the pulp, it becomes progressively poisoned (predominantly by calcium) and loses its activity. To counteract this loss, restoration through regeneration of the carbon's activity is required. Bester (2012) states that during adsorption, both organic and inorganic adsorbates can accumulate within the porous structure of activated carbon, and the objective of regeneration is to remove the accumulated adsorbates to recover the original structure and activity of the activated carbon required for the adsorption/leach circuit (Bester, 2012). The eluted carbon is pumped to thermal regeneration (usually kilns) where three vital reactions which correspond to the three types of adsorbates on the spent carbon, occur: desorption of volatile organic compounds; decomposition of organic compounds not sufficiently volatile for desorption and pyrolysis of the remaining compounds (Bester, 2012). After regeneration, the carbon is dropped into water where it is quenched before it is returned to the circuit (Bester, 2012).

2.3.9. Water utilised for tailings storage

The separated pulp from the carbon is referred to as tailings. Approximately 60 Ml/day of water is required to pump the tailings from the plant to a tailings storage facility (TSF), also referred as a slimes dam. Water recovered from the TFS is returned to the gold recovery plant where it is reused in various processes. A substantial amount of the water is, however, lost to the environment through evaporation and seepage through the TSF into the groundwater systems (Bester, 2012). It has become common practice that water is decanted off the TSF concurrently with deposition to maintain a sufficient depth to ensure adequate water clarity (Bester, 2012). The addition of vegetation to the TSF surface is encouraged as this minimises and reduces water and wind erosion from the top of the dam (Sub-Committee of Consulting Metallurgists of the Minerals Council of South Africa, 1987).

Two common techniques employed for water control within a tailing's storage facility are a decant tower and a decant barge (TailPro Consulting, 2020). A decant tower acts as an intake structure consisting of a vertical or inclined hollow tower allowing free water to be pumped out of the tower or drain by gravity via a buried conduit (TailPro Consulting, 2020). A decant barge comprises of a large floating platform that houses pumps used to reclaim water from the supernatant pond back to the processing plant (TailPro Consulting, 2020).

2.4. EFFECTS THAT CONTAMINANTS HAVE ON GOLD FLOTATION AND LEACHING

2.4.1. Metal oxides

2.4.1.1. The effect that metal-cyanide species have on gold adsorption onto activated carbon

The recovery of gold and silver from dilute caustic cyanide solutions utilising activated carbon has been well documented (Sheya & Palmer, 1989). Activated carbon has effectively demonstrated its affinity as an efficient extractor of precious metals from cyanide leach solutions. Many gold ore deposits in South Africa are leached with cyanide. Due to the presence of other metals within the ore, the formation of cyanide complexes of antimony, arsenic, cobalt, copper, iron, nickel, thallium, and zinc may appear in the leach solution that is contacted with activated carbon (Sheya & Palmer, 1989).

Rees (2000) states that copper presents difficulties in leach circuits due to the lack of selectivity of cyanide for gold over copper. Rees (2000) further states that an ore with a maximum copper content of 1 % is considered economically viable for gold recovery from a copper-gold sulfide ore by cyanidation. Cyanide forms complexes with a number of other elements, including iron, which forms $Fe(CN)_6^{4-}$, nickel forms $Ni(CN)_4^{2-}$, zinc (Zn), forms $Zn(CN)_4^{2-}$ and silver cyanide, which forms $Ag(CN)_2^-$. The result of the formation of these complexes is a decrease in the level of free cyanide present in the slurry to drive the leaching of gold (Rees, 2000).

Rees (2000) further elaborates that copper is considered the most problematic metal in gold leaching due to the rapid formation of copper cyanide complexes which may consume a great deal of the available cyanide. Another difficulty caused by copper solubility in cyanide solutions is the negative effect on gold recovery in CIP or CIL processes. In these processes, the copper cyanide complex $Cu(CN)_2^-$ competes with aurocyanide for adsorptive sites on the activated carbon (Rees, 2000). This leads to excessive cyanide consumption and a higher leaching pH to form copper cyanide with three or four ligands, as these complexes do not strongly adsorb onto activated carbon. Unfortunately these conditions do not favour the adsorption of $Au(CN)_2^-$ onto activated carbon, so a compromise is usually made to maximise gold recovery, whilst minimising cyanide use, and other associated costs, in the CIP circuit (Rees, 2000). Boehme & Potter (1983) have determined the effects of copper, ferrocyanide, silver, and thiocyanate have on the rate of gold loading during cyanide leaching. They concluded that

as the copper concentration increased, the rate of gold loading onto activated carbon decreased (Boehme & Potter, 1983).

The impact of zinc cyanides in gold recovery was investigated by Hedley and Tabachnick (1968). They stated that the zinc cyanide complex formed was deleterious to gold dissolution (and thus gold recovery). This was related to cyanide being consumed by an excess of zinc metal, as used in zinc precipitation. Therefore, the negative effect of zinc was due to insufficient free cyanide being available for the dissolution of gold to occur (Hedley & Tabachnick, 1968).

Boehme & Potter (1983) proved that silver too had an adverse effect on the loading rate of gold in ratios of silver to gold of 1:1 and 2:1. Boehme & Potter (1983) also stated that ferrocyanide had no effect on loading rate, while thiocyanate was less detrimental than copper or silver, but a decrease in loading rate was nevertheless observed.

Fleming & Nicol (1984) demonstrated that activated carbon adsorbs organic solvents that affect the rate of gold loading. Activated carbon adsorbs small quantities of cyanide, hydroxide, calcium and iron sulfide, which contaminates the carbon, resulting in a decrease in gold loading capacity (Fleming & Nicol, 1984). In addition, metal oxides comprising cobalt, copper, iron, and nickel also adsorb onto carbon. These metals reduce gold loading by competing with the gold for adsorption sites on the carbon. Fleming & Nicol (1984) also concluded that both the presence of copper and nickel may have a stronger affinity to load onto activated carbon than gold. This, however, is dependent on the chemical conditions of the solution (Fleming & Nicol, 1984).

2.4.1.2. Solubility of metal oxides

Various metals such as zinc, iron, copper, nickel, magnesium and cadmium are often present within polluted mine water systems. Methods to remove these heavy metals include precipitation, sorption and ion exchange. Due to the ease of control and low operating costs, removal of heavy metals by chemical precipitation is often practised in industry (Damons, 2001).

Chemical precipitation may be utilised to reduce the metal concentration in solution. Reduction in metal concentration by precipitation is determined by the solubility of the various species of that metal within solution. Most metals are predominantly precipitated as metal hydroxides, sulfides and carbonates because they are relatively insoluble in this form. To achieve precipitation of metals as hydroxides, the solution pH is adjusted to range between 8 and 12. Due to the amphoteric nature of metal as hydroxides, an optimal pH exists for metals removal by precipitation (Damons, 2001).

Metal ions exit as hydrated ions in solution and these hydrated metal ions react with hydroxyl ions to form hydro complexes (Benefield, et al., 1982). The hydro complexes formed may contain one or more

metal ions (mono or polynuclear complexes). Furthermore, Benefield *et al*. (1982) have proposed the following generalised reactions that describes the hydro complex formation reactions for trivalent metal ions:

$$Me_{aq}^{3+} + OH^- \leftrightarrow Me(OH)_{aq}^{2+} \quad (2.1)$$

$$Me_{aq}^{3+} + 2OH^- \leftrightarrow Me(OH)_2{}_{aq}^{+} \quad (2.2)$$

$$Me_{aq}^{3+} + 3OH^- \leftrightarrow Me(OH)_3{}_{aq}^{0} \quad (2.3)$$

$$Me_{aq}^{3+} + 4OH^- \leftrightarrow Me(OH)_4{}_{aq}^{-} \quad (2.4)$$

The above formation reactions are utilised to construct solubility diagrams and are presented in Figures 2.2, 2.3, 2.4, 2.5 and 2.6 below. The solubility curves for $Fe(OH)_2$, $Al(OH)_3$, $Fe(OH)_3$ and $Mg(OH)_2$ are presented and indicate the effect that pH has on the associated metal solubility. The solubility curves indicate that the solubility of $Al(OH)_3$ and $Fe(OH)_3$ increases under both basic and acidic conditions. Under acidic conditions (pH below 7), both $Al(OH)_3$ and $Fe(OH)_3$ form cationic species which result in an increase in the solubility of the solid phase. Under alkaline conditions (pH above 7), they form anionic species which also result in an increase in the solubility of the solid phase. The solubility curves also indicate that $Fe(OH)_2$ and $Mg(OH)_2$ will form insoluble metal hydroxides and will precipitate at a high pH value (pH above 10) (Damons, 2001).

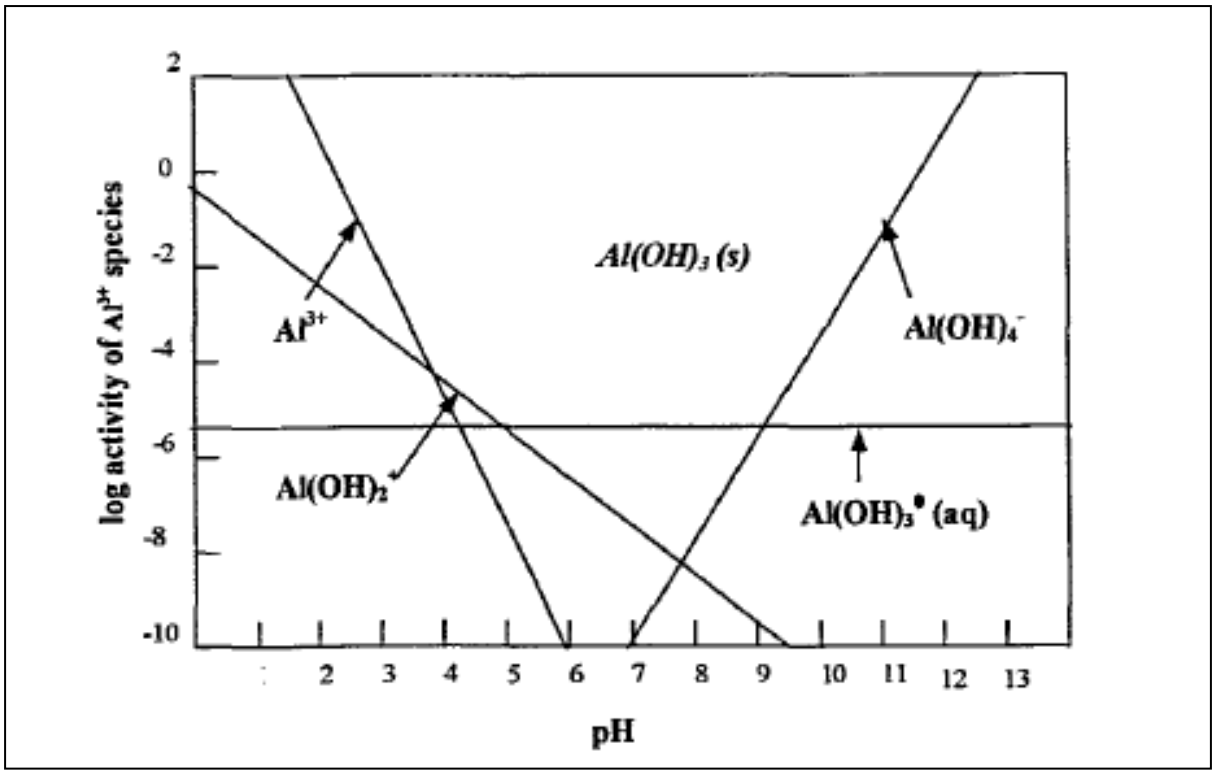

Figure 2.2: Solubility diagram for aluminium hydroxide (Damons, 2001)

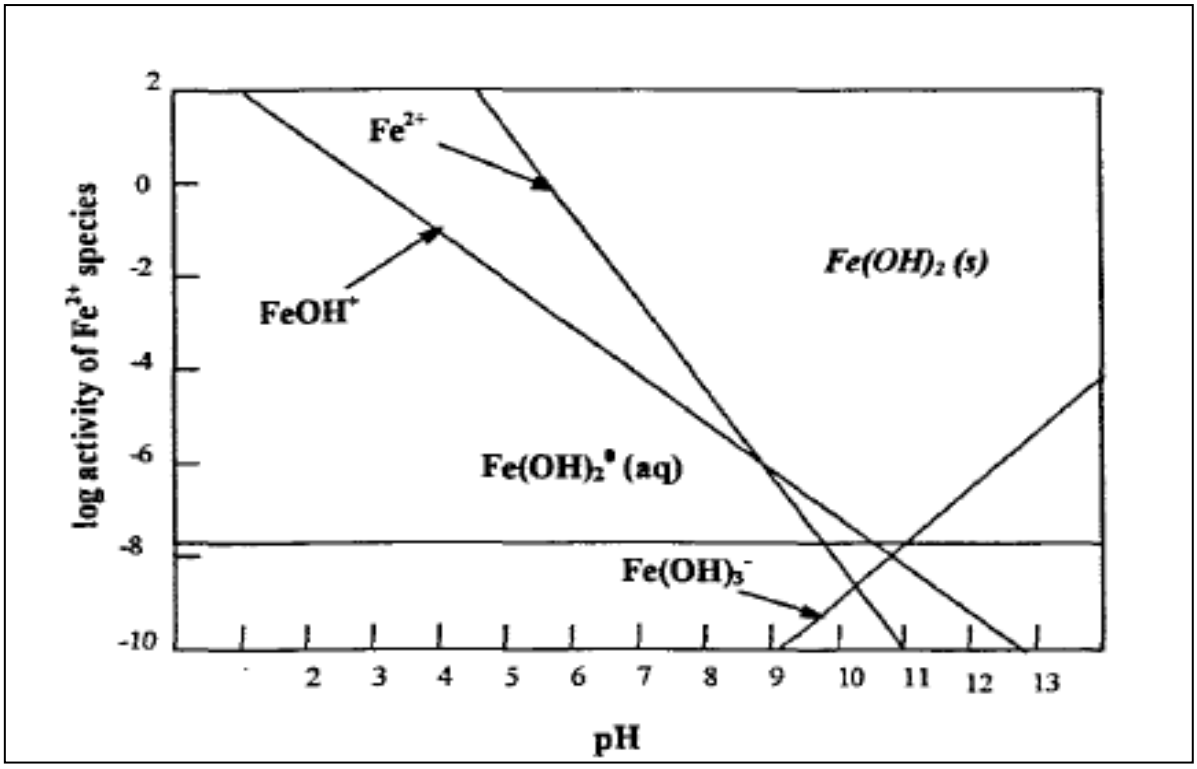

Figure 2.3: Solubility diagram for iron (II) hydroxide (Damons, 2001)

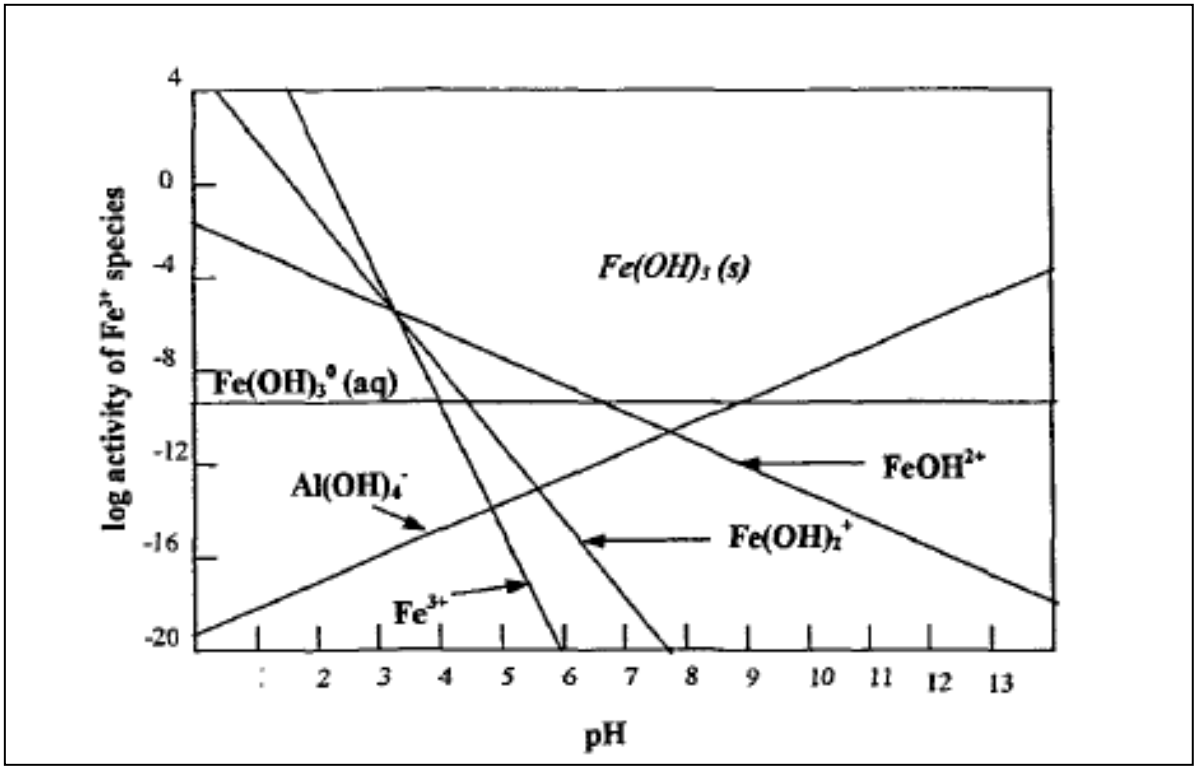

Figure 2.4: Solubility diagram for iron (III) hydroxide (Damons, 2001)

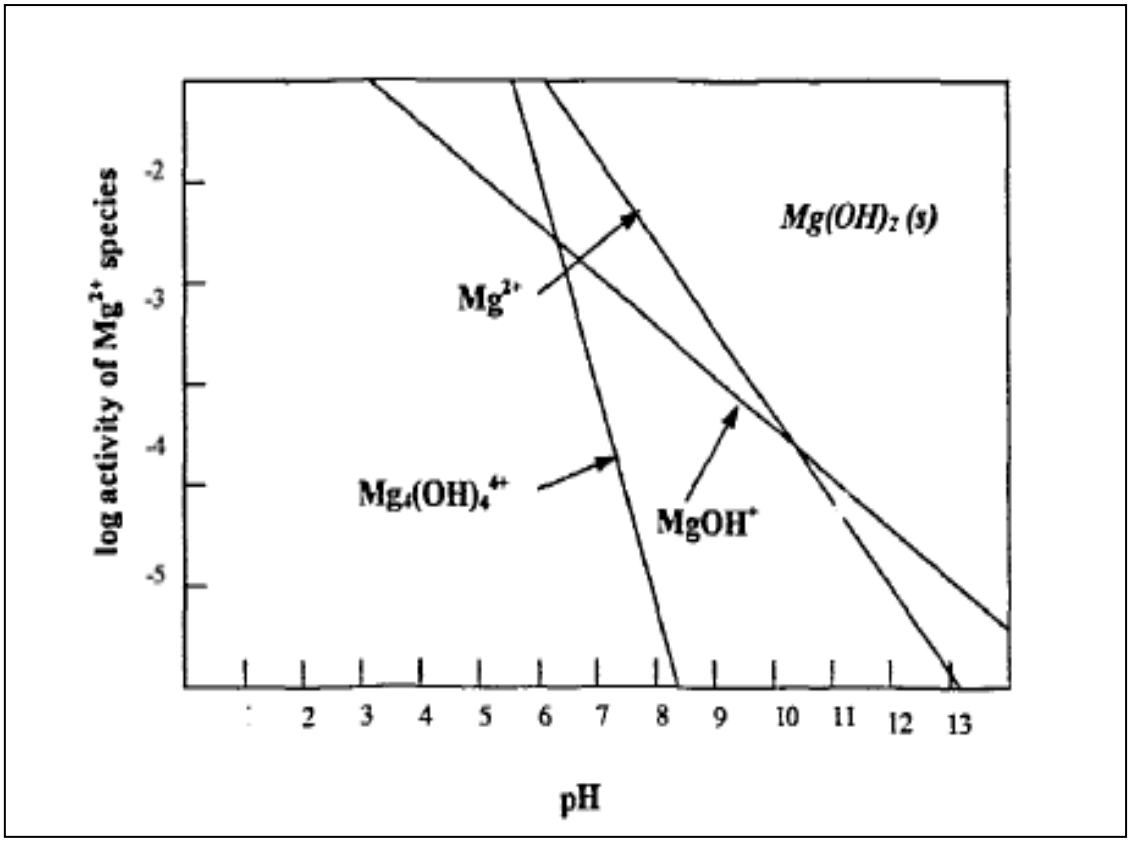

Figure 2.5: Solubility diagram for magnesium hydroxide (Damons, 2001)

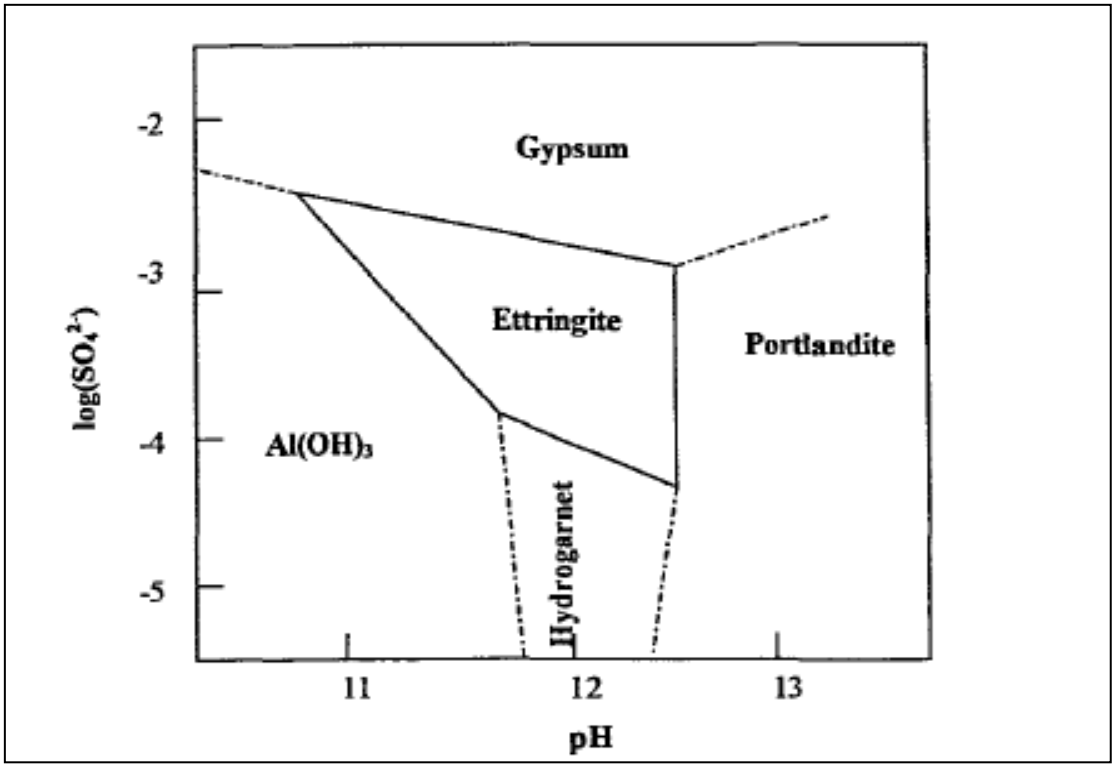

Figure 2.6: Ettringite stability in alkaline environments (Damons, 2001)

2.4.2. Calcium and sodium

2.4.2.1. The effect of calcium and sodium on gold adsorption onto activated carbon

Davidson (1974) states that the degree to which various gold cyanide complexes are adsorbed onto activated carbon is dependent on the complementary cations present. He further states that the adsorption strength decreases in the order $Ca^{2+} > Mg^{2+} > H^{+} > Li^{+} > Na^{+} > K^{+}$ (Davidson, 1974). Davidson (1974) concluded that both calcium and sodium considerably aid in gold adsorption at various pH values. Calcium addition produced the greatest effects on the activated carbon capacity at concentrations well below that of sodium. The presence of excessive calcium ions had an appreciable effect on both reaction rates and on adsorption isotherms when compared to the effects produced by the addition of sodium ions. Figures 2.7 and 2.8 illustrate the reaction rate curves obtained for gold adsorption at various pH levels in the presence of excess sodium and calcium ions respectively (Davidson, 1974).

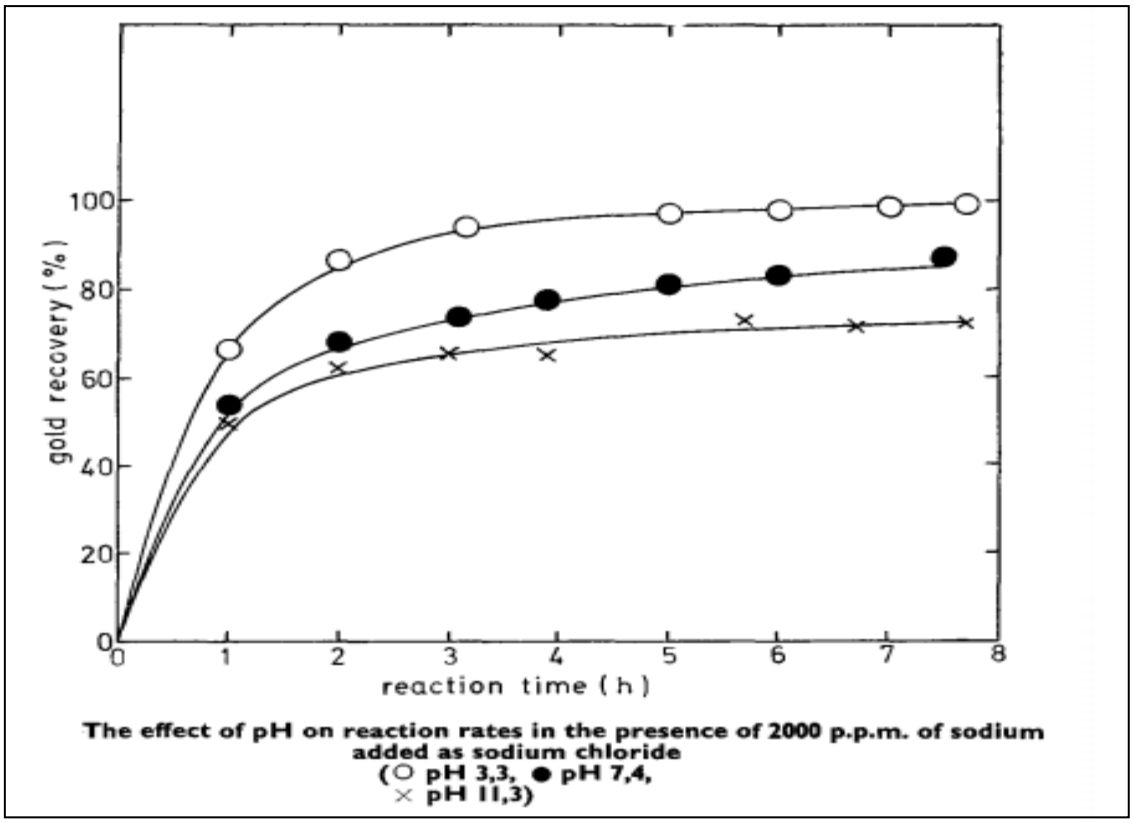

Figure 2.7: Reaction rate curve for gold adsorption in the presence of excess sodium (Davidson, 1974)

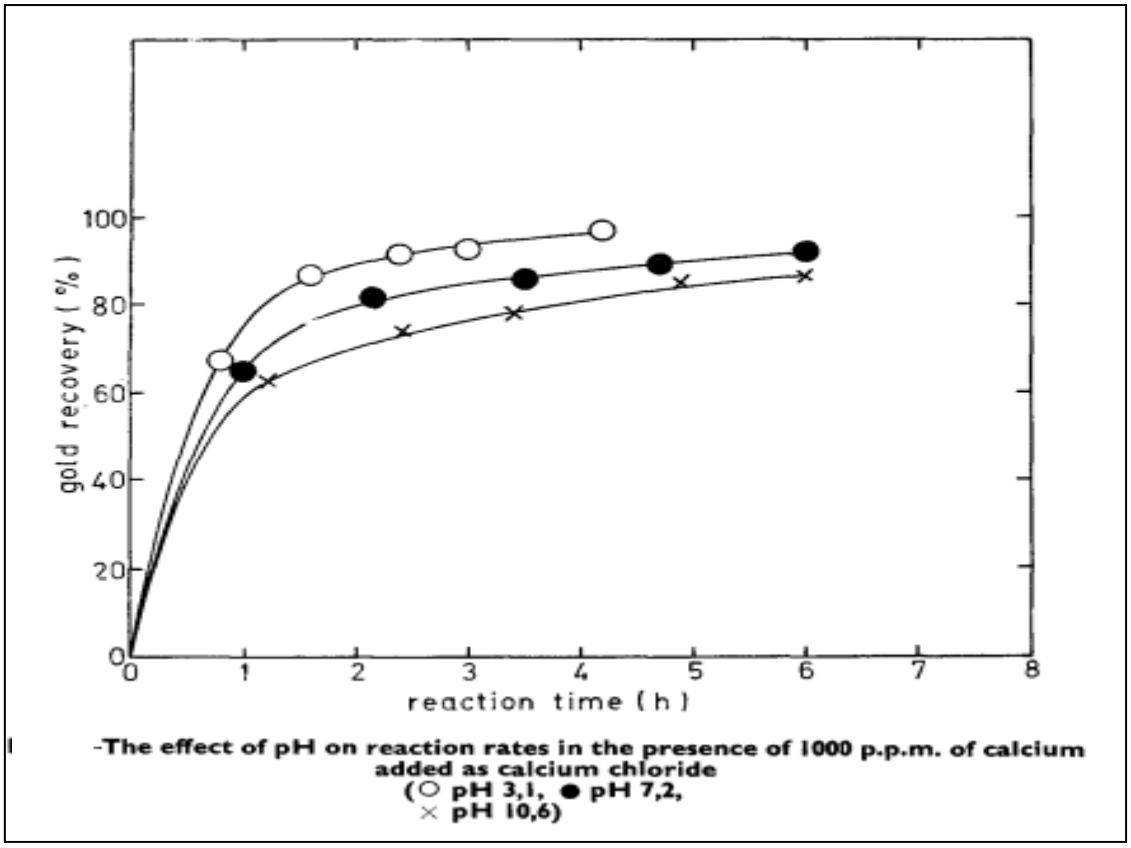

Figure 2.8: Reaction rate curve for gold adsorption in the presence of excess calcium (Davidson, 1974)

Furthermore, in the presence of an excess of the halide salts, the addition of calcium resulted in faster reaction rates within the pH range between 3 to 11 when compared to sodium (Davidson, 1974). Figures 2.7 and 2.8 illustrate the reaction rates for gold adsorption at various pH levels (3 to 11) obtained in the presence of excess sodium and calcium ions respectively (Davidson, 1974). From Figures 2.7 and 2.8, it is evident that the addition of calcium ions had a noticeable effect on reaction rate when compared with the effects produced by the addition of sodium ions. The rate of reaction is determined by the gradient of the slopes, with a steeper gradient indicating a greater reaction rate. It is evident from Figure 2.8 that the gradient of the curves are much steeper than those presented in Figure 2.7, illustrating that the presence of calcium ions has a greater influence on the kinetics of gold adsorption than sodium ions at various leach pH values. Nevertheless, the presence of sodium ions did indicate an improved reaction rate. Davidson (1974) attributed the improved leaching kinetics to the formation of a more strongly adsorbed calcium aurocyanide complex.

Cyanidation of gold ores progresses according to the Elsner equation is as follows:

$$Au^0 + 2CN^- + \frac{1}{2}H_2O + \frac{1}{4}O_2 \rightarrow [Au(CN)_2]^- + OH^- \quad (2.5)$$

Davidson & Solet (2007) presented arguments in favour of the following gold dissolution reaction over the Elsner equation during typical plant conditions:

$$2Au^0 + Ca^{2+} + 4CN^- + H_2O + \frac{1}{2}O_2 \rightarrow Ca[Au(CN)_2]_2 + 2OH^- \quad (2.6)$$

From Equation 2.6, it is observed that the end product is a calcium aurocyanide complex. Davidson & Solet (2007) proved that the calcium aurocyanide complex is rapidly and strongly adsorbed onto the activated carbon at near ambient temperatures

Slatter et al. (2009) demonstrated that sodium ions increased silver loading with negligible effect on gold loading. Slatter et al. (2009) further demonstrated that water sources with high sodium concentrations caused little to no mineral dissolution, indicating little surface alteration during gold flotation and leaching.

2.4.2.2. Calcium as a foulant on activated carbon

Calcium, as an inorganic contaminant, has detrimental effects on activated carbon activity as high concentrations lead to low gold recoveries by inhibiting the elution process. Removal of the calcium deposit is therefore an essential part of the process and is achieved by the acid wash circuit. Hydrochloric acid is used to remove calcium carbonate scale build up on the carbon, thereby opening up the activated carbon micro-pores and thus maximising the carbon's surface area. This results in improved overall gold elution efficiency in downstream processes (van Denventer & van de Merwe, 1993).

2.4.3. Summary of effects that contaminants have on gold flotation and leaching

The literature has been reviewed with respect to the effect that various metal cations have on gold recovery as well as the effects of inorganic ions such as calcium and sodium have on gold recovery. While there is a vast array of the literature that focuses predominantly on the effect and interactions of metal-cyanide complexes with the aurocyanide complex within gold ores, little is known about the effects of water quality on cyanide gold leaching. Due the growing scarcity of fresh water, the practice of recycling process water has become an economical viable option. Knowledge is lacking on the effect of water quality of the process water streams on typical sulfide ore gold processing plants. Furthermore, knowledge on the effects of excessive sulfates on gold flotation and gold recovery is lacking. Therefore, this thesis aimed at providing the water quality of process water at the ERGO Brakpan plant and these data are presented in Appendix A, Table A.1. The effects of contaminants within the process water on gold flotation and cyanide leaching are presented in Chapter 4.

2.5. WATER TREATMENT TECHNIQUES

This section provides an overview of acid mine drainage treatment techniques. Focus has been placed on treatment techniques that have shown success in the removal of heavy meal ions, sulfate ions as well as calcium and magnesium ions.

2.5.1. Chemical precipitation - crystalactor

Hard water is defined as water that causes scaling due to excess amounts of calcium and magnesium ions. Hard water may be purified by chemical softening, which entails the addition of chemicals to hard water to remove calcium and magnesium ions. The contaminants are removed from water through precipitation in the form of calcium carbonate and magnesium hydroxide. The softening reactions are illustrated below:

$$Ca(HCO_3)_2 + Ca(OH)_2 = 2CaCO_3 + 2H_2O \quad (2.7)$$

$$Mg(HCO_3)_2 + 2Ca(OH)_2 = 2CaCO_3 + Mg(OH)_2 + 2H_2O \quad (2.8)$$

As seen from Equation 2.7, the calcium will precipitate to form calcium carbonate. However, the calcium carbonate will form in a crystalline structure and with the absence of seed crystals, precipitation will be slow with scaling likely to take place. Conversely, should precipitation crystals be present, precipitation takes place almost immediately on the available crystal surface (Giesen, et al., 2009). A crystalactor may be utilised to precipitate calcium and magnesium from hard water. A crystalactor is a fluidised bed of grains on which the crystallization of $CaCO_3$ occurs. Water and lime are introduced from the base of the crystalactor and softening reactions proceed in the presence of lime and a suspended bed of fine particles, which acts as a catalyst.

The upward movement of the water causes the particles within the crystalactor to fluidise allowing the precipitated particles to attach to the surface of the fluidised particles (Giesen, et al., 2009). While the crystalactor is in operation, the particles grow, with several of the larger particles removed at the base of the reactor with the pellet discharge. Smaller particles are added at the top of the crystalactor to maintain/increase the surface area. Schutte (2006) states that it is vital that the pH is not increased as this will prevent the precipitation of magnesium hydroxide. Bester (2012) states that seeded fluidised bed reactors are utilised for several precipitation reactions including that of copper sulfide. Bester (2012) further states that a vital mechanism for softening of hard water within a fluidised bed is the

aggregation of the initially formed particles. The optimized grain size for this type of softening was found to be 0.3 to 1 mm (Bester, 2012).

Figure 2.9 illustrates the process operation of a crystalactor:

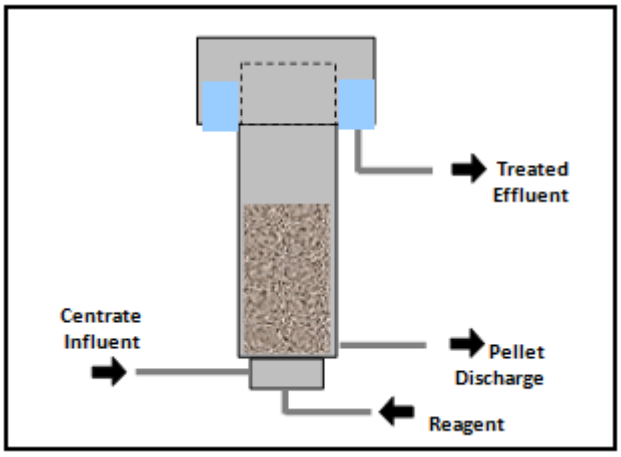

Figure 2.9: Operating principle of a crystalactor (Bester, 2012)

2.5.2. Ion exchange resins

An ion exchange resin consists of cross-linked polymers to which charged functional groups are attached through covalent bonding. The cations and anions in water are selectively removed when water is percolated through beds containing cation and anion exchange resins. The common types of resins are (Botha, et al., 2009):

- Strong Acid Cation (SAC) – absorbs all cations
- Weak Acid Cation (WAC) - reduces alkalinity (HCO_3^- and OH^-) by exchanging their hydrogen ions for the cations associated with the bicarbonate ion
- Strong Base Anion (SBA) – removes all anions including silica
- Weak Base Anion (WBA) - efficiently removes strong acids such as sulphuric and hydrochloric acid

The use of ion exchange resins has several applications in industry from the primary recovery of metals to the removal of low levels of impurities from highly concentrated solutions. Ion exchange resins are utilised to recover metals from both clarified and dirty solutions as well as pulps. In addition, ion exchange resins are utilised to reduce metal concentrations and contaminants in wastewater streams and to supply suitable quality process water. The most beneficial ion-exchange reactions are those that are reversible. The resin exchange capacity is measured as the number of fixed charged sites per volume of resin (Bochenek, et al., 2011).

The exhausted resin bed may undergo regeneration by use of excess of the required ion with no structural changes occurring during regeneration. Regeneration requires approximately 1 to 5 bed volumes of regenerant (depending on the amount of resin being regenerated). This is followed by a minimum of 2 bed volumes of rinse water with approximately 2 % of wastewater produced.

Typical resin reactions for loading are given below in Equation 2.9-2.12 (Bester, 2012):

$$\begin{bmatrix} Ca \\ Mg \\ 2Na \end{bmatrix} \cdot \begin{bmatrix} SO_4 \\ 2Cl \\ 2NO_3 \end{bmatrix} + 2R \cdot H \leftrightarrow 2R \cdot \begin{bmatrix} Ca \\ Mg \\ 2Na \end{bmatrix} + \begin{bmatrix} H_2SO_4 \\ 2HCl \\ 2HNO_3 \end{bmatrix} \quad (2.9)$$

$$\begin{bmatrix} H_2SO_4 \\ 2HCl \\ 2HNO_3 \end{bmatrix} + 2R \cdot OH \leftrightarrow 2R \cdot \begin{bmatrix} SO_4 \\ 2Cl \\ 2NO_3 \end{bmatrix} + 2H_2O \quad (2.10)$$

The regeneration reactions for resin are given below:

$$2R \cdot \begin{bmatrix} Ca \\ Mg \\ 2Na \end{bmatrix} + 2HNO_3 \leftrightarrow \begin{bmatrix} Ca \\ Mg \\ 2Na \end{bmatrix} \cdot NO_3 + 2R \cdot H \quad (2.11)$$

$$2R \cdot \begin{bmatrix} SO_4 \\ 2Cl \\ 2NO_3 \end{bmatrix} + NH_4 \cdot OH \leftrightarrow 2NH_4 \cdot \begin{bmatrix} SO_4 \\ 2Cl \\ 2NO_3 \end{bmatrix} + 2R \cdot OH \quad (2.12)$$

Resins are predominantly utilised for the softening of hard water through the removal of calcium and magnesium as well as heavy metals. The selectivity of the chosen resin determines the run time before breakthrough of the contaminant ions. Wong *et al. (*2009) state that resins that possess a chelating functional group (such as iminodiacetate), have a high affinity for hard ions and other troublesome metals such as uranium and aluminium. These resins are used to remove trace-metals in water (Wong, et al., 1996). The production of cation exchange resins is through the introduction of an acid group (such as SO_3H) into the polymer (Schutte, 2006).

Counter-current ion exchangers are the predominant configuration utilised in industry, where production can take place either through up or down-flow. The regeneration of these resins will occur in the direction opposite to production. Counter-current operation is desired as the regeneration circuit prevents ionic contamination (Degremont and Ondeo Industrial Solutions, 1995a). In addition, the circuit may be operated as a continuous process by utilisation of a multicolumn set-up operated in shifts. In addition, Bochenek *et al. (*2011) state that continuous operation may significantly outperform periodic operation.

Figure 2.10 illustrates the operating principles of a counter-current ion exchange process.

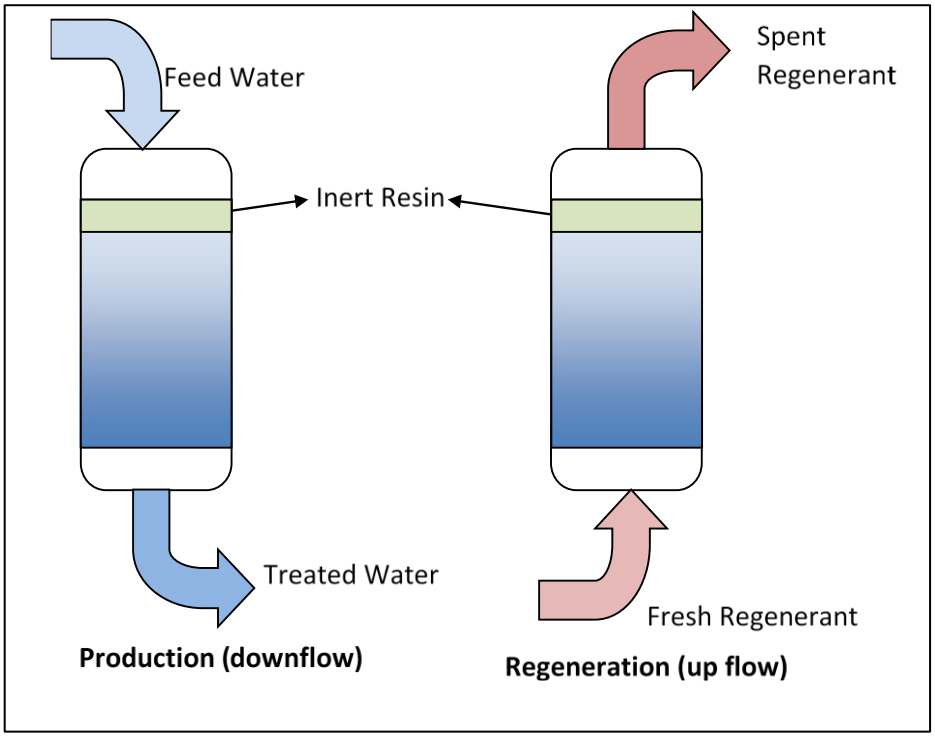

Figure 2.10: Counter-current ion exchange configuration (Bester, 2012)

In conclusion, for resin regeneration, the regenerant (brine) contains high concentrations of ions as well as excess regenerant chemicals. Care should be taken in the disposal of the waste regenerant to ensure that no pollution of surface or groundwater sources occurs. Typically, the brine is pumped to evaporation ponds and the non-hazardous gas emissions are discharged into the atmosphere (Bester, 2012).

2.5.3. Membranes – reverse osmosis

Several membrane applications and processes are utilised in water reclamation operations. The predominant practice involves pre-treatment of polluted water with coarse membranes (such as microfiltration). This is usually followed by additional processing such as precipitation and reverse osmosis (which utilises finer membrane technologies). Reverse osmosis utilises a semi-permeable membrane to proficiently remove a wide range of pollutants including ionic and non-ionic molecules. As illustrated by Figure 2.11, pure solvent (clear water) is allowed to pass to the other side, while retaining the solute on the pressurised side of the membrane.

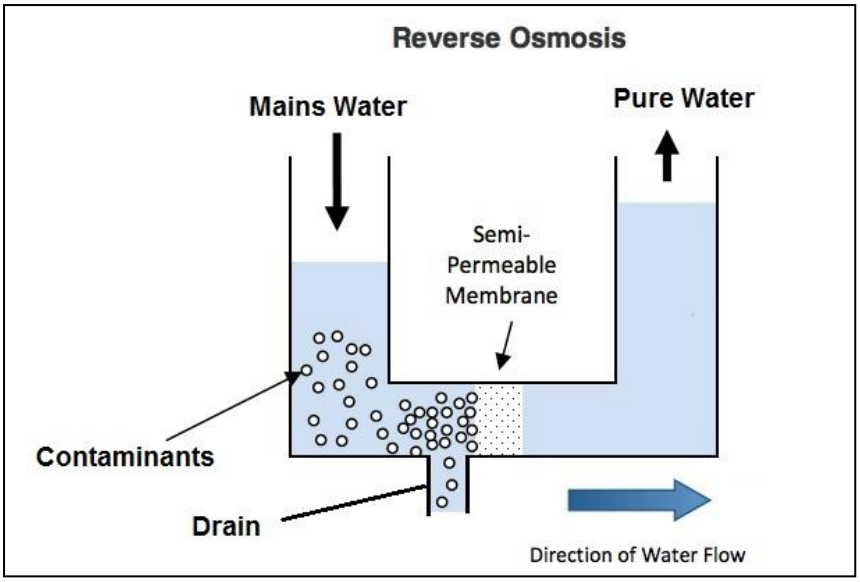

Figure 2.11: Operating principle of reverse osmosis (Puretec, n.d.)

Reverse osmosis membranes are not an absolute barrier, and some ions or molecules will pass through the membrane. Reverse osmosis is predominantly utilised for removing salts from brackish water and has been proven to effectively separate and concentrate effluents such as ammonium and nitrate effluents. Bester (2012) states that a major advantage is that permeate from the reverse osmosis process may be recycled and reused. In addition, the concentrated ammonium sulfate solution may be utilized as a liquid fertilizer (Bester, 2012).

2.5.4. Coagulation and flocculation (settling tanks)

Coagulation increases the tendency of small particles in an aqueous suspension to attach to one another, forming flocs that attach to the surfaces such as the grains in a filter bed (Bester, 2012). Coagulants are added to the water prior to the water flowing into a settling tank. Degremont and Ondeo Industrial Solutions (1995b) state that the pH of the water is an important factor and should be adjusted to ensure optimum performance by the coagulant. However, polymeric coagulants are not pH sensitive. Coagulation occurs in a rapid mixing tank over one to two minutes maximum. During coagulation, the coagulant is added to destabilize the surface charge of particles in order to promote the formation of micro flocs during the particle collisions (Minnesota Rural Water Association, 2009).

Conversely, flocculation allows for the formation of floc in quiescence. During flocculation, the solution is mixed gently, providing time and energy so that the clumps will gather together and form larger clumps. A polymer is added to the solution to strengthen the floc that formed, thus, resulting in reduced floc breakage and improved dewatering (Minnesota Rural Water Association, 2009).

The common practice in industry is the utilisation of circular settling tanks that make use of a scraper system with more than one blade. These blades rotate around the centre of the tanks and push the sludge along the floor of the tank. Water enters the settling tanks at the top in the middle of the tanks. The flocculated solid particles settle and are removed as underflow from the bottom of the tank, while the clarified water is removed as the tank overflow (Degremont and Ondeo Industrial Solutions, 1995b). A schematic of a settling tank is shown in Figure 2.12.

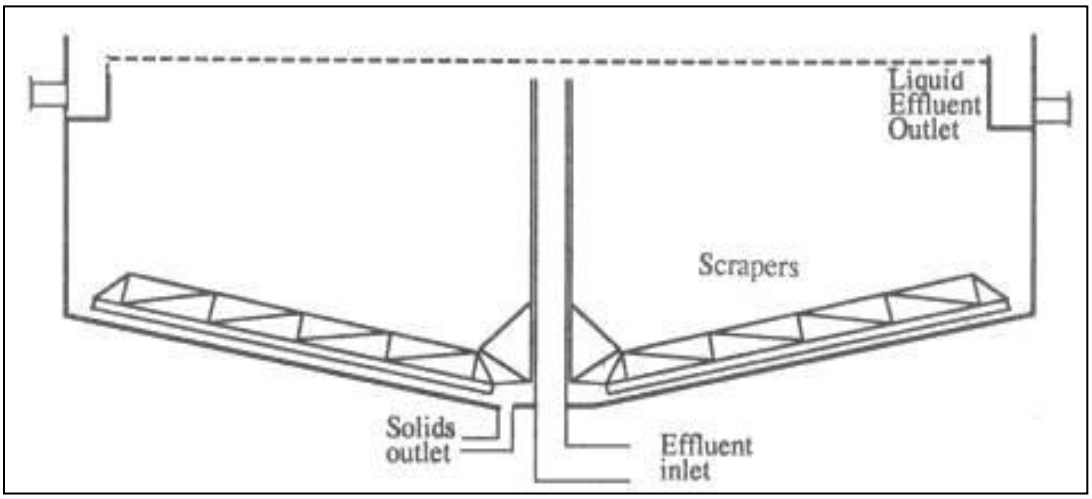

Figure 2.12: Schematic of settling tank (Bester, 2012)

2.5.5. The SAVMIN process

The SAVMIN process is a chemical precipitation process that reduces the concentration of Ca^{2+} and SO_4^{2-} by precipitation of gypsum and ettringite. The process was patented by Mintek and Savannah mining. Lime is added to acid mine drainage (AMD) to precipitate metal hydroxides with the subsequent formation of ettringite to remove sulfates and calcium by addition of aluminium hydroxide (Damons, 2001). The process consists of 5 vital stages as described below and illustrated by Figure 2.13:

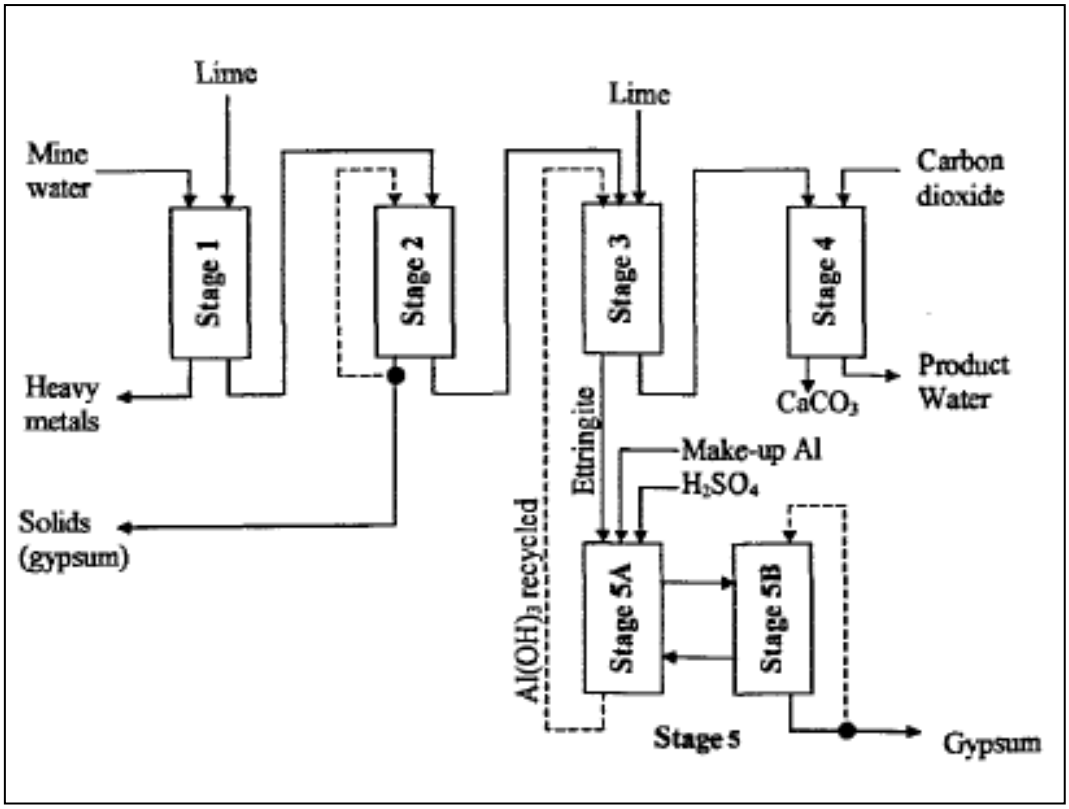

Figure 2.13: Block flow diagram of the SAVMIN process (Damons, 2001)

In stage 1, acid mine drainage at an approximate pH value of 6 is introduced into a mixing vessel and contacted with lime (CaO) to raise the pH to above 11.5. At these conditions, the heavy metals are precipitated as metal hydroxides (refer to reaction 2.13 below). Although most metal hydroxides will precipitate at relatively lower pH values (pH<8), a high pH is needed for the precipitation of magnesium. The metal hydroxide precipitants are filtered and removed from the system (Damons, 2001):

$$Me^{2+} + 2OH^- \rightarrow Me(OH)_2 \tag{2.13}$$

Where "Me" refers to divalent heavy metals such as iron, nickel, magnesium, etc.

The addition of lime in stage 1, may cause the solution phase to become supersaturated with gypsum, depending on the concentration of sulfates in the feed water. The solution from stage 1 is contacted with gypsum crystals in stage 2 in order to provide active surfaces of gypsum, which acts as a catalyst for the precipitation of the 'supersaturated' gypsum (Damons, 2001). This precipitated gypsum is thickened, filtered and leaves the process as waste or as a by-product. The formation of gypsum is described by the following stoichiometric reaction (Damons, 2001):

$$Ca^{2+} + SO_4^{2-} + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O \tag{2.14}$$

Stage 3 is the heart of the SAVMIN process. Aluminium hydroxide is added to the saturated gypsum solution from stage 2. This results in the formation of an insoluble salt known as ettringite, thereby removing both sulfate and calcium from solution (Damons, 2001). The stoichiometry for ettringite precipitation is given by the following reaction (Damons, 2001):

$$6Ca^{2+} + 2Al(OH)_3 + 3SO_4^{2-} + 37H_2O \rightarrow 3CaO \cdot 3CaSO_4 \cdot Al_2O_3 \cdot 31H_2O + 6H_3O^+ \quad (2.15)$$

For optimum ettringite formation, the solution pH value should range between 11.4 -12.4. To achieve and maintain the pH between these limits, lime is added to the solution to maximise ettringite precipitation.

The solution from stage 3, which contains little to no heavy metals, calcium and sulfates, is treated with carbon dioxide to reduce the pH to neutral (pH range between 7.0 – 8.0) in stage 4. Pure calcium carbonate is precipitated and is separated from the resultant product water by filtration (Damons, 2001). The stoichiometry for the reduction in pH is given by the following reaction (Damons, 2001):

$$H_2O + CO_2 \rightarrow CO_3^{2-} + 2H^+ \quad (2.16)$$

A carbonate ion (CO_3^{2-}) is formed as well as 2 hydronium ions. The presences of the H^+ ions will lower the pH (Damons, 2001). The CO_3^{2-} ions simultaneously react with the Ca^{2+} ions in solution to form calcium carbonate as shown in the following reaction (Damons, 2001):

$$Ca^{2+} + CO_3^{2-} \rightarrow CaCO_3(s) \quad (2.17)$$

Apart from altering the pH, precipitation of $CaCO_3$ also aids in the removal of calcium from solution.

The ettringite slurry is simultaneously transported to stage 5 and undergoes decomposition in order to regenerate the amorphous aluminium hydroxide, which is recycled back into stage 3. The decomposition of ettringite is achieved by contacting the slurry with sulphuric acid. This results in a decrease in the pH of the slurry and thereby renders it unstable. Decomposition takes place in gypsum-saturated water, at a liquid to solid ratio that allows the calcium and sulfate ions to remain in solution as supersaturated calcium sulfate (Damons, 2001). The decomposition reaction stoichiometry is the reverse of that of the ettringite formation. The resultant solution is thickened and filtered to separate the insoluble aluminium hydroxide (which is recycled to stage 3) from the supersaturated solution of calcium sulfate. The supersaturated calcium sulfate solution is contacted with gypsum crystals, as in stage 2, in order to crystallize the calcium sulfate, which is removed by thickening and filtration.

2.6. SUMMARY AND SCOPE OF WORK

The discussion in Sections 2.4 and 2.5 has reviewed the literature with regards to effects that several metal cations, sodium and calcium have on gold recovery, with successful AMD treatment techniques presented. A number of areas emphasised in the literature review influence the scope of this thesis. Specifically, these are:

- The recovery of gold from an ore is dependent on the mineralogy and amount of gold present within the ore. The creation of metal-cyanide complexes occurs when leaching reagents (such as sodium cyanide) are contacted with the ore. There exists a need for further understanding of the effect of metal-cyanide complexes on gold recovery since a significant concentration of metal ions are present within ERGO Brakpan's process water (refer to Appendix A, Table A.1).

- There is limited knowledge on the water quality of the process water streams on typical sulfide ore gold processing plants. Literature is also limited in terms of the effects of excessive sulfates on gold flotation and gold recovery. Therefore, this thesis aimed to provide an analysis of the process water at the ERGO Brakpan plant. The effects of contaminants within the process water on gold flotation and cyanide leaching are presented in Chapter 4.

- Literature is limited on the difference in gold recoveries between untreated process water and water that has been treated for reuse by processes discussed in Section 2.5. The SAVMIN process was of particular interest due to the high sulfate concentration within ERGO Brakpan's process water. The most economically viable treatment option is presented in Chapter 5.

3. CHAPTER 3: METHODOLOGY AND EQUIPMENT

This chapter describes the experimental design and procedures undertaken to remove contaminants from ERGO Brakpan's process water under laboratory conditions. Additionally, the experimental procedures for both the leach tests as well as flotation followed by leaching of the flotation tails tests are highlighted under laboratory conditions. The purpose of the experiments was to confirm the removal of contaminants by chemical precipitation under conditions specified in literature as well as to determine the associated effects on gold recovery.

3.1. EXPERIMENTAL VARIABLES

The independent variables for the experiments remained constant during each run. These variables are listed below:

- Experimental pressure (101 kPa)
- Volume of process water treated (10 L per test)
- Dosage of reagents required for flotation
- Dosage of reagents required for gold leaching
- Leaching residence time (7 hours)
- Slurry leaching pH (pH = 10.5)
- Bottle Roll speed (120 rpm)

The dependent variables depend on the value of the independent variables given above and are as follows:

- Final concentrations of contaminant after chemical precipitation
- Temperature at which water treatment is undertaken
- Mass of contaminant precipitant formed
- Overall gold recovery

3.2. ORE USED IN TESTS

A sulfide slime ore sourced from ERGO Brakpan's 4L50 Reclamation site was utilised in this study. The ore was sourced from an open pit reclamation site, which had previously undergone crushing in the crushing circuit. The ore was sampled over a period of one month in order to obtain a 60 kg sample of ore representative of that processed in the plant. Although ERGO Brakpan treats 64 000 tons per day, the sample utilised for this test work represents approximately 20 000 tons of the total daily feed into ERGO Brakpan. This is attributed to difficulties in sampling from all the reclamation sites due to restricted access as well as safety limitations.

3.3. REAGENTS REQUIRED

In the addition to potable water and ERGO Brakpan's process water being used for the experiments, a number of reagents required for leaching, flotation and water treatment were also used. These include:

- Calcium Oxide (Lime)
- Sodium Carbonate (Soda Ash)
- Sodium Cyanide
- Flotation Reagents (Sodium Normal Propyl Xanthate (SNPX), 1,1,3 Triethyloxybutane (Senfroth), Alkyl Dithiophosphate (Senkol))
- Nickel Sulfate
- Iron Sulfate
- Manganese Sulfate
- Magnesium Sulfate
- Zinc Sulfate

The chemicals utilised to perform the experiments were all of Laboratory Reagent (LR) grade.

3.3.1. Reagent dosage for chemical precipitation

Lime and soda ash softening was the proposed treatment technique undertaken on ERGO's process water. The lime and soda ash softening procedures are presented in Sections 3.8.1 and 3.8.2 respectively, with the results presented in Chapter 4, Section 4.3. Lime and soda ash dosages depends on carbonate and non-carbonate hardness in the water. Lime is used to remove carbonate hardness, and both lime and soda ash are used to remove non-carbonate hardness. If total hardness is less than or equal to total alkalinity, then hardness is attributed to carbonate hardness only with no non-carbonate hardness. If total hardness is greater than total alkalinity, the carbonate hardness is equal to the total alkalinity concentration, and the non-carbonate hardness is calculated from the difference between total hardness and total alkalinity concentrations (Minnesota Rural Water Association, 2009). If total hardness is equal to or less than total alkalinity, then the lime dosage is calculated as shown in equation 3.1 (Minnesota Rural Water Association, 2009):

$$Lime\ Dosage\ \left(\frac{mg}{l}\right) = \frac{A+B+C+D}{Lime\ Purity} \times \%EXCESS \tag{3.1}$$

Where: A = The Carbon Dioxide Concentration in source water (mg/l)

B = Bicarbonate Alkalinity in source water (mg/l)

C = Hydroxide Alkalinity in source water (mg/l)

D = Magnesium Concentration in source water (mg/l)

% Excess = Amount of lime fed in excess to ensure pH > 10.5

Lime Purity = A lime purity of 85% was utilised (as a decimal)

When treating water that contains non-carbonate hardness, soda ash is required. The amount of soda ash is estimated by the following formula:

$$Soda\ Ash\ \ (mg/l) = Non - Carbonate\ Hardness\ as\ CaCO_3 \times \frac{Molecular\ Weight\ of\ Na_2CO_3}{Molecular\ Weight\ of\ CaCO_3} \times \%\ EXCESS \quad (3.2)$$

Initial analyses using inductively coupled plasma optical emission spectroscopy (ICP-OES) and inductively coupled plasma mass spectrometry (ICP-MS) were undertaken on a composite sample of the ERGO Brakpan process water. The contaminants present in the highest concentrations were identified and were the focus of this thesis (Refer to Appendix A, Table A.1 for the summarised ICP-MS analysis).

The carbon dioxide, bicarbonate and hydroxide concentrations were obtained by titration using 0.1M sodium hydroxide (NaOH) as the titrant. The calcium carbonate ($CaCO_3$) concentration was obtained by back titration with the use of excess 0.1M hydrochloric and 0.1M sodium hydroxide (NaOH) as the titrant. The magnesium concentration was obtained from the ICP analysis and the required reagent dosages were calculated using Equations 3.1 and 3.2 (Minnesota Rural Water Association, 2009).

Table 3.1: Reagent dosage calculation

Reagent dosage							
	A	**B**	**C**	**D**	**$CaCO_3$**	**Total**	**Excess**	**Dosage**
	mg/l	mg/l		mg/l	mg/l	mg/l	%	**mg/l**
Unslaked Lime	350	67		129	-	546	120	**756**
Soda Ash	-	-		-	127	127	120	**180**

At ambient conditions, the dosage proved valid, producing the desired pH for effective chemical precipitation (>10.5). However, due to the inverse relationship between temperature and pH, it became evident that as the temperature of the process water increased, a greater amount of lime was required to achieve the desired pH.

3.4. EXPERIMENTAL EQUIPMENT

The experiments were performed in a batch configuration, using two 5000 ml glass beakers, each with a magnetic stirrer. The pH was measured with the use of a HANNA HI8314 membrane pH meter, which was calibrated before each test using buffer solutions of a pH of 7 and 10 respectively. Filtration was

undertaken using WHATMAN 320mm filter paper and a 10 000 ml pressure filter at 5 bar air pressure. The solution samples were analysed for heavy metals, sodium, potassium, magnesium and calcium by means of Atomic Absorption (AA) and the solids were analysed by means of X-Ray Diffraction (XRD) to determine crystal structure of precipitate. Sulfate concentrations were analysed by means of ICP analysis.

3.5. CARBON PREPARATION

Activated carbon, sourced from Protea Chemicals Pty (Ltd), was used for all experiments where activated carbon was needed. Care was taken in the preparation of activated carbon, as the carbon contained large amounts of fine carbon particles that are trapped within the pores of the carbon and attached to the carbon surface. It was necessary to remove these prior to its use in experimentation, as these particles are capable of adsorbing gold, and are extremely difficult to recover from the slurry. The carbon preparation required washing the carbon in distilled water several times to remove any loose fines. The carbon was then placed into several 5 L bottles and placed onto a roller machine for 7 hours at a stirring speed of 120 rpm. Following the attrition treatment, carbon fines were removed from the coarse carbon by washing the carbon through a +1.00 mm screen using distilled water. During experimentation, any carbon attrition due to the breakdown of carbon particles in the slurry was likely to be minimal. Figure 3.1 illustrates the washed activated carbon.

Figure 3.1: Washed activated carbon

3.6. ANALYTICAL TECHNIQUES

Several analytical techniques were used for the analysis of results. The techniques used are well established in both industry and academia. Details of the techniques used are given below. The results of this thesis depend largely on the accuracy of the solution analysis.

3.6.1. Atomic absorption calibration

Atomic absorption spectroscopy (AAS) was used for the quantitative determination of chemical elements using the absorption of optical radiation (light) by free atoms in the gaseous state. A blank sample and standard solutions of varying concentration were prepared for calcium (Ca), potassium (K), sodium (Na), magnesium (Mg), iron (Fe), nickel (Ni) and manganese (Mn). A linear calibration was performed using a range of at least 4 standards for each element and the composition of the standard solutions are presented in Table 3.2 below. The atomic adsorption calibration curves for the respective elements are presented in Appendix B, Figures B.1 -B.7

Table 3.2: Composition of the standard solutions for calibration of atomic absorption spectroscopy

Stock Standard Preparation	Concentration (mg/l)	Aliquot (ml)	Final Volume (ml)	Acid (ml)
Sodium & Potassium	100	10	100	1
Calcium and Magnesium	100	10	100	1
Iron, Nickel and Manganese	100	10	100	1

The composition of the standards presented in Table 3.2 corresponds to the make-up of 100 mg/l standard solutions diluted from 1000 mg/l stock solutions. The 100 mg/l standard solutions were further diluted to produce 50 mg/l, 20 mg/l, 10 mg/l and 5 mg/l standard solutions. The required dilutions from the 100 mg/l standards were calculated utilising the following equation:

$$C_1V_1 = C_2V_2 \tag{3.3}$$

Where:

V_1 = Volume of stock solution needed to make the new solution

C_1 = Concentration of stock solution

V_2 = Final volume of new solution

C_2 = Final concentration of new solution

The solution samples obtained for each leach test were diluted to achieve concentrations that would fit within the calibration range. Preparation of samples required the addition of a diluent or acid before the addition of de-ionised water to a predetermined volume. The dilution preparations are presented in Table 3.3 below:

Table 3.3: Dilution of solutions for analysis by atomic absorption spectroscopy

Element	Expected Concentration (mg/l)	Dilution	Aliquot (ml)	Diluent (ml)	Acid (ml)	Final Volume (ml)
Calcium (Ca)	≤ 100	5	10	25		50
Manganese (Mn)	≤ 20	1.25	25		0.5	50
Magnesium (Mg)	≤ 80	5	10	25		50
Sodium (Na)	≤ 200	20	5	5		50
Potassium (K)	≤ 10	1.25	40	5		50
Nickel (Ni)	≤ 10	1.25	40		0.5	50
Sulfate	≤ 300	10	2.5		0.5	50
Iron (Fe)	≤ 20	1.25	25		0.5	50

3.6.2. X-ray diffraction

XRD analysis was used to characterise the crystalline materials formed during the chemical precipitation. XRD analysis was performed using a Philips PW 1800 X-Ray Diffractometer. The XRD analysis was performed as a qualitative analysis to determine the mineral phases that had formed by making use of the distinct x-ray diffraction patterns associated with each mineral. The parameters used for the XRD analysis are described in Table 3.4:

Table 3.4: Parameters used for the XRD analysis

Parameter	Value
Type of radiation used	Cobalt-K alpha (CoKa)
Operating current	10 mA
Operating voltage	30 kV
2θ range	10°- 90°

3.6.3. Fire assay

Fire assay was used as the analytical technique for determining the gold content in the solid samples. All solid samples were submitted to Mechanical Analysis and Engineering Design (MAED) Metallurgical Laboratory for assay.

3.7. PREPARATION OF CONTAMINANT STANDARDS

One of the aims of this thesis, was to establish which of the ions, or combination of ions, present in the process water were responsible for the lower gold recovery compared to potable water. To achieve this, potable water was spiked with individual chemical species to determine the effect on gold recovery. In this way it was possible to isolate an individual ion or combination of ions that has the greatest effect on gold recovery. The contaminants with the highest concentration were identified from the initial ICP-MS analysis undertaken on a composite sample of the ERGO Brakpan process water and formed the focus of this work (Refer to Appendix A, Table A.1 for the summarised and ICP-MS analysis). Standard solutions of the identified contaminants were prepared by dissolution of a known amount of the analyte in cyanide solution. Stock solutions of 1 g/L (1000 mg/l) of Ni, Fe, Mn, Ca, Mg and sulfur (added as sulfate) were made up respectively. Iron, manganese and nickel were added as sulfates. The formation of $Ni(CN)_4^{2-}$, $Mn(CN)_2^{-}$, $Fe(CN)_6^{4-}$ in solution was achieved by the addition of the molar requirement of sodium cyanide.

Slatter et al. (2009) demonstrated that water sources with high sodium concentrations caused little to no mineral dissolution during gold flotation and leaching, therefore the potential effect that sodium has on gold recovery is assumed to be negligible. Sulphuric acid was used to make up the sulfate standard solution. Calcium and magnesium standards were created separately by making a 1 g/L (1000 mg/l) stock solution, respectively. Calcium oxide and magnesium sulfate ($MnSO_4$) were used respectively for the stock solutions. Sodium cyanide was added to these standards so that the free cyanide concentration was 190 mg/l (this corresponds to the ERGO Brakpan plant set point value). Equation 3.3 was utilised to determine the volume of stock solution required to produce 50 mg/l, 100 mg/l and 500 mg/l of each contaminant within 1 litre of slurry (mixed with potable water). The sources utilised to make up standard solutions with the respective dissociation stoichiometric reactions with sodium cyanide are summarised in Table 3.5 below:

Table 3.5: The sources utilised to make up standard solution with the respective dissociation stoichiometric reactions with sodium cyanide

Contaminant	Source	Stoichiometric Dissociation Reactions
Sulphur/ Sulfate	Sulphuric Acid H_2SO_4	$H_2SO_4 + 2H_2O \rightarrow 2H_3O^+ + SO_4^{2-}$
Iron	Iron (II) Sulfate	$Fe^{2+} + 6CN^- \rightarrow [Fe(CN)_6]^{4-}$
Manganese	Manganese (II) Sulfate	$Mn^{2+} + 2CN^- \rightarrow [Mn(CN)_2]^-$
Nickel	Nickel Sulfate	$Ni^{2+} + 4CN^- \rightarrow [Ni(CN)_4]^{2-}$
Calcium	Calcium Oxide	$Ca^{2+} + 2CN^- \rightarrow Ca(CN)_2$
Magnesium	Magnesium Sulfate	$Mg^{2+} + 2CN^- \rightarrow Mg(CN)_2$

3.8. EXPERIMENTAL PROCEDURE

3.8.1. Lime softening procedure

A diagrammatic representation of the lime softening procedure is presented in Figure 3.2. The results obtained from the lime softening procedure are presented in Chapter 4, Section 4.3. The lime softening procedure is presented below:

1. A 100 litre sample of composite process water from ERGO Brakpan was collected; from which a 10 litre sample was measured out and placed into two 5 litre glass beakers.
2. An additional 250 ml sample was collected and labelled as the head sample. This sample was analysed to determine the initial calcium and heavy metal concentrations.
3. The two 5 litre process water samples were treated with unslaked lime at the required dosage, as presented in Table 3.1, to obtain a pH of 10.5.
4. A one hour waiting period was observed to allow complete precipitation of contaminants.
5. The resulting solution was filtered through pressure filter and the precipitant was collected, dried overnight in an oven at 60°C and thereafter weighed.
6. A 250 ml portion of the filtered water was collected and labelled. This sample was analysed for calcium and heavy metal concentration.
7. The remaining 9.75 litre of filtered water was used to conduct the flotation and leach test- work to observe the effect on gold recovery.
8. Steps 2-7 above were repeated after heating the process water to the following temperatures: 20°C, 60°C and 90°C. The solubilities of silica, calcium and magnesium are reduced by increased temperature. Therefore, they are more effectively removed by an increase in temperature (Suez Water Technologies & Solutions, n.d.). Heating was achieved by placing the two 5 litre glass beakers in a water bath which was placed on a hot plate. The temperature of the process water was measured using a HANNA HI8314 membrane multi-meter. Heating was controlled manually by adjusting the dial on the hot plate.

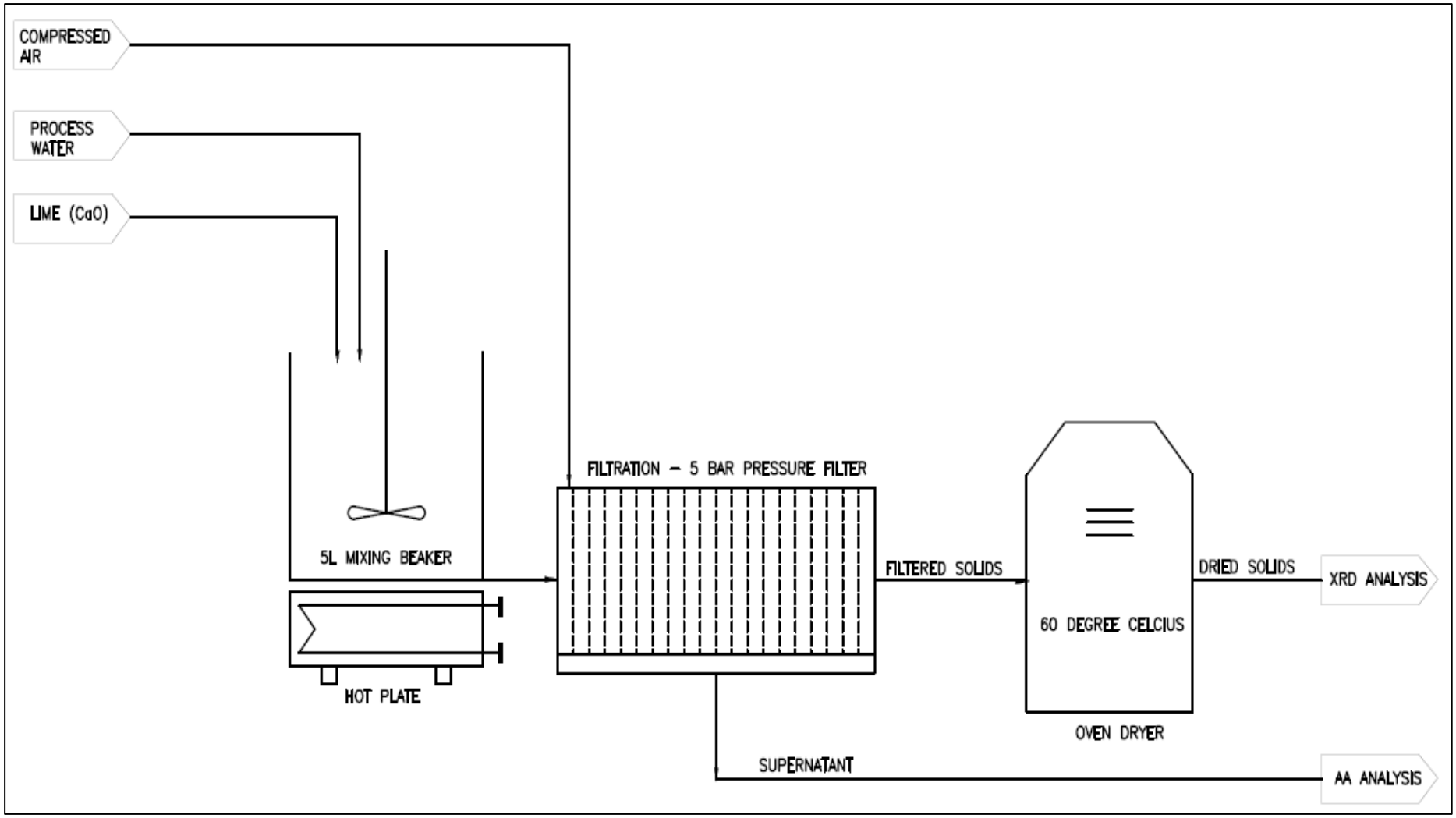

Figure 3.2: Diagrammatic representation of the lime softening experimental procedure

3.8.2. Lime and soda ash softening procedure

A diagrammatic representation of the lime-soda ash softening procedure is presented in Figure 3.3. The results obtained from the lime-soda ash softening procedure are presented in Chapter 4, Section 4.3. The lime-soda ash softening procedure is presented below:

1. A 100 litre sample of composite process water was collected from ERGO Brakpan; from which a 10 litre sample was measured out and placed into two 5 litre glass beakers.
2. An additional 250 ml sample was collected and labelled as the head sample. This sample was analysed to determine the initial calcium and heavy metal concentrations.
3. The two 5 litre process water samples were treated with unslaked lime at the required dosage. as presented in Table 3.1, to obtain a pH of 11.
4. The required dosage of soda ash, as presented in Table 3.1, was thereafter added to the process water.
5. A 1 hour waiting period was observed to allow complete precipitation of contaminants.
6. The resulting solution was filtered through a pressure filter and the precipitant was collected, dried overnight in an oven at 60°C and thereafter weighed.
7. A 250 ml portion of the filtered water was collected and labelled. This sample was analysed for calcium and heavy metal concentrations.
8. The remaining 9.75 litre of filtered water was used to conduct the flotation and leach test work to observe the effect on gold recovery.
9. Steps 2-7 above were repeated after heating the process water to the following temperatures: 20°C, 60°C and 90°C. To achieve heating, the two 5 litre glass beakers were placed in a water bath which was placed on a hot plate. The temperature of the process water was measured using a HANNA HI8314 membrane multi-meter. Heating was controlled manually by adjusting the dial on the hot plate.

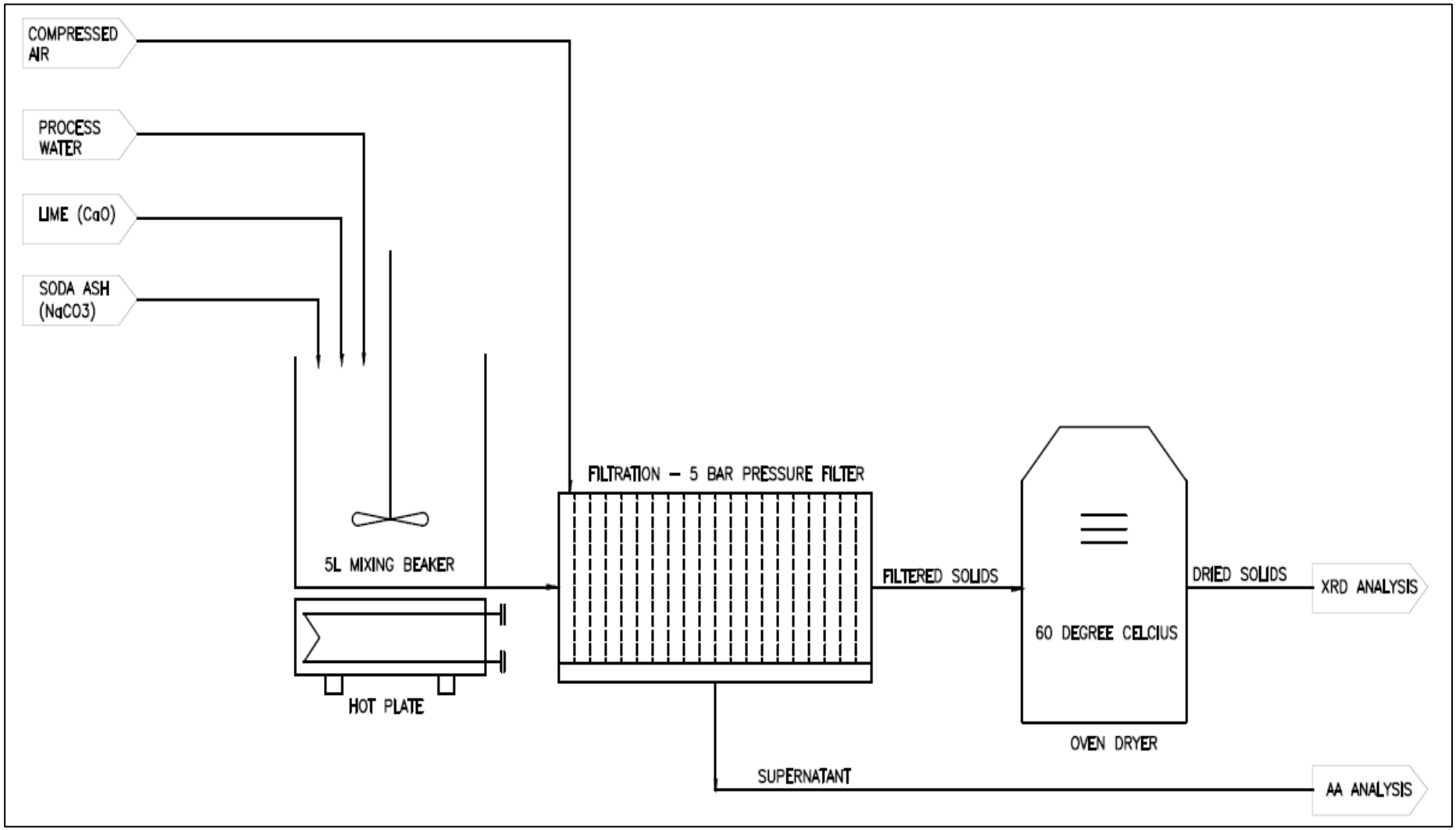

Figure 3.3: Diagrammatic representation of the lime-soda ash softening experimental procedure

3.8.3. Leach procedure

Figure 3.4 illustrates a block flow diagram of the leach test work. Leach results are presented in Sections 4.2 and 4.4. The leach test procedure is presented below:

1. A 60 kg composite dry slime sample was collected from one of the largest reclamation dumps that feeds into ERGO Brakpan.
2. The solid sample was mixed with the respective water sources (potable, untreated process water and treated process water) until a slurry density of 1.45 kg/l was obtained (this is the plant set point for effective leaching).
3. After achieving the desired density, a 1 litre head sample was collected, filtered, dried and assayed. (The head sample assay was measured as 0.20 g/t as presented in Sections 4.2 and 4.4)
4. Thereafter three, one litre samples at a density of 1.45 kg/l were placed into separate 5 litre bottles.
5. The lime, cyanide and carbon dosages shown in Table 3.6 were added and a bottle roll leach was undertaken for 7 hours, corresponding to the leach circuit residence time at ERGO Brakpan.
6. After leaching, the carbon was screened and washed prior to the slurry being filtered.
7. The filtrate was collected and submitted to Mechanical Analysis and Engineering Design (MAED) Metallurgical Laboratory for assay.
8. The filter cake was re-pulped and filtered twice, before being dried in an oven overnight at 110ºC.
9. The dried residue and carbon were submitted to MAED Metallurgical Laboratory for gold assay.

Table 3.6: Leach reagent dosage

Flotation Reagent	Plant Scale Dosage (g/t)	Lab Scale Equivalent Dosage (mg/l)
Sodium Cyanide	0.36	0.36
Lime	1.00	1.00
Activated Carbon	20	20

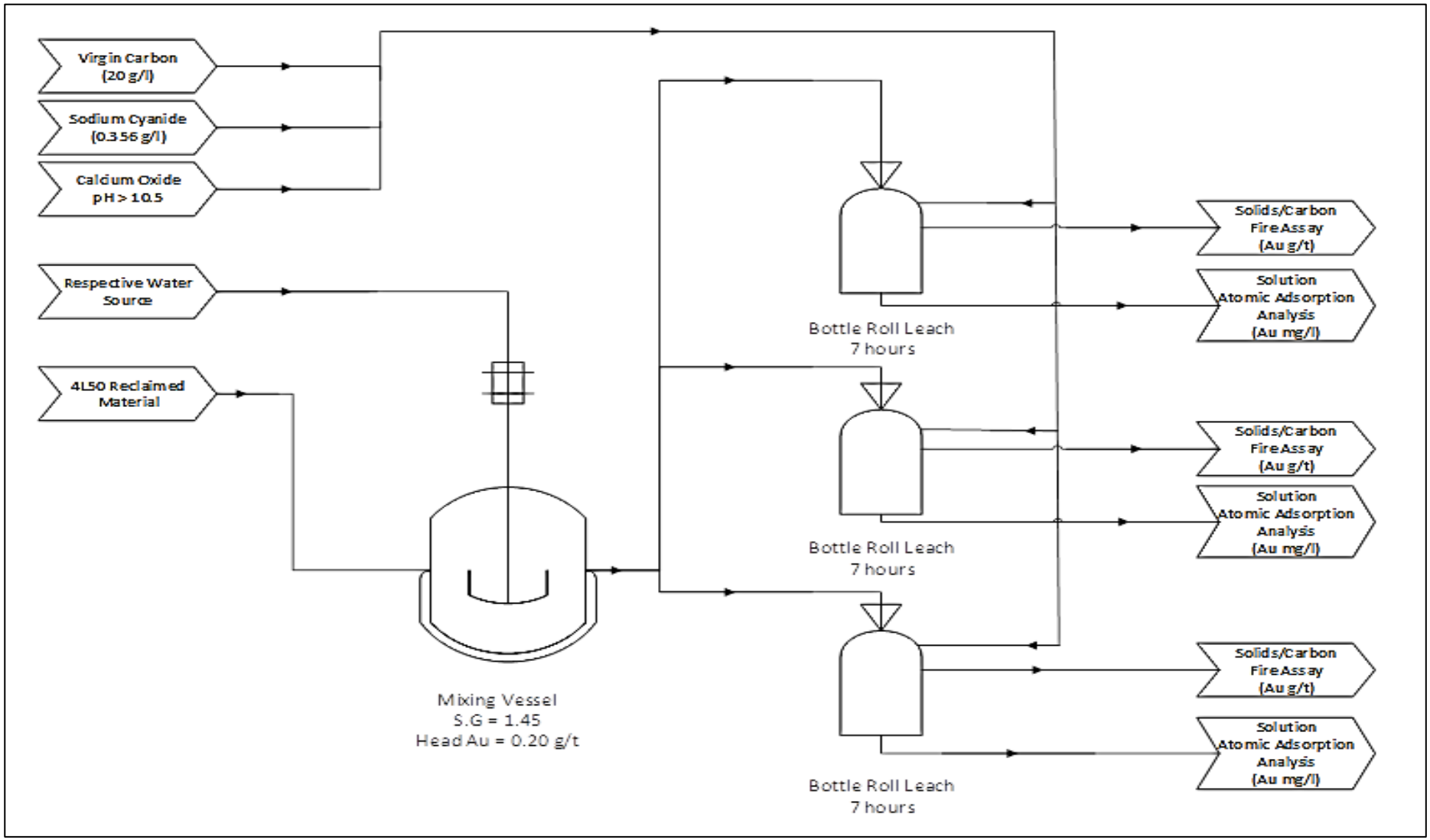

Figure 3.4: Block flow diagram illustrating the leach test work

3.8.4. Flotation and leach of flotation tails procedure

Figure 3.5 illustrates a block flow diagram summarising the flotation and tails leach test work. Flotation results are presented in Section 4.4. The flotation and tails leach test procedure is presented below:

1. After completion of preparation for the leach test work, additional water was added to the head sample to dilute the slurry to a density of 1.32 kg/l (the optimized plant condition for effective flotation).
2. A 5 litre portion of the slurry at a SG of 1.32 was transferred into a Denver Laboratory Scale Flotation Cell.
3. The initial pH of the slurry was measured. pH values outside of the optimal range between pH 6 - 8 required adjustment using lime (for pH values below 6), or sulphuric acid (for pH values above 8).
4. Thereafter the flotation reagents were added. Table 3.7 provides the flotation reagents that were added, as well as the required dosages.
5. The rougher flotation proceeded for 10 minutes after conditioning for 5 minutes.
6. The rougher concentrate was collected by scraping the froth at 10 second intervals.
7. The rougher concentrate was floated for 10 minutes, transferred into a sample dish, labelled and placed into the oven to dry.
8. The dried sample was submitted to MAED Metallurgical Laboratory for gold assay.
9. The cleaner tails were transferred into a bucket and allowed to settle before being prepared for the bottle roll leach test by decanting the water and thickening the sample to a density of 1.45 kg/l.
10. The leach test procedure followed steps 3-9 of the *leach procedure*.

Table 3.7: Flotation reagent dosage

Flotation Reagent	Plant Scale Dosage (g/t)	Lab Scale Equivalent Dosage (ml)
Sodium Normal Propyl Xanthate	20	4.8
1,1,3 Triethyloxybutane	10	2.4
Alkyl Dithiophosphate	12	0.3

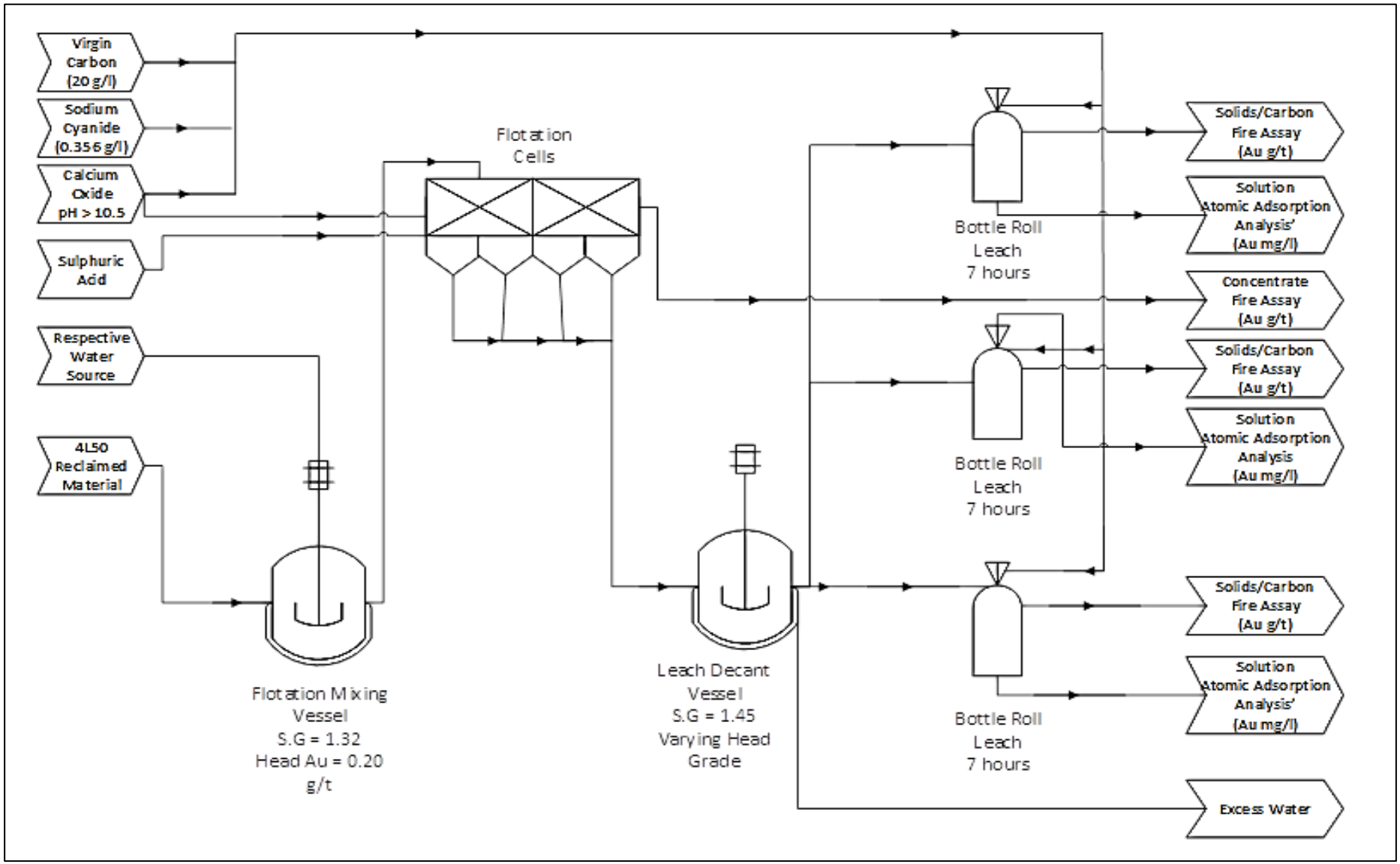

Figure 3.5: Block flow diagram illustrating the flotation and tails leach test work

4. CHAPTER 4: RESULTS AND DISCUSSIONS

4.1. STATISTICAL ANALYSIS

4.1.1. Leach results

Ten leach experimental runs were undertaken to determine if there was a significant difference in gold recovery between potable water and process water. Replication of experimental work allowed for calculation of margins of experimental error and repeatability of experimental results. The average statistical analysis results are presented in Table 4.1.

Table 4.1: Average statistical results for straight leach test work

Sample ID	Head	Washed Residue	Gold Dissolution	Gold Accountability
	g/t	g/t	%	%
Potable water	0.202 ± 0.012	0.094 ± 0.009	53.3 ± 3.3	101.3 ± 3.0
Process Water	0.202 ± 0.012	0.105 ± 0.005	47.9 ± 4.1	101.9 ± 5.9

A two-sample t-test was undertaken on the washed residues achieved to determine if there was a significant difference in gold recoveries between potable water and process water. A two-sample t-test is an inferential test that determines if there is a significant difference between the means of two data sets. The t-test determines if the two data sets come from the same population or different populations (Central Virginia Governor's School for Science and Technology, 2003). A summary of the t-test results are presented in Table 4.2. The descriptive statistical analysis is presented in Appendix C.

Table 4.2: Two-sample t-test on leach washed residue

t-Test: Paired Two Sample for Means		
	Potable water	Process Water
Mean	0.094	0,105
Variance	<0.01	<0.01
Observations	10.00	10,00
Pearson Correlation	-0.9	
Hypothesized Mean Difference	<0.01	
df	9.00	
t-statistic	6.26	
p critical one-tail	0.00	
t critical one-tail	1.83	
p critical two-tail	<0.01	
t critical two-tail	2.26	

For the t-test, as in all hypothesis testing, the computations are done assuming the null hypothesis is true. The null hypothesis for the t-test states that the means of the two washed residues are the same. The t-critical value is the cut off between retaining and rejecting the null hypothesis. When the t-statistic is further from 0 than the t-critical value, the null hypothesis is rejected and if the t- statistic is less than the t-critical value, the null hypothesis is retained (Central Virginia Governor's School for Science and Technology, 2003). Furthermore, if the p-critical value is less than 0.05, the null hypothesis is rejected, indicating that a significant difference does exist (Central Virginia Governor's School for Science and Technology, 2003).

Table 4.2 shows that the t-statistic value of 6.26 is greater than the t-critical value of 2.26, and the p-critical value obtained was 0.00015 and presented as less than 0.01; thus the null hypothesis is rejected. Therefore, the means of the washed residues for leach tests with potable water and process water respectively are not the same and there is a significant difference between them.

The same head sample was used for both tests, thus the head gold grade remained constant. A washed residue of 0.094 ± 0.009 g/t was achieved when using potable water, while a washed residue of 0.105 ± 0.005 g/t was observed when using process water. An economic evaluation (presented in Chapter 5) has shown that a small decrease in the washed residue could realise an appreciable economic benefit and approximately 10 % increase in gold recovery can be achieved when using potable water.

4.1.2. Flotation tails leach results

Flotation tests were undertaken and the flotation tails that were generated from these tests were then subjected to leaching, as described in Section 3.8.4. Statistical analysis was performed on the 10 replicates and the average are presented in Table 4.3 (Refer to Appendix C for complete statistical analysis):

Table 4.3: Average statistical results for flotation tails leach test work

Sample ID	Head	Washed Residue	Gold Dissolution	Gold Accountability
	g/t	g/t	%	%
Potable water	0.155 ± 0.013	0.084 ± 0.010	45.9 ± 4.7	102.7 ± 6.4
Process Water	0.187 ± 0.009	0.111 ± 0.007	40.5 ± 5.8	105.4 ± 5.7

A two-sample t-test was similarly undertaken on the flotation tails leach washed residues to determine if there was a significant difference in gold recoveries between potable water and process water after flotation. A summary of the t-test results is presented in Table 4.4. The descriptive statistical analysis is presented in Appendix C.

Table 4.4: Two-sample t-test on flotation tails leach washed residue

t-test: Paired two sample for means		
	Potable water	Process Water
Mean	0.084	0.111
Variance	<0.01	<0.01
Observations	10.00	10.00
Pearson Correlation	-0.22	
Hypothesized Mean Difference	0.00	
df	9.00	
t-statistic	12.10	
p critical one-tail	0.00	
t critical one-tail	1.83	
p critical two-tail	<0.01	
t critical two-tail	2.26	

The null hypothesis for the t-test states that the means of the two washed residues are the same. Table 4.4 shows that the t-statistic value of 12.10 is greater than the t-critical value of 2.26, and the p-critical value obtained was 0.0007 and presented as less than 0.01; thus the null hypothesis is rejected. Therefore, there exists a significant difference in the means of the washed residues for leach tests using potable water and process water respectively.

The average washed residue achieved after flotation using potable water was significantly lower at 0.084 ± 0.010 g/t, compared to straight leach test work, which yielded an average washed residue concentration of 0.094 ± 0.009 g/t. These results were anticipated, as most of the pyrite was expected to be concentrated during the flotation process resulting in less pyrite reporting to the flotation tails. By illustration, an initial head grade of 0,202 ± 0,012 g/t was measured before commencement of the straight leach test work, compared to the reduced flotation tails head grade of 0,155 ± 0,013 g/t achieved after flotation and before subsequent leaching.

The same behaviour was not observed for the tests conducted using the process water. An initial head grade of 0.202 ± 0.012 g/t was measured before conducting the straight leach test work, and an average washed residue of 0.105 ± 0.005 g/t was achieved. However, the reduction in the flotation tails head grade was less than that achieved when using potable water, reporting a significantly higher flotation tails head grade of 0.187 ± 0.009 g/t, thus indicating that more of the gold encapsulated pyrite had reported to the tails. As a result, a significantly higher washed residue of 0.111 ± 0.007 g/t was reported when using process water compared to the 0.084 ± 0.010 g/t achieved when using potable water. This would indicate that the flotation using potable water was significantly more efficient in concentrating the gold encapsulated pyrite.

When using potable water, an average washed residue of 0.094 ± 0.009 g/t was observed during the straight leach test work and a further reduction in the average washed residue to 0.084 ± 0.010 g/t was observed for the flotation tails leach tests. This illustrates that, for potable water, a larger overall gold recovery was achieved when performing flotation prior to leaching. The reduction in flotation efficiency with subsequent reduction in gold recovery when using the process water could be attributed to the contaminants present in this water.

4.2. THE EFFECT OF INDIVIDUAL CONTAMINANTS ON GOLD RECOVERY

The recovery of gold by cyanidation is affected by the formation of metal cyanide complexes during gold leaching. Limited literature is available on the effect of speciation of metal cyanides and the interaction between these species in gold leaching. Work undertaken by Rees (2000) provided insight to the effect of metal-cyanide complexes on gold leaching; however, his work was focused on a copper concentrate. Extensive work undertaken by Davidson & Solet (2007) clarified the major role of calcium and postulated the formation of a calcium aurocyanide complex that rapidly and strongly adsorbed onto activated carbon. Given the limited literature available, this test work was undertaken to examine the effect of the identified contaminants as well as any metal-cyanide complexes might have on gold leaching of a sulfide ore.

The contaminants identified as having the highest concentrations in the ERGO Brakpan process water are summarised in Table 4.5 below (Refer to Appendix A, Table A.1 for the complete analysis):

Table 4.5: ICP-MS analysis of ERGO Brakpan's process water

Sample ID	Mn	Fe	Ni	Na	Mg	Ca	K	Zn	Sulfate
	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l
Process Water Head Sample	24.5	135	13.3	762	193	1135	29.7	5.5	3328

Potable water was spiked with increasing concentrations of the individual contaminant species to determine the effect of each contaminant species on gold recovery. The main aim of the test work was to isolate an individual ion/group of ions that had the greatest effect of gold recovery to facilitate the consideration of more comprehensive beneficiation options rather than following a "blanket" overall treatment approach.

4.2.1. The effect of sulfur/sulfates on gold recovery

The gold processed in the 21st century is recovered from sulfide ores, and the cyanidation process has been the most important process used in the extraction of gold from its ores. Jeffrey & Breuer (2000) showed that many sulfide minerals are, to some extent, soluble in cyanide solution; thus it is expected that there will always be some sulfur species present in the leach solution. Elemental sulfur is one of the by-products of the breakdown of pyrite, and agglomerations of elemental sulfur have been observed in filtered solutions prior to ICP analysis. This sulfur is thermodynamically stable as a sulfate ion under the conditions used for cyanidation (Rees, 2000). This study aimed at establishing the effects of sulfates on the leaching efficiency of gold from a sulfide ore.

A known concentration of sulfate was mixed with potable water and used to perform laboratory scale leach tests on ERGO Brakpan's 4L50 sulfide ore. The leach test procedure presented in Section 3.8.3. was followed. The leach results as well as residual contaminant concentrations are summarised in Tables 4.6 and 4.7 respectively:

Table 4.6: Sulfate spiked potable water - leach test results

Contaminant	Sample	Head grade	Washed Residue	Gold Dissolution	Account-ability
		(g/t)	(g/t)	(%)	(%)
SULFATES	0 mg/l SO_4– addition	0.200	0.094	53.3	101.3
	50 mg/l SO_4– addition	0.200	0.113	43.5	102.6
	100 mg/l SO_4– addition	0.200	0.127	36.5	100.7
	500 mg/l SO_4– addition	0.200	0.137	31.5	102.9

Table 4.7: Initial and residual concentrations of sulfur as a contaminant

Contaminant concentrations			
Sulfates		Cyanide (as free NaCN)	
Initial (mg/l)	Final (mg/l)	Initial (mg/l)	Final (mg/l)
50	34	490	201
100	77	490	187
500	403	490	194

From Table 4.6, it is observed that addition of sulfate ions to the leach test work resulted in a significant decrease in gold recovery, with approximately 10% reduction in gold recovery observed from a 50 mg/l addition of sulfates. Furthermore, gold recovery decreased as the sulfate concentration increased. The leach pH was adjusted to above pH 10.5 using lime (CaO) to prevent volatilization of poisonous HCN gas. At the high pH, the calcium ions could have reacted with the sulfate ions to produce gypsum ($CaSO_4 \cdot 2H_2O$) which fouls sites on activated carbon, thus reducing gold recovery.

Fink & Putnam (1950) illustrated that the addition of trace amounts of sodium sulfide to the cyanide solution dramatically hindered gold leaching, postulating that the sulfide ions passivate the surface of

the gold by forming a passive layer of Au_2S. Jeffrey and Breuer (2000) showed similar results using an ore containing 5 % silver. Most naturally occurring gold sulfide ores contains trace amounts of silver which has proven to increase the cyanide leaching reaction rate (Jeffrey & Breuer, 2000). The leaching behaviour of gold in the absence of hydrosulpfide was found to be significantly different to that in its presence, indicating that gold leached much more readily when hydrosulfide was not present (Jeffrey & Breuer, 2000). In this study, this was pertinent because sulfide encapsulated within gold ores may react with hydrogen ions and passivate the surface of the gold by forming a passive layer, thus reducing gold recovery. Furthermore, the ore processed at ERGO Brakpan is predominantly sulfide in nature.

The results summarised in Table 4.6 are presented graphically in Figure 4.1.

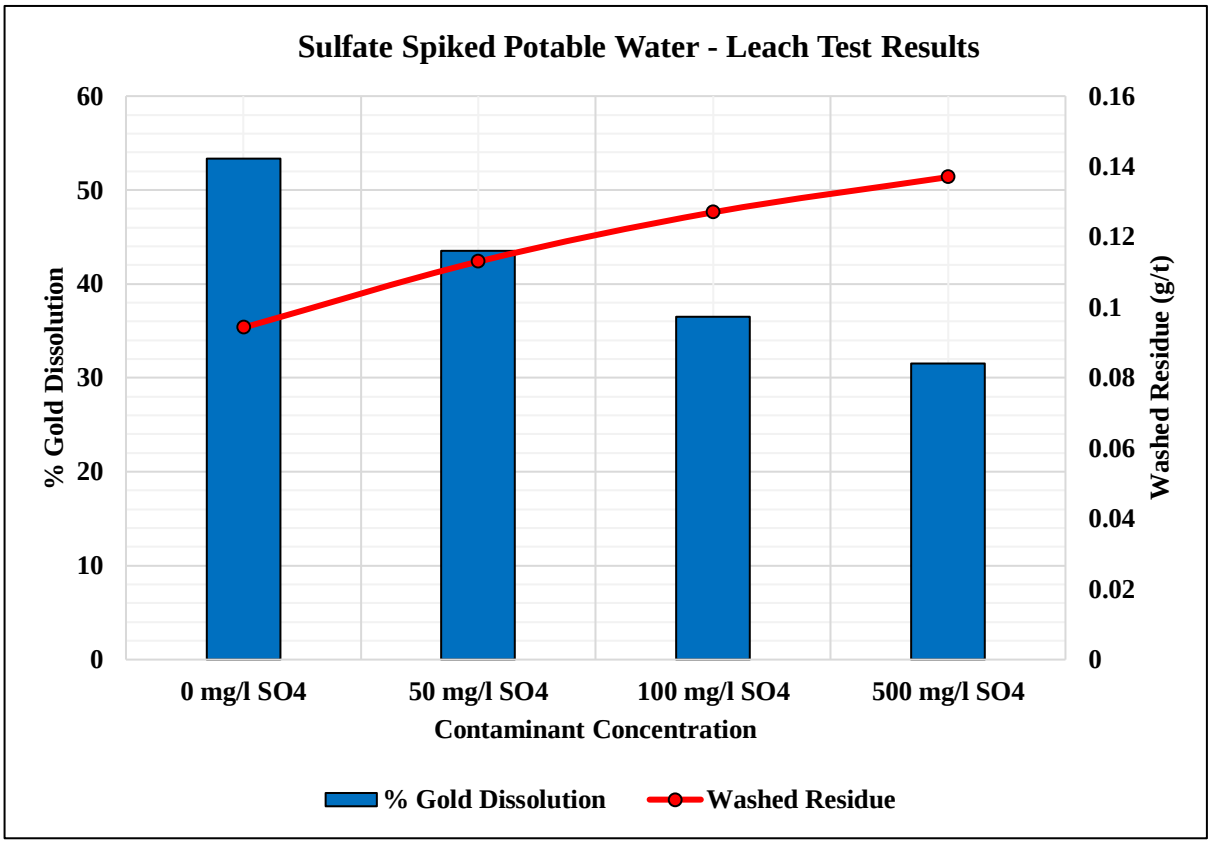

Figure 4.1: Sulfate spiked potable water - leach test results

4.2.2. The effect of calcium on gold recovery

Whilst the Elsner equation (refer to Equation 2.6) accurately describes the cyanidation of gold under controlled conditions, its validity is questioned when related to actual plant conditions. Under plant conditions, the addition of lime together with the addition of either calcium or sodium cyanide is the general practice. The addition of calcium through lime and calcium cyanide may have a significant effect on subsequent gold dissolution and on the solubility of the resulting aurocyanide complex. The present study aimed to establish the effect that calcium has on the leaching of gold from a sulfide ore.

A known concentration of calcium (added as CaO) was mixed with potable water and used to perform laboratory scale leach tests on ERGO Brakpan's 4L50 sulfide ore. The leach test procedure presented in Section 3.8.3 was followed. The leach test results as well as residual contaminant concentrations are summarised in Table 4.8 and 4.9 respectively:

Table 4.8: Calcium spiked potable water - leach test results

Contaminant	Sample	Head grade	Washed Residue	Gold Dissolution	Account-ability
		(g/t)	(g/t)	(%)	(%)
CALCIUM	0 mg/l Ca^{2+} addition	0.200	0.094	53.3	101.3
	50 mg/l Ca^{2+} addition	0.200	0.133	33.3	102.9
	100 mg/l Ca^{2+} addition	0.200	0.122	38.9	100.3
	500 mg/l Ca^{2+} addition	0.200	0.117	41.7	101.2

Table 4.9: Initial and residual concentrations of calcium as a contaminant

Contaminant concentrations			
Calcium		Cyanide (as free NaCN)	
Initial (mg/l)	Final (mg/l)	Initial (mg/l)	Final (mg/l)
50	25	665	251
100	36	833	302
500	156	2182	456

From Table 4.8, it was observed that the addition of 50 mg/l of calcium had a negative effect on gold recovery showing a 20 % decrease in gold recovery compared to that achieved when using potable water without added calcium. This may have resulted from poor gold adsorption caused by passivation

of carbon surfaces by calcium foulants. As the initial calcium concentrations were increased above 50 mg/l, gold recoveries showed improvement with an increase in gold recovery of approximately 8.35 % observed between potable water spiked with 50 mg/l and 500 mg/l of calcium respectively. Extensive work undertaken by Davidson & Solet (2007) explained the major role played by calcium in gold recovery and hypothesised the formation of a calcium aurocyanide complex that rapidly and strongly adsorbed onto activated carbon (refer to Equation 2.5). These results would support the formation of a calcium aurocyanide complex as the gold recovery increased with an increased calcium concentration. Davidson & Solet (2007) further stated that the preferential adsorption of calcium aurocyanide onto activated carbon was due to the limited solubility of the calcium aurocyanide complex, when compared with the higher solubility of potassium and sodium aurocyanide (Davidson & Solet, 2007) The residual calcium concentrations shown in Table 4.9, indicate that a large amounts of calcium were consumed during the tests. This could indicate that the calcium had partially reacted with contaminants within the ore to foul the carbon as well as to form a calcium aurocyanide complex that assisted to increase gold recovery at the higher free calcium concentrations.

The results summarised in Table 4.8 is presented graphically in Figure 4.2:

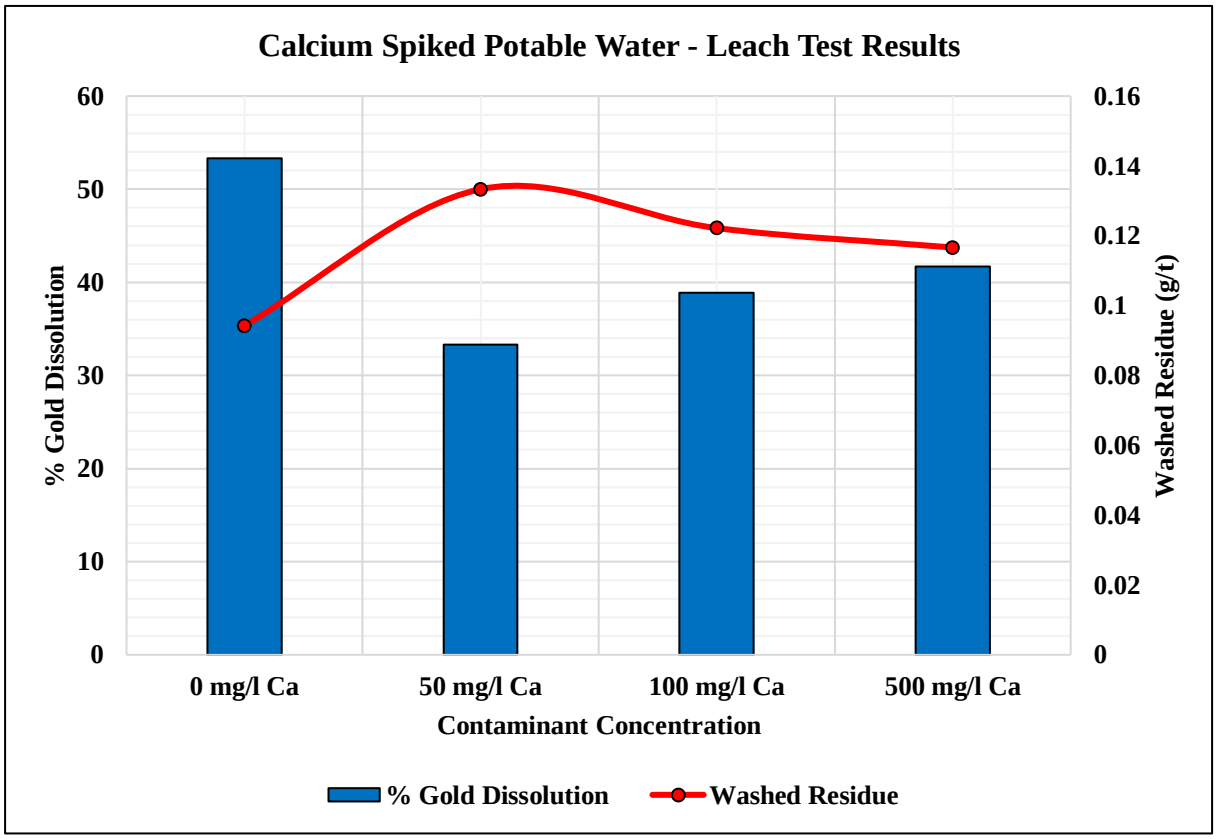

Figure 4.2: Calcium spiked potable water - leach test results

4.2.3. The effect of heavy metals on gold recovery

Literature on the effect of speciation of metal cyanides in gold leach slurries and the interaction between these metal cyanide species is limited and has not been researched extensively. The implications of these metal cyanide species on gold leaching in cyanide solutions are of great interest. In most cases, gold ores are considered as refractory ores due to their excessive cyanide consumption (Rees, 2000). If this consumption is attributed to the formation of metal cyanide species that compete with the aurocyanide complex for sites on activated carbon, then the effects of these metal cyanide species on gold leaching in cyanide solutions should be established. The role of a variety of different metal-cyanide species on gold recovery are investigated within this research.

Examination of the stability constants of various metal-cyanide species (Rees, 2000) given in Table 4.10, shows that aurocyanide is one of the most stable metal cyano complexes. Iron (II) and iron (III) also form strong complexes, with the iron (III) complex marginally stronger and the iron (II) complex marginally weaker than the gold cyanide complex (Rees, 2000). Nickel forms a strong cyanide complex in the +2 oxidation state, with zinc forming a weaker cyanide complex (Rees, 2000). Due to the stability of these species being similar, these complexes are all present to some extent in ore leach solutions, and their predominance is determined by the free cyanide concentration and is strongly pH dependent (Rees, 2000). If not removed, these metal complexes accumulate and circulate through the process affecting vital downstream processes with subsequent reduction of gold recovery. The process water on many gold processing plants also contains a considerable amount of weak acid dissociable (WAD) cyanide species.

The stability constants are listed in terms of log K. where K is the constant for the equilibrium. Ferrocyanide is shown as an example in Equation 4.1 (Rees, 2000):

$$K = \frac{[Fe(CN_6^{4-})]}{[Fe^{2+}].[CN^-]^6} \tag{4.1}$$

Table 4.10: Stability constants for metal cyanide species (Rees, 2000)

Species	Stability (log K)	Species	Stability (log K)
$Au(CN)_2^-$	39.3	$Fe(CN)_6^{4-}$	35.4
$Ag(CN)_2^-$	20.5	$Fe(CN)_6^{3-}$	43.6
$Cu(CN)_2^-$	16.3	$Ni(CN)_4^{2-}$	30.2
$Cu(CN)_3^{2-}$	21.7	$Zn(CN)_4^{2-}$	19.6
$Cu(CN)_4^{3-}$	23.1		

4.2.3.1. The effect of iron on gold recovery

Rees (2000) had demonstrated that a large number of interactions can occur between different metal ions and cyanide species. Rees (2000) further proved that these interactions were dependent on pH, and the concentrations of cyanide and metal species. The present study aimed to establish the effect that iron has on the leaching of gold from a sulfide ore.

A known concentration of iron (added as $FeSO_4$) was mixed with potable water and utilised to perform laboratory scale leach tests on ERGO Brakpan's 4L50 sulfide ore. The leach test procedure highlighted in Section 3.8.3 was followed. The leach results as well as residual contaminant concentrations are summarised in Tables 4.11 and 4.12 respectively:

Table 4.11: Iron spiked potable water - leach test results

Contaminant	Sample	Head grade	Washed Residue	Gold Dissolution	Account-ability
		(g/t)	(g/t)	(%)	(%)
Iron	0 mg/l Fe^{2+} addition	0.200	0.094	53.3	101.2
	50 mg/l Fe^{2+} addition	0.200	0.132	34.0	108.3
	100 mg/l Fe^{2+} addition	0.200	0.158	20.8	102.2
	500 mg/l Fe^{2+} addition	0.200	0.188	6.0	101.6

The data shows that an increase in iron concentration, together with the addition of a stoichiometric amount of cyanide, resulted in a decrease in gold recovery. Since the cyanide addition was stoichiometric based on the oxidative breakdown of pyrite ore, the iron is expected to be present in the ferrous form and in the presence of cyanide would predominantly form the $Fe(CN)_6^{4-}$ complex (Rees, 2000). This is a very stable complex and it has been debated in literature whether this complex would dissociate. As shown in Table 4.10, the stability constants of both aurocyanide as well as the $Fe(CN)_6^{4-}$ complex are similar (Kyle, 1997). The results summarised in Table 4.11 show that as the initial iron concentration was increased from 50 mg/l to 500 mg/l, a 28 % reduction of gold recovery occurred. These results are supported by Kyle (1997) who states that excess ferrous ions will readily displace gold from the aurocyanide complex.

Since the iron was added as iron (II) sulfate, it can be debated that the effect on gold recovery was also affected by the sulfate ions present. While this holds true, a comparison of the results in Table 4.6 and Table 4.11, show that gold recoveries obtained from sulfate addition alone were much higher than those in which iron sulfate was added. This would indicate that the detrimental effect on the gold recovery was due to the presence of the iron.

Table 4.12: Initial and residual concentrations of iron as a contaminant

Contaminant concentrations			
Iron		**Cyanide (as free NaCN)**	
Initial (mg/l)	Final (mg/l)	Initial (mg/l)	Final (mg/l)
50	15	860	416
100	26	1222	523
500	157	4128	1112

It is observed from Table 4.12 that not all of the cyanide was consumed at the conclusion of the experiment and that cyanide was still available for reaction. Rees (2000) states that a large amount of residual free cyanide was most likely attributed to weakly bound metal complexes. Also, under plant conditions, the addition of lime is general practice to maintain a pH of above 10.5. These alkaline conditions would result in metal precipitation as metal hydroxides, thus reducing the metal concentration and liberating free cyanide into solution (Damons, 2001). Therefore, the low residual iron concentrations are attributed to precipitation as a metal hydroxide under alkaline conditions and the formation of the stable $Fe(CN)_6^{4-}$ that would have competed for and occupied sites on the activated carbon, thus reducing gold recovery.

The results summarised in Table 4.11 are presented graphically in Figure 4.3:

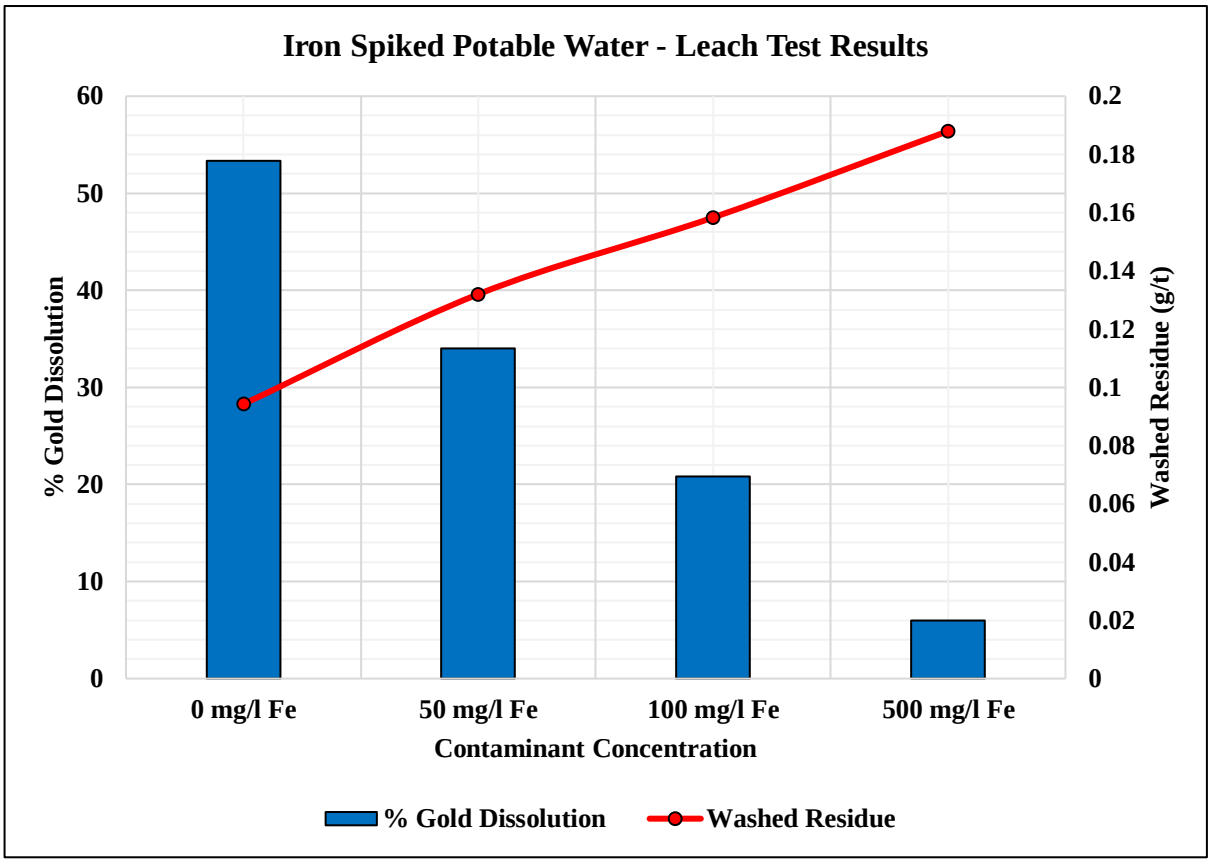

Figure 4.3: Iron spiked potable water - leach test results

4.2.3.2. The effect of nickel on gold recovery

The effect that nickel has on the leaching of gold from a sulfide ore was investigated. A known concentration of nickel (added as $NiSO_4$) was mixed with potable water and used to perform laboratory scale leach tests on ERGO Brakpan's 4L50 sulfide ore. The leach test procedure shown in Section 3.8.3 was followed. The leach results as well as residual contaminant concentrations are summarised in Tables 4.13 and 4.14 respectively:

Table 4.13: Nickel spiked potable water - leach test results

Contaminant	Sample	Head grade	Washed Residue	Gold Dissolution	Account-ability
		(g/t)	(g/t)		(%)
Nickel	0 mg/l Ni^{2+} addition	0.200	0.094	53.3	101.3
	50 mg/l Ni^{2+} addition	0.200	0.122	39.0	101.5
	100 mg/l Ni^{2+} addition	0.200	0.146	27.0	100.1
	500 mg/l Ni^{2+} addition	0.200	0.168	16.0	102.0

Results show that an increase in nickel concentration, together with the addition of the stoichiometric amount of cyanide, resulted in a detrimental effect on gold recovery. From the stability constants presented in Table 4.10, nickel indicates that a medium strength cyanide complex will be formed. This complex is nine orders of magnitude less stable than the aurocyanide complex (Kyle, 1997). Due to the lower stability of the $Ni(CN)_4^{2-}$ complex when compared to the stability of the $Fe(CN)_6^{4-}$ complex, it was expected and proven that the presence of iron has a greater detrimental effect on gold recovery, as shown in Section 4.2.3.1 above. Additionally, the $Ni(CN)_4^{2-}$ complex has 4 CN^- ions compared to 2 CN^- ions in $Au(CN)_2^-$, which results in higher hydration. This allows a better dispersion in the solution without forming clusters, which enables adsorption onto activated carbon when found in large concentrations (Sayiner & Acarkan, 2013). Similar behaviour was evident in the current study, resulting in a decrease in gold recovery as the initial nickel concentration was increased. The results in Table 4.13 indicate that the presence of nickel had a negative effect on gold recovery, probably as a result of excess nickel ions readily displacing gold from the aurocyanide complex. An increase in nickel concentration from 50 mg/l to 500 mg/l resulted in a 23 % reduction of gold recovery. Furthermore, due to the lower stability of the nickel-cyanide complex as compared to the aurocyanide complex, the precipitation of nickel hydroxide would have been favoured at lower nickel concentrations due to the alkaline leach pH.

Comparison of the results in Table 4.6 and Table 4.13, showed that the gold recoveries between sulfate addition alone and nickel sulfate addition were significantly different. Leach tests with sulfate addition alone produced significantly higher gold recoveries than those in which nickel sulfate was added. This would indicate that the presence of nickel, rather than the sulfate ions caused the negative effect on gold recovery.

Table 4.14: Initial and residual concentrations of nickel as a contaminant

Contaminant concentrations			
Nickel		**Cyanide (as free NaCN)**	
Initial (mg/l)	Final (mg/l)	Initial (mg/l)	Final (mg/l)
50	11	727	302
100	23	950	416
500	92	2800	1024

Data in Table 4.14 show large residual cyanide concentrations for all tests, which attributed to the liberation of free cyanide resulting from the metal precipitation occurring at high pH conditions. Therefore, the low contaminant residual concentrations are attributed to the precipitation of the metal hydroxide under alkaline conditions and the formation of the stable $Ni(CN)_4^{2-}$ that occupied sites on the activated carbon, and thus reducing gold recovery.

The results summarised in Table 4.13 are presented graphically in Figure 4.4:

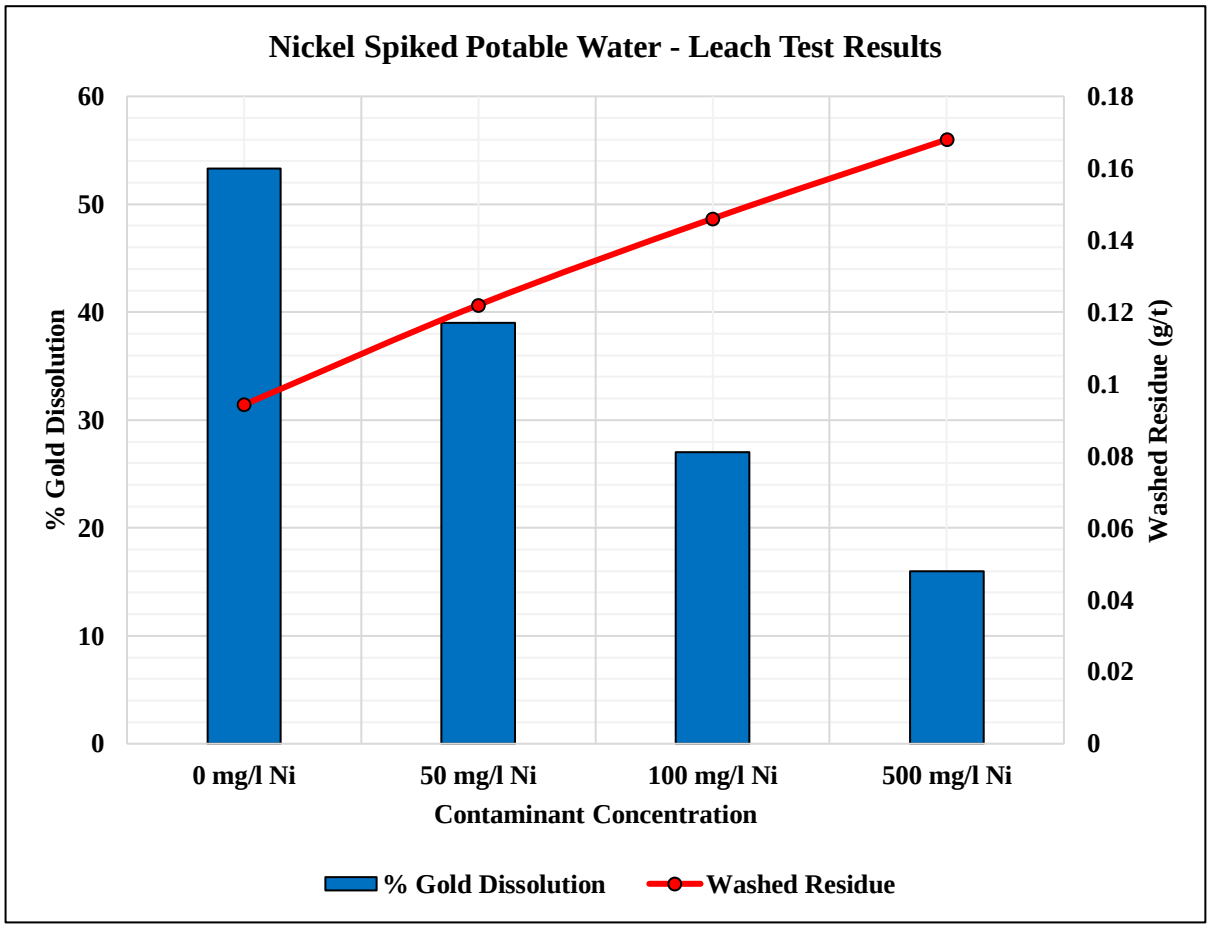

Figure 4.4: Nickel spiked potable water - leach test results

4.2.3.3. The effect of zinc on gold recovery

In gold processing plants, the process water contains a considerable amount of weak acid dissociable (WAD) cyanide species. Zinc cyanide can be a major constituent of the process water due to the cyanide leaching of zinc minerals and the zinc cementation applied to precipitate gold and silver (Merrill-Crowe process). This study aimed to establish the effect that zinc has on the leaching of gold from a sulfide ore.

A known concentration of zinc (added as $ZnSO_4$) was mixed with potable water and utilised to perform laboratory scale leach tests on ERGO Brakpan's 4L50 sulfide ore. The leach test procedure shown in Section 3.8.3 was followed. The leach results as well as residual contaminant concentrations are summarised in Tables 4.15 and 4.16 respectively.

Table 4.15: Zinc spiked potable water - leach test results

Contaminant	Sample	Head grade	Washed Residue	Gold Dissolution	Account-ability
		(g/t)	(g/t)	(%)	(%)
Zinc	0 mg/l Zn^{2+} addition	0.200	0.094	53.3	101.3
	50 mg/l Zn^{2+} addition	0.200	0.118	41.0	101.7
	100 mg/l Zn^{2+} addition	0.200	0.130	35.0	99.1
	500 mg/l Zn^{2+} addition	0.200	0.141	29.5	102.9

The results show that an increase in zinc concentration, together with the addition of the stoichiometric amount of cyanide, had resulted in a negative effect on gold recovery. As observed from data presented in Table 4.10, zinc forms a weak cyanide complex and precipitation of this species at a high pH is rapid compared to the other metal species (Rees, 2000). Kyle (1997) explains that zinc predominantly forms a tetra cyano complex at a pH of 10 that may dissociate as the cyanide concentration is decreased. It is therefore possible that the zinc in solution rapidly precipitates, thus passing its cyanide ligands to free gold in solution (Kyle, 1997). Furthermore, due to the lower stability of the $Zn(CN)_4^{2-}$ complex when compared to the stability of the both the $Ni(CN)_4^{2-}$ and $Fe(CN)_6^{4-}$ complexes, it is evident that the zinc cyanide complex negatively effects gold recovery only to a degree. The presence of both nickel and iron has proven to have a far greater detrimental effect on gold recovery. It is postulated that the bulk

of the zinc introduced would have formed zinc hydroxide and precipitated out of solution rapidly due to the high leach pH.

Comparison of the results in Tables 4.6 and 4.15, show that the gold recoveries obtained were significantly higher with sulfate addition alone was undertaken compared to tests where zinc sulfate addition was undertaken; thus indicating that the presence of zinc to have a negative effect on gold recovery.

Table 4.16: Initial and residual concentrations of zinc as a contaminant

Contaminant concentrations			
Zinc		**Cyanide (as free NaCN)**	
Initial (mg/l)	Final (mg/l)	Initial (mg/l)	Final (mg/l)
50	5	600	256
100	12	704	333
500	32	1530	652

From data shown in Table 4.16, it is evident that a large concentration of zinc was consumed in the present study. At the high leach pH, the process of metal precipitation occurs, and the associated liberation of cyanide would result in high residual cyanide values. Additionally, zinc forms a weak cyanide complex and precipitation of this species at a high pH is rapid in comparison to the other metal species (Rees, 2000). Therefore, the low residual zinc concentrations are attributed to the formation of the low stability $Zn(CN)_4^{2-}$ that occupied sites on activated carbon, and thereby reducing gold recovery, and the more predominant reduction of zinc concentration due to the rapid precipitation of zinc as a metal hydroxide under alkaline conditions. The results summarised in Table 4.15 are presented graphically in Figure 4.5.

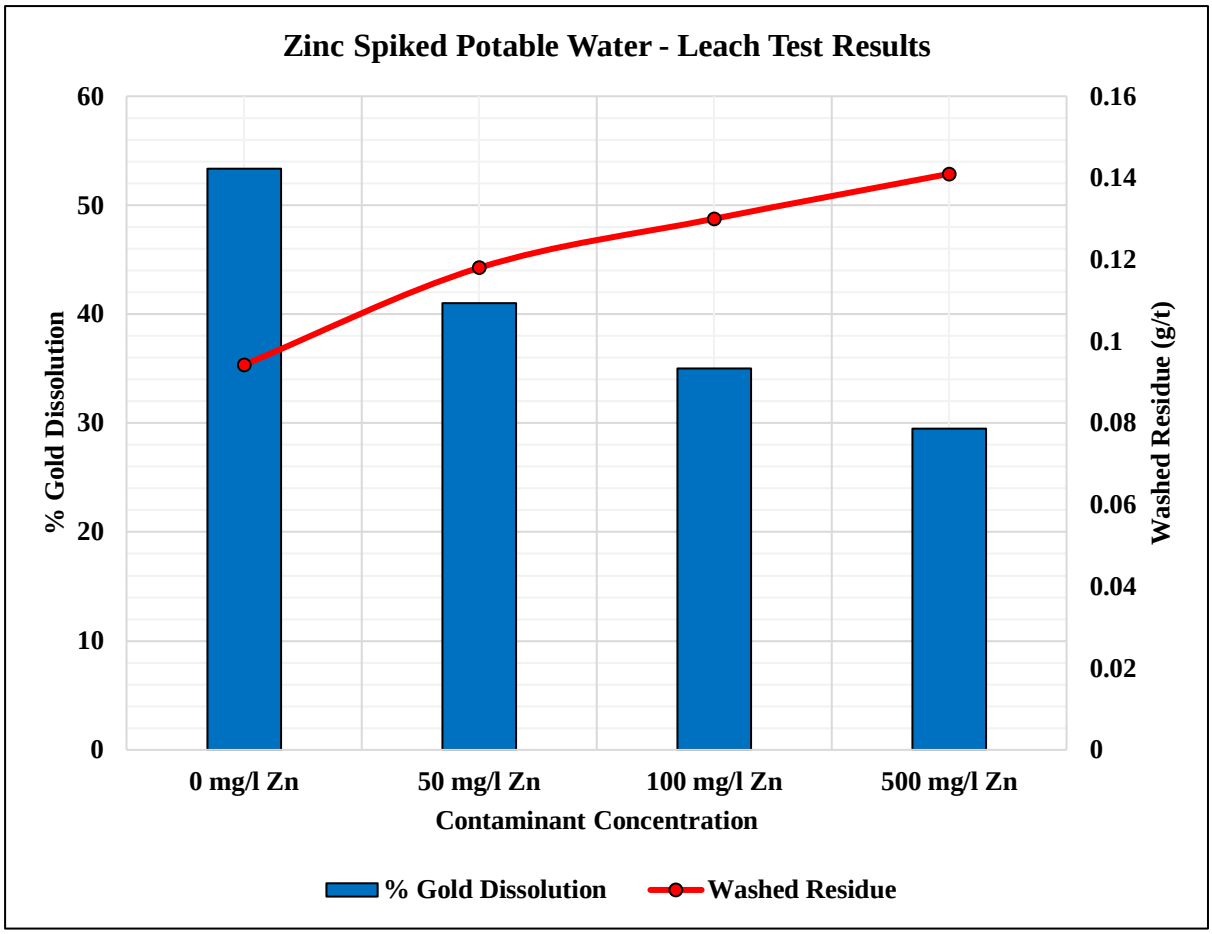

Figure 4.5: Zinc spiked potable water - leach test results

4.2.4. The effect of magnesium on gold recovery

The role of magnesium and its effect on gold leaching has not been experimentally investigated in literature before. The present study aimed to establish the effect that magnesium has on the leaching of gold from a sulfide ore.

A known concentration of magnesium (added as $MgSO_4$) was mixed with potable water and used to perform laboratory scale leach tests on ERGO Brakpan's 4L50 sulfide ore. The leach test procedure shown in Section 3.8.3 was followed. The leach results as well as residual contaminant concentrations are summarised in Tables 4.17 and 4.18 respectively:

Table 4.17: Magnesium spiked potable water - leach test results

Contaminant	Sample	Head grade	Washed Residue	Gold Dissolution	Account-ability
		(g/t)	(g/t)	(%)	(%)
Magnesium	0 mg/l Mg^{2+} addition	0.200	0.094	53.3	101.3
	50 mg/l Mg^{2+} addition	0.200	0.125	37.5	104.4
	100 mg/l Mg^{2+} addition	0.200	0.122	39.0	107.5
	500 mg/l Mg^{2+} addition	0.200	0.120	40.00	105.7

Table 4.18: Initial and residual concentrations of magnesium as a contaminant

Contaminant concentrations			
Magnesium		Cyanide (as free NaCN)	
Initial (mg/l)	Final (mg/l)	Initial (mg/l)	Final (mg/l)
50	22	775	356
100	48	1053	689
500	201	3277	1203

From Table 4.17, it was observed that the addition of 50 mg/l of magnesium had a negative effect on gold recovery showing a 15.8 % decrease in gold recovery compared to that achieved when using potable water without added magnesium. Passivation of active carbon surfaces by magnesium foulants may have caused the poor gold adsorption. As the magnesium concentrations were increased above 50 mg/l, gold recoveries showed improvement with an increase in gold recovery of approximately 2.5 % observed between potable water spiked with 50 mg/l and 500 mg/l of magnesium, respectively. These results showed that when magnesium was found in large excess, it aids gold recovery. The residual magnesium concentrations shown in Table 4.18, indicate that a large amount of magnesium was consumed during the tests.

The results summarised in Table 4.17 are presented graphically in Figure 4.6:

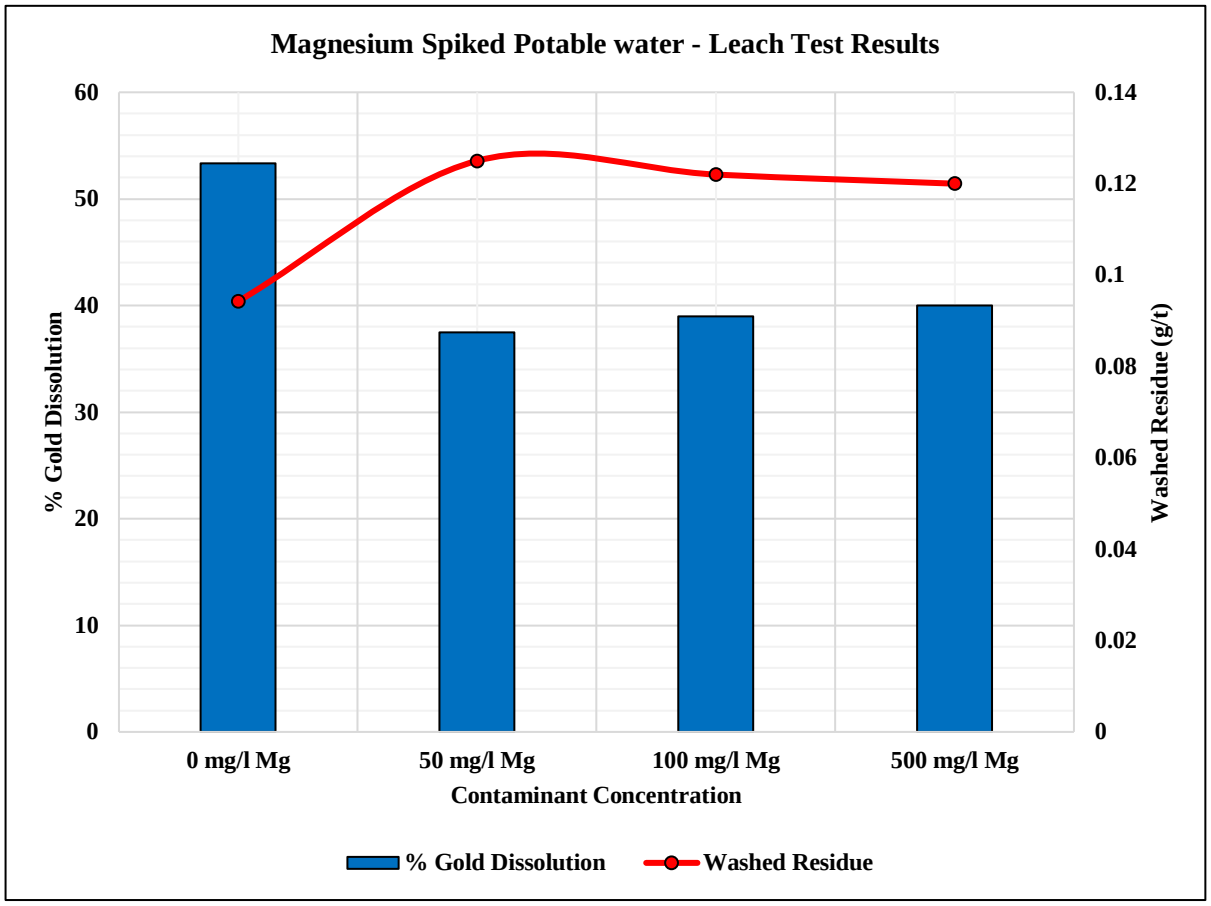

Figure 4.6: Magnesium spiked potable water - leach test results

4.3. TREATMENT OF ERGO BRAKPAN'S PROCESS WATER AND BENEFICIATION RESULTS

4.3.1. Chemical precipitation – lime and soda ash softening

The results presented in Section 4.2, show that the presence of heavy metals such as Fe, Ni and Zn have the largest negative effect on gold recovery. The presence of sulfates also illustrated a possible passivation effect influencing gold recovery. From the comprehensive analysis of ERGO Brakpan's process water, where data are summarised in Table A.1 in Appendix A, it is evident that the process water contains a considerable amount of heavy metals which could result in the formation of weak acid dissociable (WAD) cyanide species.

It is further observed, from Table 4.5, that the sulfate and calcium concentrations within the ERGO Brakpan process water are significantly higher (approximately 3400 mg/l and 1135 mg/l respectively) than the other contaminants present. The results presented in Section 4.2.2, show that a high calcium

concentration may be beneficial and aid in improving gold recovery, due to the formation of a calcium aurocyanide complex that is rapidly and strongly adsorbed onto activated carbon (refer to Equation 2.5).

In Section 4.2.1, it was evident that the presence of sulfur minerals has a negative effect on gold recovery. Elemental sulfur is one of the by-products of the breakdown of pyrite. This sulfur is thermodynamically stable as a sulfate under the conditions used for cyanidation (Rees, 2000). Results indicated that an increase in sulfur/sulfate concentration resulted in a lower recovery of gold. These results were of particular interest as most gold mining, extraction and recovery operations in South Africa are from sulfide ores. Due to the significantly higher concentration of sulfates in ERGO Brakpan's process water, it was initially proposed to utilise the SAVMIN process as a viable beneficiation option. The SAVMIN process is a chemical precipitation process that reduces the concentration of harmful Ca^{2+}, SO_4^{2-} and heavy metals by precipitation to form gypsum, ettringite and metal hydroxides. Pre-treatment involves the addition of lime to acid mine drainage (AMD) to precipitate metal hydroxides with the subsequent formation of ettringite to remove sulfates and calcium by addition of aluminium hydroxide.

While the SAVMIN process may prove to be the most viable benefication option for sulfate reduction in ERGO Brakpan's process water, from the results presented in Section 4.2, it was clearly evident that the presence of heavy metals had a greater negative effect on gold recovery than the presence of sulfates. Additionally, many operating difficulties of the SAVMIN process have been highlighted. Summers (2019) had indicated that the formation of gypsum and ettringite is highly sensitive to changes in pH, therefore high capital and maintenance costs were observed in his pilot plant study. Summers (2019) added that additional operating costs were also incurred due to excessive addition of aluminium hydroxide for successful ettringite formation as well as costs associated with flocculant addition, which was required to increase the ettringite settling rate (Summers, 2019).

ERGO Brakpan operations focus on the retreatment of vast quantities of surface mine tailings material (approximately 64 000 t/day). The process is water intensive requiring approximately 60 Ml/day of water per day. Due to the large quantity of process water that would be required for daily operation at ERGO Brakpan, it was established that the SAVMIN process would not be economically viable for sulfate reduction.

Reverse osmosis (with pre-treatment) would also remove heavy metals and sulfates and yield a water quality similar to or better than potable water. However, the complexity of the pre-treatment, capital cost and the generation of a high salinity brine (with disposal challenges) make reverse osmosis unviable.

Minnesota Rural Water Association (2009) state that chemical precipitation is one of the most common methods utilised for water purification. The addtion of chemicals such as lime (calcium hydroxide,

$Ca(OH)_2$) and soda ash (sodium carbonate, Na_2CO_3) increases the water's pH level and precipitates the ions that cause hardness. The lime and soda ash softening processes offer many benefits, including the reduction of dissolved minerals in the water and the reduction of heavy metals by metal hydroxide precipitation. Lime softening is implemented as the first step of the SAVMIN process, followed by the addition of gypsum ($CaSO_4 \cdot 2H_2O$) crystals which provide active surfaces, which act as a catalyst for the precipitation of the 'supersaturated' gypsum thus reducing the sulfur/sulfate concentration.

Since lime addition for increasing leach pH is general practice at most gold leach plants, and given its easy accessibility, it could be considered a viable and economical beneficiation option. Furthermore, the solubilities of calcium, magnesium, and silica are significantly reduced by increased temperature and they are therefore, more effectively removed by warm/hot softening than by cold/ambient softening (Suez Water Technologies & Solutions, n.d.). The following test work aimed to establish the viability of lime and soda ash softening, at various temperatures, as an economic beneficiation option. The lime softening procedure and lime-soda ash softening procedure presented in Sections 3.8.1 and 3.8.2 was followed respectively. The required dosages of lime (CaO) and soda ash (Na_2CO_3) were calculated following the procedure outlined by Minnesota Rural Water Association (2009) and summarised in Table 3.1. The lime softening and soda ash softening test work parameters and final pH and resultant precipitate masses are summarised in Table 4.19 and Table 4.20:

Table 4.19: Lime softening test parameters

Lime softening parameters						
Sample ID	**Initial pH**	**Lime Addition (mg/l)**	**Volume Treated (L)**	**Temperature (°C)**	**pH After**	**Mass of Precipitant (g)**
Ca^{2+} Removal - Test 1	5.46 ± 0.05	756.0 ± 2.0	9.50 ± 0.01	19.1± 0.5	10.90 ± 0.05	7.7 ± 0.2
Ca^{2+} Removal - Test 3	4.30 ± 0.05	1170.0 ± 2.0	10.0 ± 0.01	60.0 ± 0.5	10.92 ± 0.05	15.5 ± 0.2
Ca^{2+} Removal - Test 5	3.61 ± 0.05	1170.0 ± 2.0	10.0 ± 0.01	90.0 ± 0.5	11.20 ± 0.05	28.9 ± 0.2

Table 4.20: Lime and soda ash softening test parameters

Lime and soda ash softening parameters							
Sample ID	**Initial pH**	**Lime Addition (mg/l)**	**Soda Ash Addition (mg/l)**	**Volume Treated (L)**	**Temperature (°C)**	**pH After Addition**	**Mass of Precipitant (g)**
Ca^{2+} + Mg^{2+} Removal - Test 2	5.54 ± 0.05	756.0 ± 2.0	180.0 ± 2.0	9.50 ± 0.01	19.1 ± 0.5	11.02 ± 0.05	10.9 ± 0,2
Ca^{2+} + Mg^{2+} Removal - Test 4	4.67 ± 0.05	1170.0 ± 2.0	180.0 ± 2.0	10.0 ± 0.01	60.0 ± 0.5	11.00 ± 0.05	17.9 ± 0,2
Ca^{2+} + Mg^{2+} Removal - Test 6	3.72 ± 0.05	1170.0 ± 2.0	180.0 ± 2.0	10.0 ± 0.01	90.0 ± 0.5	11.30 ± 0.05	32.2 ± 0,2

The data in Tables 4.19 and 4.20, show that at ambient conditions, the dosage proved valid, producing the desired pH > 10.5, for effective chemical precipitation. However, due to the inverse relationship between temperature and pH, it is evident that as the temperature of the process water increased, a greater amount of lime was required to achieve the desired pH. Additionally, more lime reacted to form precipitates at higher temperature, resulting in a higher lime demand.

Figure 4.7 illustrates the relationship between process water temperature and the mass of precipitant formed:

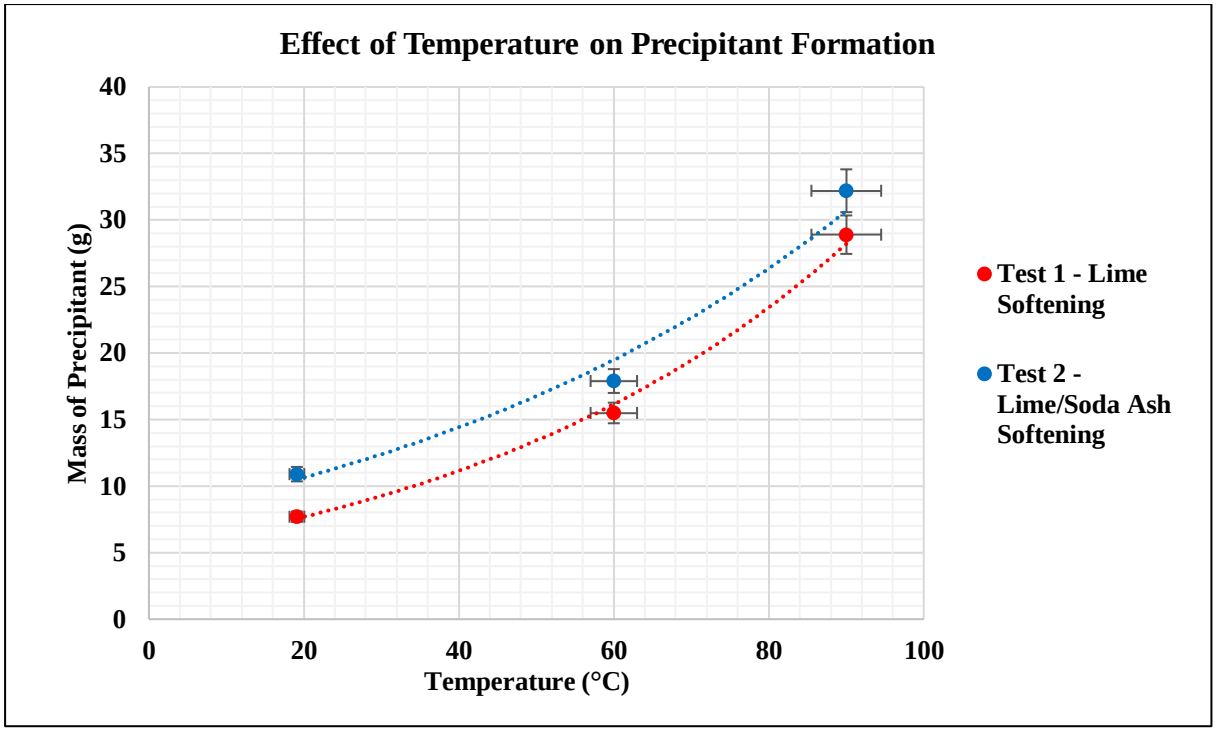

Figure 4.7: Relationship between temperature and mass of precipitant formed

From Figure 4.7, it is observed that as the temperature of process water was increased, the mass of precipitant formed also increased. This is attributed to both the additional lime, required to obtain the desired pH as temperature increased, as well as the reduced solubilities of calcium and other contaminants at higher temperature.

4.3.2. Visual results of the lime and soda ash softening process

Figure 4.8 illustrates two 5 litre process water samples that were treated with unslaked lime at the required dosage to obtain a pH of 10.5. The same procedure was repeated after heating the process water to the following temperatures: 20 °C, 60 °C and 90 °C. The two 5 litre glass beakers were placed into a warm bath which was placed on a hot plate.

Figure 4.8: Two 5 litre glass beakers of process water treated with lime (CaO)

A one hour waiting period was observed to allow complete precipitation of contaminants. Figure 4.9 illustrates the settling of precipitants formed during the lime and soda ash treatment.

Figure 4.9: Settling of precipitants

The resulting solution was filtered through a vacuum filter and the precipitate collected, dried, weighed and submitted for XRD analysis. The filtered precipitate from the lime-soda ash softening test at ambient conditions is illustrated in Figure 4.10:

Figure 4.10: Filtered precipitate for lime-soda ash softening at ambient conditions

Figure 4.11 shows a comparison between ERGO Brakpan's untreated process water and samples that have undergone lime and soda ash treatment/softening.

Figure 4.11: Comparison of process water after treatment

4.3.3. X-ray diffraction analysis of precipitants

The aim of the treatment of ERGO Brakpan's process water with lime, as well as a combination of both lime and soda ash was twofold. These chemicals would react with the hardness and natural alkalinity in the water to form insoluble compounds that precipitate out of solution and also the precipitation of relatively insoluble metal hydroxides, sulfates and carbonates, which could be removed from the water by filtration. (Damons, 2001). To achieve precipitation of metals as hydroxides, the solution pH was adjusted by lime addition to above 10.5.

X-Ray Diffraction (XRD) analysis was undertaken on the precipitates formed for each water treatment test. The XRD scans are presented in Appendix D, Figures D.1-D.6. It is evident from the XRD analysis that tests in which only lime treatment was undertaken, a large number of calcium containing compounds precipitated from ERGO Brakpan's process water. The main constituents were identified as quartz, magnesium silicate hydrate, calcium silicate, cordierite (identified as anthophyllite), basanite and gypsum. The formation of gypsum ($CaSO_4 \cdot 2H_2O$) and basanite ($CaSO_4 \cdot 5H_2O$) would consume free sulfates in solution, thus resulting in a reduction of sulfates in the process water.

XRD scans undertaken on the precipitates formed when the process water was treated with a combination of lime followed by soda ash, indicated that lime treatment alone was more efficient. At ambient conditions, it was evident that only calcium containing compounds had precipitated (namely gypsum, quartz and calcite). With the addition of soda ash, it was expected that magnesium containing compounds would also be present within the precipitate. The addition of soda ash produced precipitates with fewer identifiable calcium and magnesium containing compounds as compared to the tests in which only lime treatment had been undertaken. This behaviour was observed at all temperatures investigated. Nevertheless, an increase in temperature resulted in a larger amount of identifiable calcium and magnesium compounds within the precipitate and that the solubility of calcium and magnesium had decreased, resulting in a more effective removal by chemical precipitation.

The presence of metal hydroxides was not identified from the XRD analysis. This is attributed to the lower concentrations of heavy metals within ERGO Brakpan's process water compared to the concentrations of calcium and sulfates. The calcium and sulfate concentrations were significantly higher and would have supersaturated the process water (refer to the ICP-MS analysis of ERGO Brakpan's process water in Table 4.5). The atomic absorption data presented in Section 4.3.4 did, however, indicate that lime and soda ash addition significantly reduced the heavy metal concentrations.

4.3.4. Atomic absorption analysis

Tables 4.21 and 4.22 summarise the duplicate atomic absorption (AA) results for the potable water and composite process water samples taken from ERGO Brakpan respectively. The analysis of potable water was undertaken as a benchmark. While a larger number of contaminants were identified in the ERGO Brakpan process water (refer to Appendix A, Table A.1 for complete ICP-MS analysis), focus was placed on contaminants present in the largest concentrations.

Table 4.21: potable water contaminant concentrations (benchmark concentrations)

potable water contaminant concentration								
Sample	Mn	Fe	Ni	Na	K	Mg	Ca	Sulfate
ID	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l
Sample 1	<0.004	<0.004	0.004	18.8	1.2	6.1	26.9	11.3
Sample 2	<0.004	<0.004	0.009	18.3	1.2	7.1	22.9	11.6
Average	<0.004	<0.004	0.006	18.6	1.2	6.6	24.9	11.5

As expected, the concentrations of contaminants within potable water were minimal and are within the South African National Standard (SANS 214:2015) Drinking Water Specifications (VinLAB (Pty) Ltd, 2016)

Table 4.22: Process water contaminant concentrations

Process Water initial concentrations								
Sample	Mn	Fe	Ni	Na	K	Mg	Ca	Sulfate
ID	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l
Sample 1	24.4	119	13.2	755	1201	194	1139	3075
Sample 2	24.7	151	13.3	770	1171	192	1131	3581
Average	24.5	135	13.3	762	1186	193	1135	3328

Comparing the atomic absorption results summarised in Table 4.22 to the initial ICP-MS analysis undertaken on ERGO Brakpan's process water (Table A.1, Appendix A), it is evident that the results are in close agreement. The average zinc concentration was measured to be less than 2 mg/l and thus was assumed negligible.

Tables 4.23 and 4.24 summarise the atomic absorption results from the various process water treatment techniques (lime and soda ash treatment) at respective treatment temperatures undertaken on the ERGO Brakpan process water.

Table 4.23: Contaminant concentrations' after respective treatment – heavy metals

Sample ID	Treatment	Manganese		Iron		Nickel		Sodium	
	Temperature (°C)	mg/l	% Removal	mg/l	% Removal	mg/l	% Removal	mg/l	% Removal
Process Water Head Sample		24.5		135		13.3		762	
Lime Softening	20 ± 0.5	0.02	99.9	0.17	99.8	0.12	99.1	776	-1.8
Lime - Soda Ash Softening	20 ± 0.5	0.02	99.9	0.10	99.9	0.15	98.9	920	-20.7
Lime Softening	60 ± 0.5	0.04	99.8	0.14	99.9	0.63	95.3	784	-2.8
Lime - Soda Ash Softening	60 ± 0.5	0.02	99.9	0.08	99.9	1.25	90.6	1102	-44.6
Lime Softening	90 ± 0.5	<0.01	100.0	0.11	99.9	0.89	93.3	783	-2.75
Lime - Soda Ash Softening	90 ± 0.5	<0,01	99.9	0.05	99.9	1.26	90.5	1036	-35.9

(Note: Negative values indicate an increase in contaminant concentration)

Table 4.24: Contaminant concentrations' after respective treatment - inorganics

Sample ID	Treatment	Magnesium		Calcium		Potassium		Sulfates	
	Temperature (°C)	mg/l	% Removal	mg/l	% Removal	mg/l	% Removal	mg/l	% Removal
Process Water Head Sample		193		1135		29.7		3328	
Lime Softening	20 ± 0.5	26.5	86.3	498	56.2	21.6	-6.6	2636	20.8
Lime - Soda Ash Softening	20 ± 0.5	1.3	99.8	565	50.2	31.4	-5.7	2795	16.0
Lime Softening	60 ± 0.5	0.18	99.9	871	23.3	36.0	-21.5	2807	15.7
Lime - Soda Ash Softening	60 ± 0.5	0.12	99.9	896	21.1	30.5	-2.9	3094	7.0
Lime Softening	90 ± 0.5	0.13	99.9	599	47.2	30.7	-3.6	1799	45.9
Lime - Soda Ash Softening	90 ± 0.5	0.10	99.9	697	38.5	31.2	-5.2	1989	40.2

(Note: Negative values indicate an increase in contaminant concentration)

The data shows that lime softening/treatment alone was sufficient to remove > 99 % of heavy metals (Mn, Fe and Ni) from ERGO Brakpan's process water. The apparent increase in sodium concentration, observed with lime softening alone, was attributed to impurities within reagents and experimental error since sodium concentration should not increase with lime dosing. Additionally, the apparent increase in potassium for all tests is similarly attributed to both experimental error and impurities within the reagents used, since potassium concentration should not be effected by both lime and soda ash dosing.

The data shows that at ambient conditions, the reduction of magnesium was not as effective. This result was expected due to the higher solubility of magnesium at low temperature, as well as no addition of soda ash to aid in the reduction of the magnesium concentration. The largest reduction in calcium concentration occurred in the lime-only treatment at ambient conditions.

It is evident that treatment of process water with a combination of lime and soda ash at ambient conditions was effective in the reduction of the magnesium concentration. However, a similar reduction in magnesium concentration (> 99 %) may be achieved with just lime treatment alone at 60 °C and 90 °C respectively. It is also evident that the reduction in nickel concentration is negatively affected by both soda ash addition as well as an increase in treatment temperature.

The tests where soda ash was added showed a significantly increased sodium concentration compared to the sodium concentrations reported for the lime-only treatment tests. The increased concentrations are attributed to the additional sodium from the soda ash and due to the mono-valence of the sodium ion, and chemical precipitation being ineffective in sodium reduction. Davidson (1974) did, however, conclude that sodium considerably aided in gold adsorption onto activated carbon. Slatter et al. (2009) demonstrated that while sodium ions increased silver loading, the presence of sodium had a negligible effect on gold loading. Since extensive research has been undertaken on the effects that sodium has on gold recovery, there was no merit in further investigation of the detrimental effects of sodium on gold recoveries.

From the analysis of both the XRD results and the atomic absorption results, it can be concluded that that lime softening/treatment on its own was sufficient to significantly reduce heavy metal concentrations. Additionally, a greater reduction in calcium concentrations was observed compared to a combination treatment using lime and soda ash. Furthermore, lime addition resulted in sulfate reduction due to the precipitation of gypsum.

4.4. FLOTATION AND LEACH RESULTS FOR GOLD RECOVERY USING TREATED PROCESS WATER

4.4.1. Leach results for gold recovery

The lime and soda ash softened process water was utilised to perform laboratory scale straight leach tests, with the aim of determining the viability of the treatment options for gold recovery. Table 4.25 summarises the average data from three repeated straight leach tests undertaken using the differently treated waters at different temperatures. Figure 4.12 gives a graphical representation of these results. The leach raw data is presented in Appendix E.

Table 4.25: Summarised leach results

Straight Leach Results				
Water Sample ID	**Head grade**	**Washed Residue**	**Gold Dissolution**	**Gold Accountability**
	(g/t)	**(g/t)**	**(%)**	**(%)**
Potable water	0.200	0.094	53.3	101.3
Process Water	0.200	0.105	49.7	101.9
Lime Softened Process Water – Ambient Temperature (20 ˚C)	0.200	0.094	52.0	102.1
Lime-Soda Ash Softened Process Water – Ambient Temperature	0.200	0.104	47.5	99.2
Lime Softened Process Water - 60°C	0.200	0.085	57.3	103.8
Lime-Soda Ash Softened Process Water - 60°C	0.200	0.111	44.7	104.2
Lime Softened Process Water - 90°C	0.200	0.107	46.7	99.3
Lime-Soda Ash Softened Process Water - 90°C	0.200	0.110	44.9	104.8

The data shows that there are significant differences in gold dissolution and thus gold recovery achieved, when using potable water and ERGO Brakpan's process water for the straight leaching of 4L50 reclaimed material. potable water produced a larger dissolution of gold from solids, which in turn resulted in a lower washed residue, indicating a greater recovery of gold from the feed material.

Treatment of ERGO Brakpan's process water with lime (lime softening only), produced a significantly lower washed residue than for treatment with a combination of lime and soda ash. Tests where only lime treatment was applied to ERGO Brakpan's process water, produced a higher gold dissolution than the corresponding lime-soda ash treatment at all temperatures tested.

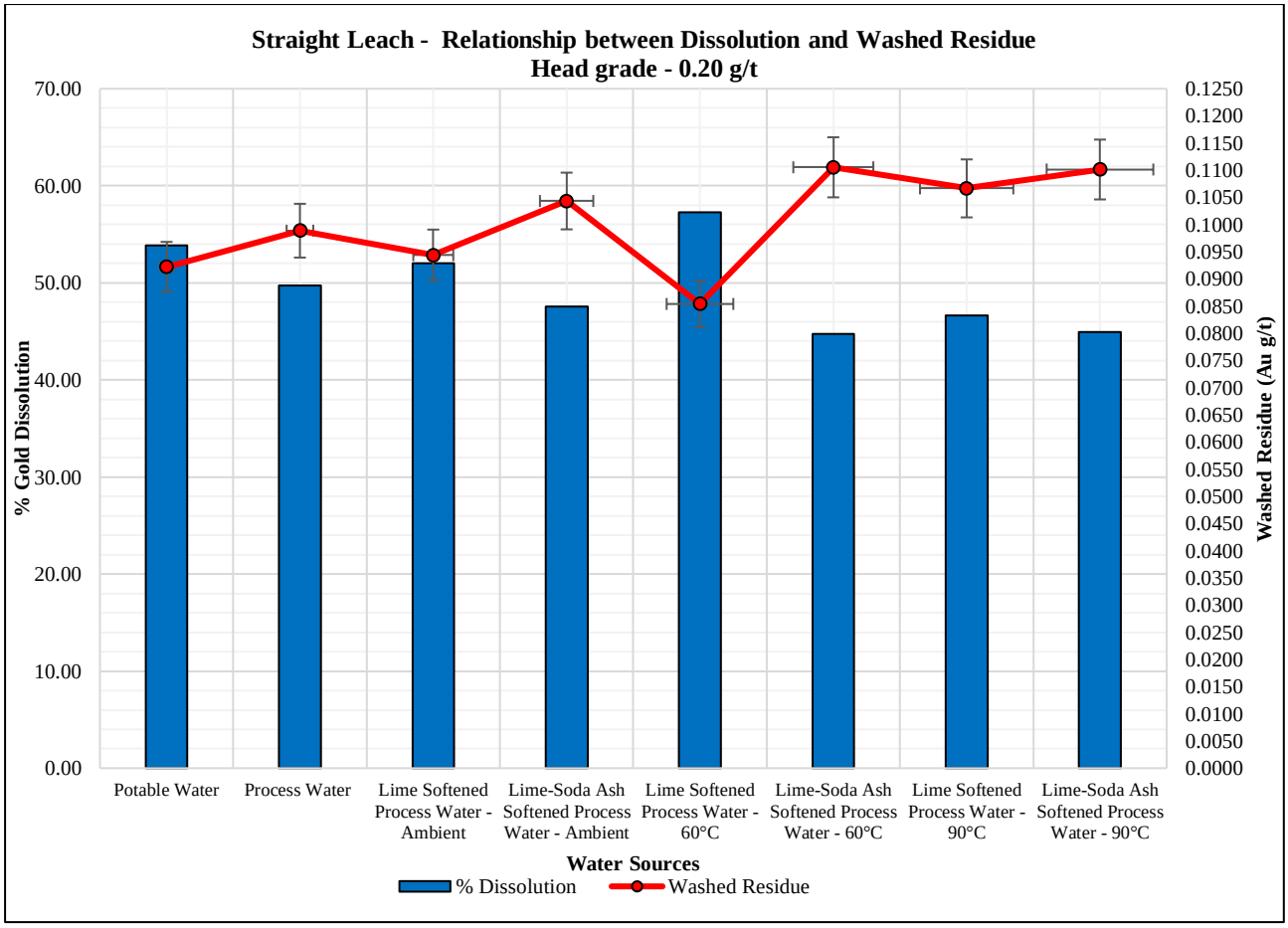

Figure 4.12: Graphical representation of straight leach results

The addition of soda ash as part of the proposed treatment of ERGO Brakpan's process water proved to be detrimental in gold recovery. Leach tests, using process water that had been treated with soda ash, produced the worst gold dissolution with the highest washed residue values. It is also noted that tests using soda ash treated process water performed more poorly than tests using ERGO Brakpan's untreated process water. This behaviour could be as a result of the carbonates from the soda ash addition reacting with calcium to produce $CaCO_3$, which is a known foulant of activated carbon which occupies active sites on the activated carbon, and thus reducing gold recovery. Additional to this, the XRD and AA analysis data (refer to Sections 4.3.3 and 4.3.4), confirm an increase in sodium concentration, when using process water treated with soda ash, which may have served to further compound the reduction of gold recovery.

It was also observed that when using process water that had been treated with lime for leaching, at both ambient and at 60 °C temperatures, lower washed residues and higher gold recoveries were achieved than when using the current untreated process water at ERGO Brakpan. This result is significant, indicating that lime treatment of process water may be sufficient to improve gold recoveries at ERGO Brakpan. At ambient conditions, lime treated water produced a 1.87 % improvement in gold dissolution/recovery compared to potable water.

Process water treated with lime at 60 °C provided the most significant improvement in gold recovery, producing a gold dissolution from solids of 57.3 % and a washed residue of 0.085 g/t, compared to reported potable water's reported gold dissolution of 53.3 % and washed residue of 0.094 g/t . The additional gold recovery when using lime treated process water at 60 °C was possibly attributed to free and residual cyanide that was available within the process water that aided in the leaching process, as well as improved leaching kinetics. A free cyanide value of approximately 50 mg/l was measured in ERGO Brakpan's process water. The major drawback is the excessive energy requirement. Heating 60 Ml of process water from ambient (20°) to 60 °C will required approximately 2790 MWh of energy on a daily basis and thus will not be economically viable (refer to Appendix F, Table F.4 for the energy balance).

4.4.2. Flotation and tails leach results for gold recovery

The lime and soda ash softened process water was used to perform lab scale leach flotation tests to assess the viability of the treatment options for gold recovery. A sulfide ore reclaimed from ERGO Brakpan's 4L50 reclamation site was utilised. Table 4.26 summarises the averaged data achieved from the three repeated flotation leach test work undertaken using the different water sources. Figures 4.13 and 4.14 give a graphical representation of these results.

Table 4.26: Summarised flotation and tails leach results

Flotation and tails leach results						
Water Sample ID	Head grade	Washed Residue	Gold Dissolution	Concentrated Gold	Mass Percentage as Concentrate	Gold Account -ability
	(g/t)	(g/t)	(%)	(g/t)	Mass %	(%)
Potable water	0.155	0.084	45.9	3.5	1.3	102.7
Process Water	0.187	0.111	40.5	1.8	1.4	105.4
Lime Softened Process Water – Ambient	0.150	0.096	36.3	2.8	1.4	102.9
Lime-Soda Ash Softened Process Water – Ambient	0.157	0.102	34.5	2.7	1.4	101.1
Lime Softened Process Water - 60°C	0.150	0.077	48.6	2.9	1.4	101.8
Lime-Soda Ash Softened Process Water - 60°C	0.150	0.113	24.4	2.9	1.5	104.1
Lime Softened Process Water - 90°C	0.170	0.090	47.1	2.2	1.3	103.8
Lime-Soda Ash Softened Process Water - 90°C	0.180	0.100	44.4	1.2	1.4	101.6

Significant differences in gold dissolution and thus gold recoveries were apparent when using potable water and ERGO Brakpan's process water for leaching of 4L50 reclaimed material flotation tails. potable water produced a larger dissolution of gold from solids resulting in a lower washed residue, which results in a greater recovery of gold from the feed material.

Results showed that ERGO Brakpan's process water that had only been treated with lime achieved significantly lower washed residue concentrations after leaching than ERGO Brakpan's process water that had been treated with a combination of lime and soda ash. This also held true for all temperatures tested. Figures 4.13 and 4.14 should be interpreted together to understand the relationship between the flotation concentrate grade and the head grade of the tails. All samples had an initial head grade of 0.200 g/t prior to flotation. An increase in the flotation concentrate grade resulted in a lower head grade of the tails, owing to gold enclosed within sulfides which were predominantly of a higher gold grade, which reported to the flotation concentrate. The flotation concentrate undergoes ultra-fine grinding prior to reporting to a high grade CIL-CIP circuit at ERGO Brakpan.

The data also showed that the addition of soda ash as part of the proposed treatment of ERGO Brakpan's process water may prove detrimental to gold recovery. Flotation tails leach tests which utilised process water treated with soda ash, produced the lowest gold dissolutions, as well as the highest washed residues. These tests also performed worse than ERGO Brakpan's current process water, with the ambient treatment and 60 °C treatment producing lower gold recoveries. It was postulated that soda ash addition (and thus carbonate addition) reacts with the calcium to produce $CaCO_3$ which is a known foulant on activated carbon that occupies active sites thus reducing gold recovery.

The data shows that process water treated with lime only at ambient conditions produced a lower gold dissolution of 36.3 %, compared to the untreated process water gold dissolution of 40.5 %. It was also observed that the head grade of the flotation tails was 0.187 g/t and 0.150 g/t for untreated process water and lime treated process water at ambient conditions respectively. This indicated that lime treated process water at ambient conditions produced a more efficient flotation recovery of higher-grade encapsulated gold. It was further observed that process water treated with lime at ambient conditions produced a significantly lower washed residue of 0.096 g/t compared to untreated process water washed residue of 0.111 g/t. Since the initial head grade of all the samples was 0.200 g/t prior to flotation, the data showed a higher overall recovery of gold with lime treated process water at ambient conditions compared to untreated process water.

Process water treated with lime at 60 °C produced a higher gold dissolution and a lower washed residue compared to that achieved using potable water. The additional gold recovery was attributed to free cyanide that was present within the process water and aided in the leach process. The concentrate grade achieved when using the water treated with lime alone at 60 °C was however not as high as that achieved when using potable water. This indicated that, while the leach of the flotation tails was efficient, the

flotation tests were not as efficient as the tests utilising potable water. This could be attributed to the sensitivity of the flotation reagents, which were sulfate selective and the increased concentrations of residual free cyanide that is a known suppressant in sulfide mineral flotation. Potable water was expected to perform better due to minimal sulfate and free cyanide concentrations present in this water. Sulfate reduction was not effectively achieved in the treated process water. Results did however confirm that sulfate concentration had an appreciable effect on gold recoveries during leach tests as shown in Section 4.2.1.

Process water treated with lime at 90 °C produced lower washed residues and higher gold recoveries than the process water utilised at ERGO Brakpan. These results proved that lime treatment alone on ERGO Brakpan's process water was sufficient to improve gold recoveries at ERGO Brakpan. At 90 °C, lime treated water produced only a 1.2 % difference in gold dissolution when compared to potable water. However, due to the high energy requirement as well as associated costs, lime treatment at 90 °C, is not considered a viable economic and practical solution.

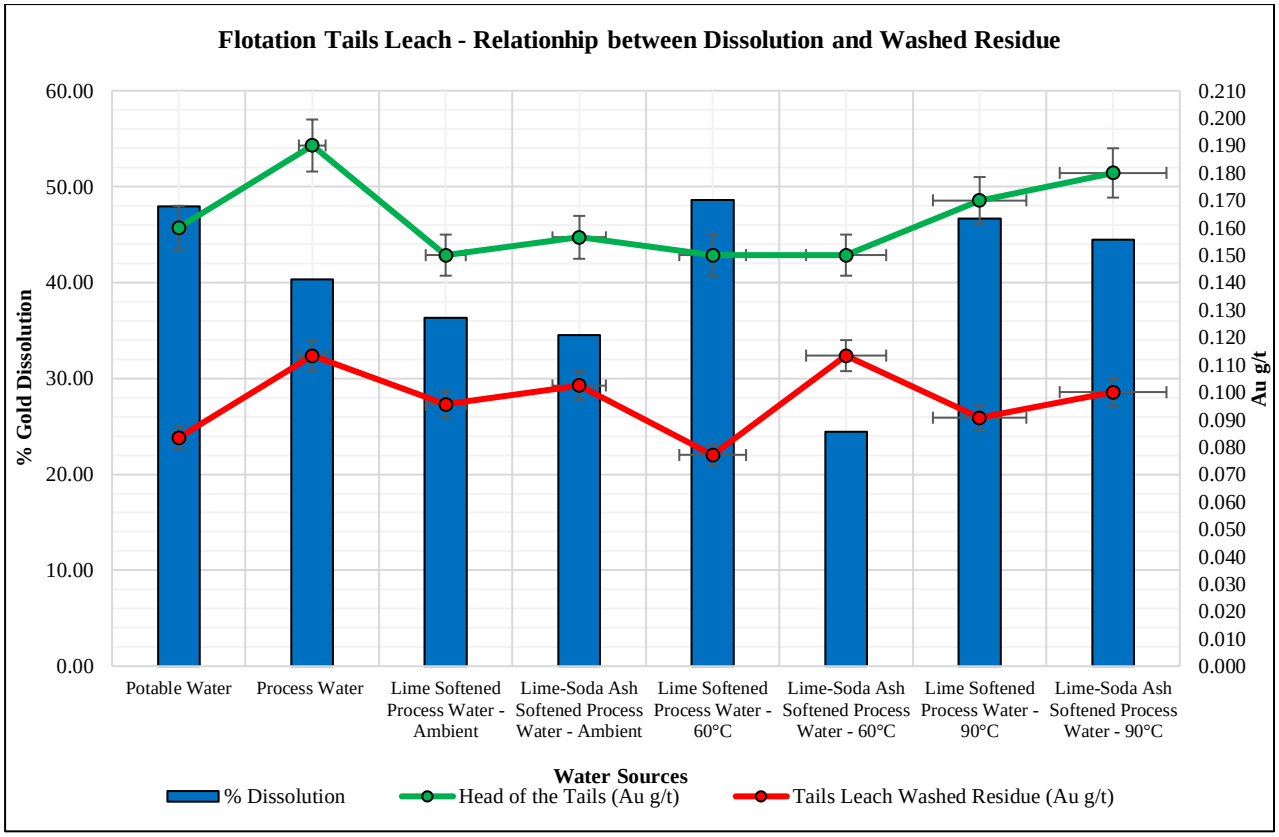

Figure 4.13: Graphical representation of flotation and tails leach results

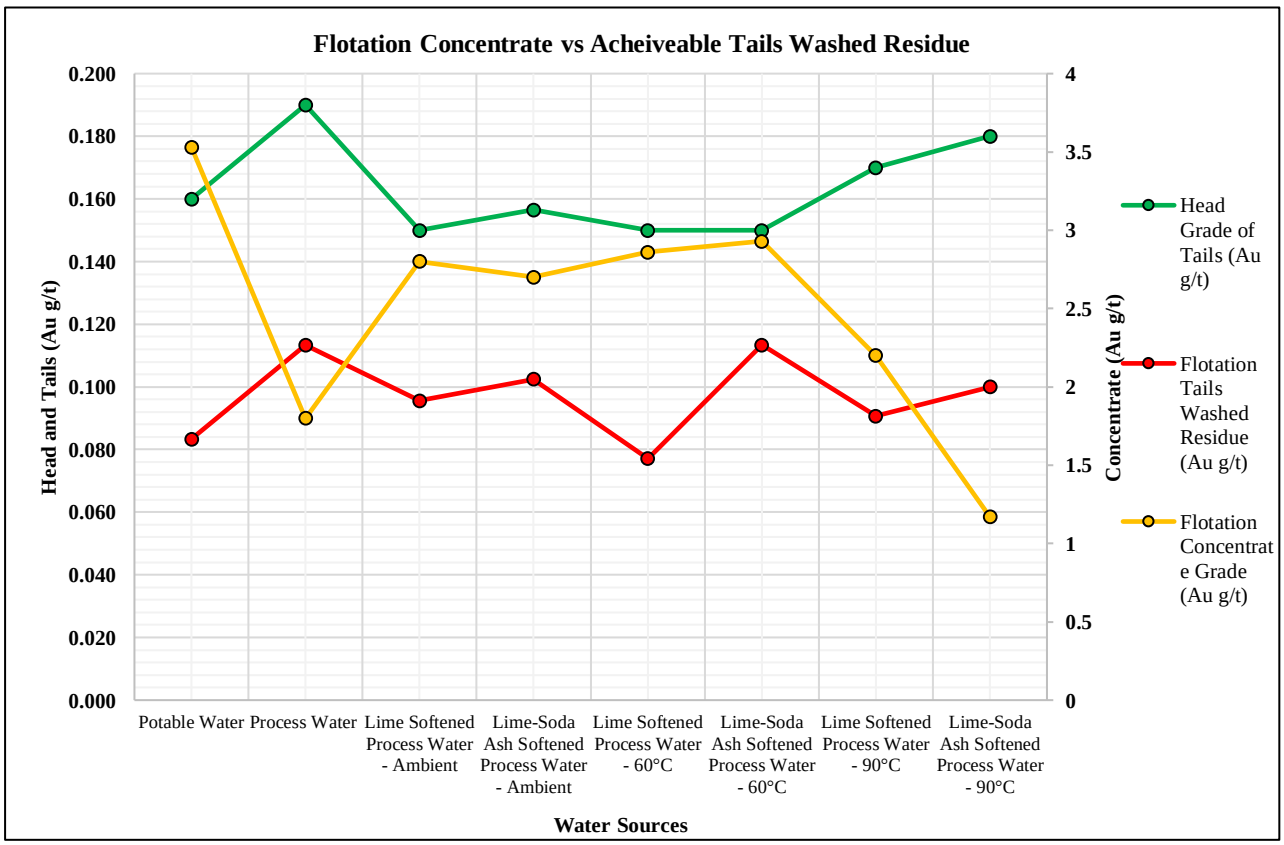

Figure 4.14: Graphical representation of flotation concentrate vs achievable tails washed residue

4.5. SUMMARY OF RESULTS

The main findings of this chapter are summarised below.

A statistical analysis had indicated that an average improvement of ~10 % in gold recovery can be achieved when using potable water over process water, during straight leach tests. Similarly, an average improvement of ~5 % in gold recovery was observed for flotation tails leach tests, with the use of potable water.

The presence of heavy metals such as Fe, Ni and Zn were observed to have the greatest negative effect on gold recovery. It was proven that the zinc cyanide complex negatively effects gold recovery only to a small degree and that the presence of both nickel and iron had a far greater detrimental effect on gold recovery.

The presence of sulfates had shown a possible passivation effect which negatively affected gold recovery, this was however not to the same degree as the negative effects of the heavy metals. The presence of calcium had shown an increase in gold recovery of ~8.35 % between potable water spiked with 50 mg/l and 500 mg/l of calcium, respectively. This was attributed to the formation of a calcium aurocyanide complex as the gold recovery increased with an increased calcium concentration. The presence of magnesium had similarly shown an increase in gold recovery of ~2.5 % between potable water spiked with 50 mg/l and 500 mg/l of magnesium, respectively.

Based on the straight leach tests result, it was evident that lime softening at ambient conditions produced a washed residue similar to potable water. At ambient conditions, the washed residues obtained were 0.0940 g/t and 0.0944 g/t for potable water and lime treated process water respectively. Due to the negligible difference in the gold recoveries (approximately 1.3 %), between the lime treated process water and potable water, the use of lime was an attractive beneficiation option. While straight leach test results indicated that lime softening process water at 60 °C produced a higher gold recovery of 57.3 % and a washed residue of 0.085 g/t , heating 60 Ml of process water from 20 °C to 60 °C requires approximately 2790 MWh on a daily basis of energy and thus will not be economically viable.

The data also showed that the addition of soda ash as part of the proposed treatment of ERGO B's process water may prove detrimental to gold recovery. Flotation tails leach tests which utilised process water treated with soda ash, produced the lowest gold dissolutions, as well as the highest washed residues.

Leaching of flotation tails showed that process water treated with lime at 60 °C and 90 °C respectively, produced lower washed residues and higher gold recoveries than the untreated process water utilised at ERGO Brakpan. These results proved that lime treatment alone on ERGO Brakpan's process water was sufficient to improve gold recoveries at ERGO Brakpan. However, due to the high energy requirement

and associated costs, lime treatment at 60 °C and 90 °C, was not considered a viable economic and practical solution.

5. CHAPTER 5: ECONOMIC EVALUATION

5.1. PROPOSED WATER TREATMENT TECHNIQUE

The potential full-scale installation of a lime treatment process followed by gypsum crystallization was investigated, based on the results obtained in Chapter 4. Van Niekerk (2002) had undertaken a full scale evalution of the SAVMIN process treating 5 Ml/day of acid mine drainage. Since the lime treatment and gypsum crystallization form part of the SAVMIN process, the economic evaluation presented in this study closely followed the approach presented by van Niekerk (2002), with costing estimates based on his work. The 4L50 reclaimed ore utilised in the tests in Chapter 4 accounts for approximately 32 % of ERGO Brakpan's daily feed; thus, the current econmic study investiagated the full-scale installation to treat 20 Ml/day process water, which accounts for approximately 32 % of ERGO Brakpan's daily consumption. This would ensure consistency since the economic evaluation (i.e. internal rate of return, Net Present Value and payback) would not be severely penalised by costing a plant for 60 Ml/day capacity, but only achieving the gold income from 32 % of the daily ore throughput. Furthermore, the flotation circuit was shut down at ERGO Brakpan due to inefficient gold recoveries and changes in the mineralogy of the feed material. Thus, the flotation circuit is bypassed, and only straight leaching is currently performed. Consequently, the economic evaluation undertaken in this chapter was based on the results obtained from straight leach tests only to accurately represent the current ERGO Brakpan operation.

5.1.1. Generalised proposed process description

Tables 4.19 and 4.20, show that ERGO Brakpan's process water is acidic in nature. The process water will therefore be fed into three 6000 m^3 high shear neutralisation reactors in which the associated metal precipitation as well as formation of excess gypsum ($CaSO_4 \cdot 2H_2O$) will be undertaken. Since lime addition to increase leach pH is general practice in most gold leach plants, it is easily accessible and therefore a viable economical beneficiation option. Consequently, slaked lime will be the preferred neutralisation alkali that will be dosed into the neutralisation reactor to increase the pH to above pH 11.5 to ensure efficient magnesium removal. The precipitated metals and excess gypsum will be removed by a thickener/settler. Sufficient slaked lime addition will ensure neutralisation of free and mineral acidity and will precipitate heavy metals as hydroxides. Sufficient contact time will be allowed to stimulate precipitation of excess gypsum, thus reducing sulfate concentration in the gypsum crystallisation reactor. The gypsum precipitation will be further catalysed by the recycle of a small portion of settled solids and gypsum sludge from the neutralisation thickener Figure 5.1 illustrates the block flow diagram for the proposed lime treatment and gypsum crystallization process.

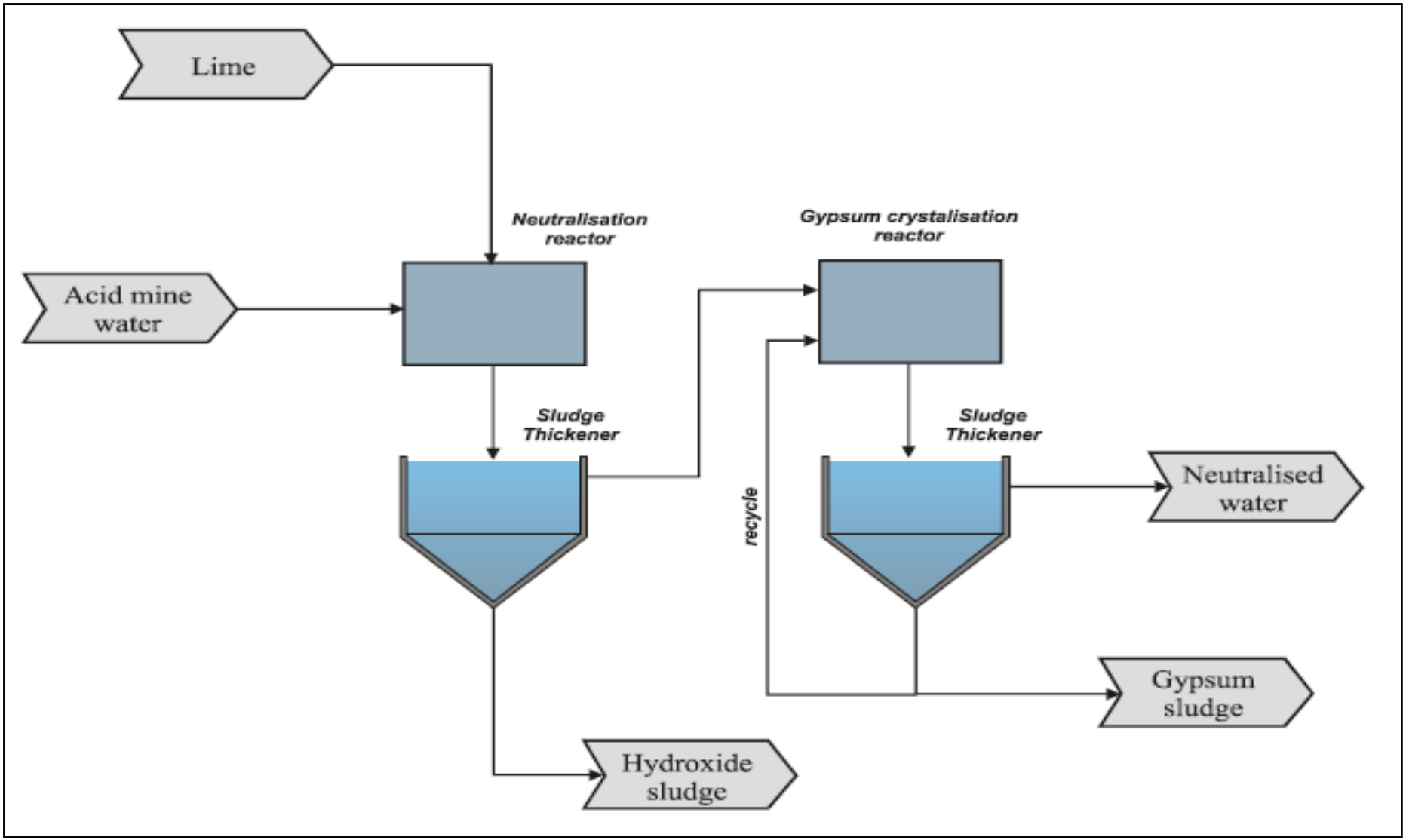

Figure 5.1: Block flow diagram for the proposed lime treatment and gypsum crystallization process (van Niekerk, 2002)

5.1.2. Process Reactions

ERGO Brakpan's process water, at an approximate pH value of 5, will be introduced into the neutralisation reactor and contacted with hydrated lime ($Ca(OH)_2$) to raise the pH to above pH 11.5. At these conditions, the heavy metals will precipitate as metal hydroxides (refer to reaction 5.1). Although most metal hydroxides will precipitate at relatively lower pH values (pH<8), a high pH is needed for the precipitation of magnesium (refer to Equation 5.2).

$$Me^{2+} + Ca(OH)_2 \rightarrow Me(OH)_2\,(s) + Ca^{2+} \quad (5.1)$$

$$Mg^{2+} + Ca(OH)_2 \rightarrow Mg(OH)_2\,(s) + Ca^{2+} \quad (5.2)$$

Where "Me" refers to divalent heavy metals such as iron, nickel, copper, zinc etc.

The resultant supernatant solution will thereafter be contacted with gypsum crystals. The addition of lime may cause the solution phase to become supersaturated with gypsum, due to the high concentration of sulfates in the feed water. The solution will be contacted with gypsum crystals in order to provide active surfaces of gypsum, which acts as a catalyst for the precipitation of the 'supersaturated' gypsum (Damons, 2001). The formation of gypsum is described by the following stoichiometric reaction:

$$Ca^{2+} + SO_4^{2-} + H_2O \rightarrow CaSO_4 \cdot 2H_2O \quad (5.3)$$

5.1.3. Process constraints

5.1.3.1. Contact time

Chemical precipitation reactions are relatively rapid, provided that optimum process conditions are maintained, such as pH maintained above 11.5 (van Niekerk, 2002). A retention time of 30 minutes will be observed after lime addition to ensure neutralisation and optimal metal precipitation. Furthermore, 30 minutes of contact time will be allowed to stimulate precipitation of excess gypsum, thus reducing the sulfate concentration. These contact times are relatively short compared to biological process requirements.

5.1.3.2. Gypsum crystallisation

The efficient and effective gypsum crystallisation is vital for two main reasons:

- Incomplete gypsum crystallisation will result in ineffective sulfate reduction resulting in excess sulfates circulating through ERGO Brakpan's leach train.
- Incomplete gypsum crystallisation will result in a high scaling potential in downstream processes. Furthermore, scale formation will require additional maintenance resources.

5.1.4. Process conditions

From their pilot plant test work, van Niekerk (2002) confirmed that the different processes of neutralisation, metals removal and gypsum crystallisation could be combined into a single reactor and thickener. Furthermore, van Niekerk (2002) confirmed that neutralisation and heavy metal removal of acid mine water was achieved by high lime treatment. According to the required unslaked lime dosage calculation procedure outlined by Minnesota Rural Water Association (2009), as well as the lime softening test work outlined in Chapter 4, the average unslaked lime consumption to raise the pH to above pH 11.0 was 756 mg/l. Accounting for the molar ratio between unslaked and slaked lime, the required slaked lime dosage is:

$Ca(OH)_2$ = 1000 mg/l
= 1.0 kg/m^3 mine water

Gypsum crystallisation will be catalysed by recycling the thickener underflow back to the neutralisation reactor to maintain an inventory of gypsum crystals. From the pilot plant study undertaken by van Niekerk (2002), as well from lime softening test work outlined in Chapter 4, the average lime gypsum sludge product is estimated as:

$CaSO_4 \cdot 2H_2O$ = 0.82 kg/m^3 mine water

5.1.5. Process controls

The treatment process control will be relatively simple and is based on manipulation of the following general process variables:

- Chemical dosages
- Recycle rates

The neutralisation process will require tight control for addition of the slaked/hydrated lime slurry. A pH sensor will monitor the neutralisation reactor and will control the lime slurry addition rate. The target pH will be typically within the range of pH 11.0 - 11.5 to ensure magnesium precipitation.

The gypsum crystallisation process requires the maintenance of a minimum solids inventory within the neutralisation reactor. This will be achieved by the thickener underflow recycle back to the reactor. Furthermore, van Niekerk (2002) states that a recycle target solids content of 5 – 10 % is typically applied. The excess gypsum sludge will report as waste to the thickener underflow. The waste solids underflow rate will be controlled from a relative density monitor as well as a feedback control loop.

5.1.6. Treatment residue

The lime treatment and gypsum crystallisation processes generate various residue streams which include:

- Metal hydroxide sludge from neutralisation
- Gypsum sludge from neutralisation
- Lime sludge from slaked lime make up

All residue streams will be thickened by their respective thickeners and pumped to ERGO Brakpan's tailings facility. No further treatment of these streams is proposed, however future analysis is proposed should the streams have value in them.

5.1.7. Lime make – up and dosing

Unslaked lime is readily available at ERGO Brakpan and is stored in bulk silos. A screw feeder will control the lime addition into the slaker. The slaked lime slurry will thereafter be treated in a lime thickener where the thickened lime slurry (approximate SG = 1.45) will be utilised for neutralisation, metal hydroxide precipitation and excess gypsum crystallisation in the high shear neutralisation reactor. The dilute milk of lime (lime thickener overflow) will be utilised in ERGO Brakpan's leach train. The lime storage, make- up and dosing process is presented as Figure G.1 in Appendix G.

The general process flow diagram is presented as Figure G.2 in Appendix G. The proposed plant layout is presented as Figure G.3 in Appendix G.

5.2.PROCESS DESIGN CRITERIA (PDC)

The process design criteria for the installation of a full-scale lime treatment process followed by gypsum crystallization is summarised in Table 5.1. Refer to Appendix F for the design calculations.

Table 5.1: Process design criteria

Process Design Criteria	Units	Design Conditions
Site Data		
Location		DRDGOLD ERGO Brakpan Site
Country		South Africa
Site Co-ordinates	Latitude/longitude	-26.28, 28.37
Altitude above sea level	metres	1627
Atmospheric Pressure	atm	Atmospheric
Topography		Undulating
Summer mean daily max temperature	°C	39
Winter mean daily min temperature	°C	12
Prevailing wind direction		Northerly to North-westerly
Annual average rainfall	mm	697
Rainfall period	Spring/Summer	October to April
Max monthly rainfall	mm	154
On Site Services		
Potable Water	On Site	Available
Power	Voltage	525
Neutralisation Reactor		
Number of Neutralisation Reactor(s)		3
Average working days per month	Days	30
Total Volume of Process Water for Treatment	Ml/day	20
	m^3/s	0.23
Retention Time	minutes	30
Volume of each Neutralisation Reactor	m^3	140
Mixing Intensity	W/m^3	70
Lime Dosing Rate	kg/m^3 mine water	1.0
	kg/s	0.23
	ton/month	600
Operating Temperature	°C	25

Process Design Criteria	Units	Design Conditions
Neutralisation Settler/Thickener		
Number of Thickener(s)		1
Feed Volumetric Flowrate	m^3/s	0.23
Up flow Rate	m/hr	0.70
Solids Loading Rate	kg/m^2hr	0.57
Retention Time	minutes	30
Volume of Neutralisation Settler	m^3	417
Thickener Underflow Density	SG	1.45
Residue Disposal		
Estimated Production of Residue Streams:		
Metal Hydroxide/Gypsum	ton/day	16.4
	kg/m^3 mine water	0.82

5.3.ECONOMIC ANALYSIS

5.3.1. Capital expenditure (CAPEX) estimate

The capital expenditure (CAPEX) for a full-scale installation of a lime treatment and gypsum crystallization treatment facility was estimated from the economic evaluation undertaken by van Niekerk (2002) for the SAVMIN process treating 5 Ml/day of acid mine drainage. The current study investigated the full-scale installation for treatment of 32% of ERGO Brakpan's daily consumption of 20 Ml/day of process water.

Tribe & Alpine (1986) state that an increase in capactiy of equipment is greater than the increase in equpiment cost, in a form that is analgous to surface area and volume. In practice, this is refered to as economies of scale.Therefore, the significance of economies of scale is its potential use in the projection of equipment costs for larger scales of plants during investment planning (Tribe & Alpine, 1986). The rule is obtained from consideration of two types of equipment, tanks and pipes.

The relationship between surface area (S) and volume (V) is given by Equations 5.4 and 5.5 for a tank and pipeline respectively, leading to the generalised overall equipment cost Equation 5.6 presented below (Tribe & Alpine, 1986):

$$\frac{S_1}{S_2} = \left(\frac{V_1}{V_2}\right)^{0.667} \tag{5.4}$$

$$\frac{S_1}{S_2} = \left(\frac{V_1}{V_2}\right)^{0.5} \tag{5.5}$$

$$\frac{C_1}{C_2} = \left(\frac{V_1}{V_2}\right)^{\alpha} \tag{5.6}$$

Where C refers to equipment cost and α denotes the scale coefficient. A value of α less than unity implies increasing returns to scale. The value of α = 0.6 is often used as a rule of thumb to obtain the investment cost of a capacity level V_2 given the cost C_1 associated with the level of capacity V_1 (Tribe & Alpine, 1986)..

The method of economies of scale was applied to the pilot plant economic evaluation undertaken by van Niekerk (2002) to estimate the CAPEX of a 20 Ml/day lime and gypsum treatment facility focusing on the following disciplines:

- Mechanical Equipment
- Civil Engineering
- Structural Steel
- Piping & Valves
- Electrical & Instrumentation
- Installation
- Transport
- Commissioning

The CAPEX estimate study is summarised below. A 5 % working capital allocation was applied on the total CAPEX for unaccounted costing contingencies. The CAPEX is estimated for a plant that is expected run for 10 years. An average inflation of 6% per annum (Sinnot, 2005) was applied to the economic evaluation undertaken by van Niekerk (2002) to estimate the CAPEX for the current study.

Table 5.2: The CAPEX for a full-scale lime treatment and gypsum crystallization facility

Lime treatment and gypsum crystallization facility – 20 Ml/day throughput						
Description	Mechanical	Civil	Electrical	Instruments	Other	**Total**
Lime storage and make -up	R13 054 579	R6 522 244				R19 576 823
Neutralisation reactor	R929 100	R6 266 302				R7 195 402
Neutralisation thickener	R929 100	R6 266 302				R7 195 402
Interconnecting pipework		R3 934 530				R3 934 530
Electrical Equipment			R3 825 235			R3 825 235
Instrumentation and Controls				R2 583 130		R2 583 130
Transportation Costs					R1 642 666	R1 642 666
Commissioning					R795 203	R795 203
Working Capital					R2 337 419	R2 337 419
Total	R14 912 779	R22 989 378	R3 825 235	R2 583 130	R4 775 288	**R49 085 809**

5.3.2. Operating expenditure (OPEX)

The operating and maintenance costs were calculated to include: chemical consumption, electrical power consumption and the required operating personnel and labour. The unit prices for chemicals and power was provided by ERGO Brakpan and reflects the price indices as of May 2020. The operational expenditure was estimated for the treatment of 20 Ml/day and was based on an average of 30 operational days per month. The number of required equipment and machinery was scaled up from 5 Ml/day undertaken by van Niekerk (2002) to 20 Ml/day for the current OPEX study. The power requirement for each major mechanical equipment was estimated and provided by ERGO Brakpan's senior electrical foreman and the estimated labour costs was provided by ERGO Brakpan's Human Resource Department. The total OPEX costs are summarised in Table 5.3.

Table 5.3: The OPEX for a 20 Ml.day full-scale lime treatment and gypsum crystallization treatment facility

Monthly operating costs				
Reagent costs				
Chemical type	Dosage (kg/m^3 process water)	Unit Cost (R/kg)	Quantity (kg)	Total
Slaked Lime	1,0	R1,84	6000000	R1 104 000
Total Monthly Reagent Costs				**R1 104 000**
Electrical Power Costs				
Equipment Item	Number installed	Unit Cost (R/kWh)	Monthly Unit Power Requirement (kWh)	Total
Process Water Feed Pumps	6	R1,00	12240	R73 440
Neutralisation Reactor Mixers	6	R1,00	7920	R47 520
Neutralisation Settler Bridge	1	R1,00	720	R720
Hydroxide/Gypsum Recycle Pumps	4	R1,00	12240	R48 960
Lime Screw Feeder	1	R1,00	1440	R1 440
Lime Slacker Tank Mixer	1	R1,00	720	R720
Lime Settler Bridge	1	R1,00	720	R720
Lime Recycle Pumps	2	R1,00	2664	R5 328
Milk of Lime Dosing Pumps	2	R1,00	10080	R20 160
Total Monthly Power Costs				**R199 008**
Labour Costs				
Personnel Category	Number Required	Unit Cost (R/hr)	Monthly Working Hours	Total
Process Engineer	1	225	160	R36 000
Process Assistants	3	110	240	R79 200
Process Operators	4	55	240	R52 800
Labourers	6	35	240	R50 400
Total Monthly Labour Costs				**R218 400**
Total OPEX Costs			**R1 512 408**	

5.3.3. Maintenance costs

Planned maintenance and associated repair work are costed as a fraction of the capital investment into different components of the plant (van Niekerk, 2002). The estimated annual maintenance cost is summarised in Table 5.4. Due to the nature of ERGO Brakpan's operations, partial plant shut down for planned maintenance is done 1 day in a 90-day cycle and is planned for a month containing 31 days.

Table 5.4: Maintenance costs for a full-scale installation of a lime treatment followed by gypsum crystallization treatment facility

Cost Component	Capital Expenditure	Annual allowance for maintenance (%)	Annual Maintenance Cost (R/year)	Monthly Maintenance Costs (R/month)
Building and Civil Works	R22 989 378	0,5	R114 947	R9 579
Mechanical Maintenance	R14 912 779	2	R298 256	R24 855
Electrical	R3 825 235	3	R114 757	R9 563
Instrumentation and other	R7 358 418	3	R220 753	R18 396
Total Maintenance Costs			**R748 712**	**R62 393**

5.3.4. Estimated revenue

Based on the straight leach tests results summarised in Section 4.4, the potential monthly revenue of the project was based on the difference in the attainable washed residues between the process water, which produced an average washed residue of 0.105 g/t, and lime treated process water at ambient conditions, which produced an average washed residue of 0.094 g/t (refer to Table 4.25).

In April 2020, at the time at which the economic evaluation was undertaken for the current study, the gold price was above R1 000 000 per kg (Gold Price, 2020). A conservative gold price of R950 000 per kg was utilised for the estimated revenue of the project, summarised in Table 5.5 below. An estimated revenue of approximately R6 270 000/month may be realised from processing of the 4L50 reclaimed ore with lime softened process water, which accounts for approximately 32 % of ERGO Brakpan's daily feed.

Table 5.5: Estimated project revenue

Revenue		
		Units
Washed Residue with Process Water	0,105	g/t
Washed Residue with Lime Softened Water (Ambient)	0,094	g/t
Potential Additional Gold Recovery	0,011	g/t
4L50 - Reclamation Site Daily throughput	20 000	t/day
Average Working Days per month	30	days
Gold Price per kg	R950 000	R/kg
Potential Additional Gold Recovery	220	g/day
	6,60	kg/month
Estimated Monthly Additional Revenue	**R6 270 000**	R/month

5.3.5. Discounted Cash Flow analysis

A discounted cash flow (DFC) analysis was undertaken to determine the net present value (NPV) of the project. A discounted cash flow is a valuation method used to estimate the value of an investment or project based on its future cash flows and attempts to determine the value of an investment or project today, adjusted for the time value of money (Chen, 2020). The net present value is the difference between the present value of cash inflows and the present value of cash outflows over a period of time (Chen, 2020), and is calculated using Equation 5.7. Chen (2020) further states that a positive net present value indicates that the projected earnings generated by the project or investment exceed the anticipated costs. It is therefore assumed that a positive net present value will show a profitable project while a negative will result in a net loss (Chen, 2020).

$$NPV = \sum_{t=1}^{n} \frac{R_t}{(1+i)^t} \tag{5.7}$$

where:

Rt = Net cash inflow-outflows during a single period t

i = Discount rate

t = Number of timer periods

The discounted cash flow analysis determines the present value of expected future cash flows by using a discount rate as illustrated in Equation 5.7. The discount rate is essentially the interest rate used to determine the present value by predicting future returns. For the current study, a yearly discount rate of 10 % was chosen, which corresponds to the average long-term cost of funds for ERGO Brakpan from an equity.

From the discounted cash flow analysis, the internal rate of return (IRR) is also calculated. The internal rate of return is the discount rate that makes the net present value of a particular project equal to zero (Hayes, 2020), and is calculated by equating Equation 5.7 to zero. Hayes (2020) states that the internal rate of return of a project must exceed the chosen discount rate to achieve the expected return on investment. Hayes (2020) further states that the higher a project's rate of return, the more desirable it is to undertake the project.

The discounted cash flow analysis for the full-scale lime treatment and gypsum crystallization facility is presented in Table 5.6. The average yearly escalations were provided by ERGO Brakpan with the analysis undertaken over the lifetime of the project, namely, 10 years. Additionally, an average increase in the Rand gold price of 10 % has been observed (Gold Price, 2020) over the last five years and has therefore been the chosen annual income escalation rate for the project evaluation.

Table 5.6: Discounted Cash Flow for a lime treatment and gypsum crystallisation plant

Discounted Cash Flow for proposed lime and gypsum crystallisation plant - (R'million)												
Income/Expenditure	**Yearly Escalations**	Years										
	%	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**
Capital Investment		46,75										
Working Capital		2,34	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00
Income from Sales	**10**		75,24	82,76	91,04	100,14	110,16	121,17	133,29	146,62	161,28	177,41
Variable Costs												
Lime Consumption	6,8		13,25	14,15	15,11	16,14	17,24	18,41	19,66	21,00	22,42	23,95
Utilities	5		2,39	2,51	2,63	2,76	2,90	3,05	3,20	3,36	3,53	3,70
Fixed costs												
Maintenance	10		0,75	0,82	0,91	1,00	1,10	1,21	1,33	1,46	1,60	1,77
Operating Labour	8		2,62	2,83	3,06	3,30	3,57	3,85	4,16	4,49	4,85	5,24
Total Costs			19,01	20,31	21,71	23,20	24,80	26,51	28,35	30,31	32,41	34,66
Net Income before Tax			56,23	62,45	69,33	76,94	85,36	94,66	104,95	116,31	128,88	142,75
Income Tax	**28**		15,75	17,49	19,41	21,54	23,90	26,51	29,39	32,57	36,09	39,97
Cash Flow		-49,09	40,49	44,97	49,92	55,40	61,46	68,16	75,56	83,75	92,79	102,78
Discount Rate (%)	**10**											
Periodic Rate		1,00	0,91	0,83	0,75	0,68	0,62	0,56	0,51	0,47	0,42	0,39
Present Value		-49,09	36,81	37,16	37,51	37,84	38,16	38,47	38,78	39,07	39,35	39,63
Net Present Value (NPV)		333,68										
Internal Rate of Return (%)	**76**											
Payback Period (Years)	**2**											

From the discounted cash flow, it is evident that a net present value of approximately R 333 000 000 may be achieved over the lifetime of the project. A positive net present value further indicates that the investment costs to implement the proposed project are justifiable as the project will generate a positive discounted cash flow above the initial cost. Furthermore, the internal rate of return obtained of 76 %, is significantly higher than the chosen discount rate of 10 % clearly indicating that the project is profitable.

A sensitivity analysis was undertaken to determine the effect on the internal rate of return, should major unaccounted changes occur to the cost drivers. The results are summarised in Table 5.7 and are presented graphically in Figure 5.2.

Table 5.7: Sensitivity of internal rate of return for capital, income and costs changes

Cost Driver	Internal Rate of Return (IRR)		
	+5 %	0 %	-5 %
Capital Investment	72%	76%	79%
Negative Income	71%	76%	81%
Total Cost of Sales	74%	76%	77%

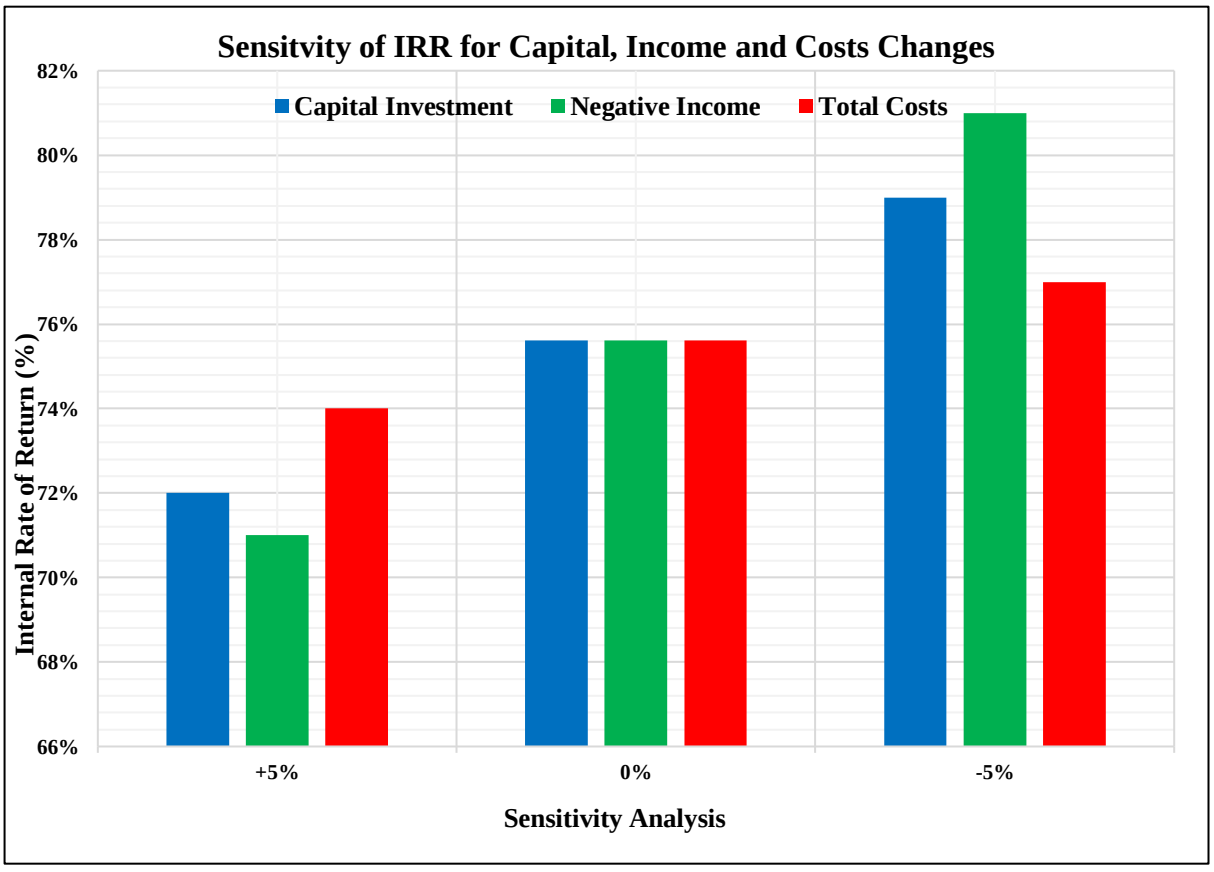

Figure 5.2: Sensitivity analysis on the internal rate of return

The results show that the internal rate of return (IRR) is most sensitive to changes in the project income. A 5 % increase in the projected income resulted in an increase in the IRR by 5 %. Conversely, a 5 % decrease in the projected income could decrease the IRR by 5 %. Nevertheless, the results show that the IRR is significantly higher than the project discount rate. Therefore, should possible fluctuations occur from cash flow streams, the project would still yield a significant profit over the life period. Fortunately, trends over the last five years have indicated an annual gold price increase between 8 and 18 % (Gold Price, 2020). Additionally, the gold price is quite sensitive to the rand to dollar (ZAR/USD) exchange rate, however, over the last five years, an average increase of 5.35 % in the ZAR/USD has been observed (X-Rates, 2020) . Should similar trends be observed in the future, profitability for implementation of the proposed lime treatment and gypsum crystallisation plant would be significantly larger.

6. CHAPTER 6: CONCLUSIONS AND RECOMMENDATIONS

A summary of the major findings, which significantly improve the understanding of the effect of water quality on gold leaching are presented below.

- Experimental work and statistical analysis indicated a significant difference in gold recoveries achieved between using potable water and untreated process water. An average improvement of ~10 % in gold recovery was achieved when using potable water over untreated process water, during straight leach tests. Similarly, an average improvement of ~5 % in gold recovery was achieved for flotation tails leach tests, with the use of potable water.

- Analysis of a composite sample of ERGO Brakpan's process water, confirmed the presence of appreciable concentrations of heavy metals which could result in the formation of weak acid dissociable (WAD) cyanide species during leaching. The metal species were identified as iron, nickel, zinc and manganese. The contaminants present in the highest concentrations were sulfate and calcium.

- A study of the effect of the identified contaminants on gold recovery was undertaken by spiking potable water with a known concentration of the respective contaminant. Leach tests showed that iron, nickel and zinc have the largest negative effect on gold recovery, with iron and nickel having a greater detrimental effect on gold recovery than zinc.

- Sulfates indicated possible passivation effects, which negatively influenced gold recoveries; however, this was to a lesser extent than that caused by the heavy metals.

- The increased gold recoveries observed as calcium concentrations increased agree with the possible formation of a calcium aurocyanide complex.

- The presence of excess magnesium showed a minimal increase in gold recovery.

- Due to the insignificant difference in the gold recoveries (approximately 1.3 %), between the lime treated process water and potable water (which was used as the benchmark), the use of lime was chosen as an attractive beneficiation option. Furthermore, flotation tails leach tests showed that process water treated with lime, produced lower washed residues and higher gold recoveries than the untreated process water utilised at ERGO Brakpan. These results proved that lime treatment alone on ERGO Brakpan's process water was sufficient to improve gold recoveries at ERGO Brakpan.

- Lime softening proved efficient in the reduction of heavy metal concentration by metal hydroxide precipitation as well as showing a reduction in sulfate concentration by the precipitation of gypsum.

- While the leach results indicated that lime treated process water at elevated temperatures produced higher gold recoveries; the heating requirement for 60 Ml of process water on a daily basis is not economically viable.

- Using a conservative gold price of R950 000/kg, an estimated monthly revenue of approximately R6 270 000 may be realised from processing of the 4L50 reclaimed ore with lime softened process water, which accounts for approximately 32 % of ERGO Brakpan's daily feed.

- An economic evaluation for a full-scale installation (20 Ml/day) of a lime treatment and gypsum crystallization facility showed a CAPEX of approximately R R49 085 809 and a monthly OPEX of R1 521 408.

- A discounted cash flow analysis showed a Net Present Value of approximate R333 000 000 within the project life (10 years) with an internal rate of return of 76 %

The following recommendations for future research, with regards to the effects on water quality of gold recovery:

- Since reclaimed material from only the 4L50 site was used for this thesis, it is recommended to establish the differences in gold recovery between the use of potable water and untreated process water on a composite sample of ERGO Brakpan's total feed lines.

- Prior to full scale implementation, the design and commissioning of a 60 m^3/day lime treatment and gypsum crystallisation pilot plant is recommended. The resultant treated water should be mixed with ERGO Brakpan feed material and proceed to a pilot plant leach train. Washed residues and gold recoveries obtained from the pilot plant study should be compared to plant set points and the results obtained from this investigation.

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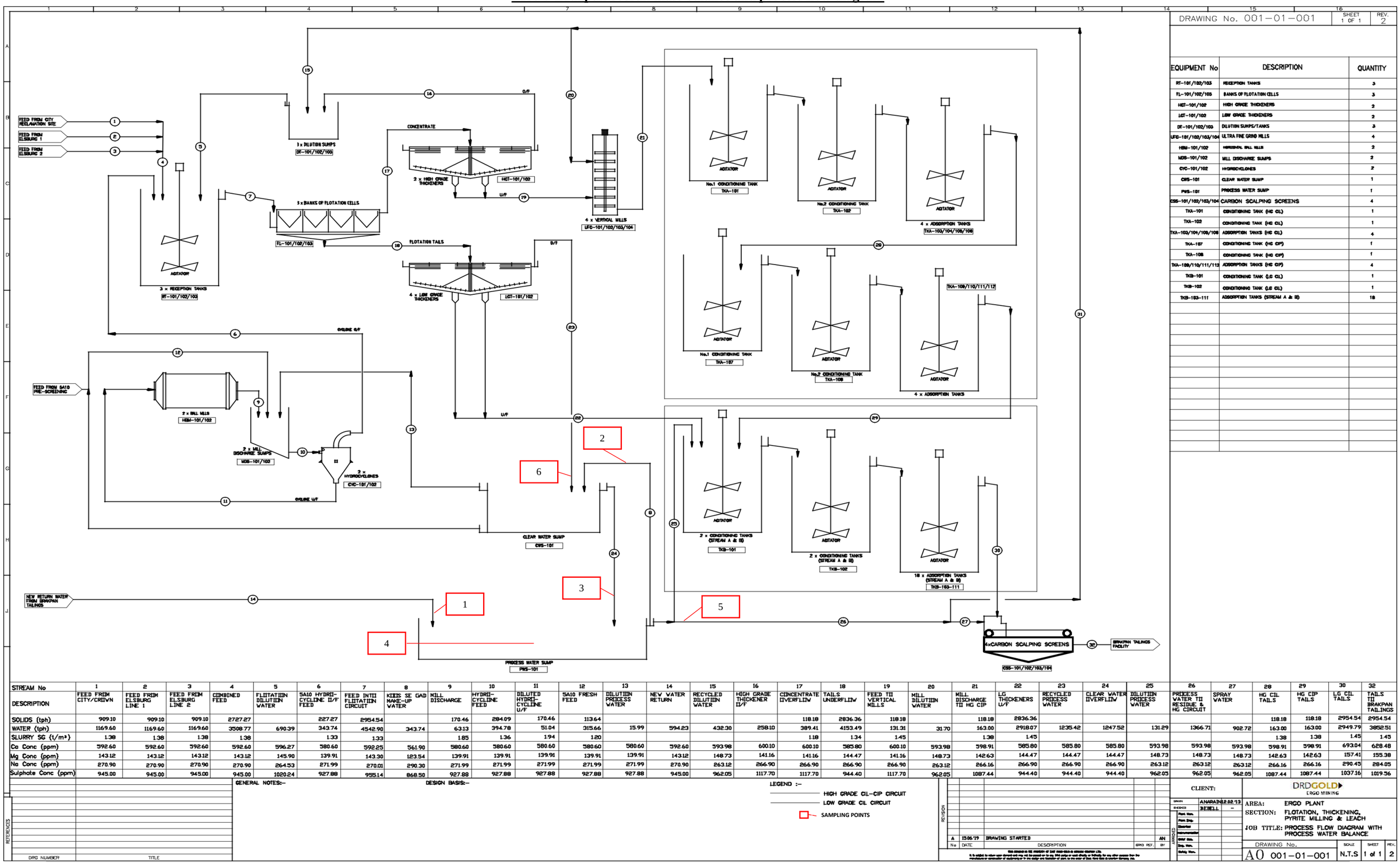
APPENDIX A – ICP-MS ANALYSIS OF COMPOSITE SAMPLES OF ERGO BRAKPAN'S PROCESS WATER

Table A.1 summaries the concentrations on contaminants within ERGO Brakpan's process water. Samples were collected over a period of one month from six various sampling points within the ERGO Brakpan plant and were submitted to the University of Johannesburg for ICP-MS analysis. The sampling points are highlighted in the process flow diagram and are presented as Figure A.1 below:

Table A.1: ICP-MS analysis of ERGO Brakpan's process water

Element Name	Calcium	Iron	Magnesium	Manganese	Sodium	Potassium	Silicon	Sulfur	Phosphorus	Lithium	Boron	Aluminium	Cobalt	Nickel	Copper	Zinc
Sampling Point	Ca	Fe	Mg	Mn	Na	K	Si	S	P	Li	B	Al	Co	Ni	Cu	Zn
	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	µg/l	µg/l	µg/l	µg/l	µg/l	µg/l	µg/l
1	593	112	143	43.2	271	38.4	19.5	945	0.5	134	344	2827	9538	21648	156	5400
2	562	7.1	124	37.7	290	38.3	13.9	869	0.5	108	255	1153	9780	18363	388	4317
3	586	104	145	45.3	267	39.0	15.5	944	0.5	133	236	1834	10201	21827	225	5428
4	593	51.3	127	39.4	289	37.7	13.9	947	0.6	111	214	1543	9490	18682	347	4377
5	595	18.6	124	39.5	283	38.2	13.0	1012	0.6	113	201	1504	9858	20283	595	4622
6	600	148	141	46.2	267	38.1	13.9	1118	0.6	139	209	5406	11458	25069	175	8508

ERGO Brakpan overall mass balance and process flow diagram



EQUIPMENT No	DESCRIPTION	QUANTITY
RT-101/102/103	RECEPTION TANKS	3
FL-101/102/103	BANKS OF FLOTATION CELLS	3
HGT-101/102	HIGH GRADE THICKENERS	2
LGT-101/102	LOW GRADE THICKENERS	2
DF-101/102/103	DILUTION SUMPS/TANKS	3
UFG-101/102/103/104	ULTRA FINE GRIND MILLS	4
HBM-101/102	HORIZONTAL BALL MILLS	2
MDS-101/102	MILL DISCHARGE SUMPS	2
CYC-101/102	HYDROCYCLONES	2
CWS-101	CLEAR WATER SUMP	1
PWS-101	PROCESS WATER SUMP	1
CSS-101/102/103/104	CARBON SCALPING SCREENS	4
TKA-101	CONDITIONING TANK (HG CIL)	1
TKA-102	CONDITIONING TANK (HG CIL)	1
TKA-103/104/108/108	ADSORPTION TANKS (HG CIL)	4
TKA-107	CONDITIONING TANK (HG CIP)	1
TKA-108	CONDITIONING TANK (HG CIP)	1
TKA-108/110/111/112	ADSORPTION TANKS (HG CIP)	4
TKB-101	CONDITIONING TANK (LG CIL)	1
TKB-102	CONDITIONING TANK (LG CIL)	1
TKB-103-111	ADSORPTION TANKS (STREAM A & B)	18

STREAM No	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30	32
DESCRIPTION	FEED FROM CITY/CRUSH	FEED FROM ELSBURG LINE 1	FEED FROM ELSBURG LINE 2	COMBINED FEED	FLOTATION DILUTION WATER	SAG HYDRO-CYCLONE U/F FEED	FEED INTO FLOTATION CIRCUIT	KIBBS SE GAP MAKE-UP WATER	MILL DISCHARGE	HYDRO-CYCLONE FEED	DILUTED HYDRO-CYCLONE U/F	SAG FRESH FEED	DILUTION PROCESS WATER	NEW WATER RETURN	RECYCLED DILUTION WATER	HIGH GRADE THICKENER U/F	CONCENTRATE OVERFLOW	TAILS UNDERFLOW	FEED TO VERTICAL MILLS	MILL DILUTION WATER	MILL DISCHARGE TO HG CIP	LG THICKENERS U/F	RECYCLED PROCESS WATER	CLEAR WATER OVERFLOW	DILUTION PROCESS WATER	PROCESS WATER TO RESIDUE & HG CIRCUIT	SPRAY WATER	HG CIL TAILS	HG CIP TAILS	LG CIL TAILS	TAILS TO BRAKPAN TAILINGS
SOLIDS (tph)	909.10	909.10	909.10	2727.27		227.27	2954.54		170.46	284.09	170.46	113.64						118.18	2836.36	118.18		118.18	2836.36					118.18	118.18	2954.54	2954.54
WATER (tph)	1169.60	1169.60	1169.60	3508.77	690.39	343.74	4542.90	343.74	63.13	394.78	51.04	315.66	15.99	594.23	432.30	258.10	389.41	4153.49	131.31	31.70	163.00	2918.07	1235.42	1247.52	131.29	1366.71	902.72	163.00	163.00	2949.79	3852.51
SLURRY SG (t/m³)	1.38	1.38	1.38	1.38		1.33	1.33		1.85	1.36	1.94	1.20					1.18	1.34	1.45		1.38	1.45						1.38	1.38	1.45	1.45
Ca Conc (ppm)	592.60	592.60	592.60	592.60	596.27	580.60	592.25	561.90	580.60	580.60	580.60	580.60	580.60	592.60	593.98	600.10	600.10	585.80	600.10	593.98	598.91	585.80	585.80	585.80	593.98	593.98	593.98	598.91	598.91	693.04	628.48
Mg Conc (ppm)	143.12	143.12	143.12	143.12	145.90	139.91	143.30	123.54	139.91	139.91	139.91	139.91	139.91	143.12	148.73	141.16	141.16	144.47	141.16	148.73	142.63	144.47	144.47	144.47	148.73	148.73	148.73	142.63	142.63	157.41	155.38
Na Conc (ppm)	270.90	270.90	270.90	270.90	264.53	271.99	270.01	290.30	271.99	271.99	271.99	271.99	271.99	270.90	263.12	266.90	266.90	266.90	266.90	263.12	266.16	266.90	266.90	266.90	263.12	263.12	263.12	266.16	266.16	290.45	284.05
Sulphate Conc (ppm)	945.00	945.00	945.00	945.00	1020.24	927.88	955.14	868.50	927.88	927.88	927.88	927.88	927.88	945.00	962.05	1117.70	1117.70	944.40	1117.70	962.05	1087.44	944.40	944.40	944.40	962.05	962.05	962.05	1087.44	1087.44	1037.16	1019.56

GENERAL NOTES:-

DESIGN BASIS:-

LEGEND :-
- HIGH GRADE CIL-CIP CIRCUIT
- LOW GRADE CIL CIRCUIT
- SAMPLING POINTS

REFERENCES

DRG NUMBER	TITLE

No	DATE	DESCRIPTION	ERGO REF.	BY
A	15.06.19	DRAWING STARTED		AN

CLIENT:

DRDGOLD
ERGO MINING

DRAWN	ANARAINE	02.07.19
CHECKED	BETTELL	--
Plant Mgr.		
Plant Engr.		
Chemist		
Metallurgist		
Metal Mgr.		
Engr. Mgr.		
Geology Mgr.		

AREA: ERGO PLANT
SECTION: FLOTATION, THICKENING, PYRITE MILLING & LEACH
JOB TITLE: PROCESS FLOW DIAGRAM WITH PROCESS WATER BALANCE

DRAWING No.	SCALE	SHEET	REV.
A0 001-01-001	N.T.S	1 of 1	2

Figure A.1: ERGO Brakpan overall mass balance and process flow diagram

APPENDIX B - ATOMIC ABSORPTION CALIBRATION CURVES

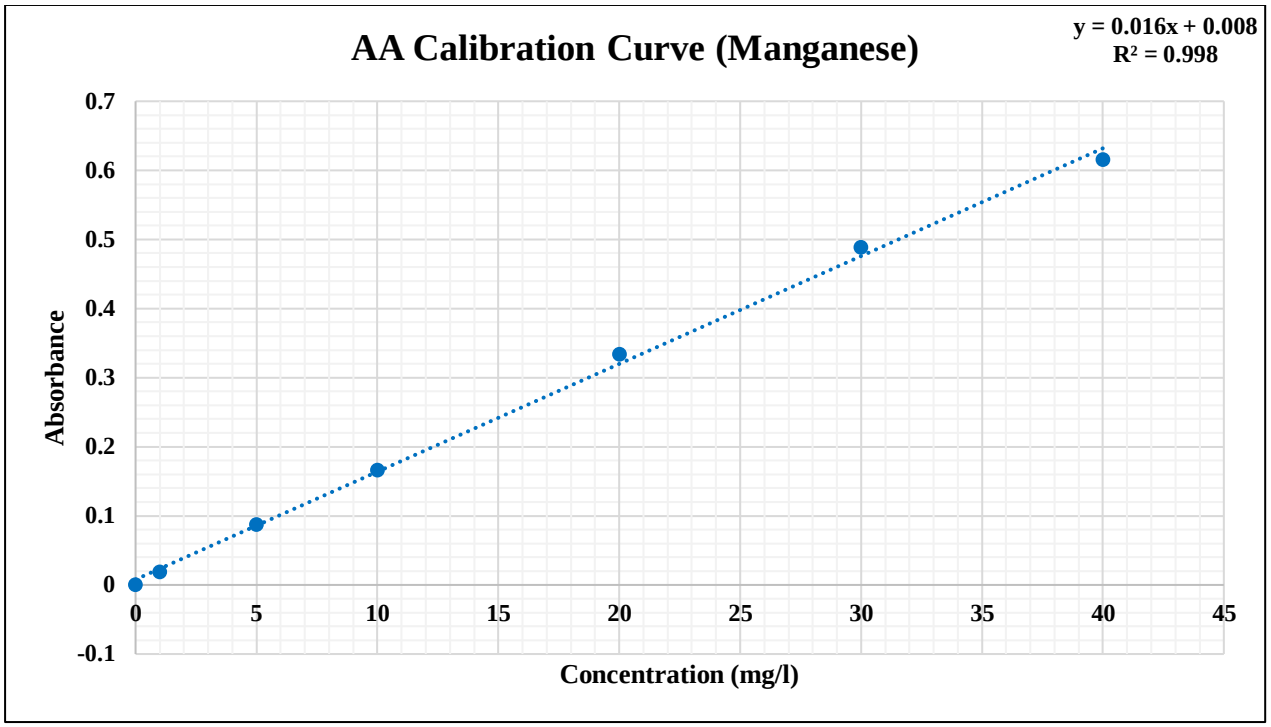


Figure B.1: AA calibration curve - Manganese

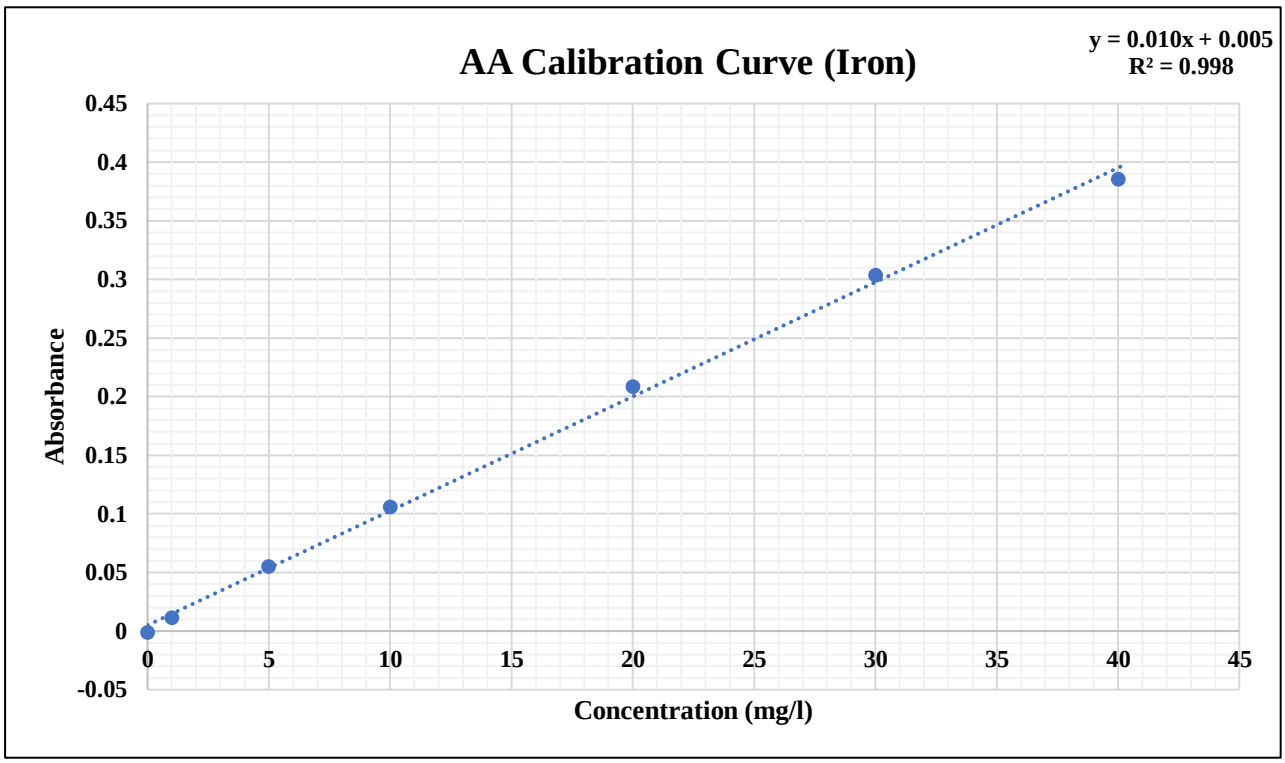


Figure B.2: AA calibration curve - Iron

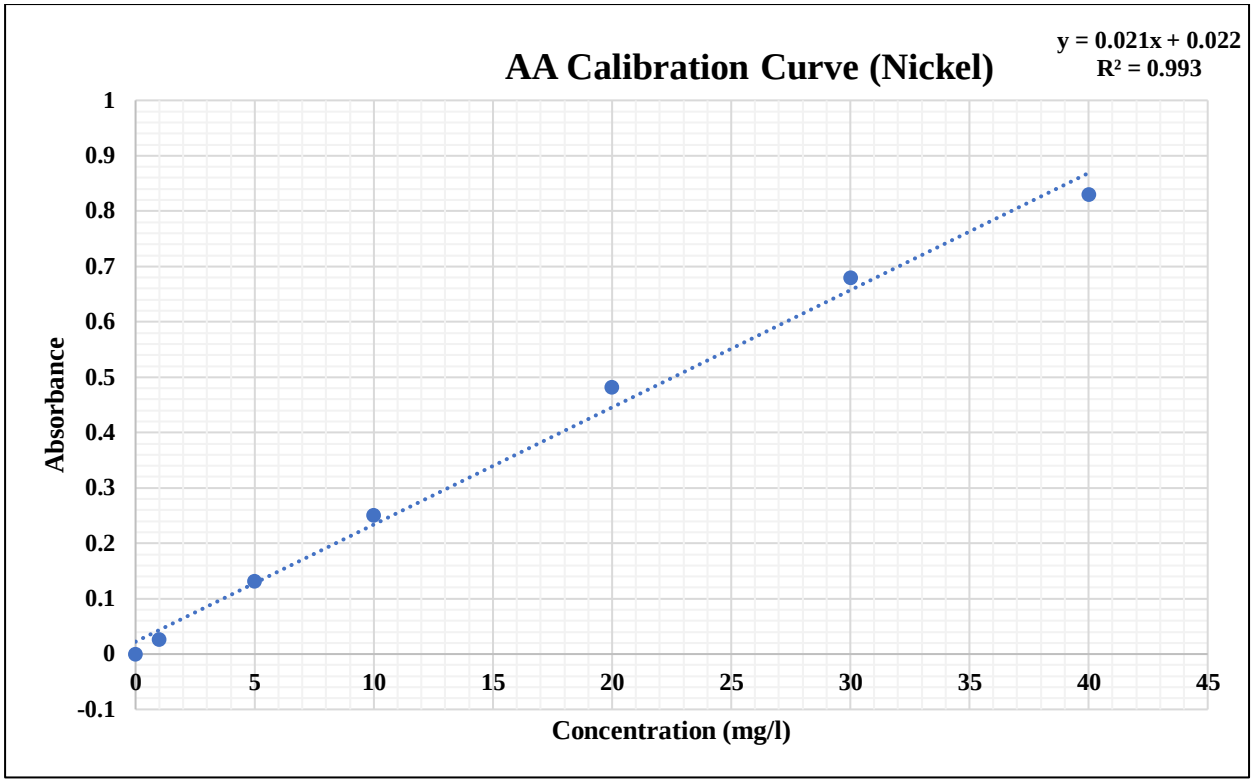


Figure B.3: AA calibration curve - Nickel

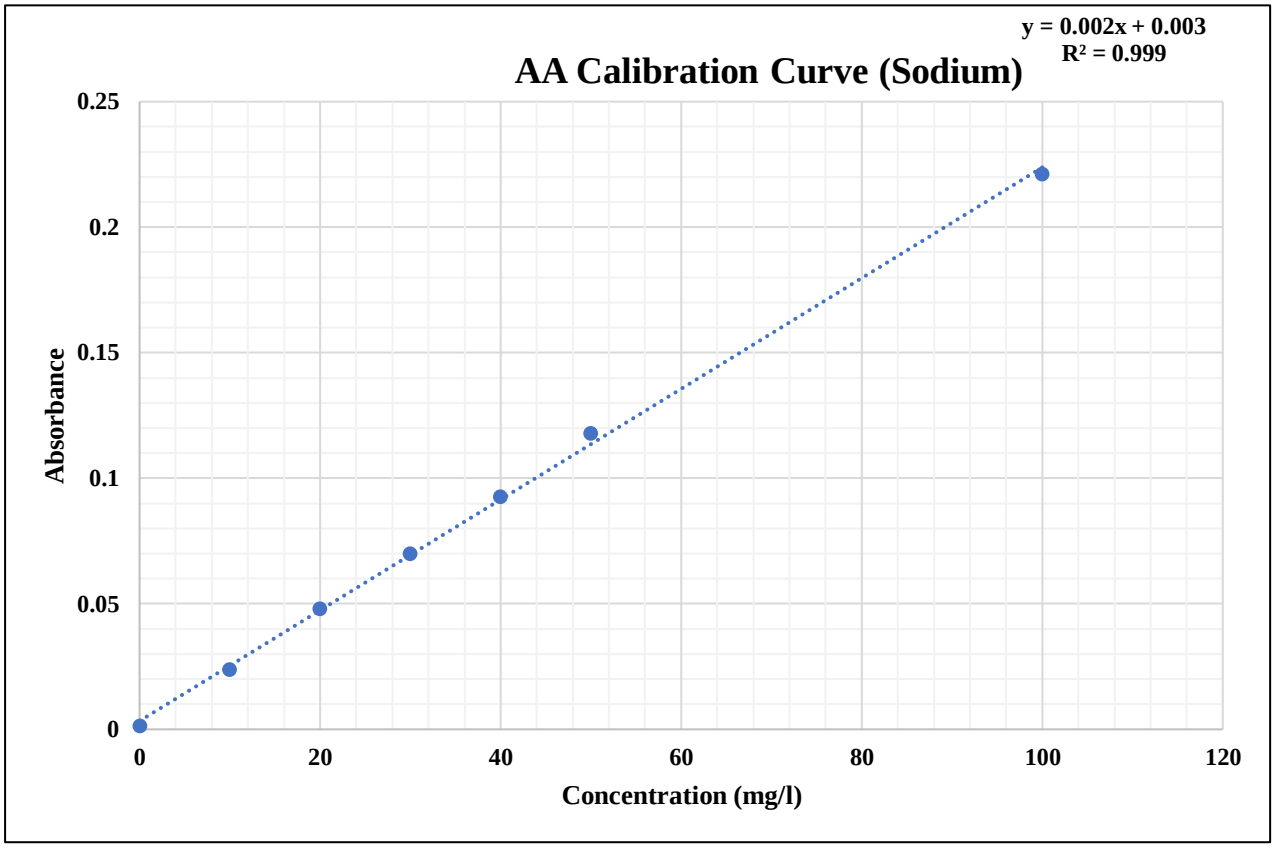


Figure B.4: AA calibration curve - Sodium

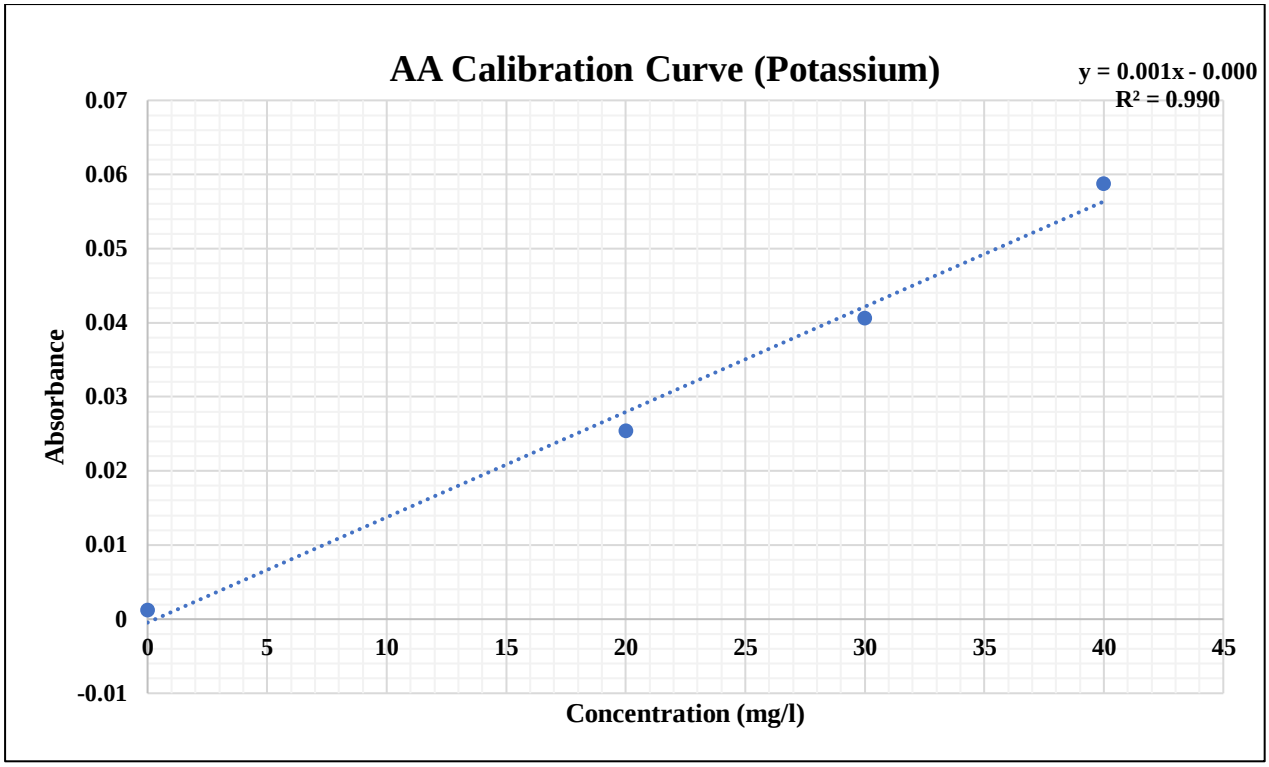


Figure B.5: AA calibration curve - Potassium

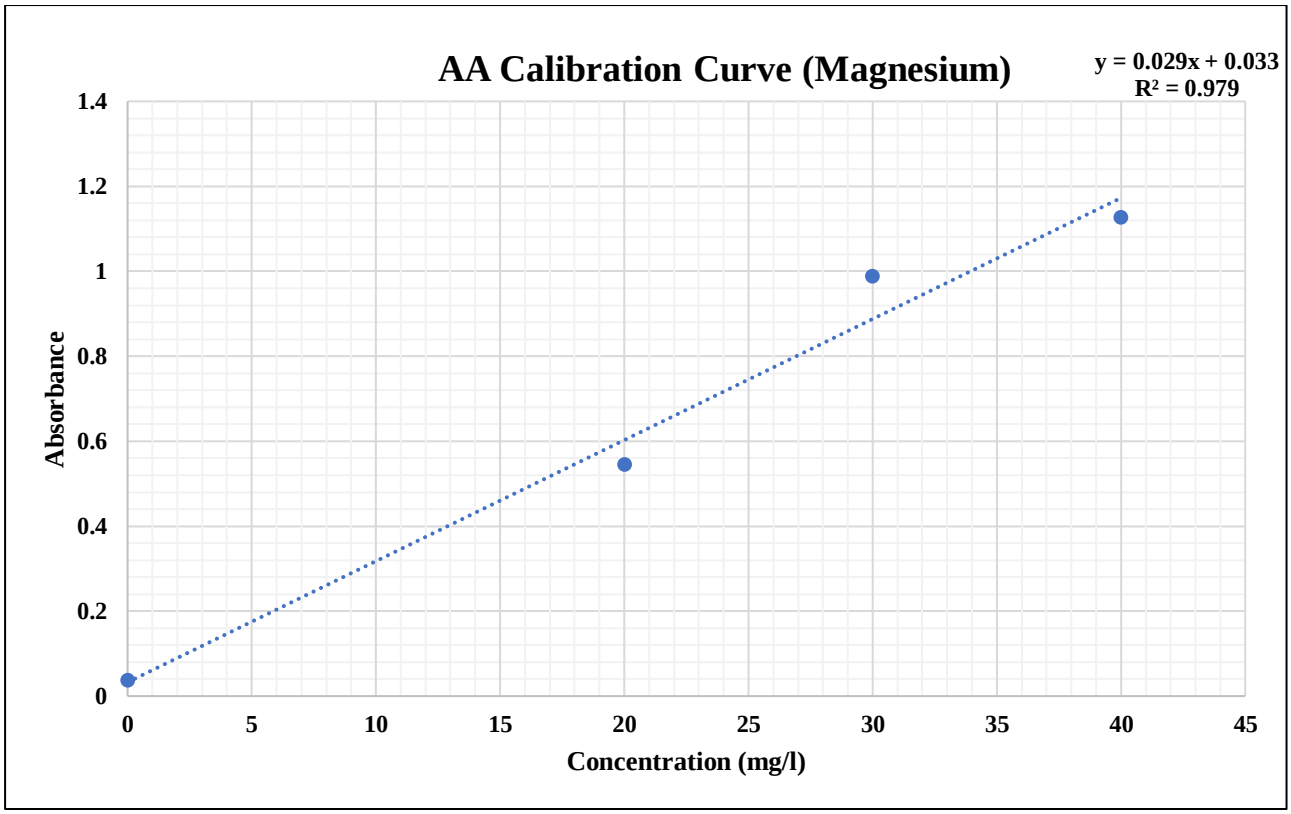


Figure B.6: AA calibration curve - Magnesium

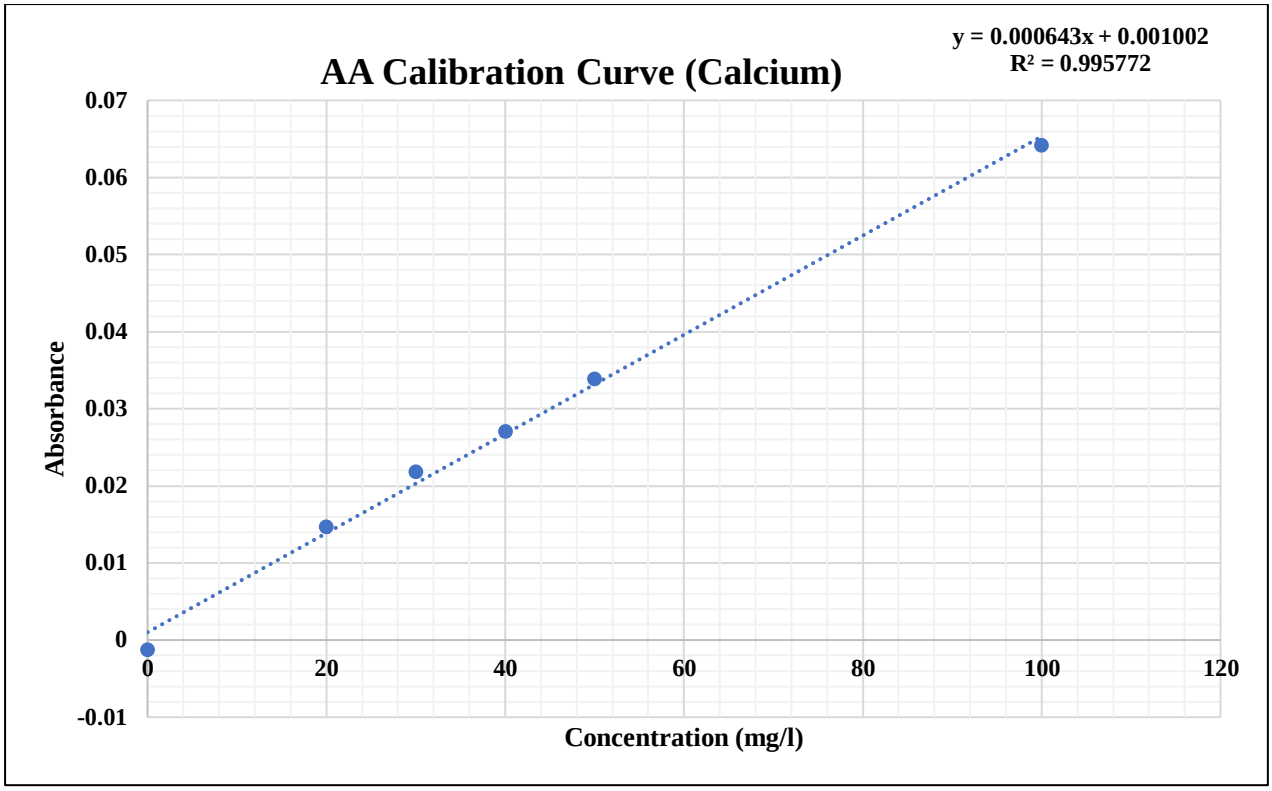


Figure B.7: AA calibration curve – Calcium

APPENDIX C – STATISTICAL ANALYSIS DATA

The descriptive statistical analysis for the results presented in Section 4.1 is presented below.

Leach statistical analysis data

Table C.1: Statistical analysis of straight leach test data using potable water

Potable water Leach Tests				
Test Number	Head	Washed Residue	Gold Dissolution	Gold Accountability
	g/t	g/t	%	%
1	0.200	0.097	51.4	103.1
2	0.200	0.095	52.5	99.0
3	0.200	0.098	51.0	103.2
4	0.210	0.098	53.2	101.0
5	0.190	0.086	54.7	103.3
6	0.210	0.095	55.0	100.7
7	0.200	0.096	52.2	100.0
8	0.200	0.093	53.7	102.3
9	0.200	0.087	56.6	100.0
10	0.210	0.099	53.0	100.0
Mean	**0.202**	**0.094**	**53.3**	**101.3**
Standard Deviation	0.006	0.004	1.6	1.5
95% Confidence	0.012	0.009	3.2	3.0

Table C.2: Statistical analysis of straight leach test data using process water

Recycled Process Water Leach Tests				
Test Number	Head	Washed Residue	Gold Dissolution	Gold Accountability
	g/t	g/t	%	%
1	0.200	0.106	47.0	105.1
2	0.200	0.108	46.0	101.0
3	0.200	0.102	49.1	104.2
4	0.210	0.103	50.8	102.0
5	0.190	0.106	44.2	98.3
6	0.210	0.105	50.2	100.7
7	0.200	0.107	46.7	108.0
8	0.200	0.100	50.0	100.3
9	0.200	0.107	46.6	98.0
10	0.210	0.109	48.3	101.0
Mean	**0.202**	**0.105**	**47.9**	**101.9**
Standard Deviation	0.006	0.003	2.0	2.9
95% Confidence	0.012	0.005	4.1	5.9

The raw data the straight leach t-test results presented in Section 4.1 is presented below.

Table C.3: t-Test raw data for straight leach tests

t-Test Raw Data for Leach Tests		
	Washed Residue	
Test Number	Potable water	Process Water
1	0.097	0.106
2	0.095	0.108
3	0.098	0.102
4	0.098	0.103
5	0.086	0.106
6	0.095	0.105
7	0.096	0.107
8	0.093	0.100
9	0.087	0.107
10	0.099	0.109
Mean	0.094	0.105

An in-depth statistical analysis of the data presented in Table C.4.

Table C.4: Descriptive statistics on washed residues during leach tests

Descriptive Statistics on Washed Residues during Leach Tests			
Potable water		**Process Water**	
Mean	0.094	Mean	0.105
Standard Error	<0.01	Standard Error	<0.01
Median	0.10	Median	0.11
Mode	0.10	Mode	0.11
Standard Deviation	<0.01	Standard Deviation	<0.01
Sample Variance	<0.01	Sample Variance	<0.01
Kurtosis	0.05	Kurtosis	-0.27
Skewness	-1.11	Skewness	-0.74
Range	0.01	Range	0.01
Minimum	0.09	Minimum	0.10
Maximum	0.10	Maximum	0.11
Sum	0.94	Sum	1.05
Count	10.00	Count	10.00

Flotation leach statistical analysis data

Table C.5: Statistical analysis of flotation tails leach test data using potable water

Potable water Flotation Tails Leach				
Test Number	Head	Washed Residue	Gold Dissolution	Gold Accountability
	g/t	g/t	%	%
1	0.16	0.082	48.6	101.5
2	0.15	0.079	47.6	99.4
3	0.16	0.092	42.3	104.3
4	0.16	0.080	49.9	107.6
5	0.15	0.082	45.4	99.7
6	0.16	0.090	43.7	100.2
7	0.14	0.076	46.0	105.2
8	0.16	0.083	47.8	99.3
9	0.15	0.085	43.2	102.1
10	0.16	0.088	45.0	108.2
Mean	**0.155**	**0.084**	**45.9**	**102.7**
Standard Deviation	0.007	0.005	2.4	3.2
95% Confidence	0.013	0.010	4.7	6.4

Table C.6: Statistical analysis of flotation tails leach test data using process water

Recycled Process Water Flotation Tails Leach				
Test Number	Head	Washed Residue	Gold Dissolution	Gold Accountability
	g/t	g/t	%	%
1	0.19	0.113	40.4	105.5
2	0.18	0.109	39.7	103.6
3	0.19	0.112	40.9	102.3
4	0.18	0.116	35.8	106.6
5	0.19	0.111	41.8	103.7
6	0.19	0.110	41.9	110.2
7	0.18	0.116	35.8	104.2
8	0.19	0.103	45.6	108.6
9	0.19	0.115	39.4	101.1
10	0.19	0.108	43.2	108.7
Mean	**0.187**	**0.111**	**40.4**	**105.4**
Standard Deviation	0.005	0.004	2.9	2.9
95% Confidence	0.009	0.007	5.8	5.7

The raw data the flotation tails leach t-test results presented in Section 4.1 is presented below.

Table C.7: t-Test raw data for flotation tails leach tests

t-Test Raw Data for Flotation Tails Leach Tests		
	Washed Residue	
Test Number	Potable water	Process Water
1	0.082	0.113
2	0.079	0.109
3	0.092	0.112
4	0.080	0.116
5	0.082	0.111
6	0.090	0.110
7	0.076	0.116
8	0.083	0.103
9	0.085	0.115
10	0.088	0.108
Mean	0.084	0.111

An in-depth statistical analysis of the data presented in Table C.8.

Table C.8: Descriptive statistics on washed residues during leach tests

Descriptive Statistics on Washed Residues during Leach Tests			
Potable water		**Process Water**	
Mean	0.084	Mean	0.111
Standard Error	<0.01	Standard Error	<0.01
Median	0.08	Median	0.11
Mode	0.08	Mode	0.11
Standard Deviation	0.01	Standard Deviation	<0.01
Sample Variance	<0.01	Sample Variance	<0.01
Kurtosis	-0.62	Kurtosis	0.21
Skewness	0.21	Skewness	-0.72
Range	0.02	Range	0.01
Minimum	0.08	Minimum	0.10
Maximum	0.09	Maximum	0.12
Sum	0.84	Sum	1.11
Count	10.00	Count	10.00

APPENDIX D - X-RAY DIFFRACTION (XRD) ANALYSIS

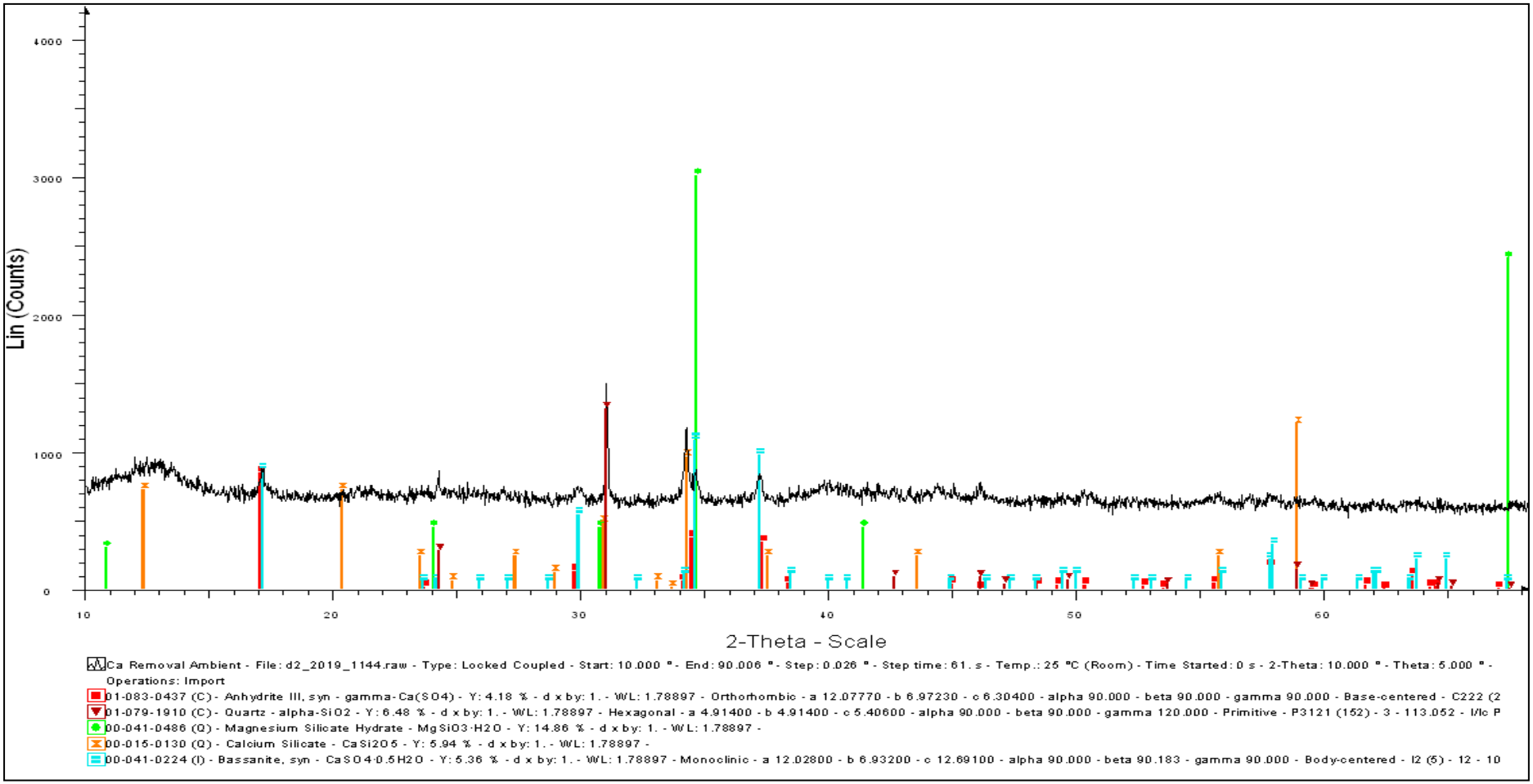


Figure D.1: XRD scan for lime softening at ambient temperature (20°C)

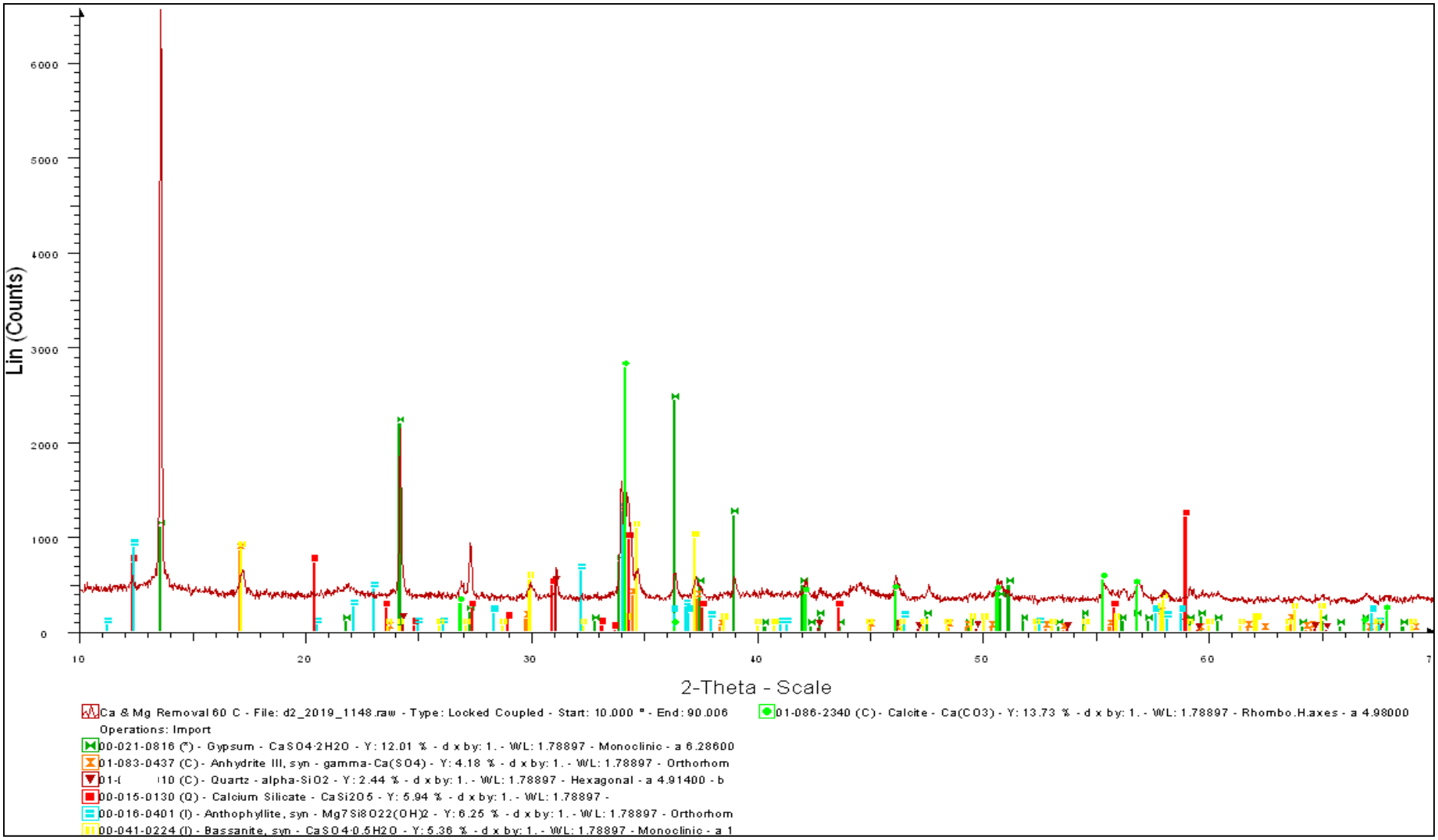


Figure D.2: XRD scan for lime softening at 60 °C

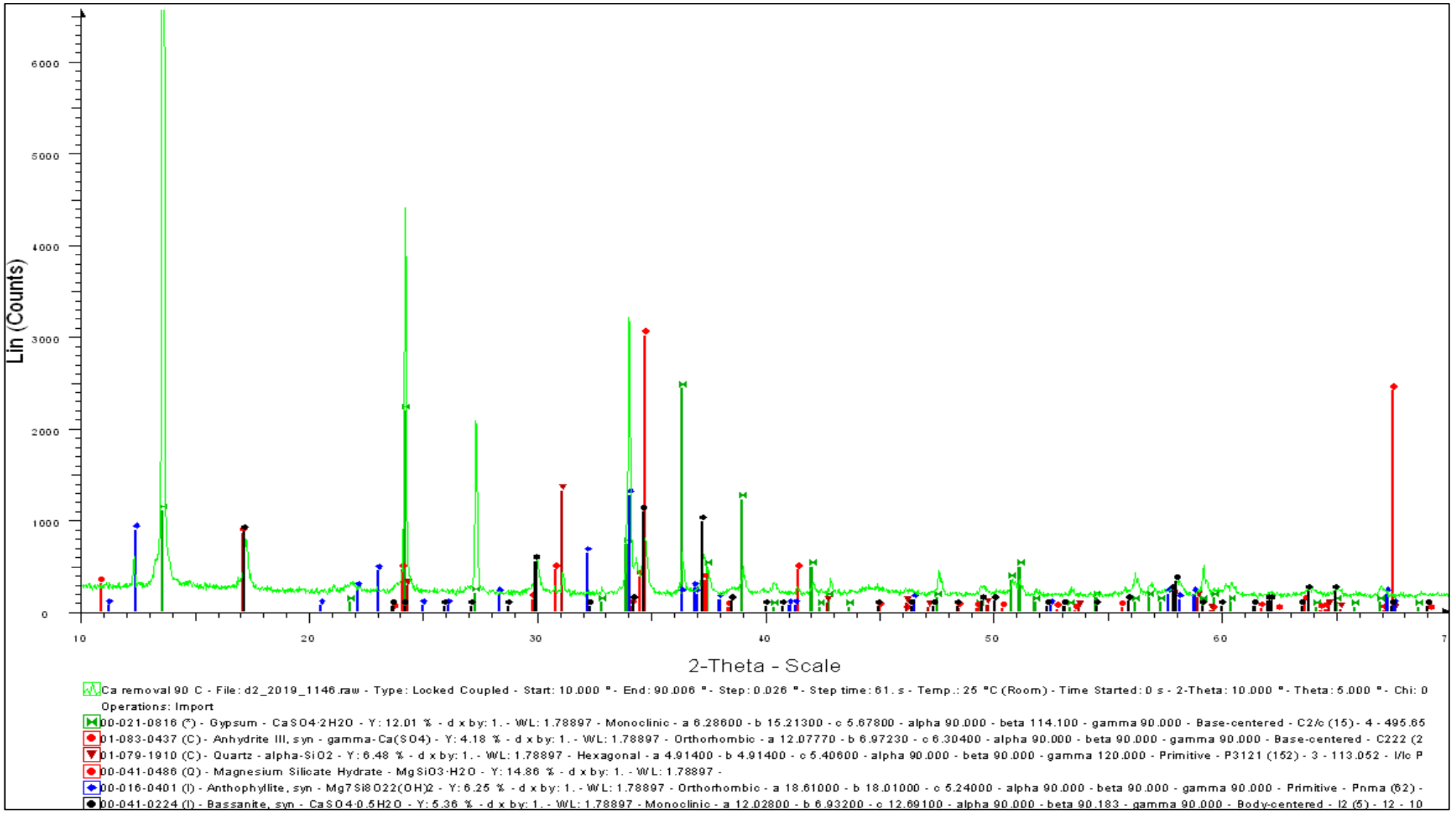


Figure D.3: XRD scan for lime softening at 90 °C

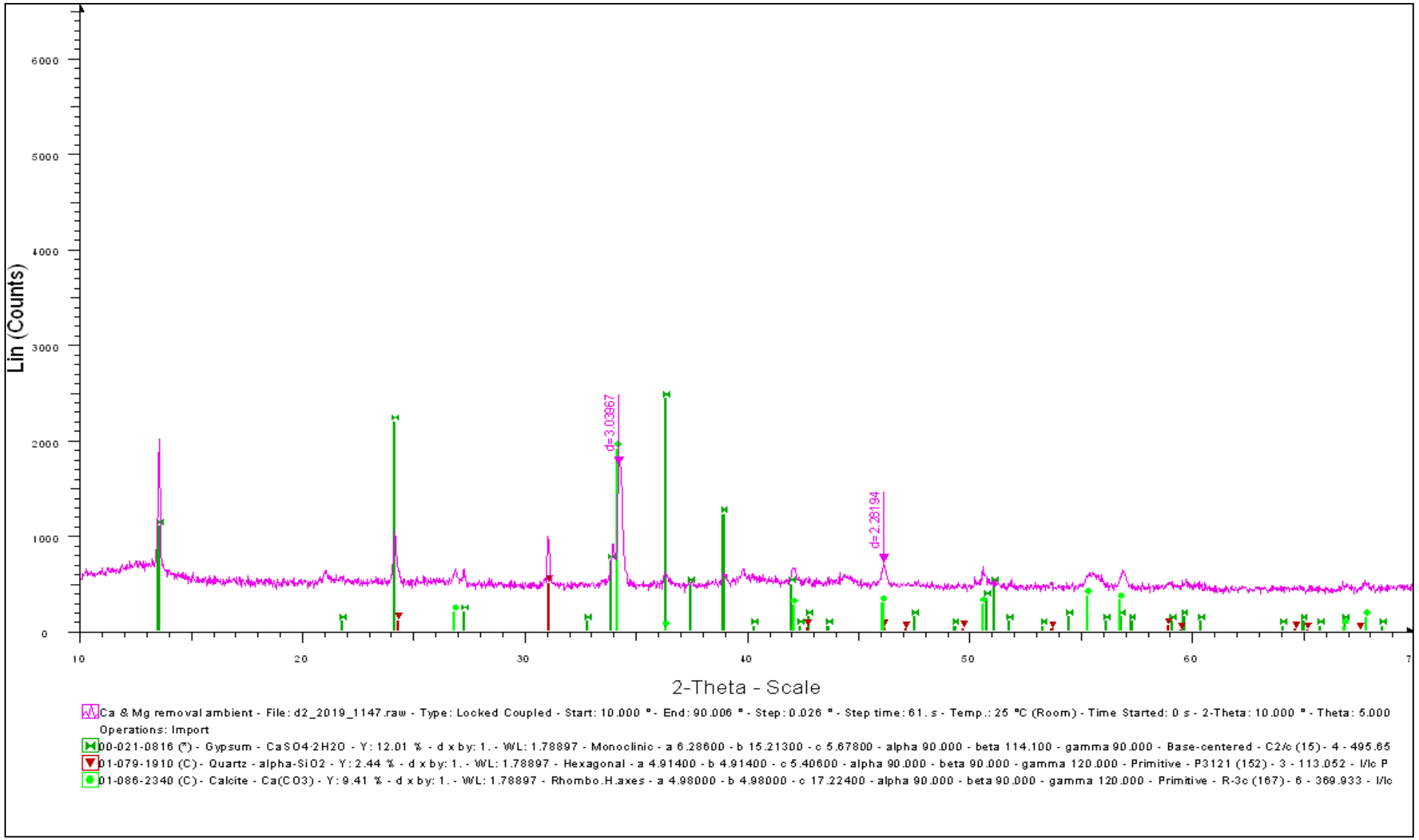


Figure D.4: XRD scan for lime and soda ash softening at ambient temperature (20°C)

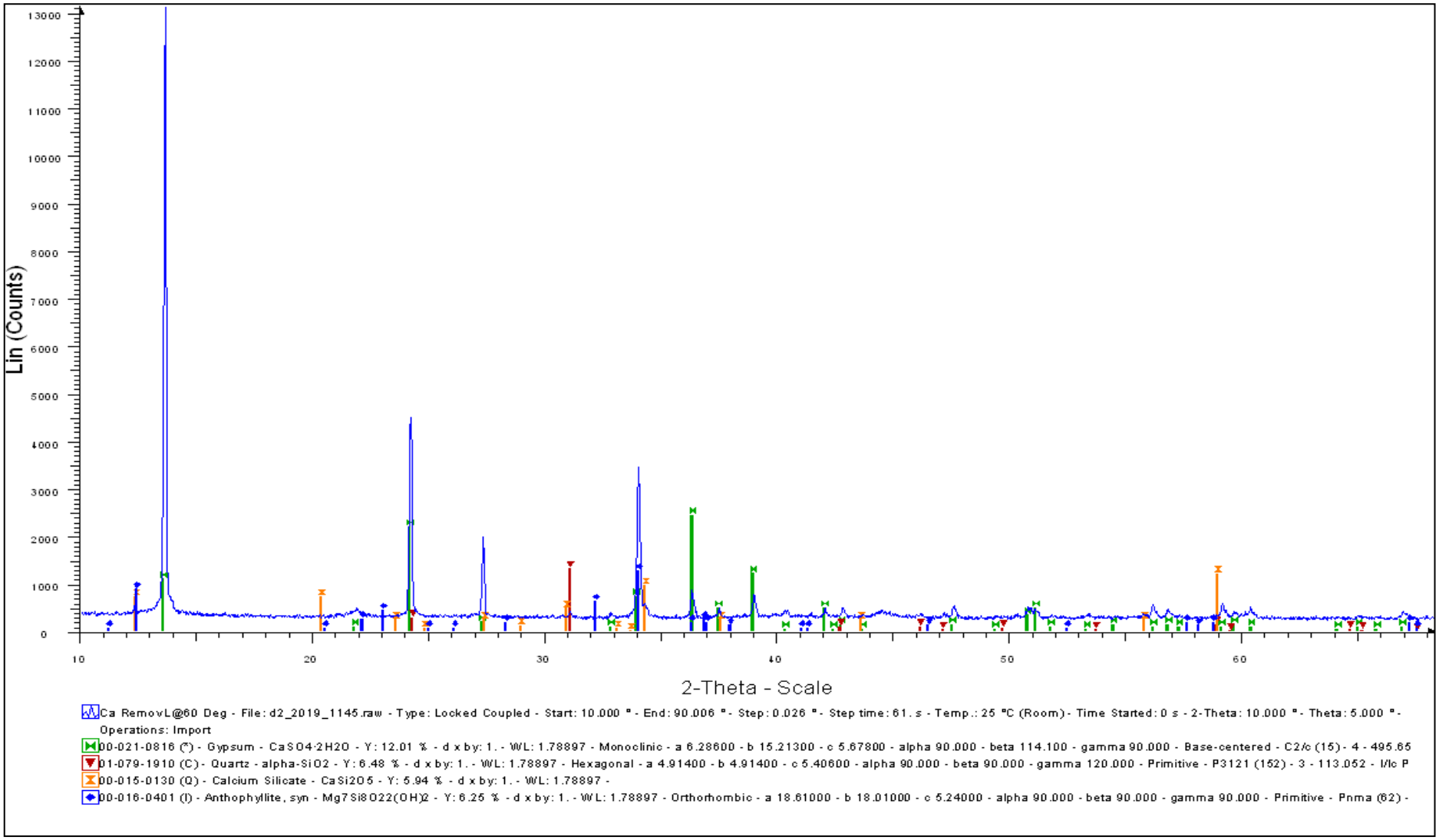


Figure D.5: XRD scan for lime and soda ash softening at 60 °C

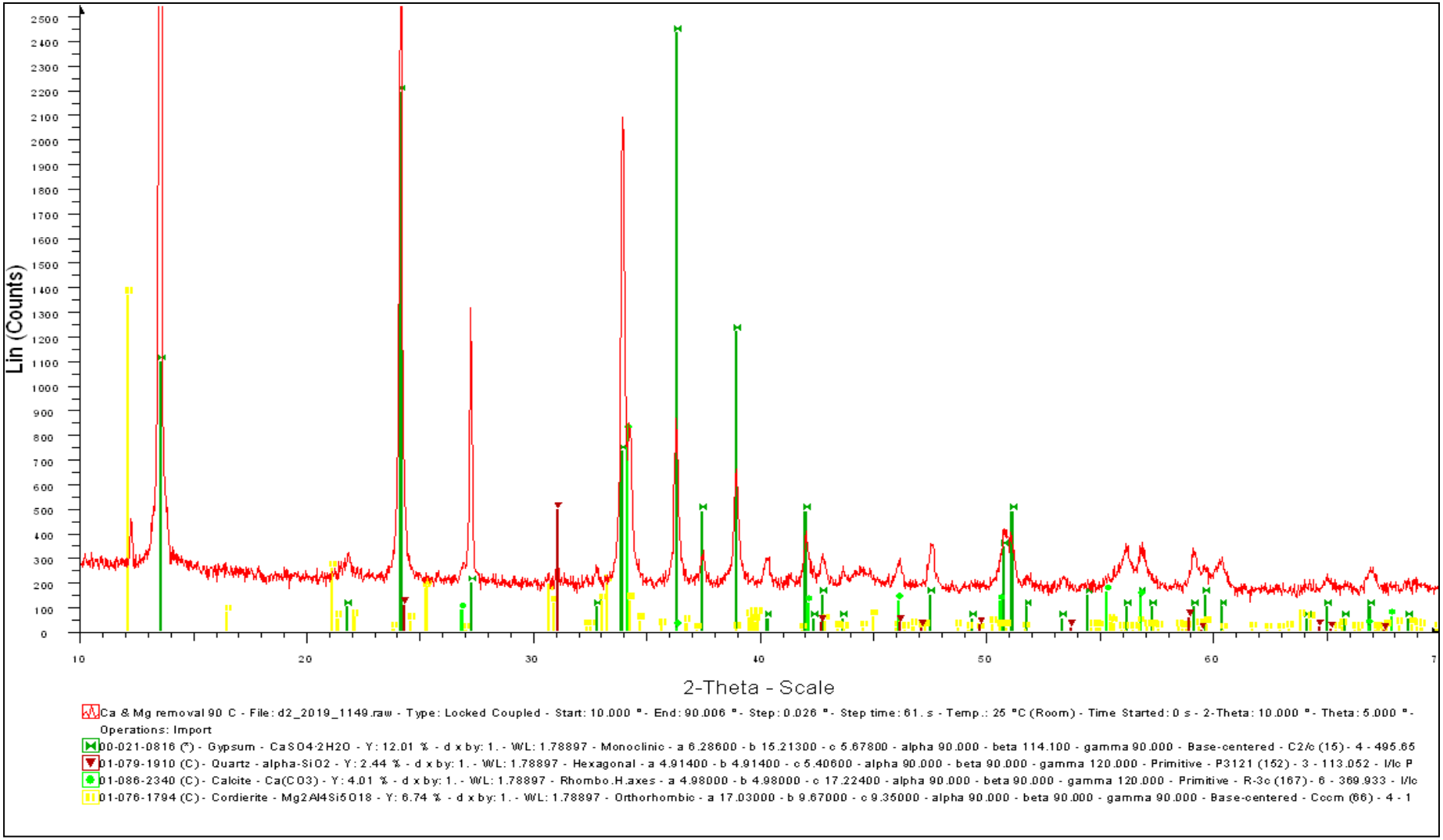


Figure D.6: XRD scan for lime and soda ash softening at 90 °C

APPENDIX E – LEACH TESTS RAW DATA

Contaminant spiked potable water leach results

The raw data for the tests in which potable water was spiked with individual contaminants are presented below. Refer to Appendix F for the gold mass balance and accountability sample calculations.

Table E.1: Leach results for sulfate spiked potable water

Leach Results for Sulfate spiked potable water						
Sample Name	Head grade	Washed Residue	Gold Dissolution	Gold in Filtrate	Gold Loaded Carbon	Gold Account-ability
	g/t	g/t	%	mg/l	g/t	%
50 mg/l Sulfate Addition	0.200	0.113	43.5	0.002	3.30	103.6
100 mg/l Sulfate Addition	0.200	0.127	36.5	0.002	2.60	100.7
500 mg/l Sulfate Addition	0.200	0.137	31.5	<0.001	2.50	102.9
Mean	0.200	0.126	37.2	0.001	2.80	102.4
Standard Deviation	0.000	0.01	6.03	0.001	0.44	1.50
95% Confidence	0.000	0.02	12.06	0.002	0.87	3.01

Table E.2: Leach results for calcium spiked potable water

Leach Results for Calcium spiked potable water						
Sample Name	Head grade	Washed Residue	Gold Dissolution	Gold in Filtrate	Gold Loaded Carbon	Gold Account-ability
	g/t	g/t	%	mg/l	g/t	%
50 mg/l Calcium Addition	0.200	0.133	33.3	0.003	2.65	102.9
100 mg/l Calcium Addition	0.200	0.122	38.9	<0.001	2.68	100.3
500 mg/l Calcium Addition	0.200	0.117	41.7	<0.001	3.10	101.2
Mean	0.200	0.124	38.0	0.001	2.81	101.5
Standard Deviation	0.000	0.01	4.25	0.001	0.25	1.28
95% Confidence	0.000	0.02	8.50	0.002	0.50	2.57

Table E.3: Leach results for iron spiked potable water

Leach Results for Calcium spiked potable water						
Sample Name	Head grade	Washed Residue	Gold Dissolution	Gold in Filtrate	Gold Loaded Carbon	Gold Account-ability
	g/t	g/t	%	Mg/l	g/t	%
50 mg/l Iron Addition	0.200	0.132	34.0	0.002	3.05	108.3
100 mg/l Iron Addition	0.200	0.158	20.8	0.001	1.60	102.3
500 mg/l Iron Addition	0.200	0.188	6.00	<0.001	0.60	101.6
Mean	0.200	0.159	20.3	0.001	1.75	104.1
Standard Deviation	0.000	0.03	14.01	0.001	1.23	3.69
95% Confidence	0.000	0.06	28.02	0.002	2.46	7.38

Table E.4: Leach results for nickel spiked potable water

Leach Results for Calcium spiked potable water						
Sample Name	Head grade	Washed Residue	Gold Dissolution	Gold in Filtrate	Gold Loaded Carbon	Gold Account-ability
	g/t	g/t	%	Mg/l	g/t	%
50 mg/l Nickel Addition	0.200	0.122	39.0	0.002	2.95	101.5
100 mg/l Nickel Addition	0.200	0.146	27.0	0.002	1.81	100.1
500 mg/l Nickel Addition	0.200	0.168	16.0	0.002	1.35	102.0
Mean	0.200	0.145	27.3	0,002	2.04	101.2
Standard Deviation	0.000	0.02	11.50	0,002	0.82	0.96
95% Confidence	0.000	0.05	23.01	0,004	1.65	1.92

Table E.5: Leach results for zinc spiked potable water

Leach Results for Calcium spiked potable water						
Sample Name	Head grade	Washed Residue	Gold Dissolution	Gold in Filtrate	Gold Loaded Carbon	Gold Account-ability
	g/t	g/t	%	Mg/l	g/t	%
50 mg/l Zinc Addition	0.200	0.118	41.0	0.001	3.30	101.7
100 mg/l Zinc Addition	0.200	0.130	35.0	0.001	2.86	103.2
500 mg/l Zinc Addition	0.200	0.141	29.5	<0,001	2.45	103.0
Mean	0.200	0.130	35.2	0.001	2.87	102.6
Standard Deviation	0.000	0.01	5.75	0.001	0.43	0.82
95% Confidence	0.000	0.02	11.50	0.002	0.85	1.63

Table E.6: Leach results for magnesium spiked potable water

Leach Results for Calcium spiked potable water						
Sample Name	Head grade	Washed Residue	Gold Dissolution	Gold in Filtrate	Gold Loaded Carbon	Gold Account-ability
	g/t	g/t	%	Mg/l	g/t	%
50 mg/l Mg Addition	0.200	0.125	37.50	0.002	3.05	104.4
100 mg/l Mg Addition	0.200	0.122	39.00	0.002	3.20	107.5
500 mg/l Mg Addition	0.200	0.120	40.00	<0,001	3.30	105.7
Mean	0.200	0.122	38.8	<0.001	3.18	103.7
Standard Deviation	0.000	<0.01	1.26	<0.001	0.126	1.59
95% Confidence	0.000	<0.05	2.52	<0.002	0.25	3.18

Treated process water leach results

The raw data for the straight leach and flotation tails leach is presented below.

Table E.7: Lime softening at ambient conditions - straight leach tests

Lime Softening at Ambient Conditions - Straight Leach Tests						
Sample Number	Head grade	Washed Residue	Gold Dissolution	Gold in Filtrate	Gold Loaded Carbon	Gold Account-ability
	g/t	g/t	%	Mg/l	g/t	%
Test 1	0.200	0.100	49.2	0.003	3.69	102.6
Test 2	0.200	0.096	51.0	0.003	3.85	101.9
Test 3	0.200	0.087	55.9	0.002	5.06	101.7
Mean	0.200	0.094	52.0	0.003	4.20	102.1
Standard Deviation	0.000	0.007	3.47	0.001	0.75	0.47
95% Confidence	0.000	0.014	6.94	0.001	1.50	0.95

Table E.8: Lime softening at ambient conditions - flotation tails leach tests

Lime Softening at Ambient Conditions - Flotation Tails Leach Tests						
Sample Number	Head grade	Washed Residue	Gold Dissolution	Gold in Filtrate	Gold Loaded Carbon	Gold Account-ability
	g/t	g/t	%	Mg/l	g/t	%
Test 1	0.150	0.100	33.3	0.002	2.28	103.7
Test 2	0.150	0.100	33.3	0.001	2.32	103.9
Test 3	0.150	0.087	42.3	0.001	3.44	101.2
Mean	0.150	0.096	36,3	0.001	2.68	102.9
Standard Deviation	0.000	0.008	5.16	0.001	0.66	1.50
95% Confidence	0.000	0.015	10.3	0.001	1.32	3.00

Table E.9: Lime-soda ash softening at ambient conditions - straight leach tests

Lime-Soda Ash Softening at Ambient Conditions - Straight Leach Tests						
Sample Number	Head grade	Washed Residue	Gold Dissolution	Gold in Filtrate	Gold Loaded Carbon	Gold Account-ability
	g/t	g/t	%	Mg/l	g/t	%
Test 1	0.200	0.113	43.5	0.001	3.30	98.8
Test 2	0.200	0.096	52.2	0.001	3.95	101.5
Test 3	0.197	0.104	46.9	0.001	3.60	97.3
Mean	0.199	0.104	47.5	0.001	3.62	99.2
Standard Deviation	0.002	0.009	4.37	0.000	0.33	2.12
95% Confidence	0.004	0.017	8.75	0.000	0.65	4.25

Table E.10: Lime-soda ash softening at ambient conditions - flotation tails leach tests

Lime-Soda Ash Softening at Ambient Conditions - Flotation Tails Leach Tests						
Sample Number	Head grade	Washed Residue	Gold Dissolution	Gold in Filtrate	Gold Loaded Carbon	Gold Account-ability
	g/t	g/t	%	Mg/l	g/t	%
Test 1	0.157	0.104	33.5	0.002	2.28	101.9
Test 2	0.157	0.100	36.1	0.001	2.32	99.6
Test 3	0.157	0.103	34.0	0.001	3.44	101.9
Mean	0.157	0.102	34.5	0.001	2.68	101.1
Standard Deviation	0.000	0.002	1.37	0.001	0.66	1.34
95% Confidence	0.000	0.004	2.74	0.001	1.32	2.67

Table E.11: Lime softening at 60 °C- straight leach tests

Lime Softening at 60 °C- Straight Leach Tests						
Sample Number	Head grade	Washed Residue	Gold Dissolution	Gold in Filtrate	Gold Loaded Carbon	Gold Account-ability
	g/t	g/t	%	Mg/l	g/t	%
Test 1	0.200	0.088	56.3	0.001	4.40	103.3
Test 2	0.200	0.094	53.1	0.002	4.30	104.4
Test 3	0.200	0.075	62.5	0.001	4.40	103.8
Mean	0.200	0.085	57.3	0.001	4.37	103.9
Standard Deviation	0.000	0.010	4.77	0.001	0.06	4.33
95% Confidence	0.000	0.019	9.54	0.001	0.12	8.65

Table E.12: Lime softening at 60 °C - flotation tails leach tests

Lime Softening at 60 °C - Flotation Tails Leach Tests						
Sample Number	Head grade	Washed Residue	Gold Dissolution	Gold in Filtrate	Gold Loaded Carbon	Gold Account-ability
	g/t	g/t	%	Mg/l	g/t	%
Test 1	0.150	0.075	50.0	0.002	3.00	101.2
Test 2	0.150	0.075	50.0	0.001	3.20	102.3
Test 3	0.150	0.081	45.8	0.001	2.80	100.5
Mean	0.150	0.077	48.6	0.001	3.00	101.8
Standard Deviation	0.000	0.004	2.41	0.001	0.20	0.91
95% Confidence	0.000	0.007	4.82	0.001	0.40	1.81

Table E.13: Lime- soda ash softening at 60 °C- straight leach tests

Lime- Soda Ash Softening at 60 °C- Straight Leach Tests						
Sample Number	Head grade	Washed Residue	Gold Dissolution	Gold in Filtrate	Gold Loaded Carbon	Gold Account-ability
	g/t	g/t	%	Mg/l	g/t	%
Test 1	0.200	0.124	37.9	<0.001	2.80	106.2
Test 2	0.200	0.097	51.4	<0.001	3.80	104.2
Test 3	0.200	0.110	44.8	<0.001	3.20	102.2
Mean	0.200	0.111	44.7	<0.001	3.27	104.2
Standard Deviation	0.000	0.013	6.75	<0.001	0.50	2.00
95% Confidence	0.000	0.027	13.49	<0.001	1.01	4.00

Table E.14: Lime-soda ash softening at 60 °C - flotation tails leach tests

Lime-Soda Ash Softening at 60 °C - Flotation Tails Leach Tests						
Sample Number	Head grade	Washed Residue	Gold Dissolution	Gold in Filtrate	Gold Loaded Carbon	Gold Account-ability
	g/t	g/t	%	Mg/l	g/t	%
Test 1	0.150	0.110	26.7	<0.001	1.80	103,3
Test 2	0.150	0.120	20.0	0.001	1.60	104,4
Test 3	0.150	0.110	26.7	0.001	1.80	104,8
Mean	0.150	0.113	24.4	0.001	1.73	104,1
Standard Deviation	0.000	0.006	3.85	0.001	0.12	2,08
95% Confidence	0.000	0.012	7.70	0.001	0.23	4,16

Table E.15: Lime softening at 90 °C- straight leach tests

Lime Softening at 90 °C- Straight Leach Tests						
Sample Number	Head grade	Washed Residue	Gold Dissolution	Gold in Filtrate	Gold Loaded Carbon	Gold Account-ability
	g/t	g/t	%	Mg/l	g/t	%
Test 1	0.200	0.100	50.0	0.001	3.00	94.5
Test 2	0.200	0.110	45.0	<0.001	3.30	103.7
Test 3	0.200	0.110	45.0	<0.001	3.00	99.8
Mean	0.200	0.107	46.7	<0.001	3.10	99.3
Standard Deviation	0.000	0.006	2.89	<0.001	0.17	4.62
95% Confidence	0.000	0.012	5.77	0.001	0.35	9.24

Table E.16: Lime softening at 90 °C - flotation tails leach tests

Lime Softening at 90 °C - Flotation Tails Leach Tests						
Sample Number	Head grade	Washed Residue	Gold Dissolution	Gold in Filtrate	Gold Loaded Carbon	Gold Account-ability
	g/t	g/t	%	Mg/l	g/t	%
Test 1	0.170	0.082	51.8	0.001	3.80	105.3
Test 2	0.170	0.094	44.7	0.001	3.40	101.8
Test 3	0.170	0.096	43.5	0.001	2.80	103.3
Mean	0.170	0.091	46.7	0.001	3.33	103.8
Standard Deviation	0.000	0.008	4.45	0.000	0.50	4.36
95% Confidence	0.000	0.015	8.91	0.000	1.01	8.72

Table E.17: Lime-soda ash softening at 90 °C- straight leach tests

Lime-Soda Ash Softening at 90 °C- Straight Leach Tests						
Sample Number	Head grade	Washed Residue	Gold Dissolution	Gold in Filtrate	Gold Loaded Carbon	Gold Account-ability
	g/t	g/t	%	Mg/l	g/t	%
Test 1	0.200	0.113	43.5	0.004	3.20	104.4
Test 2	0.200	0.113	43.5	0.003	3.40	106.2
Test 3	0.200	0.104	47.8	0.003	4.00	103.8
Mean	0.200	0.110	44.9	0.003	3.53	104.8
Standard Deviation	0.000	0.005	2.51	0.000	0.42	1.25
95% Confidence	0.000	0.010	5.02	0.001	0.83	2.50

Table E.18: Lime-soda ash softening at 90 °C - flotation tails leach tests

Lime-Soda Ash Softening at 90 °C - Flotation Tails Leach Tests						
Sample Number	Head grade	Washed Residue	Gold Dissolution	Gold in Filtrate	Gold Loaded Carbon	Gold Account-ability
	g/t	g/t	%	Mg/l	g/t	%
Test 1	0.180	0.110	38.9	0.001	2.42	100.2
Test 2	0.180	0.100	44.4	0.002	2.90	101.9
Test 3	0.180	0.090	50.0	0.002	3.30	102.8
Mean	0.180	0.100	44.4	0.002	2.87	101.6
Standard Deviation	0.000	0.010	5.56	0.000	0.44	1.32
95% Confidence	0.000	0.020	11.1	0.001	0.88	2.65

APPENDIX F – SAMPLE CALCULATIONS

The sample calculations for leach tests are provided below.

Gold mass balance

The flotation and leach set conditions with associated set-points are presented in Tables F.1 and F.2.

Table F.1: Leach test work conditions

	TARGET	Units
Solids	49.30	%
Leach time	7.0	hrs
Initial pH	5 to 8	
Final pH	10.50	
Pulp SG	1.45	
Dry SG	2.7	
L/s Ratio	1:1	
Volume of Sample	1.0	litres
Mass of dry solids/head	0.71	kg
Mass of water	0.74	kg
Cyanide dosage	0.35	kg/t
Lime Dosage	2.80	kg/t
Mass of Cyanide added	0.25	g
Mass of Lime added	2.0	g

Table F.2: Flotation test work conditions

	TARGET	Units
Cell Size	5.0	litres
Flotation SG	1.30	
Sample Mass	2.41	kg
pH Float	6 to 8	
Rougher Flotation Time	10	minutes
Rougher Mass Pull Target	4	%
SNPX	20	g/t
Senkol 5	10	g/t
Conditioning Time	5	minutes
Frother Senfroth XP200	12	g/t
Cleaning	No	

Gold is leached off the solids and reports to filtrate/solution as an aurocyanide complex. The addition of high activity carbon results in adsorption of the aurocyanide complex. Therefore, the gold mass balance accounts for the gold leached off solids, residual gold within the filtrate as well as gold leaded onto activated carbon.

For all leach tests, a leach SG of 1.45 was the set point which produces a slurry with 49.3 % dry solids. Therefore, in a 1 litre sample, the expected washed residue dry solids are approximately 715 g. The carbon dosage rate used in the tests were 20 g/L.

The overall gold balance is presented below.

$$\%\ Gold\ Accountability = \frac{A + B + C}{Total\ Gold\ in\ Feed\ (\mu g)} \times 100 \qquad (E.1)$$

Where:

A = Residual Gold on Solids (µg)
B = Gold in filtrate/solution (µg)
C= Gold loaded onto activated carbon (µg)

The calculations are presented for the leach test in which potable water was spiked with 50 mg/l of sulfates. The leach results are presented in Table F.3 below.

Table F.3: Leach results for sulfate spiked potable water

Leach Results for Sulfate spiked potable water						
Sample Name	Head grade	Washed Residue	Gold Dissolution	Gold in Filtrate	Gold Loaded Carbon	Gold Account-ability
	g/t	g/t	%	mg/l	g/t	%
50 mg/l Sulfate Addition	0.200	0.113	43.5	0.002	3.30	103.6

The calculation of variables A, B and C are presented below.

Residual Gold on solids

$$A\ (\mu g) = Washed\ Residue\left(\frac{g}{t}\right) \times Mass\ of\ solid\ sample\ (kg) = 0.113 \times \frac{715}{1000} = 0.081\ \mu g$$

Gold in filtrate/solution

$$B\ (\mu g) = Gold\ in\ filtrate\ (ppm) \times Mass\ of\ water\ (kg) = 0.002 \times 0.74 = 0.001\ \mu g$$

Gold Loaded onto activated carbon

$$C\ (\mu g) = Carbon\ gold\ loading\ (\frac{g}{t}) \times Mass\ of\ carbon\ (kg) = 3.30 \times \frac{20}{1000} = 0.066\ \mu g$$

The total gold in the feed was calculated as follows:

$$Feed(\mu g) = Head\ Grade\left(\frac{g}{t}\right) \times Mass\ of\ solid\ sample\ (kg) = 0.200 \times \frac{715}{1000} = 0.143\ \mu g$$

Therefore, the overall gold balance is calculated as:

$$\%\ Gold\ Accountability = \frac{A + B + C}{Total\ Gold\ in\ Feed\ (\mu g)} \times 100$$

$$= \frac{0.081 + 0.001 + 0.066}{0.143} \times 100$$

$$= \frac{0.148}{0.143} \times 100$$

$$= 103.6\ \%$$

Energy balance

In Chapter 4, the results showed that lime softening at elevated temperatures to be beneficial in gold recovery, however significant operating costs would be associated with the heating requirement for 60 Ml/day of process water. The energy balance equation utilised to calculate the energy requirement is presented below:

$$** Energy\ Requirement\ (KJ) = m \cdot Cp \cdot \Delta T \qquad (E.2)$$

Where:

m = mass of eluant

Cp = heat capacity of the eluant

ΔT = change in eluant temperature

** 3600 KJ is equivalent to 1 kWh

A summary of the energy balance calculation for heating process water from ambient conditions to 60 °C is summarized in Table F.4 below:

Table F.4: Daily water heating requirement

Daily Water Heating Requirement			
Initial Temperature		20	°C
Required Heating Temperature		60	°C
Cp (Heat Capacity)		4.18	J/g.°C
Volume		60 000	m3/day
Heating Requirement		10 046 400	MJ
Conversion: MJ → kWh		$\frac{MJ}{3600}$	
Energy/Heating Requirement	per day	2790	MWh
	monthly	83 700	MWh

Should electrical energy be used to achieve the energy requirement and based in the unit cost per kilowatt as illustrated in Chapter 5, Table 5.3, the estimated monthly costs are presented in Table F.5.

Table F.5: Monthly heating costs to heat process water from ambient to 60 °C

Monthly Heating/ Electricity Costs			
	Unit Cost (R/kWh)	Quantity (kWh)	Total
Water Heating Requirement	R1,00	83 700 000	R83 700 000,00

From Table F.5, it is evident that the heating of the process water would not be economically viable.

Neutralisation Reactor Design Calculations

Following the design outlined by van Niekerk (2002), a total of three neutralisation reactors were chosen, each with a design retention time of 30 minutes. Thus, the volume of each reactor was calculated as follows:

$$Total\ Volumetric\ Flowrate\ of\ Process\ Water = 20\frac{Ml}{day} \times \frac{1000}{24\ \times 60\ \times 60} = 0.23\frac{m^3}{s}$$

$$Volume\ of\ Neutralisation\ Reactor = \frac{Retention\ Time\ \times Volumetric\ Flowrate}{Number\ of\ Reactors}$$

$$Volume\ of\ Neutralisation\ Reactor = \frac{30\ minutes \times 0.23\frac{m^3}{s} \times 60}{3}$$

$$Volume\ of\ Neutralisation\ Reactor = 138.9\ m^3 \approx 140\ m^3$$

Neutralisation Settler/Thickener Design Calculations

Following the design outlined by van Niekerk (2002), one neutralisation settler/thickener was chosen, with a design retention time of 30 minutes and up-flow velocity of 0.7 m/hr Thus, the volume of the settler/ thickener was calculated as follows:

$$Total\ Volumetric\ Flowrate\ of\ Process\ Water = 20\frac{Ml}{day} \times \frac{1000}{24 \times 60 \times 60} = 0.23\frac{m^3}{s}$$

$$Volume\ of\ Neutralisation\ Settler = \frac{Retention\ Time \times Volumetric\ Flowrate}{Number\ of\ Settlers}$$

$$Volume\ of\ Neutralisation\ Reactor = \frac{30\ minutes \times 0.23\frac{m^3}{s} \times 60}{1}$$

$$Volume\ of\ Neutralisation\ Reactor = 416.7\ m^3 \approx 417\ m^3$$

Lime Design Dosage and Consumption

From Section 5.1.4., the required slaked lime dosage was established as 1.0 kg/m^3 mine water. Therefore, the monthly consumption is calculated as follows:

$$Lime\ Cosumption = Lime\ Dosage \times Volumetric\ Flowrate$$

$$Lime\ Cosumption = 1.0\frac{kg}{m^3} \times 20\frac{Ml}{day} \times \frac{1000 \times 30}{1000}$$

$$Lime\ Cosumption = 600\frac{ton}{month}$$

APPENDIX G - PROCESS FLOW DIAGRAMS

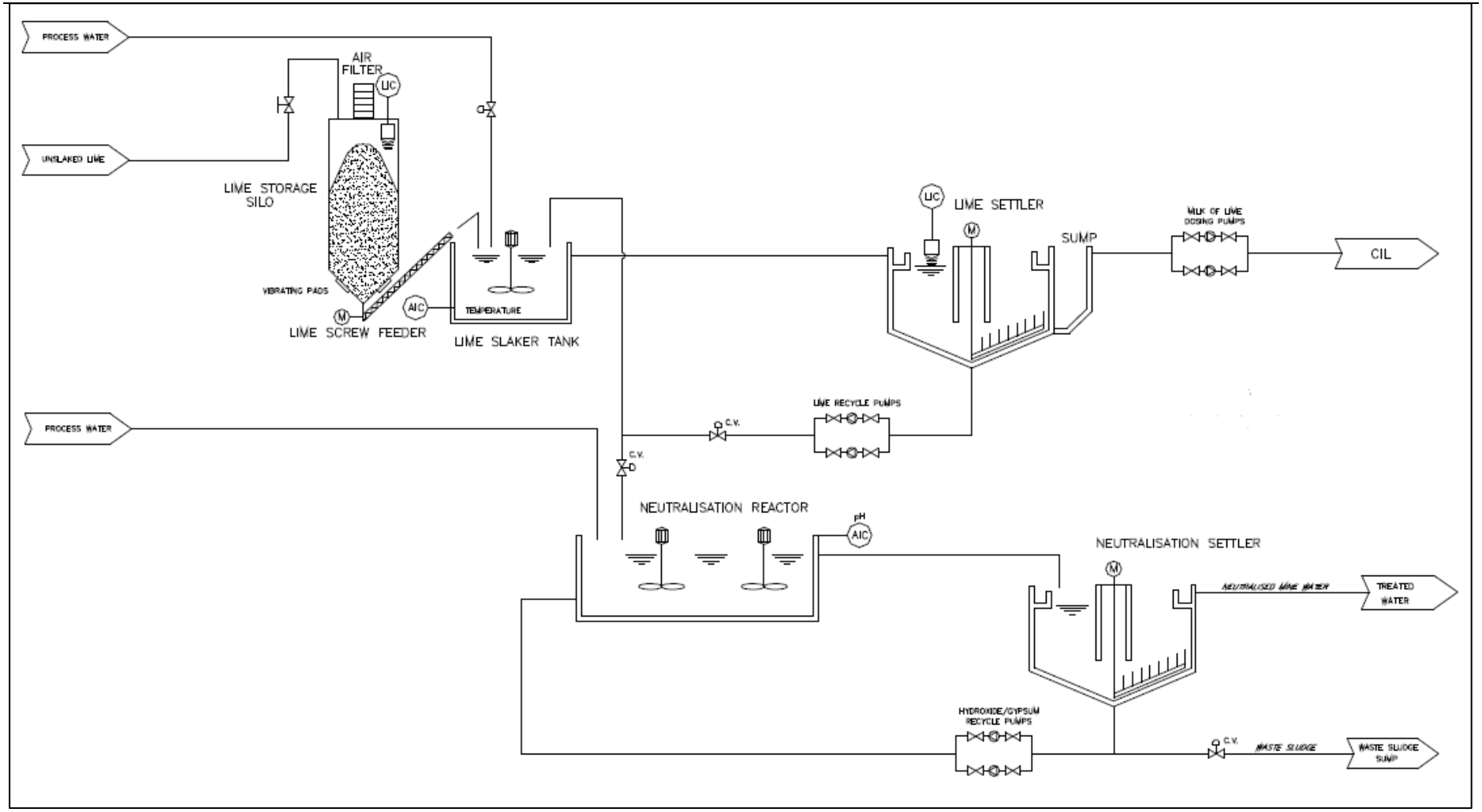


Figure G.1: Neutralisation, lime storage, make- up and dosing process flow diagram (van Niekerk, 2002)

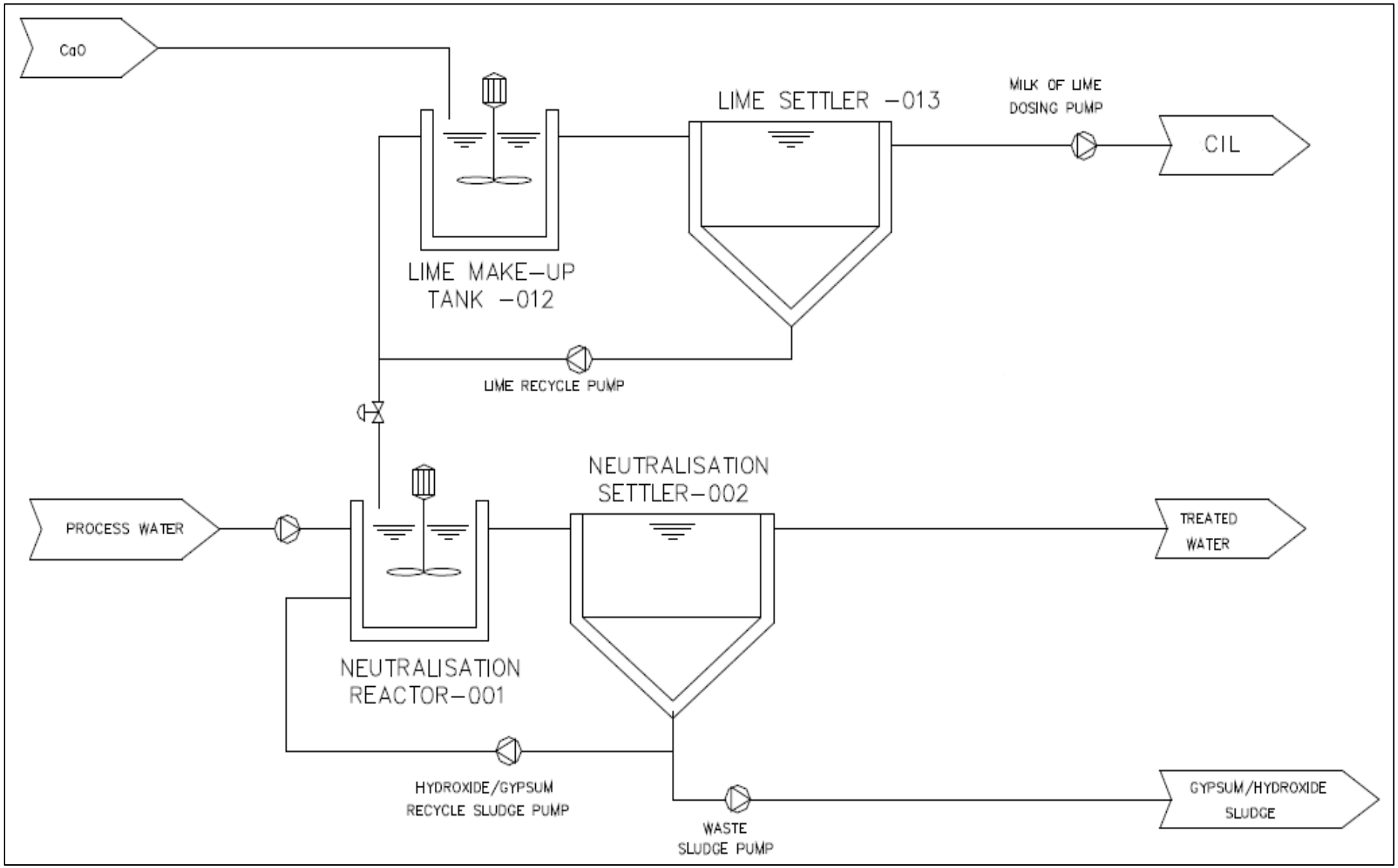


Figure G.2: Proposed treatment process flow diagram (van Niekerk, 2002)

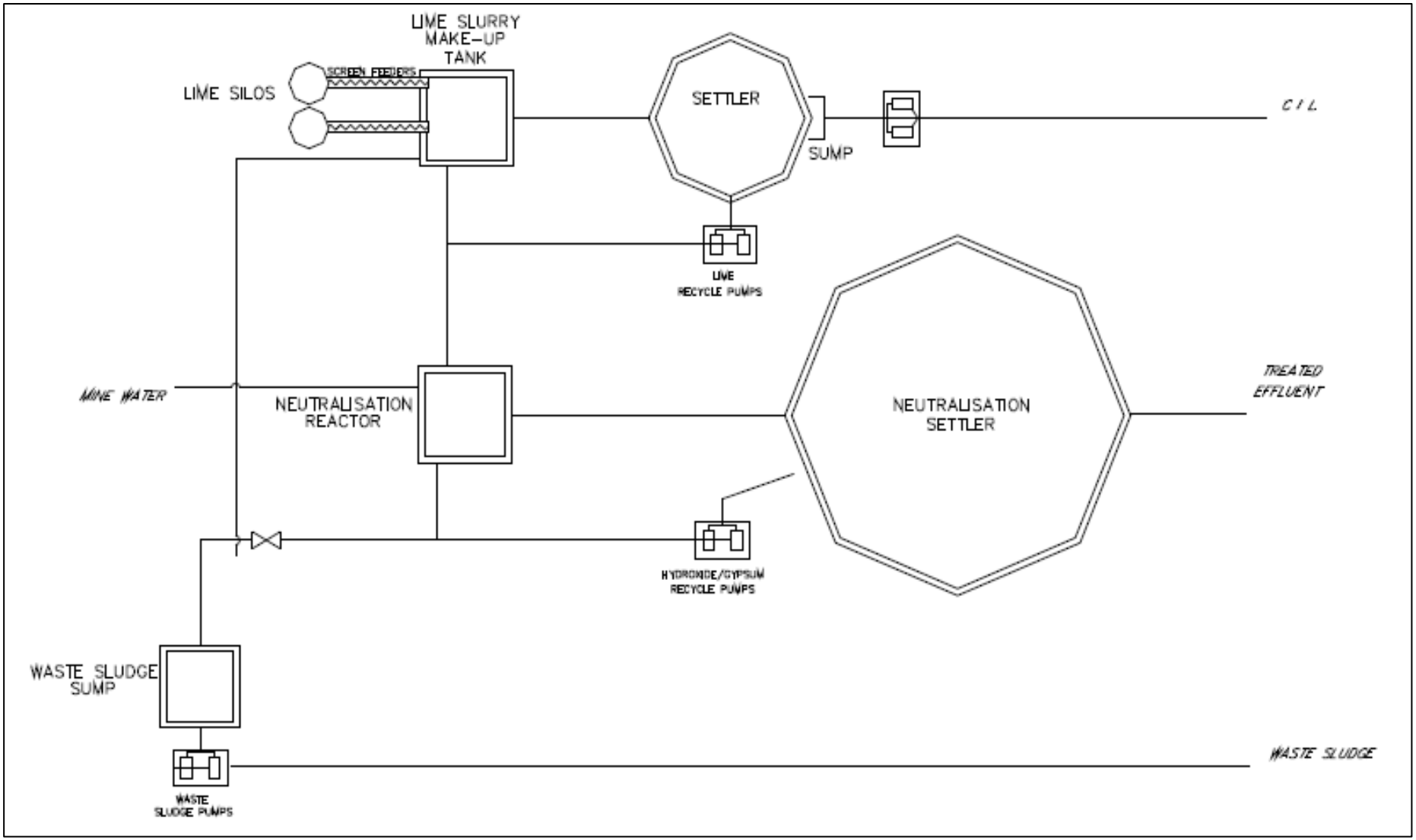


Figure G.3: The proposed plant layout (van Niekerk, 2002)