

---

# Techno-economical comparison of three process routes for the treatment of Gamsberg zinc ore



By

Zethu Dlamini

**Supervisor**

A/ Professor Jochen Petersen

Submitted in partial fulfillment for the award of the degree of Master of Science in Chemical Engineering

February 2015



The copyright of this thesis vests in the author. No quotation from it or information derived from it is to be published without full acknowledgement of the source. The thesis is to be used for private study or non-commercial research purposes only.

Published by the University of Cape Town (UCT) in terms of the non-exclusive license granted to UCT by the author.

## **Acknowledgements**

First and foremost, I thank God who made it possible for me to finish this study. It wasn't by my might nor by my power, it is all because of his Grace.

Special thanks go to my supervisor, Associate Professor Jochen Petersen, for his valuable input and sustained commitment throughout the duration of this project. Most importantly, for realising my skills and placing me in a project aligned with my passion. I am also grateful to Professor Jean Paul Franzidis for funding this project through the Minerals to Metals Initiative.

I would like to thank the CeBER BM2 team for their constructive criticism and fruitful discussion in all the presentations I had for this study.

I am also grateful to Charles Kohler, Jonathan Ledgerwood, Diane Taggart, Richard Beck and Professor Mike Nicol for the technical advice they brought into the project. To my friends and family, thank you for your support and encouragement.

To Hlayisi Baloyi and Jemitias Chivavava, thank you for being my first correctors for this document. I appreciate the time you spent reading through each and every page of this document.

Christopher Chikochi, Thobile Dube and Excellent Gwebu, thank you for your support, kindness and most importantly, for facilitating the submission of this work.

To Salebona Shongwe, thank you for always availing yourself when I needed encouragement and support through this MSc journey. You have dedicated your time and energy like I was part of your family. I am humbled by such dedication.

I dedicate this work to my parents, Bheka and Thuli Dlamini, for their continued support and motivation throughout the completion of my degree. I'm indebted to them for always believing in me and expect what I didn't even know I could do.

## Declaration

1. I know that plagiarism is wrong. Plagiarism is to use another's work and pretend that it is one's own.
2. I have used the Harvard-UCT (Author-Date).2014 convention for citation and referencing. Each contribution to, and quotation in, this report from the works of other people has been attributed, and has been cited and referenced.
3. This report is my own work.
4. I have not allowed, and will not allow, anyone to copy my work with the intention of passing it off as his or her own work.

Signature

Signed by candidate

Date

25 May 2015

## Nomenclature

### List of abbreviations

BT	Bleed treatment
BZS	Basic zinc sulphate
CI	Cost index
D	Diameter
Di	Inside diameter
DCF	Discounted cash flow
EW	Electrowinning
FCI	Fixed Capital Investment
ft	Feet
H	Height
HAL	High acid leach
HL	Heap leaching
IRR	Internal Rate of Return
kPa	Kilopascal
L	Length
LAL	Low acid leach
lb	Pound
LME	London Metal Exchange
$M_{ar}$	Minimum acceptable rate of return
NPV	Net Present Value
NT	Neutralization
PBP	Payback period
PLS	Pregnant leach solution
PVR	Present value ratio
RLE	Roast-leach-electrowinning
ROI	Return on investment
Sc	Scrubbing
SHG	Special high grade
SX	Solvent extraction

TCI	Total capital investment
tpa	Tonne per annum
US\$	United States dollar
USGS	United States Geological Survey
WC	Working Capital

### **List of elements and compounds**

Ca	Calcium
Cd	Cadmium
Co	Cobalt
Cu	Copper
D2EHPA	Di(2-ethylhexyl) phosphoric
Fe	Iron
H <sub>2</sub>	Hydrogen gas
H <sub>2</sub> O <sub>2</sub>	Hydrogen peroxide
H <sub>2</sub> SO <sub>4</sub>	Sulphuric acid
HCl	Hydrochloric acid
Mg	Magnesium
Mn	manganese
Pb	Lead
S	Sulphur
SiO <sub>2</sub>	Silicon dioxide
SO <sub>2</sub>	Sulphur dioxide
Zn	Zinc
ZnS	Sphalerite

## **Abstract**

There is an abundant availability of zinc sulphide (sphalerite) ore (160 million tons at 7.40 % Zn) at Gamsberg, Northern Cape in South Africa. The ore body is South Africa's greatest unexploited base metal resource. Regardless of its size, the low zinc and high manganese content of the sphalerite combined with the low zinc price prohibits the development of the deposit. Sphalerite is the most common zinc mineral, hence 95 % of the world's zinc production is from this mineral. Sphalerite is currently processed by crushing-milling-flotation, followed by the roast-leach-electrowinning (RLE) process. This route has major detrimental impacts on the environment, it produces SO<sub>2</sub>, and cannot treat ores of low grade or higher complexity. Therefore, alternative processes are being sought in order to circumvent the RLE process.

This study compares three different process routes in the context of processing ore from the Gamsberg deposit for refining 3.4 million tpa ore in order to produce special high grade (SHG) zinc (>99.995% Zn). These routes include heap leaching and refining locally (route 1), preparing a flotation concentrate and refining it locally (route 2) and lastly, preparing a flotation concentrate and shipping it for toll refining in Europe (route 3). Zinc heap leaching has not yet been commercialised due to the absence of solvent extraction reagents which can selectively extract zinc from a low tenor acidic pregnant leach solution without incorporating the neutralization stage. Therefore, route 1 has higher risk as compared to the other routes. A desktop model which provides a comparison of capital cost, operating cost, NPV, IRR and PVR has been developed. Parameters such as average zinc grade, process recovery and zinc price are provided as inputs. The effects of fluctuations in important parameters such as working capital, and zinc price on NPV are assessed using Matlab.

The total capital investment is 12.4, 15.0 and 7.55 billion Rand for routes 1, 2 and 3, respectively. The corresponding NPVs are: 1.57, 0.23 and 1.09 billion Rand. Milling and electrowinning stages are the major contributors to capital and operating costs, hence route 1 and 2 have high capital and operating cost. The NPVs are more sensitive to the computed discount rate and fixed capital, and to a lesser extent the construction time. Low capital and operating costs, low environmental impact and maximum return to investors are key parameters for investment decision in the mining industry. Therefore, if there is no capital rationing and there is

enough energy to supply a green-field project, route 1 is the most attractive route for Gamsberg. If not, route 3 is the better choice.

## Table of Contents

Acknowledgements.....	ii
Declaration.....	iii
Nomenclature.....	iv
List of abbreviations.....	iv
List of elements and compounds.....	v
Abstract.....	vi
List of tables.....	xi
List of figures.....	xi
1 Introduction.....	1
1.1 Background and formulation of research problem.....	1
1.2 Aim.....	5
1.3 Research questions.....	6
2 Literature review.....	7
2.1 Gamsberg ore deposits.....	7
2.1.1 Deposit geology.....	7
2.1.2 Effect of ore mineralogy on zinc processing.....	10
2.2 Zinc processing.....	10
2.2.1 Chloride media.....	11
2.2.2 Sulphate media.....	11
2.3 Process routes being studied.....	12
2.3.1 Production of zinc concentrates.....	12
2.3.2 Options for treating zinc concentrates.....	13

2.3.3	The HydroZinc™ Process.....	17
2.4	Heap leaching.....	19
2.4.1	Use of heap leaching process.....	21
2.4.2	Efficiency factors in heap leach process.....	22
2.4.3	Heap bioleaching .....	23
2.4.4	Heap design.....	27
2.5	Design of tailings dams.....	32
2.6	Solution purification.....	34
2.6.1	Iron precipitation from leach solutions.....	34
2.6.2	Comparison of the various iron removal processes .....	38
2.6.3	Solvent extraction (SX) process.....	40
2.7	Metal recovery.....	46
2.7.1	Zinc Electrowinning (EW).....	46
2.8	Economics of zinc production.....	49
2.8.1	Market.....	49
2.8.2	Zinc selling price.....	51
2.8.3	Cost of innovation.....	53
2.8.4	Concentrate marketing.....	54
2.9	Assessing Profitability of Investments.....	57
2.9.1	Discounted profitability criteria.....	57
2.10	Summary .....	59
3	Methodology.....	60
3.1	Process routes.....	60
3.2	Flowsheet development and material balances .....	61
3.2.1	Major assumptions.....	61

3.2.2	Route 1 assumptions .....	61
3.2.3	Route 2 assumptions .....	64
3.3	Capital cost estimates .....	64
3.3.1	Heap .....	64
3.3.2	Concentrator plant.....	65
3.3.3	Refinery.....	65
3.3.4	Total fixed cost .....	66
3.4	Operating cost estimates.....	66
3.5	Financial analysis .....	66
4	Results and discussion.....	68
4.1	Capital and operating cost estimates .....	68
4.1.1	Capital costs .....	68
4.1.2	Operating costs.....	70
4.2	Ranking the process options.....	72
4.2.1	Resolving contradictions between IRR and NPV rankings .....	74
4.2.2	Shareholder value.....	75
4.3	Energy crisis in South Africa .....	76
4.4	Sensitivity analysis.....	77
4.4.1	Zinc Price .....	79
4.4.2	Fixed capital investment .....	80
4.4.3	Construction time.....	81
4.4.4	Working Capital.....	82
4.4.5	Electricity Price.....	83
4.4.6	Discount rates.....	84
4.5	Gamsberg-Skorpion Integrated Zinc Project.....	86

5	Conclusions and Recommendations .....	88
5.1	Conclusions .....	88
5.2	Recommendations .....	89
6	References .....	91

## List of tables

Table 1: Zinc and manganese concentrations per ore type (Schouwstra et al., 2010).....	8
Table 2: Advantages and Disadvantages of heap leaching (Ghorbani et al., 2011) .....	21
Table 3: Factors and parameters affecting heap bioleaching and metal recovery (Pradhan et al., 2008) .....	24
Table 4: Advantages and disadvantages of multi-lift and on/ off heaps.....	27
Table 5: Scaling up of laboratory results for full-scale heaps (John, 2011) .....	28
Table 6: Comparison of Embankment Types (Vick, 1990).....	34
Table 7: Comparison between the different iron removal processes (Claassen, 2002) .....	39
Table 8: Representative values for advanced electrolyte (Cole & Sole, 2003) .....	48
Table 9: World’s 10 biggest countries in refined zinc production (Zinc Investing News, 2013a; Zinc Investing News, 2014).....	49
Table 10: Top zinc miners by 2010 (PwC & MCX, 2011).....	51
Table 11: Top refined zinc producers by 2010 (PwC & MCX, 2011) .....	51
Table 12: Zinc mine depletions (and total attritions) between 2012-2016 (Barbir, 2012) .....	53
Table 13: Risk profiles for different investment types (Cohen, 2009) .....	58
Table 14: Parameters considered for cost analysis .....	67
Table 15: Rankings for the three routes .....	73
Table 16: Comparison of route 1 and 3 .....	75
Table 17: Range of variation of Factors Affecting the profitability of a Chemical Process (Turton et al., 2008) .....	78

## List of figures

Figure 1: End use of zinc (International Zinc Association, 2011c) .....	2
Figure 2: Location and geology of Gamsberg deposit (Schouwstra et al., 2010).....	7
Figure 3: Generalized block flow diagram for producing concentrates (Davenport et al., 2002a) .....	13
Figure 4: Process routes to treat concentrates.....	14
Figure 5: RLE process flow diagram. NL (Neutral leach), HAL (Hot Acid Leach) (Fuls, 2011) .....	15
Figure 6: Stand-alone atmospheric process (Fuls, 2011).....	17

Figure 7: The Teck Comico HydroZinc™ process (Lizama et al., 2003) .....	18
Figure 8: a) Heap slope section with typical angle-of-repose ore lift slopes and bench setbacks for overall flatter slope b) Simplified schematic cross-section of a heap leach facility (Ulrich, Andrade & Gardner, 2003) .....	29
Figure 9: Double Composite Liner (Lupo & Morrison, 2007) .....	31
Figure 10: Phase diagram for iron oxides precipitation from 0.5M ferric sulphate solutions. Hydroxy salts = basic iron sulphates e.g. Jarosites. Fe(OH) <sub>3</sub> = iron oxy-hydroxide(s) (Claassen et al., 2002). .....	35
Figure 11: General Solvent Extraction Circuit (Ritcey & Ashbrook, 1984) .....	40
Figure 12: Simplified flowsheet for zinc electrowinning (Nicol et al., 2008).....	46
Figure 13: Zinc prices for the period 1989-2014 (IndexMundi, 2014).....	52
Figure 14: Process routes being studied. Solvent Extraction (SX) and Electrowinning (EW).....	60
Figure 15: Capital costs for the three process routes. EW (Electrowinning), BT (Bleed treatment), SX (Solvent extraction) and NT (Neutralization) .....	69
Figure 16: Operating costs for the three process routes. EW (Electrowinning), BT (Bleed treatment), SX (Solvent extraction) and NT (Neutralization) .....	71
Figure 17: Sensitivity of NPV to zinc price.....	79
Figure 18: Sensitivity of NPV to Fixed Capital (FC) Cost.....	80
Figure 19: Sensitivity of NPV to construction time.....	81
Figure 20: Sensitivity of NPV to Working Capital.....	82
Figure 21: Sensitivity of NPV and IRR to the Electricity Price .....	83
Figure 22: Sensitivity of NPV to discount rate.....	84
Figure 23: Zoomed in Sensitivity of NPV to discount rate in order to display crossover values clearly.....	85

## CHAPTER 1

### 1 Introduction

#### 1.1 Background and formulation of research problem

Zinc is a natural component of the earth's crust and environment. It is found in the rocks, soil, water and biosphere. It is the 23<sup>rd</sup> most abundant element in the earth's crust and is the fourth most used metal after iron, aluminum and copper (CPRA, 2007). Zinc ore deposits are widely spread throughout the world and the ore is extracted in more than 50 countries. China, Australia, and Peru are the three largest zinc producers with Australia having the highest reserves (Zinc Investing News, 2014). According to the International Zinc Association (2011a), there are four major zinc deposits types which are:

- **Volcanic hosted sulphides (VMS):** these deposits are polymetallic and they are main sources of copper and zinc. They are associated with significant amounts of silver, gold, cadmium, bismuth or tin.
- **Carbonate hosted:** these deposits have limestone and dolomite as the most common host rocks. The amount of zinc-lead in the ore ranges between 5 % and 10 % with zinc being predominant over lead. Copper, silver and barite or fluorite may also be present in the ore.
- **Sediment hosted (sedex deposits):** shale, siltstone and sandstone are the most common host rocks for the deposits. Some of the world's largest sources of zinc, lead and silver are from this type of deposit. The zinc-lead content ranges between 10 % and 20 %.
- **Intrusion related:** these deposits are found in carbonate rocks in conjunction with magmatic-hydrothermal systems and they are main sources of lead and zinc with lead being predominant. They are associated with calcium, silver and magnesium.

Zinc mining is mostly done underground. In 2011, underground mining accounted for 80%, open pit mines accounted for 8 % and the combination of underground and open pit accounted for 12 % of existing mines. However, in terms of volumes, open pit mines accounted for 15 %, underground accounted for 64 % and the combination of underground and open pit accounted for 21 % (International Zinc Association, 2011b).

Minerals associated with zinc include sphalerite (zinc sulphide), smithsonite (zinc carbonate), hemimorphite (zinc silicate) and franklinite (a zinc spinel) (CPRA, 2007; BCS, 2002). The most common zinc mineral is sphalerite, hence 95 % of the world's zinc production is from this mineral (International Zinc Association, 2011a). The most common metals associated with zinc are: lead, copper, iron, gold and silver.

Zinc plays a major role in a wide range of industrial and consumer products that keep our society moving. The annual worldwide zinc mine production was 13.5 million metric tons by 2013 (Zinc Investing News, 2014). More than half of the annual global production is used for galvanizing steel. Galvanized steel is ideal for construction since the zinc layer protects the steel from rusting and corroding (Burgess, 2006). Approximately 14 % goes into the production of zinc base alloys, mainly to supply the die casting industry and 10 % to produce brass and bronze (International Zinc Association, 2011c). Significant amounts are also utilized in rolled zinc applications including roofing, gutters and down-pipes. The remainder is consumed in compounds such as zinc oxide and zinc sulfate. Zinc oxide is widely used in cosmetics. The above mentioned end uses have been shown in Figure 1.

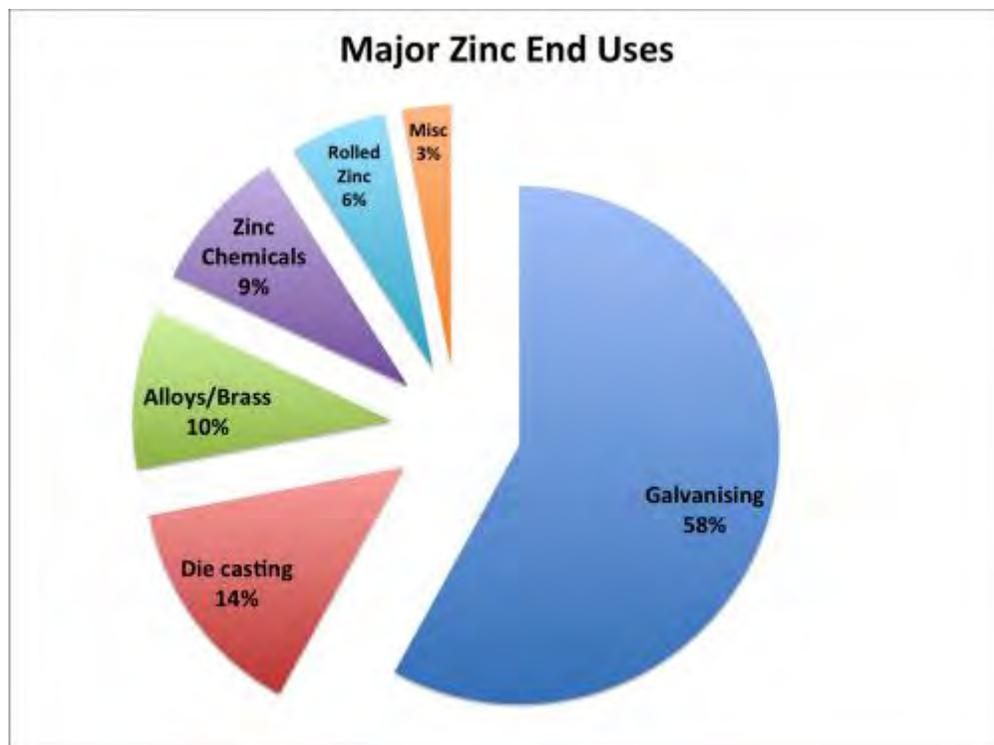


Figure 1: End use of zinc (International Zinc Association, 2011c)

Developing countries are the key drivers in zinc consumption (Williams, 2013). China is the largest world's zinc consumer; this high demand is due to the construction industry and increased motor vehicle production in China. Throughout the nineties, China has been a major zinc supplier but due to the country's great demand for galvanized metal it has now become a net zinc importer (Burgess, 2006). The zinc demand is mainly driven by China's economic affluence and USA's large economy (CPRA, 2007).

Metallic zinc can be recovered either through hydrometallurgical or pyrometallurgical processes. The major difference between these two processes is that the former produces pure zinc directly whereas the latter produces lower grade zinc which contains significant impurities that have to be removed by thermal refining (CPRA, 2007). As a result, 85% of primary zinc production is from hydrometallurgical processes (de Souza et al., 2007; Harvey et al., 2002). The traditional route for processing zinc sulphide ore is by crushing-milling-flotation followed by roast-leach-electrowinning (RLE) process (Babu et al., 2002). The RLE process has a major negative impact on the environment; it produces SO<sub>2</sub> gas during the roasting step. Also, it cannot treat ores of low grade or higher complexity. The SO<sub>2</sub> gas produced can be converted to sulphuric acid; however this is only economically viable when a market for sulphuric acid exists close to site.

Due to great pressure exerted on companies to be environmentally friendly and the fact that ore reserves are not only diminishing but being of lower grade and higher complexity, purely hydrometallurgical processes are being sought in order to circumvent the RLE route. Several processes such as bioleaching and chemical leaching as an integrated process, persulphate leaching, heap bioleaching, sulphide and silicate leaching integration have been proposed in order to avoid producing SO<sub>2</sub> (de Souza et al., 2007). From both economic and environmental points of view, zinc processes that exclude the roasting step are more feasible. Purely hydrometallurgical processes avoid the roasting step by directly dissolving zinc from the sulphide ore followed by purification and electrolysis.

Since the 90s, the zinc industry has had an oversupply of metal and concentrate; this has led to zinc prices being low. Zinc has been trading below US\$1.00/lb except for the brief period from 2005 to 2007 (Williams, 2013). Only processes with low costs are sustainable under this economic condition. Due to the eroding profitability, new technologies have been sought in order to sustain green field zinc projects in the industry since conventional processes can only sustain

brown field projects (Lizama et al., 2003). There has been a lack of zinc exploration and there is a scarcity of capital in the zinc sector due to the low prices of zinc. On the other hand, leading zinc companies are expected to close down due to reserve exhaustion between 2014 and 2016 and some will experience a fall in production (Environmental Resources Management, n.d.b; Williams, 2013). This will cause deterioration of zinc supply, yet there is a forecasted zinc demand growth at a rate of 3.7 % per annum over the period 2015-2025 with the global demand exceeding production by 503kt/a by year 2015 (Wood Mackenzie, 2012). The increase in zinc demand coupled with reduced supply will increase the prices in the future. The lack of exploration and current financing will impact zinc production and guarantee zinc receives significant focus in the future.

Black Mountain Mining (Pty) Ltd (herein referred to as Black Mountain), part of the global Vedanta mining group, intends to establish an open pit zinc mine in the Namakwa District, Northern Cape Province. The proposed site is Gamsberg and is characterized by an oval shaped inselberg 220 metres above the surrounding plains (Environmental Resources Management, n.d.a). The mine will have an output capacity of approximately 10 million tpa (beneficiation tonnage) and a life span of approximately 20 years (Environmental Resources Management, n.d.a). At its peak time, the company is expected to introduce approximately 500 000 tpa of zinc into the global market (Environmental Resources Management, n.d.b). The proposed project will produce zinc concentrate which will be exported through the port of Saldanha to Europe and Asia for refining and distribution. The presence of existing transportation infrastructure provides a strategic advantage and market viability for Black Mountain, while limiting the footprint of the proposed development (Environmental Resources Management, n.d.b).

Black Mountain operates a mine in Aggeneys, the volume of ore extracted comprises of zinc (64 700 tpa), lead (74 600 tpa), copper (10 200 tpa) and silver (5 400 tpa) (Environmental Resources Management, n.d.b). The ore body at Gamsberg is between 100 to 500 m in depth and has a lateral extent of 3500 m (Environmental Resources Management, n.d.b). The major minerals for this ore body include sphalerite, pyrite, pyrrhotite, silicate and quartz with traces of chalcopyrite and galena (Ghorbani, 2012). 60 000 tpa ore is being mined at Gamsberg and transferred to a concentrator plant in Aggeneys where it is then processed together with ore from the Black Mountain Deeps Mine (Environmental Resources Management, n.d.b).

Different hydrometallurgical processes to treat sphalerite exist; the study done by Fuls (2011) identified these processes. Most sphalerite processes involve concentrate leaching compared to heap leaching. The three most favourable concentrate leaching processes are: Atmospheric leach, Dynatec pressure leach and Albion leach. These processes have favourable economics (low capital and operating cost), low technical risk (processes have been commercialized), low safety, health and environment impact and low energy and water consumption (Fuls, 2011).

Not considered in Ful's study was the HydroZinc™ process which recovers zinc from sulphide ore through heap leaching, neutralization, solvent extraction and electrowinning (Filippou, 2004; Lizama et al., 2003). A previous study done by Lizama et al., (2003) concluded that this process could produce zinc which is more competitive than the operating cost of all conventional processes. Regardless of zinc heap leaching being practicable with recoveries of over 80 % achieved within a leach time of 730 days in pilot test work at Teck Cominco (Lizama et al., 2003); it has not been commercialized yet. This is not due to heap leaching issues but it is because there are no solvent extraction (SX) reagents which can selectively extract zinc from a low tenor acidic pregnant leach solution (PLS) without any neutralization (John, 2011; Cole & Sole, 2003). Although processes that include heap leaching are cost-effective, the additional neutralization step has a major impact on the economics of the HydroZinc™ process.

Profitability assessments for a project can take a number of different forms, e.g. time-based, value-based, and rate-based. Time-based assessment will determine how fast the investment will be paid back, value-based assessment will determine how much the investment will grow and rate-based assessment will determine the yearly return as a percentage of the investment. Before any project can be implemented, the project undergoes a feasibility study phase to evaluate the economics of the project and to determine whether or not the proposed project meets investment guidelines. If the result is positive, the project goes to construction and operation.

## **1.2 Aim**

The objective of the study is to do a techno-economical comparison of three different process routes in the context of Gamsberg for refining 3.4 million tpa ore in order to produce special high grade (SHG) zinc (>99.995% Zn). These routes include preparing a flotation concentrate and shipping it for toll refining in Europe, preparing a concentrate and refining it locally and lastly, heap leaching and refining locally. This study aims to developing a model which generates

information pertaining to the feasibility of these routes and thereafter identify the key parameters significant to the feasibility of these routes (sensitivity analysis). Research questions have been formulated in order to compare these routes; these have been listed in the section below.

### **1.3 Research questions**

- At what conditions does the HydroZinc™ process become more attractive over the other process routes?
- At what conditions does the use of a particular route become justifiable over the other process routes in the context of Gamsberg?
- How sensitive are the processes to the following variables?
  - Zinc and electricity price
  - Fixed capital investment
  - Construction time
  - Working capital

## CHAPTER 2

### 2 Literature review

#### 2.1 Gamsberg ore deposits

##### 2.1.1 Deposit geology

The Gamsberg Zn-Pb deposit (160 million tons at 7.40 % Zn, 0.55 % Pb) is one of the four Broken Hill-type deposits (Black Mountain, Big Syncline, Broken Hill and Gamsberg) that are located in the Central North Western Cape, South Africa (Stalder & Rozendaal, 2004). It is located 16 km to the east of Broken Hill (Porter GeoConsultancy, 2001). This deposit is shown in Figure 2 below.

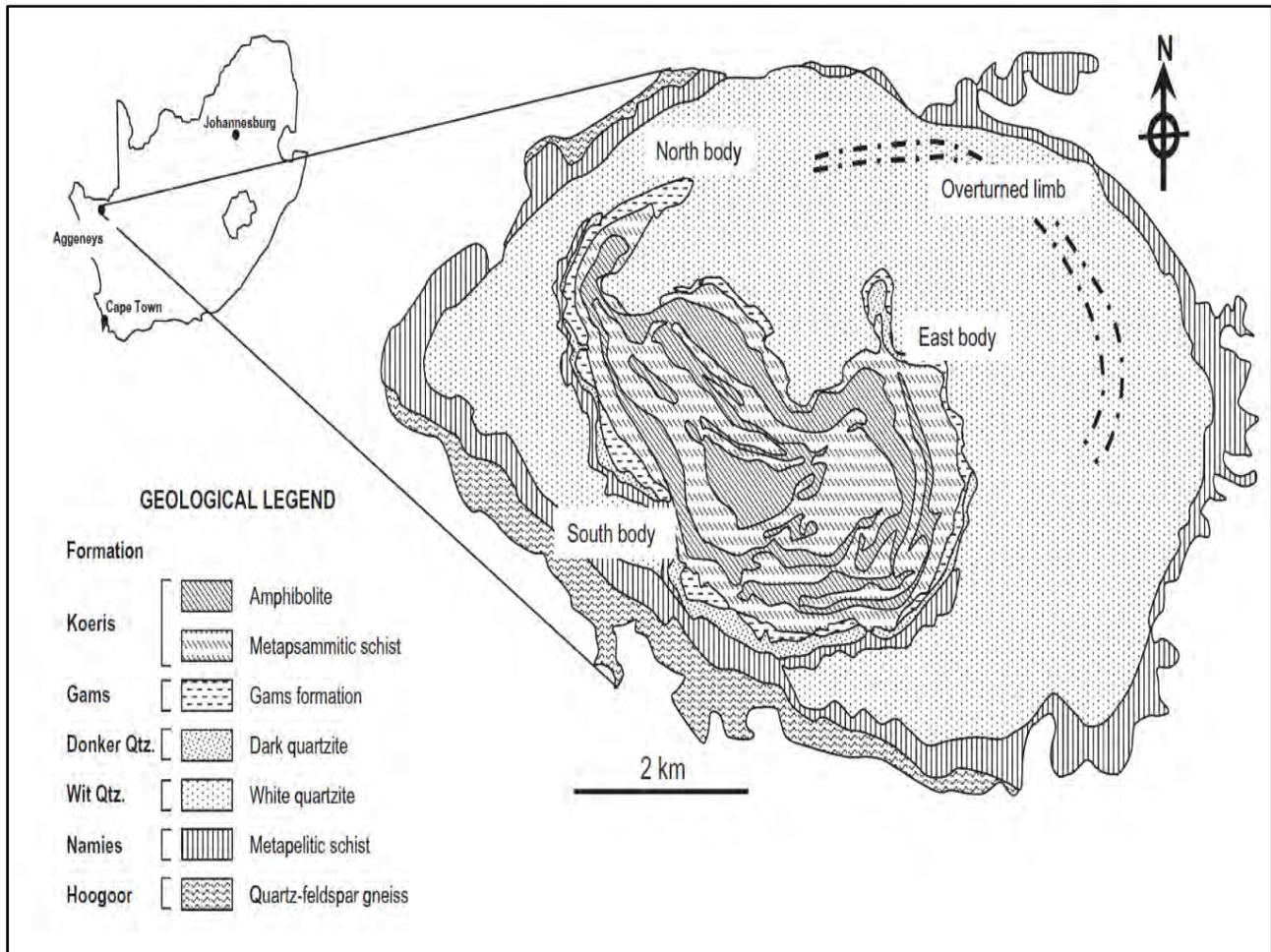


Figure 2: Location and geology of Gamsberg deposit (Schouwstra et al., 2010)

The Broken Hill deposits form part of the Bushmanland Group which is a volcano-sedimentary sequence that hosts ore deposits (Reid et al., 1997). This group has metapelitic rocks that are Ca-Na-Mg depleted. The group's ore body is divided into two types: a pyrrhotite-galena-sphalerite-chalcopryrite ore hosted by quartz-garnet-rich gangue (Black Mountain and Broken Hill) and a pyrite-pyrrhotite-sphalerite ore hosted by quartz-biotite-sillimanite gangue (Big Syncline and Gamsberg). The principal ore minerals are sphalerite, galena and chalcopryrite with garnet and apatite being minor constituents. Barite horizons occur as lateral equivalents of the sulphide ores at Black Mountain and Gamsberg and calcitic marble bands are associated with the Big Syncline and Gamsberg (Moore et al., 1990).

Despite the discovery of Gamsberg zinc deposit in 1972, the ore body continues to be South Africa's greatest unexploited base metal resource. Regardless of its size, the low zinc and high manganese content of the sphalerite combined with the low zinc price prohibits the development of the deposit. The high manganese content of the sphalerite makes it unsuitable for processing by conventional methods i.e. roasting, leaching/ solvent extraction and/or electrowinning (McClung & Viljoen, 2011). The ore body has two distinct ore types, namely pelitic and garnet-magnetite ore. The zinc and manganese concentrations for these ore bodies are shown in Table 1. The pelitic ore is characterized by a high Fe-sulphide content whilst the garnet-magnetite ore have higher levels of garnet, FeMn-silicates and magnetite (Schouwstra et al., 2010).

**Table 1: Zinc and manganese concentrations per ore type (Schouwstra et al., 2010)**

	wt % Zinc			wt % Manganese		
	Minimum	Maximum	Average	Minimum	Maximum	Average
<b>Garnet-magnetite</b>	3.7	14.9	9.6	1.1	9.5	6.9
<b>Pelitic</b>	4.7	15.8	8.3	1.1	4.7	1.9

The economic value of the deposit is dependent on the mineralogical characteristics and associations present in the ore assemblage, i.e. the loss of sphalerite to tailings in flotation circuits due to inadequate particle separation reduces the economic viability of the process option. Given the low grade of the ore and the sensitivity to manganese level in concentrates, the pelitic ore is the more attractive option to process hence this ore type is discussed in detail below.

Mineralization for this ore reserves occurs within the Gams formation and is hosted by a thin, less than 1 km thick deformed and metamorphosed, continental margin volcano-sedimentary rocks (McClung & Viljoen, 2011). According to Porter GeoConsultancy (2001), the Gams ore comprises three units, namely:

- A lower member composed of a diverse suite of quartz-felspar-garnet-clinopyroxene rock, garnet-clinopyroxene-feldspar marble and garnet-clinopyroxene-quartz-magnetite rocks.
- A middle sulphide zone with quartz-garnet-amphibole rocks and graphitic quartz-sillimanite-muscovite-feldspar containing major amounts of pyrite, pyrrhotite, sphalerite and galena.
- An upper unit of diverse garnet, pyroxenoid, clinopyroxene, orthopyroxene, amphibole, quartz, apatite, carbonate, magnetite, hematite and barite bearing rocks.

The Gams ore is underlain by white quartzite with minor schists and conglomerate which succeeds the Hoogoor Suite leuco-gneisses. The deposit is located at the contact between the underlying quartzite and overlying Koeris formation (Porter GeoConsultancy, 2001). It is confined to a single, large sheath-fold that forms a steep-sided inselberg measuring 7 km by 5 km (McClung & Viljoen, 2011). Zinc is only found in sulphides in the Gams ore body; hence, the middle zone is discussed in detail.

The sulphide zone is irregularly present within the shear fold structure and is weakly mineralized containing 1 to 4 % Zn. The intervals of weaker mineralization include a number of higher grade cores with 7 % Zn embracing smaller cores containing of above 10 % Zn. The sphalerite ore rich in pyrrhotite and pyrite is located near the centre of the Gams ore Formation, flanked by iron sulphides, predominantly pyrite below and pyrrhotite above. These iron sulphides are in turn sandwiched by magnetite and hematite rich zones towards the outer margins of Gams Formation (Porter GeoConsultancy, 2001). There are six sphalerite populations in the Gams ore: zinc-, cobalt-, iron- and manganese-rich, as well as manganese-enriched and manganese-poor. The zinc- and cobalt-rich populations display high zinc contents and low concentrations of iron and manganese, while the iron-rich population displays higher iron contents. Dominating the

sphalerites of the mineralized pelitic schist is the manganese-enriched population (McClung & Viljoen, 2011).

### **2.1.2 Effect of ore mineralogy on zinc processing**

Few studies have focused on the correlation between the mineralogy of the deposit and its implications on mineral processing. Sphalerite is easily upgraded to form high grade concentrate by froth flotation and the concentrate can be easily refined by the traditional RLE process at most smelters. However, the high degree of stoichiometric variability, resulting in high levels of impurities in zinc concentrates, renders the concentrates unsuitable for processing by RLE processes. A study done by McClung & Viljoen (2011) on the manganese conundrum of the orebody determined that milling the ore to a finer grain size will not remove or reduce the iron and manganese content due to the absence of discrete iron and manganese mineral phases. However, the use of solvent extraction in downstream processing could control the manganese content and simultaneously recover zinc and manganese for sale. Hence, the recovery of manganese as a by-product could potentially enhance the economics of the deposit. Additionally, McClung & Viljoen (2011) stated that the mineral chemistry of the sphalerite might cause some sphalerite to display a delayed response during flotation or report to the lead concentrate. As a result, the option for shipping concentrate for refining elsewhere is less attractive since miners get penalized by smelters for manganese present in concentrates.

## **2.2 Zinc processing**

Meeting process criteria along with complying with environmental legislation are important issues that must be addressed when designing the flowsheet for a zinc refinery. As mentioned above, the traditional route for zinc extraction is through crushing-milling-flotation followed by RLE process (Filippou, 2004; Cole & Sole, 2002b). The direct use of hydrometallurgical route (ore-heap leaching-metal) allows utilization of ores that yield low upgrading recoveries whereas hydrometallurgical processing of concentrates avoids the generation of SO<sub>2</sub>. The zinc electrowinning stage is sensitive to presence of impurities in feed solution, hence the inclusion of solvent extraction stage accommodates the treatment of complex ores, e.g. ores that have high lead or halide contents (Cole & Sole, 2002a). Zinc processing involves using two types of media during the leaching stage namely, chloride and sulphate.

### **2.2.1 Chloride media**

Few processes that involve treatment of sulphides with chloride media have been commercialized due to corrosion problems associated with chlorine (Cole & Sole, 2003). Chloride leaching requires corrosion resistant materials of construction and also demands special care to avoid environmental pollution (Babu et al., 2002). Although modern materials of construction have improved, there are no plants using this technology (Cole & Sole, 2002a). There are two processes involving treatment with chloride media. The Zinclor process involves leaching concentrates with ferric chloride in which the zinc in the pregnant leach solution (PLS) is then extracted using pentylphosphonate prior to being electrowon. The electrowinning stage utilizes a divided cell which produces 99.99% zinc cathode and regenerates chlorine that is recycled to the leaching circuit. The second process is the CENIM-LNETI, This process involves the leaching of concentrates with concentrated ammonium chloride followed by zinc extraction using D2EHPA modified with isodecanol (Cole & Sole, 2003).

### **2.2.2 Sulphate media**

Many processes applied commercially use this medium. The sulphate medium can also involve bacterial leaching to extract valuable metals. Bioleach liquors can produce ultrapure electrolytes suitable for the production of special high-grade zinc metal (Cole & Sole, 2003). A study done by Fuls (2011) identified all processes using sulphate medium. However, it focused mainly on processes viable for Skorpion Zinc, an integrated zinc mine and refinery located near Rosh Pinah in southern Namibia. Elimination of other zinc sulphide processes was due to the remote location of Skorpion and zinc sources available (Black Mountain and Gamsberg in the Northern Cape, Rosh Pinah Zinc Sulphide deposits) taking into consideration the impurities contained within these ore bodies. Viable options had to be easily integrated with the existing refinery, have favourable economic outcome (low capital and operating cost with positive returns) and be able to produce zinc by 2016-2017.

Ful's study considered pyrometallurgical, electrolytic and hydrometallurgical processes. Both pyrometallurgical and electrolytic processes were deemed not feasible as they produce sulphuric acid as by-product. Due to the remote location of the existing plant, transport cost could prevent the sulphuric acid produced to be sold at a competitive price. Within the hydrometallurgical process category, the processes studied are: Dynatec pressure leach, Atmospheric leach, Albion leach, stirred tank bioleach and bio-heap leach. The stirred tank bioleach process was eliminated

due to dilute sulphuric acid and low zinc tenors produced which required an introduction of a high operating cost SX stage. The bio-heap leach process was rejected based on the lack of industrial development and the short time frame available to have it ready for production by 2016/2017. Consequently, the study developed and established the viability of Dynatec pressure leach, Atmospheric leach and Albion leach process. After investigation, the Albion process was discarded as an option due to high operating and capital costs. Compared to the Atmospheric leach process, the pressure leach process had low operating and high capital costs. Both processes were considered feasible and economically viable for Skorpion Zinc. This study will focus mostly on the heap leach process (HydroZinc™) which was rejected in the early stages of Ful's study.

## **2.3 Process routes being studied**

### **2.3.1 Production of zinc concentrates**

Direct smelting of zinc ores requires too much energy and furnace capacity and hence ores are first concentrated prior to smelting. Flotation is the most preferred concentration process (BCS, 2002). Ore is initially crushed and milled to a size where zinc mineral grains are liberated from the gangue. Very large grain sizes can enclose valuable minerals in the gangue thus preventing their flotation and very fine grain sizes cause slime formation. Therefore, the optimum grain size is approximately between 10-100 µm. The flotation feed particle size is controlled by hydrocyclones; coarser particles are recycled for further grinding (Davenport et al., 2002a).

After milling, the pulp is treated with various chemical reagents that promote flotation (collectors, frothers, activators and depressants) and water to control the flotation cell feed density (BCS, 2002b; Kawatra, n.d.). These chemical reagents modify surface properties of the mineral such that the valuable mineral attach to the air bubbles and float and thus are collected to the concentrate stream (Kawatra, n.d.; King et al., 2011). Flotation occurs in a series of steps, and cells are arranged from roughers to scavengers and then cleaners. Roughers make a coarse separation of values from gangue, and scavengers remove smaller quantities of the remaining values (BCS, 2002; Kawatra, n.d.). Concentrate from the cleaner cells is sent to thickeners where the concentrate is thickened by settling. The thickened concentrate is dewatered and dried prior to being fed to a roasting operation or sent for export. The block flow diagram for the production of concentrates is shown in Figure 3.

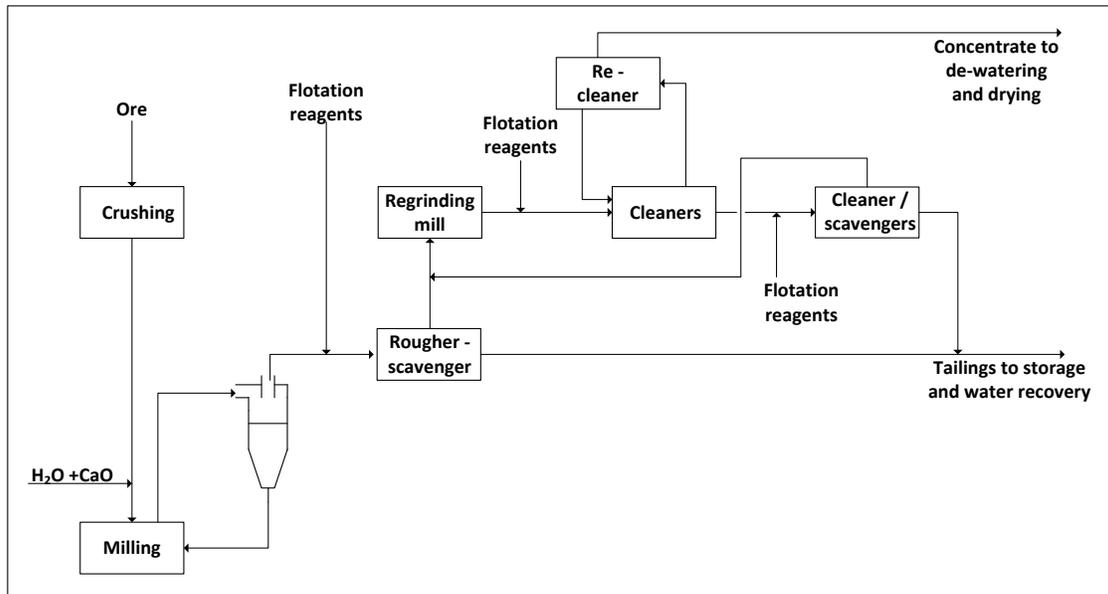
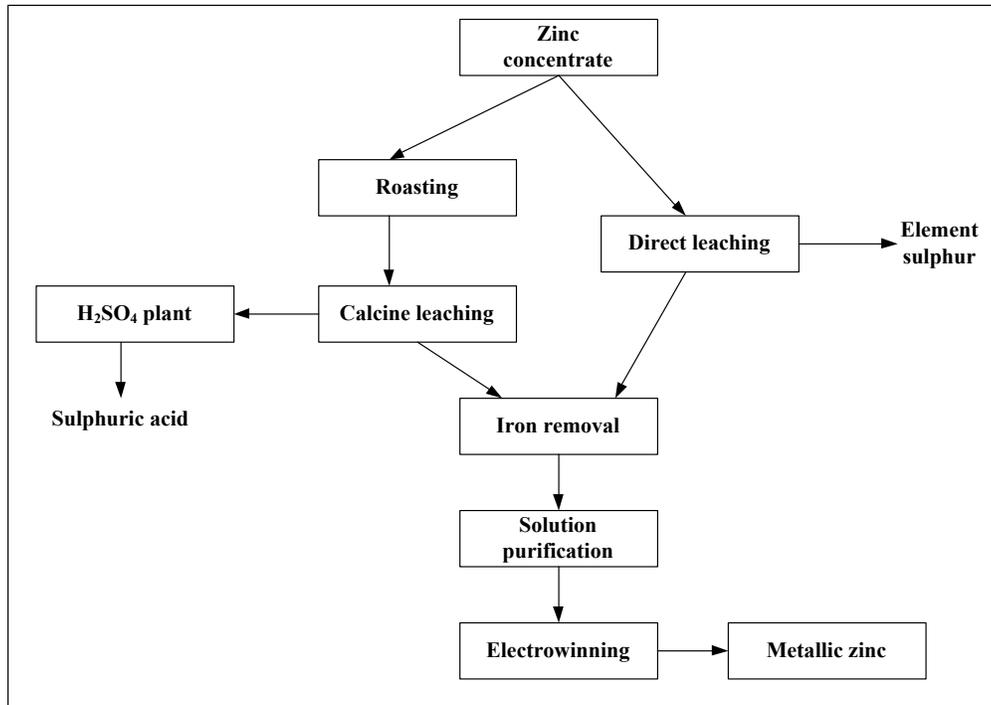


Figure 3: Generalized block flow diagram for producing concentrates (Davenport et al., 2002a)

### 2.3.2 Options for treating zinc concentrates

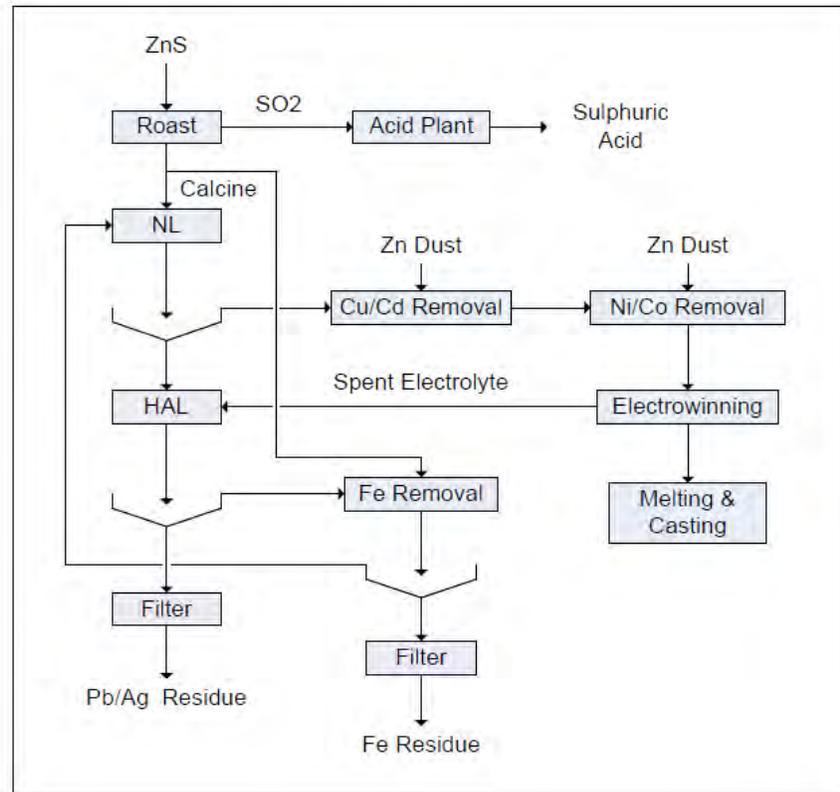
There are two process routes used to treat zinc concentrates, as shown in Figure 4.



**Figure 4: Process routes to treat concentrates**

### **2.3.2.1 The roast-leach electrowinning (RLE) process**

As mentioned above, RLE is the most common process used to produce zinc from sphalerite ore. This process consists of four stages: roasting of the zinc concentrates in order to remove sulphur and form a leachable zinc oxide, leaching of the zinc oxide, purification of the leach liquor and the electrolytic recovery of zinc (Moors & Dijkema, 2006; BCS, 2002a). During leaching, iron is also dissolved together with zinc and must be removed from the solution prior to electrowinning in order to produce a marketable zinc product. Iron is precipitated as jarosite during solution purification, approximately one ton of jarosite is formed per two tons of zinc produced (Moors & Dijkema, 2006). Other by-products from the process include gypsum, sulphuric acid, cadmium metal, silver/lead products and copper and cobalt enriched products (Moors & Dijkema, 2006). Figure 5 presents the process flow diagram.



**Figure 5: RLE process flow diagram. NL (Neutral leach), HAL (Hot Acid Leach) (Fuls, 2011)**

Concentrates containing approximately 50 % zinc as zinc sulphide and 30 % sulphur are roasted between 900 and 950 °C to give a zinc oxide calcine (Martín et al., 2002; Swarnkar et al., 1996). Iron present in concentrates is between 5 and 10 % (Buban et al., 1999). During roasting, most of the iron present in concentrates combines with zinc oxide to form zinc ferrite ( $ZnO.Fe_2O_3$ ) which does not dissolve during subsequent leaching (Swarnkar et al., 1996). Depending on the level of silica in the feed material, willemite ( $Zn_2SiO_4$ ) can also form during this stage. Moreover, most of the halides ( $Cl^-$  and  $F^-$ ) in the zinc concentrate volatilize during roasting and they are combined with the  $SO_2$  gas produced, these impurities are scrubbed from the  $SO_2$  stream prior to acid generation. The  $SO_2$  off-gas is then converted to 98% sulphuric acid via an effluent treatment plant (Wyethe et al., 2008).

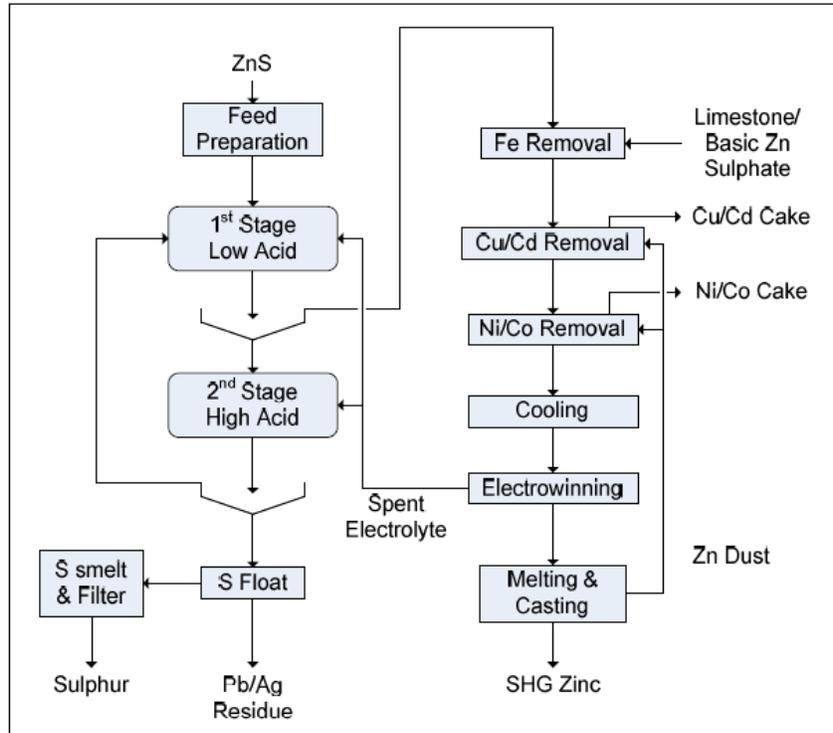
Most of the calcine from the roasting stage is subsequently leached with spent electrolyte in a neutral leaching (NL) stage to a final pH of 4.4 at 60 °C. Zinc oxide and willemite dissolves, thereby generating a zinc sulphate solution while zinc ferrite remains insoluble and separates as residue when the neutral leach slurry is separated. This residue contains approximately 16-18 %

zinc, 27-28 % iron and 0.2 % cadmium and other impurities such as arsenic, antimony and magnesium (Swarnkar et al., 1996). To recover this zinc, the residue is subjected to a hot acid leach (HAL) since it is not easily amenable to normal leaching. This stage is operated at approximately 100 g/l H<sub>2</sub>SO<sub>4</sub> and a temperature of 92 °C. The successive zinc sulphate solution obtained contains high level of iron; hence, this iron is precipitated into a filterable crystalline compound before the solution can be recycled back to the NL stage. Significant zinc losses occur in this stage due to adsorption to the iron precipitate. Ferric iron is precipitated by neutralization with calcine at pH 3.2 and 60 °C, jarosites are mostly precipitated under these conditions (Wyethe et al., 2008).

The zinc sulphate solution from the NL stage containing approximately 140 g/l zinc is purified before being fed to the cell-house where zinc is plated at current densities between 508-559 A/m<sup>2</sup> (Wyethe et al., 2008). Zinc dust is added to the solution to cement only copper and cadmium which precipitate out the solution as a sludge. This solution is further treated with zinc dust again which act as a catalyst to cement nickel and cobalt (BCS, 2002a). After the two purification stages, the solution is ready for electrowinning.

### **2.3.2.2 Direct leaching -Atmospheric leach process**

Different processes are used to directly leach zinc concentrates; the focus of this work is on the atmospheric leach process. The process was developed by Outotec in the middle of 1990s and has been practiced in industrial scale with several zinc plants since 1998 (Lahtinen et al., 2008). Unlike the RLE process, this process is environmentally friendly as it does not generate sulphur dioxide and hence avoids the production of additional sulphuric acid (Lahtinen et al., 2008). Moreover, it is highly cost-effective and extremely flexible with respect to different raw materials and variations of feed materials (Outotec, 2014; Saxén & Björn, 2008). When compared to other direct leaching processes e.g. pressure leaching, it offers low capital and operation cost, low power consumption and low need for maintenance (Saxén & Björn, 2008). Atmospheric direct leaching can be constructed as a standalone process or combined with calcine leaching and iron can be removed either as jarosite or goethite. The two stage process is illustrated in Figure 6.

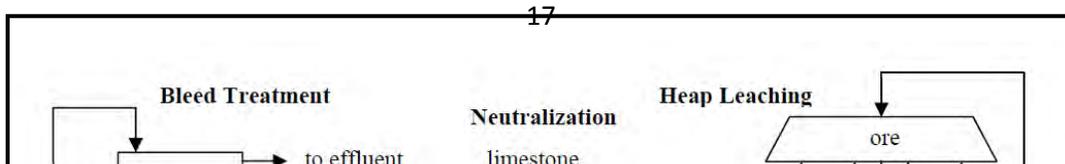


**Figure 6: Stand-alone atmospheric process (Fuls, 2011)**

The first stage leaching reactor operates at a temperature of 100 °C and a total overpressure less than 20 kPa. In this neutral leach stage, 75 – 78 % of zinc is extracted with the discharge acid concentration of 5-8 g/l. Partially leached solids are then fed to the HAL to give an overall zinc extraction of 95 % or more. The time required to extract zinc is dependent on the initial iron and lixiviant concentration, however, 10-20 h is needed to extract more than 95 % zinc (Filippou, 2004). The HAL stage operates at similar conditions to NL besides the acid concentration discharge which is 30 g/l. For this process, element sulphur produced is in its solid state and hence there are no problems associated with sulphur coating on reactive mineral surfaces (Lahtinen et al., 2008).

### 2.3.3 The HydroZinc™ Process

The HydroZinc™ process was developed by Teck Cominco in 1999 and was later demonstrated using ore from Red Dog mine between 2000 and 2002 (Lizama et al., 2003). The demonstration plant produced 1- t/day zinc cathode. This process treats low grade zinc sulphide ore by an entirely hydrometallurgical route. It consists of five stages, namely: heap leaching (HL),



neutralization (NT), solvent extraction (SX) Electrowinning (EW) and Bleed treatment (BT). A schematic diagram for the process is shown in Figure 7.

**Figure 7: The Teck Comico HydroZinc™ process (Lizama et al., 2003)**

Zinc sulphide ore is initially crushed, agglomerated and stacked on a line pad for heap leaching. The ore is reduced to 100 % passing 50 mm with the most preferable size being 12 mm. The heap is aerated with air at the bottom and irrigated at the top with an acidic raffinate solution from the SX stage. The SX raffinate contains approximately 15-30 g/l sulphuric acid and the irrigation rate is at least 5 l/min per m<sup>2</sup> of heap surface (Harlamovs et al., 2004). The pregnant leach solution (PLS) is rich in zinc and low in acid; it contains other metal ions solubilized during leaching e.g. ferric iron. Zinc delta (difference between raffinate and PLS) through the heap is 10 g/l (Filippou, 2004). The PLS is collected at the bottom of the heap and treated in the neutralization stage. Heap leaching technology is discussed in detail in sections 2.4.

In the neutralization stage, limestone is added to raise the pH to 3.8 and remove ferric iron and sulphate ions as iron hydroxide and gypsum respectively (Lizama et al., 2003). The neutralization slurry is sent to a sedimentation thickener where a flocculant is added. The thickener underflow is filtered, washed and stocked whereas the overflow is filtered to remove particulate material prior to SX.

The SX circuit consists of three stages: extraction, scrubbing and stripping. This stage is shown in detail in Figure 11. Di-(2-ethyl hexyl) phosphoric acid (DEHPA) dissolved in kerosene is used to extract zinc from the aqueous feed solution (Filippou, 2004; Lizama et al., 2003). Zinc is extracted in two stages in which organic raffinate (stripped organic) flows counter-current to the aqueous feed solution. The extraction zinc delta is 20 g/l (Lizama et al., 2003) and 30-50 % zinc is extracted (Filippou, 2004). The loaded organic solvent is washed in three counter-current scrub stages to remove entrained and co-extracted impurities. The scrubbing solution consists of water and spent/advanced electrolyte from electrowinning. After washing, the loaded organic is sent for stripping.

Zinc is then stripped from the loaded organic with spent electrolyte in two counter-current stripping stages. The zinc delta for the electrolyte (spent electrolyte to advance electrolyte) is 50 g/l (Lizama et al., 2003). The stripped organic is recycled to the extraction stage, thus, completing the organic cycle. To avoid poisoning the organic solvent with ferric, a portion of the organic raffinate is contacted with metallic zinc shot to reduce ferric to ferrous iron prior to recycling. Ferrous iron can be easily stripped from the organic phase using water and electrolyte (O'keefe, 1993). Entrained organic is removed from the advanced electrolyte prior to the EW circuit using dual media filtration and activated carbon treatment. Zinc is electrowon from the advance electrolyte and the spent electrolyte is recycled to the SX stripping stage. Iron levels in loaded electrolyte in excess of 5 mg/l reduce zinc current efficiency (Nicol et al., 2008). To control impurity build up in the electrolyte, a portion is bled off (purged) from the circuit.

Minerals found in zinc ores are not co-extracted with zinc during solvent extraction so they are recycled with the raffinate to the leach circuit. A bleed treatment stage is added in the process to control the impurity level, remove excess water and recover zinc from the water. The raffinate stream is neutralized with lime and/ or limestone to recover zinc as a basic zinc sulphate (BZS) and precipitate impurities. BZS is separated from impurity precipitates by settling in a clarifier. The underflow (BZS) is recycled to the neutralization stage and the overflow is treated in an effluent treatment plant.

## **2.4 Heap leaching**

With high grade ore resources declining and the economic and environmental cost of energy increasing, less energy intensive processes are becoming more attractive. Such processes include

heap leaching. Heap leaching is usually applied to marginal ores that cannot be economically processed by the concentrate route. It involves the stacking of metal-bearing ore into a heap on an impermeable pad. The ore is irrigated at the top for an extended period of time (days, months or years) with a chemical solution in order to dissolve targeted metals. The leachate becomes increasingly loaded as it percolates through the heap and is then collected from the base of the heap. The success of heap leaching is measured by the degree to which target levels and rates of recovery are achieved (Kappes, 2002). To enhance metal recovery several methods are used. This includes solution recycling, use of chemical and biological reactants and/or physical alteration of the ore (Ulrich et al., 2003).

The economics of the process are strongly governed by the trade-off between the slow rate and limited extent of leaching from large particles and the cost of crushing finer (Ghorbani et al., 2011). When operated correctly, heap leaching offers an economic recovery of metals at a low cost compared to current conventional processes (Petersen & Dixon, 2007). When using physical geometry, heap leach facilities are divided into four types (Jansen & Taylor, 2002):

1. ***Dump leach facility***- These facilities use run-of-mine (ROM) ore placed in a defined topographic area and rely on natural low permeability barriers underlying the stack to control the flow of solution. Ore is usually placed in lifts greater than 10 m and total heights reaching 200 m or more.
2. ***ROM leach pad***- These facilities use ROM ore placed on a designed pad that is lined with a low permeability layer and fitted with a solution collection system. Ore lifts of 7 to 8 m are used. Heap total heights range from 100 to 150 m.
3. ***Permanent heap leach pad***- These facilities use physically altered ore placed on a designed pad that is lined with a low permeability layer and fitted with a solution collection system. Heaps consist of 4 to 7 m lifts and total heights ranging from 80 to 100 m.
4. ***On/off heap leach pad***- These facilities use similar pads and collection systems as the permanent pad but each lift is removed after it has been leached. The spent ore is then placed in large lifts in a separate waste dump.

The choice on the facility to be used is dependent on the site constraints and the nature of the ore. The focus of this work is on using heap leach pad facilities.

### 2.4.1 Use of heap leaching process

Heap leaching is mostly chosen based on financial reasons as it represents the best return on investment (ROI) on low-grade materials. Moreover, heap leach projects can be the first stage towards developing a long term project since they have a shorter construction time. They can allow small companies to establish an operation which can pay for further resource development and fund the next stage of development (John, 2011). For operations in unstable political areas or small operations, heap leaching may be chosen due to the manageable level of capital investment (Kappes, 2002). Additionally, some ore reserves are not enough to justify a conventional mill thus making heap leaching the only viable option for processing them (John, 2011). Advantages and disadvantages of heap leaching are shown in Table 2.

**Table 2: Advantages and Disadvantages of heap leaching (Ghorbani et al., 2011)**

Advantages	Disadvantages
Low capital and operating costs	Lower recoveries than mill/float or mill/leach
Milling step is not needed, however, crushing and agglomeration may required	Long leach cycles and hold-up
Simplicity of atmospheric leach processes	Lengthy pilot test programme
Can be used to treat low-grade ores, wastes and small deposits	Large footprint
Less intensive environmental regulatory concerns	Environmental release of PLS

Metal tenor may be built up by recycling solution over heaps
Simplicity in equipment and operation
Shorter start-up times

The use of agitation leach tanks (using similar lixivants) has preceded heap leaching due to some heap leaching processes lacking the ability to achieve high recovery rates of metal from low grade solutions without destroying the lixiviant. As a result, some heap leaching processes remain novel ideas (Zn, Ni & Co). Although these metals have achieved recoveries of over 85 % in heap leach pilot testwork, they are still not commercialized due the lack of SX reagents or IX exchange resins to selectively extract these metals without significant pH modification to the PLS and/or neutralization of the sulphuric acid lixiviant (John, 2011). This additional stage affects the overall economics of the heap leaching processes.

#### **2.4.2 Efficiency factors in heap leach process**

The recovery rate from heaps is characterized by the percentage of metal of interest in the ore transferred into the leach solution and the time required for metal dissolution. Factors that influence the degree of metal dissolution are discussed below.

##### **2.4.2.1 Degree of liberation**

*Ore particle size:* ore is crushed to different particle size ranges in order to increase exposure of the mineral grains to lixivants as leaching rates are proportional to the surface area. Ore is crushed in multiple stages; in primary crushing, the particle size is reduced to a size range of 10 to 15 cm (Zanbak, 2012). The ore is further crushed to optimum heap sizes during secondary/tertiary crushing. Permeable ores require little or no crushing at all. John (2011) categorized heaps to comprise of ores with p80 of 5 to 100 mm, however, according to Ghorbani and co-authors (2011) most heap leach facilities use a top size range of 12-25 mm. All particle class sizes initially have a rapid leaching rate; the conversion is almost complete for smaller sizes and partial for large particles over a 1-2 year period (Ghorbani et al., 2011).

*Lixiviant contact with ore particles:* the lixiviant rate through the heap should be slow enough to provide sufficient contact with the ore particles to dissolve the metals. According to John (2011), irrigation rates between 2-15 Lm<sup>-2</sup>h<sup>-1</sup> are optimum for heaps. The presence of fine particles in the

crushed ore decreases heap permeability and hence decreasing the percolation rate of the lixiviant through the heap (Zanbak, 2012). However, their presence is critical to ensure ore-liquid contact through capillary action.

#### **2.4.2.2 Agglomeration**

During crushing, fine particles are generated and an excessive amount of these particles can clog pores between large ore particles thus leading to an uneven distribution of the leaching solution. Agglomeration binds these fines into larger particles thereby improving permeability and recovery of sought metals (Zanbak, 2012; John, 2011; Breitenbach, 2005).

#### **2.4.2.3 Permeability**

Permeability is the most important aspect of heap leaching. Failure to achieve a permeable heap reduces recovery due to incomplete lixiviant access to the ore, decreased percolation rates, increased residence time and decreased oxygen level in the heap (Ghorbani et al., 2011; John, 2011). Uneven permeability in heaps results in lixiviant channeling thus leaving some parts of the heap without immediate contact. Typical recovery in base metal sulphide heaps is between 60 and 70 % over a period of 1-2 years (Ghorbani et al., 2011). Most compaction in heaps is caused by unsuitable heap stacking and the presence of fines which tend to block or channel the percolation of the leach solution.

#### **2.4.2.4 Solution delivery**

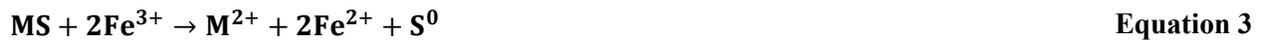
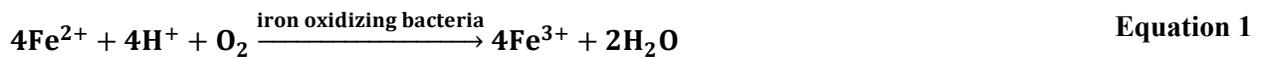
Lixiviant addition rate into a heap is also critical for the successful operation of a heap leach facility. Heap leaching is partially limited by the supply and diffusion of sulphuric acid which is a key reagent for ferrous oxidation reaction. Lixiviant addition is achieved by using a network of pumps, pipes, tubing, valves and irrigation drippers or sprinklers. Occasionally, operators will combine drippers and sprinklers (Ulrich et al., 2003). Heat is generated during sulphide leaching; an increase irrigation rate can improve the operating conditions of the heap by removing excess heat (Petersen & Dixon, 2007).

#### **2.4.3 Heap bioleaching**

Bioleaching technology offers a significant potential to deliver attractive environmental and social benefits in the mining industry (Pradhan et al., 2008). Heap bioleaching is established in copper and gold industries and has been demonstrated for the zinc industry (Lizama et al., 2005). This microorganism based metal recovery process has a competitive edge over other processes in

the face of increased concern regarding the effect of mining on the environment (Rawlings, 2004).

Bioleaching of sulphide minerals involves three sub-processes: bio-oxidation of ferrous into ferric (Equation 1), microbial oxidation of sulphur into an acid (Equation 2) and ferric leaching of the metal sulphide (*MS*) mineral (Equation 3). Acidithiobacillus Ferrooxidans or similar bacteria are mostly used to catalyze the oxidation of sulphide minerals during biological leaching (John, 2011; Pradhan et al., 2008). According to Pradhan and co-authors (2008), mineral bio-oxidation operations are advantageous since they are not usually subject to contamination by the generated unwanted microorganisms. However, heap bioleaching of sulphides is slow compared to pyrometallurgical process routes and other high intensity hydrometallurgical processes due to the slow rate of oxygen transfer in heaps (Watling, 2006).



#### 2.4.3.1 Factors affecting heap bioleaching

Like any other process involving living organisms, heap bioleaching is influenced by environmental, biological and physico-chemical factors (Table 3). For the growth of microorganisms, optimal conditions of humidity, pH, temperature, energy sources and nutrients have to exist along with the absence of possible inhibitors (Pradhan et al., 2008). Moreover, heap geometry may also affect bioleaching processes and rate of recovery.

**Table 3: Factors and parameters affecting heap bioleaching and metal recovery (Pradhan et al., 2008)**

Factor	Parameters affecting bioleaching
<b>Physical and chemical parameters</b>	Temperature, pH, redox potential, CO <sub>2</sub> and O <sub>2</sub> content, nutrient availability, oxygen availability, homogenous mass transfer, Fe (III) concentration and the presence of inhibitors.
<b>Biological parameters</b>	Microbial diversity, population density, microbial activities, metal tolerance, spatial distribution of microorganisms, attachment to ore particles, adaptation abilities of microorganisms and inoculum

<b>Ore characteristics</b>	Composition, mineral type, acid consumption, grain size, mineral dissemination, surface area, porosity, hydrophobic galvanic interactions and formation of secondary minerals
----------------------------	---

#### 2.4.3.1.1 Type of ore material

The bioleaching process depends largely on the nature of ore material, since microbiological activity inside the heap is conditioned by its mineral composition (Muñoz et al., 1995). A porous ore allows the lixiviant to penetrate more into the ore body. High carbonate content in the ore increases the pH of the lixiviant thus suppressing the bacterial activity (Bosecker, 1997). In addition, the ore should provide sufficient quantity of nutrients to the microorganisms for growth. Quartzic and granitic ore are susceptible to microbial leaching (Pradhan et al., 2008). For maximum bacterial activity, heaps have to be maintained at a pH range between 1 and 2 (Watling, 2006).

#### 2.4.3.1.2 Aeration

Most metal leaching microorganisms are aerobic. Aeration supply both O<sub>2</sub> and CO<sub>2</sub> to the heap. CO<sub>2</sub> serves as a source of carbon needed for biomass generation (Pradhan et al., 2008). The lack of oxygen in the heap can dispose it to anaerobic conditions. Aeration accelerates bio-oxidation as oxygen is very important for viability and activity of leaching microorganisms. Increasing the rate of aeration increases the leaching rate. However, if oxygen is sufficiently present throughout the heap, an increase in aeration does not have an effect.

#### 2.4.3.1.3 Irrigation

A study by Lizama and co-authors (2005) identified the irrigation to height ratio, L/h, as a key parameter in heap bioleaching. This ratio controls colonization and steady-rate stages of bioleaching, acid and iron balances (Lizama et al., 2005). There is an inverse relationship between heap height and metal extraction (Padilla et al., 2008). When two columns of different heights are irrigated at the same rate, the shorter column receives high amounts of reagent per mass of reacting ore in a given time period. As a result, the shorter column which has a high L/h ratio extracts faster.

#### 2.4.3.1.4 pH

Solution pH is critical to the availability of ferric ion which is a reagent for leaching most sulphide minerals. High pH is detrimental to ferrous ion oxidizing ability by certain microorganism and may also reduce heap permeability due to ion precipitation within the heap. A pH greater than 2.0 has a negative effect on the microbial population, such levels causes bacterial de-activation and thus result in loss of microbial culture (Ojumu & Petersen, 2011; Du Plessis et al., 2006).

#### 2.4.3.1.5 Temperature

Microorganisms are classified by the temperature range in which they can survive; optimum temperatures of 30–40 °C for mesophiles, around 50 °C for moderate thermophiles and above 65 °C for extreme thermophiles. Below the optimum temperature, microbes are inactive and at temperatures above it, they are denatured. Biooxidation of sulfide minerals is an exothermic process; it produces significant heat in stirred tank reactors and heaps (Ojumu et al., 2009; Pradhan et al., 2008). According to Ojumu et al., (2008), the exothermic reactions can make the internal zones in heaps reach temperatures above 60 °C. Seasonal changes can cause modifications in the microbial population while favoring the development of mesophilic or thermophilic bacteria (Pradhan et al., 2008). Thus, the heap temperature, due to the exothermic reactions, can be a controlling factor in some commercial operations.

#### 2.4.3.1.6 Other factors

Metal oxidation mediated by microorganisms can be inhibited by other factors such as organic compounds, surface-active agents, solvents, or specific metals (Pradhan et al., 2008).

##### 2.4.3.1.6.1 *Jarosites formation*

The bio-oxidation step in bioleaching is sometimes inhibited by the formation of an intermediate sulfide passivation layer. The passivation layer complex is formed by  $\text{Fe}(\text{OH})_3$ , which tends to form the jarosite  $[\text{KFe}_3(\text{SO}_4)_2(\text{OH})_6]$  and subsequently coats the unreacted material (Pradhan et al., 2008). The elemental sulfur formed during the process passivates the ore through a similar mechanism, thus decreasing the effectiveness of reagent and mineral surface interaction. The formation of ferric iron precipitate and jarosite is highly pH dependent, with precipitation of jarosite favored in pH range 1.5 to 3 and sulfate concentration higher than 3000 mg/l (Murad & Rojik, 2004). The pH favoring formation of jarosite is also ideal for microbial activity and

jarosite precipitation increases with an increase in temperature (Guise & Castro, 1998). Equation 4 shows the formation of jarosites (Daoud & Karamanev, 2006).



*2.4.3.1.6.2 Build-up of metal ion/organic matter concentration*

High concentrations of heavy metals such as copper, zinc, arsenic and iron are toxic to microorganism; they deleteriously affect the rates of bio-oxidation. Dissolved ions interfere with the microbial ferrous iron oxidation which is a critical sub-process in bioleaching (Ojumu et al., 2008). A study done on multi-metal ion tolerance of *Acidithiobacillus Ferrooxidans* with respect to bioleaching of sulphide mineral established that among  $\text{Fe}^{3+}$ ,  $\text{Cu}^{2+}$  and  $\text{Zn}^{2+}$ ,  $\text{Fe}^{3+}$  is the most toxic whereas  $\text{Zn}^{2+}$  is the least (Das et al., 1997). Likewise, high concentration of sulfate and other ions in the re-circulating irrigation solution on the heaps has deleterious effect to the bacteria. Moreover, the organic compounds used in solvent extraction might be introduced into the raffinate thus generating adverse conditions to bacteria (Pradhan et al., 2008).

**2.4.4 Heap design**

Heaps are either constructed in a multi-lift or on/off pattern. Multi-lift heaps consist of an initial lift which is built on an impermeable surface and subsequent lifts which are built on top of the first one after it has been leached. On/off heaps consist of a single lift built on an impermeable surface. They are removed after leaching and replaced by a new lift. The advantages and disadvantages of the two types of heap construction are shown in Table 4 (Davenport et al., 2002b):

**Table 4: Advantages and disadvantages of multi-lift and on/ off heaps**

Multi-lift heaps	On/off heaps
<b>Advantages</b>	
Ore need only be moved once onto the heap	Simple to construct
Lixiviant flows through all the lifts until	Aeration and pregnant solution pipe-work

leaching is moved to another area thus permitting recovery of metal from leaching minerals in the lower lifts.	can be maintained when ore is emptied from the pad.
	Base need not be as strong as those needed for multi-lift heaps
<b>Disadvantages</b>	
Require a strong impermeable base	Ore has to be moved twice (on and off)
Require a large initial base because the heaps are pyramidal	
Require a versatile heap building equipment	
Impermeable zones in lower lift may cause high grade solution from upper lifts to be trapped, causing a severe reduction in recovery rate and possibly in overall recovery.	

#### **2.4.4.1 Heap scale up**

For the design of full-scale heaps, several column tests and pilot heap tests are required. However, these test programs do not guarantee accurate prediction of full scale heap behavior (Lizama et al., 2005). Table 5 shows the rule of thumb design criteria when scaling up laboratory results.

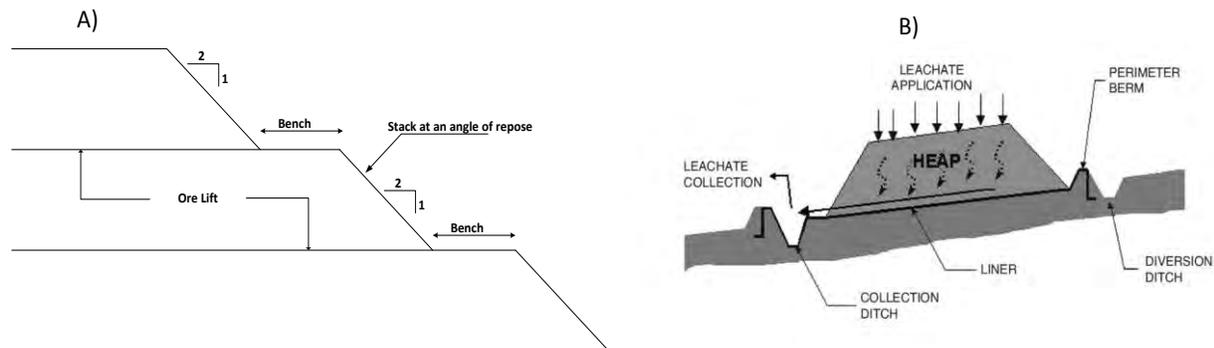
**Table 5: Scaling up of laboratory results for full-scale heaps (John, 2011)**

Laboratory Results	Rule of Thumb Design Criteria
<b>Leach period ‘T’ for certain recovery ‘R’</b>	Leach period is 3T to achieve a recovery of 0.8R

<b>Pre-lixiviation / agglomeration addition of 'L' kg per tonne</b>	Design 120 % L kg /t for all reagents
<b>PLS grade</b>	Calculate assuming leach period 3.T, 0.8.R, lift height, nominal irrigation rate and ore grade after mining dilution. Do not use test work figures for PLS grades.
<b>Ore grade</b>	Assume 80 % grade of predicted reserves

#### 2.4.4.2 Heap construction

Heap ore lifts are between 5 and 10 m and have maximum heights in the range of 30 to 60 m (John, 2011; Breitenbach, 2005). To establish an overall stable design slope for a heap, each lift is offset with benches along the exterior slope. The highest heap stack is above 150 m (Breitenbach, 2005). The exterior ore lifts slopes are constructed at an angle of repose of 2H: 1V (Breitenbach & Thiel, 2005; Van Zyl, Henderson & Cobb, 1990). Heap designs are shown in Figure 8.



**Figure 8: a) Heap slope section with typical angle-of-repose ore lift slopes and bench setbacks for overall flatter slope b) Simplified schematic cross-section of a heap leach facility (Ulrich, Andrade & Gardner, 2003)**

#### 2.4.4.3 Ore preparation

Preparation of ore for heap leaching involves placement of run-of-mine (ROM) ore on the leach heaps with or without trickling a strong  $H_2SO_4-H_2O$  solution through the heap (acid curing). The variant of this involves crushing ROM ore and agglomerating it with strong sulfuric acid prior to placing the agglomerates on a heap pad. The former method is cheaper but gives the slowest leaching rates hence lower metal recovery. Acid curing quickly dissolves metals from soluble minerals and acidifies the heap, thereby preventing ferric sulfate precipitation during subsequent leaching. Approximately, 10 or 20 kg of strong sulfuric acid per ton of ore are supplied to the

heap over a period of 10 days; this period is shorter for oxide ores and longer for sulfide ores (Davenport et al., 2002b). Acid cure economically enhances extraction rate and efficiency.

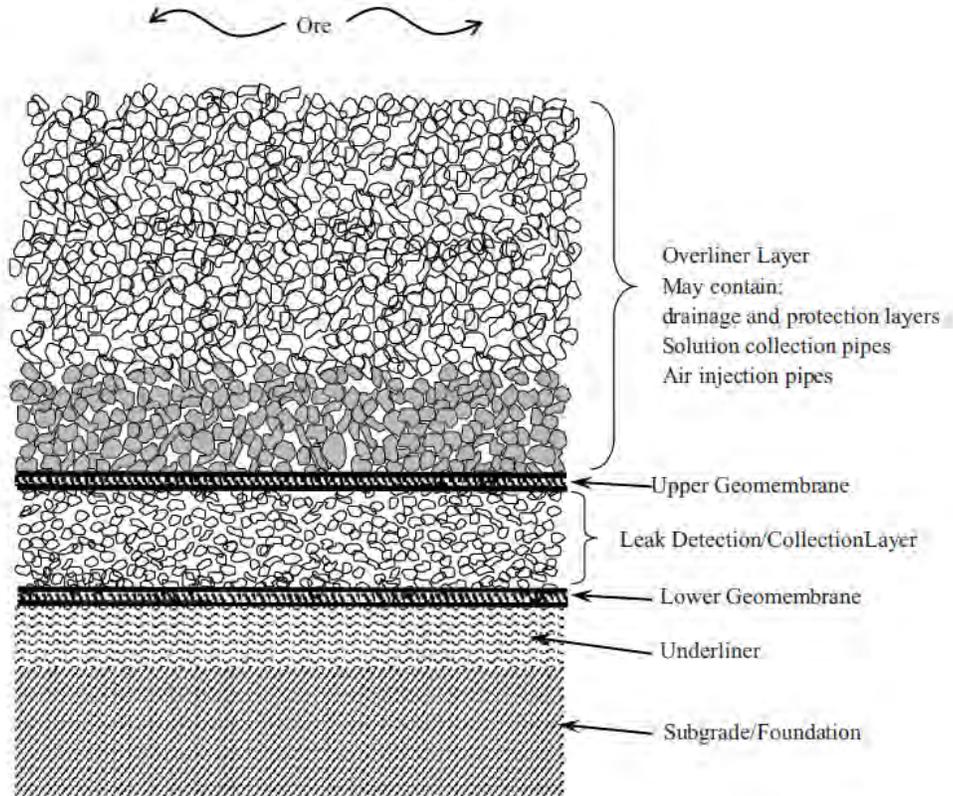
Metal extraction and efficiency increase with a decrease in ore sizes. Fine particles within a heap help decrease the percolation rate of a lixiviant and hence increasing extraction rates. However, too much fines in a heap can end up clogging it. Consequently, there needs to be a balance in terms of fines in a heap.

#### **2.4.4.4 Pad construction and drainage system**

Pad construction/ liner system contributes significantly to heap leach capital and operating costs (Van Zyl, Henderson & Cobb, 1990). A properly designed liner system offers environmental and operational benefits by controlling the leachate within the heap facility thus enhancing solution recovery (Lupo, 2005). Liner system can be used for 10 years or more (Breitenbach, 2005). The base of the heap is sloped between 0.5 % and 2 % (2 m drop in 100 m horizontal) in order to direct the PLS to a collection basin (ponds) and avoid slippage of the heap on the polymer liner (Breitenbach & Thiel, 2005; Ulrich, Andrade & Gardner, 2003; Kappes, 2002). Liner systems configurations depend on the leach facility, site conditions (climate, topography) and ore type.

Liner system designs consist of either single composite or double composite system Figure 9. The liner systems are named using the number of low-permeability layers in the system. Regulatory authorities determine the pad construction to be used; double-lined pads are more required by regulatory agencies (Van Zyl, Henderson & Cobb, 1990). Single composite liner systems are used where the hydraulic head for the leach solution is low; the inverse is true for a double composite liner (Lupo, 2005). As a result, ponds are constructed with double composite liners (Kappes, 2002).

Heaps with low permeability ore sometimes require inter-lift liners due to the significant amount of time taken by the leachate to reach the base of the heap. These liners can be designed to partially leak and re-leach lower lifts. However, most of the leachate (>90 %) from the fresh ore lift is sent through the drain system to the process ponds (Breitenbach, 2005). Generally, there are no inter-lift liners for copper and zinc heap leaching (Breitenbach & Thiel, 2005).



**Figure 9: Double Composite Liner (Lupo & Morrison, 2007)**

#### **2.4.4.5 Ore placement**

Heaps are constructed either by dumping ore from trucks or by stacking using a conveyor. The use of trucks makes construction simple although the vehicles tend to compact heaps as they run over it to their dumping destination. Large trucks noticeably compact ore to a depth of 2 m. Moreover, truck stacking can result in the tie-up of a large tonnage of ore below the truck roadways. Since the width of truck roadways is nearly the same regardless of the daily production rate, this has major effects on small operations compare to large ones. As a result, truck stacking is mostly used for operations processing at least 100 000 tonnes/day. Conveyors avoid the compaction problem hence they are extensively used worldwide (Davenport et al., 2002b). Conveyor stackers are mostly operated from the base of the lift; however, they may be located on top of the lift and dumping over the edge. Such stackers are used for heaps processing up to 50 000 tonnes/day (Kappes, 2002).

#### **2.4.4.6 Aeration**

Sulphide leaching requires O<sub>2</sub> from air for the leaching reactions unlike oxide leaching which only requires lixiviant. The air is provided by perforated polymer pipes which are placed on the heap base and blow air upwards through the heap. In order to avoid filling air pipes with lixiviant, Davenport and co-authors (2002) recommends that pipes should be 1 m above the heap base, provide a drainage manifold at their low end and they should be sloping and blowing downwards in the direction of solution flow.

#### **2.4.4.7 Solution application equipment**

There are four standard types currently being used for heap leaching and these are: drip emitters (most common), wobbler sprinklers, reciprocating sprinklers (mostly used for side slopes) and high rate evaporative sprinklers (used in cases where rapid evaporation is needed, i.e. heap closure). Regardless of the system used, capital and operation costs for solution handling are low. On the heap, high Density Polyethylene (HDPE) header pipes up to 400 mm diameter are placed every 30 to 60 m across the heap. From the header pipes, distribution pipes of polyvinyl chloride (PVC) or UV stabilized PVC follows. These are between 75 mm and 150 mm in diameter and they are also placed on similar (30 to 60 m) spacing. From these, drip emitter lines up to 60 m long cross the heap on 1 m centers, or sprinkler manifold pipes (25 to 50 mm PVC) up to 60 m long cross the heap on 6 to 8 meter centers (Kappes, 2002).

#### **2.4.4.8 PLS collection**

The solution from the heap flows by gravity through approximately 10 cm polymer drain pipes on the sloping heap base to a collection trench. This solution then flows by pipeline from the collection trench to a pond. High density polyethylene pipes are used for low pressure flows. 316L stainless steel pipe is used for high pressure pumped flows (Davenport et al., 2002b).

### **2.5 Design of tailings dams**

80% of the ore end up as tailings in the zinc industry and hence tailings disposal is of significance in the overall mining and milling operations (Kohad, 1998). According to Brown (2002), failures of tailing facilities have resulted to the closure of mining operations, devastating environmental damage and loss of lives. The main purpose of tailings impoundments is to retain fine-grained tailings with a secondary or co-purpose of conserving water for use in the mill and mine (U.S. Environmental Protection Agency, 1994).

Tailings disposal includes discarding dry or thickened tailings in impoundments, backfilling underground mines and open pits. The most common method is the disposal of tailings slurry in impoundments (U.S. Environmental Protection Agency, 1994). Designs for tailings disposal is dependent on the quantity and characteristics of the tailings, disposal site (climate, topography, geology, hydrogeology) and regulatory requirements related to safety and environmental performance. Three factors are of importance during design: cost, stability and environmental performance.

Tailings are retained in impoundments either by raised embankment or retention dam. The difference between the two structures is that retention dams are constructed at full capacity at the beginning yet raised embankments are constructed in phases over the life of the impoundment. Hence, raised embankments make it possible to address problems that may arise during the life of the impoundment and they have lower initial capital cost. Moreover, raised embankments impoundments have a wide variety for construction material compared to retention dams, this is due to the smaller quantities needed at a particular time. Due to the reasons above, raised embankments are more common, hence they will be discussed in detail.

Raised embankments can be constructed using upstream, downstream, or centerline methods. These methods are named depending on how the centerline of the embankment crest moves with each phase, i.e. downstream embankment is constructed such that the centerline of the embankment crest moves downstream with each phase (Brown, 2002). These methods are compared in Table 6.

**Table 6: Comparison of Embankment Types (Vick, 1990)**

Embankment Type	Upstream	Downstream	Centerline
<b>Mill Tailings Requirements</b>	At least 60% of the tailings should be sand.	Suitable for any type of tailings	Sands or low plasticity slimes
<b>Water-Storage Suitability</b>	Not suitable for significant water storage	Good	Not recommended for permanent storage.
<b>Raising Rate Restrictions</b>	Less than 30 ft/yr. Over 50 ft/yr can be hazardous	None	Height restrictions for individual's raises may apply.
<b>Embankment Fill Requirements</b>	Native soil, sand tailings, waste rock	Sand tailings, waste rock, native soils	Sand tailings, waste rock, native soil
<b>Relative Embankment Cost</b>	Low	High	Moderate

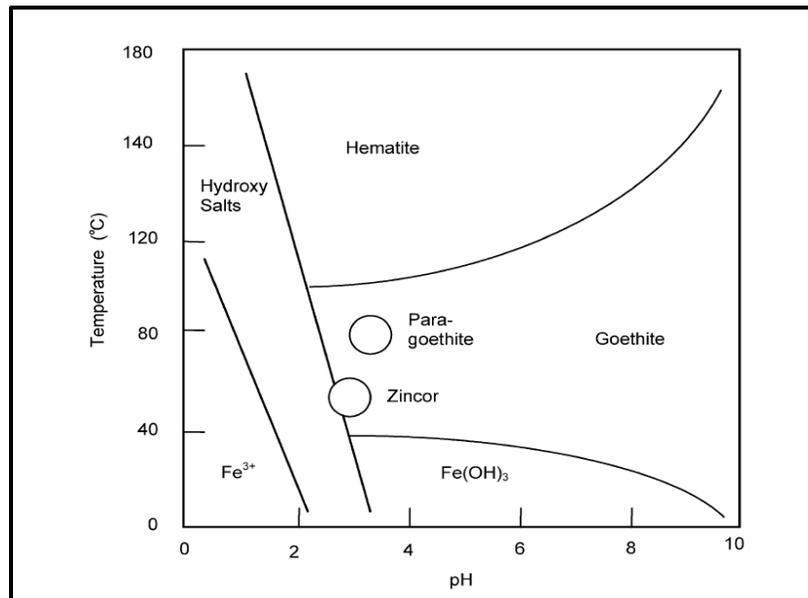
## 2.6 Solution purification

### 2.6.1 Iron precipitation from leach solutions

Iron is one of the most abundant elements on earth and ends up in many process solutions. It is mostly found in the form of pyrite, pyrrhotite and chalcopyrite. It is an undesirable constituent of zinc ores, concentrates and calcine oxides hence its presence in leach solutions is a major problem in zinc hydrometallurgy, and it needs to be removed before electrolysis. Zinc concentrates usually contain between 3% and 18% iron. The presence of iron in zinc concentrates necessitates the incorporation of at least two extra processing stages and determines the amount of zinc that can be recovered (Claassen et al., 2002).

During leaching, iron is dissolved together with the desired metal and is mostly in the form of ferric in the PLS. Iron is subsequently precipitated from the PLS as an oxide, oxy-hydroxide or a hydroxyl salt e.g. jarosite (Claassen et al., 2002). Iron oxides, hydroxides or oxy-hydroxides are collectively referred to as iron oxides. These iron oxides have a strong adsorptive capacity;

therefore, their precipitation involves co-removal of organic and inorganic impurities (Loan et al., 2006; Claassen, 2002). However, this benefit can be out-weighed by loss of valuable metals and the fact that the precipitates are metastable. The precipitates are also difficult to settle and filter. Removing iron as a hydroxide results in poorly filterable, voluminous and gelatinous precipitates (Claassen, 2002). Moreover, the solid separation and washing stages become difficult since these stages are efficient if the precipitate is dense with a crystalline character. The relative stabilities of iron oxide precipitates are shown in Figure 10.



**Figure 10: Phase diagram for iron oxides precipitation from 0.5M ferric sulphate solutions. Hydroxy salts = basic iron sulphates e.g. Jarosites.  $Fe(OH)_3$  = iron oxy-hydroxide(s) (Claassen et al., 2002).**

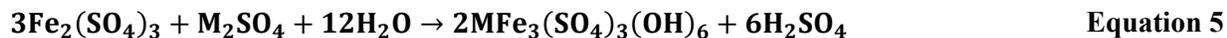
The efficient removal of iron from leach liquors is a critical step in zinc processing. Therefore, precipitate quality, that is particle size, morphology and density, plays an important role in iron precipitation. The precipitate size is determined by the nucleation rate and crystal growth. Supersaturation is the critical parameter controlling the nucleation rate (Myerson, 2002). At high supersaturation levels, nucleation is favored over growth thus small particles are formed while low supersaturation levels promotes growth, hence coarser particles (Jarvenin, 2008; Myerson, 2002). Using hot, dilute solutions keeps the level of supersaturation low and the slow addition of reagents to the impure hot PLS ensures that the dilute conditions are maintained (Claassen, 2002). The precipitate size influences the economics of the process as it impacts the cost of

downstream processing and the loss of valuable metals in residue streams through adsorption (Claassen & Sandenbergh, 2007).

In most existing zinc plants, iron is removed as jarosite because of its simplicity and low costs (Ismael & Carvalho, 2003; Dutrizac, 1996). However, the large volume of residues produced has become questionable on environmental grounds. The jarosite process was the first iron removal process that produced filterable residue on a commercial scale (Claassen, 2002). The goethite and hematite processes are more costly thus they are less utilized (Loan et al., 2006). These processes are discussed in detail below.

### **2.6.1.1 Jarosite process**

Jarosites are iron-oxyhydroxysulfate material and are mostly found in acidic and sulphate rich environments (Zhu et al., 2013). Their formation is represented by the following equation (Klauber, 2008; Ismael & Carvalho, 2003):



Where **M** represents any of the ions:  $\text{H}_3\text{O}^+$ ,  $\text{Na}^+$ ,  $\text{Li}^+$ ,  $\text{K}^+$ ,  $\text{NH}_4^+$ ,  $\frac{1}{2}\text{Pb}^{2+}$ . The formation of jarosite is temperature, pH and Eh dependant. The stability zone is within pH 1 to 3 and temperatures between 20 and 200 °C (Ismael & Carvalho, 2003). At low and high pH values, there is no precipitation. In this process, iron is precipitated from a hot acid leach solution in the presence of sodium or ammonium ions which act as jarosite seeds. The presence of jarosite seeds eliminate induction periods and promote the rate of precipitation (Dutrizac, 1996). The ideal conditions for the process are temperatures between 95 to 100 °C, pH 1.5 to 1.8, vigorous agitation and the presence of seed material. The acid formed during precipitation is neutralized with zinc calcine. During precipitation, zinc and other valuable metals are co-precipitated with jarosite. The zinc recovery is between 96 and 98% when iron is recovered as jarosite (Ismael & Carvalho, 2003). The advantages of the process includes the production of filterable iron precipitates, minimum losses of zinc in jarosite and the ease of integration into a variety of hydrometallurgical flowsheets (Dutrizac, 1996). The weakness of the process is the co-precipitation of other metal ions present in solution. As a result, jarosite has to be impounded in controlled tailing ponds since exposure of such residues contaminated with heavy metals to atmospheric conditions cause environmental problems (Ismael & Carvalho, 2003).

### **2.6.1.2 Goethite process**

Goethite is the second most stable iron oxide, after hematite. It is the end product of many transformations particularly under hydrous soil conditions (Cornell & Schwertmann, 1996). It is the most stable iron oxide at high pH values (Figure 10) and its stability is enhanced by substitution of cations (Murad & Rojik, 2004). When goethite contains traces of aluminum (Al) it becomes more stable than hematite (Chen & Cabri, 1986). Polymorphs of FeOOH are divided into four categories, namely: goethite  $\alpha$ -FeOOH, lepidocrocite  $\gamma$ -FeOOH, akaganeite  $\beta$ -FeOOH, feroxyhyte  $\delta'$ -FeOOH (Cornell & Schwertmann, 1996). Of these minerals, goethite is the most common and most stable. The precipitation reaction is shown below (Ismael & Carvalho, 2003):



In this process, hot acid leach solution is sent to a pre-neutralisation stage where ferric iron is reduced to ferrous iron by adding ZnS concentrate. The solution obtained from the liquid-solid in the pre-neutralisation stage is transferred to the goethite precipitation stage where calcine is added to maintain pH at 2.5 (Claassen, 2002). For the process to be implemented, the iron concentration in solution should not exceed 1 g/l (Ismael & Carvalho, 2003). The ideal process conditions are temperatures between 80 and 90 °C, pH 2 to 3 (Ismael & Carvalho, 2003). The rate of precipitation is critical for the process; slow precipitation produces well-crystallized precipitate ( $\alpha$ -FeOOH) while higher rates of precipitation produces akagenite ( $\beta$ -FeOOH) precipitates. The formation of akagenite is undesirable since it is difficult to filter. Just like in the jarosite process, the acid produced needs to be neutralized. This process produces low volume of waste compared to the jarosite process. Unlike jarosite, goethite can be further treated in a pyrometallurgical process to produce an inert material that could be used in the construction industry (Ismael & Carvalho, 2003; Lahtinen et al., 2008).

### **2.6.1.3 Para-goethite process (PG)**

For the goethite process, the iron concentration has to be low; this is achieved by slow oxidation and the use of dilute solutions. The latter concept led to the development of the Paragoethite process. This process involves the addition of concentrated acidic ferric iron liquor to promote dilution; precipitation is then induced by the addition of calcine slurry (Loan et al., 2006). The residue produced is finely grained and poorly crystalline, so it is difficult to characterise but closely resembles goethite (Claassen et al., 2002). The operating conditions are temperatures between 70 and 80 °C, pH range of 3 to 3.2 as well as intensive mixing. Excessive mixing is

detrimental to the filtration residues. The major factors impacting the process operation is the presence of silica and high supersaturation (very low solubility of ferric iron). Compared to other processes, the PG process has the lowest capital cost and it is easy to operate. However, this advantage is offset by the poor solid-liquid separation characteristics and high calcine requirements for neutralisation. Moreover, the residues produced retain significant levels of zinc of non-calcine origin (Loan et al., 2006).

#### **2.6.1.4 Hematite process**

Hematite is the most stable iron compound in the natural environment, and hence the end member of transformations of other oxides. There are two polymorphs of  $\text{Fe}_2\text{O}_3$ : hematite  $\alpha\text{-Fe}_2\text{O}_3$  and maghemite  $\gamma\text{-Fe}_2\text{O}_3$  (Chen & Cabri, 1986). The latter is metastable with respect to hematite. The polymorph  $\alpha\text{-Fe}_2\text{O}_3$  is the one precipitated from iron bearing zinc solutions (Claassen, 2002). Hematite is formed by hydrolysis of ferrous sulphate:



In this process, ferrous is oxidised and precipitated under oxidizing atmosphere ( $\text{PO}_2 > 5\text{bar}$ ) and at high temperatures ( $T > 185\text{ }^\circ\text{C}$ ). The reaction is dependent on the acid concentration in solution; at concentrations above 65 g/l,  $\text{FeOH}\text{SO}_4$  is preferentially formed. Consequently, different mixtures of  $\text{Fe}_2\text{O}_3/\text{FeOH}\text{SO}_4$  are formed during this process depending on the initial iron concentration. The phase boundary between  $\text{Fe}_2\text{O}_3$  and  $\text{FeOH}\text{SO}_4$  can be displaced to a higher acid level and higher temperatures by the presence of zinc sulphate in solution. Iron removed as hematite can be used in cement, iron industries or as a pigment. The hematite process operate at high temperatures, this operating condition translate into high capital and operating cost. However, storage costs are minimised because hematite can be sold to cement makers. The quality produced by the process cannot be used in the steel industry since this industry requires pure hematite. Pure hematite is precipitated only from a pure iron solution (Ismael & Carvalho, 2003).

#### **2.6.2 Comparison of the various iron removal processes**

The choice of iron removal process is dependent on three factors: cost of operation, zinc losses and environmental considerations. The comparison for the above processes is shown in Table 7 below.

**Table 7: Comparison between the different iron removal processes (Claassen, 2002)**

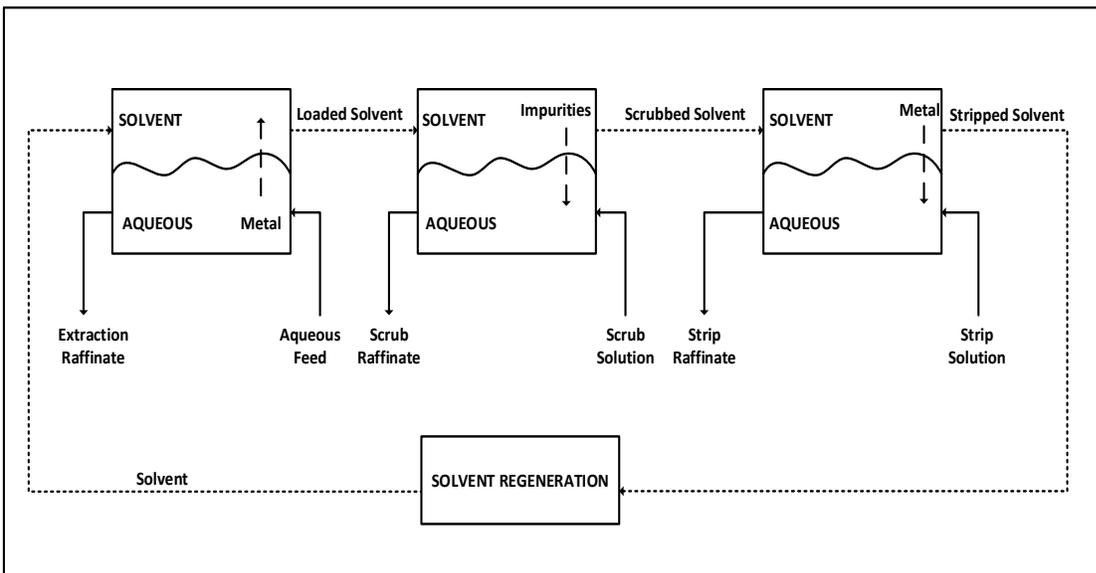
Variable	Item	Sub-item	PG	Goethite	Jarosite	Hematite
<b>Environment</b>	Ponding	Management	Easier	Easier	Difficult	Easier
		Rehabilitation	Safer	Safer	Difficult	Safer
	Bleed	Fluorine	Yes	Yes	Partially	No
	Residue composition	% Fe	35 - 38	40 - 45	25 - 28	58 - 60
		% S	4 - 7	2.5 - 5	10 - 12	2 - 4
		% Zn	8 - 9	5 - 8	4 - 6	0.5 - 1
<b>Economics</b>	Reagents	Oxygen (kg/t Zn)	None	32	None	More than goethite
		NH <sub>3</sub> (kg/t Zn)	None	None	9	None
		Acid (kg/t Zn)	None	None		More than goethite
	Energy	Steam (t/ Zn)	None to 1.2	1.2	1.55	More than goethite
			1.2			
	CAPEX		Low	Medium	Medium	High
OPEX		Low	Medium	Medium	High	
<b>Zn recovery</b>	Overall	% Zn recovery	94 – 95	96 – 97	97 – 98	98 - 99

In terms of the environment, the jarosite process is less desirable. Jarosites are difficult to dry and compact. Most residues are disposed at pH values higher than 2, jarosite is not stable under these conditions thus it transforms to goethite (Chen & Cabri, 1986). During the transformation, ions are liberated and there is a drop in the pH of the solution as well. This change has negative impact on the environment and makes the jarosite process even less attractive. Economically, the PG process is more favorable compared to the other processes. From an environmental and metal extraction perspective, the hematite process is preferable. However, the high CAPEX and OPEX for the process make it less desirable. Claassen (2002) recommend the goethite process as the best choice for greenfields operations; it has lower cost, more environmentally friendly residue as well as relative good zinc recoveries. The use of calcine in the process during

neutralisation could further increase zinc recovery thus making the process more attractive. This increase in recovery occur only if the calcine is dissolved, undissolved calcine leads to zinc losses.

### 2.6.3 Solvent extraction (SX) process

SX or liquid-liquid extraction, as applied to hydrometallurgy, comprises of three steps: extraction, scrubbing and stripping. The general solvent extraction flowsheet is shown in the Figure 11.



**Figure 11: General Solvent Extraction Circuit (Ritcey & Ashbrook, 1984)**

Extraction: a metal bearing aqueous feed (SX feed) and a solvent are fed, usually in counter-current flow into a contactor. The metal of interest is transferred from the aqueous to the organic phase. The extraction rate is dependent on the nature of solvent and extraction conditions chosen, in certain cases the metal of interest could be more or less completely transferred to the organic phase (Gupta & Mukherjee, 1990b). The extraction step has several contacting units in order to increase its efficiency. The aqueous phase from a contacting unit is sent to another contacting unit downstream, the reverse holds for the solvent phase; this network creates the counter-current flow. The aqueous phase (raffinate) from the extraction stage is further treated for the recovery of other metals or recycled to some point upstream or downstream from the SX circuit.

The loaded solvent goes to another contactor where it is scrubbed with a suitable aqueous solvent to remove impurities entrained or co-extracted in the extraction stage. This is known as the scrubbing (Sc) stage. The scrubbing solvent is either a solution of an acid or alkali salt (Gupta & Mukherjee, 1990b). The aqueous phase from the scrubbing stage (scrub raffinate) is mixed with SX feed in the aqueous solvent extraction feed tank. This stage also has numerous contacting units and the system is designed in order to have counter current flow between the aqueous phase and organic phase.

From the scrubbing stage, the loaded solvent passes through the stripping (S) stage in which the desired metal is stripped from the organic phase by a suitable aqueous solution, such as electrolyte, producing a highly concentrated solution of the metal. The lean aqueous stream is known as the strip solution or spent electrolyte and the rich stream is the strip raffinate or advance electrolyte. The stripped solvent is treated prior to entering the extraction stage. The treatment stage is for removing impurities that cannot be scrubbed in the scrubbing stage, yet they are co-extracted with zinc in the extraction stage e.g. ferric iron in D2EHPA solvent (Cole & Sole, 2003; Lizama et al., 2003; ). The different treatment methods will be discussed in detail below. Just like the stages above, the stripping stage has several contacting units.

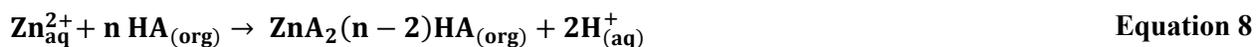
All the stages described above, extraction stage, scrubbing stage, stripping stage, require suitable use of either organic solvents or aqueous solvents. The nature of the extractable metal species determines the type of solvents to be used for the solvent extraction circuit (Ritcey & Ashbrook, 1984).

#### **2.6.3.1 Types of extractants**

The most common key impurities in zinc processes are: iron, copper, cadmium, manganese, cobalt and nickel (Cole & Sole, 2003). The type of zinc extractant used in a process is determined by the impurities present. A suitable extractant should have the following characteristics:

- Provide high zinc delta under fairly acidic conditions
- Selective for zinc over impurities
- Amenable to easy stripping of zinc

There are two types of acids used for extraction in sulphate media, namely: organophosphorus acids and carboxylic acids. Carboxylic acids are not commonly used for extraction in sulphate media because zinc extraction in these takes place at high pH values, yet most processes are at low pH ranges, as a result only organophosphorus acids will be discussed in detail below. Organophosphorus acids are divided into four: di(2-ethylhexyl) phosphoric (D2EHPA), a phosphonic-acid extractant (Ionquest 801), a phosphinic-acid extractant (CYANEX 272), and its dithio analogue, CYANEX 301 (Cole & Sole, 2003). The complexation of zinc by these extractants is represented by Equation 8 below:



Where HA represents the acidic extractant and  $n=2$  to 4. Compared to Ionquest 801 and CYANEX 301, D2EHPA and CYANEX 272 are more selective for zinc over cobalt and nickel. Moreover, zinc can be easily stripped from these extractants by moderate concentrations of sulphuric acid. However, all four extractants are not selective for zinc over ferric iron. Ferric iron complexes formed with the extractants are more stable than zinc complexes. Sulphuric acid can be used to strip ferric iron from Ionquest 801 and CYANEX 272. D2EHPA becomes progressively poisoned in the presence of ferric iron and has to be treated with concentrated HCl (5 – 6 M) (Cole & Sole, 2003).

CYANEX 301 has the highest zinc delta but due to poor stripping and lower selectivity for zinc over cobalt and nickel, it is not a viable option for zinc extraction (Cole & Sole, 2003). Amongst the four acids, D2EHPA is the only extractant feasible for the zinc process in sulphate medium. As mentioned above, it is more selective for zinc over impurities and zinc can be easily stripped by a spent zinc electrolyte from the organic phase (Balesini et al., 2011). Most of the metals that cannot be tolerated in zinc EW (e.g. copper, cobalt, nickel and cadmium) are not extracted by D2EHPA. The selectivity series of D2EHPA is shown in Equation 9 below (Flett, 2005):



Although D2EHPA does not have high zinc delta, the zinc transfer can be maximised by using high concentration (Cole & Sole, 2003). The use of this extractant requires the removal of iron from the PLS prior to SX.

### **2.6.3.2 Removal of ferric iron in D2EHPA**

As mentioned above, D2EHPA is not selective of Fe (III) over Zn. Fe (III) is hard to strip from D2EHPA whereas Fe (II) is easily stripped from the organic (Lizama et al., 2003). The easy removal of ferrous iron compared to ferric from the organic is due to the lower stability of ferrous in D2EHPA (Chang et al., n.d.). The build-up of ferric iron in the organic affects the efficiency of the extraction process, thus preventing the direct re-utilization of the organic (Lupi & Pilone, 2000). The conventional method to strip ferric iron is through the use of concentrated acidic solutions. Other processes include reductive stripping, galvanic stripping, and hydrolytic stripping. The Hydrozinc™ process uses galvanic stripping thus this method will be discussed in detail.

### **2.6.3.3 Stripping by concentrated acidic solutions**

The most common acidic solutions used are: hydrochloric acid, sulphuric acid and nitric acid (Lupi & Pilone, 2000). This method requires the use of concentrated solution and where lower concentrations are used the process is complex e.g. the use of combinations of extractants (O'Keefe, 1993). With HCl, an increase in acid concentration increases the stripping rate; however, the use of highly concentrated acidic solution such as 6M HCl can degrade the organic phase (Lupi & Pilone, 2000). The degradation of D2EHPA is not only detrimental to reuse but also blocks zinc metal dissolution (Chang et al, n.d.). H<sub>2</sub>SO<sub>4</sub> can also be used to strip ferric iron; however, this process is not feasible if iron is extracted from highly acidic solution. Compared to sulphuric acid, nitric acid is a less efficient stripping agent. This is due to weak ability for ferric iron to undergo aqueous complex formation with nitrate anion. The use of ammonium bifluoride (NH<sub>4</sub>HF<sub>2</sub>) solution increases the rate of stripping in D2EHPA but this process is not commonly used because of the many operational steps involved (Lupi & Pilone, 2000).

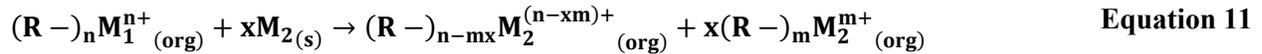
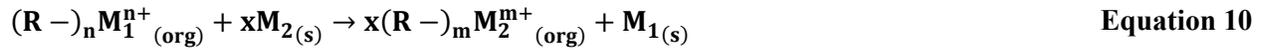
### **2.6.3.4 Reductive stripping**

Reductive stripping is achieved by gaseous reductants such as SO<sub>2</sub> and H<sub>2</sub>. These gases reduce ferric to ferrous hence the easy removal of iron in the organic (Chang et al., n.d.). This process has feasibility problems due to the high operational pressures and temperature (O'Keefe, 1993). Moreover, the SO<sub>2</sub> dissolved in the organic needs to be removed with water washes or 3% H<sub>2</sub>O<sub>2</sub> prior to reuse (Chang et al., n.d.). When using H<sub>2</sub> as a reductant, there is a need for hydrogen activation catalyst such as nickel.

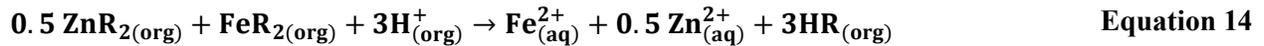
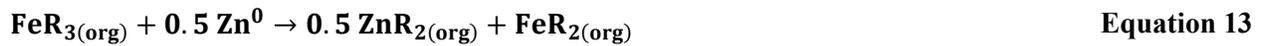
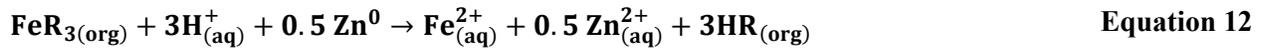
### 2.6.3.5 Galvanic stripping

Galvanic stripping involves the use of solid metal reductants in organic solvent system to spontaneously remove contained impurity ions. This process operates at ambient pressure and temperature (Chang et al., n.d.). Although organic solvents solutions are poor electrolytic conductors, research studies have demonstrated the feasibility of conducting redox reactions in solvent extraction organic solutions (O’Keefe et al., 2002). The organic solvents impart a high degree of polarization to the system and force the reactions to occur over a relatively short range because of the high solution resistivity. The presence of oxygen in the process is beneficial to initiate zinc dissolution however, it is detrimental to the efficiency of the process as the ferrous iron can re-oxidise.

Galvanic stripping can either be partial (Equation 11) or complete reduction (Equation 10) of an  $M_1$  cations by the less noble metal  $M_2$ , the reactions are shown below (O’Keefe, 1993).



The reactions are strongly influenced by system kinetics thus the degree and types of polarization inherent to the system often dictate the extent and efficiency of the reactions. Galvanic stripping can be carried out in two ways: simultaneous stripping or separate stripping. In simultaneous stripping, reduction and stripping occur in one stage. The organic solvent, solid metal reductant and the aqueous stripping phase react together before settling and separating. This technique is more effective for partial reduction separations (Cole & Sole, 2002b). In separate stripping, the reduction and stripping is accomplished in two stages. This technique is more efficient for low concentration cation impurity removal, metal deposition or seed crystal nucleation for microelectronic applications. Equation 12, Equation 13 and Equation 14 show the reactions for simultaneous and separate stripping respectively (O’Keefe, 1993).



For the application of galvanic stripping, the correct combination of organic, ion and metal is necessary. The selection criterion is based on the solubility and reactivity in the organic and the thermodynamic driving force generated by the two metal half cells (Cole & Sole, 2002b). The

removal of ferric iron from zinc sulphate electrolytes by galvanic stripping offers some attractive advantages. Ferrous iron (from the reduced ferric iron) can be easily stripped from an organic solvent and can further be recovered from the aqueous stripping phase into a form that could be of commercial value. The aqueous strip solution can be used in water treatment applications (Chang et al, n.d.).

### **2.6.3.6 Additives in solvent systems**

#### 2.6.3.6.1 Diluents

Diluent is an organic liquid used to dissolve an extractant in order to alter the extractant's physical properties for better mixing and separation with aqueous phases. Diluents decrease the viscosity of the extractant and provide a suitable concentration for metal extraction. They also improve dispersion and coalescence properties of the solvent (Ritcey & Ashbrook, 1984). Diluents are aliphatic, aromatic or a mixture of the two; the most used commercially is kerosene (Gupta & Mukherjee, 1990b). Diluents are as important as extractants because of the physical, chemical and economical effect they exhibit on the SX stage and thus proper selection is required. Below are the requirements of a diluent (Ritcey & Ashbrook, 1984).

- Be mutually soluble with an extractant or modifier
- Have high solvency for an extracted metal species, hence minimizing formation of third phases and low loading capacity of a solvent
- Have low volatility and a high flash point, thus decreasing losses due to volatility as well as fire hazards
- Be insoluble in the aqueous phase
- Have a low surface tension
- Be cheap and readily available

The extraction of zinc in sulphate media with D2EHPA uses kerosene as a diluent (Long et al., 2010). The organic concentration ranges between 20 and 30% v/v D2EHPA (Filippou, 2004).

#### 2.6.3.6.2 Modifiers

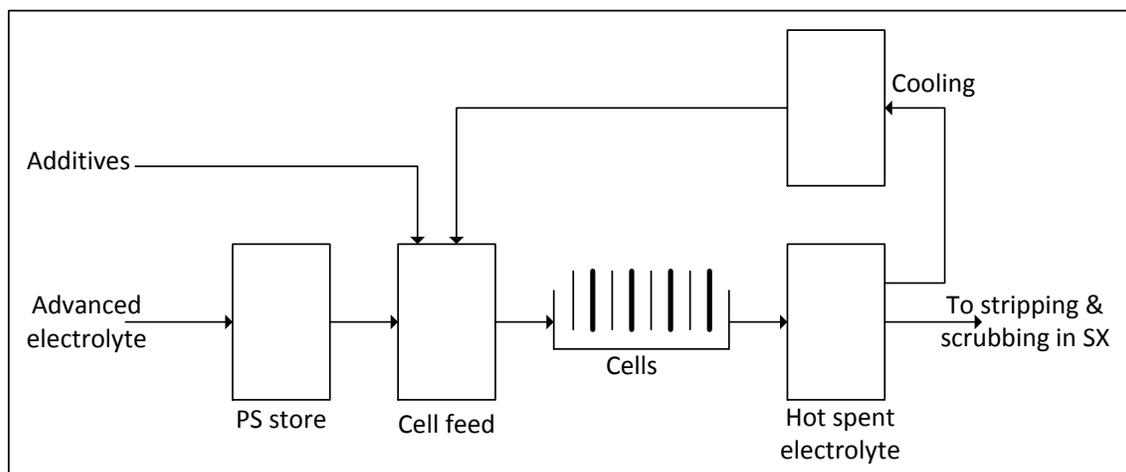
The use of diluents sometimes tend to form a third phase (two organic and one aqueous) in SX systems, this is most likely to occur when using aliphatic diluents. This formation is overcome by the addition of a modifier to the solvent. Modifiers improve the solubility of a metal complex

in the solvent system (Ritcey & Ashbrook, 1984). According to various sources, these reagents are either alcohols or extractants of a neutral category. The use of kerosene as a diluent does not require a modifier.

## 2.7 Metal recovery

### 2.7.1 Zinc Electrowinning (EW)

Electrowinning involves immersing metal cathodes and inert, conductive anode in an electrolyte and thereby applying an electric potential between the anode and cathode in order to plate a pure metal from the electrolyte onto the cathode. Most of the zinc metal produced is through the electrolysis of  $ZnSO_4$  solutions (Nicol et al., 2008; Gupta & Mukherjee, 1990b). The anode and cathode of the cell is made of lead-silver sheet and high-grade aluminium sheet respectively (Gupta & Mukherjee, 1990b). Figure 12 below show an electrowinning section of a plant.



**Figure 12: Simplified flowsheet for zinc electrowinning (Nicol et al., 2008)**

The cells used for Zn electrowinning are made of wood or concrete lined with lead and they are either connected in series or parallel. For series connection, cells are placed in cascade and the solution flows under gravity from one cell to another. The discharge from the last cell then goes to return-acid launders (Gupta & Mukherjee, 1990b). However, with parallel connection, each cell discharges directly to the return-acid launders.

#### 2.7.1.1 Electrowinning reactions

The cathode, cathode's side reaction, anode and overall reactions are shown below (Gupta & Mukherjee, 1990b):



The electrowinning products are: pure zinc metal at the cathode, oxygen at the anode and sulphuric acid is generated in the solution (Equation 17). The theoretical cell potential ( $E^{\circ}$ ) for the decomposition of water and  $\text{ZnSO}_4$  solution is -1.23 V and -1.99 V respectively. These values indicate that electrolysis of  $\text{ZnSO}_4$  solution would result in gaseous products,  $\text{O}_2$  and  $\text{H}_2$  at the anode and cathode respectively. Since the cell potential has to be maintained at a value above 1.23 V before Equation 18 occurs, this overpotential ensures that zinc deposition takes place in preference to hydrogen evolution at the cathode (Gupta & Mukherjee, 1990b). Once Zn has been deposited, the cathode with zinc is replaced with a clean one and then washed and stripped. The oxygen enters the atmosphere and the acid is re-circulated to the SX circuit.

In convectional cells, zinc deposition has 90% current efficiency and the power consumption is 3.67 kWh/kg of Zn (Gupta & Mukherjee, 1990b). The quality and type of zinc deposit is dependent on temperature, current density, pH, additives and presence of impurity cations and anions (Saba & Elsherief, 2000). Temperature and current density have substantial influence on the process.

### **2.7.1.2 Current density and temperature**

Zn processes either operate at low current density, 269 to 430 amperes/m<sup>2</sup> or high current density, 861 to 1076 amperes/m<sup>2</sup>. With each current density, there is a narrow range of acidity for the most efficient Zn deposition. When operating at high current density, a high Zn concentration (>170 g/l Zn), high acid concentration (200 g/l  $\text{H}_2\text{SO}_4$ ), a high purity electrolyte and rate of circulation are necessary. Most plants operate at low current densities and acidities (Gupta & Mukherjee, 1990b).

The ideal temperature range is between 30 to 40°C (Gupta & Mukherjee, 1990b). At high temperatures, current efficiency deteriorate and corrosion of lead anode by  $\text{H}_2\text{SO}_4$  is enhanced subsequently contaminating the cathodic zinc with lead.

### 2.7.1.3 Electrolyte

Zinc EW from sulphate media is highly sensitive to the presence of trace impurities and requires a highly purified electrolyte (Cole & Sole, 2002a). Therefore, it is necessary to bleed the electrolyte to control the composition. Typical compositions of advance electrolyte are shown in Table 8.

**Table 8: Representative values for advanced electrolyte (Cole & Sole, 2003)**

Element	Permitted concentration		Element	Permitted concentration	
	Advanced Electrolyte (mg/l)	Zinc cathode (%)		Advanced Electrolyte (mg/l)	Zinc cathode (%)
Zn	>90 000	99.995	Pb		0.0015
Mn	2000		Cu		0.001
Cd	<0.05	0.0015	Se	<0.1	
Co	<0.05		Cl <sup>-</sup>	<100	
Ni	<0.05		F	<20	
Sb	<0.02		Sn		0.001
Ge	<0.02		Al		0.001
Fe	<5.0	0.001			

Impurities affect the zinc electro-deposition process by altering the deposit's morphology and decreasing zinc current efficiency and the product purity (Mureşan et al., 1996b; Venkateswaran et al., 1996). At high concentrations they cause total re-dissolution of the deposited zinc metal (Ivanov, 2004). Zn has high occupancy in the Electromotive Force (EMF) series hence all metallic elements below Zn present in solution are co-deposited with Zn during electrolysis (Mureşan et al., 1996b; Gupta & Mukherjee, 1990b). The presence of copper and iron has more negative effects, even at low concentrations they decrease the current efficiency and worsen the quality of the electrodeposited zinc hence they both need to be removed during purification. These impurities serve as micro-cathodes upon which hydrogen is evolved and they negatively affect the cathodic current efficiency (C.E) and quality of deposited zinc (Saba & Elsherief, 2000). Other impurities such as Fe, Cu, Cd and As favour the re-dissolution of deposited Zn in the electrolyte (Gupta & Mukherjee, 1990b). Anions like Cl<sup>-</sup> and F<sup>-</sup> have deleterious effect as

well; these ions interact with the Pb anode and Al cathode respectively. To reduce the effects of impurities, the electrolyte is purified prior to electrowinning to decrease the concentration of metallic ions below a certain limit when their harmful effect is no longer present.

#### 2.7.1.4 Additives

During electrowinning there is acid formation; this has an effect on the quality of the cathode deposit and leads to acid mist formation (Saba & Elsherief, 2000). Moreover, the discharge of  $H^+$  affects the efficiency (Mureşan et al., 1996a; Venkateswaran et al., 1996). A decrease in hydrogen current efficiency increases zinc current efficiency and vice versa (Ivanov, 2004). Additives are added during the process in order to improve the cathode quality and to prevent the acid mist formation. Some organic substances such as triethyl-benzyl-ammonium chloride (TEBA) and hydroxyethylated-2-butyne-1,4 -diol (EAA) inhibit the reverse dissolution of zinc which is promoted by impurities (Mureşan et al., 1996a). Glue, saponin and antimony are the most commonly used additives. The presence of antimony in the electrolyte inhibits hydrogen evolution and as a result zinc re-dissolution is lessened (Venkateswaran et al., 1996).

## 2.8 Economics of zinc production

### 2.8.1 Market

According to the Zinc Investing News, (2014) the global zinc mine production remained unchanged from 2012; it was at 13.5 million tons by 2013, whereas the refined zinc production increased by 3 % to 13 million tons. Meanwhile, the world's consumption increased by 5 % to 12.9 million tons. The 2012/2013 Top 10 Zinc-producing Countries are presented in Table 9; these numbers are reported by US Geological Survey (USGS).

**Table 9: World's 10 biggest countries in refined zinc production (Zinc Investing News, 2013a; Zinc Investing News, 2014)**

Country	2012 zinc metal production (tons)	2013 zinc metal production (tons)
<b>China</b>	4 600 000	5 000 000
<b>Australia</b>	1 490 000	1 400 000
<b>Peru</b>	1 270 000	1 200 000
<b>India</b>	690 000	800 000

<b>United States</b>	748 000	760 000
<b>Mexico</b>	630 000	600 000
<b>Canada</b>	640 000	550 000
<b>Bolivia</b>	430 000	400 000
<b>Kazakhstan</b>	420 000	370 000
<b>Ireland</b>	345 000	330 000

China has been the world's largest zinc producer for the past two years and in each year the production accounts for more than 40 % of the global production. Despite China's rise in production for 2013, the country still experience production shortfall due to Chinese consumers spending record amounts on products that contain zinc (Zinc Investing News, 2014). According to the USGS, the United States was the leading consumer of zinc by 2012 using about one-seventh of the world's total output (Zinc Investing News, 2013a). The US is a net importer of zinc yet it has one of the world's largest zinc mines, Alaska Red Dog mine (Zinc Investing News, 2013a). However, China's continued industrialization has led the country to be the leading consumer of the metal.

The world's largest zinc mine is Rampura Agucha in Rajasthan, India. The open pit mine was first commissioned in 1991 and by end of March 2013 it had reserves of 110 million tons of ore. It mines 6.15 million tpa of ore which is milled into zinc and lead concentrate. Century mine in Northwest Queensland is the world's second largest and Australia's largest open pit mine. This mine produces zinc and lead concentrate. By 2013, it was producing between 480 000 and 490 000 tpa of zinc in the form of zinc concentrate (Zinc Investing News, 2013b). The mine is expected to close down by 2016 due to resource depletion (Barbir, 2012; Zinc Investing News, 2013b; Zinc Investing News, 2014). Red Dog mine in Alaska is also one of the world's largest producers of zinc concentrate; it represents 5 % of global zinc mine production and 79 % of the US zinc production. The open pit mine began operation in 1989 and initially had 77 million tons ore containing 17 % zinc (Zinc Investing News, 2013b).

Nyrstar is the world's largest zinc producing company; it is a collaboration between Australian and Belgian companies (Zinc Investing News, 2013b). It has nine mining operations and six

smelters; it produces 1.1 million tpa zinc metal and 475 000 tpa zinc concentrate (De Leeuw, 2011). However, the top zinc mines are not owned by this company. Table 10 and Table 11 below show the top companies that mine and produce zinc, respectively.

**Table 10: Top zinc miners by 2010 (PwC & MCX, 2011)**

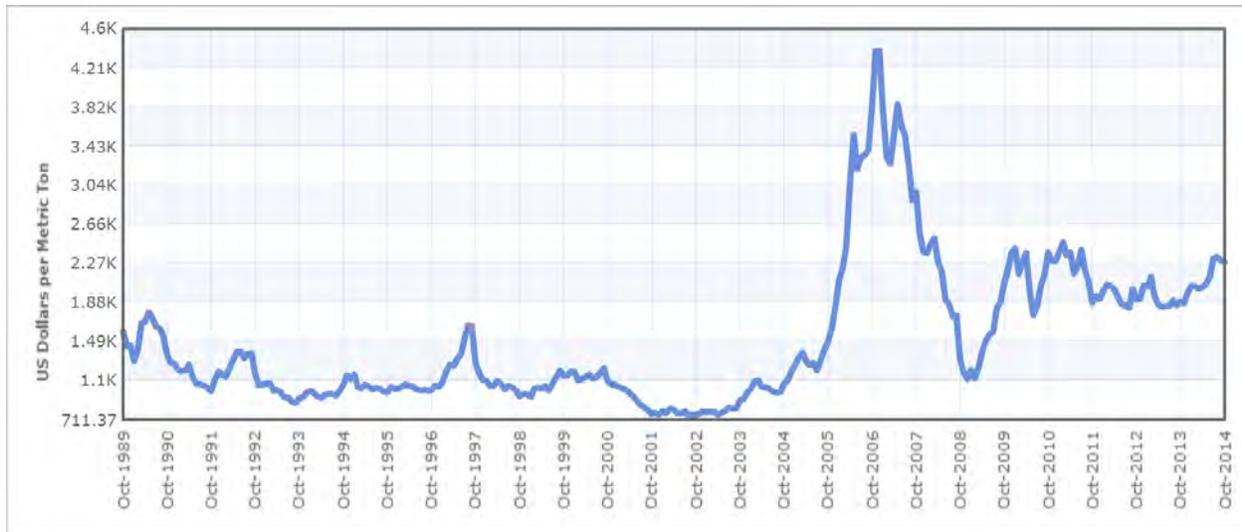
Name	Production (tpa)
<b>Glencore</b>	734 000
<b>Hindustan Zinc</b>	714 000
<b>Xstrata</b>	666 000
<b>Minmetals Australia (MMG)</b>	664 000
<b>Teck Resources</b>	646 000

**Table 11: Top refined zinc producers by 2010 (PwC & MCX, 2011)**

Name	Production (tpa)
<b>Nyrstar</b>	1 077 000
<b>Korea Zinc Group</b>	940 000
<b>Hindustan Zinc</b>	669 000
<b>Votorantim</b>	488 000
<b>Boliden</b>	456 000

### 2.8.2 Zinc selling price

One of the main places used to gauge base metal supplies as well as the overall performance of the sector is the London Metal Exchange (LME) (Burgess, 2006). It is one of the largest non-ferrous metals exchange in the world. Figure 13 below shows the price of zinc in US\$ / t of zinc from 1989-2014.



**Figure 13: Zinc prices for the period 1989-2014 (IndexMundi, 2014)**

There is an inverse relationship between LME stocks and zinc prices. At high stock levels the zinc prices are low due to abundant supply and at low stock levels the prices are high due to high demand. Generally, large amounts of zinc are stored in LME warehouses and when the price is high they get sold out, hence the decrease in stock levels with high zinc prices. In the past decades, the zinc market used to have brief periods of high prices that occurred nearly once in a decade. The lowest zinc price experienced so far was in 2002 (Figure 13); this was due to abundant zinc supply. During the period from 2005-2007, the LME had a deficit in their stock. By 2007, it had 108 kt which was estimated to be sufficient for only four days of global consumption (CPRA, 2007). This deficit was caused by a decline in production while demand was relentlessly increasing (Burgess, 2006). The rise in global demand and concerns over supply were the driving force towards the high zinc prices during this period.

It has been forecasted that zinc demand will outpace the supply; hence, high prices are expected in the future. The supply will decrease due to mine depletions and attrition; these results are presented in Table 12. CRU consulting has predicted a price of US\$ 4 000/t for the period of 2016/2017 (Zinc Investing News, 2014b). With zinc currently trading at US\$ 2 400/t spot on the LME, this represents an increase of 66 % in the next 2-3 years. Based on Figure 13 which shows an increase of 22 % over one year period, the forecasted increase is plausible.

**Table 12: Zinc mine depletions (and total attritions) between 2012-2016 (Barbir, 2012)**

Mine (location)	Zinc metal production (kt)
Century (Australia)	500
Brunswick (Canada)	200
Lisheen (Ireland)	167
Skorpion (Namibia)	162
Perseverance (Canada)	128
Pomorzany-Olkusz (Poland)	65
Mae Sod (Thailand)	45
Bairendaba Yindu (China)	45
Other depletions and attritions	402
<b>Total depletions and attritions</b>	<b>1 714</b>

### 2.8.3 Cost of innovation

Innovation comprises both risks and opportunities. Commercialization is the most costly part of product/process development; however, the economic benefits of an innovation are never fully realized until the innovation is introduced to the market (Luoma, 2008). To meet the sustainability global challenge, Moors & Dijkema (2006) recommends companies to implement corporate strategies and adopt novel technologies. However, these technologies need to produce better products and at a competitive cost in order to be marketable. Many great technological innovations never make it to the market even though some have been tested beyond the laboratory in small, pilot scale operations (Luoma, 2008).

Different organisations face different barriers and enablers to innovation. Primarily, base metals companies often have long-term contracts with their raw material suppliers and customers, or they form technological alliances with other companies. As a result, they keep using their conventional production processes because of these contractual obligations they have made. Moreover, the present industrial production systems (large scale of operations and the high degree of technical complexity) together with the mature technologies being used makes it difficult to implement new technologies (Moors et al., 2005). Presently, it is difficult to create new alternative technologies that fit within the existing infrastructure. Consequently,

manufacturers tend to favor the development that can be incorporated into an existing production process with only slight changes as well as improvement of existing production processes rather than a completely new process. For the zinc industry, intense process innovation has focused on producing zinc at lower costs (Moors & Dijkema, 2006).

New technologies require high investments and entail high financial risks (Fernandes & Paunov, 2012; Jurimae, 2010; Shi et al., 2008; Moors & Dijkema, 2006). Moreover, they require different types of training, learning, and regulatory legislation. Financial constraints play a prominent role in the rigidity of the conventional base metals production processes. In addition, there are risks related to safety, health and the environment associated with the use of as yet unproven technology. The capital intensity of large production installations also acts as a barrier for commercialization of innovative technologies. Most of the capital invested in large scale operations is usually recovered between 15 and 20 years, yet the company life span may be as long as 30 to 50 years and the commodities produced also have a long economic life span (Moors & Dijkema, 2006). It is more practicable to continue production in 20 years using installations for which the initial capital has been fully recovered. Hence, well established processes have large-scale advantages, and often are still very profitable, giving adequate returns on investments, after the machinery has already been depreciated. With most firms, quarterly profit-figures are becoming increasingly important, which could also hinder long-term decisions yet all capital investments for metal industry are long-term (Moors et al., 2005). All the above mentioned factors increase the perceived risk for base metals producing companies considering using innovative technologies.

#### **2.8.4 Concentrate marketing**

Most zinc mines do not own downstream processing facilities (smelter or refinery) to produce zinc metal. These types of mines are known as custom mines and they sell their concentrates to smelters. Concentrates are either treated on behalf of the mine (toll smelting), or they can be completely sold to the smelter (Outright sale). In the case of toll smelting, the refined metals can also be sent to any other location chosen by the mine (Lusinga, 2011).

The zinc concentrate composition varies; some of the metals found in the concentrate are useful as they are separated during smelting (Ag, Au) whereas others are cannot be recovered during

smelting (Fe, Mn, Si). Miners get penalized for non-recoverable metals present in concentrates and they benefit from recoverable metals (Ekaterina, 2012).

#### **2.8.4.1 Zinc contracts**

The pricing of zinc concentrates is categorized into three: pricing clause, treatment charge and penalties. These will now be discussed further

##### **Pricing clause**

According to Söderström (2008), zinc concentrate buyers pay for 85% of the final zinc content subject to a minimum deduction of 8%-units. The minimum deduction is used to compensate smelters for processing low grade concentrates.

##### **Treatment charge (TC)**

Reference TC (TCr) is the fee paid to smelters for the treatment of concentrates based on an agreed zinc selling price, this price is known as the base price. TCr is dependent on metal price and price escalators (Ekaterina, 2012; Teck, 2012). Price escalators are used to increase/decrease TCr due to a variation in the zinc price above/below base price. The current TCr is \$ 230/t of concentrate with a base price of \$2,000/t and the price escalators are 6% for zinc prices at \$2,000-2,500/t, 5% for a price at \$2,500-3,000/t, 2% for \$3,000-3,500 and flat for levels over that mark (Platts, 2014). De-escalators settled at -2% for a zinc price at \$2,000-1,500/t and flat for prices below that range. TCr, base price and price escalators are negotiated between miners and smelters on an annual basis. Benchmark terms are established when major mines and smelters have agreed on the annual terms.

##### **Penalties**

Penalties are applied for the impurities harmful to the zinc smelting processes. For instance, for each 1% unit by which the final iron assay exceeds 8%, miners pay a penalty charge of US\$2/t of concentrates (Söderström, 2008). Concentrates with high manganese content are regarded as unsaleable due to the manganese penalty. US\$1/t has to be paid for each 0.1% Mn above 0.5% (Kohler, personal communication 2014). This penalty combined with market volatility has made Gamsberg to remain unexploited because it has high manganese concentration.

#### **2.8.4.2 Revenue for concentrates**

The revenue received by the miners for their concentrates after smelting is evaluated using Equation 19-21 (Ekaterina, 2012).

$$Pc = \sum_{i=1}^n Pmi * Qmpi - TC * Qc - \sum_{k=1}^m Pk * Qc \quad \text{Equation 19}$$

Where  $Pc$  payment for the concentrate,  $Pmi$  metal price for the metal  $i$ ,  $Qmpi$  quantity of payable metal  $i$ ,  $TC$  treatment charge,  $Qc$  concentrate quantity,  $P$  penalty for non-recoverable metal,  $i=1, \dots, n$  where  $n$  is the number of metals recovered,  $k=1, \dots, m$  where  $m$  is the number of non-recoverable metals triggering penalties.

$$Qmpi = \frac{(Mci - ai)}{100} \times Qc \quad \text{Equation 20}$$

The quantity of the payable metal  $i$  is the difference between the metal content ( $Mci$ ) and the unit deduction ( $ai$ ), divided by 100, the variables  $Mci$  and  $ai$  are expressed in percentage. This result is then multiplied to the concentrate quantity. As mentioned above, the minimum unit deduction has to be 8% for zinc concentrates.

$$TC = (P_{Zn\_spot} - P_{Zn\_base}) \times Scale + TCr \quad \text{Equation 21}$$

The treatment charge is the sum of  $TCr$  and price participation which is caused by the variation in the zinc price above/below base price.

#### 2.8.4.3 Zinc smelters

Zinc smelter's revenue is the combination of  $TC$ , by products and free metal. The main by product is sulphuric acid. As mentioned above, miners receive 85% of payable zinc, the remainder counts as free metal and it benefits smelters. The value of the free zinc in zinc concentrates is dependent on the smelter's recovery rate and LME zinc price and is evaluated using Equation 22. The typical recovery rate for a zinc smelter is 95.5 % (Söderström, 2008).

$$\text{free Zn} = \text{Zn content} \times \text{LME price} \times (\text{recovery (\%)} - \text{payable zinc (\%)}) \quad \text{Equation 22}$$

#### 2.8.4.4 Vertical integration

Mining companies that own their smelting and other downstream processing facilities are known as vertically integrated companies. Vertical integration is very risky; it is complex, expensive, and hard to reverse, hence companies should not do it unless it is absolutely necessary to create or protect value. The four main reasons to vertically integrate are (Stuckey & White, 1993):

- The market is too risky and unreliable

- Companies with downstream processing have more market power than companies without it
- Integration would create or exploit market power by raising barriers to entry or allowing price discrimination across customer segments
- The market is undeveloped and the company must forward integrate to develop a market

Additionally, vertical integration may be the only economic means for mines that are located far inland since transporting of concentrates to custom smelters could be very expensive (Lusinga, 2011).

## **2.9 Assessing Profitability of Investments**

For projects to be profitable, the income earned during the plant's operational time should be greater than the investment made at the beginning of the project. As a result, the time value of money should be considered when evaluating profitability. According to Cohen (2009), profitability assessment takes three forms: time-based (how fast will the investment be paid back?), value-based (how much will the investment grow to?) and rate-based (what is the yearly return as a percentage of the investment). For these bases, discounted or non-discounted techniques are used for profitability evaluations (Cohen, 2009; Turton et al., 2008). However, non-discounted techniques do not account for the time value of money hence they are not recommended for evaluating new large projects (Turton et al., 2008). Due to the nature of this study, only the discounted profitability technique will be discussed.

### **2.9.1 Discounted profitability criteria**

With this technique, future cash flow values are discounted into the present and the resulting discounted cumulative cash flow is then used to evaluate profitability. Discount rate ( $i$ ) is determined by corporate managements and they represent the minimum acceptable rate of return ( $M_{ar}$ ) a company will accept for new investment. Many factors influence the determination of discount rates e.g. the level of risk on the project, Table 13.

**Table 13: Risk profiles for different investment types (Cohen, 2009)**

Investment type	Level of risk	M <sub>ar</sub> (%/ yr)
<b>Basic very safe investment</b>	Safe	4-8
<b>New capacity with established market position</b>	Low	8-16
<b>New product entering into established market or new process technology</b>	Medium	16-24
<b>New product or process in a new application</b>	High	24-32
<b>Everything new, high research &amp; development and marketing effort</b>	Very high	32-48+

**2.9.1.1 Profitability assessment**

**Time-based:** the payback period (PBP) is used for this criterion. This is defined as the time required after start-up to recover the fixed capital investment (FCI) for the project. The shorter the payback period, the more profitable the project is.

$$PBP = \frac{FCI}{\text{Average Annual Cash Flow}} \quad \text{Equation 23}$$

**Value-based:** the net present value (NPV) is used for this criterion. It is defined as the cumulative discounted cash at the end of the project. NPV is greatly influenced by the level of fixed capital investment.

$$NPV = \sum_{j=1}^n PWF_j \times \text{Annual Cash Flow}_j - \sum_{j=-b}^n PWF_j \times TCI \quad \text{Equation 24}$$

Where: *TCI* is the total capital investment (FCI + working capital (WC)) and *PWF<sub>j</sub>* is the present worth factor for year *j*. The PWF for year *j* is  $(1 / (1+i)^j)$ .

For better comparison of projects with different *TCI*, the present value ratio (PVR) is used (Turton et al., 2008). This ratio is calculated as shown in Equation 25.

$$PVR = \frac{\text{Present Value of All Positive Cash Flows}}{\text{Present Value of All Negative Cash Flows}} \quad \text{Equation 25}$$

A value of unity for PVR represents a break-even situation. The further away the ratio is from unity, the more profitable the process is, whereas ratios less than unity represent unprofitable projects.

**Rate-based:** the internal rate of return (IRR) is used for this criterion. It represents the highest interest/ discount rate for which a project can break even (NPV =0, PVR=1). A project is considered profitable when IRR is greater than  $M_{ar}$ .

$$0 = \sum_{j=1}^n \frac{1}{(1 + IRR)^j} \times \text{Annual Cash Flow}_j - \sum_{j=-b}^n \frac{1}{(1 + IRR)^j} \times \text{TCI} \quad \text{Equation 26}$$

When the time value of money is not taken into account, return on investment (ROI) is also used for this criterion. It represents the non-discounted rate at which money is made from FCI.

$$\text{ROI} = \frac{\text{Average Annual Net Profit}}{\text{Fixed Capital Investment (FCI)}} \quad \text{Equation 27}$$

The average net profit is calculated over the life of the plant after start-up.

## 2.10 Summary

The Gamsberg ore body has been described together with the problems associated with it. The three proposed routes to treat this ore body have been discussed in detail as well. Processes associated with each route and the operating conditions have been stated. Moreover, financial tools used to evaluate the feasibility of a process have been discussed. Doing a material balance and calculating the capital and operating costs associated with each route is the next phase of the study. The financial tools will then be used to evaluate the different process routes in the context of Gamsberg in order to recommend the better option to refine the high manganese and low zinc ore body.

## CHAPTER 3

### 3 Methodology

#### 3.1 Process routes

A simplified zinc leach flowsheet comparing the three routes being studied is presented in Figure 14 and the block flow diagrams for route 1 and 2 are shown in appendix A. Route 3 has not been shown in the appendix since it is a sub-section of route 2.

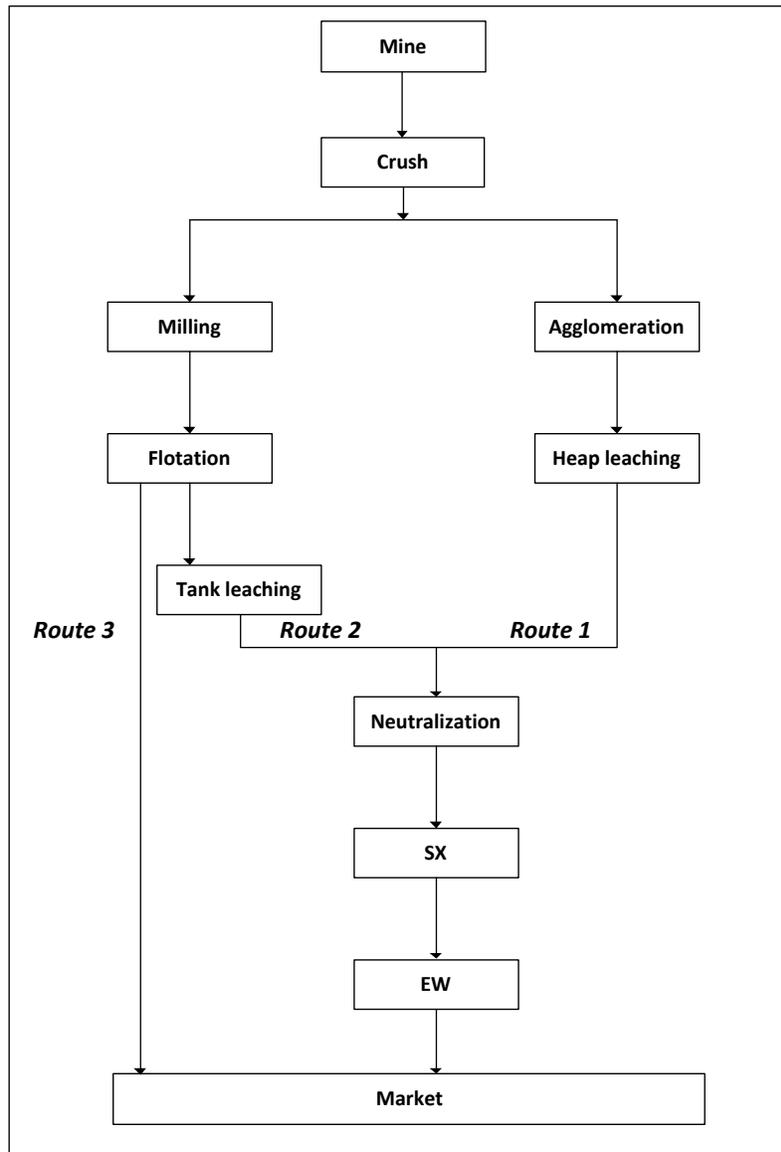


Figure 14: Process routes being studied. Solvent Extraction (SX) and Electrowinning (EW)

Route 1 comprises of heap leaching and refining locally. The refinery stages include neutralization (NT), solvent extraction (SX) and electrowinning (EW). To control impurity levels and remove excess water in the process, the bleed treatment (BT) stage is added. Route 2 comprises of crushing, milling, thickening and tank leaching, thereafter it follows the same refinery stages as Route 1. Lastly, Route 3 includes crushing, milling, thickening, and flotation to produce a concentrate and shipping it for toll refining in Europe.

## **3.2 Flowsheet development and material balances**

The financial analysis associated with each route has not been presented in the literature review, hence the need to develop the flowsheet for each route and establish the operating and capital cost models required in order to make a recommendation for the Gamsberg ore body.

### **3.2.1 Major assumptions**

The routes are compared based on the same amount of ore being mined; hence these routes have different production rates due to the different zinc extractions and recoveries. A mining rate of 3.4 million tpa is used as a basis to develop the model. Since the zinc composition varies within the ore body (6.27-12.6 wt%), the aim is to produce a minimum of 150 kt/a SHG zinc for route 1 with the lowest possible grade. This production rate is the same as Skorpion Zinc production. According to Stalder and Rozendaal (2004), the zinc ore has an average grade of 7.40 wt%, therefore, this zinc grade has been used for the model. Route 1 has a zinc production of 176 kt/y, Route 2 yields 204 kt/y and Route 3 has a Zn equivalent of 216 kt/y. Route 1 and Route 2 produces special high grade (SHG) zinc whereas Route 3 produces 472 kt/y of concentrates having 45.9 wt% zinc.

### **3.2.2 Route 1 assumptions**

#### **3.2.2.1 Heap leaching (HL)**

The heap leaching stage is developed using three particle class sizes A (+5.25/-6.75 mm), B (+14/-16mm), C (+23/-25mm). These are the same class sizes used by Ghorbani (2012) when conducting experiments with the Gamsberg ore. These experiments were done over 314 days, hence the same operation time has been assumed for the heap. The three class sizes A, B, C had zinc extractions of 79%, 69% and 56%, respectively. There is no literature on size distribution for heaps in the zinc industry, however, a study done by Ogbonna (2006) which investigated the effect of size class distribution on the recovery of copper concluded that a ratio of 0.6:0.2:0.2 for

0.5 mm, 2.5 mm and 5.0 mm class sizes, respectively, lead to higher copper extraction. Consequently, the same ratio of 0.6:0.2:0.2 for the mass of ore for A, B, C class sizes, correspondingly, is used to calculate the overall zinc extraction rate for the heap. The mineralogy of the ore before and after leaching the three particle class sizes is shown in appendix B. This mineralogy, together with the zinc extractions is used to calculate the extraction rates of the other metals present in the ore. The multiple reactions occurring in the heap are also shown in appendix B. The sulfuric acid make-up stream is calculated based on the concentration of acid to be recycled into the heap. The recycle stream should contain 15-30 g/l sulphuric acid.

### **3.2.2.2 Neutralization (NT)**

When refining the pregnant leach solution (PLS) from the heap leaching stage, it is firstly treated in the neutralization circuit. Limestone is added in this circuit to raise the pH and precipitate ferric as an iron hydroxide (goethite) and sulphate as gypsum. Most of the ferric iron present in the PLS is precipitated. Iron concentration in SX feed has to be less than 300 mg/L. Basic Zinc Sulphate (BZS) precipitate from the Bleed Treatment (BT) stage is recycled to this stage. All BZS precipitates are re-dissolved in this stage. The NT circuit has three stages in series; the tanks used for the three stages have a total residence time of 105 min with the pH of the last tank being 3.5 (Lizama et al., 2003). The neutralization slurry from the tanks is flocculated using aluminum hydroxide alum. The thickener underflow is filtered, washed in a countercurrent stage and then pumped to a residue pond for storage. The filtrate is recycled back to the neutralization tank. The thickener overflow is clarified and sent to SX stage. The amount of limestone and flocculent added in the stirred tank reactors and clarifier, respectively, is dependent on the reactions shown in appendix B. 98% of the limestone is utilized. The thickener and filter residues have 60-65 wt% solids and 85-90 wt% solids (Nicol et al., 2008).

### **3.2.2.3 Bleed treatment (BT)**

Some impurities are not co-extracted with zinc during solvent extraction, they are recycled back to the leaching circuit with the raffinate stream. In order to control impurity levels and overall water balance, the BT stage is added. The first unit for this stage is Copper cementation circuit. Zinc dust is added in stoichiometric amounts into a cementation tank, the equation for copper cementation is in appendix B. Cadmium, Cobalt and Nickel cementation is considered negligible due to the minor contents present in the ore, and hence there is no cementation stage for these metals, which is usually found in zinc processes. Copper is completely cemented out in the

cementation stage. The residence time for the tank reactor is 40 min. Products from the cementation circuit are filtered and the Cu/Zn cake is stocked. A pressure filter is used in this circuit. The filtrate is sent to a BZS circuit. The BZS circuit has two stages and the reactors have a residence time of 2 h in total. Limestone is used in the first stage to precipitate goethite and lime is used in the second stage for gypsum and basic zinc sulphate precipitation, reactions are shown in appendix B. Both limestone and lime are added at stoichiometric amounts. The BZS slurry is thickened and the thickener underflow is sent back to the NT stage. The thickener overflow is sent to an effluent treatment plant. Zinc concentration has to be less than 1.0 g/l for the thickener overflow stream. Wash filtrate from the neutralization stage is treated in this stage (Lizama et al., 2003).

#### **3.2.2.4 Solvent extraction**

Zinc extraction takes place in three stages in which an organic stream flows counter current to an aqueous stream. Zinc concentration in the SX feed is between 20 and 30 g/L with the extraction zinc delta being 20 g/L. 50% and 5% of zinc and iron, respectively, is extracted in the SX circuit. The iron concentration is maintained at less than 0.3 g/L as the electrowinning circuit is sensitive to impurities (Nicol et al., 2008). Mixer-settlers are used for the SX circuit, and the retention time for the mixer is 3 min. During extraction, phase separation does not reach completion and minor amounts of aqueous phase remains entrained in the loaded organic and vice-versa. The loaded organic is then sent to a scrubbing unit to remove entrained impurities in the organic stream. Process water and an electrolyte from the electrowinning stage is added to this unit. The scrubbing unit has an organic to aqueous (O: A) ratio of 20:1. The aqueous stream from this unit is mixed with the SX feed before the extraction unit. The organic stream is sent to the stripping unit where zinc is stripped from the loaded organic using sulphuric acid. The H<sub>2</sub>SO<sub>4</sub> concentration used for stripping zinc from D2EHPA is 180 g/L and the resulting advance electrolyte has a zinc concentration greater than 90 g/L. This electrolyte is then sent to the EW stage. The organic is recycled back to the extraction unit to complete the organic cycle. However, 10% (v/v) of the stripped D2EHPA goes to the iron removal circuit. This unit avoids the build-up of ferric iron in the organic stream. Zinc shot, process water and sulphuric acid is added to this unit. The ferric removal stage has an organic to aqueous (O: A) ratio of 3:1 (Filippou, 2004; Lizama et al., 2003). The reactions associated with the SX stage are shown in appendix B.

### **3.2.2.5 Electrowinning**

Conventional zinc electrowinning is used to produce special high grade zinc cathodes. Water and sulphuric acid are added to this stage. Water fed to this circuit is in stoichiometric amounts to the zinc being electrowon, the equation for the reaction is shown in appendix B.  $H_2SO_4$  added is, dependent on the amount of  $H_2SO_4$  present in the bleed stream. 5% (v/v) of the stream being recycled to the stripping stage of the SX circuit is bled off. This is done to ensure that impurity levels remain very low since this stage is very sensitive to impurities. Zinc delta for the circuit is 8 g/L. Based on the study done by Lizama et al., (2003), which had a 96.5 % solution to metal recovery for this process, the same recovery has been assumed.

### **3.2.3 Route 2 assumptions**

The assumptions for the SX, NT, BT and EW circuits stated above for route 1 stay the same for route 2. For the concentrator plant, ore is crushed, milled and then concentrated by flotation. The zinc recovery is assumed to be between 82-86% during the flotation stage (Kohad, 1998). The composition of the concentrates is shown in appendix B (Kohler, personal communication 2014). The concentrates are thickened and filtered before being leached. The Atmospheric leach process is used to treat the concentrate; this circuit consists of two leaching stages and thickeners. 98% zinc extraction is achieved in the leaching stage over a period of 16 hrs (Filippou, 2004). The extractions for the other metals in the concentrates are in appendix B, these values have been adapted from a study done by Fuls in 2011.

Route 3 assumptions are the same as the concentrator plant for route 2.

## **3.3 Capital cost estimates**

Capital cost each stage for the three routes has been shown in appendix C.

### **3.3.1 Heap**

The costs associated with heap leaching are based on the foot print of the heap. These include leach pad cost, irrigation and aeration lines. The heap design takes into account the angle of repose of  $26^\circ$ , ore density and shape (frustum of a pyramid) (Breitenbach & Thiel, 2005; Van Zyl et al., 1990). The cost of stacking ore into a heap is based on the amount of ore mined per day, this capacity is compared to an existing plant and the six-tenth rule is used to estimate the cost of stacking Gamsberg ore (Cohen, 2009; Kappes 2002).. The cost of irrigation lines and aeration lines are based on data sourced from Kappes (2002) which represents them per square meter.

Knowing the heap foot print, the total cost can thus be calculated. The above-mentioned costs and the detailed calculations for the foot print of the heap are shown in appendix C.

### **3.3.2 Concentrator plant**

The costs for the concentrator are calculated using the O'Hara Method for processing plants (Mular, 2002). These costs are based on the throughput for each stage. This method accounts for different factors depending on the stage of the concentrator plant, e.g. for the flotation section, it account for the ore type if its low/high grade ores or complex ores. The different factors used in each stage is dependent on the location, nature and type of the deposit discussed in section 2.1. A site factor of 1.5 and a rock factor of 1.0 has been assumed for plant clearing/ excavation and for the type of foundation support required for the deposit, respectively. The grinding factor has been assumed to be 1.5. The process factor for the flotation and thickening/filtration stages has been assumed to be 1.6. The above assumptions are based on literature sourced from Environmental Resources Management, (n.d.a), Environmental Resources Management, (n.d.b) and Stalder & Rozendaal, (2004). The summary of the plant cost estimates together with the factors of each stage are shown in appendix C.

### **3.3.3 Refinery**

Sizing and costing of all mechanical equipment is based on the material balance. Settling and filtration rates sourced from Seider et al., (2003) were used when sizing thickeners and filters. 1,500 lb/d-ft<sup>2</sup> and 5,000 lb/d-ft<sup>2</sup> is assumed for fine particles (produced by precipitation) and coarse particles, correspondingly. For mixer-settlers, the cost is based on the diameter and orientation of the tank. It has been assumed that mixers are horizontal liquid drums with a residence time of 3 min and an H/D ratio of 1. Stirred tanks reactors are assumed to be vertical and have an H/D ratio of 2. Due to the orientation of the mixing tanks, inside diameter (Di) has to be less 12 ft and 21 ft for horizontal and vertical tanks, respectively (Seider et al., 2003). Settler costs are based on the settling area. The settling rate is assumed at 4 m<sup>3</sup>/h/m<sup>2</sup> (Anderson et al., 2002). The cost for agitators is based on the horsepower drawn by the turbine and this horsepower is dependent on the function of the turbine. 0.985 kW/m<sup>3</sup> and 1.97 kW/m<sup>3</sup> is need for liquid-liquid extraction and suspension of solids particles, respectively (Seider et al., 2003). Electrowining and melting and casting circuits costs are based on costs of existing plants and due to the difference in capacities, the six-tenth rule was used to calculate the costs for the plant (Cohen, 2009; Lizama et al., 2003). In estimating the cost of the plant equipment, the limiting

size of individual units of equipment has been recognized. Therefore, duplication has been accounted for where necessary. Detailed costs are shown in appendix C.

#### **3.3.4 Total fixed cost**

Detailed factorial estimates have been used to calculate the total capital cost of the plants from the total purchase costs of major equipment. These estimates include the cost of equipment erection, piping, electrical, buildings, utilities, storages, site-development and ancillary buildings (Cohen, 2009). The values associated with each estimate are shown in appendix C.

### **3.4 Operating cost estimates**

Operating costs consist of reagent costs, power consumption, depreciation and maintenance. The operating costs for the concentrator plant are based on correlations which are presented as either equations or graphs. Therefore, there is no material balance done for reagents used in this section since these costs are dependent on the ore/concentrate flow rates. The milling cost curve is shown in appendix C. The overall plant maintenance and depreciation is estimated as 2% and 10%, respectively, of the fixed capital costs. The effective life of all fixed assets is ten years. The cost of raw materials used is estimated from the mass balance and the cost per ton for the raw materials used. The operating expenditure of each stage for the three routes has been shown in appendix C.

### **3.5 Financial analysis**

PVR, IRR, NPV and PBP are chosen as financial indicators to gauge the feasibility of the three routes. The ramp up capacity is 30% and 60% for year 1-2 and year 3-4, respectively, full production start in year 5. The plant is assumed to have an operational time of 20 years (Environmental Resources Management, n.d.a). The zinc price escalation is calculated to be 6.75 % per annum over the 20 yrs, based on the yearly average zinc price change for the past four years. The annual escalation rates for the cost is 6.25 %, this is based on the 2014 average inflation rate for South Africa (Inflation.eu, 2014). In African countries, green field projects usually have discount rates between 10-13% (Ledgerwood, personal communication 2014). A discount rate of 13% has been used for the analysis. Ideally, the three projects will have different discount rates due to the different risk levels associated with them. However, for better comparison of NPVs, one discount rate has been used for all three routes. The yearly cash flow analysis for the three routes have been shown in appendix C. The sensitivity of electricity price,

zinc price, FCI, WC and construction time on NPV is evaluated. A Matlab code for the sensitivity of zinc price has been shown in appendix C as a sample for the sensitivity analysis.

Detailed cost estimates are shown in appendix C and assumptions used for the cost analysis are shown in Table 14.

**Table 14: Parameters considered for cost analysis**

Parameter	Value
<b>Zinc Price</b>	2, 780 \$/t
<b>Exchange rate</b>	10.4 R/\$
<b>Tax rate</b>	28 %
<b>Discount rate</b>	13 %
<b>Escalation rate</b>	6.25 %
<b>Depreciation</b>	10 years
<b>Price escalation</b>	6.75 %
<b>Scrap value</b>	0

## CHAPTER 4

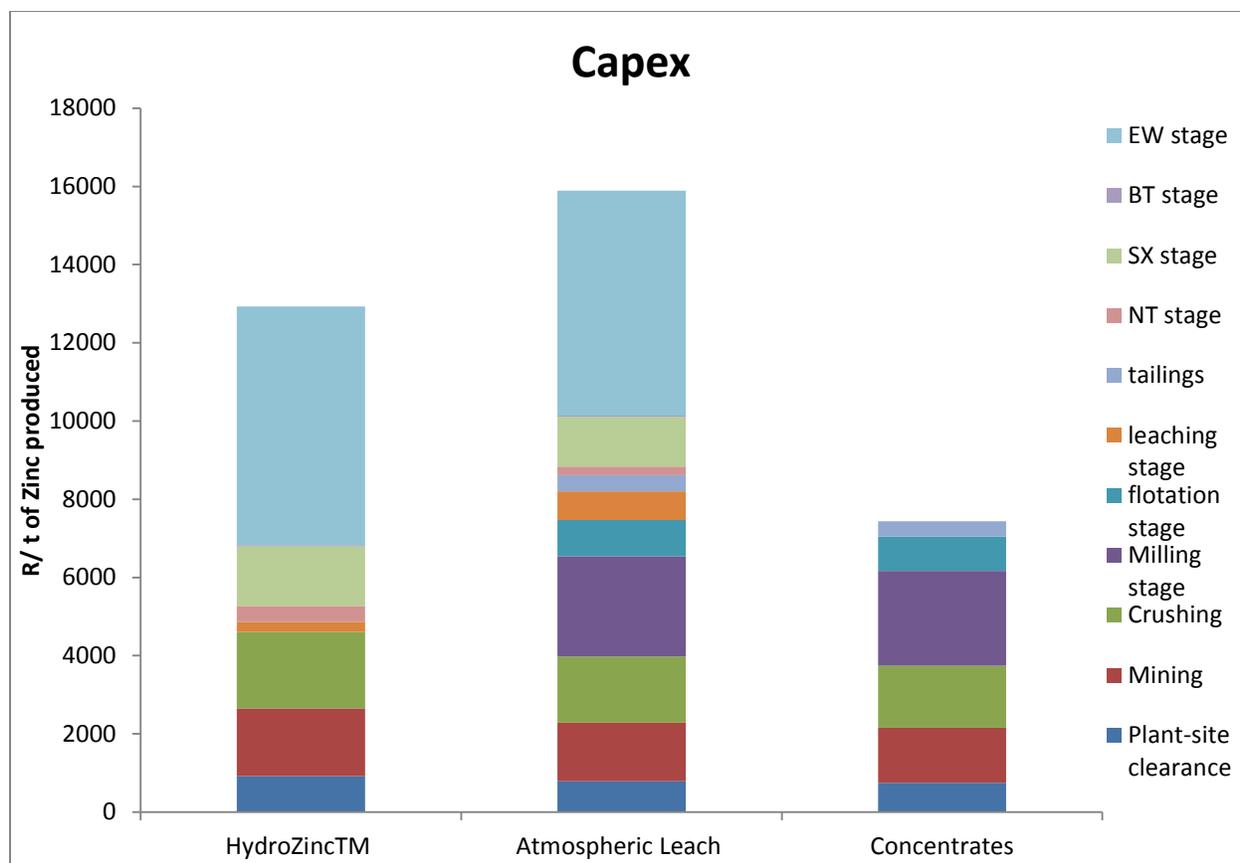
### **4 Results and discussion**

This chapter outlines the costs associated with the processes discussed in chapters 1-3 above and the factors that affect the economic performance of these processes. However, due to inflation and other factors which change cost prices, such as differences in currency values between countries, it is difficult to provide cost figures that will remain constant with time. All data presented is expressed in South African Rand as of January 2015, it can be adjusted by the cost index (CI) rates from this date.

#### **4.1 Capital and operating cost estimates**

##### **4.1.1 Capital costs**

Capital costs play a critical role in influencing the choice of process technology (du Plessis et al., 2006). Capital costs provided for the processes below are for a green field zinc plant. For estimating the cost of the different plant units, the limiting size of individual units and their orientation was recognized and hence duplication was included where necessary. Figure 15 illustrates the breakdown of the capital cost (Capex) for the three processes.



**Figure 15: Capital costs for the three process routes. EW (Electrowinning), BT (Bleed treatment), SX (Solvent extraction) and NT (Neutralization)**

For the heap, the ore is crushed to -25 mm and stacked in 6 m lifts on a double composite liner with leak detection and collection. The heap has a foot-print of 151 000 m<sup>2</sup> and the exterior ore lifts slopes are constructed at an angle of repose of 26.6 °. For tank leaching, the ore is milled to -150 µm.

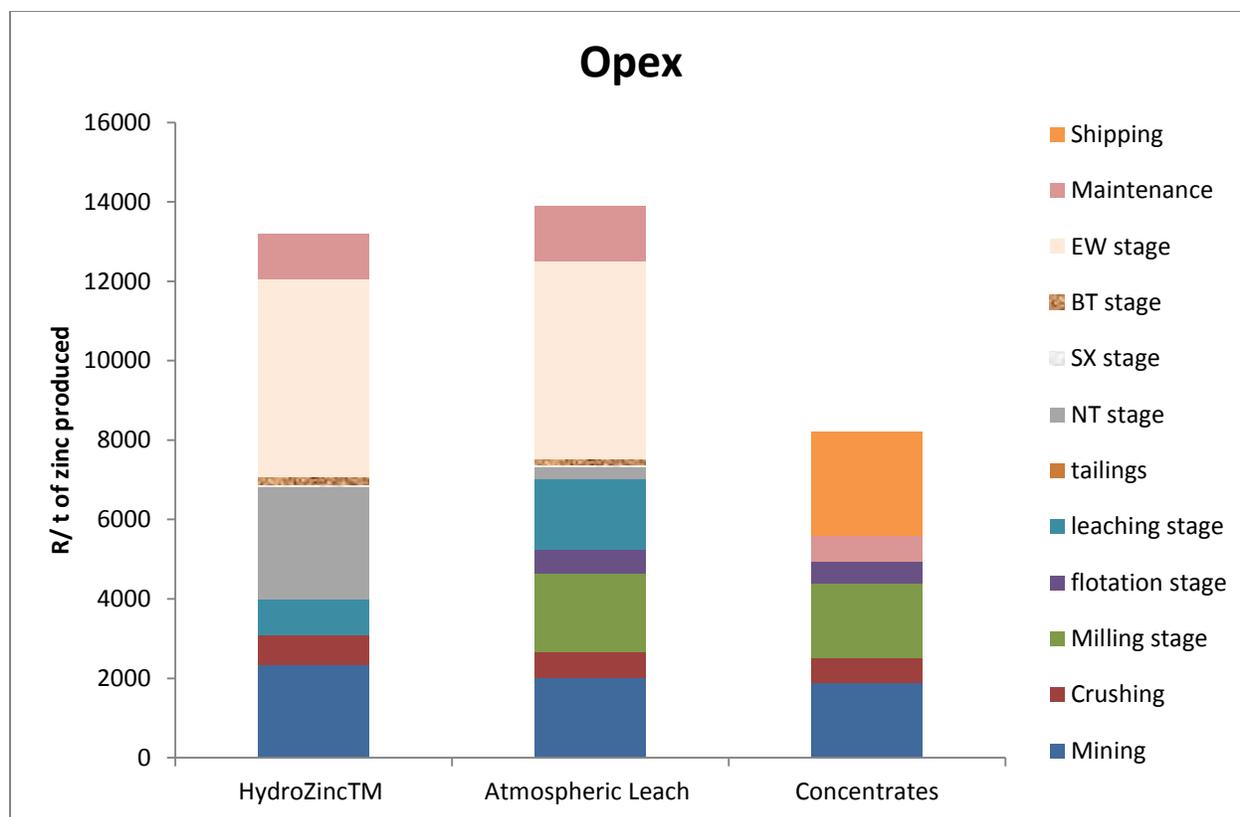
As stated by Stuckey & White (1993), vertical integration is expensive and companies should not do it unless it is absolutely necessary to create or protect value. This is noted from the costs of route 1 and 2 which are much higher compared to route 3. The electrowinning (EW), crushing and milling stages determine the major capital costs. This is due the type of equipment used in these stages. Mills, crushers and furnaces are more expensive compared to tanks and thickeners which are used in the other stages. A significant amount is required for the EW stage due to the many process components: electrolytic cells including anodes, rectifier, process tank, filters, cooling system and pumps. The materials of construction used for this stage are also costly,

anodes are made of lead and they are coated with lead-silver alloy. Titanium is used for cathode blanks onto which the zinc is plated. Titanium is the most expensive material of construction (Seider et al., 2003). Unlike the other stages, this stage requires a separate cell house to be built. All the factors mentioned above contributes to the high capex for EW. As a result, route 2 is the most capital intensive since it has both the milling and the electrowinning stages.

Although Solvent Extraction (SX) and Neutralization (NT) stages have similar equipment (mixing tanks), the SX stage is more capital intensive. This is due to the multiple mixing units within the SX stage. The SX stage has eight mixer-settler units and the NT stage has three units (Lizama et al., 2003). The capex for SX and NT stages are higher in route 1 than route 2 due to the fact that streams in this system are more dilute, and hence there are higher flow rates in route 1. The zinc concentration in the PLS for route 1 and 2 are 19.1 g/l and 73.2 g/l, respectively. The Bleed Treatment (BT) stage is the least expensive even though it has many units within it, due to the small flow rates of the bleed stream. Only 5% of the filtrate from neutralization clarifier is bled off from the process. As anticipated, the capex for heap leaching stage is significantly lower than the tank leaching stage. Heaps are known to have low capex compared to agitated leach tanks (Ghorbani et al., 2011; John, 2011; Kappes, 2002).

#### **4.1.2 Operating costs**

Operating costs are usually divided into two: manufacturing (includes depreciation, raw materials etc) and general expenses (includes administration, marketing, etc). Only manufacturing costs were considered for the model. Power is estimated from the consumption figures for each stage and the power cost is taken as 86 cents per kWh (Eskom, 2014). The major power cost is for electrolysis and will not vary greatly between route 1 and 2. However, consumption from the other areas varies significantly, depending on the area and the type of process. Figure 16 shows the operating cost (Opex) for the processes.



**Figure 16: Operating costs for the three process routes. EW (Electrowinning), BT (Bleed treatment), SX (Solvent extraction) and NT (Neutralization)**

The cost of treating zinc concentrate locally is R 7,260/ t of zinc produced and the charges when refined in Europe is R 6,140/ t of zinc (base charge, penalties and price participation). As noted from the two charges, refining concentrates locally is expensive. The two charges are not competitive. Another major drawback with refining zinc locally is the high capital that has to be invested, as compared to the smelters in Europe which are old plants that already have their initial investment recouped. As a result, sending concentrates for refining elsewhere has an advantage over on-site refining. However, in remote areas where concentrate transport costs are high, on-site concentrate leaching is economically attractive. When the variable cost is lower than fixed operating costs, the contribution to profit is greater. Companies with higher fixed costs relative to variable costs have a high operating leverage and they turn out to be more profitable (Harvard Business School Press, 2002a). The fixed operating costs for route 1 and 2 are much higher than route 3, since there is still depreciation to be accounted for and these processes are capital intensive.

As with capital costing, the EW stage has the highest operating cost. This stage requires considerably more electrical energy compared to other stages. Reduction is energy intensive, between 3000 and 4000 kWh/t of zinc is required (Nicol et al., 2008; Gupta & Mukherjee, 1990b). Mining, crushing and milling stages are also energy intensive, and hence there are high operating costs associated with them. As expected, a heap has lower operating costs than tank leaching. Tank leaching has significant operating costs due to the energy required by motors to drive mixing turbines. The energy drawn by turbines is relative to its application e.g. blending of liquids require less energy compared to mixing solid and liquid particles (Seider et al., 2003). Turbines for the leaching stage require  $1.97 \text{ kW/m}^3$  yet turbines in SX mixers require  $0.985 \text{ kW/m}^3$ , the volume is  $232 \text{ m}^3$  and  $43.5 \text{ m}^3$ , respectively.

As the volumetric flow rates of route 1 are higher as compared to route 2, the energy drawn for SX, BT and NT stages will be higher for route 1 than route 2. Consequently, route 1 has higher operating cost for these stages. Even though the extraction rates for Fe are higher in route 2 than route 1, 85% and 45%, respectively, the operating cost for NT are higher in route 1 than 2. This is due to the high Fe concentration in the PLS, which is a result of the higher concentration of Fe in the ore compared to the concentrate, where much of it has been removed in the flotation stage. As the concentrate will be exported to Europe in route 3, high shipping cost are associated with it.

## **4.2 Ranking the process options**

Net Present Value (NPV), Present Value Ratio (PVR) and Internal Rate of Return (IRR) are used to rank the three routes in order to recommend the best option for Gamsberg. These results are shown in Table 15. As mentioned in the literature review, for a project to be profitable, NPV should be greater than zero, PVR should be greater than unity and IRR should be greater than or equal to the Minimum Acceptable Rate of Return ( $M_{ar}$ ).  $M_{ar}$  is the rate of earning that must be achieved by an investment in order for it to be acceptable to an investor and it differs with the risk level of each investment. When ranking process options, more emphasis is placed on IRR and NPV (Shapiro, 2004a; Johnson, 1999a). The preferable route for Gamsberg will be a route that requires minimum investment and produces a higher return on investment. However, if the incremental capital associated with another route having a larger investment can be justified with respect to its incremental savings then the route with larger investment will be chosen.

Route 2 and route 3 are well established technologies, and hence these investment are associated with low level of risk. The description on risk levels is shown in Table 13, Chapter 2. A  $M_{ar}$  of 12 % is considered for these two routes. Although heap leaching is a well-established technology in the copper industry, it still remains a novel idea in the zinc industry, and thus the HydroZinc™ process is associated with medium level of risk. Therefore, a  $M_{ar}$  of 16 % has been considered for route 1.

**Table 15: Rankings for the three routes**

	Route 1 (HydroZinc™)	Route 2 (Atmospheric Leach)	Route 3 (Concentrates)
<b>Zinc production (kt/y)</b>	176	204	216
<b>TCI (R,Billion)</b>	12.4	15.0	7.55
<b>PBP (Years)</b>	2.76	3.33	3.14
<b>NPV (R, Billion)</b>	1.57	0.23	1.09
<b>IRR (%)</b>	14.4	13.2	14.6
<b>PVR</b>	1.14	1.02	1.16

Ranking by NPV is in dimensional form hence it is not helpful in judging the strength of an investment (Lang, 1989). However, IRR and PVR give results in ratios thus enabling judging the relative strength of an investment and the strength of increments of investment. Consequently, all three financial indicators are used for the ranking process. Based on the results in Table 15, route 3 has the lowest Total Capital Investment (TCI). Consequently, it could be expected that route 3 will have the shortest pay-back period (PBP), but this is not the case with the results above. This is due to the low zinc revenue generated by route 3 given the low zinc content in the concentrates as well as the high manganese penalties associated with Gamsberg concentrates.

The ranking order by NPV for the three routes is route 1, followed by route 3 and then route 2. However, the ranking order by IRR and PVR is route 3, followed by route 1 and then route 2; this ranking differs from the ranking by NPV. Contradictions between IRR and NVP rankings are known to happen for projects requiring different amounts of capital investment or projects having different lives (Shapiro, 2004b; Harvard Business School Press, 2002a; Johnson, 1999a;

Lang, 1989). The inconsistency in this case is caused by different amounts of capital investment required for each project. It can be resolved by using the weighted PVR average and incremental investment analysis (Johnson, 1999a).

#### **4.2.1 Resolving contradictions between IRR and NPV rankings**

Due to the discrepancy in the results between the method used for ranking the alternatives, route 1 and 3 are further investigated. PVR and incremental investment analysis is used to resolve this discrepancy.

##### **4.2.1.1 PVR**

From Table 15, it appears that route 3 is slightly more favourable than route 1 since the PVR values are 1.16 and 1.14, respectively. However, the initial investment for the two processes differs significantly. If route 3 is selected, the R 4.85 billion (difference between route 3 and 1 TCI) not required for investment is most likely to be invested in the capital market at the same rate as the discount rate (13%). As stated above, when the return is equal to the investment, the value of PVR is unity. The results on weighted average for the two processes are shown below.

$$\text{Route 1 PVR} = \left(\frac{12.4}{12.4}\right) \times (1.14) = 1.14 \quad \text{Equation 28}$$

$$\text{Route 3 PVR} = \left(\frac{7.55}{12.4}\right) \times (1.17) + \left(\frac{4.85}{12.4}\right) (1) = 1.10 \quad \text{Equation 29}$$

From the results above, route 1 is the more desirable alternative. Investing R 12.4 billion in route 1 implies receiving R 1.14 per Rand invested whereas investing R 7.55 billion in route 3 and investing R 4.75 in capital market implies receiving R 1.10 per Rand invested. This analysis is done to resolve conflicts between financial indicators before making a decision on an investment and is more useful provided there is no limited capital budget (capital rationing). Based on the results above, NPV and weighted PVR result in the same ranking for the three routes. However, if there is capital rationing (the more likely scenario), the incremental outlay of R 4.85 billion for the HydroZinc<sup>TM</sup> process relative to making concentrate needs to be considered for decision making.

##### **4.2.1.2 Incremental investment**

Incremental investment analysis identifies the difference in net cash flows between projects (the project with lower investment cost is subtracted from the project with higher investment cost) and the resultant cash flow is discounted and then analyzed in terms of IRR, PVR and NPV

(Johnson, 1999a; Lang, 1989). The lower investment cost project is known as the defender and the higher investment cost project is known as the challenger (Lang, 1989). In this case, the difference in cash flows for route 1 and 3 is used. These results are shown in Table 16.

**Table 16: Comparison of route 1 and 3**

Defender	Challenger	TCI (R, Billion)	NPV (R, Billion)	PVR	IRR
Route 3 (Concentrates)	Route 1 (HydroZinc™)	4.85	0.484	1.11	14.2%

According to Johnson (1999b), the higher investment project should be accepted if and only if the incremental NPV is positive, PVR is above unity and IRR is greater than the discount rate (i). Based on results in Table 16, route 3 is rejected in favour of route 1 because of the positive NPV, IRR above the discount rate and PVR is above unity. The combination of route 3 and the incremental difference serve as route 1. For projects with different TCI, if the incremental difference is acceptable, then the larger project should be accepted (Johnson, 1999b). The ranking order when using IRR and PVR is now identical to NPV, that is, route 1, followed by route 3, with route 2 excluded from this analysis.

#### **4.2.2 Shareholder value**

Investment decisions are dependent on the shareholder's objectives for an investment, and hence when making such decisions, there is an obligation to act in the shareholder's best interest. In practice, most investment decisions are based on maximizing the wealth of shareholders. In order to reach this goal, investment analysis should place higher weight on earlier cash flow than on the more distant future. Moreover, an investment should penalize heavily the expected cash flows of riskier projects and value all cash inflows and outflows associated with the project. Companies that do not focus on creating shareholder wealth are likely to have a value gap, the difference between the value of the company if it were optimally managed and the actual value of the company. Such companies are prime takeover targets and candidates for a forced corporate restructuring, therefore, maximizing the shareholder value provides defense against hostile takeover. This principle of maximizing shareholder's wealth thereby provides a rational guide to financial decision making (Shapiro, 2004b).

NPV is the only value that is consistent with the goal of shareholder wealth maximization, it accounts for risk, time value of money and focuses on cash flows, thus evaluating projects the same way as investors do (Shapiro, 2004a). As a result, NPV overcomes the deficiency noted in other financial indicators (Johnson, 1999a). The main drawback with applying NPV is computing the proper discount rate. When faced with contradiction in rankings between financial indicators, and there is no capital rationing, the best project is the one that adds more value to the firm, thus NPV is the better criterion to use. However, if capital rationing exists, PVR then provides superior investment decisions (Shapiro, 2004a; Johnson, 1999a; Lang, 1989).

### **4.3 Energy crisis in South Africa**

Since late 2007, the country started experiencing blackouts as demand exceeded supply, Eskom has been struggling to meet the electricity demands costing the economy billions of rands in lost production and economic growth (Vecchato & Mantshantsha, 2013; Wikipedia, 2009). Furthermore, it has been forecasted that the security of electricity supply will continue to be problematic due to plant decommissioning (Vecchiato & Mantshantsha, 2013). Previous tariff rates have shown an increase in electricity price over the previous years since 1988 (Eskom, 2013; Ramokgopa, 2007; Ramokgopa, 2001). This tariff history coupled with the forecasted insecurity of electricity supply makes it improbable for the electricity price to decrease in the future. Therefore, electricity prices are expected to increase rather than decrease.

Currently, two power stations are being built (Kusile and Medupi) which are expected to be complete by 2018. Unit 6 of Medupi power plant was supposed to be synchronized in the national grid in December 2014 with the remaining units following at nine-month intervals (Eskom, 2014; News24, 2014a; News24, 2014b). This date was a year later than the previous committed date by former Eskom CEO, Brian Dames, and four years later than initially committed to by the Eskom board when the project started. Despite earlier assurances about the synchronization, Eskom missed the deadline and a revised date has been given as mid June 2015 (Gori, 2015; Yelland, 2015).

Although Kusile and Medupi projects will generate 9 600MW in combined power to the national grid, there is still a need for more new power stations since after completing these projects Eskom will start decommissioning some old structures that have reached the end of their design life. This will further reduce installed generating capacity, pushing it below the required reserved

margin which is 15% of total capacity (Vecchato & Mantshantsha, 2013). Decommissioning will start by 2021 with 32 028 MW reduction in the national grid by 2040. Majuba, Medupi and Kusile will be the only plants still in operation by 2040 (Cohen, 2013). The aging infrastructure, decline in sales, billions spent in diesel to power the grids and delays to new power stations are all burdening the already strained parastatal and severely constrained grid. The power utility admits there is no light at the end of the tunnel (Gori, 2015). Therefore, electricity supply will continue to be problematic in the future, and hence energy intensive green-field projects are not justifiable. Brownfield projects are favored because there is power allocated to such projects. Even if power was readily available, Eskom prioritizes the sale of electricity to other countries before South Africa e.g. Skorpion Zinc get some of its power from South Africa (Kohler, personal communication 2014; Chiloane, 2012).

#### **4.4 Sensitivity analysis**

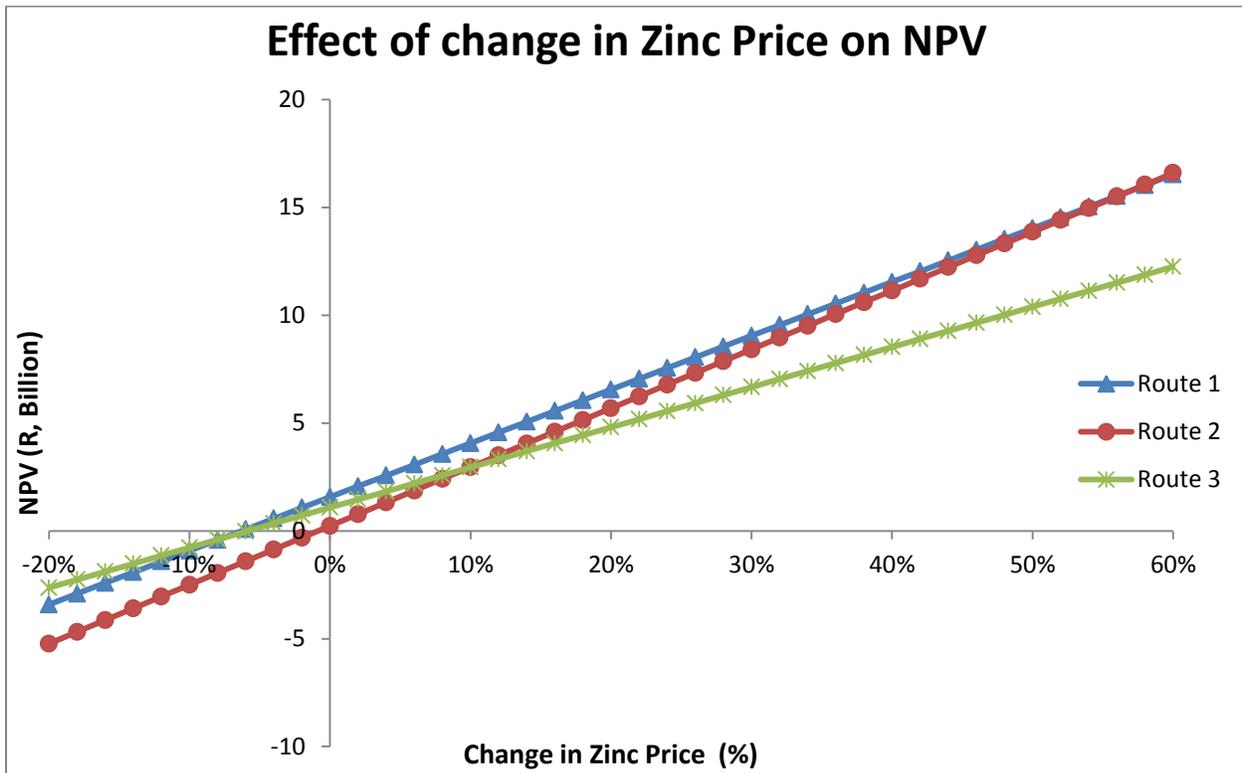
Until now, it has been assumed the financial analysis is deterministic, yet many of the costs and parameters used in evaluating the profitability of a chemical process are estimates that are subject to error because the future is unpredictable. Therefore, decision makers have to study in advance how potential errors will affect NPV. Moreover, it is worthwhile to determine whether key parameters have a significant impact on NPV before a thorough project risk analysis. The baseline estimate has a high probability of not exceeding a certain value and of not falling below a certain value. These maximum and minimum values define ranges in which the baseline estimate may vary, and hence sensitivity analysis is the technique to study the effect of change between these limits. Many factors affect profitability and these are shown in Table 17.

**Table 17: Range of variation of Factors Affecting the profitability of a Chemical Process (Turton et al., 2008)**

<b>Factor in profitability analysis</b>	<b>Probable variation from forecasts (%)</b>
<b>Cost of fixed capital investment</b>	-10 to +25
<b>Construction time</b>	-5 to +50
<b>Start-up costs and time</b>	-10 to +100
<b>Sales volume</b>	-50 to +150
<b>Inflation rates</b>	-10 to +100
<b>Price of product</b>	-50 to +20
<b>Income tax rates</b>	-5 to +15
<b>Plant replacement and maintenance costs</b>	-10 to +100
<b>Interest rates</b>	-50 to + 50
<b>Working capital</b>	-20 to +50
<b>Raw material availability and price</b>	-25 to +50
<b>Salvage value</b>	-100 to +10
<b>Profit</b>	-100 to +10

The sensitivity analysis is presented from Figure 17 to Figure 23 with respect to zinc price, FC, WC, construction time, electricity cost and discount rates. The factors chosen for the sensitivity analysis are based on the likelihood of deviation from the estimates used above. Only those that can have significant changes have been studied. This decision is based on the zinc industry and the location of the plant.

#### 4.4.1 Zinc Price

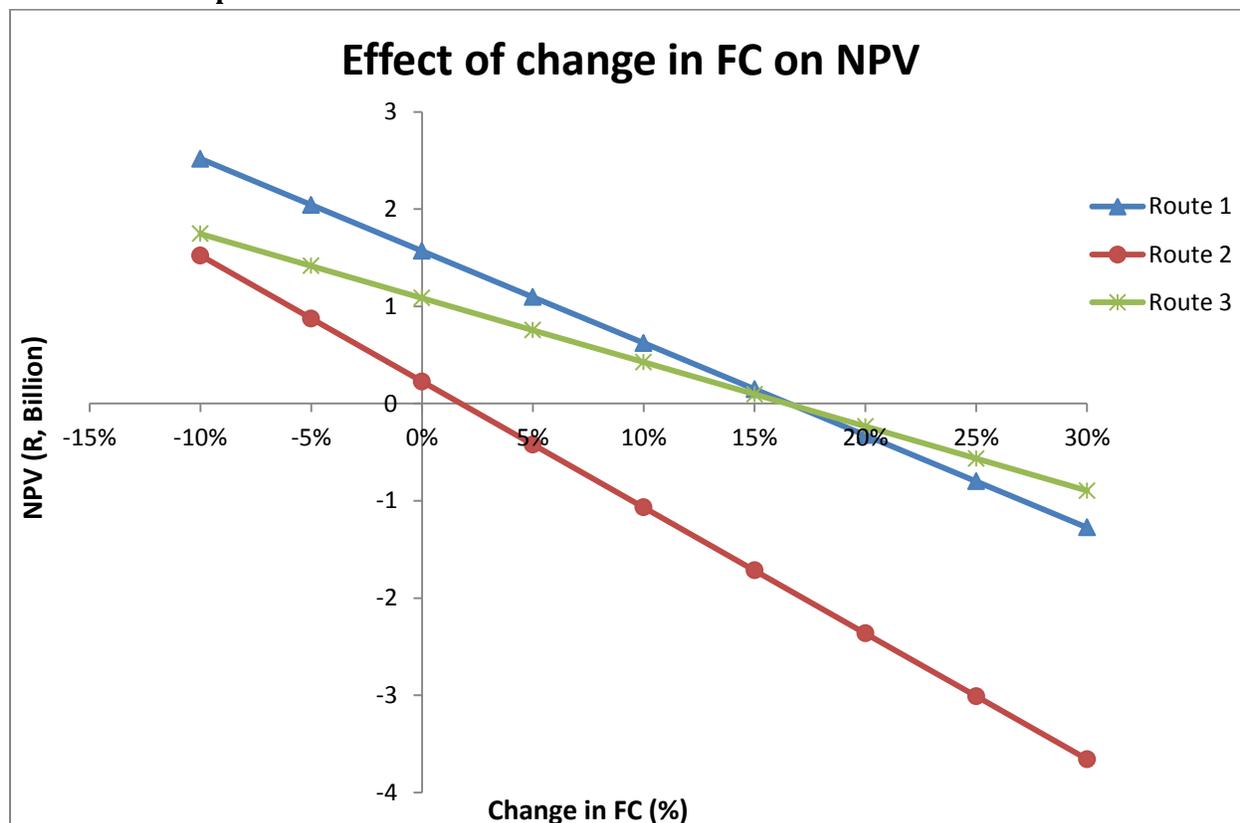


**Figure 17: Sensitivity of NPV to zinc price**

As the Zinc price increases, the difference between the NPVs for route 1 and route 2 decreases. This is due to the lower Zn production rate in route 1, hence the zinc price influences the profitability of each process significantly. Conversely, the difference between route 3 and the other routes increases. This is due to the price participation charge found in zinc concentrate contracts. As the zinc prices increase, the treatment charge is also escalated. On the other hand, from Figure 17 it can be noted that a 7.66 % decrease from the current price makes route 3 preferable, however, the NPVs are negative for all processes at this point. Therefore, route 1 is the most attractive route between a zinc price of US\$2,130/ t (6.30 % decrease) and US\$3,570/ t (56.9 % increase over the current price). Beyond that, route 2 becomes the better choice. It is unlikely for zinc prices to considerably decrease from the current price. Zinc last traded at US\$2,130/ t in June 2014. Ever since July 2014, zinc prices have been trading above this price (IndexMundi, 2014; Westmetall, 2014). CRU consulting predicted that zinc will be trading at US\$4,000/t during 2016/2017 (Zinc Investing News, 2014b). Moreover, zinc supply will be disrupted due to mine closures. Century in Australia, Skorpion in Namibia and the Lisheen mine

in Ireland are closing down in 2015 (Mining, 2014). The decrease in supply coupled with China's robust demand makes it improbable for zinc prices to decrease in the future.

#### 4.4.2 Fixed capital investment

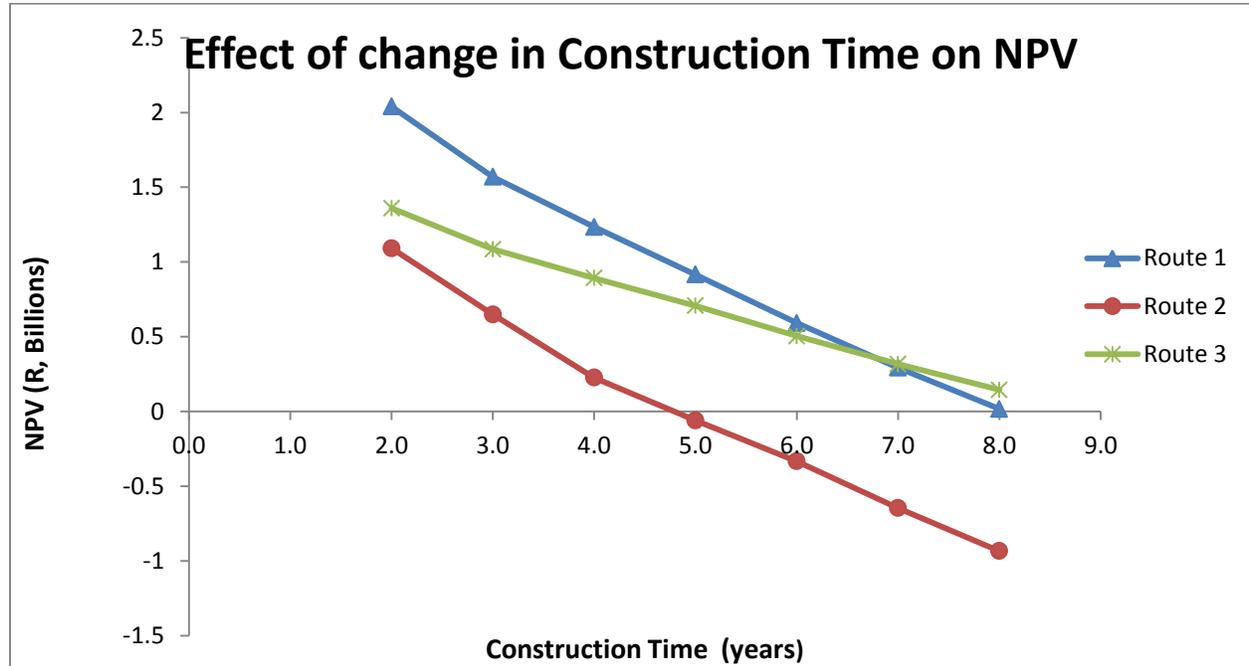


**Figure 18: Sensitivity of NPV to Fixed Capital (FC) Cost**

Fixed costs for feasibility studies are not always on target, these estimates are almost always wrong and hence there is a need to establish a range of likely cost scenarios (Harvard Business School Press, 2002b). From Figure 18, NPV decreases as the FC cost increase, with the NPV of route 2 decreasing faster as compared to the other routes. This change can be noted from the gradient of the line together with the point at which route 2 cuts the x-axis (NPV=0). This significant change is due to the bigger total investment associated with route 2, and hence a slight change in FC has a significant impact on NPV. At 16.8 % increase, the order for the preferred process route change. According to Lang (1989) and Turton and co-authors (2008), the expected range of variation for fixed cost of chemical plants is between -10 and 25 % (see Table 17), therefore, the 16.8 % increase fall within the expected range for changes in FC. As illustrated by Figure 18, a 16.8 % increase in FC makes route 3 preferable, however, the NPVs

are negative for all processes at this point. Therefore route 1 is the most attractive route below 16.6 % increase in FC (NPV of route 1 is zero).

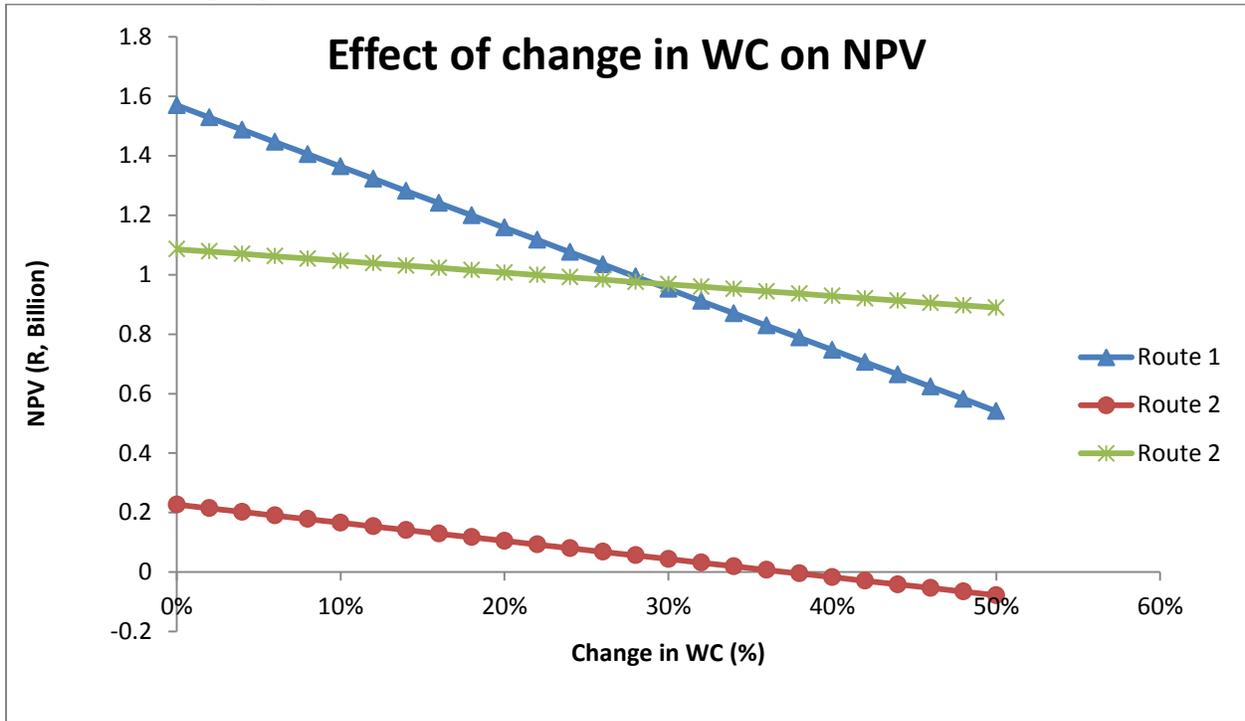
#### 4.4.3 Construction time



**Figure 19: Sensitivity of NPV to construction time**

The problem of delays is a common phenomenon for construction projects worldwide; few projects finish within the contracted deadlines (Challal & Tkiouat, 2012b). The main causes for such delays are: design changes, under-budgeting and unforeseen construction site hazards (Challal & Tkiouat, 2012a). Construction delays have adverse effects on projects since they often result in cost overruns, thus affecting the feasibility and profitability of a project (Afshari et al., 2010). Provided the construction time is the same for all the routes, route 1 is preferable as long as the construction time is less than 6.80 years, thereafter, route 3 is preferable. However, if the construction time differs, the preferable route has a higher NPV corresponding to its construction time. This result can be easily deduced from Figure 19 above.

#### 4.4.4 Working Capital



**Figure 20: Sensitivity of NPV to Working Capital**

Working capital is the amount of capital required to start up the plant and finance the first few months of operation before starting to generate revenues. Working capital for chemical plants is between 10% and 20 % of the fixed capital (Cohen, 2009; Turton et al., 2008). The time taken between mining the ore and receiving revenue for the sold product is known as inventory, or sometimes referred as the pipeline. Changes in inflation, electricity and reagents costs result in changes in working capital, however, WC is primarily related to inventory. Compared to tank leaching, HL has slow extraction rates thus requiring large mined inventories to be under leach before their full extraction value is realized. As a result, HL has high working capital than tank leaching.

Vertical integration results in higher working capital due to the longer pipeline, the sale of concentrates raises revenue much more quickly. Consequently, route 1 and route 2 have higher working capital as compared to route 3. Heaps are operated for a year or more (John, 2011) yet only 10-20 hours is needed for atmospheric leaching (Filippou, 2004). As a result, it takes much longer to ramp up heaps to full production capacity. Therefore, heaps have much higher working capital than tank leaching processes, this can be noted from the gradient of the lines. Route 1 has

a steeper slope, and hence a slight increase in WC considerably affects the profitability of the process. However, the WC for route 1 is more likely to decrease due to heap optimization. Heap optimization will reduce the residence time and hence reduce WC. From Figure 20, route 1 is preferable to the other routes as long as the change in working capital remains less than 29.1 %, thereafter, route 3 is the better route. Even if the residence time for the heap was to increase by three months, the heap will still be the better choice compared to the other routes.

#### 4.4.5 Electricity Price

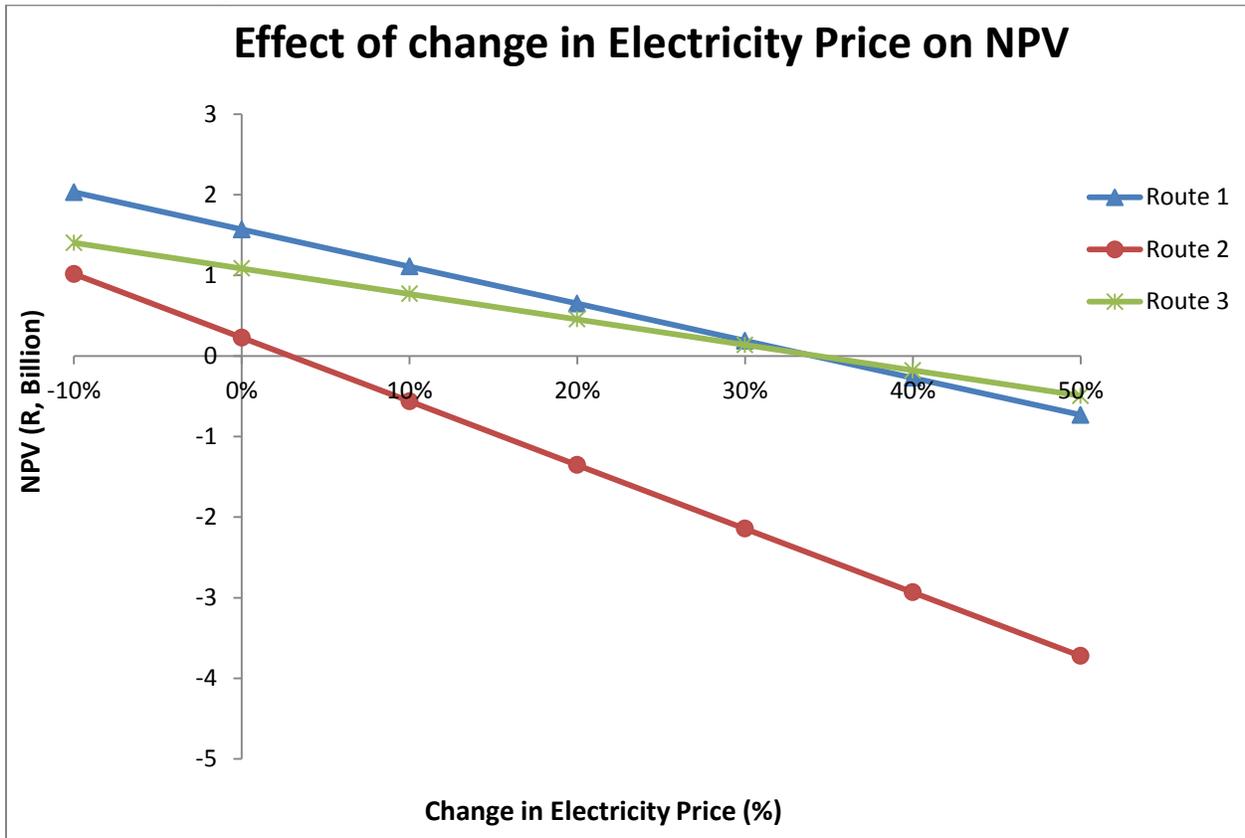


Figure 21: Sensitivity of NPV and IRR to the Electricity Price

For route 1 and route 3, an increase in electricity cost leads to a decrease in the difference between NPVs. This is due to the different process stages associated with each route. Although both routes have energy intensive stages, i.e. electrowinning and milling, the electrowinning stage is more energy intensive than milling. This can be noted from the line gradients of the two routes. Route 1 has a steeper slope, and hence a slight increase in electricity price considerably affects the profitability of the process. Route 2 is the most energy intensive as compared to the

other routes, it has both the milling and electrowinning stages. Therefore, an increase in electricity cost results in an increase in the difference between NPVs of the other routes.

Previous tariff history coupled with the forecasted insecurity of electricity supply makes it improbable for the electricity price to decrease in the future. Therefore, electricity prices are expected to increase rather than decrease (Eskom, 2013; Ramokgopa, 2007; Ramokgopa, 2001). From Figure 21, it can be noted that route 1 is the more preferable for electricity prices below R 1.15 per kWh, 33.5 % increase over the current price. Route 3 is only preferable between 33.5 % increase and 34.3 %, thereafter, NPV for route 3 is negative.

#### 4.4.6 Discount rates

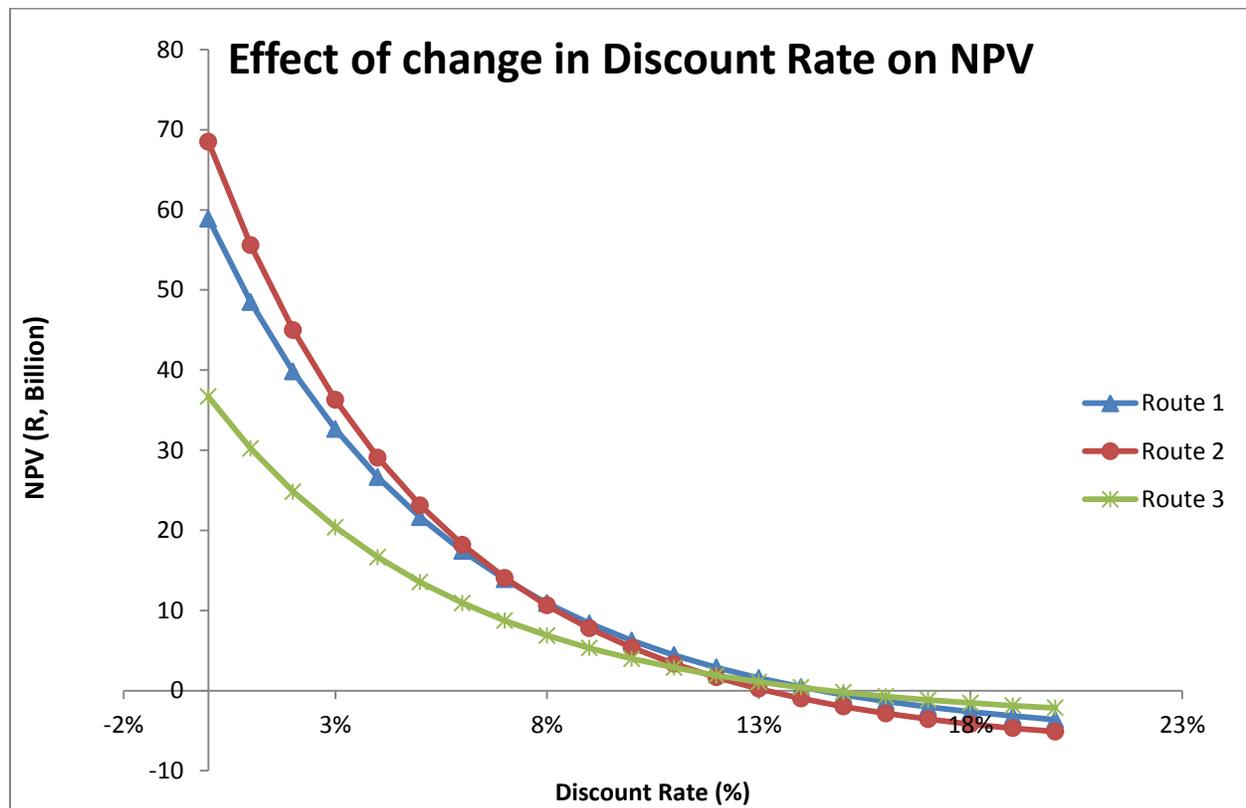
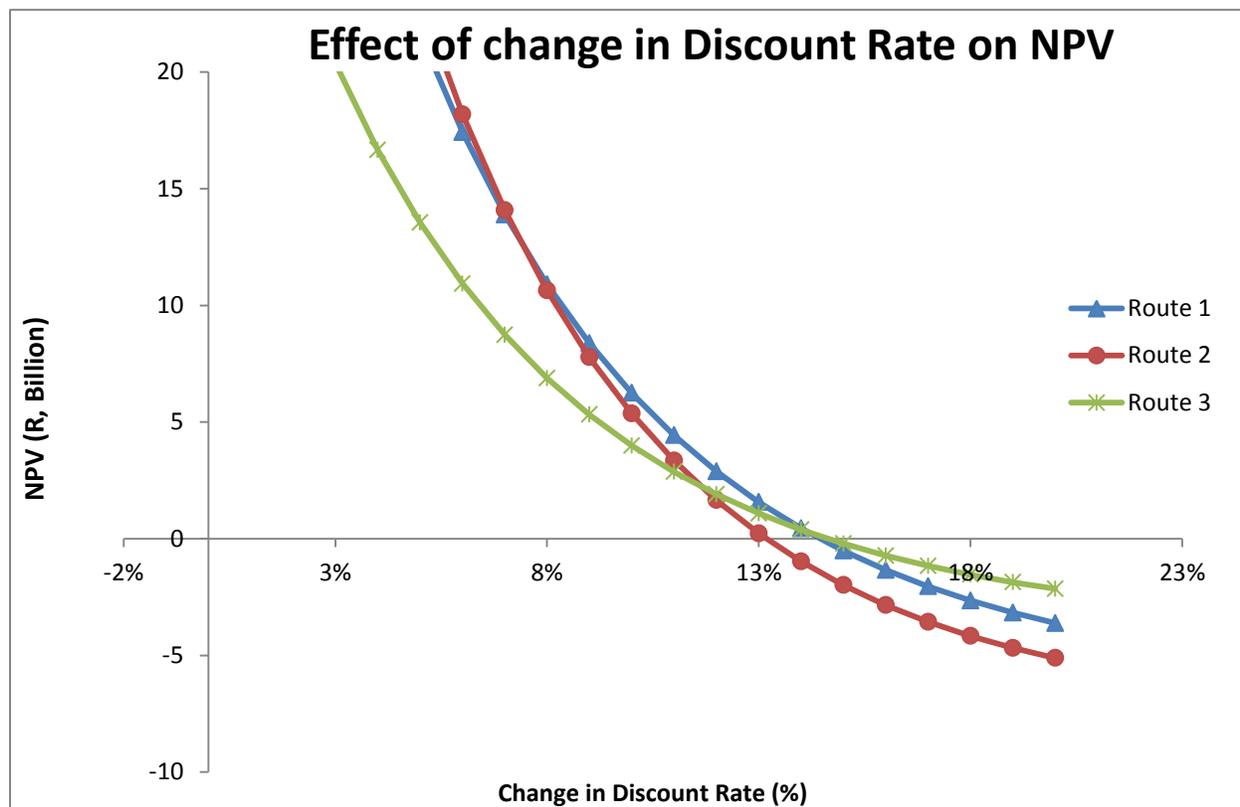


Figure 22: Sensitivity of NPV to discount rate



**Figure 23: Zoomed in Sensitivity of NPV to discount rate in order to display crossover values clearly**

From Figure 22 and Figure 23, NPV decreases as the discount rate increases which implies that the future cash flows decrease. However, the NPV of route 2 decreases faster as compared to route 1 and route 3. This is due to the bigger investment associated with route 2, and hence it takes a longer period for this route to make a profit. R 15.0 Billion is required in order to invest in route 2 as compared to R 12.4 Billion and R 7.55 Billion required for route 1 and route 3. Cash flows received at later stages are more sensitive to changes in discount rates (Johnson, 1999b). Route 3 cash flows are greater in the early years of the project and are less sensitive to change in discount rates as compared to route 1 and route 2.

The discount rate at which NPVs of different projects are equal is known as the crossover rate, and there are three crossover rates for these projects. The first crossover rate is for route 1 and route 2, it is 7.39%. The subsequent crossover rate is for route 2 and route 3, it is 11.6% and the last crossover rate is for route 1 and 3, it is 14.2%. Therefore, route 2 is the more preferable route at all discount rates less than 7.39% and route 1 is preferable for rates between 7.39% and 14.2%.

Route 3 is preferable for discount rates in excess of 14.2 % and below 14.6 % (IRR of route 3). In African countries, green field projects usually have discount rates between 10-13% (Ledgerwood, personal communication 2014). A discount rate of 13% has been used for the analysis. Therefore, route 1 is better choice for this project.

#### **4.5 Gamsberg-Skorpion Integrated Zinc Project**

As of 13 November 2014, Vedanta has approved the development of Gamsberg-Skorpion Integrated Zinc Project in order to refine the Gamsberg ore body (Mining Weekly, 2014; Mining-Technology, 2014; Miningmx, 2014). Skorpion Zinc, the largest zinc refinery complex in Africa, is expected to close down by 2015 and a plan to invest US\$782 m (R8, 090 m) over a three year period in a Gamsberg-Skorpion integrated project has been approved (Mining Weekly, 2014; Mining-Technology, 2014). According to various sources, approximately US\$630 m (R6, 520 m) will be used to develop a 250 kt/y open pit zinc mine, concentrator plant and associated infrastructure at Gamsberg with the remainder being used to upgrade the Skorpion Zinc plant to refine concentrates from Gamsberg into special high grade zinc metal. The upgrade includes installing a roaster, an acid plant and calcine leach section to process the sulphide concentrates.

This integration is a brownfields RLE process. A study done by Fuls in 2011 recommended direct zinc sulphide concentrate leaching methods when upgrading the Skorpion Zinc refinery to treat Gamsberg concentrates. It could be a better option to directly leach the zinc concentrates, treating them using the atmospheric leach process rather than installing a roaster which has high operating and capital costs. Moreover, some of the units which have to be installed can be eliminated e.g. acid plant, thus further lowering the capex and opex for upgrading the Skorpion Zinc plant. However, the RLE technology is well established and hence the process risk is minimal.

Considering the availability of zinc sulphide concentrates, together with the existing smelting and refining infrastructure, there is an opportunity to sustain refined zinc production at Skorpion. The proposed integration offers strategic advantages since it will sustain employment at Skorpion and treat high manganese Gamsberg concentrate. Moreover, Namibia has a reliable infrastructure and a stable political and economic environment. According to Mr Albanese, Vedanta's CEO, the output from Gamsberg coupled with the extended life of the refinery at the

Skorpion site will make the Southern African region one of the most important suppliers of refined zinc globally (Zinc International, 2014).

This proposal will offset the fall in production volume resulting from the end of life of the Lisheen mine in Ireland which is part of Vendata stable of mines. The first phase of Gamsberg mine is expected to have a total lifespan of approximately 13 years with the first ore likely to be produced between 2017 and 2018. There is a potential for much longer mine life. The construction of the new mine is expected to commence during the first quarter of 2015 (Sesa Sterlite, 2014). Various sources have stated that the Gamsberg project will be operated by Black Mountain Mining (BMM) and it will create 1500 jobs during construction and 500 permanent jobs (Reuters, 2014). BMM has been in operation for over three decades and mine life is expected to end in 2021. Approximately 1.6 Mt ore are hoisted at Black Mountain per annum, Gamsberg has the potential to triple BMM's output (Mining-Technology, 2014; Zinc International, 2014).

## CHAPTER 5

### 5 Conclusions and Recommendations

#### 5.1 Conclusions

The objective of the study was to compare three process routes in the context of processing ore from the Gamsberg deposit for refining 3.4 million tpa ore in order to produce special high grade (SHG) zinc (>99.995% Zn). These routes include heap leaching and refining locally (route 1), preparing a flotation concentrate and refining it locally (route 2) and lastly, preparing a flotation concentrate and shipping it for toll refining in Europe (route 3). The amount of mined ore was calculated based on producing a minimum of 150 kt/a SHG zinc for route 1, with the lowest possible zinc grade at 6.27 wt%. The zinc production rate is the same as name-plate production at Skorpion Zinc.

A desk-top model was developed using Microsoft Excel to compare the economic viability of the three routes. Sensitivity analysis of the net present value (NPV) to key variables was simulated using Matlab. The base case assumptions for the model are: zinc price of US\$ 2,330/t, zinc extraction of 72.6% and 98% for heap and tank leaching, respectively, 96.5% zinc recovery from the pregnant leach solution, 86% metal recovery during flotation, an average zinc grade of 7.4 wt% and a 20 year plant life. The findings of these studies are listed below:

- ❖ Route 1 has a zinc production of 176 kt/y, route 2 yields 204 kt/y and route 3, 216 kt/y.
- ❖ For both capital and operating costs, route 2 has the highest costs followed by route 1 and then route 3. The milling and electrowinning stages are the major contributors to capital and operating costs.
- ❖ The total capital investment is 12.4, 15.0 and 7.55 billion Rand for routes 1, 2 and 3, respectively. The corresponding NPVs are 1.57, 0.23 and 1.09 billion Rand under base case conditions.
- ❖ Route 2 has the longest pay-back period (3.33 years) followed by route 3 (3.14 years) whereas route 1 has the shortest with 2.76 years.
- ❖ The internal rate of return is 14.4 % for route 1, 13.2 % for route 2 and 14.6 % for route 3.

- ❖ With decreasing order, the sensitivity of NPV to key variables is as follows: discount rate, fixed capital, working capital, electricity price, zinc price and the least sensitive variable is construction time.

In conclusion, route 1 is the most attractive route for Gamsberg. Although this route is risky as compared to the other routes, it produces zinc which is highly competitive to many mine-to-metal processes. The economics show that route 3 is the better option for processing ore from the Gamsberg deposit when an investor is not willing to take the risk associated with route 1. There is an energy crisis in South Africa and operating a refinery is energy intensive. Provided there is enough energy to supply a green-field project (i.e. one with no previous energy allocation) and there is no capital rationing, route 1 is the attractive option for Gamsberg. If not, route 3 becomes the better choice. Although route 1 is preferable, heap leaching still remains as a novel technology in the zinc industry and hence it is the most risky route. Route 2 and 3 are well established, and hence less risky.

## **5.2 Recommendations**

This study focused on the economic viability of the three process routes. It is recommended that the environmental and social implications associated with each route should be considered before making a decision on the processing route for Gamsberg. Moreover, the higher risk associated with route 1 compared to the other routes should be carefully evaluated. The only source of revenue considered for the model is from the sale of zinc metal or zinc concentrate, therefore, it would be worthwhile to account for all revenue streams associated with each route e.g. the copper cake produced in route 1 and 2 could be leached to recover zinc and upgrade the copper content. Likewise, all expenses should be accounted for as the model did not include general expenses e.g. administration. An economic evaluation would have to be made for the proposed route in order to compare it against the other three routes studied.

For route 1, pilot test work will have to be conducted to obtain accurate design parameters for the heap, since zinc heap leaching remains a novel technology. Moreover, it is worthwhile to establish the effect of particle size distribution on the recovery. Metal recovery during flotation changes with grade. The effect of change in grade on the metal recovery should be studied and the sensitivity of NPV to the metal recovery should be analysed for route 2 and 3.

As a final commentary, this study has built up on the work done by Fuls by evaluating the HydroZinc™ process (route 1) which was discarded earlier in his analysis. Therefore, provided there is no capital rationing and a proposed project will vertically integrate; it is recommended that the HydroZinc™ process be considered amongst possible processing options.

## CHAPTER 6

### 6 References

- Afshari, H., Khosravi, S., Ghorbanali, A., Borzabadi, M. & Valipour, M. 2010. Identification of causes of non-excusable delays of construction projects. International Conference on E-Business Management and Economics. 42.
- Anderson, C.G., Giralico, M.A., Posf, T.A., Robinson, T.G. & Tinkle, O.S. 2002. Selection and sizing of copper solvent extraction and electrowinning equipment and circuits. Mineral Processing Plant Design, Practice, and Control Proceedings.SME. 21709-1743.
- Babu, M.N., Sahu, K.K. & Pandey, B.D. 2002. Zinc recovery from sphalerite concentrate by direct oxidative leaching with ammonium, sodium and potassium persulphates. Hydrometallurgy. 64(2): 119-129.
- Balesini, A., Razavizadeh, H. & Zakeri, A. 2011. Solvent Extraction of Zinc from Acidic Solution Obtained from Cold Purification Filter Cake of Angouran Mine Concentrate Using D2EHPA. Iranian Journal of Chemical Engineering. 8(3):
- Barbir, K. 2012. Zinc Market Outlook. Available:  
[https://www.google.co.za/url?sa=t&rct=j&q=&esrc=s&source=web&cd=4&ved=0CD0QFjAD&url=http%3A%2F%2Fwww.nyrstar.com%2F\\_layouts%2Fdownload.aspx%3FSourceUrl%3D%2Finvestors%2Fen%2FNyr\\_Documents%2FEnglish%2Fantaik%2BLead%2B%2BZinc%2BConference%2BNovember%2B2012.pdf&ei=NW5fU\\_fGB-md0QXL54GgBg&usg=AFQjCNGCMLZYdGZIy8yxUig\\_vX\\_pB0YD0w&bvm=bv.65397613,d.bGE&cad=rja](https://www.google.co.za/url?sa=t&rct=j&q=&esrc=s&source=web&cd=4&ved=0CD0QFjAD&url=http%3A%2F%2Fwww.nyrstar.com%2F_layouts%2Fdownload.aspx%3FSourceUrl%3D%2Finvestors%2Fen%2FNyr_Documents%2FEnglish%2Fantaik%2BLead%2B%2BZinc%2BConference%2BNovember%2B2012.pdf&ei=NW5fU_fGB-md0QXL54GgBg&usg=AFQjCNGCMLZYdGZIy8yxUig_vX_pB0YD0w&bvm=bv.65397613,d.bGE&cad=rja) [29 April 2014].
- BCS, I. 2002a. Energy and Environmental Profile of the U.S. Mining Industry - Lead and Zinc. Available: <http://infohouse.p2ric.org/ref/46/45410.pdf> [10 April 2014].
- BCS, I. 2002b. Energy and Environmental Profile of the U.S. Mining Industry -Lead and Zinc.
- Berezowsky, T. 2011. Commodity Review and Second-Half Outlook – Part Two. Available: <https://agmetalmminer.com/2011/07/commodity-review-and-second-half-outlook-part-two/> [16 April 2014].
- Bosecker, K. 1997. Bioleaching: metal solubilization by microorganisms. FEMS Microbiology Reviews. 20(3-4): 591-604.

- Breitenbach, A. & Thiel, R. 2005. A tale of two conditions: heap leach pad versus landfill liner strengths. Proc.GRI. 19.
- Breitenbach, A.J. 2005. Heap leach pad design and construction practices in the 21st Century. Vector Colorado LLC.
- Brown, B. 2002. Management of Tailings Disposal on Land. Proceedings of the Mineral Processing Plant Design, Practice, and Control. A.L Mular, D.N Halbe & D. J. Baratt, Eds. Society for Mining, Metallurgy, and Exploration (SME). 1890-1827.
- Buban, K., Collins, M. & Masters, I. 1999. Iron control in zinc pressure leach processes. JOM. 51(12): 23-25.
- Burgess, L. 2006. The Most Undervalued Metal on the Market Today. WealthDaily.
- Challal, A. & Tkiouat, M. 2012a. Identification of the causes of deadline slippage in construction projects: State of the art and application. Journal of Service Science and Management. 5151.
- Challal, A. & Tkiouat, M. 2012b. Qualitative approach risk period in construction projects. Journal of Financial Risk Management. 1(03): 42.
- Chang, C., Gu, H. & O'Keefe, T.J. n.d. Review of the galvanic stripping process for use in treating oxidized metal wastes. Available: <http://www.engg.ksu.edu/HSRC/96Proceed/chang.pdf> [9 September 2013].
- Chen, T.T. & Cabri, L.J. 1986. Mineralogical overview of iron control in hydrometallurgical processing. In Iron control in hydrometallurgy. J.E. Dutrizac & A.J. Monhemius, Eds. Chichester, West Sussex: Ellis Horwood Limited. 19-46.
- Chiloane, L.D. 2012. Solar Energy in the Minerals Processing Industry: Identifying the first opportunities
- Claassen, J.O. 2002. Characterisation and Optimisation of the Zincor Iron Removal Process.
- Claassen, J.O. & Sandenbergh, R. 2007. Influence of temperature and pH on the quality of metastable iron phases produced in zinc-rich solutions. Hydrometallurgy. 86(3): 178-190.
- Claassen, J.O., Meyer, E.H.O., Rennie, J. & Sandenbergh, R.F. 2002. Iron precipitation from zinc-rich solutions: defining the Zincor Process. Hydrometallurgy. 67(1-3): 87-108.
- Cohen, B. 2013. The future of coal: existing power stations considerations for new stations. Available:

<http://stias.ac.za/wp-content/uploads/2013/03/WS-10-Cohen-Future-of-Coal.pdf> [11 December 2014]

- Cohen, B. 2009. Engineering Economics Analysis. Department of Chemical Engineering, University of Cape Town:
- Cole, P. & Sole, K. 2002a. Solvent extraction in the primary and secondary processing of zinc. *Journal of the South African Institute of Mining and Metallurgy*. 102(8): 451-456.
- Cole, P.M. & Sole, K.C. 2002b. Solvent extraction in the primary and secondary processing of zinc. *Proceedings of the International Solvent Extraction Conference*. 17 to 21 March. K.C. Sole, P.M. Cole, J.S. Preston & D.J. Robinson, Eds. Chris van Rensburg Publications (Pty) Ltd. 863.
- Cole, P.M. & Sole, K.C. 2003. Zinc solvent extraction in the process industries. *Mineral Processing and Extractive Metallurgy Review*. 24(2): 91-137.
- Cornell, R.M. & Schwertmann, U. 1996. *The iron oxides: structure, properties, reactions, occurrence and uses*. New York: Weinheim: VCH.
- CPRA 2007. *Zinc Industry Report*. Export Credit Insurance Corporation of South Africa Limited.
- Daoud, J. & Karamanev, D. 2006. Formation of jarosite during  $\text{Fe}^{2+}$  oxidation by *Acidithiobacillus ferrooxidans*. *Minerals Engineering*. 19(9): 960-967.
- Das, A., Modak, J.M. & Natarajan, K.A. 1997. Studies on multi-metal ion tolerance of *Thiobacillus ferrooxidans*. *Minerals Engineering*. 10(7): 743-749.
- Davenport, W.G., King, M., Schwarz, M. & Biswas, A.K. 2002a. Concentrating Copper Ores. In *Extractive Metallurgy of Copper*. 4th ed. The Boulevard, Langford Lane Kidlington, Oxford OX5 1GB, UK: Elsevier Science Ltd. 31-54.
- Davenport, W.G., King, M., Schwarz, M., Biswas, A.K. & Salomon-de-Friedberg, H. 2002b. Hydrometallurgical Copper Extraction: Introduction and Leaching. In *Extractive Metallurgy of Copper*. 4th ed. The Boulevard, Langford Lane Kidlington, Oxford OX5 1GB, UK: Elsevier Science Ltd. 289-303.
- De Leeuw, M. 2011. China continues to drive demand for zinc. Available: [http://www.nyrstar.com/investors/en/Nyr\\_Documents/English/DB%20Presentation%20Nyrstar%20BRICS%20November%202011%20FINAL.pdf](http://www.nyrstar.com/investors/en/Nyr_Documents/English/DB%20Presentation%20Nyrstar%20BRICS%20November%202011%20FINAL.pdf) [29 April 2014].

- De Souza, A.D., Pina, P.S. & Leão, V.A. 2007. Bioleaching and chemical leaching as an integrated process in the zinc industry. *Minerals Engineering*. 20(6): 591-599.
- Du Plessis, C.A., Batty, J.D. & Dew, D.W. 2006. Commercial Applications of Thermophile Bioleaching. *Biomining*. D.E. Rawlings & B.D. Johnson, Eds. London: Springer.57-80
- Dutrizac, J.E. 1996. The effect of seeding on the rate of precipitation of ammonium jarosite and sodium jarosite. *Hydrometallurgy*. 42(3): 293-312.
- Ekaterina, V. 2012. Russian Metal Industry In The Scope Of The International Pricing Rules: An Example Of Zinc Concentrate Market. Available: [http://lare-efi.u-bordeaux4.fr/IMG/docx/Vostretsova\\_Russian\\_metal\\_industry\\_April\\_2012\\_Conf\\_on\\_Economic\\_Development.docx](http://lare-efi.u-bordeaux4.fr/IMG/docx/Vostretsova_Russian_metal_industry_April_2012_Conf_on_Economic_Development.docx). [09 May 2014].
- Environmental Resources Management n.d.a. Environmental and Social Impact Assessment [ESIA] for Proposed Construction of the Gamsberg Zinc Mine and Associated Infrastructure, Northern Cape Province. Available: [www.sahra.org.za/sites/default/files/additionaldocs/BLACK%20MOUNTAIN%20BID%20Eng-email%20version.pdf](http://www.sahra.org.za/sites/default/files/additionaldocs/BLACK%20MOUNTAIN%20BID%20Eng-email%20version.pdf) [15 April 2014].
- Environmental Resources Management n.d.b. Scoping Report for Gamsberg Zinc Mine. Available: [www.erm.com/PageFiles/6669/Chapter%203%20A.pdf](http://www.erm.com/PageFiles/6669/Chapter%203%20A.pdf) [19 June 2013].
- Eskom 2013. Tariffs & Charges Booklet - 2013/2014.
- Eskom 2014. Medupi Power Station Project. Available: [http://www.eskom.co.za/Whatweredoing/NewBuild/MedupiPowerStation/Pages/Medupi\\_Power\\_Station\\_Project.aspx](http://www.eskom.co.za/Whatweredoing/NewBuild/MedupiPowerStation/Pages/Medupi_Power_Station_Project.aspx) [24 October 2014].
- Fernandes, A.M. & Paunov, C. 2012. The risks of innovation: Are innovating firms less likely to die?
- Filippou, D. 2004. Innovative Hydrometallurgical Processes for the Primary Processing of Zinc. *Mineral Processing and Extractive Metallurgy Review*. 25(3): 205-252.
- Flett, D.S. 2005. Solvent extraction in hydrometallurgy: the role of organophosphorus extractants. *Journal of Organometallic Chemistry*. 690(10): 2426-2438.
- Fuls, H.F. 2011. Evaluation of Processing Options for the Treatment of Zinc Sulphide Concentrates at Skorpion Zinc.

- Ghorbani, Y., Becker, M., Mainza, A., Franzidis, J. & Petersen, J. 2011. Large particle effects in chemical/biochemical heap leach processes – A review. *Minerals Engineering*. 24(11): 1172-1184.
- Ghorbani, Y., Petersen, J., Harrison, S.T.L., Tupikina, O.V., Becker, M., Mainza, A.N. & Franzidis, J. 2012. An experimental study of the long-term bioleaching of large sphalerite ore particles in a circulating fluid fixed-bed reactor. *Hydrometallurgy*. 129–130(0): 161-171.
- Gori, Hasina. 2015. Eskom needs billions to keep lights on. Available: <http://www.sabc.co.za/news/a/9209ef0046e2dbf48b77ab618fc35e0b/Eskom-needs-billions-to-keep-lights-on-20151001> [12 January 2015]
- Guise, L. & Castro, F. 1998. Purification of sulphuric acid solutions from the leaching of nickelliferous of laterites.
- Gupta, C.K. & Mukherjee, T.K. 1990a. *Hydrometallurgy in Extraction Processes, Volume 1*. CRC Press.
- Gupta, C.K. & Mukherjee, T.K. 1990b. *Hydrometallurgy in Extraction Processes, Volume II*. CRC Press.
- Harlamovs, Juris R., Ashman, Daniel W., Dominguez, Jose Alberto Gonzalez., Lizama, Hector M., Makwana, Dilipkumar D. & Stradling, Andrew W. 2004. Heap bioleaching process for the extraction of zinc.
- Harvard Business School Press 2002a. *Practical Tools for Management Decisions*. In *Finance for Managers*. Harvard Business Review Press. 138-148.
- Harvard Business School Press 2002b. *The Time Value of Money*. In *Finance for Managers*. Harvard Business Review Press. 150-170.
- Harvey, T.J., Van Der Merwe. W. & Afewu, K. 2002. The application of the GeoBiotics GEOCOAT<sup>®</sup> biooxidation technology for the treatment of sphalerite at Kumba resources' Rosh Pinah mine. *Minerals Engineering*. 15: 823-829.
- IndexMundi 2014. Zinc Monthly Price - US Dollars per Metric Ton. Available: <http://www.indexmundi.com/commodities/?commodity=zinc&months=300> [28 October 2014].
- Inflation.eu 2014. Inflation South Africa 2014. Available: <http://www.inflation.eu/inflation-rates/south-africa/historic-inflation/cpi-inflation-south-africa-2014.aspx> [26 November 2014].

- International Zinc Association 2011a. Zinc - Natural Occurrence. Available: [http://www.zinc.org/basics/zinc\\_natural\\_occurrence](http://www.zinc.org/basics/zinc_natural_occurrence) [08 March 2014].
- International Zinc Association 2011b. Zinc Production - From Ore to Metal. Available: [http://www.zinc.org/basics/zinc\\_production](http://www.zinc.org/basics/zinc_production) [08 March 2014].
- International Zinc Association 2011c. Zinc Uses. Available: [http://www.zinc.org/basics/zinc\\_uses](http://www.zinc.org/basics/zinc_uses) [08 March 2014].
- InvestmentMine 2014. 1 Year Zinc Prices and Price Charts. Available: <http://www.infomine.com/investment/metal-prices/zinc/1-year/> [09 September 2014].
- Ismael, M. & Carvalho, J. 2003. Iron recovery from sulphate leach liquors in zinc hydrometallurgy. *Minerals Engineering*. 16(1): 31-39.
- Ivanov, I. 2004. Increased current efficiency of zinc electrowinning in the presence of metal impurities by addition of organic inhibitors. *Hydrometallurgy*. 72(1): 73-78.
- Jansen, M. & Taylor, A. 2002. A new approach to heap leach modeling and scale-up. Proceedings of the ALTA Copper-7 Conference. [Http://www. Altamet. Com. au/Technical% 20Papers% 20and% 20Articles/ALTA% 20Copper/New% 20Approach% 20to% 20Heap% 20Leach% 20Modelling% 20and% 20Scale-Up. Pdf.](http://www.altamet.com.au/Technical%20Papers%20and%20Articles/ALTA%20Copper/New%20Approach%20to%20Heap%20Leach%20Modelling%20and%20Scale-Up.Pdf)
- Jarvenin, G. 2008. Precipitation and crystallization processes.
- John, L. 2011. The Art of Heap Leaching - The Fundamentals. In *Percolation Leaching: The status globally and in Southern Africa*. Misty Hills: The Southern African Institute of Mining and Metallurgy (SAIMM). 17-42.
- Johnson, H.J. 1999a. Capital Budgeting Techniques. In *Making Capital Budgeting Decisions: Maximizing the Value of the Firm* Corporate finance manual. Prentice Hall: Financial Times. 37-57.
- Johnson, H.J. 1999b. Special Situations. In *Making Capital Budgeting Decisions: Maximizing the Value of the Firm* Corporate finance manual. Prentice Hall: Financial Times. 93-114.
- Jurimae, T. 2010. Risk management in the procurement of innovation. (EUR 24229 EN). Luxembourg: European Union.
- Kappes, D.W. 2002. Precious metal heap leach design and practice. Proceedings of the Mineral Processing Plant Design, Practice, and Control. Proceedings of the Mineral

Processing Plant Design, Practice, and Control. A.L Mular, D.N Halbe & D. J. Baratt, Eds. Society for Mining, Metallurgy, and Exploration (SME). 1606-1630.

- Kawatra, S.K. n.d. Froth Flotation – Fundamental Principles. Available: [http://www.chem.mtu.edu/chem\\_eng/faculty/kawatra/Flotation\\_Fundamentals.pdf](http://www.chem.mtu.edu/chem_eng/faculty/kawatra/Flotation_Fundamentals.pdf) [06 June 2014].
- King, M.J., Sole, K.C. & Davenport, W.G. 2011. Extractive metallurgy of copper. Elsevier.
- Klauber, C. 2008. A critical review of the surface chemistry of acidic ferric sulphate dissolution of chalcopyrite with regards to hindered dissolution. *International Journal of Mineral Processing*. 86(1–4): 1-17.
- Kohad, V. 1998. Flotation of Sulphide Ores-HZL Experience.
- Lahtinen, M., Svens, K., Haakana, T. & Lehtinen, L. 2008. Zinc plant expansion by outotec direct leaching process. 47th Annual Conference of Metallurgists of CIM, Winnipeg, Manitoba Canada, Zinc and Lead Metallurgy. 167.
- Lang, H. 1989. Cost Analysis for Capital Investment Decisions. CRC Press.
- Lizama, H.M., Harlamovs, J.R., Belanger, S. & Brienne, S.H. 2003. The Teck Cominco Hydrozinc™ process. In *Hydrometallurgy 2003 - Firth International Conference in Honour of Professor Ian M. Ritchie - Volume 2: Electrometallurgy and Environmental Hydrometallurgy*. C.A. Young, and others, Eds. volume 2 ed. TMS (The Minerals, Metals & Material Society). 1503 -1516.
- Lizama, H.M., Harlamovs, J.R., McKay, D.J. & Dai, Z. 2005. Heap leaching kinetics are proportional to the irrigation rate divided by heap height. *Minerals Engineering*. 18(6): 623-630.
- Loan, M., Newman, O.M.G., Cooper, R.M.G., Farrow, J.B. & Parkinson, G.M. 2006. Defining the Paragoethite process for iron removal in zinc hydrometallurgy. *Hydrometallurgy*. 81(2): 104-129.
- Long, H., Chai, L., Qin, W. & Tang, S. 2010. Solvent extraction of zinc from zinc sulfate solution. *Journal of Central South University of Technology*. 17760-764.
- Luoma, T., Paasi, J. & Nordlund, H. 2008. Managing commercialisation risks in innovation development: Linking front end and commercialisation. *Proceedings of the XIX ISPIM Annual Conference. Open Innovation: Creating Products and Services through Collaboration*. Tours, France.

- Lupi, C. & Pilone, D. 2000. Reductive stripping in vacuum of Fe(III) from D2EHPA. *Hydrometallurgy*. 57(3): 201-207.
- Lupo, J. & Morrison, K. 2007. Geosynthetic design and construction approaches in the mining industry. *Geotextiles and Geomembranes*. 25(2): 96-108.
- Lupo, J.F. 2005. Heap Leach facility liner design. Proceedings of the North American Geosynthetics Society (NAGS)/GRI19 Conference, Las Vegas, Nev. 14.
- Martín, D., Díaz, G., García, M. & Sánchez, F. 2002. Extending zinc production possibilities through solvent extraction. *Journal of the South African Institute of Mining and Metallurgy*. 102(8): 463-467.
- McClung, C.R. & Viljoen, F. 2011. A detailed mineralogical assessment of sphalerites from the Gamsberg zinc deposit, South Africa: The manganese conundrum. *Minerals Engineering*. 24(8): 930-938.
- Mining 2014. Doubts creeping in over stellar zinc price run. Available: [www.mining.com/doubts-creeping-in-over-stellar-zinc-price-run-44070](http://www.mining.com/doubts-creeping-in-over-stellar-zinc-price-run-44070) [05 September 2014]
- Mining Weekly 2014. Vedanta approves \$782 Gamsberg-Skorpion zinc project development. Available: [http://www.engineeringnews.co.za/article/vedanta-to-invest-782m-to-develop-gamsbergskorpion-zinc-project-2014-11-13/article\\_comments:1](http://www.engineeringnews.co.za/article/vedanta-to-invest-782m-to-develop-gamsbergskorpion-zinc-project-2014-11-13/article_comments:1) [11 December 2014].
- Miningmx 2014. Vedanta has cracked Gamsberg: Kumar.
- Mining-Technology 2014. Sesa Sterlite to invest \$78 in Gamsberg-Skorpion Intergrated Zinc Project. Available: <http://www.mining-technology.com/news/newssesa-sterlite-to-invest-782m-in-gamsberg-skorpion-integrated-zinc-project-4438682> [11 December 2014].
- Moore, J.M., Watkeys, M.K. & Reid, D.L. 1990. The regional setting of the Aggeneys/Gamsberg base metal deposits Namaqualand, South Africa. *Regional Metamorphism of Ore Deposits and Genetic Implications*. 9-19 July 1989. P.G. Spry & T.L. Bryndzia, Eds. Utrecht, Netherlands: VSP BV. 77-95.
- Moors, E.H. & Dijkema, G.P. 2006. Embedded industrial production systems: lessons from waste management in zinc production. *Technological Forecasting and Social Change*. 73(3): 250-265.

- Moors, E.H.M., Mulder, K.F. & Vergragt, P.J. 2005. Towards cleaner production: barriers and strategies in the base metals producing industry. *Journal of Cleaner Production*. 13(7): 657-668.
- Mular, A.L. 2002. Major Mineral Processing Equipment Costs and Preliminary Capital Cost Estimations. *Proceedings of the Mineral Processing Plant Design, Practice, and Control*. A.L. Mular, D.N Halbe & D. J. Baratt, Eds. Society for Mining, Metallurgy, and Exploration (SME). 310-326.
- Muñoz, J.A., González, F., Blázquez, M.L. & Ballester, A. 1995. A study of the bioleaching of a Spanish uranium ore. Part I: A review of the bacterial leaching in the treatment of uranium ores. *Hydrometallurgy*. 38(1): 39-57.
- Murad, E. & Rojik, P. 2004. Jarosite, schwertmannite, goethite, ferrihydrite and lepidocrocite: the legacy of coal and sulfide ore mining. 3rd Australian-New Zealand Soils Conference.
- Mureşan, L., Maurin, G., Oniciu, L. & Avram, S. 1996a. Effects of additives on zinc electrowinning from industrial waste products. *Hydrometallurgy*. 40(3): 335-342.
- Mureşan, L., Maurin, G., Oniciu, L. & Gaga, D. 1996b. Influence of metallic impurities on zinc electrowinning from sulphate electrolyte. *Hydrometallurgy*. 43(1-3): 345-354.
- Myerson, A. 2002. *Handbook of industrial crystallization*. Butterworth-Heinemann.
- News24 2014a. Eskom needs maintenance, not just new plants.
- News24 2014b. Medupi unit gets its boiler certificate.
- Nicol, M.J., Senanayake, G. & Welham, N. 2008. *Hydrometallurgy-Theory and Practice*. (Unpublished).
- Ogbonna, N. 2006. *Mathematical Modelling of Agglomerate Scale Phenomena in Heap Bioleaching*.
- Ojumu, T.V. & Petersen, J. 2011. The kinetics of ferrous ion oxidation by *Leptospirillum ferriphilum* in continuous culture: The effect of pH. *Hydrometallurgy*. 106: 5-11.
- Ojumu, T.V., Hansford, G.S. & Petersen, J. 2009. The kinetics of ferrous-iron oxidation by *Leptospirillum ferriphilum* in continuous culture: The effect of temperature. *Biochemical Engineering*. 46: 161-168.

- Ojumu, T.V., Petersen, J. & Hansford, G.S. 2008. The effect of dissolved cations on microbial ferrous-iron oxidation by *Leptospirillum ferriphilum* in continuous culture. *Hydrometallurgy*. 94: 69-76
- O'Keefe, T.J. 1993. Method for Stripping Metals in Solvent Extraction
- O'Keefe, T., O'Keefe, M., Fang, R., Sun, J. & Dahlgren, E. 2002. Novel electrochemical processing using conventional organic solvents. Proceedings of the International Solvent Extraction Conference. 17-21 March 2002. K.C. Sole, P.M. Cole, J.S. Preston & D.J. Robinson, Eds. Chris Van Rensburg Publications (Pty) Ltd. 459.
- Outotec 2014. Outotec Hydrometallurgical Zinc Plants and Processes. Available: [http://www.outotec.com/ImageVaultFiles/id\\_1161/d\\_1/cf\\_2/OTE\\_Outotec\\_Hydrometallurgical\\_Zinc\\_Plants\\_and\\_Pro.PDF](http://www.outotec.com/ImageVaultFiles/id_1161/d_1/cf_2/OTE_Outotec_Hydrometallurgical_Zinc_Plants_and_Pro.PDF). [10 August 2014].
- Padilla, G.A., Cisternas, L.A. & Cueto, J.Y. 2008. On the optimization of heap leaching. *Minerals Engineering*. 21(9): 673-678.
- Petersen, J. & Dixon, D. 2007. Modelling zinc heap bioleaching. *Hydrometallurgy*. 85(2): 127-143.
- Porter GeoConsultancy 2001. Ore deposit description. Available: <http://www.portergeo.com.au/database/mineinfo.asp?mineid=mn017> [23 April 2014].
- Pradhan, N., Nathsarma, K.C., Srinivasa Rao, K., Sukla, L.B. & Mishra, B.K. 2008. Heap bioleaching of chalcopyrite: A review. *Minerals Engineering*. 21(5): 355-365.
- PwC & MCX 2011. Data for Ready Reference Non-Agricultural Commodities. In Commodity insights yearbook 2011. Available: <http://www.mcxindia.com/knowledgehub/overview/PDF/2011/NON-Agricultural/Zinc.pdf> ed.138-143.
- Ramokgopa, B. 2001. Tariff Histry. Available: <http://www.eskom.co.za/CustomerCare/TariffsAndCharges/Documents/TariffHistry.pdf> [30 March 2014].
- Ramokgopa, B. 2007. Tariff History: 2002 - 2007. Available: [http://www.eskom.co.za/CustomerCare/TariffsAndCharges/Documents/Tariff\\_History\\_2007\\_Edited\\_Draft.pdf](http://www.eskom.co.za/CustomerCare/TariffsAndCharges/Documents/Tariff_History_2007_Edited_Draft.pdf) [30 March 2014].
- Rawlings, D.E. 2004. Microbially-assisted dissolution of minerals and its use in the mining industry. *Pure and Applied Chemistry*. 76(4): 847-859.

- Reid, D.L., Smith, C., Watkeys, M., Welke, H. & Betton, P. 1997. Whole rock radiometric age patterns in the Aggeneys-Gamsberg ore district, central Bushmanland, South Africa. Department of Geological Sciences, University of Cape Town.
- Reuters 2014. UPDATE 1-India's Sesa Sterlite to invest \$782 mln for zinc deposits in Africa. Available: <http://af.reuters.com/article/southAfricaNews/idAFL3N0T34DY20141113> [11 December 2014]
- Ritcey, G.M. & Ashbrook, A.W. 1984. Solvent extraction: principles and applications to process metallurgy. New York: Elsevier.
- Saba, A. & Elsherief, A. 2000. Continuous electrowinning of zinc. *Hydrometallurgy*. 54(2): 91-106.
- Saxén & Björn 2008. Zinc Direct Leaching. Available: [http://www.outotec.com/Global/Investors/2008/CMD\\_2008/4%20Zinc%20direct%20leaching%20Bjorn%20Saxen%20Aug27%202008.pdf](http://www.outotec.com/Global/Investors/2008/CMD_2008/4%20Zinc%20direct%20leaching%20Bjorn%20Saxen%20Aug27%202008.pdf) [10 August 2014].
- Schouwstra, R., de Vaux, D., Hey, P., Malysiak, V., Shackleton, N. & Bramdeo, S. 2010. Understanding Gamsberg – A geometallurgical study of a large stratiform zinc deposit. *Minerals Engineering*. 23(11–13): 960-967.
- Seider, W.D., Seader, J.D. & Lewin, D.R. 2003. Cost accounting and Capital Cost Estimation. In *Product & Process Design Principles*. 2nd ed. John Wiley & Sons Ltd. 472-558.
- Sesa Sterlite 2014. Sesa Sterlite Announces Development of Gamsberg-Skorpion Integrated Zinc Project. Available: <http://www.mining-technology.com/news/newssesa-sterlite-to-invest-782m-in-gamsberg-skorpion-integrated-zinc-project-4438682> [11 December 2014]
- Shapiro, A.C. 2004a. Capital-Budgeting Principles and Techniques. In *Capital Budgeting and Investment Analysis*. Prentice Hall. 13-35.
- Shapiro, A.C. 2004b. Introduction to Capital Budgeting. In *Capital Budgeting and Investment Analysis*. Prentice Hall. 1-12.
- Shi, H., Peng, S., Liu, Y. & Zhong, P. 2008. Barriers to the implementation of cleaner production in Chinese SMEs: government, industry and expert stakeholders' perspectives. *Journal of Cleaner Production*. 16(7): 842-852.

- Söderström, U. 2008. Zinc smelter revenue stream. Available: [http://investors.boliden.com/afw/files/press/boliden/Kokkola-2008\\_5\\_Smelters\\_Zinc\\_US.pdf](http://investors.boliden.com/afw/files/press/boliden/Kokkola-2008_5_Smelters_Zinc_US.pdf) [09 May 2014].
- Stalder, M. & Rozendaal, A. 2004. Apatite nodules as an indicator of depositional environment and ore genesis for the Mesoproterozoic Broken Hill-type Gamsberg Zn–Pb deposit, Namaqua Province, South Africa. *Mineralium Deposita*. 39(2): 189-203.
- Stuckey, J. & White, D. 1993. When and when not to vertically integrate. Available: [www.mckinsey.com/insights/strategy/when\\_and\\_when\\_not\\_to\\_vertically\\_integrate](http://www.mckinsey.com/insights/strategy/when_and_when_not_to_vertically_integrate) [05 May 2014].
- Swarnkar, S.R., Gupta, B.L. & Sekharan, R.D. 1996. Iron control in zinc plant residue leach solution. *Hydrometallurgy*. 42(1): 21-26.
- Teck 2012. Modelling Workshop. Available: <http://www.teck.com/DocumentViewer.aspx?elementId=201905&portalName=tc>. [09 May 2014].
- Turton, R., Bailie, R.C., Whiting, W.B. & Shaeiwitz, J.A. 2008. Profitability Analysis. In *Analysis, Synthesis, and Design of Chemical Processes*. 3rd ed. Pearson Education. 297-345.
- U.S. Environmental Protection Agency 1994. Design and Evaluation of Tailings Dams. (EPA 530-R-94-038).
- Ulrich, B., Andrade, H. & Gardner, T. 2003. Lessons learnt from heap leaching operations in South America-An update. *Journal-South African Institute of Mining and Metallurgy*. 103(1): 23-28.
- Van Zyl, D., Henderson, M. & Cobb, B. 1990. Economic aspects of pad construction costs on heap leach projects. *International Journal of Mining and Geological Engineering*. 8(4): 275-286.
- Vecchiato, P. & Mantshantsha, S. 2013. Third new coal power station to help remove energy constraints. Available: <http://www.bdlive.co.za/business/energy/2013/08/23/third-new-coal-power-station-to-help-remove-energy-constraints> [11 December 2014]
- Venkateswaran, K., Srinivasan, G. & Nandakumar, V. 1996. Electrowinning of zinc-Effect of metallic impurities and addition agents. *Bulletin of Electrochemistry*. 12(05): 349-351.
- Vick, S.G. 1990. Planning, Design, and Analysis of Tailings Dams. BiTech.

- Watling, H. 2006. The bioleaching of sulphide minerals with emphasis on copper sulphides—a review. *Hydrometallurgy*. 84(1): 81-108.
- Westmetall 2014. Market data - Zinc. Available: [http://www.westmetall.com/en/markdaten.php?action=show\\_table&field=LME\\_Zn\\_cash](http://www.westmetall.com/en/markdaten.php?action=show_table&field=LME_Zn_cash) [28 October 2014].
- Wikipedia 2009. Eskom. Available: <http://en.wikipedia.org/wiki/Eskom> [28 October 2014].
- Williams, M. 2013. *Commodities and Market Outlook – Zinc Presentation*. (Unpublished).
- Wyethe, J.P., Vegter, N.M. & Rodrigues, J.M. 2008. Recent Recovery Improvement Initiatives at ZINCOR: The use of Zinc Oxide Fume and Low-Ferrite Calcine for Iron Removal to Reduce Un-dissolved Zinc losses.
- Yelland, Chris. 2015. Eskom’s Medupi mess puts South Africa at risk. Available: <http://mybroadband.co.za/news/energy/116462-eskoms-medupi-mess-puts-south-africa-at-risk-html> [12 January 2015]
- Zambak, C. 2012. Heap Leaching Technique in Mining Within the Context of Best Available Technigues (BAT). Available: <http://www.euromines.org/sites/default/files/content/files/mining-techniques/batforheapleaching-feb2013-c.zambak-euromines.pdf> [20 October 2013].
- Zhu, J., Gan, M., Zhang, D., Hu, Y. & Chai, L. 2013. The nature of Schwertmannite and Jarosite mediated by two strains of *Acidithiobacillus ferrooxidans* with different ferrous oxidation ability. *Materials Science and Engineering: C*. 33(5): 2679-2685.
- Zinc International 2014. Vendata Zinc International Gamsberg-Skorpion Integrated Zinc Project South Africa & Namimbia. Available: [http://www.vedanta-zincinternational.com/pdf/2.2-Gamsberg-Skorpion-Integrated-Project%20PR-Presentation-\(20141113\)-vF1.pdf](http://www.vedanta-zincinternational.com/pdf/2.2-Gamsberg-Skorpion-Integrated-Project%20PR-Presentation-(20141113)-vF1.pdf) [11 December 2014]
- Zinc Investing News 2013a. Top 10 Zinc-producing Countries of 2012. Available: <http://zincinvestingnews.com/6534-top-10-zinc-producing-countries-2012-china-australia-united-states-canada-peru.html> [25 April 2014].
- Zinc Investing News 2013b. The world's Top 3 Zinc Mines: An Overview. Available: <http://zincinvestingnews.com/6212-top-3-zinc-mines-australia-alaska-glencore-xstrata-mmg-teck.html> [25 April 2014].

- Zinc Investing News 2014a. 2013 Top 10 Zinc-producing Countries. Available: <http://zincinvestingnews.com/8275-top-zinc-producing-countries-2013-china-australia-peru.html> [03 June 2014].
- Zinc Investing News 2014b. Zinc to Hit \$ 4,000 per Tonne in the next few years. Available: <http://zincinvestingnews.com/8820-zinc-to-hit-4000-per-tonne-in-next-few-years-cru-consulting.html> [12 September 2014]

# Appendix A

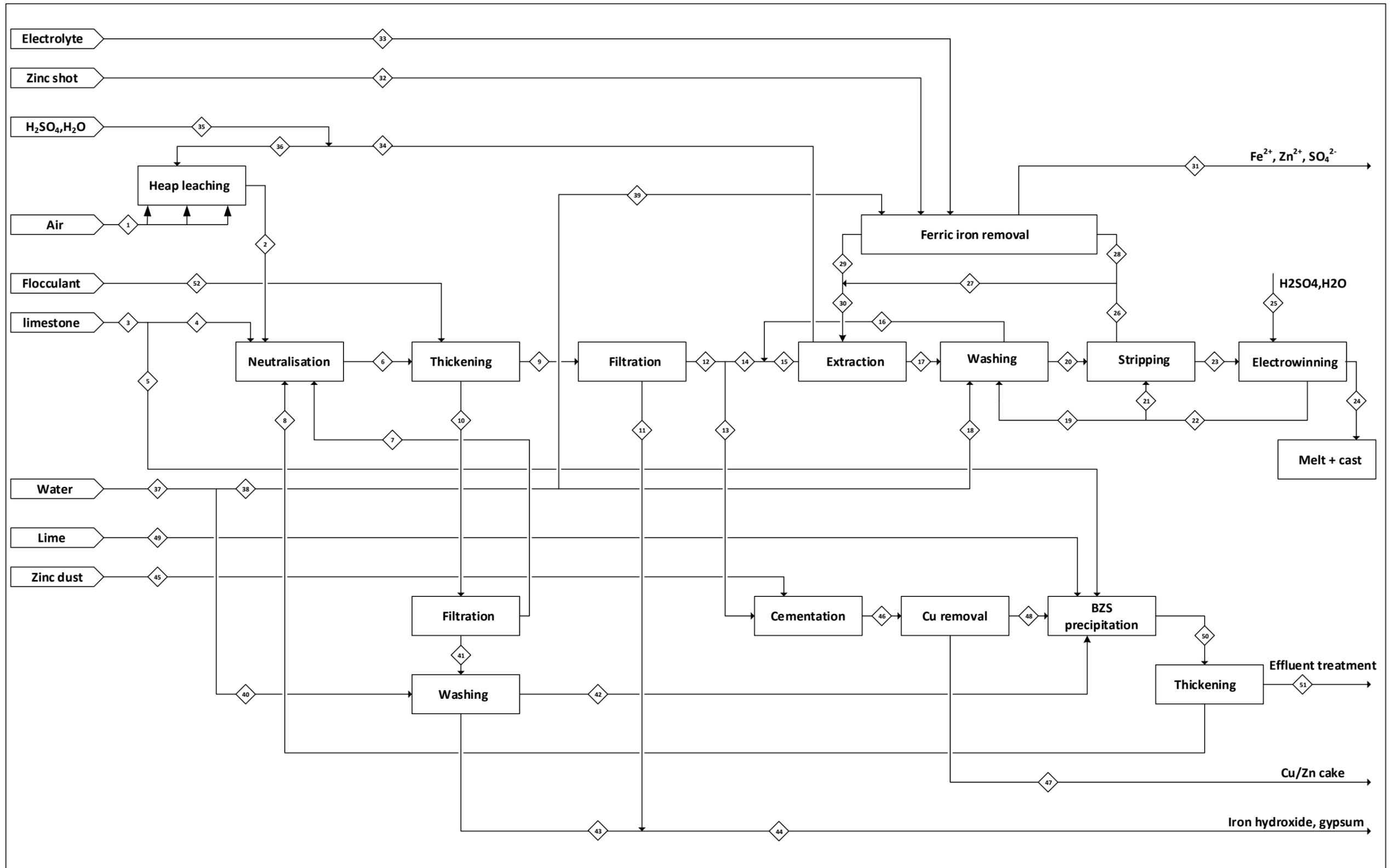


Figure A-1: Route 1 Block Flow Diagram

**Table A-1: Stream Table for route 1**

stream number		1	2	3	4	5	6	7	8	9	10	11	12	13
stream name		Air into HL	PLS	Total limestone	Limestone to NT Stage	Limestone to BZS	ZnS slurry	filtrate from NT thickner U/F	BSZ thickner U/F	NT thickner O/F	NT thickner U/F	Filter residue	filtrate from NT thickner O/F	BT
Flow	t/h	111.00	4720.00	245.00	245.00	0.07	5010.00	89.20	11.80	4640.00	363.00	5.08	4640.00	232.00
Flow	m3/h	93800.00	2420.00	132.00	132.00	0.04	2390.00	44.10	4.52	2230.00	154.00	0.22	2230.00	112.00
Solids	wt %	0%	0%	50%	50%	50%	5%	0%	62%	0%	65%	91%	0%	0%
Solids	t/h	0.00	0.00	123.00	123.00	0.04	243.00	0.00	7.26	4.61	237.00	4.61	0.00	0.00
Aq. Liquids	t/h	0.00	4720.00	123.00	123.00	0.04	4770.00	89.20	4.51	4640.00	126.00	0.46	4640.00	232.00
Gases	t/h	111.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Org. Liquids	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Zn	g/L	0.00	19.10	0.00	0.00	0.00	20.60	20.90	0.81	21.50	0.00	0.00	21.50	21.50
Aq. Liquids: Fe	g/L	0.00	16.80	0.00	0.00	0.00	0.11	0.11	0.00	0.11	0.00	0.00	0.11	0.11
Aq. Liquids: H2SO4	g/L	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Zn	t/h	0.00	0.00	0.00	0.00	0.00	49.90	0.00	1.93	0.00	0.00	0.00	0.00	0.00
Solids: Fe	t/h	0.00	0.00	0.00	0.00	0.00	40.40	0.00	0.01	0.00	40.40	0.00	0.00	0.00
Solids: Cu	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Ca	t/h	0.00	0.00	49.10	49.10	0.01	50.60	0.00	0.89	0.00	50.60	0.00	0.00	0.00
Solids: Mn	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Mg	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: H2O	t/h	0.00	0.00	0.00	0.00	0.00	5.14	0.00	0.02	0.00	0.02	0.00	0.00	0.00
Solids: SO4	t/h	0.00	0.00	0.00	0.00	0.00	121.00	0.00	2.86	0.00	121.00	0.00	0.00	0.00
Aq. Liquids: Zn	t/h	0.00	46.40	0.00	0.00	0.00	49.20	0.92	0.00	47.90	1.30	0.00	47.90	2.40
Aq. Liquids: Fe	t/h	0.00	40.70	0.00	0.00	0.00	0.26	0.00	0.00	0.25	0.01	0.00	0.25	0.01
Aq. Liquids: Cu	t/h	0.00	0.40	0.00	0.00	0.00	0.40	0.01	0.00	0.39	0.01	0.00	0.39	0.02
Aq. Liquids: Ca	t/h	0.00	0.77	0.00	0.00	0.00	0.16	0.00	0.00	0.16	0.00	0.00	0.16	0.01
Aq. Liquids: Mn	t/h	0.00	11.90	0.00	0.00	0.00	12.10	0.23	0.00	11.80	0.32	0.00	11.80	0.59
Aq. Liquids: Mg	t/h	0.00	2.76	0.00	0.00	0.00	2.81	0.05	0.00	2.74	0.07	0.00	2.74	0.14
Aq. Liquids: H2O	t/h	0.00	3780.00	123.00	123.00	0.04	3980.00	74.40	4.33	3870.00	105.00	0.39	3870.00	194.00
Aq. Liquids: H2SO4	t/h	0.00	13.90	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq.SO4	t/h	0.00	805.00	0.00	0.00	0.00	700.00	13.10	0.17	681.00	18.50	0.07	681.00	34.00
aq.H+	t/h	0.00	24.60	0.00	0.00	0.00	25.00	0.47	0.01	24.40	0.66	0.00	24.40	1.22
Aq. Liquids: R2Zn	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: R3Fe	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
aq.SO4	g/L	0.00	332.00	0.00	0.00	0.00	305.00	305.00	38.40	305.00	120.00	305.00	305.00	305.00
aq.H+	g/L	0.00	702.00	0.00	0.00	0.00	10.90	10.90	1.49	10.90	10.90	10.90	10.90	10.90
Aq. Liquids: Cu	g/L	0.00	0.16	0.00	0.00	0.00	0.18	0.18	0.00	0.18	0.18	0.18	0.18	0.18
Aq. Liquids: Ca	g/L	0.00	0.32	0.00	0.00	0.00	21.30	0.07	0.03	0.07	328.00	0.07	0.07	0.07
Aq. Liquids: Mn	g/L	0.00	4.90	0.00	0.00	0.00	5.28	5.28	0.72	5.28	5.28	5.28	5.28	5.28
Aq. Liquids: Mg	g/L	0.00	1.14	0.00	0.00	0.00	1.23	1.23	0.17	1.23	1.23	1.23	1.23	1.23
Aq. Liquids: H2O	g/L	0.00	1560.00	0.00	0.00	0.00	1670.00	1690.00	958.00	1730.00	683.00	1730.00	1730.00	1730.00

**Table A-1: Stream Table for route 1**

stream number		14	15	16	17	18	19	20	21	22	23	24	25	26
stream name		PLS to SX	SX Feed	Scrub Raffinate	Loaded Solvent	Scrub H2O	Scrub electrolyte	Scrubbed solvent	Strip solution	Strip + Scrub solution	Advance electrolyte	Zn product	Acid + H2O to EW	Stripped solvent
Flow	t/h	4410.00	4800.00	396.00	702.00	99.20	39.50	444.00	750.00	790.00	786.00	23.40	36.40	409.00
Flow	m3/h	2120.00	2490.00	370.00	1980.00	99.20	146.00	1860.00	2780.00	2920.00	2780.00	3.28	146.00	1860.00
Solids	wt %	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	100%	0%	0%
Solids	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	23.40	0.00	0.00
Aq. Liquids	t/h	4410.00	4800.00	396.00	702.00	99.20	39.50	444.00	750.00	790.00	786.00	0.00	36.40	46.10
Gases	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Org. Liquids	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	363.00
Aq. Liquids: Zn	g/L	21.50	29.40	74.60	26.00	0.00	90.00	20.00	90.00	90.00	103.00	0.00	0.00	20.00
Aq. Liquids: Fe	g/L	0.11	0.10	0.00	0.02	0.00	0.00	0.01	0.00	0.00	0.00	0.00	0.00	0.01
Aq. Liquids: H2SO4	g/L	0.00	0.00	0.00	0.00	0.00	180.00	0.00	180.00	180.00	0.00	0.00	180.00	0.00
Solids: Zn	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	23.40	0.00	0.00
Solids: Fe	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Cu	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Ca	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Mn	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Mg	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: H2O	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: SO4	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Zn	t/h	45.50	73.10	27.60	51.60	0.00	13.20	37.20	250.00	263.00	286.00	0.00	0.00	37.20
Aq. Liquids: Fe	t/h	0.24	0.25	0.01	0.03	0.00	0.00	0.02	0.00	0.00	0.00	0.00	0.00	0.02
Aq. Liquids: Cu	t/h	0.37	0.58	0.21	0.38	0.00	0.00	0.18	0.01	0.01	0.01	0.00	0.00	0.17
Aq. Liquids: Ca	t/h	0.15	0.23	0.08	0.15	0.00	0.00	0.07	0.00	0.00	0.00	0.00	0.00	0.07
Aq. Liquids: Mn	t/h	11.20	17.50	6.29	11.50	0.00	0.01	5.24	0.24	0.26	0.27	0.01	0.00	5.21
Aq. Liquids: Mg	t/h	2.60	4.06	1.46	2.68	0.00	0.00	1.22	0.06	0.06	0.06	0.00	0.00	1.21
Aq. Liquids: H2O	t/h	3680.00	3980.00	298.00	199.00	99.20	0.00	0.00	0.00	0.00	0.00	0.00	10.10	0.00
Aq. Liquids: H2SO4	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	26.30	0.00
Aq.SO4	t/h	647.00	708.00	61.20	35.40	0.00	25.80	0.00	490.00	515.00	490.00	0.00	25.80	0.00
aq.H+	t/h	23.10	24.90	1.78	1.25	0.00	0.54	0.00	10.20	10.70	9.07	0.00	0.54	0.00
Aq. Liquids: R2Zn	t/h	0.00	0.00	0.00	398.00	0.00	0.00	398.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: R3Fe	t/h	0.00	0.00	0.00	2.27	0.00	0.00	2.27	0.00	0.00	0.00	0.00	0.00	2.27
aq.SO4	g/L	305.00	284.00	165.00	17.90	0.00	176.00	214.00	176.00	176.00	176.00	0.00	176.00	0.00
aq.H+	g/L	10.90	10.00	4.82	0.63	0.00	3.67	1.22	3.67	3.67	3.27	0.00	3.67	0.00
Aq. Liquids: Cu	g/L	0.18	0.23	0.57	0.19	0.00	0.00	0.09	0.00	0.00	0.00	0.00	0.00	0.09
Aq. Liquids: Ca	g/L	0.07	0.09	0.22	0.08	0.00	0.00	0.04	0.00	0.00	0.00	0.00	0.00	0.04
Aq. Liquids: Mn	g/L	5.28	7.02	17.00	5.81	0.00	0.09	2.82	0.09	0.09	0.10	0.00	0.00	2.81
Aq. Liquids: Mg	g/L	1.23	1.63	3.95	1.35	0.00	0.02	0.66	0.02	0.02	0.02	0.00	0.00	0.65
Aq. Liquids: H2O	g/L	1730.00	1600.00	806.00	100.00	1000.00	0.00	0.00	0.00	0.00	0.00	0.00	69.20	0.00

**Table A-1: Stream Table for route 1**

stream number		27	28	29	30	31	32	33	34	35	36	37	38	39
stream name		Stripped solvent split to extraction	Stripped solvent for Fe removal	Stripped solvent from Fe removal	Solvent Feed to extraction	electrolyte from Fe reduction	Zinc shot	Electrolyte for Fe reduction	Raffinate to heap	make-up acid + H2O	Recycle to heap	Total water in the process	water for scrub +Fe reduction	Water for Fe reduction
Flow	t/h	368.00	40.90	40.30	408.00	62.60	0.01	0.03	4510.00	14.50	4520.00	926.00	161.00	62.00
Flow	m3/h	1670.00	186.00	186.00	1860.00	62.00	0.00	0.02	2370.00	58.30	2420.00	926.00	161.00	62.00
Solids	wt %	0%	0%	0%	0%	0%	100%	0%	0%	0%	0%	0%	0%	0%
Solids	t/h	0.00	0.00	0.00	0.00	0.00	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids	t/h	41.50	4.61	3.83	45.30	62.60	0.00	0.03	4510.00	14.50	0.00	926.00	161.00	62.00
Gases	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Org. Liquids	t/h	326.00	36.30	36.50	363.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Zn	g/L	20.00	20.00	20.00	20.00	0.12	0.00	0.00	9.36	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Fe	g/L	0.01	0.01	0.01	0.01	0.20	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: H2SO4	g/L	0.00	0.00	0.00	0.00	0.00	0.00	0.00	100.00	180.00	15.20	0.00	0.00	0.00
Solids: Zn	t/h	0.00	0.00	0.00	0.00	0.00	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Fe	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Cu	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Ca	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Mn	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Mg	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: H2O	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: SO4	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Zn	t/h	33.50	3.72	3.72	37.20	0.01	0.01	0.00	22.10	0.00	37.20	0.00	0.00	0.00
Aq. Liquids: Fe	t/h	0.02	0.00	0.00	0.02	0.01	0.00	0.00	0.22	0.00	0.02	0.00	0.00	0.00
Aq. Liquids: Cu	t/h	0.16	0.02	0.00	0.16	0.01	0.00	0.00	0.36	0.00	0.16	0.00	0.00	0.00
Aq. Liquids: Ca	t/h	0.06	0.01	0.00	0.06	0.01	0.00	0.00	0.14	0.00	0.06	0.00	0.00	0.00
Aq. Liquids: Mn	t/h	4.69	0.52	0.09	4.78	0.43	0.00	0.00	10.80	0.00	4.78	0.00	0.00	0.00
Aq. Liquids: Mg	t/h	1.09	0.12	0.02	1.11	0.10	0.00	0.00	2.50	0.00	1.11	0.00	0.00	0.00
Aq. Liquids: H2O	t/h	0.00	0.00	0.00	0.00	62.00	0.00	0.00	3780.00	4.04	3780.00	926.00	161.00	62.00
Aq. Liquids: H2SO4	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.03	0.00	10.50	36.80	0.00	0.00	0.00
Aq.SO4	t/h	0.00	0.00	0.00	0.00	0.03	0.00	0.03	673.00	10.30	683.00	0.00	0.00	0.00
aq.H+	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	24.80	0.21	25.00	0.00	0.00	0.00
Aq. Liquids: R2Zn	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: R3Fe	t/h	2.04	0.23	0.00	2.04	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
aq.SO4	g/L	0.00	0.00	0.00	0.00	0.52	0.00	0.01	284.00	176.00	282.00	0.00	0.00	0.00
aq.H+	g/L	0.00	0.00	0.00	0.00	0.00	0.00	0.00	10.50	3.67	10.30	0.00	0.00	0.00
Aq. Liquids: Cu	g/L	0.09	0.09	0.02	0.09	0.23	0.00	0.00	0.15	0.00	0.07	0.00	0.00	0.00
Aq. Liquids: Ca	g/L	0.04	0.04	0.01	0.03	0.09	0.00	0.00	0.06	0.00	0.03	0.00	0.00	0.00
Aq. Liquids: Mn	g/L	2.81	2.81	0.47	2.57	7.01	0.00	0.00	4.55	0.00	1.97	0.00	0.00	0.00
Aq. Liquids: Mg	g/L	0.65	0.65	0.11	0.60	1.63	0.00	0.00	1.06	0.00	0.46	0.00	0.00	0.00
Aq. Liquids: H2O	g/L	0.00	0.00	0.00	0.00	1000.00	0.00	0.00	0.00	69.20	1560.00	1000.00	1000.00	1000.00

Table A-1: Stream Table for route 1

stream number		40	41	42	43	44	45	46	47	48	49	50	51	52
stream name		Fe residue wash water	Belt filter residue	Wash water to BZS precipitation	Washed residue	solids stream out	Zn dust to cementation	cementation products	Cementation filter residue	Cementation filtrate	Lime to BZS precipitation	BZS product	BZS thickener O/F	Flocculant to NT thickener
Flow	t/h	765.00	274.00	675.00	365.00	370.00	0.02	232.00	0.02	232.00	3.30	910.00	898.00	4.86
Flow	m3/h	765.00	110.00	721.00	155.00	155.00	0.00	112.00	0.06	112.00	2.06	834.00	830.00	0.00
Solids	wt %	0%	86%	0%	65%	65%	100%	0%	91%	0%	50%	1%	0%	100%
Solids	t/h	0.00	237.00	0.00	237.00	242.00	0.02	0.02	0.02	0.00	1.65	7.26	0.00	4.86
Aq. Liquids	t/h	0.00	37.10	675.00	128.00	128.00	0.00	232.00	0.00	232.00	1.65	903.00	898.00	0.00
Gases	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Org. Liquids	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Zn	g/L	0.00	21.50	0.00	0.00	0.90	0.00	21.60	0.35	21.70	0.00	0.88	0.88	0.00
Aq. Liquids: Fe	g/L	0.00	366.00	0.00	0.00	0.00	0.00	0.11	0.00	0.11	0.00	0.00	0.00	0.00
Aq. Liquids: H2SO4	g/L	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Zn	t/h	0.00	0.00	0.00	0.00	0.00	0.02	0.00	0.00	0.00	0.00	1.93	0.00	0.00
Solids: Fe	t/h	0.00	40.40	0.00	40.40	40.40	0.00	0.00	0.00	0.00	0.00	0.01	0.00	0.00
Solids: Cu	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.02	0.02	0.00	0.00	0.00	0.00	0.00
Solids: Ca	t/h	0.00	50.60	0.00	50.60	50.60	0.00	0.00	0.00	0.00	0.89	0.89	0.00	0.00
Solids: Mn	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Mg	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: H2O	t/h	0.00	5.14	0.00	5.14	5.14	0.00	0.00	0.00	0.00	0.00	0.02	0.00	0.00
Solids: SO4	t/h	0.00	121.00	0.00	121.00	121.00	0.00	0.00	0.00	0.00	0.00	2.86	0.00	0.00
Aq. Liquids: Zn	t/h	0.00	0.38	0.25	0.13	0.14	0.00	2.42	0.00	2.42	0.00	0.73	0.73	0.00
Aq. Liquids: Fe	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.01	0.00	0.01	0.00	0.00	0.00	0.00
Aq. Liquids: Cu	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Ca	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.01	0.00	0.01	0.00	0.02	0.02	0.00
Aq. Liquids: Mn	t/h	0.00	0.09	0.06	0.03	0.03	0.00	0.59	0.00	0.59	0.00	0.65	0.65	0.00
Aq. Liquids: Mg	t/h	0.00	0.02	0.01	0.01	0.01	0.00	0.14	0.00	0.14	0.00	0.15	0.15	0.00
Aq. Liquids: H2O	t/h	765.00	31.00	671.00	125.00	126.00	0.00	194.00	0.00	194.00	1.65	865.00	861.00	0.00
Aq. Liquids: H2SO4	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq.SO4	t/h	0.00	5.45	3.54	1.91	1.98	0.00	34.00	0.00	34.00	0.00	34.70	34.60	0.00
aq.H+	t/h	0.00	0.20	0.13	0.07	0.07	0.00	1.22	0.00	1.22	0.00	1.34	1.34	0.00
Aq. Liquids: R2Zn	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: R3Fe	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
aq.SO4	g/L	0.00	49.40	0.00	0.00	0.00	0.00	305.00	4.88	305.00	0.00	41.60	41.70	0.00
aq.H+	g/L	0.00	1.77	0.00	0.00	0.00	0.00	10.90	0.17	10.90	0.00	1.61	1.61	0.00
Aq. Liquids: Cu	g/L	0.00	0.03	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Ca	g/L	0.00	0.01	0.00	0.00	0.00	0.00	0.07	0.00	0.07	0.00	0.03	0.03	0.00
Aq. Liquids: Mn	g/L	0.00	0.86	0.00	0.00	0.00	0.00	5.28	0.08	5.29	0.00	0.78	0.78	0.00
Aq. Liquids: Mg	g/L	0.00	0.20	0.00	0.00	0.00	0.00	1.23	0.02	1.23	0.00	0.18	0.18	0.00
Aq. Liquids: H2O	g/L	1000.00	281.00	0.00	0.00	0.00	0.00	1730.00	27.80	1740.00	803.00	1040.00	1040.00	0.00

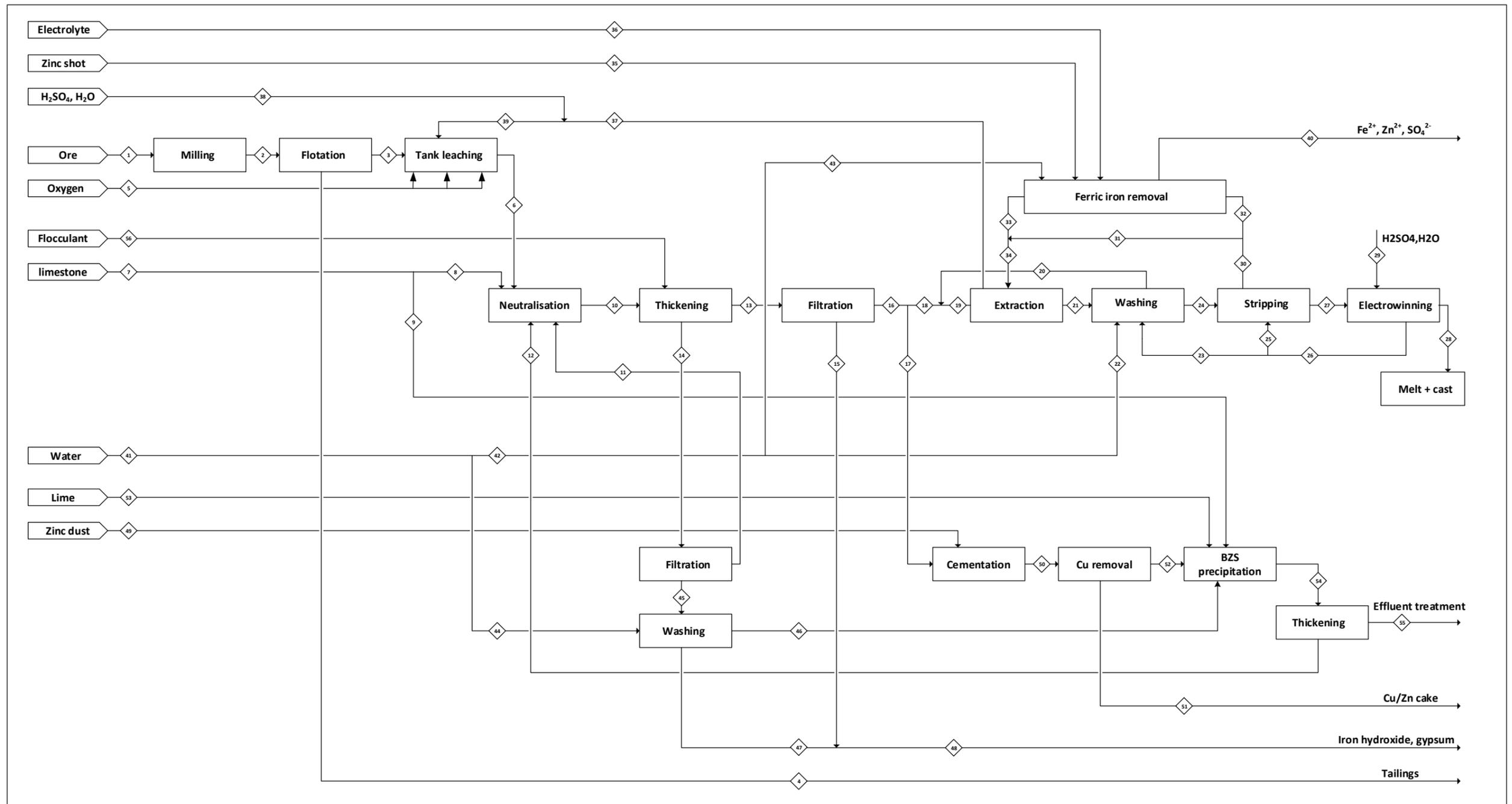


Figure A-2: Route 2 Block Flow Diagram

**Table A-2: Stream Table for route 2**

stream number		1	2	3	4	5	6	7	8	9	10	11	12	13
stream name		Crushed Ore	Milled Ore	Concentrate	Tailings	O2 into TL	PLS	Total limestone	Limestone to NT Stage	Limestone to BZS	ZnS slurry	filtrate from NT thickner U/F	BSZ thickner U/F	NT thickner O/F
Flow	t/h	451.00	451.00	62.60	389.00	9.50	3740.00	24.30	24.30	0.03	3790.00	13.30	21.70	3740.00
Flow	m3/h	115.00	115.00	0.00	0.00	7260.00	1170.00	13.10	13.10	0.02	929.00	4.04	7.05	911.00
Solids	wt %	100.00%	100.00%	100.00%	0.00%	0.00%	0.00%	50.00%	50.00%	50.00%	1.00%	0.00%	62.00%	0.00%
Solids	t/h	451.00	451.00	62.60	0.00	0.00	0.00	12.10	12.10	0.02	32.90	0.00	13.40	0.46
Aq. Liquids	t/h	0.00	0.00	0.00	0.00	0.00	3740.00	12.10	12.10	0.02	3760.00	13.30	8.34	3740.00
Gases	t/h	0.00	0.00	0.00	0.00	9.50	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Org. Liquids	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Zn	g/L	0.00	0.00	0.00	0.00	0.00	73.20	0.00	0.00	0.00	96.40	78.20	4.59	97.90
Aq. Liquids: Fe	g/L	0.00	0.00	0.00	0.00	0.00	4.84	0.00	0.00	0.00	0.13	0.11	0.00	0.13
Aq. Liquids: H2SO4	g/L	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Zn	t/h	33.40	33.40	28.70	4.68	0.00	0.00	0.00	0.00	0.00	47.20	0.00	3.57	0.00
Solids: Fe	t/h	82.60	82.60	5.68	76.90	0.00	0.00	0.00	0.00	0.00	5.54	0.00	0.01	0.00
Solids: Cu	t/h	0.45	0.45	0.22	0.23	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Ca	t/h	0.84	0.84	0.12	0.72	0.00	0.00	4.86	4.85	0.01	6.47	0.00	1.65	0.00
Solids: Mn	t/h	2.22	2.22	1.10	1.12	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Mg	t/h	1.59	1.59	0.03	1.56	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: H2O	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.08	0.00	0.08	0.00
Solids: SO4	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	15.50	0.00	5.28	0.00
Aq. Liquids: Zn	t/h	0.00	0.00	0.00	0.00	0.00	85.60	0.00	0.00	0.00	89.60	0.32	0.03	89.10
Aq. Liquids: Fe	t/h	0.00	0.00	0.00	0.00	0.00	5.66	0.00	0.00	0.00	0.12	0.00	0.00	0.12
Aq. Liquids: Cu	t/h	0.00	0.00	0.00	0.00	0.00	2.10	0.00	0.00	0.00	2.11	0.01	0.00	2.10
Aq. Liquids: Ca	t/h	0.00	0.00	0.00	0.00	0.00	1.62	0.00	0.00	0.00	1.66	0.01	0.00	1.65
Aq. Liquids: Mn	t/h	0.00	0.00	0.00	0.00	0.00	10.60	0.00	0.00	0.00	10.70	0.04	0.02	10.60
Aq. Liquids: Mg	t/h	0.00	0.00	0.00	0.00	0.00	0.30	0.00	0.00	0.00	0.30	0.00	0.00	0.30
Aq. Liquids: H2O	t/h	0.00	0.00	0.00	0.00	0.00	2610.00	12.10	12.10	0.02	2640.00	9.32	6.91	2630.00
Aq. Liquids: H2SO4	t/h	0.00	0.00	0.00	0.00	0.00	1.15	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids SO4	t/h	0.00	0.00	0.00	0.00	0.00	989.00	0.00	0.00	0.00	984.00	3.47	1.34	979.00
Aq. Liquids H+	t/h	0.00	0.00	0.00	0.00	0.00	26.60	0.00	0.00	0.00	26.70	0.09	0.04	26.60
Aq. Liquids: R2Zn	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: R3Fe	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
aq. Liquids SO4	g/L	0.00	0.00	0.00	0.00	0.00	845.00	0.00	0.00	0.00	1070.00	1070.00	190.00	1070.00
aq. Liquids H+	g/L	0.00	0.00	0.00	0.00	0.00	759.00	0.00	0.00	0.00	29.20	29.20	5.76	29.20
Aq. Liquids: Cu	g/L	0.00	0.00	0.00	0.00	0.00	1.80	0.00	0.00	0.00	2.31	2.31	0.00	2.31
Aq. Liquids: Ca	g/L	0.00	0.00	0.00	0.00	0.00	1.39	0.00	0.00	0.00	8.76	1.45	0.39	1.82
Aq. Liquids: Mn	g/L	0.00	0.00	0.00	0.00	0.00	9.08	0.00	0.00	0.00	11.70	11.70	2.30	11.70
Aq. Liquids: Mg	g/L	0.00	0.00	0.00	0.00	0.00	0.26	0.00	0.00	0.00	0.33	0.33	0.07	0.33
Aq. Liquids: H2O	g/L	0.00	0.00	0.00	0.00	0.00	2230.00	0.00	0.00	0.00	2840.00	2310.00	980.00	2890.00

**Table A-2: Stream Table for route 2**

stream number		14	15	16	17	18	19	20	21	22	23	24	25	26
stream name		NT thickner U/F	Filter residue	filtrate from NT thickner O/F	BT	PLS to SX	SX Feed	Scrub Raffinate	Loaded Solvent	Scrub H2O	Scrub electrolyte	Scrubbed solvent	Strip solution	Strip + Scrub solution
Flow	t/h	51.10	0.62	3740.00	187.00	3550.00	3880.00	334.00	694.00	111.00	45.80	516.00	869.00	915.00
Flow	m3/h	18.20	0.00	911.00	45.50	865.00	1210.00	340.00	2210.00	111.00	169.00	2150.00	3220.00	3390.00
Solids	wt %	63.00%	74.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Solids	t/h	32.30	0.46	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids	t/h	18.80	0.16	3740.00	187.00	3550.00	3880.00	334.00	694.00	111.00	45.80	516.00	869.00	915.00
Gases	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Org. Liquids	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Zn	g/L	0.00	0.00	97.90	97.90	97.90	70.30	0.11	12.60	0.00	90.00	20.00	90.00	90.00
Aq. Liquids: Fe	g/L	0.00	0.00	0.13	0.13	0.13	0.10	0.00	0.01	0.00	0.00	0.01	0.00	0.00
Aq. Liquids: H2SO4	g/L	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	180.00	0.00	180.00	180.00
Solids: Zn	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Fe	t/h	5.54	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Cu	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Ca	t/h	6.47	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Mn	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Mg	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: H2O	t/h	0.08	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: SO4	t/h	15.50	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Zn	t/h	0.45	0.00	89.10	4.46	84.60	84.70	0.04	27.90	0.00	15.20	43.10	290.00	305.00
Aq. Liquids: Fe	t/h	0.00	0.00	0.12	0.01	0.11	0.12	0.01	0.03	0.00	0.00	0.02	0.00	0.00
Aq. Liquids: Cu	t/h	0.01	0.00	2.10	0.11	2.00	2.54	0.54	2.09	0.00	0.00	1.56	0.07	0.08
Aq. Liquids: Ca	t/h	0.01	0.00	1.65	0.08	1.57	1.99	0.42	1.65	0.00	0.00	1.23	0.06	0.06
Aq. Liquids: Mn	t/h	0.05	0.00	10.60	0.53	10.10	12.80	2.72	10.60	0.00	0.02	7.87	0.36	0.38
Aq. Liquids: Mg	t/h	0.00	0.00	0.30	0.02	0.29	0.36	0.08	0.30	0.00	0.00	0.22	0.01	0.01
Aq. Liquids: H2O	t/h	13.20	0.16	2630.00	131.00	2500.00	2740.00	248.00	137.00	111.00	0.00	0.00	0.00	0.00
Aq. Liquids: H2SO4	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids SO4	t/h	4.92	0.00	979.00	48.90	930.00	1010.00	80.40	50.50	0.00	29.90	0.00	567.00	597.00
Aq. Liquids H+	t/h	0.13	0.00	26.60	1.33	25.20	27.20	1.98	1.36	0.00	0.62	0.00	11.80	12.40
Aq. Liquids: R2Zn	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	461.00	0.00	0.00	461.00	0.00	0.00
Aq. Liquids: R3Fe	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.10	0.00	0.00	1.10	0.00	0.00
aq. Liquids SO4	g/L	271.00	0.00	1070.00	1070.00	1070.00	838.00	236.00	22.80	0.00	176.00	0.00	176.00	176.00
aq. Liquids H+	g/L	29.20	0.00	29.20	29.20	29.20	22.60	5.82	0.61	0.00	3.67	0.00	3.67	3.67
Aq. Liquids: Cu	g/L	2.31	0.00	2.31	2.31	2.31	2.10	1.58	0.95	0.00	0.00	0.00	0.02	0.02
Aq. Liquids: Ca	g/L	356.00	0.00	1.82	1.82	1.82	1.65	1.24	0.74	0.00	0.00	0.00	0.02	0.02
Aq. Liquids: Mn	g/L	11.70	0.00	11.70	11.70	11.70	10.60	7.99	4.77	0.00	0.00	0.00	0.11	0.11
Aq. Liquids: Mg	g/L	0.33	0.00	0.33	0.33	0.33	0.30	0.23	0.14	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: H2O	g/L	726.00	0.00	2890.00	2890.00	2890.00	2280.00	728.00	62.00	1000.00	0.00	0.00	0.00	0.00

**Table A-2: Stream Table for route 2**

stream number		27	28	29	30	31	32	33	34	35	36	37	38	39
stream name		Advance electrolyte	Zn product	Acid + H2O to EW	Stripped solvent	Stripped solvent split to extraction	Stripped solvent for Fe removal	Stripped solvent from Fe removal	Solvent Feed to extraction	Zinc shot	Electrolyte for Fe reduction	Raffinate to heap	make-up acid + H2O	Recycle to Tank
Flow	t/h	910.00	27.10	42.20	475.00	428.00	47.50	47.00	475.00	0.00	0.02	3660.00	6.23	3670.00
Flow	m3/h	3220.00	3.80	169.00	2150.00	1940.00	215.00	215.00	2150.00	0.00	0.01	1150.00	25.00	1170.00
Solids	wt %	0.00%	100.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	100.00%	0.00%	0.00%	0.00%	0.00%
Solids	t/h	0.00	27.10	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids	t/h	910.00	0.00	42.20	55.00	49.50	5.50	4.86	54.40	0.00	0.02	3660.00	6.23	0.00
Gases	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Org. Liquids	t/h	0.00	0.00	0.00	420.00	378.00	42.00	42.10	420.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Zn	g/L	103.00	0.00	0.00	20.00	0.00	0.00	0.00	20.00	0.00	0.00	50.30	0.00	0.00
Aq. Liquids: Fe	g/L	0.00	0.00	0.00	0.01	0.00	0.00	0.00	0.01	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: H2SO4	g/L	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	100.00	180.00	29.90
Solids: Zn	t/h	0.00	27.10	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Fe	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Cu	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Ca	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Mn	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Mg	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: H2O	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: SO4	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Zn	t/h	332.00	0.00	0.00	43.10	38.80	4.31	4.31	43.10	0.00	0.00	57.50	0.00	0.00
Aq. Liquids: Fe	t/h	0.00	0.00	0.00	0.02	0.02	0.00	0.00	0.02	0.00	0.00	0.11	0.00	0.00
Aq. Liquids: Cu	t/h	0.08	0.00	0.00	1.55	1.40	0.16	0.08	1.48	0.00	0.00	1.92	0.00	0.00
Aq. Liquids: Ca	t/h	0.06	0.00	0.00	1.22	1.10	0.12	0.06	1.16	0.00	0.00	1.51	0.00	0.00
Aq. Liquids: Mn	t/h	0.40	0.02	0.00	7.83	7.05	0.78	0.40	7.45	0.00	0.00	9.68	0.00	0.00
Aq. Liquids: Mg	t/h	0.01	0.00	0.00	0.22	0.20	0.02	0.01	0.21	0.00	0.00	0.28	0.00	0.00
Aq. Liquids: H2O	t/h	0.00	0.00	11.70	0.00	0.00	0.00	0.00	0.00	0.00	0.00	2610.00	1.73	2610.00
Aq. Liquids: H2SO4	t/h	0.00	0.00	30.50	0.00	0.00	0.00	0.00	0.00	0.00	0.02	0.00	4.50	35.00
Aq. Liquids SO4	t/h	567.00	0.00	29.90	0.00	0.00	0.00	0.00	0.00	0.00	0.02	960.00	4.41	964.00
Aq. Liquids H+	t/h	10.50	0.00	0.62	0.00	0.00	0.00	0.00	0.00	0.00	0.00	27.10	0.09	27.20
Aq. Liquids: R2Zn	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: R3Fe	t/h	0.00	0.00	0.00	1.10	0.99	0.11	0.00	0.99	0.00	0.00	0.00	0.00	0.00
aq. Liquids SO4	g/L	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
aq. Liquids H+	g/L	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	23.30
Aq. Liquids: Cu	g/L	0.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Ca	g/L	0.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.32	0.00	0.00
Aq. Liquids: Mn	g/L	0.13	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	8.46	0.00	0.00
Aq. Liquids: Mg	g/L	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.24	0.00	0.00
Aq. Liquids: H2O	g/L	0.00	0.00	69.20	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	69.20	0.00

Table A-2: Stream Table for route 2

stream number		40	41	42	43	44	45	46	47	48	49
stream name		electrolyte from Fe reaction	Total water in the process	water for scrub +Fe reduction	Water for Fe reduction	Fe residue wash water	Belt filter residue	Wash water to BZS precipitation	Washed residue	total outlet stream	Zn dust to cementation
Flow	t/h	72.30	293.00	182.00	71.80	110.00	37.80	98.50	49.70	50.40	0.11
Flow	m3/h	71.80	293.00	182.00	71.80	110.00	14.10	107.00	17.80	0.00	0.00
Solids	wt %	0.00%	0.00%	0.00%	0.00%	0.00%	85.00%	0.00%	65.00%	65.00%	100.00%
Solids	t/h	0.00	0.00	0.00	0.00	0.00	32.30	0.00	32.30	32.80	0.11
Aq. Liquids	t/h	72.30	293.00	182.00	71.80	0.00	5.52	98.50	17.40	17.60	0.00
Gases	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Org. Liquids	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Zn	g/L	0.05	0.00	0.00	0.00	0.00	97.90	0.00	0.00	0.00	0.00
Aq. Liquids: Fe	g/L	0.08	0.00	0.00	0.00	0.00	392.00	0.00	0.00	0.00	0.00
Aq. Liquids: H2SO4	g/L	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Zn	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.11
Solids: Fe	t/h	0.00	0.00	0.00	0.00	0.00	5.54	0.00	5.54	0.00	0.00
Solids: Cu	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Ca	t/h	0.00	0.00	0.00	0.00	0.00	6.47	0.00	6.47	0.00	0.00
Solids: Mn	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Mg	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: H2O	t/h	0.00	0.00	0.00	0.00	0.00	0.08	0.00	0.08	0.00	0.00
Solids: SO4	t/h	0.00	0.00	0.00	0.00	0.00	15.50	0.00	15.50	0.00	0.00
Aq. Liquids: Zn	t/h	0.00	0.00	0.00	0.00	0.00	0.13	0.09	0.04	0.00	0.00
Aq. Liquids: Fe	t/h	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Cu	t/h	0.08	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Ca	t/h	0.06	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Mn	t/h	0.38	0.00	0.00	0.00	0.00	0.02	0.01	0.01	0.00	0.00
Aq. Liquids: Mg	t/h	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: H2O	t/h	71.80	293.00	182.00	71.80	110.00	3.88	97.50	16.80	0.00	0.00
Aq. Liquids: H2SO4	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids SO4	t/h	0.02	0.00	0.00	0.00	0.00	1.45	0.96	0.48	0.00	0.00
Aq. Liquids H+	t/h	0.00	0.00	0.00	0.00	0.00	0.04	0.03	0.01	0.00	0.00
Aq. Liquids: R2Zn	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: R3Fe	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
aq. Liquids SO4	g/L	0.22	0.00	0.00	0.00	0.00	102.00	0.00	0.00	0.00	0.00
aq. Liquids H+	g/L	0.00	0.00	0.00	0.00	0.00	2.78	0.00	0.00	0.00	0.00
Aq. Liquids: Cu	g/L	1.06	0.00	0.00	0.00	0.00	0.22	0.00	0.00	0.00	0.00
Aq. Liquids: Ca	g/L	0.83	0.00	0.00	0.00	0.00	0.17	0.00	0.00	0.00	0.00
Aq. Liquids: Mn	g/L	5.34	0.00	0.00	0.00	0.00	1.11	0.00	0.00	0.00	0.00
Aq. Liquids: Mg	g/L	0.15	0.00	0.00	0.00	0.00	0.03	0.00	0.00	0.00	0.00
Aq. Liquids: H2O	g/L	1000.00	1000.00	1000.00	1000.00	1000.00	275.00	0.00	0.00	0.00	0.00

Table A-2: Stream Table for route 2

stream number		50	51	52	53	54	55	56
stream name		cementation products	Cementation filtrate	Cementation residue	Lime to BZS precipitation	BZS product	BZS thickener O/F	Flocculant to NT thickener
Flow	t/h	187.00	187.00	0.11	6.10	291.00	270.00	0.48
Flow	m3/h	45.50	45.50	0.02	3.80	156.00	149.00	0.00
Solids	wt %	0.00%	0.00%	99.00%	50.00%	5.00%	0.00%	0.00%
Solids	t/h	0.11	0.00	0.11	3.05	13.40	0.00	0.48
Aq. Liquids	t/h	187.00	187.00	0.00	3.05	278.00	270.00	0.00
Gases	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Org. Liquids	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Zn	g/L	100.00	100.00	1.60	0.00	6.90	7.01	0.00
Aq. Liquids: Fe	g/L	0.13	0.13	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: H2SO4	g/L	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Zn	t/h	0.00	0.00	0.00	0.00	3.57	0.00	0.00
Solids: Fe	t/h	0.00	0.00	0.00	0.00	0.01	0.00	0.00
Solids: Cu	t/h	0.11	0.00	0.11	0.00	0.00	0.00	0.00
Solids: Ca	t/h	0.00	0.00	0.00	1.65	1.65	0.00	0.00
Solids: Mn	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: Mg	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids: H2O	t/h	0.00	0.00	0.00	0.00	0.08	0.00	0.00
Solids: SO4	t/h	0.00	0.00	0.00	0.00	5.28	0.00	0.00
Aq. Liquids: Zn	t/h	4.56	4.56	0.00	0.00	1.08	1.05	0.00
Aq. Liquids: Fe	t/h	0.01	0.01	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Cu	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Ca	t/h	0.08	0.08	0.00	0.00	0.09	0.09	0.00
Aq. Liquids: Mn	t/h	0.53	0.53	0.00	0.00	0.54	0.53	0.00
Aq. Liquids: Mg	t/h	0.02	0.02	0.00	0.00	0.02	0.01	0.00
Aq. Liquids: H2O	t/h	131.00	131.00	0.00	3.05	230.00	223.00	0.00
Aq. Liquids: H2SO4	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids SO4	t/h	48.90	48.90	0.00	0.00	44.60	43.30	0.00
Aq. Liquids H+	t/h	1.33	1.33	0.00	0.00	1.35	1.31	0.00
Aq. Liquids: R2Zn	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: R3Fe	t/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00
aq. Liquids SO4	g/L	1070.00	1080.00	17.20	0.00	286.00	290.00	0.00
aq. Liquids H+	g/L	29.20	29.20	0.47	0.00	8.67	8.81	0.00
Aq. Liquids: Cu	g/L	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Aq. Liquids: Ca	g/L	1.82	1.82	0.03	0.00	0.58	0.59	0.00
Aq. Liquids: Mn	g/L	11.70	11.70	0.19	0.00	3.47	3.52	0.00
Aq. Liquids: Mg	g/L	0.33	0.33	0.01	0.00	0.10	0.10	0.00
Aq. Liquids: H2O	g/L	2890.00	2890.00	46.20	803.00	1480.00	1500.00	0.00

# **Appendix B**

## Mineralogy of Gamsberg ore

**Table B-1: Major elements (wt%) of the ore in the different class sizes (A, B, C) before and after leaching. A (+5.25/-6.75 mm), B (+14/-16 mm), C (+23/-25 mm) (Ghorbani, 2012)**

Chemical assay of the leach reactors feed (wt %)	Chemical assay of residue (wt %)					
	Leach reactors			Leach reactors		
	A	B	C	A	B	C
<b>SiO<sub>2</sub></b>	34.88	35.23	36.07	55.42	46.24	42.91
<b>TiO<sub>2</sub></b>	0.3	0.3	0.3	0.53	0.4	0.4
<b>Al<sub>2</sub>O<sub>3</sub></b>	7.15	7.06	7.04	14.08	11.35	12.11
<b>Fe<sub>2</sub>O<sub>3r</sub></b>	2.81	0.16	1.83	7.27	7.69	7.27
<b>MnO</b>	0.55	0.46	0.51	0.16	0.2	0.21
<b>MgO</b>	0.62	0.6	0.54	0.5	0.4	0.43
<b>CaO</b>	0.3	0.27	0.21	0.03	0.05	0.16
<b>Na<sub>2</sub>O</b>	0	0	0	0.25	0.55	0.87
<b>K<sub>2</sub>O</b>	0.85	0.89	0.8	1.93	1.53	1.31
<b>P<sub>2</sub>O<sub>5</sub></b>	0.22	0.22	0.17	0.06	0.07	0.15
<b>Cr<sub>2</sub>O<sub>3</sub></b>	0.04	0.05	0.05	0.04	0.05	0.03
<b>NiO</b>	0	0	0	0	0	0
<b>S</b>	21.17	22.41	21.03	9.39	15.1	16.08
<b>Zn</b>	11.9	12.36	13.53	2.16	4.19	5.87
<b>Pb</b>	0.57	0.35	0.22	0.49	0.43	0.15
<b>Cu</b>	0.01	0.01	0.01	0.01	0.02	0.02
<b>Fe-sul</b>	18.25	19.5	17.17	7.18	11.29	11.46
<b>Total</b>	<b>99.62</b>	<b>99.87</b>	<b>99.48</b>	<b>99.5</b>	<b>99.56</b>	<b>99.43</b>

**Table B-2: Compositions of concentrates (Kohler, personal communication 2014)**

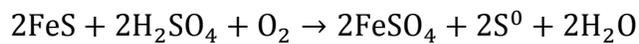
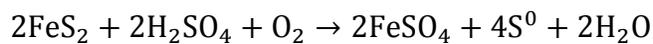
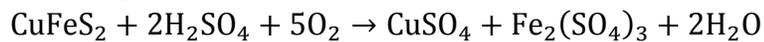
Metal	Composition (%)
<b>Zn</b>	45.9%
<b>Cu</b>	0.35%
<b>Cd</b>	0.09%
<b>Pb</b>	2.41%
<b>Fe</b>	9.07%
<b>Co</b>	0.01%
<b>Mn</b>	1.76%
<b>Mg</b>	0.05%
<b>S</b>	29.3%
<b>Ca</b>	0.19%
<b>SiO<sub>2</sub></b>	4.73%

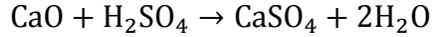
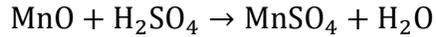
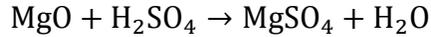
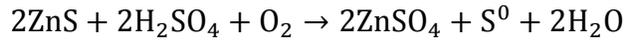
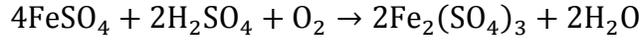
**Table B-3: Extraction rates for metals in concentrates (Fuls, 2011)**

Metal	Extraction (%)
<b>Fe</b>	98%
<b>Cu</b>	85%
<b>Ca</b>	98%
<b>S</b>	45%
<b>Mn</b>	85%
<b>Mg</b>	85%

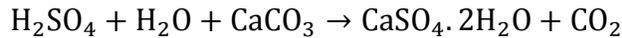
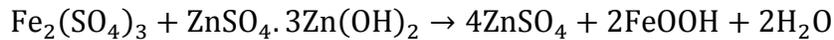
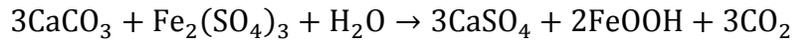
## Process Reactions

### Leaching:

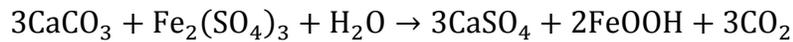




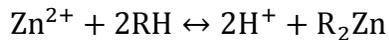
**Fe Precipitation:**



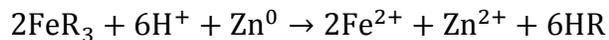
**Gypsum and Basic Zinc Precipitation:**



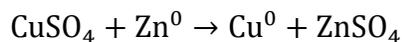
**Fe, Zn extraction and stripping:**



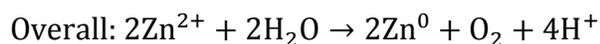
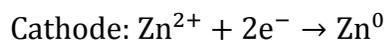
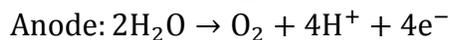
**Fe reduction:**



**Cu cementation:**



**Zn Electrowinning:**



# Appendix C

## Sample calculations

### Heap base area for route 1

Angle of repose =  $26.6^\circ$  C (2H: 1V)

$$V \text{ (m)} = h + 6$$

$$\text{Diagonal (m)} = 2 \times 2 \times (h + 6)$$

Assuming square based pyramid

$$a^2 + a^2 = [4 \times (h + 6)]^2$$

$$a^2 = 8 \times (h + 6)^2$$

$$b^2 = 8 \times h^2$$

$$\text{Volume of pyramid} = \frac{1}{3} \times l \times w \times h$$

$$V_a = \text{big pyramid} = \frac{8}{3} \times (6 + h)^3$$

$$V_b = \text{small pyramid} = \frac{8}{3} \times (h)^3$$

$$V_{\text{heap}} = V_a - V_b$$

$$\frac{m}{\rho} = 3.92 \text{ t/m}^3 = \frac{8}{3} \times (6 + h)^3 - \frac{8}{3} \times (h)^3$$

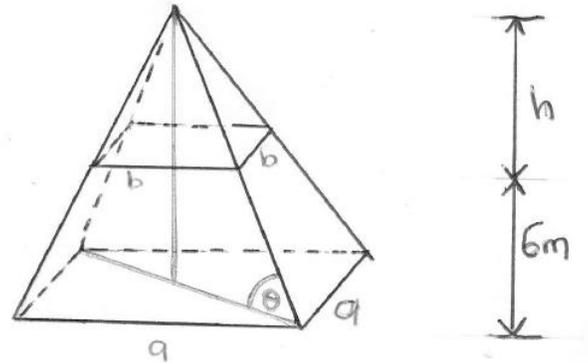
Solve iteratively and find value for h or use the expansion method for cubic functions and then solve the quadratic function in order to calculate h.

$$h = 131 \text{ m}$$

$$a^2 = 151,151 \text{ m}^2$$

$$b^2 = 138,244 \text{ m}^2$$

For checking purposes, volume of a heap is equal to the volume of a frustum of a pyramid.



$$V_{\text{fustum of pyramid}} = \frac{h}{3} \times (a^2 + b^2 + \sqrt{a^2b^2})$$

Table C-1 displays the installed cost per square meter of pad surface for route one. Table C-2 displays the installed cost per square meter of solution irrigation lines and air injection pipelines. Since the foot print of the heap has been calculated above, the total leach pad costs can thus be establish.

**Table C-1: Leach pad – double composite liner costs (Kappes, 2002)**

Construction element	units	Cost
<b>Heap area</b>	m <sup>2</sup>	151,000
<b>Preliminary earthworks</b>	\$/m <sup>2</sup>	1
<b>Underliner</b>	\$/m <sup>2</sup>	1.5
<b>Lower geomembrane</b>	\$/m <sup>2</sup>	3
<b>Leak detection</b>	\$/m <sup>2</sup>	0.5
<b>Upper geomembrane</b>	\$/m <sup>2</sup>	1.5
<b>Drain pipes</b>	\$/m <sup>2</sup>	0.5
<b>Drainage protection layer</b>	\$/m <sup>2</sup>	0.5
<b>Total pad cost</b>	\$/m <sup>2</sup>	8.5
<b>Total cost</b>	\$	1,280,000

**Table C-2: Solution irrigation lines and air injection pipelines costs (Kappes, 2002)**

Construction element	Units	Costs
<b>Heap area</b>	m <sup>2</sup>	151,000
<b>Irrigation line cost</b>	\$/m <sup>2</sup>	0.6
<b>Air injection pipe cost</b>	\$/m <sup>2</sup>	0.5
<b>Total cost</b>	\$	166,000

The six tenth rule adapted from Cohen et al., (2009) has been used to calculate the cost of stacking ore at Gamsberg. The results are shown in Table C-3.

**Table C-3: Conveyor cost for stacking ore (Kappes, 2002)**

	Units	Costs
<b>S<sub>1</sub></b>	t ore/d	10000
<b>S<sub>2</sub></b>	t ore/d	10,800
<b>C<sub>1</sub></b>	\$	1,500,000
<b>C<sub>2</sub></b>	\$	1,570,000

$$C_2 = C_1 \times \left(\frac{S_2}{S_1}\right)^{0.6}$$

C<sub>1</sub>: Capital cost of item 1

C<sub>2</sub>: Capital cost of item 2

S<sub>1</sub>: Capacity of item 1

S<sub>2</sub>: Capacity of item 2

Table C-4 displays the total heap costs based on the figures presented in Table C-1 to Table C-3. The difference in cost indexes (C.I.) has been used to calculate the current heap leaching cost. This method has been adapted from and Cohen et al., (2009).

**Table C-4: Total heap leaching cost**

	Units	Costs
<b>C<sub>1</sub></b>	\$	3,020,000
<b>Exchange rate</b>	R/\$	10.4
<b>C<sub>1</sub></b>	R	31,400,000
<b>C.I.<sub>1</sub>, 2002</b>		396
<b>C.I.<sub>2</sub>, 2015</b>		575
<b>C<sub>2</sub></b>	R	45,600,000

$$C_2 = C_1 \times \left(\frac{C.I._2}{C.I._1}\right)$$

C.I.<sub>1</sub>: Cost index for year 1

C.I.2: Cost index for year 2

**O'Hara method**

Table C-5 presents the different stages for a concentrator plant. For each stage, a cost equation and capacity range is given. Since the cost equations are associated with certain factors, these factors are shown in Table C-6. The costs equations are applied for Route 2 and Route 3. The costs for the concentrator plant are adapted from Mular (2002).

**Table C-5: Summary of Mineral Processing Plant Capital Cost Estimation (Mular, 2002)**

Cost item	Parameter	Graph Range	Cost Equation	Comment
<b>Clearing/ excavation</b>	T=capacity, stpd	500-7000	$C_1=86924 F_s T^{0.3}$	$F_s$ = site factor
<b>Foundation</b>	T=capacity, stpd	500-7000	$C_2=43463 F_c T^{0.5}$	$F_c$ = rock factor
<b>Crushing/ conveyors</b>	T=capacity, stpd	500-7000	$C_3=97790 T^{0.5}$	
<b>Grinding/storage</b>	T= capacity, stpd	500-7000	$C_5=17386 F_g T^{0.7}$	$F_g$ = grind factor
<b>Flotation/ processing</b>	T= capacity, stpd	500-7000	$C_6=5433 F_p T^{0.7}$	$F_p$ = processing factor
<b>Thickening/filtering</b>	T= capacity, stpd	500-7000	$C_7=10866 F_t T^{0.5}$	$F_t$ = thickening factor

**Table C-6: O'Hara Factors for Table C-5 (Mular, 2002)**

Factor	Value	Application
$F_s$ = site factor	1.0	Flat sites; less than 10 ft of overburden
	1.5	Moderate slopes; some blasting required
	2.5	Steep slopes; extensive blasting required
$F_c$ = rock factor	1.0	Solid rock for foundation support
	1.8	Gravel/sand as support
	3.5	Moist soil as support; piled foundations
$F_w$ = climate factor	1.0	Mild climate
	1.8	Cold climate
	2.5	Severe climate
$F_g$ = grind factor	1.0	Soft ores; 55%-200 mesh; work index under 12
	1.5	Medium ores; 70%-200 mesh; work index=15
	1.8	Hard ores; 80%-200 mesh;
	1.0	Au ores; cyanidation
$F_p$ = process factor	1.2	Flotation; course low grade Cu ores
	1.6	Flotation; hi grade Cu/Zn ores
	2.0	Selective flotation; complex base metal ores
	3.0	Complex Au ores; float, roast, cyanide
	5.0	Gravity concentration
$F_t$ = process factor	1.0	Low grade Cu ores
	1.6	Hi grade Cu/Zn ores
	2.0	Complex Pb/Zn/Ag or Cu/Zn/Pb ores
	3.0	Cyanided Au ores

Table C-7 shows a sample calculation of the flotation circuit for Route 2 and Route 3 (ore mined is the same for all the routes). This calculation uses an equation from Table C-5 and a flotation process factor from Table C-6. Cost indexes method has been adapted from Cohen (2009).

**Table C-7: Capital cost for the flotation circuit (Cohen, 2009; Mular, 2002)**

	units		Comment
<b>T</b>	tpd	10,800	T: capacity
<b>conversion</b>	stpd/tpd	1.10	
<b>T</b>	stpd	11,900	500 < T < 7000
<b>T for each stage</b>		5,950	
<b>Fp</b>		1.6	
<b>C.I.1, 1980</b>		261.8	
<b>C.I.2, 2015</b>		575	
<b>Cost</b>	\$	16,800,000	
	R	175,000,000	

$$C = 6520 \times F_p \times T^{0.5}$$

Table C-8 shows a sample calculation for costing a settler. The settler for washing/ scrubbing stage in Route 2 has been used for the results presented.

**Table C-8: Capital cost for a settler (Anderson et al., 2002)**

Settler design	units	
<b>Flow through settler</b>	m <sup>3</sup> /h	3360
<b>Settling rate</b>	m <sup>3</sup> /h/m <sup>2</sup>	4
<b>Settling area</b>	m <sup>2</sup>	840
<b>Conversion</b>	ft <sup>2</sup> /m <sup>2</sup>	10.8
<b>Settling area =S<sub>2</sub></b>	ft <sup>2</sup>	9040
<b>C.I.1, 2000</b>		394
<b>C.I.2, 2015</b>		575
<b>S<sub>1</sub></b>	ft <sup>2</sup>	4200
<b>C<sub>1</sub></b>	\$	1,080,000
<b>C<sub>2</sub></b>	\$	1,706,000
<b>Exchange rate</b>	R/\$	10.4
<b>Number of stages</b>		2
<b>Total cost</b>	R	35,500,000

$$C_2 = C_1 \times \left(\frac{S_2}{S_1}\right)^{0.6} \times \left(\frac{C. I. 2}{C. I. 1}\right)$$

The same calculation method used by Anderson et al., (2002) has been used in Table C-8. Table C-9 displays the calculation cost of a mixer which is in the extraction stage for Route 2. The equations below have been used for the calculation, they are all adapted from Seider et al., (2003).

$$D_i = \sqrt[3]{4V/\pi}$$

$$C_P = C_V + C_{PL}$$

$$C_V = \exp\{8.717 - 0.2330[\ln(W)] + 0.04333[\ln(W)]^2\}, \text{ for } 4,200 < W < 920,000 \text{ lb}$$

Where:  $C_V$  is the cost of the vessel including nozzles

$$C_{PL} = 1,580(D_i)^{0.20294}, \text{ for } 3 < D_i < 12 \text{ ft}$$

Where,

$C_{PL}$ : cost for platforms and ladders

$D_i$ : inside diameter, ft

For tank reactors, vertical vessels are used.

$$C_V = \exp\{8.717 - 0.2330[\ln(W)] + 0.04333[\ln(W)]^2\}, \text{ for } 1,000 < W < 920,000 \text{ lb}$$

$$C_{PL} = 285.1(D_i)^{0.73960}(L)^{0.70684}, \text{ for } 3 < D_i < 21 \text{ ft and } 12 < L < 40 \text{ ft}$$

**Table C-9: Capital cost for a mixer (Seider et al., 2003)**

Mixer-design	units	Value	Comment
<b>L/D</b>		1	
<b>Residence time</b>	min	3	
<b>Flow rate</b>	m <sup>3</sup> /min	56.0	
<b>Number of mixers</b>		5	
<b>Flow rate through each mixer</b>	m <sup>3</sup> /min	14	
<b>Volume of each mixer</b>	m <sup>3</sup>	42	
<b>D</b>	m	3.50	
<b>Conversion</b>	ft/m	3.28	
<b>D</b>	ft	11.5	3 < D <sub>i</sub> < 12 ft
<b>L</b>	ft	11.5	
<b>ts</b>	in	1.25	ts: shell thickness.
<b>Conversion</b>	ft/in	0.08	
<b>ts</b>	ft	0.1	
<b>ρ</b>	Ib/ft <sup>3</sup>	490	Carbon steel density
<b>W</b>	Ib	38300	4,200 < W < 920,000 Ib
<b>CPL</b>	\$	2,590	
<b>CV</b>	\$	65,200	
<b>CP</b>	\$	67,800	
<b>Cost of all mixers</b>	\$	339,000	
<b>Number of stages</b>		2	
<b>Total cost, 2000</b>	\$	678,000	
<b>C.I.1</b>		394	
<b>C.I.2</b>		575	
<b>Exchange rate</b>	R/\$	10.4	
<b>Total cost, 2015</b>	R	9,890,000	

Table C-10 displays the capital costs of agitators for Route 1. These agitators are used in the extraction stage. The parameters displayed in the table and the equation used are sourced from Seider et al., (2003).

**Table C-10: Capital cost for an agitator (Seider et al., 2003)**

Agitator-design	units		Comment
<b>Application</b>			
<b>Liquid-liquid extraction</b>	Hp/1,000	4	
<b>Suspension of solid particles</b>	gallons	10	
<b>Number of agitators</b>		12	
<b>S</b>	Hp/gal	0.004	
<b>V</b>	gal	9,570	
<b>S</b>	Hp	38.3	Should be less than 75
<b>C<sub>p</sub></b>	\$	18,500	
<b>Total cost</b>	\$	223,000	
<b>Total cost</b>	R	2,310,000	

$$C_p = 2590S^{0.54}$$

A sample calculation for the cost of a thickener is shown in Table C-11. The method used is from Seider et al., (2003). The results presented have been extracted from Route 1 and they are for the neutralization thickener.

**Table C-11: Capital cost for a thickener (Seider et al., 2003)**

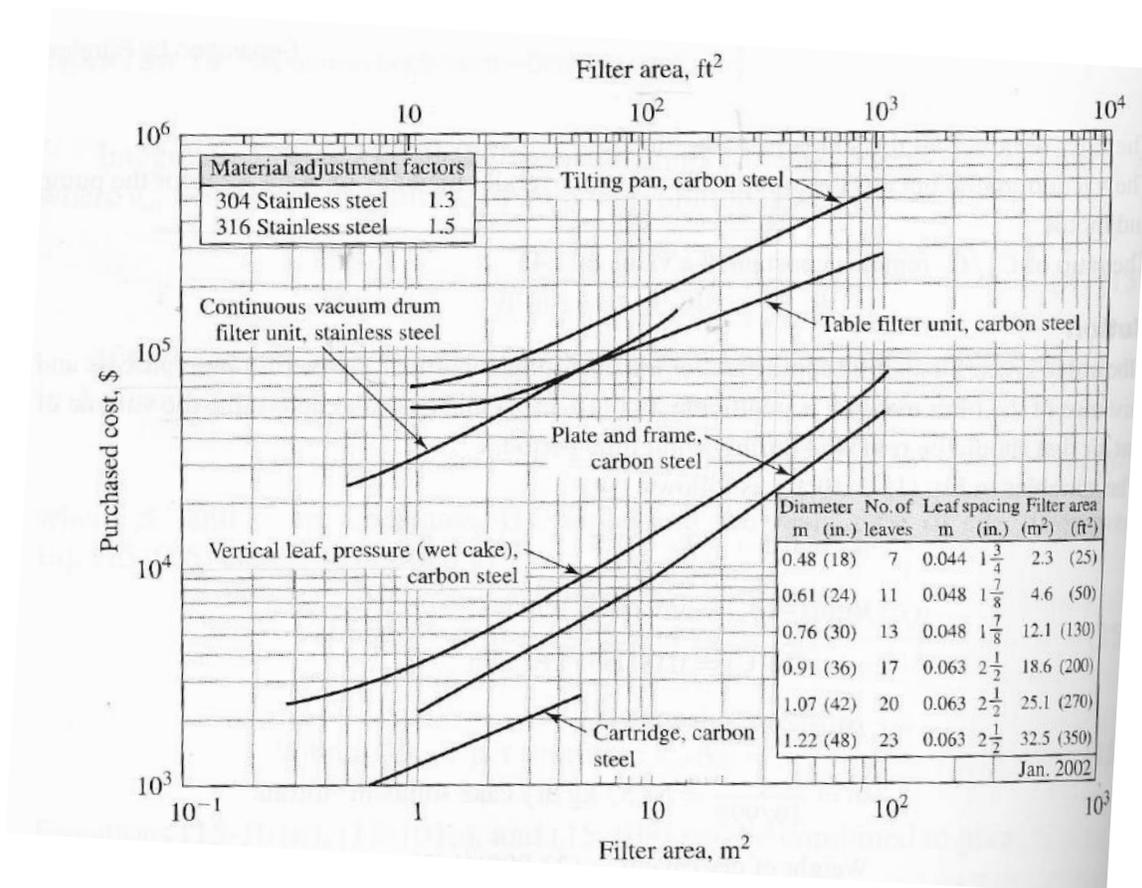
	Units	Costs
<b>A</b>	t solids/d	5836
<b>C<sub>p</sub></b>	\$	405,000
<b>C.I.1, 2000</b>		394
<b>C.I.2, 2015</b>		575
<b>Final cost</b>	R	6,150,000

$$C_p = 2650A^{0.54}$$

Table C-12 shows a sample calculation for costing a filter. The results presented are for a plate and frame filter used to remove Cu/Zn cake in Route 1. The filtrate rate has been sourced from Seider et al., (2003). Figure C-1 displays the purchase cost of the filter, the figure has been sourced from Peters et al., (2004).

**Table C-12: Capital cost for a filter (Seider et al., 2003)**

	Units	
<b>Filtrate</b>	lb/d	12,300,000
<b>Filtrate rate</b>	lb/d-ft <sup>2</sup>	1,500
<b>Filter area</b>	ft <sup>2</sup>	8,180
<b>Conversion</b>	m <sup>2</sup> /ft <sup>2</sup>	0.093
<b>Filter area</b>	m <sup>2</sup>	760
<b>Cp</b>	\$	393,000
	R	4,080,000



**Figure C-1: Purchase cost of fixed and continuous filters (Peters et al., 2004)**

An existing plant was used to cost an electrowinning and melting and casting circuits. The figures are shown in Table C-13 and Table C-14, respectively. The costs for the existing plant were sourced from Lizama et al., (2003).

**Table C-13: Capital cost for electrowinning Circuit (Lizama et al., 2003)**

	Units	
S <sub>1</sub>	t/y	120,000
S <sub>2</sub>	t/y	176,000
C <sub>1</sub>	\$	43,300,000
C <sub>2</sub>	\$	54,500,000
C.I. <sub>1</sub>		402
C.I. <sub>2</sub>		575
C <sub>2</sub>	R	811,000,000

$$C_2 = C_1 \times \left(\frac{S_2}{S_1}\right)^{0.6} \times \left(\frac{C.I._2}{C.I._1}\right)$$

**Table C-14: Capital cost for melting and casting circuit (Lizama et al., 2003)**

	Units	
S <sub>1</sub>	t/y	120,000
S <sub>2</sub>	t/y	176,000
C <sub>1</sub>	\$	14,300,000
C <sub>2</sub>	\$	18,000,000
C.I. <sub>1</sub>		402
C.I. <sub>2</sub>		575
C <sub>2</sub>	R	268,000,000

Table C-15 shows the capital expenditure (CAPEX) of each stage for the three routes. Route 1 has a zinc production of 176 kt/y, Route 2 yields 204 kt/y and Route 3 has a Zn equivalent of 216 kt/y.

**Table C-15: Capital Expenditure for the three process routes**

	Route 1 (R/ t of Zn produced)	Route 2 (R/ t of Zn produced)	Route 3 (R/ t of Zn produced)
<b>Plant-site clearance</b>	916	790	746
<b>Mining</b>	1,740	1,500	1,410
<b>Crushing</b>	1,950	1,690	1,590
<b>Milling stage</b>	0	2,560	2,420
<b>flotation stage</b>	0	932	880
<b>leaching stage</b>	258	730	0
<b>tailings</b>	0	418	394
<b>NT stage</b>	406	216	0
<b>SX stage</b>	1,520	1,270	0
<b>BT stage</b>	35.5	32.3	0
<b>EW stage</b>	6,100	5,750	0

**Fixed Capital Investment**

Table C-16 provides detailed factorial estimates used to calculate the fixed capital of each route. These estimates were adopted from Cohen, (2009). The use of these estimates is shown in Table C-17; Route 1 has been used as an example.

**Table C-16: Factors for estimation of project fixed capital cost. PCE (Plant Cost Equipment) (Cohen, 2009)**

Item	Process type		
	Fluids	Fluids-Solids	Solids
<b>Major equipment, total purchase cost</b>	PCE	PCE	PCE
<b>f1 Equipment erection</b>	0.4	0.45	0.5
<b>f2 Piping</b>	0.7	0.45	0.2
<b>f3 Instrumentation</b>	0.2	0.15	0.1
<b>f4 Electrical</b>	0.1	0.1	0.1
<b>f5 Buildings, process</b>	0.15	0.1	0.05
<b>f6 Utilities</b>	0.5	0.45	0.25
<b>f7 Storages</b>	0.15	0.20	0.25
<b>f8 Site development</b>	0.05	0.05	0.05
<b>f9 Ancillary buildings</b>	0.15	0.2	0.3
<b>Total physical plant cost (PPC)</b>			
<b>PPC = PCE (1 + f1+ ....f9)</b>			
<b>= PCE x</b>	3.40	3.15	2.80
<b>f10 Design and Engineering</b>	0.3	0.25	0.2
<b>f11 Contractor's fee</b>	0.05	0.05	0.05
<b>f12 Contingency</b>	0.1	0.1	0.1
<b>Fixed capital = PPC (1 + f10 + f11 + f12)</b>			
<b>= PPC x</b>	1.45	1.40	1.35

**Table C-17: Fixed Capital costs for Route 1 (Cohen, 2009; Peter et al., 2004; Seider et al., 2003; Anderson et al., 2002; Lizama et al., 2003; Kappes, 2002; Mular, 2002)**

Stage	Cost
<b>Plant-site clearance</b>	R 161,000,000
<b>Mining</b>	R 306,000,000
<b>Crushing</b>	R 344,000,000
<b>Heap Leaching stage</b>	R 45,500,000
<b>Neutralization stage</b>	R 71,600,000
<b>Solvent Extraction stage</b>	R 269,000,000
<b>Bleed Treatment stage</b>	R 6,260,000
<b>Electrowinning stage</b>	R 1,070,000,000
<b>PCE</b>	R 2,280,000,000
<b>x</b>	3.15
<b>PCE<sub>x</sub></b>	R 7,180,000,000
<b>x</b>	1.4
<b>Fixed Capital (PPC<sub>x</sub>)</b>	R 10,000,000,000

### **Operational cost**

The costs for operating agitators are shown in Table C-18. Solvent extraction agitators for Route 1 have been used as an example. The electricity tariff have been adapted from Eskom's schedule standard prices for 2014/2015 (Eskom, 2014).

**Table C-18: Operating costs for agitators (Eskom, 2014)**

Agitators for SX	Units	
<b>Energy drawn by equipment</b>	Hp	1,540
<b>Conversion</b>	kW/Hp	0.735
<b>Energy drawn by equipment</b>	kW	1,130
<b>Operational hours</b>	h/yr	7536
<b>Energy drawn by equipment</b>	kWh/yr	8,510,000
<b>Price</b>	R/kWh	0.863
<b>Cost</b>	R	7,340,000

Costs for operating the electrowinning circuit are given in Table C-19. The cost of power consumption per ton of zinc produced has been adapted from Nicol et al., (2008).

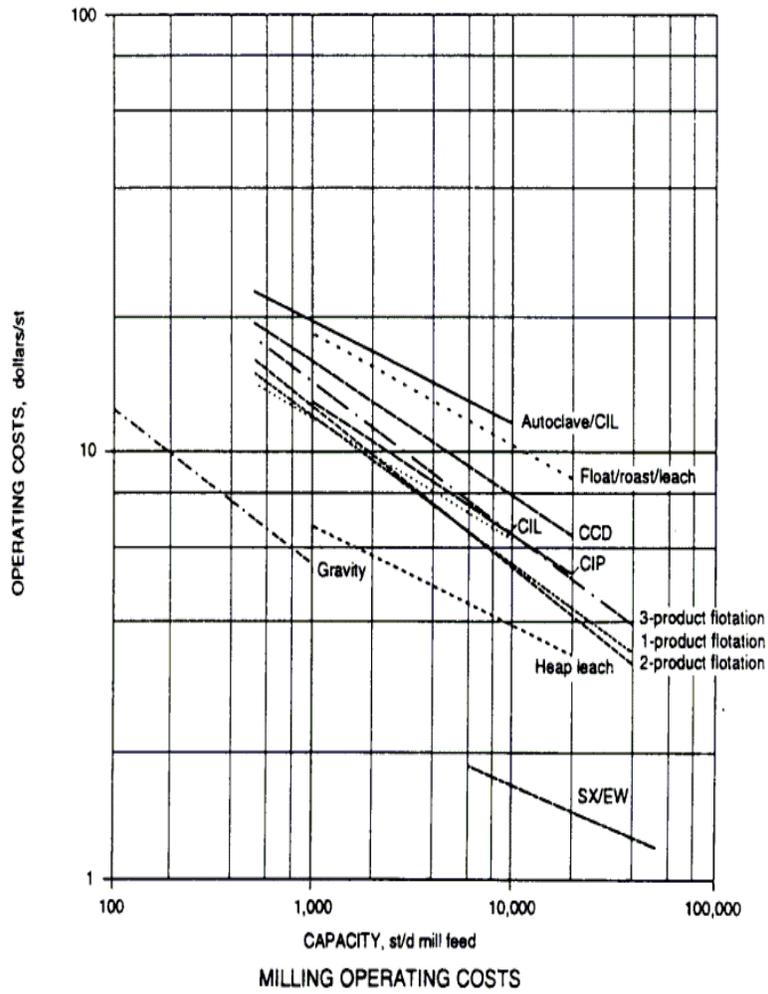
**Table C-19: Operating costs for electrowinning Stage (Nicol et al., 2008)**

EW	Units	
<b>Power consumption</b>	kWh/t Zn	3200
<b>Zinc production</b>	t Zn/ yr	176,000
<b>Power consumption</b>	kWh/y	564,000,000
<b>Cost</b>	R	486,000,000

A sample calculation for reagents costs is shown in Table C-20. Zinc dust cost from Route 1 has been used as an example. Milling costs for Route 2 and Route 3 are shown in Figure C-2. The figure has been sourced from Camm (1991).

**Table C-20: Reagents costs- Zinc dust**

Reagent	Unit	
<b>Zinc dust</b>	t/h	0.02
<b>Price</b>	R/t	4,140
<b>Cost</b>	R/hr	82.80
<b>Operational time</b>	h/y	7536
<b>Cost</b>	R	624,000



**Figure C-2: Milling Operating Costs (Camm, 1991)**

Table C-21 shows the operating expenditure (Opex) of each stage for the three routes.

**Table C-21: Operating Costs for the three routes**

	HydroZinc™ (R/ t of Zn produced)	Atmospheric Leach (R/ t of Zn produced)	Concentrates (R/ t of Zn produced)
<b>Mining</b>	2320	2000	1890
<b>Crushing</b>	775	668	631
<b>Milling stage</b>	0	1970	1860
<b>flotation stage</b>	0	601	567
<b>leaching stage</b>	883	1770	0
<b>tailings</b>	0	0	0
<b>NT stage</b>	2850	320	0
<b>SX stage</b>	46	36.1	0
<b>BT stage</b>	202	154	0
<b>EW stage</b>	4970	4970	0
<b>Maintenance</b>	1140	1400	656
<b>Shipping</b>	0	0	2610

**Revenue**

Table C-22 displays the revenue generated by Route 1 for selling zinc metal. The zinc price is at October 2014 and has been sourced from Westmetall, (2014).

Revenue for zinc metal = production × Zn grade × Zn price

**Table C-22: Revenue estimation for zinc metal (Westmetall, 2014)**

	Unit	
<b>Production</b>	t/y	176,000
<b>Grade</b>	%	100
<b>Price</b>	\$/ t	2,277
<b>Exchange rate</b>	R/\$	10.4
<b>Revenue</b>	R	4,153,000,000

Gross profit = Revenue – Expenses (Operating Cost)

Net profit = (Gross profit – Depreciation) × (1 – tax rate(%))

Cash flow = Net profit + Depreciation

Discounted Cash Flow (DCF) = Cash flow / (1 + discount rate (%))<sup>n</sup>

n is the number of year

Cash flow analysis for Route 1, Route 2 and Route 3 are shown from Table C-23 to Table 25.

**Table C-23: Route 1's Cash Flow over the plant life**

Year	Zn Production (kt/y)	Revenue (Billions)	Expenses (Billions)	Gross Profit (Billions)	Depreciation (Billions)	Net Profit (Billions)	Cash Flow (Billions)	DCF (Billions)
<b>1</b>							-R 4.40	-R 3.90
<b>2</b>							-R 4.70	-R 3.60
<b>3</b>							-R 4.90	-R 3.40
<b>4</b>	52.9	R 1.62	R 0.89	R 0.73	R 1.00	-R 0.28	R 0.73	R 0.45
<b>5</b>	52.9	R 1.73	R 0.94	R 0.78	R 1.00	-R 0.22	R 0.78	R 0.43
<b>6</b>	106	R 3.69	R 2.01	R 1.68	R 1.00	R 0.49	R 1.49	R 0.72
<b>7</b>	106	R 3.94	R 2.13	R 1.81	R 1.00	R 0.58	R 1.58	R 0.67
<b>8</b>	176	R 7.01	R 3.77	R 3.23	R 1.00	R 1.60	R 2.61	R 0.98
<b>9</b>	176	R 7.48	R 4.01	R 3.47	R 1.00	R 1.77	R 2.78	R 0.93
<b>10</b>	176	R 7.98	R 4.26	R 3.72	R 1.00	R 1.96	R 2.96	R 0.87
<b>11</b>	176	R 8.52	R 4.53	R 4.00	R 1.00	R 2.15	R 3.16	R 0.82
<b>12</b>	176	R 9.10	R 4.81	R 4.29	R 1.00	R 2.36	R 3.37	R 0.78
<b>13</b>	176	R 9.71	R 5.11	R 4.60	R 1.00	R 2.59	R 3.60	R 0.73
<b>14</b>	176	R 10.40	R 5.43	R 4.94	R 0.00	R 3.56	R 3.56	R 0.64
<b>15</b>	176	R 11.10	R 5.77	R 5.30	R 0.00	R 3.82	R 3.82	R 0.61
<b>16</b>	176	R 11.80	R 6.13	R 5.69	R 0.00	R 4.10	R 4.10	R 0.58
<b>17</b>	176	R 12.60	R 6.51	R 6.10	R 0.00	R 4.39	R 4.39	R 0.55
<b>18</b>	176	R 13.50	R 6.92	R 6.55	R 0.00	R 4.71	R 4.71	R 0.52
<b>19</b>	176	R 14.40	R 7.35	R 7.03	R 0.00	R 5.06	R 5.06	R 0.50
<b>20</b>	176	R 15.30	R 7.81	R 7.54	R 0.00	R 5.43	R 5.43	R 0.47
<b>21</b>	176	R 16.40	R 8.30	R 8.09	R 0.00	R 5.82	R 5.82	R 0.45
<b>22</b>	176	R 17.50	R 8.82	R 8.67	R 0.00	R 6.25	R 6.25	R 0.42
<b>23</b>	176	R 18.70	R 9.37	R 9.30	R 0.00	R 6.70	R 6.70	R 0.40

**Table C-24: Route 2's Cash Flow over the plant life**

Year	Zn Production (kt/y)	Revenue (Billions)	Expenses (Billions)	Gross Profit (Billions)	Depreciation (Billions)	Net Profit (Billions)	Cash Flow (Billions)	DCF (Billions)
<b>1</b>							-R 4.00	-R 3.50
<b>2</b>							-R 4.20	-R 3.30
<b>3</b>							-R 4.50	-R 3.10
<b>4</b>							-R 4.80	-R 2.90
<b>5</b>	61.3	R 2.00	R 1.15	R 0.85	R 1.43	-R 0.58	R 0.85	R 0.46
<b>6</b>	61.3	R 2.14	R 1.22	R 0.91	R 1.43	-R 0.52	R 0.91	R 0.44
<b>7</b>	123	R 4.56	R 2.60	R 1.96	R 1.43	R 0.38	R 1.81	R 0.77
<b>8</b>	123	R 4.87	R 2.77	R 2.11	R 1.43	R 0.49	R 1.92	R 0.72
<b>9</b>	204	R 8.67	R 4.90	R 3.77	R 1.43	R 1.68	R 3.11	R 1.04
<b>10</b>	204	R 9.25	R 5.20	R 4.05	R 1.43	R 1.89	R 3.32	R 0.98
<b>11</b>	204	R 9.88	R 5.53	R 4.35	R 1.43	R 2.10	R 3.53	R 0.92
<b>12</b>	204	R 10.5	R 5.87	R 4.67	R 1.43	R 2.33	R 3.76	R 0.87
<b>13</b>	204	R 11.3	R 6.24	R 5.02	R 1.43	R 2.58	R 4.01	R 0.82
<b>14</b>	204	R 12.0	R 6.63	R 5.39	R 1.43	R 2.85	R 4.28	R 0.77
<b>15</b>	204	R 12.8	R 7.04	R 5.78	R 0.00	R 4.16	R 4.16	R 0.67
<b>16</b>	204	R 13.7	R 7.49	R 6.21	R 0.00	R 4.47	R 4.47	R 0.63
<b>17</b>	204	R 14.6	R 7.95	R 6.67	R 0.00	R 4.80	R 4.80	R 0.60
<b>18</b>	204	R 15.6	R 8.45	R 7.16	R 0.00	R 5.15	R 5.15	R 0.57
<b>19</b>	204	R 16.7	R 8.98	R 7.68	R 0.00	R 5.53	R 5.53	R 0.54
<b>20</b>	204	R 17.8	R 9.54	R 8.25	R 0.00	R 5.94	R 5.94	R 0.52
<b>21</b>	204	R 19.0	R 10.1	R 8.85	R 0.00	R 6.37	R 6.37	R 0.49
<b>22</b>	204	R 20.3	R 10.8	R 9.50	R 0.00	R 6.84	R 6.84	R 0.47
<b>23</b>	204	R 21.6	R 11.4	R 10.2	R 0.00	R 7.34	R 7.34	R 0.44
<b>24</b>	204	R 23.1	R 12.2	R 10.9	R 0.00	R 7.88	R 7.88	R 0.42

**Table C-23: Route 3's Cash Flow over the plant life**

Year	Zn Production (kt/y)	Revenue (Billions)	Expenses (Billions)	Gross Profit (Billions)	Depreciation (Billions)	Net Profit (Billions)	Cash Flow (Billions)	DCF (Billions)
<b>1</b>							-R 2.70	-R 2.40
<b>2</b>							-R 2.80	-R 2.20
<b>3</b>							-R 3.00	-R 2.10
<b>4</b>	65	R 1.12	R 0.68	R 0.44	R 0.71	-R 0.27	R 0.44	R 0.27
<b>5</b>	65	R 1.20	R 0.72	R 0.48	R 0.71	-R 0.23	R 0.48	R 0.26
<b>6</b>	130	R 2.56	R 1.53	R 1.02	R 0.71	R 0.23	R 0.94	R 0.45
<b>7</b>	130	R 2.73	R 1.63	R 1.10	R 0.71	R 0.28	R 0.99	R 0.42
<b>8</b>	216	R 4.86	R 2.89	R 1.97	R 0.71	R 0.91	R 1.62	R 0.61
<b>9</b>	216	R 5.19	R 3.07	R 2.12	R 0.71	R 1.02	R 1.73	R 0.58
<b>10</b>	216	R 5.54	R 3.26	R 2.28	R 0.71	R 1.13	R 1.84	R 0.54
<b>11</b>	216	R 5.91	R 3.46	R 2.45	R 0.71	R 1.25	R 1.96	R 0.51
<b>12</b>	216	R 6.31	R 3.68	R 2.63	R 0.71	R 1.39	R 2.10	R 0.48
<b>13</b>	216	R 6.74	R 3.91	R 2.83	R 0.71	R 1.53	R 2.24	R 0.46
<b>14</b>	216	R 7.20	R 4.15	R 3.04	R 0.00	R 2.19	R 2.19	R 0.40
<b>15</b>	216	R 7.68	R 4.41	R 3.27	R 0.00	R 2.35	R 2.35	R 0.38
<b>16</b>	216	R 8.20	R 4.69	R 3.51	R 0.00	R 2.53	R 2.53	R 0.36
<b>17</b>	216	R 8.76	R 4.98	R 3.77	R 0.00	R 2.72	R 2.72	R 0.34
<b>18</b>	216	R 9.35	R 5.29	R 4.05	R 0.00	R 2.92	R 2.92	R 0.32
<b>19</b>	216	R 9.98	R 5.62	R 4.35	R 0.00	R 3.13	R 3.13	R 0.31
<b>20</b>	216	R 10.70	R 5.98	R 4.68	R 0.00	R 3.37	R 3.37	R 0.29
<b>21</b>	216	R 11.40	R 6.35	R 5.02	R 0.00	R 3.62	R 3.62	R 0.28
<b>22</b>	216	R 12.10	R 6.75	R 5.39	R 0.00	R 3.88	R 3.88	R 0.26
<b>23</b>	216	R 13.00	R 7.17	R 5.79	R 0.00	R 4.17	R 4.17	R 0.25

## Sensitivity analysis- Zinc Price

### Route 1

function [IRRs,ROIs,NPVs,PBPs] = OP1

percentages = -0.2:0.01:0.6;

leng=length(percentages);

ZP =2276.83\*(ones(1,leng)+percentages );

IRRs = zeros(1,leng);

ROIs = zeros(1,leng);

NPVs = zeros(1,leng);

PBPs = zeros(1,leng);

PVRs = zeros(1,leng);

P=percentages\*100;

for ii=1:leng

x=3401104.04327373; %ore processed per year

X=x\*2; % ore mined per year. Assuming strip ratio of 1:1 for ore and waste

zr=0.725635274646744; % zinc extraction rate in %

rr=0.965; % solution to metal recovery rate

oe=zr\*rr; %overall zinc recovery from ore

g= 0.074;% ore grade

ep=0.8623; % electricity price

zp= ZP(ii);%zinc price

pr=floor(x\*g\*oe); % production rate per year

n=20;

zpe=6.75456278032081; % zinc price escalation in %

e=6.25; % expenses escalation rate in %

exr = 10.3501; % ex rate

mng=60; % mining costs (R/t)

tr=28; % tax rate in %

dr=13; % discount rate in %

FC=9985265100.83306; % fixed capital cost

RC=1009799164.29905; % Reagents  
 EC=568369670.802325; % Electricity cost  
 MT=0.02\*FC;% maintenance cost  
 M=mng\*X; % mining cost  
 crush=136506264.691698; % crushing cost  
 OC=EC+RC+MT+M+crush; % operating cost  
 WC=OC;% working capital  
 TCI=FC+WC; % total capital investment  
 d=FC/10; % depreciation  
 rev=exr \* pr\*zp; %revenue per year  
 b= 3; %building time,years  
 a=TCI/b;%amount spent building each year

Years = 1:n+b;  
 Revenue = [zeros(1,b) rev\*0.3\*(1+zpe/100).^(b+1:b+2) rev\*0.6\*(1+zpe/100).^(b+3:b+4)  
 rev\*(1+zpe/100).^(b+5:n+b)];  
 Expenses = [zeros(1,b) OC\*0.3\*(1+e/100).^(b+1:b+2) OC\*0.6\*(1+e/100).^(b+3:b+4)  
 OC\*(1+e/100).^(b+5:n+b)];  
 GrossProfit = Revenue - Expenses;  
 Depreciation = d\*[zeros(1,b) ones(1,10) zeros(1,n-10)] ;  
 NetProfit = (GrossProfit - Depreciation)\*(1-tr/100);  
 Cashflow = NetProfit+Depreciation-a\*(1+e/100).^(1:n+b).\*[ones(1,b) zeros(1,n)];  
 DCF = zeros(1,n+b);  
 for k=1:n+b  
 DCF(k) = Cashflow(k).\*((1+dr/100).^(-k));  
 end

IRR = irr(Cashflow);  
 ROI = mean(NetProfit(b:end))/TCI;  
 NPV = sum(DCF);  
 PBP = FC/mean(Cashflow(b:end));

```
PVR=-sum((DCF>0).*DCF)/sum((DCF<0).*DCF);
table = [Years' Revenue' Expenses' GrossProfit' Depreciation' NetProfit' Cashflow' DCF];
```

```
IRRs(ii) = IRR*100;
```

```
ROIs(ii) = ROI;
```

```
NPVs(ii) = NPV;
```

```
PBPs(ii) = PBP;
```

```
PVRs(ii) = PVR;
```

```
end
```

### **Route 2**

```
function [IRRs,ROIs,NPVs,PBPs] = OP2
```

```
percentages = -0.2:0.01:0.6;
```

```
leng=length(percentages);
```

```
ZP =2276.83*(ones(1,leng)+percentages );
```

```
IRRs = zeros(1,leng);
```

```
ROIs = zeros(1,leng);
```

```
NPVs = zeros(1,leng);
```

```
PBPs = zeros(1,leng);
```

```
PVRs = zeros(1,leng);
```

```
P=percentages*100;
```

```
for ii=1:leng
```

```
x=3401104.04327373; %ore processed per year
```

```
X=x*2; % ore mined per year. Assuming strip ratio of 1:1 for ore and waste
```

```
CR=0.138649237472767; % Concentrate recovery
```

```
CF=CR*x; % concentrate produced per year
```

```
zr=0.977733659818771; % zinc concentrate extraction rate in %
```

```
rr=0.965; % solution to metal recovery rate
```

```
oe=zr*rr; %overall zinc recovery from ore
```

$g = 0.459$ ; % concentrate grade  
 $ep = 0.8623$ ; % electricity price  
 $zp = ZP(ii)$ ; % zinc price  
 $pr = \text{floor}(CF * g * oe)$ ; % production rate per year  
 $n = 20$ ;  
 $zpe = 6.75456278032081$ ; % zinc price escalation in %  
 $e = 6.25$ ; % expenses escalation rate in %  
 $exr = 10.3501$ ; % ex rate  
 $mng = 60$ ; % mining costs (R/t)  
 $tr = 28$ ; % tax rate in %  
 $dr = 13$ ; % discount rate in %  
 $FC = 14170383621.7002$ ; % fixed capital cost  
 $RC = 590770208.186378$ ; % downstream processing reagents  
 $EC = 1370720666.10009$ ; % energy cost  
 $flot = 122709490.050763$ ; % flotation cost  
 $MT = 0.02 * FC$ ; % maintenance cost  
 $M = mng * X$ ; % mining cost  
 $OC = RC + EC + M + MT + flot$ ; % operating cost  
 $WC = OC * 0.25$ ; % working capital  
 $TCI = FC + WC$ ; % total capital investment  
 $d = FC / 10$ ; % depreciation  
 $rev = exr * pr * zp$ ; % revenue per year  
 $b = 4$ ; % building time, years  
 $a = TCI / b$ ; % amount spent building each year

Years = 1:n+b;

Revenue = [zeros(1,b)  $rev * 0.3 * (1 + zpe/100)^{(b+1:b+2)}$   $rev * 0.6 * (1 + zpe/100)^{(b+3:b+4)}$   $rev * (1 + zpe/100)^{(b+5:n+b)}$ ];

Expenses = [zeros(1,b)  $OC * 0.3 * (1 + e/100)^{(b+1:b+2)}$   $OC * 0.6 * (1 + e/100)^{(b+3:b+4)}$   $OC * (1 + e/100)^{(b+5:n+b)}$ ];

GrossProfit = Revenue - Expenses;

```

Depreciation = d*[zeros(1,b) ones(1,10) zeros(1,n-10)] ;
NetProfit = (GrossProfit - Depreciation)*(1-tr/100);
Cashflow = NetProfit+Depreciation-a*(1+e/100).^(1:n+b).*[ones(1,b) zeros(1,n)];
DCF = zeros(1,n+b);
for k=1:n+b
DCF(k) = Cashflow(k).*((1+dr/100).^(-k));
end

IRR = irr(Cashflow);
ROI = mean(NetProfit(b:end))/TCI;
NPV = sum(DCF);
PBP = FC/mean(Cashflow(b:end));
PVR=-sum((DCF>0).*DCF)/sum((DCF<0).*DCF);
table = [Years' Revenue' Expenses' GrossProfit' Depreciation' NetProfit' Cashflow' DCF'];

IRRs(ii) = IRR*100;
ROIs(ii) = ROI;
NPVs(ii) = NPV;
PBP(ii) = PBP;
PVRs(ii) = PVR;
end

```

### **Route 3**

```
function [IRRs,ROIs,NPVs,PBPs] = OP32
```

```

percentages = -0.2:0.01:0.6;
leng=length(percentages);
ZP =2276.83*(ones(1,leng)+percentages );
IRRs = zeros(1,leng);
ROIs = zeros(1,leng);
NPVs = zeros(1,leng);

```

```

PBPs = zeros(1,leng);
PVRs = zeros(1,leng);
P=percentages*100;

for ii=1:leng
x=3401104.04327373; %ore processed per year
X=x*2; % ore mined per year. Assuming strip ratio of 1:1 for ore and waste
CR=0.138649237472767; % Concentrate recovery
CF=CR*x;% concentrate produced per year
g= 0.459;% concentrate grade
ep= 0.8623;% electricity price
zp= ZP(ii);% zinc price
n=20;
zpe=6.75456278032081; % zinc price escalation in %
e=6.25; % expenses escalation rate in %
exr = 10.3501; % ex rate
md= 0.08; % minimum deduction in zinc contracts
tc= 230; % treatment charge ($/t)
bp=2000;% base zinc price
ed= 0.1;% escalator/ descaltor for price participation
mng=60; % mining costs (R/t)
spng=1200; %shipping costs (R/t)

Fe=2;% Fe charge per 1% above 8%($/t)
Mn=1;% Mn charge per 0.1% above 0.5%($/t)
Fec=9.07; % iron content (%) in concentrates
Mnc=1.76; % Mn content (%) in concentrates

tr=28; % tax rate in %
dr=13; % discount rate in %
FC=7102908330.12366; % fixed capital cost

```

crush=136506264.691698; % crushing cost  
 mil=402447479.621666; %milling cost  
 flot=122709490.050763; % flotation cost  
 M=mng\*X; % mining cost  
 S=spng\*CF;% shipping cost  
 MT=0.02\*FC;% maintenance cost  
 OC=M+S+MT+mil+flot+crush; % operating cost  
 WC=OC\*0.25;% working capital  
 TCI=FC+WC; % total capital investment  
 d=FC/10; % depreciation  
 pm= (g-md)\*exr\*CF\*zp;% amount for payable metal  
 bc= tc\*CF\*exr;% amount for treatment charge  
 pp=ed\*(zp-bp)\*CF\*exr;% price participation  
 cFe=(Fec-8)\*Fe\*CF\*exr;% charge for iron content in concentrates  
 cMn=((Mnc-0.5)/0.1)\*Mn\*CF\*exr;%charge for Mn content in concentrates  
 rev=pm-bc-pp-cFe-cMn;% revenue generated from concentrate sales  
 b= 3; %building time,years  
 a=TCI/b;%amount spent building each year

Years = 1:n+b;

Revenue = [zeros(1,b) rev\*0.3\*(1+zpe/100).^(b+1:b+2) rev\*0.6\*(1+zpe/100).^(b+3:b+4) rev\*(1+zpe/100).^(b+5:n+b)];

Expenses = [zeros(1,b) OC\*0.3\*(1+e/100).^(b+1:b+2) OC\*0.6\*(1+e/100).^(b+3:b+4) OC\*(1+e/100).^(b+5:n+b)];

GrossProfit = Revenue - Expenses;

Depreciation = d\*[zeros(1,b) ones(1,10) zeros(1,n-10)] ;

NetProfit = (GrossProfit - Depreciation)\*(1-tr/100);

Cashflow = NetProfit+Depreciation-a\*(1+e/100).^(1:n+b).\*[ones(1,b) zeros(1,n)];

DCF = zeros(1,n+b);

for k=1:n+b

```
DCF(k) = Cashflow(k).*((1+dr/100).^(-k));
```

```
end
```

```
IRR = irr(Cashflow);
```

```
ROI = mean(NetProfit(b:end))/TCI;
```

```
NPV = sum(DCF);
```

```
PBP = FC/mean(Cashflow(b:end));
```

```
PVR=-sum((DCF>0).*DCF)/sum((DCF<0).*DCF);
```

```
table = [Years' Revenue' Expenses' GrossProfit' Depreciation' NetProfit' Cashflow' DCF'];
```

```
IRRs(ii) = IRR*100;
```

```
ROIs(ii) = ROI;
```

```
NPVs(ii) = NPV;
```

```
PBPs(ii) = PBP;
```

```
PVRs(ii) = PVR;
```

```
end
```

### ***Plotting of graphs***

```
clc
```

```
close all
```

```
clear all
```

```
percentages = -0.2:0.01:0.6;
```

```
leng=length(percentages);
```

```
ZP =2276.83*(ones(1,leng)+percentages );
```

```
[IRR1,ROI1,NPV1,PBP1]=OP1;
```

```
[IRR2,ROI2,NPV2,PBP2]=OP2;
```

```
[IRR3,ROI3,NPV3,PBP3]=OP32;
```

```
P=percentages*100;
```

```
figure(1)
plot(P,NPV1,'.r')
title('Effect of change in Zinc Price on NPV')
xlabel('Change in Zinc Price (%)')
ylabel('NPV (Rands)')
hold on
plot(P,NPV2,'.b')
hold on
plot(P,NPV3,'.g')
legend('Route 1','Route 2','Route 3', 'location','northwest')
hold off
```