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Cyanide use, recovery, and reuse at the DRDGOLD ERGO
Brakpan Gold Tailings Mine

Submitted by

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Executive Summary

Cyanide-containing effluent streams for gold ore leaching represents a potential avenue for cyanide recovery and reuse in gold mining, potentially improving operational profitability and environmental sustainability. The reuse or recycling of cyanide effluent was examined within the context of the DRDGOLD ERGO Brakpan gold tailings reprocessing plant operating in East Gauteng, South Africa. An operational assessment of the plant identified two cyanide-containing effluent streams suitable for testing. In addition, it also identified an available gold bearing coarse material unsuitable for leaching in the existing Carbon-in-Leach (CIL) circuit. The cyanide effluent streams of interest are from the CIL effluent and from the “Policeman Column” effluent. The CIL effluent contains 50 mg/L of free cyanide in solution at an average production of 100 000 tonnes of slurry (50% solids) per day. The policeman column effluent is a clear stream of approximately 400 mg/L free cyanide at an average production of 250 m³ per day. A literature review of current cyanide recovery methods indicated that HCN gas production was an important intermediary process step in the recovery of cyanide. HCN gas production is prohibited at the ERGO plant. The policeman column effluent provides a working process stream without need for adjustment or alteration due to its high free cyanide concentration and clarity. The reuse of this stream as process water was therefore investigated in the leaching of gold from the available ERGO plant Coarse material for revenue generation.

A representative sample was taken from the Coarse material heap for testing. XRD analysis indicated that the sample is primarily composed of silica and the Coarse material contains particle sizes predominantly +5.6 mm in diameter determined by sieve analysis. Initial leaching tests confirmed the presence of preg-robbing carbonaceous material within the Coarse material, accounting for 8.1 % of the ore as determined by Dense Media Separation (DMS) analysis

During initial leaching experiments with no agitation, it was observed that the rate of gold extraction from the coarse material exceeded the rate of gold adsorption by carbonaceous material present in the coarse material within the first 30 min of the leaching process. Adding aeration doubled the extraction. However, there was severe preg-robbing leading to a final gold extraction of 1.3 % and 2.5 % after 24 h, compared to peak gold extraction values of 2.5 % and 5.6 % at 30 min for unaerated and aerated leaching respectively. In the case of agitated leaching with both

coarse and pulverized Coarse material the severity of preg-robbing increased resulting in negligible gold extraction into the solution.

With the knowledge of the carbonaceous component of the Coarse material literature on the methods for preg-robbing mitigation were reviewed. The existing method of using competing activated carbon in CIL-like conditions was chosen for experimentation. The competing activated carbon allowed for gold extraction of 17 % to 19 % for activated carbon samples of different activity.

The column percolation leach with reused cyanide effluent experiment was tested with results showing gold extraction of 26.7% over 4 days. The absence of milling and higher extraction favored the column percolation leach process for a preliminary design and economic analysis.

The proposed column percolation leach process would operate for 18 years and produce a yearly revenue of R42.1 Million, and yearly cost of operation of R13.8 Million, with an upfront capital cost of R85.1 Million. The payback period is 6.3 years with a Return on Investment (ROI) of 15.7 %. The Net Present Value (NPV) at the end of the process is also positive at R1.9 Million.

The effect of reused cyanide effluent is seen in the reduction in cost of R2.1 million per year. Without the reused cyanide effluent, the payback period and ROI are negatively affected at 7.1 years and 14.3 % respectively. However, a negative value of -R4,726,805 for the NPV signifies that the investment has been negatively affected, making it unappealing for investment without reusing cyanide effluent. The application of the column percolation leach with reused cyanide effluent process at the ERGO Brakpan plant can allow for an increase in profit of 3.3 % based on ERGO Brakpan plant financial results for the financial year 1 July 2021 – 30 June 2022.

Candidates Declaration

I declare that this thesis is my own unaided work. It is submitted for the degree of Master of Science in Engineering at the University of Witwatersrand Johannesburg. It has not been submitted before for any other degree or examination at any other university.

Dedication

I would like to express appreciation and gratitude towards God, my parents, and my brother for supporting me in my pursuit of an MSc Degree in Engineering. I would also like to express gratitude towards the following individuals and organizations who have generously shared their knowledge and provided invaluable support:

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Nomenclature

Symbol	Description	Unit
AAS	Atomic Adsorption Spectroscopy	
AC	Activated Carbon	
AVR	Acidification, Volatilization and Reneutralization	
BFD	Block Flow Diagram	
CaO	Calcium oxide/Lime	
Ca(OH) ₂	Calcium hydroxide/Milk of Lime	
CIL	Carbon-in-leach Process	
CIP	Carbon-in-pulp Process	
Coarse Material	ERGO On-Site Carbonaceous Gold Material	
CN ⁻	Free Cyanide	
DMS	Dense Media Separation	
EDTA	Ethylenediaminetetraacetic acid	
EMSTM	Engineered Membrane Separation system	
ERGO Plant	The DRDGOLD ERGO Brakpan gold tailings plant	
HCN	Hydrogen Cyanide	
MAD	Medium Acid Dissociable	
mg/L	Milligram per liter	mg/L
mL/s.	Milliliters per second	mL/s
MNR	Metallgesellschaft Natural Resources	
NaCN	Sodium Cyanide	
NaOH	Sodium Hydroxide	
NPV	Net Present Value	
PFD	Process Flow Diagram	
ppm	Parts per Million	g/t
ROI	Return on Investment	
SAD	Strong Acid Dissociable	
SART	Sulfidization, acidification, recycling, and thickening	
SG	Specific Gravity	
WAD	Weak Acid Dissociable	
XRD	X-ray Diffraction	

Chapter 1: Introduction

Gold is an important and rare metal that has been in use as currency, in technology, and in cultural traditions for thousands of years (Kirkemo, et al., 1998). Mining techniques have changed over the past thousand years as native or easily extracted gold from alluvial places in rivers and surface mining has become rarer. However, gold mining techniques developed over the past century, such as the Carbon-in-leach (CIL) and Carbon-in-pulp (CIP) has allowed for the reprocessing of tailings from previous gold mines. The CIL and CIP process use activated carbon to adsorb gold from solution in the cyanidation process. These processes are remarkably similar with the only difference coming from the single-stage simultaneous leaching and adsorption of gold in the CIL process, rather than two separate stages for leaching and adsorption practiced in the CIP Process (Stange, 1999). Each process uses an excess of cyanide to enhance the reaction of the cyanide with gold in the slurry and improve the extraction percentage of gold. DRDGOLD stands as a prominent leader in the reprocessing of gold tailings through the CIL process at its Brakpan ERGO plant and has considerable interest in the sustainable reuse of cyanide.

1.1. Background on DRDGOLD ERGO Mining Proprietary Limited

DRDGOLD, a South African gold producer, is a global leader in the recovery of gold from the retreatment of surface tailings. Its subsidiary, ERGO Mining, specializes in reprocessing gold mine tailings at multiple sites spanning 163 km across South Africa's Witwatersrand region. The ERGO operations include the flagship Brakpan plant and 2 former plants, City Deep and Knights, which have been commissioned as pump and milling stations. In addition to its ERGO Mining operations, DRDGOLD expanded its tailings reprocessing capabilities in 2018 with the acquisition of Far West Gold Recoveries (FWGR) from Sibanye-Stillwater. Previously known as the West Rand Tailings Retreatment Project, FWGR was acquired by DRDGOLD in late July 2018. While DRDGOLD continues to grow its reprocessing operations, this thesis focuses specifically on the activities and techniques utilised at DRDGOLD's ERGO Plant in Brakpan, South Africa. The ERGO plant is equipped with the capability to reprocess gold tailings material at a rate of up to 60 000 tonnes per day and therefore requires substantial cyanide use. This cyanide is largely consumed by side reactions of the cyanide with base metals and other contaminants during leaching unreacted free cyanide through tailings. At the ERGO plant cyanide consumption has

been identified as a possible area of research by investigating the ways in which cyanide consumption can be reduced or free unreacted cyanide can be utilised economically.

1.1.1. The ERGO plant Carbon-in-Leach Process Description

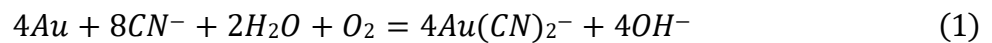
Figure 1-1 gives a block flow diagram (BFD) of the ERGO plant. The diagram was adapted from a scanned process flow diagram (PFD) of the ERGO plant provided by ERGO operations (Appendix A, Figure A-1).

Plant Feed Description

The ERGO plant is fed by four lines from different fine sand tailings dams across the southeastern Gauteng area as well as a feed of coarse sand feed that requires milling. The fine feed lines pass through the old CIP circuit (no longer in use) to distribution boxes found at the feed of the 4 thickeners, 3 of which are active simultaneously. The coarse feed is first screened by 1 mm linear screens before entering the mill. The oversize +1 mm material is heaped on site and the -1mm material is milled to -75 µm and sent to the thickener distribution box mixing with the fine feed. The thickeners are used to adjust the density of the slurry to a specific gravity (SG) of 1.45, while thickener overflow is reused in the plant as it consists of clear process water. The thickened slurry is sent to the desanding circuit where sharp material any other oversized material is removed with four 800 µm linear screens.

Carbon in Leach Circuit

The ERGO plant CIL circuit has two streams, A and B. Each stream consists of eleven 2000 m³ tanks. Calcium oxide (CaO) is added to the first tank of each stream to bring the pH up to 10.5. In both streams, oxygen is added in the second tank along with cyanide. The cyanide concentration in the second tank is set at 250 mg/L to react with the gold in the slurry, over the slurry residence time of 7 hours, via the Elsner Equation (Eq. 1).



Carbon, regenerated and fresh, is fed into the last tank (tank eleven) in each stream. The carbon is migrated upstream from tank eleven to tank three, counter-current to the flow of the slurry. A cylindrical 850 µm screen in each tank is used to retain carbon so that carbon can be pumped from tank to tank. The gold loaded carbon is removed from tank three by pumping it to a carbon hopper.

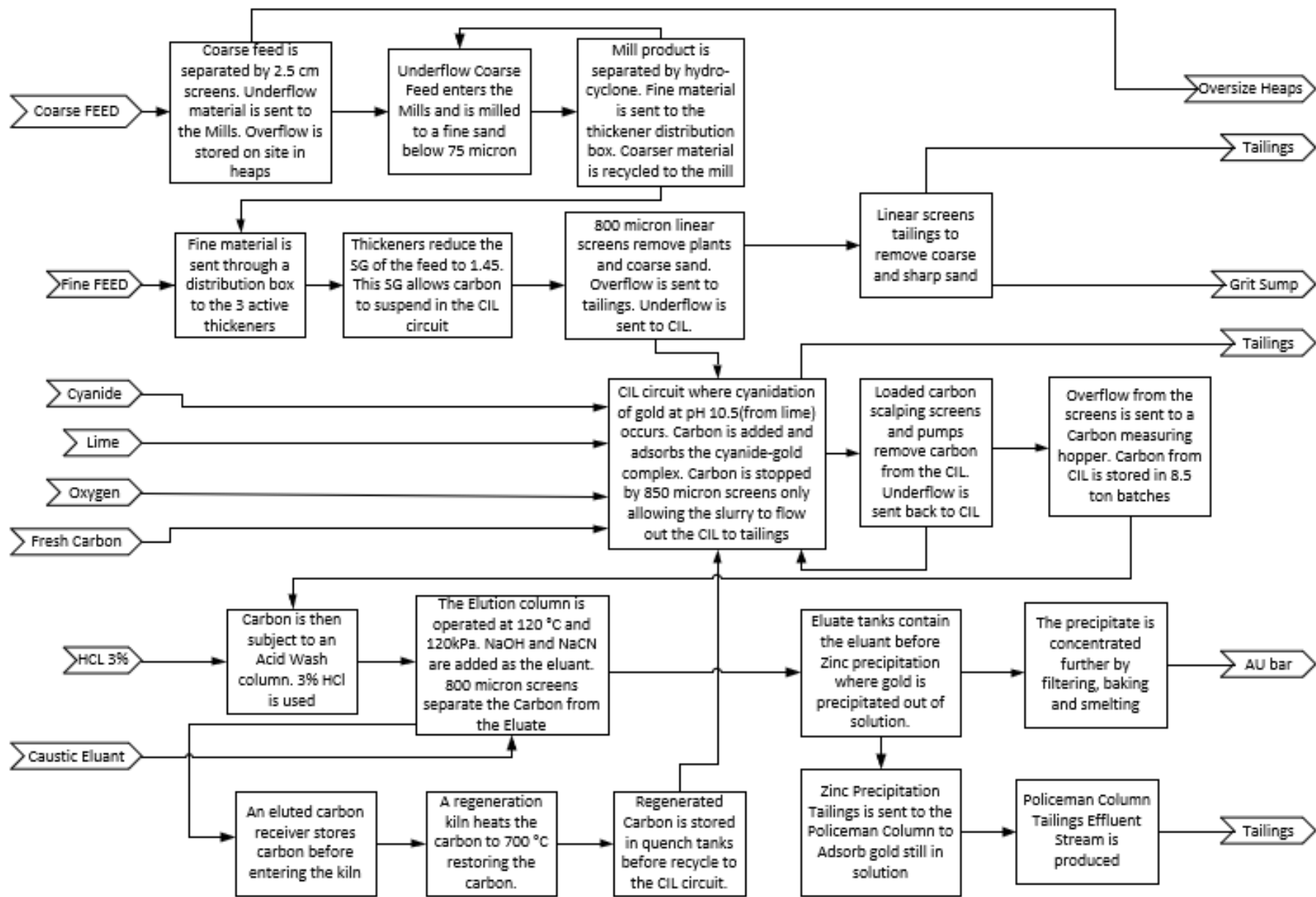


Figure 1-1: Block Flow Diagram of the ERGO plant Gold Extraction Process

Acid Wash, Elution and Zinc Precipitation

The loaded carbon collected in 8.5-ton batches and then subject to acid washing in 3 acid wash columns using a 3% hydrochloric acid solution to remove foulants such as calcium that may have built up on the surface of the carbon. The washed carbon is then sent to 3 elution tanks where the gold is eluted at 120 °C and 120 kPa in a solution constituting of 1% NaCN and 2% NaOH to form a gold rich eluate solution. The eluate is sent for zinc precipitation while the carbon is sent for regeneration by a carbon regeneration kiln operating at 800 °C before being recycled. The precipitate is filtered out of the solution and is baked and then smelted to form gold bullion bars which are sold to Rand Refinery and the waste zinc precipitation solution sent to the Policeman Column.

The subsequent section will delve into the financial aspects associated with the ERGO plant gold reprocessing plant.

1.1.2. ERGO Brakpan Economic Review and Reagent Costs

The recycling of any mass, heat, or momentum in a process is only done when the value of the item recycled outweighs the cost of inserting and running a recycling operation on the item. Therefore, a review of the costs, incomes, assets, and liabilities for the six months of July - December 2021 of the ERGO Brakpan plant will be used to understand the current operation and profitability of the plant. In Table 1-1 a recreation of ERGO Brakpan financial results from the CONDENSED CONSOLIDATED UNAUDITED INTERIM RESULTS for the period July to December 2021 has been created (excerpt found in Figure A-2, Appendix A).

Table 1-1: DRD ERGO Plant: Revenue, Cost and Operating Profit (July - December 2022)

Operating Segments	ZAR (Millions)
Revenue	1 804.7
Cash Operating Cost	1 471.8
Operating profit	352.9

The amount of interest for this study is the cash operating cost of the plant. As described in section 1.1.1 the current practice of cyanide usage at the ERGO Plant is to use fresh cyanide in the process and disregard any excess cyanide through the tailings. The total consumption and cost of each reagent used at the ERGO Plant is given in Table 1-2 for the period 1 July to 31 December 2021.

Reagents consumed in the process includes Sodium Cyanide, Oxygen, Calcium Hydroxide, Sodium Hydroxide, Hydrochloric acid, Flocculant and Activated Carbon (Raw data: Appendix A, Table A-1).

Table 1-2: Reagent consumption and cost at ERGO Plant for the period 1 July – 31 December 2021

Reagent	Consumption (tonnes)	Cost per Tonne (ZAR/tonne)	Total Reagent Cost (ZAR)
Cyanide	2 482	26 000	64 532 000
Oxygen	7 130	7 500	53 475 000
Calcium Hydroxide	16 738	1 600	26 780 800
Sodium Hydroxide	850	350	297 500
Hydrochloric acid	1 476	477	704 502
Flocculant	72	6 740	485 954
Activated Carbon	312	3 665	11 434 800

From Table 1-2 one can see that the costs related to cyanide for the months of June to December 2021 was R26 per kilogram and the total cost of cyanide for the period amounted to R64.5 million. Consequently, the cost of cyanide made up 4.18% of the entire plant's operating cost of R1.471 billion and made up 41% of all reagent costs.

A reduction of cyanide usage through recycling or reuse will therefore directly influence the cash operating expenses and a reduction in fresh cyanide usage can potentially reduce reagent costs. Conversely, reuse or recycling of cyanide for use in an independent process for higher plant gold extraction will positively influence revenue.

1.1.3. Identification of Cyanide Containing Streams for Potential Cyanide Reduction or Reuse at the ERGO Plant

The use of cyanide mainly occurs in the CIL and elution circuit. Cyanide enters the CIL circuit using a set point concentration of CN^- of 250 mg/L based on solid material tonnage. Approximately 13 tonnes of cyanide are used in the CIL circuit per day, considering both tank stream A and B, with the specific quantity varying based on the ore throughput. The gold grade of the solid material is 0.34 ppm which equates to 17 kg of extractable gold per day at an average daily tonnage of 50 000. The solid material at the exit of the CIL still has a gold grade of 0.17 ppm. This represents an

extraction of only 50% of the gold. According to the Elsner reaction equation (Eq. 1) for 17 kg of gold to bond with cyanide, only 8.5 kg of sodium cyanide is needed for all the gold entering the CIL circuit to be bonded with cyanide. This effectively means that a large amount of cyanide is added in excess.

Excess cyanide in the CIL circuit is lost in large part to side reactions that occur with other metals present in the gold ore, and free unreacted cyanide leaving the plant. At the ERGO plant the largest loss is due to the formation of other metal complexes such as copper and iron in the material entering the plant. Table 1-3 presents common cyanide metal complexes which cause the greatest losses of cyanide within a gold processing plant (Vapur, et al., 2005; Hoecker & Muir, 1987).

Table 1-3: Compounds of Cyanide Commonly Found in the Leach Solution of Gold Tailings

Classification	Cyanide Compounds
Free Cyanide	CN^-
Simple Compounds	$\text{Zn}(\text{CN})_2$, CuCN , CuCN_2 ,
WAD complex	$\text{Zn}(\text{CN})_4^{2-}$, $\text{Zn}(\text{CN})_2(\text{OH})_2$
MAD complex	$\text{Cu}(\text{CN})_2^{-1}$, $\text{Cu}(\text{CN})_3^{-2}$, $\text{Cu}(\text{CN})_4^{-3}$, $\text{Ni}(\text{CN})_4^{-2}$, $\text{Ag}(\text{CN})_2$, $\text{Fe}(\text{CN})_6^{-3}$
SAD complex	$\text{Fe}(\text{CN})_6^{-4}$, $\text{Au}(\text{CN})_2^{-1}$, $\text{Hg}(\text{CN})_4^{-2}$, SCN^-

The implementation of a cyanide remediation process should concentrate on including the recovery of cyanide metal complexes such as the $\text{Cu}(\text{CN})_3^{-2}$ and $\text{Cu}(\text{CN})_4^{-3}$ as the formation of these compounds results in a large loss of cyanide in the CIL at the ERGO plant (Max Gerke, personal communication, 2022).

Another section of the plant that uses cyanide is the elution circuit. In this section a 1% NaCN solution is used. As this is a strong cyanide solution 2% NaOH must be added to maintain the pH at an acceptable level to prevent hydrogen cyanide (HCN) gas production. The consumption of cyanide per day in this section amounts to an average of 2.5 tonnes of sodium cyanide. After the eluate is sent for zinc precipitation it is recycled to a 3rd column where fresh carbon is added to adsorb any residual gold and ensures that all the gold is extracted from the zinc precipitation tails solution. The column is known as the policeman column and the treated eluate stream from said

column is called the policeman column tailings (PCT). Therefore, there are two cyanide streams where possible value can be obtained from: There is the CIL effluent stream with an average daily free cyanide concentration of 50 mg/L in solution at an average production rate of 100 000 tonnes of slurry per day and the policeman column effluent stream with an average free cyanide concentration in solution of 400 mg/L and 250 m³ of clear solution per day.

1.1.4. The ERGO Plant's On-site Coarse Material

As mentioned in section 1.1 there is gold bearing material, heaped on site at the ERGO plant, that is currently unleachable in the existing CIL circuit. This material is carbonaceous in nature, with large varying particle sizes and this is the reason for its exclusion from leaching. The presence of carbonaceous material in the CIL circuit typically leads to preg-robbing and therefore a reduced recovery of the gold within the ore. The material however offers a high gold grade (in comparison to ore typically processed at the ERGO plant) of up to 2.37 g/tonne with each ore site listed in Table 1-4.

Table 1-4: Carbonaceous DRDGOLD ERGO Brakpan Resources

High Carbon Content Sites			
Site	Available Tonnes	Gold Grade	Carbon Grade (%)
City Deep	264 690	1.7	6.0
ERGO Oversize	211 476	1.3	5.5
ERGO Mills	256 860	2.4	22.4
Cason	151 035	1.9	7.3
Elsburg	235 000	0.7	4.9
Ezekiel & 4A18	129 800	1.3	12.5
4A6	57 000	1	8.8
Subtotal	1 330 861	-	-
Low Carbon Content Sites			
Ergo Ash Dump	50 000	1.1	1.0
4A9/4L25	200 477	0.8	0.9
4A30/4L30	80 000	0.8	0.9
Angelo Dump 4L43	100 302	0.5	0.7
Subtotal	430 779	-	-
Total	1 761 640	-	-

This reserve contains about 1.76 million tonnes of Coarse material (Table 1-4) and offers potential for additional revenue for the ERGO plant. An independent process to leach this ore could allow for the extraction of gold from the ore without negatively affecting the main ERGO CIL plant.

1.2. Problem Statement

Cyanide plays a critical role in gold mining operations, yet its cost and potential environmental repercussions underscore the need for efficient cyanide recovery methods. The recovery of cyanide can provide a cost-effective and environmentally friendly solution. While various cyanide recovery techniques exist, no universally preferred method has emerged due to the multifaceted considerations involved in selecting an appropriate approach. To implement cyanide recovery on a large scale, requires it to prove more economically advantageous than the purchase of fresh cyanide or cyanide destruction methods. The utilization of cyanide from the ERGO plant's unaltered effluent streams for gold ore leaching could offer an alternative solution for the use of cyanide from effluent streams, potentially enhancing the profitability of gold mining operations. In the context of gold mines, it is not uncommon to encounter gold-bearing material that is incompatible with the main operating or mining process. Hence, investigation into the feasibility and profitability of utilising cyanide in cyanide effluent streams for the leaching of such material would contribute valuable insights to the existing literature on gold mining and cyanide re-utilisation.

1.3. Research Questions

To better understand the possible cyanide recovery and reuse methods available and to assess the economic and technical viability for use at the ERGO plant. The research questions listed below were the driving force for this thesis:

- What cyanide recovery methods already exist in published and patent literature?
- Can additional gold extraction be realized by leaching of waste gold-bearing material with cyanide from effluent streams of the ERGO plant?
- Is using cyanide from effluent streams of the ERGO plant for leaching of waste gold-bearing material economically feasible?

1.4. Research Aim and Objectives

From the Research Questions, the following Research objectives were formulated:

- To review the published and patent literature and identify cyanide recovery and reuse methods.
- To examine the technical feasibility of using cyanide from cyanide-containing effluent streams at the ERGO plant to extract gold from a previously stockpiled gold-bearing waste material.
- To determine the economic feasibility of using cyanide from a cyanide-containing effluent stream to extract gold from previously stockpiled waste material using different leaching methods.

1.5. Outline of Thesis

This thesis consists of six chapters. Chapter 1 serves as an introduction to the thesis with the background and motivation of the study discussed and the objectives stated. Chapter 2 provides the literature review emphasizing the current cyanide recovery methods available from research and patent literature to better understand the design and procedures associated with cyanide recovery. The review also includes a section on gold percolation leaching and preg-robbing to understand the leaching of carbonaceous gold ore.

Chapter 3 describes the experimental design and procedures for the leaching of the carbonaceous gold ore under laboratory conditions. The experimental procedures for pre-leaching, CIL leaching, and percolation leaching are described with the purpose of the experiments to determine the leach conditions with the greatest extraction.

Chapter 4 demonstrates the results from the experimental procedures in Chapter 3 and compares the gold recoveries between the different leaching methods/conditions. The leaching methodologies tested include batch leaching, agitated batch leaching, the effect of aeration, the effect of comminution, and the effect of the CIL with the addition of competing activated carbon. The results of the column percolation leach experiment are also discussed.

In Chapter 5 a preliminary design of the column percolation leach process is presented which is used to analyse the profitability of the process. The process is outlined using Process Flow Diagrams. Included are detailed descriptions of equipment and their sizing followed by a cost estimation of the capital-intensive equipment. Thereafter plant revenue and costs are outlined under

variable and fixed costs. Also included is a concise section that delves into the implications for the profitability of the ERGO Brakpan plant when introducing this process.

Chapter 6 gives the conclusions of the research and offers recommendations for future research in this area.

Chapter 2: Theoretical Background and Literature Review

In Chapter 1 (Section 1.1 and 1.2) the substantial impact of cyanide usage on the cost of a gold tailings plant was highlighted. To address this, the potential attractiveness of remediation and recycling of cyanide emerges as a strategic option to enhance the profitability of gold mining operations. Section 2.1 and 2.2 will offer a thorough literature review on cyanide metal reaction chemistry and cyanide recovery, focusing specifically on methods conducive to cyanide recycling from mine effluent. Notably this discussion excludes cyanide destruction methods as the focus is primarily on sustainable practices.

The research objectives, however, also include investigating the feasibility of using cyanide from cyanide effluent streams for additional gold recovery at the ERGO plant. Therefore, in Section 2.3 value-add alternatives for revenue generation from literature are discussed highlighting work on the reuse of cyanide containing effluent process water. Following this, Section 2.4 will present a literature review on the mechanism and fluid flow of percolation leaching. This discussion forms an integral part of the investigation into the potential reuse of a cyanide from cyanide effluent streams for leaching of the ERGO plant Coarse material, employing column percolation leaching as a key methodology. Section 2.5 will address preg-robbing and explore preg-robbing mitigation strategies due to the carbon rich nature of the Coarse material.

2.1. Gold Ore in Cyanide Solution Chemistry

2.1.1. Gold Cyanide Complexes

The extraction of gold is the primary goal of any gold mine. The reaction of gold with cyanide is governed by the Elsner equation (Eq. 1) and forms a $\text{Au}(\text{CN})_2^-$ complex when in solution. This complex is the most stable form of gold in cyanide solution at a pH of roughly 9.5 and above according to the Pourbaix diagram (Figure 2-1) for a system containing gold, cyanide, and water at 298.15K (Marsden & House, 2006).

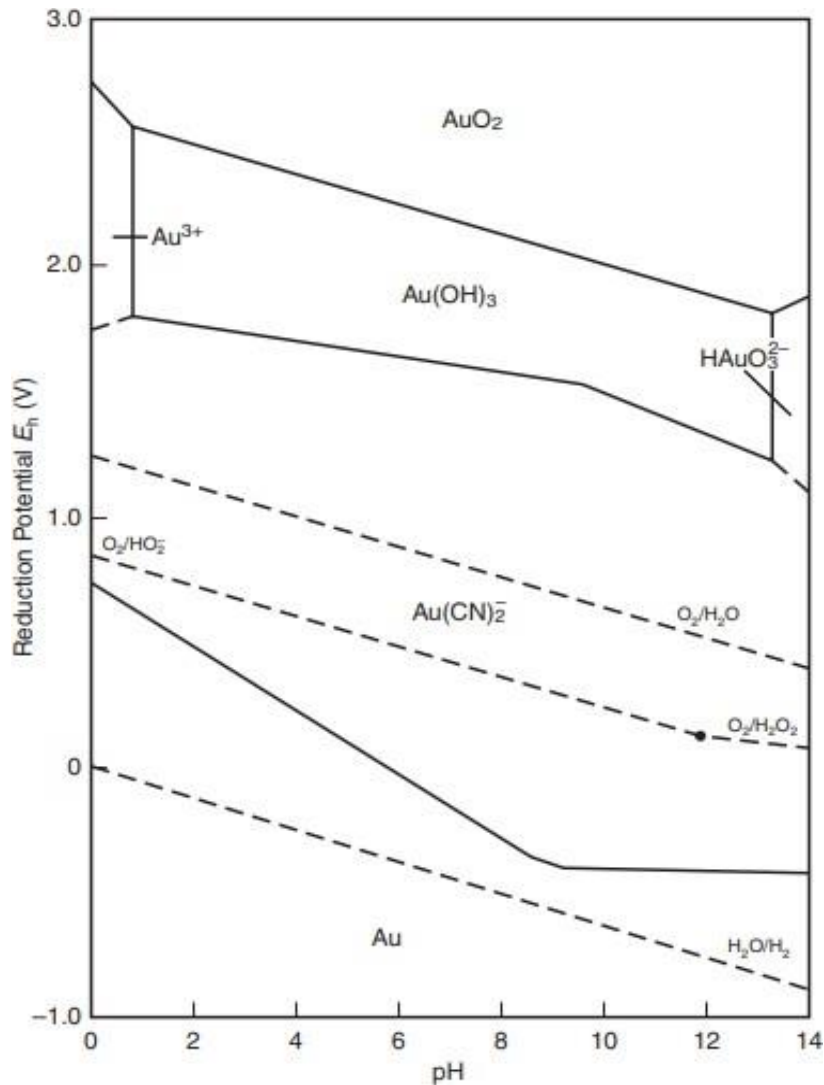


Figure 2-1: Pourbaix Diagram for gold in aqueous cyanide solution at 25 °C. Concentrations of soluble gold species = 1.0×10^{-4} M (Marsden & House, 2006).

The carbon used in the CIL plant is selective for the Au(CN)_2^- compared with other metal cyanide complexes when the metals present are in similar concentration. However, other metals such as copper and silver also load onto activated carbon, thus removing active adsorption sites for the extraction of gold. Nickel and iron complexes do not appreciably adsorb onto the activated carbon, while zinc is rarely found in gold ores and does not adsorb as well onto activated carbon as copper and silver (Marsden & House, 2006). Silver can be sold as a by-product of the gold extraction process, but the adsorption of copper should be avoided, as the increased cost of separation of the copper from a gold bar is typically uneconomic. The following section on the reaction chemistry of the copper cyanide complexes will detail the factors that lead to the decrease of copper loading onto activated carbon.

2.1.2. Copper Cyanide Complexes

Gold containing copper gold ores, such as chalcopyrite, is a common type of ore treated with cyanide primarily for the extraction of gold. The reaction and subsequent dissolution of copper in a gold mining operation are generally detrimental due to the consumption of cyanide and oxygen; two of the most expensive reagents for metal cyanide reactions. Copper forms various species with cyanide: CuCN , CuCN_2 , $\text{Cu}(\text{CN})_2^-$, $\text{Cu}(\text{CN})_4^{2-}$, $\text{Cu}(\text{CN})_3^{2-}$, and $\text{Cu}(\text{CN})_4^{3-}$. Cupric cyanide complexes are not stable in a CIL circuit due to the prevalence of oxygen and excess cyanide (Baxendale & Westcott, 1959). Therefore, $\text{Cu}(\text{CN})_2$ and $\text{Cu}(\text{CN})_4^{2-}$ do not show up in a speciation diagram for copper cyanide reactions.

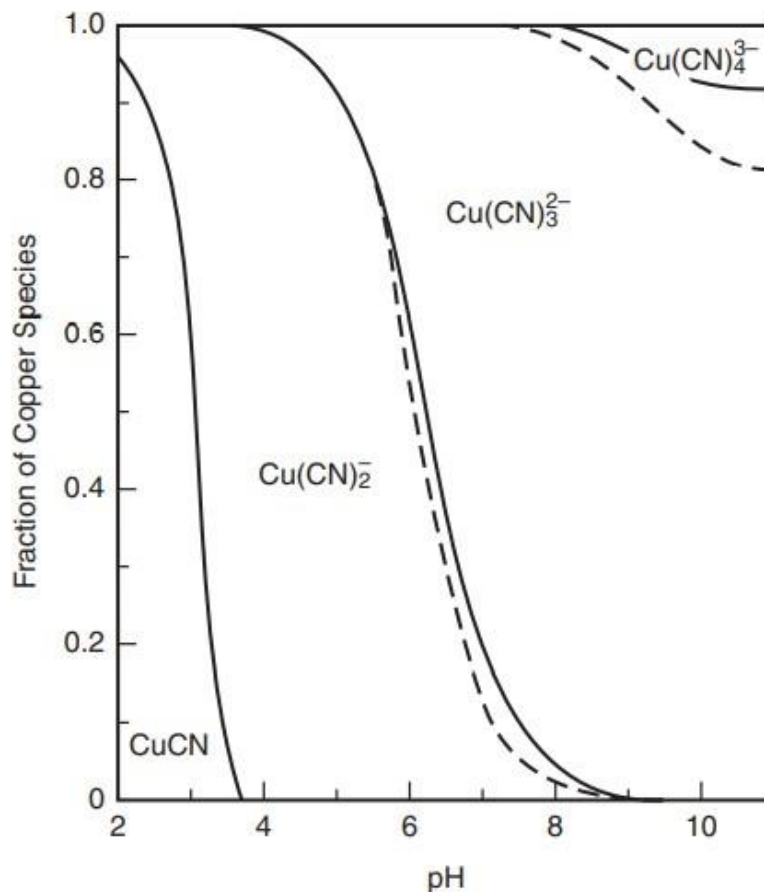


Figure 2-2: Influence of pH on the Copper Cyanide Distribution (Marsden & House, 2006)

Figure 2-2 illustrates the mole fraction of the present copper cyanide complexes in solution at constant cyanide concentration for different pH values and the effect of 80 mg/L of Zn represented by the dashed lines. Due to the use of cyanide as the leaching agent in gold mining, the solutions

are kept in the high pH range of 10.5 – 11 to ensure there is no production of HCN gas. A pH of 10.5-11, however, also works in favour of the production of the Cu(CN)_3^{2-} and Cu(CN)_4^{3-} complexes which do not adsorb onto activated carbon as well as the Cu(CN)_2^- species does. This increases gold adsorption onto the activated carbon as well as the ratio of gold to copper found on the activated carbon surface when it exits the leaching circuit (Fleming & Nicol, 1984). The same effect can be seen by the addition of excess cyanide to a leaching circuit, a decision made to ensure the maximum rate of reaction for gold dissolution. Excess cyanide in solution also promotes the production of the Cu(CN)_3^{2-} and Cu(CN)_4^{3-} over the Cu(CN)_2^- cyanide complex (Marsden & House, 2006) and reduces the competition for adsorption space on the activated carbon.

2.1.3. Iron Cyanide Complexes

Iron oxides, iron silicates, and metallic iron have a slow reaction rate with cyanide, compared with iron sulfides, in an alkaline cyanide solution. Iron compounds that do react form the stable Fe(CN)_6^{4-} metal cyanide complex. The Pourbaix diagram in Figure 2-3 illustrates the stability of the Fe(CN)_6^{4-} complex in an iron, cyanide, and sulfur solution.

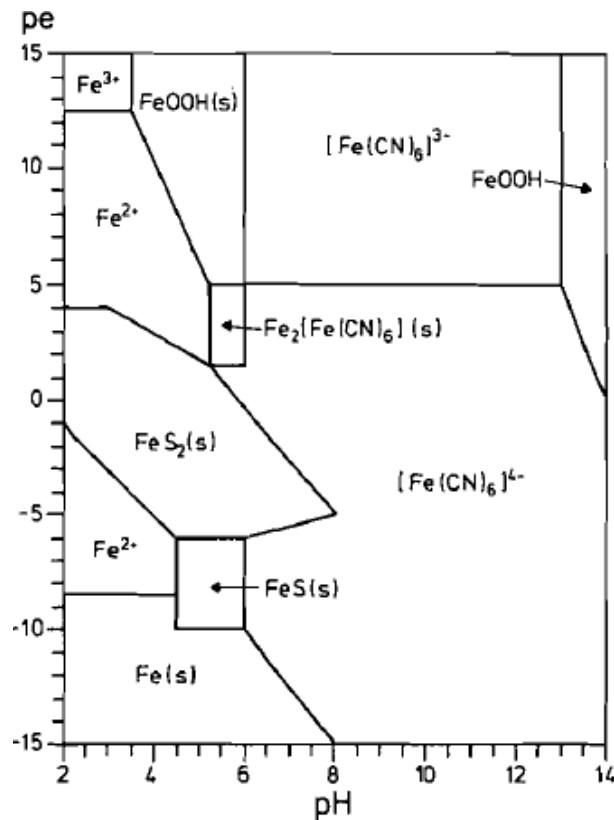


Figure 2-3: Pourbaix diagram for iron and sulfide in cyanide solution at $[\text{CN}] = 5.0 \times 10^{-3} \text{ M}$, $[\text{Fe}] = 1.0 \times 10^{-4} \text{ M}$, and $[\text{S}] = 1.0 \times 10^{-3} \text{ M}$ (Wang & Forssberg, 1990)

In the pH range of 8-14, it can be seen that $\text{Fe}(\text{CN})_6^{-4}$ is the most stable form of iron compound in the solution. The pH of a typical alkaline cyanide solution in a gold mine is in the range of 9.5-11. Therefore, the predominant iron species that exists when iron is present in a gold mine leach circuit is the $\text{Fe}(\text{CN})_6^{-4}$ complex. The presence of iron sulfides and the formation of the iron cyanide complex adversely affect the leaching of gold in two ways. First, 6 moles of cyanide are needed for every mole of iron cyanide complex created. This leads to large amounts of cyanide being consumed by undesired side reactions. Secondly, iron sulfides reacting with cyanide can negatively affect the rate of gold dissolution thereby also increasing cyanide use (Marsden & House, 2006).

2.2. Cyanide Recovery Methods

With cyanide contributing a significant amount to the cost of any gold mine, there has been research into cyanide remediation for over 90 years. There is therefore substantial research into the area with a multitude of remediation methods developed in this time. Cyanide remediation methods are grouped based on the mechanisms by which the cyanide is recovered.

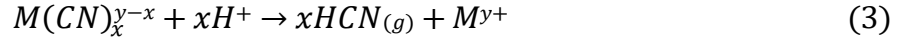
2.2.1. Cyanide Acidification Methods

Acidification, Volatilization and Reneutralization.

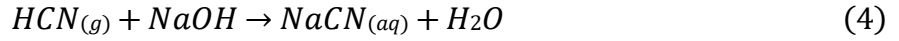
If cyanide is used in any process the pH of the system is always carefully monitored to ensure the safety of using cyanide in the process. This is due to the production of HCN gas at a pH as high as 9.3 (Goldstone & Mudder, 1998b; Vapur, et al., 2005). To avoid the production of HCN gas mines like ERGO Brakpan ensure that the slurry has a pH of at least 10.5, however, it is the volatility of HCN and therefore subsequent creation of HCN gas that provides the mechanism for acidification and volatilization processes to work.

The Acidification, Volatilization and Reneutralization (AVR) process is the first remediation process that was developed and has been in use for over 50 years. (SGS Mineral Services, 2022) In the acidification step, the first stage of the AVR process, an acid is added to reduce the pH of the slurry or solution to volatilize the cyanide in the solution. Depending on the target cyanide compounds, the pH can be optimized to ensure excess acid is not used to decrease the pH of the solution further than needed to volatilize the cyanide compounds present in the solution. Volatilization of cyanide increases with a decrease in the pH and at a pH of less than 2.5 even Strong Acid Dissociable (SAD) cyanide complexes can be volatilized. For the volatilization of

Weak Acid Dissociable (WAD) and Medium Acid Dissociable (MAD) cyanide complexes a pH of < 6 and 4 respectively are required (Goldstone & Mudder, 1998b; Vapur, et al., 2005). Volatilization of cyanide is governed by two equations (Eq. 2 and 3):



Where M = metals bonded with cyanide. As can be seen in Equation 3, when cyanide is volatilized HCN gas is the product. HCN gas is then captured, usually in an absorption tower. NaOH or other hydroxide salts can then be used to neutralize the HCN gas and form cyanide salt compounds shown in Equation 4:



Depending on the pH required for the AVR process and the regulations related to effluent pH, the effluent wastewater can be re-neutralized using lime (Vapur, et al., 2005). During the re-neutralization of the effluent metals liberated from cyanide form metal hydroxides, which precipitate out of solution as shown in Equation 5 (Young & Jordan, 1995):



Where M = metals bonded with cyanide.

A study into the effect of lowering the pH of the cyanide-containing effluent on copper showed that only 2 thirds of the cyanide can be extracted from the prevalent and stable copper tri-cyanide complex. The copper tri-cyanide complex, $Cu(CN)_3^{-2}$, dissociates at pH as high as 6 but this leads to the precipitation of CuCN (Goldstone & Mudder, 1998b; Vapur, et al., 2005; Fleming, 2016)

There exist several variations of AVR in the patent literature and these are listed in Table 2-1. The ability of each method to extract cyanide complexes is included together with the reference of each patent.

Table 2-1: AVR variations from Patent Literature

Identification	Operating pH	Recovery of				Reference
		CN ⁻	WAD	MAD	SAD	
CANMET	2.4-2.5	Yes	Yes	Yes	Yes	(Worldwide Patent No. WO1990004655A1, 1990)
Cyanisorb	6	Yes	Some	No	No	(SGS Mineral Services, 2013)
CRP	6-9.5	Yes	Some	No	No	(Worldwide Patent No. WO1990004655A1, 1990)
W088/08408	3.5	Yes	Yes	Yes	No	(Worldwide Patent No. WO1988008408A1, 1988)
US3592586A	2-4	Yes	Yes	Yes	Yes	(United States Patent No. US3592586A, 1971)

Cyanisorb and CRP are two AVR processes that have been used in precious metal plants (Fleming, 2005). The inherent advantage of these two processes is that the pH only goes as low as 6. This results in lower acid consumption and lower base consumption to raise the pH of the solution after the AVR process. Vapur et al. (2005) showed that acid consumption rises exponentially as the pH decreases below 6 and this will have a negative effect on the costs required to run the AVR process. Simple and some WAD cyano complexes such as $\text{Zn}(\text{CN})_2$, CuCN , $\text{Ni}(\text{CN})_2$, NaCN , KCN , and $\text{Ca}(\text{CN})_2$, were shown to dissociate to HCN at a pH of 6 (Vapur, Bayat, Mordogan, & Poole, 2005).

Reaction Eq. 3 and Eq. 4 produce HCN gas are rapid. Whereas in Reaction Eq. 2, the volatilization of the HCN gas is the rate-limiting step in the AVR process (Kuyuck & Akcil, 2013). Increasing the kinetics of the volatilization step can be done by increasing the air flow rate, decreasing the pH, or altering the gas-to-liquid ratio (Vapur, et al., 2005; Kabasakal, et al., 2004).

Increasing the air flow rate allows for greater concentration differences, thus increasing the rate at which HCN in solution is volatilized. An airflow of 2065 m³/min was shown to reach a NaCN recovery of 95 % after 1 hour, while an airflow of 1005 m³/min only allowed for the recovery of 85 % over 1 hour in a 1 m³ batch system (Vapur, et al., 2005). Decreasing the pH results in faster volatilization due to the increased concentration of hydrogen ions available for the formation of HCN, thereby increasing the production of HCN gas. When the pH of the solution was lowered from 5.5 to 2.7 there was an increase in the recovery of both free and complex cyanides while also dropping the volatilization times necessary to reach the same % recovery of cyanide (Kabasakal, et al., 2004). Increasing the gas-to-liquid ratio showed a positive correlation with the % of cyanide recovery. Therefore, it is beneficial to increase the gas-to-liquid ratio for greater cyanide recovery in the AVR process (Vapur, et al., 2005).

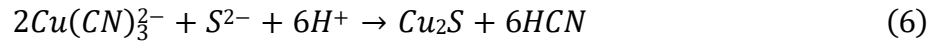
The processing of slurries introduces a complication in the AVR process. This is due to the increase in volatilization time of HCN when present in a slurry solution. At reaction times over 2 h the volatilization of HCN was reduced from 96.5 % from a solution compared with 54 % from a slurry solution showing that the presence of solids in a slurry can have a measurable effect on volatilization time (Dai, et al., 2012; Fleming & Trang, 1998). However, Vapur et al. (2005) showed that a cyanide recovery of 97.3 % from a slurry containing 40 % solids can be reached by optimizing parameters such as the gas-to-liquid ratio. Slurries, however, increase acid consumption due to the acid being consumed by other compounds in the slurry that are not present in a clear solution (Dai, et al., 2012).

To ensure safety for the operators the pH of an AVR process is always carefully monitored and controlled. The AVR process must take place in a sealed system, due to HCN gas production, and the need for large amounts of acid to reduce the pH results in a process that can be expensive to operate (SGS Mineral Services, 2022). Generally, a targeted recovery of 95% of cyanide in the slurry is needed for a profitable AVR process (Goldstone & Mudder, 1998b; Vapur, et al., 2005).

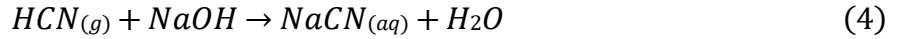
Sulfidization Cyanide Recovery Processes

Sulfidization processes make use of sulfur, by the addition of sulfuric acid and sulfide ions, as an acidification agent while also precipitating copper out of the solution as Cu₂S. The acidifying of the slurry also results in the production of HCN gas as described in the AVR process. Therefore, using sulfidization can result in the recovery of cyanide from copper, the recovery of copper metal,

and free cyanide in solution. Sulfidization processes are designed to regenerate cyanide from high cyanide copper complex concentration slurries (Estay, 2018). There exist two prominent sulfidization processes, namely the Metallgesellschaft Natural Resources (MNR) and sulfidization, acidification, recycling, and thickening (SART) processes. Sulfidization processes use sulfide ions, according to reaction Eq. 6 (Dai, et al., 2012), to create the desired copper sulfide precipitate. The reaction given in reaction Eq. 6 governs both the SART and MNR sulfidization cyanide recovery processes.



Sulfidization also involves the reaction Eq. 4 which shows the re-neutralization of the recovered HCN gas.



The SART process was first put into commercial operation at the Telfer gold mine in Australia in 2006 (Estay, 2018). SART has been shown to provide up to 90% copper recovery and free cyanide recovery of up to 90% (Estay, et al., 2012; Ford, et al., 2008). The sulfidization and acidification stages of SART use acids like H_2SO_4 to reduce the pH of the slurry to 4-5 promoting the dissociation of cyanide from WAD cyanide complexes, while NaS is added to provide the sulfide ions needed for sulfidization (Bustos, et al., 2010). The solids produced by sulfidization are then removed by thickening and filtering the solution. The solution can then be neutralized to a pH of 11 by a base such as $Ca(OH)_2$ to create $CaCN_2$ which can be recycled back into the system. The addition of lime, however, leads to the formation of gypsum which must be removed by thickening and filtration of the solution (Estay, 2018).

SART and AVR represent distinct approaches. In the AVR process, cyanide present in the mine's effluent is effectively removed, whereas SART operates by recycling the cyanide within the same effluent solution. Consequently, AVR may provide a cheaper way to reach the cyanide concentration set by the mine for its cyanidation process. SART cannot treat slurries of greater than 10% solids as a necessary function is copper recovery recovered by filtration. SART is therefore not suitable for use in an agitation/CIL leaching process for cyanide recovery unless coupled to another cyanide recovery method such as AVR (Estay, 2018).

MNR was the first sulfidization method and works much in the same way as SART but only makes

use of filtering when separating the metal precipitate from the solution and does not use thickening. Solids can make up as little as 0.5% of the solution in MNR and this results in large filtration systems being needed. Using thickening SART can increase the solids % to 10-15 % which allows for a far smaller filtration system to be used (Fleming, 2005).

Sulfidization cyanide recovery methods must ensure that excess sulfide ions are not recirculated to the leaching feed. Sulfide can interact with cyanide and form SCN^- , thereby decreasing the free cyanide available in the leaching circuit. Many gold mines also make use of Ca(OH)_2 to raise the pH of their leaching section of the plant and the use of sulfuric acid can lead to the production of gypsum, which can lead to scaling problems in the leaching circuit (Dai, et al., 2012; Young, et al., 2001). Consequently, sulfides and sulfates in the cyanide recycle stream must be kept low to mitigate these problems. Sulfates in the cyanide recycle stream can be avoided using another acidifying reagent in the SART process or by controlling and not exceeding the stoichiometric ratio of sulfide ions to the copper cyanide complex outlined in Reaction Eq. 6.

Gas Membranes

Gas membranes, also known as gas-filled membranes, provide the basis for a separation technique where 2 solutions are partitioned by a hydrophobic microporous membrane with pores filled with air. This approach is well suited for separating volatile compounds from solution (Shen, et al., 1997) and materials like hollow fiber polypropylene microporous membrane are popular for use in a gas-filled membrane system. During operation the air remains in the pores if the pressure difference between the solutions on either side of the membrane is less than the breakthrough pressure (Han, et al., 2006). Nevertheless, as such membranes can be adversely affected by fouling from solids and particulate matter in the effluent wastewater, it is best to have a clarified stream entering the gas-filled membrane system. As a result, the use of coagulation and filtration as pre-treatment steps can be an important tool to keep the membrane free of fouling and operating at its best (Han, et al., 2006; Shen, et al., 1997). In the case of the ERGO CIL, the exit stream has substantial daily throughput and would require large filtration facilities if cyanide recovery by gas membranes were to be implemented. This requirement may diminish the overall attractiveness of this process. Using gas membranes for the recovery of cyanide involves acidifying the solution so that HCN is the dominant cyanide species in the solution. Gas membranes can therefore be seen as like the AVR process. A gas membrane cyanide recovery method begins with acid pumped through an anion exchanger counter current to the cyanide-containing solution. This results in the

concentration of cyanide and subsequent volatilization of cyanide into HCN. HCN then transfers by diffusion from the solution to the wall of the gas membrane before diffusion through the gas membrane. The final step involves HCN diffusion into a stripping solution containing a salt hydroxide where it reacts to form a salt cyanide compound that can then be recycled (Kenfield, et al., 1988). Transfer by gas membranes is primarily dictated by diffusion and has been modelled using the resistance in series model and double-film theory. The overall mass transfer resistance ($1/K$) consists of the feed liquid film resistance ($1/k_L$), membrane resistance ($1/k_m$), and stripping liquid film resistance ($1/k_{L'}$) which is expressed in reaction Eq. 7 (Shen, et al., 1997):

$$\frac{1}{K} = \frac{1}{k_L} + \frac{1}{k_m} + \frac{1}{k_{L'}} \quad (7)$$

It was found that the membrane resistance and the stripping liquid film resistance were negligible when compared to that of the feed liquid resistance. Based on the overall mass transfer resistance and Fick's law an equation to describe the gas membrane transfer system has been derived as shown in by reaction Eq. 8:

$$\ln \left(\frac{[HCN]_{F,0}}{[HCN]_{F,t}} \right) = \left(\frac{KA}{V_{F,0}} \right) t \quad (8)$$

Where $[HCN]_{F,0}$ represents the initial HCN concentration in the bulk feed and $[HCN]_{F,t}$ represents the concentration at a time t . K is the overall mass transfer resistance, A is the total area of the membrane with $V_{F,0}$ representing the initial feed volume, and t representing the time (Han, et al., 2006).

It has been shown that fouling of the membrane results in a drastic decrease in the performance of the gas membrane cyanide recovery method. An experiment to understand the effect of fouling was carried out by Han et al. (2006) using a gas membrane cyanide recovery method carried out at pH 6. When an artificial clarified industrial wastewater stream was used the recovery of cyanide was 84% after 20 min compared with real industrial wastewater where only 40% of the cyanide was removed from the solution after the same time. Even after 60 min of operation the recovery of cyanide from real industrial wastewater only reached 60%. Therefore, it is important that when membranes are involved in the treatment of wastewater processes, the input solution must be purified of contaminants to prevent fouling. To demonstrate this Han et al. subjected the real wastewater to coagulation to remove contaminants and this improved the recovery of cyanide to

98% and with minimal loss of cyanide due to the coagulation step.

2.2.2. Adsorption Cyanide Recovery Processes

Adsorption as a mechanism for the recovery of gold has been in use for over 50 years (Dai, et al., 2012). Adsorption of cyanide and its complexes occur in the same way that gold cyanide complexes are adsorbed in a traditional CIL process (Young & Jordan, 1995). Research into adsorption as a mechanism for cyanide recovery has centered on the use of activated carbon and ion exchange resins as adsorbents.

Activated Carbon

Activated carbon (AC) has been the predominant adsorbent for the gold cyanidation CIL process. The use of adsorbents allows for the treatment of slurries with far less impairment to cyanide recovery than the previously discussed SART and AVR cyanide recovery processes and therefore does not need a solid-liquid separation step before the process can take place (Dai, et al., 2012). AC can also be used for the adsorption of other metals such as copper and zinc. Activated carbon can adsorb free cyanide but removal is limited to 5 mg of CN^- per gram of carbon. Efficient recovery of cyanide using AC requires that the cyanide be bonded with metals such as copper, zinc, or nickel (Marsden & House, 2006). When gold mines use AC for gold recovery excess cyanide is added to promote the adsorption of gold over other metal cyanide complexes. For example, copper cyanide complex adsorption affinity decreases as the copper is laden with more cyanide ions which further increases the negative charge of the complex. Therefore, the $\text{Cu}(\text{CN})_3^{-2}$ and $\text{Cu}(\text{CN})_4^{-3}$ species do not load significantly onto carbon in contrast to the $\text{Cu}(\text{CN})_2^-$ species does (Fleming & Nicol, 1984). The use of AC as a cyanide recovery method would therefore require that the copper in solution be predominantly the $\text{Cu}(\text{CN})_2^-$ complex to appreciably recover cyanide and copper from the tailings. Thus, copper is added to the tailings to convert free cyanide and the $\text{Cu}(\text{CN})_3^{-2}$ and $\text{Cu}(\text{CN})_4^{-3}$ species to the $\text{Cu}(\text{CN})_2^-$ complex so that copper and cyanide can be readily adsorbed onto the AC (Dai & Breuer, 2009). This allows the adsorption circuit to be utilized in a similar way to any other CIL circuit. The AC loaded with copper cyanide can then be eluted with a cyanide and NaOH solution. Thereafter another system such as AVR or SART can be implemented to separate copper and cyanide to create copper bars and free cyanide to be recycled to the feed of the gold CIL circuit or the copper can be separated by SART, or electro-winning and the free cyanide sent for recycling. Moreover, copper from electrowinning can be sold or recycled for use in the AC cyanide recovery process (Dai, et al., 2012).

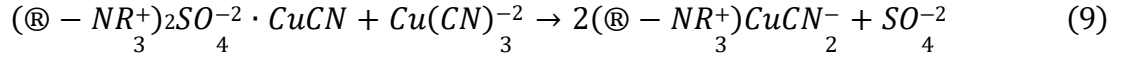
Ion Exchange Resins

The use of ion exchange resins in the gold mining industry was first put into practice in the 1970s in Uzbekistan where the first resin in pulp gold extraction plant was commissioned. It has, however, lagged the widespread use of AC as an adsorbent in the gold mining industry. Ion exchange resins that provide economically feasible operations are not as selective as AC and attrition of the resin beads is higher than AC. As a result of the low selectivity of ion exchange resins a sophisticated elution circuit is needed (Marsden & House, 2006). Ion exchange resins are classified as strong or weak depending on whether the functional group is made of a strong or weak acid/base and whether they are anion or cation resins (SAMCO, 2017). In the gold industry using cyanidation creates negatively charged anions in the form of metal cyanide complexes and so the ion exchange resins needed will be anion resins if ion-exchange is to occur.

The first process and pilot plant using ion exchange for cyanide recovery was recorded in 1959 by Eric Goldblatt (Goldblatt, 1959). The process starts with 2 adsorption columns. The first column adsorbs metal cyanide complexes, found in the effluent of the mine, onto the strong anion exchange resin. Effluent from the first column is then sent to a second column containing an anion exchange resin already conditioned with CuCN. In the second column, free cyanide in the mine effluent is complexed with the CuCN and therefore absorbed on the anion exchange resin. The resin beads from both columns are then sent for elution with 1% H₂SO₄ solution. The CuCN conditioned onto the resin beads is kept at a steady state as following elution CuCN re-precipitates onto the anion exchange resin. If copper cyanide complexes are found in the feed of this system (in the mine effluent) copper can quickly become a foulant due to lack of elution of copper cyanide and this quickly decreases the ability for metal complexes to be adsorbed on to the ion exchange resin (Goldblatt, 1959). From this process, the patented Augment process was derived.

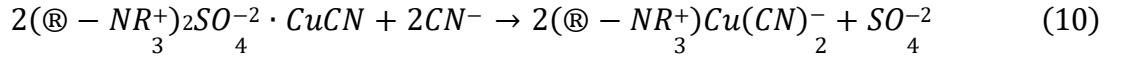
The Augment process was developed to recover cyanide and selectively recover copper in gold mine tailings using ion exchange resins (Fleming, 2016). The process depends on the chemistry of the cyanide-to-copper solution ratio (CN/Cu). At high CN/Cu values the prominent copper cyanide complexes are the Cu(CN)₃⁻² and Cu(CN)₄⁻³ cyanide copper complexes and at low CN/Cu values the Cu(CN)₂⁻ is the predominant cyanide complex. The Augment process makes use of the resin-in-pulp (RiP) process to contact the resin with the mine effluent (Fleming, 2016). The best loading of copper onto the resin is achieved by having a CN/Cu ratio of 2:1 as this promotes the Cu(CN)₂⁻ complex for adsorption. Only one resin adsorption site is needed for Cu(CN)₂⁻ compared with two

and three for the $\text{Cu}(\text{CN})_3^{-2}$ and $\text{Cu}(\text{CN})_4^{-3}$ complexes respectively (Dai, et al., 2012; Fleming, et al., 1998). Like the very first ion exchange cyanide recovery technique, the Augment process also precipitates CuCN onto the pores of the strong base resin. The CuCN reacts with the $\text{Cu}(\text{CN})_3^{-2}$ complex in the tailings and forms $\text{Cu}(\text{CN})_2^-$ which is the most readily adsorbed of the copper complexes (Fleming & Nicol, 1984). The reaction governing the production and adsorption of $\text{Cu}(\text{CN})_2^-$ is dictated by Eq. 9 (Fleming, 2016):



Where $\textcircled{R}$ represents the chosen anion base resin and $-\text{NR}^+$ represents one resin site.

Once the loading of the copper cyanide complex has built up sufficiently, free cyanide is also able to be extracted from the mine effluent as shown by Eq. 10 (Fleming, 2016):



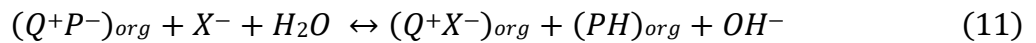
Once the adsorption step has taken place the resin is sent for elution. The eluant consists of a copper cyanide solution with a CN/Cu ratio of four or greater (SGS Mineral Services, 2013). $\text{Cu}(\text{CN})_2^-$ is converted to the $\text{Cu}(\text{CN})_3^{-2}$ complex on the surface of the resin beads and since $\text{Cu}(\text{CN})_3^{-2}$ uses an extra resin site half of the copper cyanide found on the resin beads is eluted as $\text{Cu}(\text{CN})_3^{-2}$. The remaining $\text{Cu}(\text{CN})_3^{-2}$ found on the resin beads is treated with H_2SO_4 to regenerate the CuCN on the resin beads while also creating 2 molecules of HCN per molecule of $\text{Cu}(\text{CN})_3^{-2}$ treated which is then sent for recycling while the resin is also made ready for re-use. The solution containing copper tri-cyanide created in the elution is sent to an electro-winning cell where copper is plated onto stainless steel and more cyanide is regenerated for recycling (Fleming, 2016). The Augment process can recover both copper and cyanide from slurries or solutions which makes it versatile. Nonetheless, the Augment process includes changes in pH when regenerating the resin which can cause osmotic shock and therefore losses related to bead breakage (Dai, et al., 2012). Due to the technical challenges of the Augment process, it may require that the operators of the system are highly skilled to ensure the process proceeds as required, which can lead to increased labour costs (Fleming, 2005).

The Hannah process is another strong anionic resin-based cyanide recovery method that is also capable of recovering copper as in the Augment process. It can also recover zinc which may increase the profitability of the process over the Augment process by using a filtering process. The

inner workings remain proprietary to the creators, SGS Lakefield Research and Thorpe Consulting, but up to 95% of cyanide in mine effluent has been claimed to be recovered by the Hannah process (SGS Mineral Services, 2013; Fleming, 2016).

2.2.3. Solvent Extraction Cyanide Recovery Methods

A potential alternative to ion exchange resins is using a solvent to conduct ion exchange and thereby separating metal cyanide anions from a cyanide-containing solution. Solvent extraction only works with solutions as it is the same as liquid-liquid extraction where two immiscible liquids interface. Therefore, costly solid-liquid separation is needed as a precursor to solvent extraction for cyanide recovery if the effluent is a slurry (Davis, et al., 1998). Due to the similarity of solvent extraction to the ion exchange resin cyanide recovery methods, discussed earlier, solvent extraction has a comparable inability to extract free cyanide to any useful degree, unless complexed with a metal cyanide prior to the solvent extraction process (Fleming, 2016). The solvent extraction system proposed by Davis et al. (1998) makes use of a mixture containing quaternary amine and 4-nonylphenol (LIX 7820) reagent produced by Cognis. Nonylphenol is a weak acid, and above pH of 10.5, such as is found in alkaline leach effluent, the nonylphenol begins to convert to the phenoxide anion and bonds to the quaternary amine cation forming an ion pair. At pH of 10.5 and below is where extraction of anions occurs due to the quaternary amine cation needing an anion to remain neutral and nonylphenol not yet converting to phenoxide (Davis, et al., 1998). The reversible ion exchange reaction is presented by reaction Eq. 11 and represents the extraction (reverse reaction) and the stripping (forward reaction) of anions (Xie & Dreisinger, 2009).



Where Q^+ represents the quaternary amine cation, PH represents the nonylphenol, and X^- represents the anion.

Increasing the CN:Cu ratio decreases the extraction efficiency of copper and cyanide due to the formation of $Cu(CN)_4^{-3}$ over $Cu(CN)_3^{-2}$. $Cu(CN)_3^{-2}$ has been shown to have preferential extraction by the LIX 7820 reagent with little extraction of free cyanide. Metals such as zinc and nickel are also extracted but this leads to competition between metal cyanide complexes decreasing the ability to target a specific metal bonded with cyanide (Xie & Dreisinger, 2009). Furthermore, during solvent extraction increasing copper content leads to a decrease in extraction efficiency (Xie & Dreisinger, 2010).

Stripping of the metal cyanides can be done efficiently and easily using 1M NaOH to strip (recover) above 90% of the copper and cyanide into a low volume that can then be sent for metal cyanide separation such as SART or electrowinning (Xie & Dreisinger, 2009; Dai, et al., 2012).

Solvent extraction has three drawbacks. The first is the low efficiency of extraction when the solution contains high concentrations of copper. Secondly, there is a decreasing level of extraction with each stage as the CN:Cu ratio increases due to the poor extraction of free cyanide (Dai, et al., 2012). Finally – as mentioned earlier – is the inability to allow for extraction of cyanide from slurries.

2.2.4. Membrane Cyanide Recovery Methods

Gas-filled membranes have been covered in the acidification section due to the need for acidification as a process step. This section provides details on the use of membranes as a cyanide-concentrating process where membranes provide the underlying mechanism. Recovery of materials always requires that there be a selectivity of one species over another to provide efficient recovery. Membranes are designed to provide selective recovery of targeted materials by controlling the flow between two adjacent phases. The driving forces of membrane separation are chemical potential, pressure, or an electric field (Sceresini & Breuer, 2016). Membrane operations require that the feed is free of solids so that efficient recovery of the target material can be achieved (Dai, et al., 2012). Membrane separation by pressure and electric potential for cyanide recovery is discussed in the next sections.

Membrane Separation by Pressure

Pressure as a driving force as a separation process can be done by gas separation, reverse osmosis, nano-filtration, and micro-filtration (Sceresini & Breuer, 2016). Frequently different membrane separation techniques are used in series to enhance the recovery of the target material. In reverse osmosis, pressure is applied to an effluent or wastewater to force water to travel through a semi-permeable membrane. This creates a permeate stream, representing treated wastewater, and a concentrated stream, representing the separated target material. Reverse osmosis is used to great effect in 80% of desalination plants across the world to remove salt from seawater (Aladwani, et al., 2021). Reverse Osmosis can also be adapted for the removal of cyanide ions from effluent waste streams (after solid-liquid separation).

A pilot plant and subsequent industrial plant using reverse osmosis was installed at the Gedabek

gold-copper mine in Azerbaijan (Hedjazi & Monhemius, 2018). The process starts with process water being sent through two ultrafiltration modules to remove suspended solids from the solution. An anti-scalant is added to the solution before being subjected to reverse osmosis at a pressure of 30 Bar. At pilot plant scale the process showed an extraction of copper, gold, and silver of up to 99.2, 96.3, and 98.5 % respectively into the concentrate. Cyanide recovery in the concentrate was also found to reach 99 %. There was little fouling associated with the plant, due to the anti-scalant, even with large concentrations of gypsum reporting in the process water (Hedjazi & Monhemius, 2018). The resulting concentrate can then be treated by electro-winning or SART to recover metals and cyanide for recycling.

In a similar way, the HW Process Technologies, Inc Engineered Membrane Separation System (EMSTM) using nanofiltration for cyanide recovery was tested in a pilot plant in Durango State, Mexico. The nanofiltration membrane used differentiates between multi-valent and mono-valent salts, that results in a concentrate of copper and base metal cyanide complexes that are unable to pass through the membrane and a permeate containing free cyanide, gold, and silver cyanide complexes (Lien, 1999; Dai, et al., 2012). Typically, a permeate:concentrate ratio of 9:1 was achieved resulting in a concentrate stream with a 10 times increase in copper concentration compared with the input process solution whereas the pilot plant reached up to 90% recovery of copper cyanide complexes which reported to the concentrate. Up to 90% of gold was recovered to the permeate (Lombardi & Bernard, 2001; Fleming, 2016). The concentrate can then be subject to electrowinning or SART to extract the copper and the permeate treated by zinc precipitation to extract gold and the free cyanide can then be recycled (Fleming, 2016). EMSTM was affected by fouling due to calcium in the process water (usually due to the addition of lime during leaching). This was solved in the pilot plant test using a sodium-cycle softener resin process which reduced the calcium concentration to 150 mg/L, thus decreasing the chance of calcium salts precipitation (Lombardi & Bernard, 2001; Fleming, 2016).

Microfiltration is not typically used for the separation of aqueous ionic compounds like cyanide and cyanide metal complexes due to its large pore sizes. It can, however, be used for the decontamination of the process water prior to separation by nanofiltration or reverse osmosis (Hedjazi & Monhemius, 2018) or for the capturing of metal precipitates from a SART (Estay, et al., 2021) or zinc precipitation process.

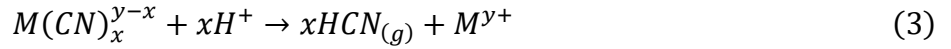
Membrane Separation by Electric Field

Using a membrane subject to an applied voltage or current for the separation and concentration of compounds in solution is known as electrodialysis. Electrodialysis differs from electro-winning as membranes stop the flow of the charged particles from reaching the surface of the electrodes but works in the same way by attracting anions to the cathode and attracting cations to the anode. An electrodialysis system to recover copper and cyanide was tested using both simulated and actual mine effluent (Zheng, et al., 2015) and the system made use of both cationic and anionic-charged ion exchange membranes. The voltage of the set-up is an important parameter of any electrolysis system and for the tested system increasing the voltage resulted in a large increase in energy consumption. At 45V the energy consumption increased to 28.47 kWh/m³ compared with 3.72 kWh/m³ for a voltage of 15V. For this system 25V was found to offer the best current efficiency to energy efficiency but with a roughly a 10 % decrease in removal rate of cyanide compared with 45V after 100 min. The initial concentration of copper and cyanide also affects the removal rate and increasing the initial concentration results in a decrease in current efficiency unless the initial concentration is very low (>121 mg/L for this system) (Zheng, et al., 2015). Increasing the system's flow rate can potentially enhance recovery if it remains within limits that do not excessively reduce the residence time needed for ions to permeate the membranes, however, increasing the flow rate also carries the drawback of elevating pressure levels, which can lead to membrane damage. (Zheng, et al., 2015). In the investigation by Zheng et al (2015) it was found that a removal rate of > 99 % of cyanide and copper was achieved at the 25V, 47 mg/L cyanide, 242 mg/L copper, and flow rate of 4.17 mL/s.

2.2.5. Cyanide recovery from Thiocyanate.

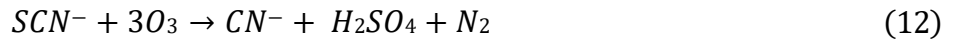
In gold mining, the interaction between cyanide and sulfur-containing minerals like pyrite or pyrrhotite produces thiocyanate. This chemical reaction consumes cyanide, which would otherwise be available for the purpose of reacting with gold. (Kuyuck & Akcil, 2013). Several processes have been proposed for the regeneration of cyanide from thiocyanate such as electrochemical oxidation, biological degradation, and ozone oxidation (Botz, et al., 2001). Among these processes, ozone oxidation has been demonstrated to be a promising technique for the recovery of cyanide from thiocyanate. It has also been shown that ozone exhibits promise in the regeneration of cyanide from thiocyanate that forms when sulfur-containing materials are subjected to cyanidation (Botz, et al., 2001).

The oxidation process begins with adjustment of the solution containing thiocyanate to a pH range of 3-5 by adding an acid like sulfuric acid or hydrochloric acid. This ensures that ozone can react effectively with thiocyanate and that cyanide can be stabilized as HCN gas (Soto, et al., 1995). This reaction was shown earlier in reaction Eq. 3:



If copper is present in solution, it should be removed before applying ozone oxidation, as it can interfere with the reaction and reduce the efficiency of cyanide regeneration (Soto, et al., 1995). Copper can be removed by using hydrogen peroxide to oxidize it to copper sulfate, which can be separated by filtration or precipitation. Alternatively, copper can be removed by precipitating it with a cationic complexing agent such as ethylene-diamine-tetra-acetic acid (EDTA), which forms a soluble complex with copper that can be extracted by solvent extraction or ion exchange (Soto, et al., 1995).

Ozone can be generated by passing oxygen gas through an electric discharge or a UV lamp. The ozone gas is then bubbled through the solution containing thiocyanate at a controlled rate and temperature. The ozone reacts with thiocyanate to form cyanide, sulfuric acid, and nitrogen gas according to reaction Eq. 12 (Jara, et al., 1995):



The reaction is exothermic and releases heat, which can increase the temperature of the solution and affect the solubility of cyanide, therefore, cooling may be required to maintain optimal conditions for cyanide recovery (Jara, et al., 1995). The cyanide produced by ozone oxidation is present as hydrogen cyanide gas in equilibrium with aqueous cyanide. The recovery of the cyanide as hydrogen cyanide is done in the same way as the AVR process described in section 3.2.1. An air stream is used to promote the volatilization of HCN from solution into the stream by stripping. The air stream containing HCN is then sent through an absorber where HCN is absorbed into a basic solution of NaOH or Ca(OH)₂ (Jara, et al., 1995). If industrial effluent contains a relatively high concentration of copper ions, it will need to be converted to copper oxide with hydrogen peroxide prior to ozonation and thiocyanate recovery as the available copper would react with cyanide increasing the loss of cyanide in the effluent. In solutions of thiocyanate, up to 85% cyanide recovery is possible when keeping the copper concentration below 10 mg/L, a pH < 7.0 and at a consumption of 1kg of ozone per kilogram of thiocyanate (Soto, et al., 1995).

2.3. Value-Add Alternative to Cyanide Recovery

In Section 2.2 a comprehensive overview of various cyanide recovery methods, highlighting their performance and process operation. In the present section, it is crucial to evaluate these methods on safety and technical feasibility to implement cyanide recovery at the ERGO plant.

HCN gas production at the ERGO plant is strictly prohibited in any environment, including the plant or laboratory due to safety concerns and past accidents. This removes the viability of any cyanide recovery method that involves using acidification for cyanide concentration along any of the process steps.

In addition, it is important to note that several cyanide recovery methods necessitate the inclusion of filtration or solid-liquid separation as a process step. At daily solids throughputs of up to 60 000 t/day (slurry throughput of up to 120 000 t/day) the filtration requirements are technically feasible but would increase complexity of the recovery process as well as increase costs that put the financial viability of the cyanide recovery method at risk. This removes the viability of cyanide recovery methods that require minimal solids in solutions from being implemented on the CIL effluent. The cyanide recovery methods are listed in Table 2-2 with a distinction made on whether it requires acidification and/or filtration.

Table 2-2: Cyanide Recovery Methods Process Requirements

	HCN production	Slurry Solid-liquid Separation
AVR	Yes	No
SART and MNR	Yes	Yes
Gas Membranes	Yes	Yes
Adsorption by AC/Resin	Yes	No
Solvent Extraction	No	Yes
Membrane Extraction	No	Yes
From Thiocyanate	Yes	No

Based on Section 2-2 the cyanide recovery method that can be best suited for the CIL effluent of the ERGO plant is adsorption by such as the Augment Process, however, there are challenges that put financial viability at risk. The process would require its own tanks, acid wash columns, elution columns and ion exchange resin regeneration facilities (Fleming, 2016). Moreover, the 50 mg/L cyanide concentration in the ERGO plant CIL effluent would not provide cost savings to cover the cost of cyanide extraction by the adsorption cyanide extraction method.

There is, however, another cyanide effluent stream - the policeman column tailings with a free cyanide concentration of 400 mg/L. Cyanide recovery on the policeman column tailings is feasible by all the cyanide recovery methods presented in Section 2.2 due to it being a clear solution without solids. Furthermore, at 400 mg/L the stream does not need concentration and the effluent stream can be used in its current state for leaching. In Section 1.1.4 attention was brought to the discarded Coarse material stockpiles on-site at the ERGO plant was highlighted. Therefore, using the policeman column tailings as the leaching solution for the leaching of the previously discarded and gold-bearing Coarse material may provide a “value-add” alternative to cyanide recovery by using the cyanide within the policeman column tailings to generate additional revenue for the ERGO plant. Gbedemah et al. (2022) conducted a study examining the impact of recycling cyanide-containing effluent for use as process water. They observed that reusing this effluent resulted in a decrease in gold recovery from 87% to 80%, however, it was found that 40 % of the initial cyanide was recovered by reusing the cyanide effluent stream. They highlighted the potential for generating revenue and saving costs on fresh cyanide purchases by reusing cyanide-containing effluent as process water within leaching processes (Gbedemah, et al., 2022).

Therefore, the following sections will focus on gold extraction processes viable for gold extraction from the Coarse material. The Coarse material consists of particles that are both large and rich in carbon, but it has a low concentration of gold. Percolation leaching is an extraction method suited for material with large and varying particles sizes as well as being cost effective due to its simplicity (Thenepalli, et al., 2019). Nonetheless, alternative methods may be required due to carbonaceous material in the Coarse material. Preg-robbing is the adsorption and thereby loss of gold to preg-robbing material in the ore with the ability to adsorb gold (Angove & Acar, 2006). Typically, preg-robbing occurs due to carbonaceous material within the ore. Even naturally formed AC has been found present in carbonaceous ores and ores with naturally occurring AC show the greatest affinity for preg-robbing (Aylmore & Muir, 2001). Therefore, a review of the literature on both percolation leaching and preg-robbing will be done in the following sections.

2.4. Percolation Leaching Operations

Percolation leaching is a leaching method whereby the leaching solution is allowed to flow through a body of ore unagitated and is predominantly used within the mining industry as a heap leaching

operation. Percolation leaching also allows for the leaching of material with varying particle size such as the Coarse material at the ERGO plant. Percolation leaching is predominantly associated with heap leaching and therefore most of the literature concentrates on the heap leach process. The first heap leaching of gold and silver ore was conducted in 1969 in the USA and is a process where crushed metal-bearing ore is loaded, in the form of a heap, on a large impermeable pad while pouring solutions containing lixivants such as cyanide over the heap to dissolve the target metals and collect the resulting solution from the bottom of the pad (Manning & Kappes, 2016). The heap leaching process is well-regarded for mineral liberation across the world due to the low capital cost, ease of implementation, and low water usage (Thenepalli, et al., 2019). The use of columns for percolation leaching is generally used for the understanding of the mineralogy and extraction efficiency of a specific ore to then be used in modelling a large heap leaching operation. Therefore, the work referenced in 2.4.1 and 2.4.2 on the ore characterization and fluid flow within heaps also applies to the percolation operations within a column.

2.4.1. Ore Characteristics and its Influence on Percolation Leaching

The ore fed to a percolation leach operation will have a major effect on the design of an operation. The ore characteristics are segmented into physical, chemical, and mineralogical properties. A percolation leaching process is likely to employ ore with unique physical properties such as a wide range of particle diameters as opposed to the controlled particle sizes allowed to enter a tank leaching system such as a CIL process. When designing a percolation leach operation, column percolation leaching tests are conducted to help understand the physical and chemical properties of the ore to then be upscaled if proven feasible for a full-scale heap leach operation. Experiments typically include varying particle sizes to define the optimal metal recovery, metal recovery rate, and reagent requirement of the ore under simulated heap leach laboratory conditions (Angove & Acar, 2006).

2.4.2. Percolation Leaching Fluid Flow

Percolation leaching is very flexible as a process where leaching can be carried out on ore sizes ranging from large rock pieces to finely crushed ore. When irrigation of the ore body begins the cyanide solution slowly seeps towards the bottom of the body to an impermeable pad or bottom of the column. The rate at which the lixiviant seeps through the body should be controlled to ensure that the lixiviant has enough residence time to contact, react, and dissolve the metals into the lixiviant solution while also allowing for enough irrigation flow to allow for sufficient lixiviant to

reach the desired reaction rate (Ilankoon & Neethling, 2012). The parameters affecting the liquid flow in a percolation leach are the packing of the ore body, the fluid properties of the solution, and the gas flow rate from beneath the ore body. The predominant factor however is the packing characteristic of the ore. Therefore, it is important that the packing of the body allows for uniform permeability. Fluid flow only begins once pores and small channels in the heap become flooded, allowing the solution to begin to seep through the body along the surface of the particles of the body (Bouffard & Dixon, 2001). Therefore, it is important to understand the wash efficiency of a percolation leach. The wash efficiency equates to the solution that is sprayed on top of the ore body that replaces the solution in the ore. The wash efficiency of a well-milled and packed ore body is typically 70% in practice and therefore three displacement washes are needed to remove the solution in the heap to get recovery of the solution to 95%. A total of four displacements is therefore needed, the first being the first application of the lixiviant solution to saturate the ore (Manning & Kappes, 2016). Within the ore body air-filled pores/voids all exist. If the liquid flow rate is too high these air-filled voids begin to flood which can lead to a reduction of the mass transfer of gas to target metals in the ore. This also leads to greater solution use, which increases costs. The ore body can never be truly uniform even under lab conditions and so high and low permeable regions exist. Solution flow reaches low permeability zones, but gas flow will flow predominantly through high permeability zones, meaning that low permeability zones can become oxygen deficient (Bartlett & Prisbrey, 1996).

2.5. Preg-Robbing Phenomenon in Gold Mining

Preg-robbing is a phenomenon in gold processing wherein the gold recovery is hindered due to the presence of carbonaceous matter, which adsorbs gold cyanide complexes from the leach solution, thereby reducing gold extraction efficiency. The Coarse material described in section 1.1.4 contains some carbonaceous material and therefore a review on the effect of carbonaceous material on gold leaching will need to be understood as well as a review on the strategies to mitigate its impact on gold recovery.

2.5.1. Preg Robbing Mechanism

The preg-robbing behavior of carbonaceous matter has been attributed to various mechanisms. One of the most widely accepted explanations is the presence of graphitic or activated carbon in the ore, which can strongly adsorb gold cyanide complexes (Marsden & House, 2006). The preg-robbing capacity depends on the nature and amount of carbon in the ore. Higher concentrations of

organic carbon with a porous structure will have a higher gold-adsorbing capacity. Additionally, organic matter such as humic acids has been identified as a preg-robbing agent due to its chelating and complexing properties (Ellis & Senanayake, 2004)

2.5.2. Methods to Mitigate Carbonaceous Preg-Robbing

Several strategies have been proposed and implemented to mitigate the impact of preg-robbing on gold recovery. The most common methods include:

1. **Pre-treatment of ores:** Pre-treatment methods, such as oxidation of the carbonaceous matter can be applied to react with and remove the carbonaceous matter and thereby prevent preg-robbing (La Brooy, et al., 1994). Roasting processes involve heating the gold ore at high temperatures to oxidise the carbonaceous matter, while other oxidation processes, such as pressure oxidation or bacterial oxidation, aim to oxidize the carbonaceous matter and render it inactive. Acid washing involves treating the ore with a strong acid, such as hydrochloric acid or sulfuric acid, to dissolve the carbonaceous matter (Afenya, 1991).
2. **Addition of adsorption inhibitors:** Adsorption inhibitors, such as kerosene, diesel fuel, or polyethylene glycols, can be added during the leaching process to reduce the affinity of carbonaceous matter for gold cyanide complexes (Afenya, 1991). These inhibitors compete with gold cyanide complexes for adsorption sites on the carbonaceous matter, thereby reducing preg-robbing. However, adsorption inhibitors also impact the adsorption capability of the activated carbon used for extraction of the gold and therefore the addition of adsorption inhibitors is generally avoided (Rees & van Deventer, 2000).
3. **Use of alternative lixiviants:** Alternative lixiviants, such as thiosulfate or chloride, can be used instead of cyanide to extract gold from preg-robbing ores (Aylmore & Muir, 2001). Some alternative lixiviants exhibit a lower affinity for carbonaceous matter and clay minerals, thereby reducing preg-robbing and improving gold recovery. However, this affects AC adsorption negatively and an alternative adsorption material may be required.
4. **Addition of competing AC in the CIL process:** When dealing with carbonaceous ores where most of the gold can be directly extracted through cyanidation the method of choice is often the CIL process. However, the effectiveness of the CIL process on preg-robbing ore can vary depending on the strength and quantity of the preg-robbing material (Kotze, et al., 2016). AC is therefore added in large quantities to a CIL circuit to encourage gold cyanide complexes to adsorb onto the AC rather than carbonaceous material in the ore. Known

disadvantages of this method include the need for larger elution and carbon regeneration equipment driving up the cost of the process as well as milling required for large particle size ore driving costs up further (Ryan, et al., 2016).

2.6. Literature Insights

A literature review on the cyanide recovery methods discussed in Sections 2.2 highlights the potential to recover spent cyanide potentially mitigating some operational costs associated with cyanide usage. As mentioned in section 1.1.3 there are two cyanide containing effluent streams on the ERGO plant from which economic value might be obtained. Cyanide recovery from the CIL effluent may prove difficult due to factors such as: low free cyanide concentration, high solids percentage and substantial daily throughput. The policeman column tailings however are a solid-free, clear solution with high free cyanide concentration which could work well under any of the cyanide recovery methods discussed in Chapter 2. Alternatively, utilizing it directly as a leaching process stream without modification owing to its abundant free cyanide concentration has the potential to enhance profitability by eliminating the need for additional processing steps and expensive equipment required for cyanide recovery. The safety risk involved in cyanide recovery processes is also a major drawback and not a viable option for the ERGO plant.

Therefore, a discussion was included on a value-add alternative to cyanide recovery in Section 2.3. Specifically, the focus was on utilizing cyanide-containing effluent for leaching waste gold-bearing material. The literature review, therefore, shifted focus towards leaching methods for coarse gold ores as well as preg-robbing gold ores. In Section 2.4 percolation leaching is presented as an effective method for treating low-grade ore for its cost efficiency in comparison to more complex mining processes. These lower costs may offset the anticipated reduction in gold recovery from the Coarse material due to preg-robbing. Section 2.5 delves into an exploration of preg-robbing and proposes mitigation strategies. The use of high activity AC in a CIL circuit produces a possible alternative to percolation leaching of the Coarse Material but will require milling a large expense for any mining operation.

Chapter 3: Methodology and Equipment

3.1. Materials

3.1.1. Leach Ore Sample

Coarse material was sourced from the ERGO plant reclamation site listed as ERGO Mills. This material contains a significant amount of carbon and is a waste product separated from the feed entering the mills. A 100kg composite sample was collected from the ERGO Mills oversize heap and consisted of a blended mix of samples collected from different areas of the heap to create a representative sample.

3.1.2. Cyanide Containing Effluent

Policeman column effluent was collected from the ERGO Brakpan policeman column exit stream. The stream is a clear solution with 400 mg/L cyanide, which was adjusted to a pH of 12 using NaOH.

3.2. Health, Safety and Environmental Impact of Using Cyanide

When handling cyanide in a laboratory, strict safety regulations are required due to the high toxicity of cyanide. Personal protective equipment, such as gloves, an acid-resistant jacket, and eye protection were worn when working with cyanide compounds. Handling and moving cyanide powders was done with extreme caution to prevent accidental inhalation or contact. All powders were handled in a well-ventilated fume hood to minimize inhalation risks (Thermo Fischer Scientific, 2010). Cyanide solutions were prepared at a pH greater than 10.5 to ensure that no HCN gas is produced. Accidental spills of cyanide into rivers and streams have led to the degradation of fish populations and aquatic biota (Eisler & Wiemeyer, 2004). Therefore, in compliance with the ERGO plant rules and regulations guided by South African regulatory frameworks, cyanide containing components were disposed of and stored in the correct manner to prevent accidental exposure to the surrounding environment. Cyanide containing effluents were correctly labelled for disposal and placed within the necessary pick-up area and all cyanide containing samples were correctly labelled for analysis.

3.3. Experimental Procedures

3.3.1. Procedure to Determine the Particle Size Distribution

A 2 kg representative sample was taken from the 100 kg stockpile and dried in an oven at 105 °C for 24 h. A 500 g sample was weighed to an accuracy of 0.1g and used for particle size analysis in a MACSALAB sieve shaker shown in Figure 3-1. The shaker was stacked with 5.6 mm, 4 mm, 2 mm, 1.7 mm, 1 mm, 850 µm and 300 µm sieves and allowed to shake for 10 min. After sieving the fractions were collected and weighed separately. Similarly for samples of the pulverized Coarse material, the dried material was first pulverised until an 80% percent passing of 300 µm sieve was achieved. The pulverised sample was then placed on the top sieve and shaken for 10 min with 1 mm, 850 µm, 300 µm, 150 µm, 106 µm and 75 µm sieves. After sieving the fractions were collected and weighed separately.



Figure 3-1: Sieve Analysis Setup

3.3.2. Initial Leaching Experiments

2 kg of representative Coarse material was taken from the 100kg stockpile and dried in an oven at 105 °C for 24 h. 1 kg of the dried sample was placed into a 16 L lab-scale Pachuca tank (Figure 3-2) with WHATMAN 320 mm filter paper already situated at its base. 1.6 L of water was treated with CaO until a pH of 11 was reached. The 1.6 L of treated water was separated into 500 ml and 1 L samples and 0.48 and 0.96 grams of NaCN (500 mg/L) added to each sample respectively.

The 1 L solution sample was added to the tank and the produced slurry agitated, adjusting the pH with additional CaO until the slurry reached a pH of 11. When aeration was added, an airline was connected, and the airline opened until a slight bubbling was attained at an airflow rate of 2 L/min. Leaching occurred over 24 hours leaching with sample collection at various intervals. Each time, 100 mL of leachate was collected from the 16-liter column, and 100 mL of the 500 mL sample was transferred to the column. The slurry was filtered, and the filtrate was subjected to titration and AAS for free cyanide and gold analysis respectively. The solid filter product was washed, filtered, and dried overnight before being submitted for fire assay at the MAED Laboratory.



Figure 3-2: 16 L Lab-scale Pachuca Tank

3.3.3. Bottle-Roll Leaching of Coarse Material

A 5 kg sample from the representative 100 kg stockpile was mixed and divided into five 500 g samples. Each 500 g sample was placed in a bottle and 500 mL of water was added. The bottles remained open, as shown in Figure 3-3, for the duration of the experiment. CaO was added until a pH of 11 was reached and this was repeated for all five slurries.

The 5 x 5-liter bottles were rolled for 1 h at 60 rpm to ensure proper mixing of CaO and thereafter additional CaO was added to ensure that a pH of 11 was maintained. 0.48 g of NaCN was added to each bottle and the bottles were returned to the rollers and the slurry leached for varying time intervals (30 minutes, 1 hour, 2 hours, 3 hours, and 5 hours). After leaching, the slurry was filtered at 5 Bars using WHATMAN 320mm filter paper. The filtrate was collected, and the residue solids

were washed with water, filtered, and dried overnight. All solid filter products were labelled and sent for fire assay at the MAED laboratory while all solution samples were labelled and subjected to titration and AAS.



Figure 3-3: Bottles used in Bottle-Roll leaching Experiments.

3.3.4. Bottle-Roll Leaching of Pulverized Coarse Material

3 samples of 3.5 kg Coarse material were collected from the representative stockpile. Each sample was dried and pulverized until an 80% percent passing of 300 μm sieve was achieved. Each 3.5kg sample was then divided into 5 x 500 g samples. Each 500 g sample is placed in bottles (Figure 3.3.) together with 500 mL of water. CaO was added to each bottle until a pH of 11 was reached and then rolled for 1 h at 60 rpm. 0.48 grams of NaCN was added to each bottle and the leaching commenced. The bottles remained open to the atmosphere during experiments simulating CIL conditions included 25 g of activated carbon. The bottles underwent leaching for different time intervals (30 minutes, 1 hour, 2 hours, 3 hours, and 5 hours). Post-leaching, the slurry was filtered at a pressure of 5 Bars using WHATMAN 320 mm filter paper. In cases where carbon was present, the slurry was first screened to remove carbon which was washed and dried. After filtration, the solid residue was washed and dried overnight. The solid filter products were sent for fire assay at the MAED Laboratory while the filtrate was subjected to titration and AAS analysis to determine the free cyanide and gold concentration respectively.

3.3.5. Column Leaching of the Coarse Material

A 10-liter agitated mix tank was prepared by cleaning, sterilizing, and adding policeman column tailings solution and 25 g of activated carbon (AC). Continuous mixing was applied via a propeller at 60 rpm and gentle bubbles formed by attaching an airline at an air flowrate of 1 L/min.

The setup of the percolation leaching column involved using a 16-liter Pachuca tank with WHATMAN 320 mm filter paper already situated at its base (Figure 3.2). A 1 kg sample representative of the Coarse material was loaded into the column and levelled for uniform leaching solution flow. The Pachuca tank was then connected to the agitated mix tank via a peristaltic pump, with the solution pumped at a rate of 0.23 mL/s into the column. The solution exiting the column bottom was returned to the agitated mix tank. The percolation leaching process ran for 4 days to ensure sufficient contact between the Coarse material and leaching solution.

The agitated mix tank was emptied every 24 h, and the AC and solution filtered. The solutions were collected and analysed for their gold content by the MAED laboratory while the AC was washed, dried, and then sent for gold analysis. After the 4 days were complete, the coarse material in the Pachuca tank was washed, filtered, dried overnight, and sent for fire assay by the MAED laboratory.

3.4. Analytical Techniques

3.4.1. Fire Assay

Fire assay was used as the analytical technique for determining the gold content in solid samples. All solid samples were submitted to the MAED laboratory for fire assay. Fire assay at the MAED begins with smelting the sample of gold ore or gold containing activated carbon and uses lead oxide as flux when added to the molten sample of gold-containing ore/activated carbon concentrating the metals to form a lead bead. A cupel is used to remove the lead thereafter where the molten lead mixture is poured and the lead oxidized and absorbed leaving a bead of gold mass that can be compared with the original sample mass to produce a value for the gold concentration in the sample.

3.4.2. Atomic Absorption Spectroscopy

Atomic Absorption Spectroscopy (AAS) was employed as the analytical technique for quantifying the gold content in liquid samples. All liquid samples were submitted to MAED for AAS analysis to determine their gold concentration. A blank sample and standard solutions were prepared by

diluting 1000 mg/L stock gold solution concentrations, to concentrations from 0.001 mg/L to 0.05 mg/L with intervals of 0.005 mg/L starting from 0.005 mg/L up to 0.05 mg/L which were used for calibration. Multiple sets of the same sample were prepared and tested at different calibration sets i.e. 0.001 mg/L to 0.02 mg/L due to gold dissolution being influenced by process conditions, necessitating several attempts to ensure accurate and reliable results.

3.4.3. Titration

The cyanide concentration in solution was determined by carrying out titration with silver nitrate (AgNO_3). A solution with a AgNO_3 concentration of 0.01 mol/L was used as the titrant to react with the cyanide ions present in the sample solution. The titration is based on reaction Eq. 13 (Environmental Protection Agency, 2014).



20 mL of sample solution was placed within an Erlenmeyer Flask. Thereafter, 5 drops of Rhodanine (p-dimethylamino-benzal-rhodanine) indicator was added to the flask to detect the endpoint of the titration, shown by a colour change of the solution from orange to pink.

3.4.4. X-ray Diffraction

X-ray diffraction (XRD) analysis was employed to categorize the minerals present in the Coarse material under investigation. The purpose of the XRD analysis was to determine the mineral phases by utilizing their distinct X-ray diffraction patterns. The XRD analysis was conducted using a Philips PW 1800 X-Ray Diffractometer using Cobalt-K alpha radiation in the 2θ range of 10° - 90° . By subjecting the gold bearing material to XRD analysis, the diffraction patterns obtained from the interaction of X-rays with the mineral lattice structures enabled the identification and categorization of different minerals within the Coarse material.

3.4.5. Dense Media Separation

Dense media separation (DMS) is an analytical method employed for the separation and concentration of minerals or particles based on the difference in their densities and involves the use of a dense medium, typically a dense liquid, which acts as a separating medium (Bosman, 2014). The sample material is introduced into the dense medium and under the influence of gravity the particles settle, or float based on their density. A zinc chloride solution with a specific gravity of 1.55 and 1.65 was used to determine the amount of porous (preg-robbing) carbon present in the Coarse material sample.

Chapter 4: Results and Discussion

4.1. Characterization of ERGO Coarse Material

The Coarse material from the ERGO Mills heap was subject to analysis to understand its characteristics. Proper characterization of this ore is a critical first step in a leaching process. In the case of carbonaceous gold ore material, which are known to be challenging to leach due to the presence of organic carbon, a thorough understanding of the material's properties is important. By examining the Coarse material properties, the aim is to better understand the behaviour of the Coarse material during leaching and identify potential challenges that may need to be addressed to achieve optimal leaching results. The analysis included size distribution (sieve analysis), mineral phases (XRD) and preg-robbing carbon content (DMS).

4.1.1. Gold Grade

The gold grade can affect the rate of reaction and therefore design of a mining operation. At a measured gold grade of 0.65 ± 0.18 g/t, the Coarse material is considered a low-grade ore.

4.1.2. Ore Particle Size Distribution

Particle size distribution is an important factor to consider in the characterization of gold ores for leaching (de Andrade Lima & Hodouin, 2005). The size distribution of particles in the ore affects the accessibility of the ore to leaching solutions, as smaller particles have a larger surface area per unit mass and allow for more interaction with the leach solution. The particle size distribution of the Coarse material was obtained by a sieve analysis analytical method. The results of the sieve analysis are shown in Figure 4-1 (Raw Data: Appendix B, Table B-1). The particle size distribution shows that there is a significant quantity of material that is very large in particle size as most of the mass reports to the large + 5.6 mm sieve range. There was, however, a significant portion of the ore which reported to the sub 300 μm particle size range. This originates from dust and finer material that was coated onto and lodged in between the larger particle-sized material.

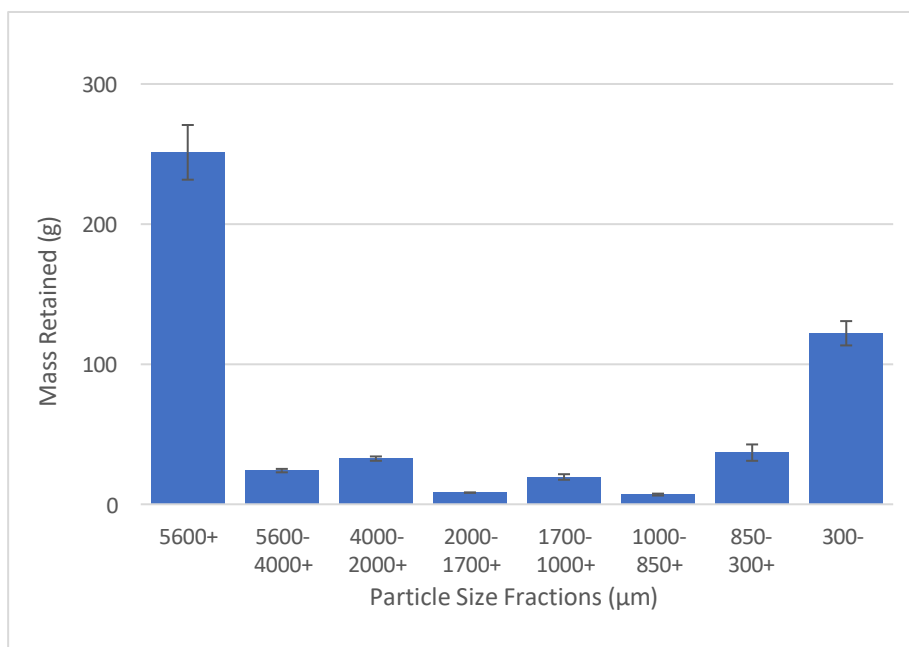


Figure 4-1: Particle Size Distribution for the Coarse Material

The Coarse material exhibited significant variation in size, and it was noted that the composition consisted predominantly of either large rocks or finer particles resembling dust. Due to the large rock sizes, some as big as 15 cm, it was not possible to classify the ore in terms of the percentile values of particle size distribution like d50. Typically, ore of this type is ground by milling to smaller and more uniform particle sizes to produce a larger surface area that helps liberate the target metals to be extracted from the ore in the leaching operation (Wills, 1990). Consequently, the Coarse material was pulverised before use and the particle size determined as reported in Section 4.1.3.

4.1.3. Coarse Material Particle Size Distribution after Pulverization

As part of the investigation into the Coarse material mineral, the substance was subject to pulverisation until 80 percent of the material passed through the 300 µm screen. The particle size of the pulverised Coarse material was determined by sieve analysis - as shown in Figure 4-2 (Raw data: Appendix B, Table B-2). Approximately 80 % of the ore fell within the particle size range of 300 µm and 106 µm. The median size (d50), at approximately 140 µm is indicated by the black line in Figure 4-2. This decrease in particle size usually indicates that better leaching yields can be expected due to the higher surface area of the ore particles and therefore better external mass

transfer of the leaching solution to the gold within the ore. The influence of particle size reduction on leaching was explored in the bottle roll leaching tests.

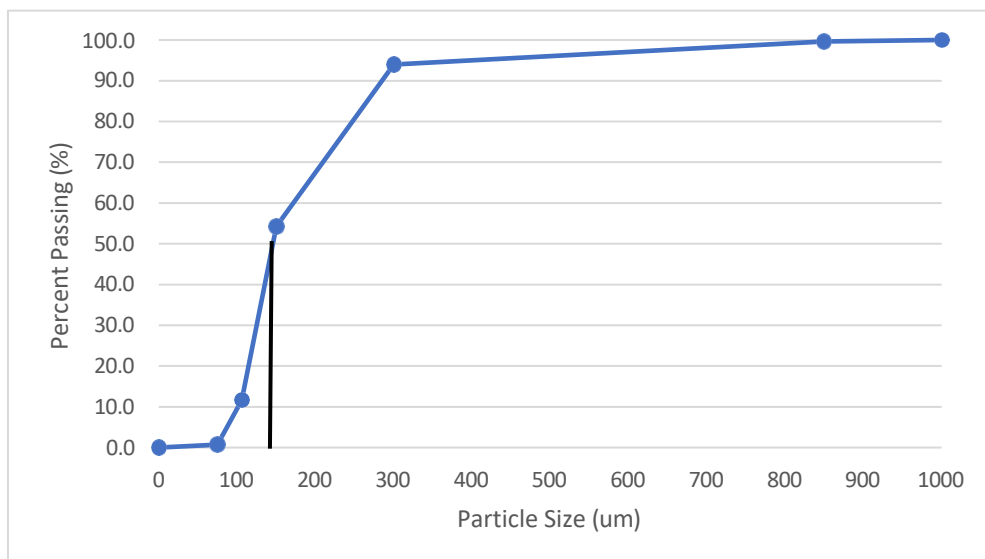


Figure 4-2: Cumulative Percent Passing Particle Size Distribution of Pulverized Coarse material.

4.1.4. Mineralogy by X-ray Diffraction Analysis

An XRD analysis was done to identify the minerals present in the Coarse material. XRD was undertaken at the School of Chemistry of the University of the Witwatersrand and the results can be seen in Figure 4-3. The phase analysis (Software Analysis: Appendix B, Figure B-2) reveals that the Coarse material primarily consists of quartz (SiO_2) and based on these results the gold leaching process for this Coarse material should not necessitate additional pre-leaching steps.

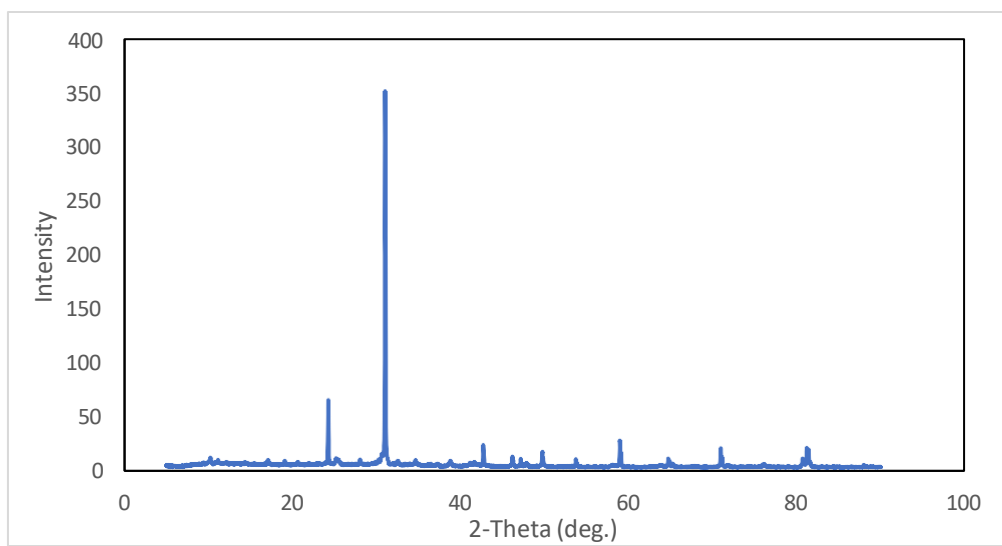


Figure 4-3: X-ray Diffraction Spectrum for the Coarse Material

4.1.5. Analysis of the Carbonaceous Component in the Coarse Material

Table 1-4 indicates that the Coarse material contains carbonaceous material. The presence of carbon within a gold ore has a preg-robbing affect. To characterize the carbonaceous content within the Coarse material, Dense Media Separation (DMS) was used to determine – by separation – the amount of preg-robbing carbon present in the Coarse material, since carbon with the ability to absorb gold has a lower density than silica, the main component of the Coarse material.

In DMS, the material that floats is referred to as the “float material” while the material that sinks is called the “sink material”. DMS is an analytical method that can be used to understand the carbon component of the Coarse material by separating components in the ore of different densities. At ERGO the specific gravity (SG) of the slurry is controlled at 1.45 to ensure the activated carbon (AC) remains in suspension. At densities above a SG of 1.45 the AC will float. Although, carbon has a skeletal SG of 2.2 and would be expected to sink at an SG of 1.45, the porous and hydrophobic nature of AC leads to an ‘apparent’ density in the slurry lower than its skeletal density (Cen, et al., 2013). The activity of AC is directly proportional to its porosity, while its bulk density is inversely proportional to an increase in porosity and thereby its activity (Ganjoo, et al., 2023). Carbon within the Coarse material with greater porosity, and therefore lower apparent density and high affinity for preg-robbing will therefore be separated out.

The DMS was carried out using a zinc chloride solution with a specific gravity of 1.55 to separate and recover preg-robbing material in the sample ore. The float material, which accumulated along the top of the solution was removed and was found to resemble AC in appearance and feel as can be seen in Figure 4-4. Obvious stones were removed and the carbon in the floats were washed dried and weighed. The carbon in the float material came to around 3.8% of the sample mass (Appendix B, Table B-3). However, a true separation was never achieved as there was a great deal of finer solids in suspension at a SG of 1.55, even after 30 minutes. These solids were smaller- sized particles, but similar in feel and look to the carbon in the floats, which were all > 1mm in size.



Figure 4-4: Float Material by DMS at a SG of 1.55

Consequently, another DMS was undertaken at SG of 1.65 where the ore was separated into -1mm and +1 mm particles. A cleaner separation was achieved and 6.2 % of the particles by mass in the +1mm were found to be in the float material, while the size fraction containing the -1mm sized particles had float material consisting of 11.3% by mass. Therefore, using the mass in each size fraction (Appendix B, Table B-4), preg-robbing material is 8.1 % of the sample. An ore is deemed significantly carbonaceous if its preg-robbing carbon content exceeds 1% (Bermeo, et al., 2020) hence, the Coarse material was classified as highly carbonaceous.

4.2. Cyanide Leaching of the ERGO Coarse Material

Along with the Coarse material characterization, the Coarse material was leached by initial batch-mode leach experiments as an understanding and intermediary step before experimentation by bottle roll, CIL, and Column Percolation leaching. Slurries were produced in initial, bottle-roll and CIL leaching experiments under the following conditions: 50 wt.% solids with pulp specific gravity of 1.4, 500 mg/L free cyanide concentration, and pH of 11.

4.2.1. Initial Leaching

Initial leaching experiments were done on unaltered Coarse material with large particles size using fresh cyanide solution (500 mg/L). In the absence of any additional aeration the Coarse material was allowed to interact with the leach solution for 24 hours. In Figure 4-5 The gold extraction into solution has a rapid rise and subsequently decreases before a slow rise and settling at a final gold extraction to solution of 1.3 % (Raw data: Appendix C, Table C-1). From the results it appears as if the gold initially leaches into solution and is then adsorbed with time. The results were expected

as the Coarse material was classified as highly carbonaceous ore, and the presence of carbonaceous material has a preg-robbing effect.

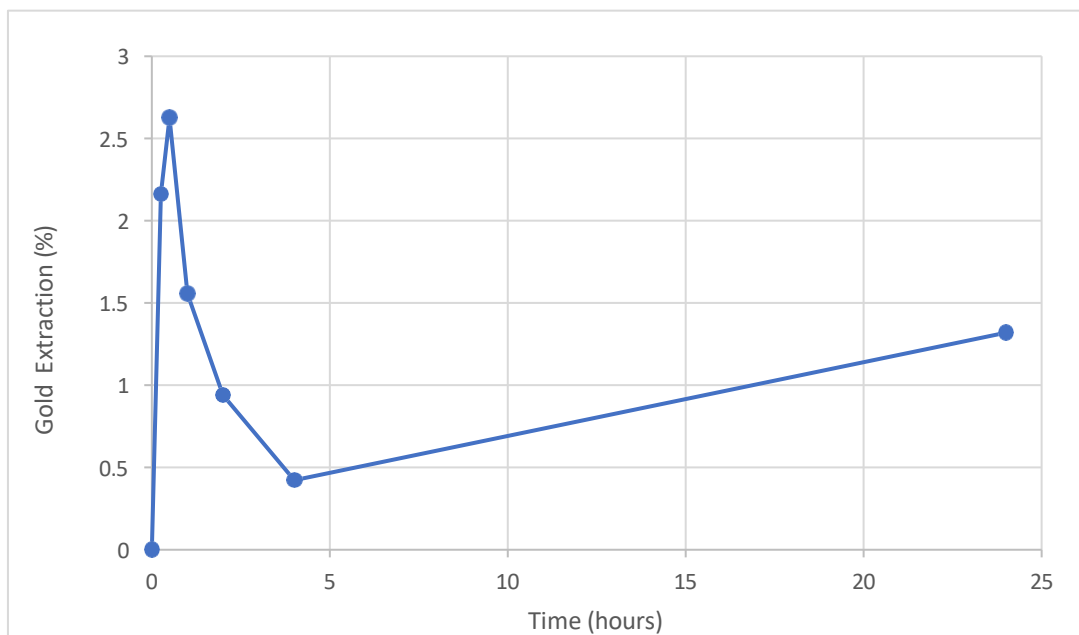


Figure 4-5: Initial Leaching Gold Extraction Percentage Profile ($[\text{CN}^-] = 500 \text{ mg/L}$, $\text{pH} = 11$)

The effect of aeration on gold extraction was also investigated under the same conditions as the initial leaching experiment with air being bubbled through the slurry at 2 L/min. Figure 4-6 shows that adding aeration results in a doubling of the gold extracted to the solution (Raw data: Appendix C, Table C-2). Aeration introduces a degree of agitation but the likely reason for the increase in the rate of the Elsner equation is due to the increase in dissolved oxygen in the leaching solution. However, there is still a great degree of preg-robbing as the gold extraction decreases in a similar manner after roughly 30 minutes of leaching.

The results for both the aerated and non-aerated leaching experiments shows that the extraction into solution is favored by contact times of less than 30 min. Above 30 min the rate of gold adsorption onto the preg-robbing material is higher than the rate of gold dissolution if one is to avoid removing the preg-robbing material from the Coarse material prior to leaching. The leaching process would need to allow a barren solution to enter and react with the Coarse material with a residence time of less than 30 min. The solution can be subjected to adsorption by AC and then be reintroduced to the ore to continue the extraction of gold. This could potentially lead to a decrease in preg-robbing.

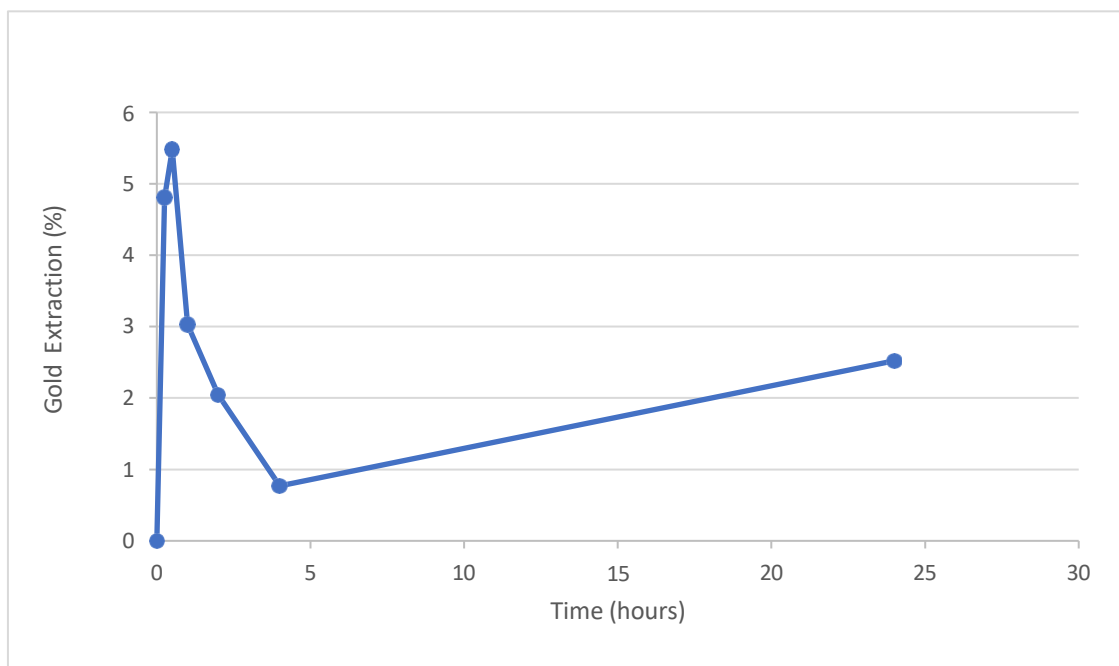


Figure 4-6: Aerated Initial Leaching Gold Extraction Percentage Profile ($[\text{CN}^-] = 500 \text{ mg/L}$, $\text{pH} = 11$, Air Flowrate = 2 L/min)

The initial leaching of the Coarse material indicates the severity of the preg-robbing associated with leaching this ore. The effect of agitation, comminution on gold extraction as well as methods to counteract preg-robbing are investigated in the following sections.

4.2.2. Bottle-Roll Leaching

With an initial understanding of the leaching of the Coarse material bottle-roll leaching was used to add agitation to the experiments and bottles were open allowing for aeration. Agitation by bottle-roll leaching allows for a more rigorous preconditioning before cyanide is added to the slurry and for better external mass transfer kinetics.

Coarse Material Bottle-Roll Leaching

The Coarse material, in its original form from the heap, underwent cyanide leaching using the bottle roll leaching technique without any alterations. The bottle-roll leach test of unaltered Coarse material, with large particles size, was carried out using solution with 500 mg/L free cyanide at a pH of 11. The test involved agitation of the mixture by rolling the slurry at a speed of 60 rpm . By using the same conditions as the initial leach experiment the effect of agitation on the cyanidation of the Coarse material could be determined. Agitation increases the external mass transfer in the leaching system as agitation introduces the ability to suspend solids for better mixing contact

between the leaching solution and solid ore (Wu, et al., 2021). If there is to be a higher level of gold extraction to the solution, there is likely also to be a greater degree of preg-robbing by the carbonaceous material found in the ore. Figure 4-7 shows the results of the bottle leaching test as gold extraction with time (Raw data: Appendix C, Table C-3). From the results one can see a similar trend with the initial leach tests where little extraction occurred, after 5 h the gold in solution was zero. However, there are two distinctions between agitated and unagitated leaching. In unagitated leaching, gold remains in the solution even at the end of the leaching period and continues to increase after the 5 h period. Additionally, it is evident that unagitated leaching exhibits higher extraction of gold in the solution within the same time period, but both experiments show little extraction of gold, particularly at the end of the bottle-roll leach with zero discernible gold dissolution.

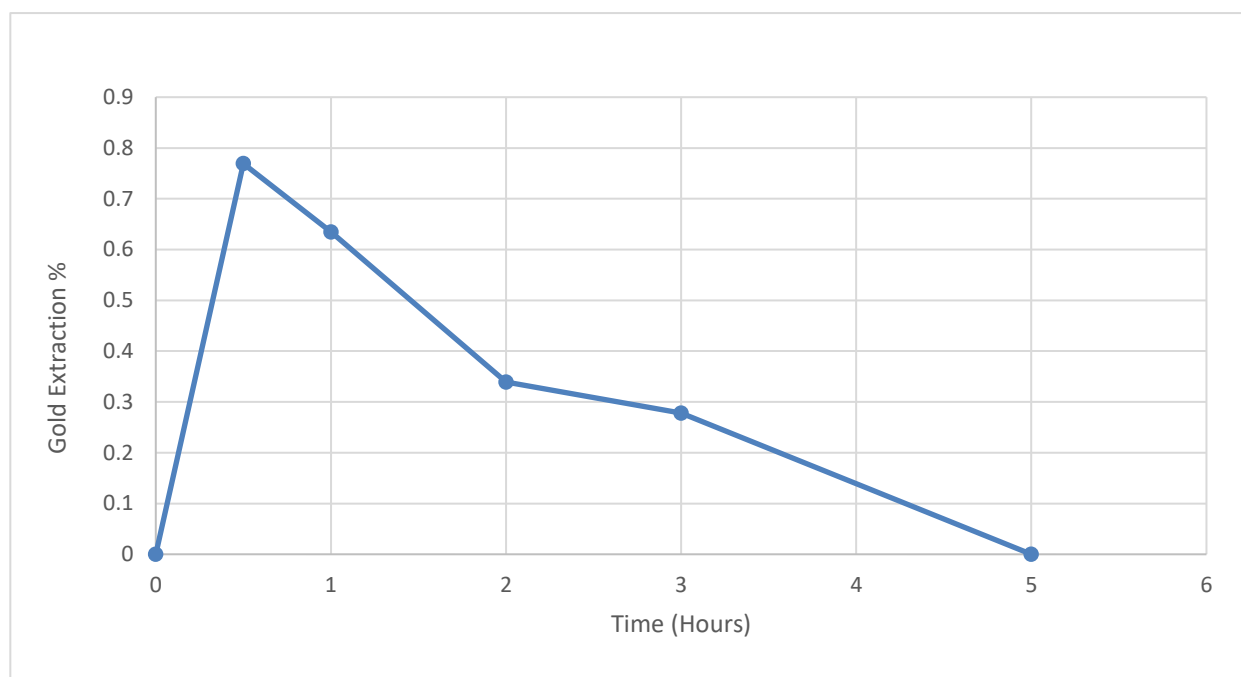


Figure 4-7: Bottle-Roll Leaching of the Coarse Material ($[\text{CN}^-] = 500 \text{ mg/L}$, $\text{pH} = 11$, 60 rpm)

It appears as if agitation increases the negative impact of preg-robbing, and this results in an overall decrease in gold extraction into the solution compared with unagitated leaching, though neither had significant gold extraction to the solution. The decrease in the gold recovery from 2.5% to 1% upon introducing agitation is likely due to an increase in external mass transfer. This enhancement promotes the adsorption of gold from the solution onto the carbonaceous material within the Coarse material. Kyzas et al. (2020) observed a similar effect, reporting that Pb^{2+} adsorption onto activated carbon increased from 94 mg/g to 171 mg/g with agitation, supporting the hypothesis

that agitation enhances $\text{Au}(\text{CN})_2^-$ adsorption onto carbonaceous material.

Pulverised Bottle-Roll Leaching

To investigate the influence of particle size on the leaching process, the Coarse material was pulverised to a d_{50} of 140 μm , resulting in a reduction of the ore particle size. The bottle-roll leach test of pulverised Coarse material was carried out using a leach solution with 500 mg/L free cyanide at a pH of 11 with rolling agitation at 60 rpm. Results from this experiment showed no detectable gold in solution over the leach duration at any time interval (Raw data: Appendix C, Table C-4).

As the particle size decreases, it is expected that the overall kinetics of both the gold dissolution reaction and the adsorption of gold cyanide onto the carbon surface will increase (de Andrade Lima & Hodouin, 2005). Depending on the proportional increase between the rate of dissolution and the rate of adsorption there could be an increase in gold cyanide found in the solution if the rate of dissolution increases in greater proportion compared with the rate of adsorption of gold cyanide onto the carbonaceous material within the Coarse material. However, it is important to consider that the relationship between particle size and the leaching process can have a different outcome. In some cases, the rate of adsorption may increase to a greater extent than the rate of gold dissolution, resulting in a reduced amount of gold being found in the solution. In the case of the leaching of the pulverized Coarse material the rate of adsorption of gold cyanide onto the carbon within the ore shows no extraction of gold. The effect of agitation as well as pulverization provided poor extraction outcomes. Bottle roll tests are designed to simulate leaching in an agitated series tank system. Therefore, if one intends to leach the Coarse material in an agitated series tank system, the carbonaceous material will need to be removed prior to leaching.

4.2.3. Gold Extraction by the CIL Process

With an understanding of the effect of the carbonaceous material present in the Coarse material and the effect of comminution on the Coarse material a decision was made to examine preg-robbing mitigation by addition of competing activated carbon. The CIL process experiments were performed to simulate the conditions of a CIL process and involved using a cyanide solution with 500 mg/L free cyanide and a pH of 11, along with an agitation speed of 60 rpm. Additionally, 60 g of AC per kg of Coarse material was added during the experiments. The carbon samples utilized in this research were sourced from UDEC. The samples, PICAGOLD G209AS (CTC activity 55%) and PICAGOLD G210AS (CTC activity 58%), have an apparent density of 450-550 kg/m^3 and a

moisture content < 5% each.

To determine the effect of the AC in solution, the AC was analyzed for gold content. In Figure 4- 8 (Raw Data: Appendix C, Table C-5, and C-6) the extraction of gold adsorbed onto the PICAGOLD G209AS and PICAGOLD G210AS AC over the leaching period is shown. The carbon becomes loaded with gold and shows an increase in gold extraction over the leaching period. Introducing AC to compete with the carbonaceous materials showed a positive gain in the extraction of gold as it would be possible to separate the AC from the slurry using a linear screen. When attempting to extract gold to the solution in previous experiments, preg-robbing occurred resulting in little extraction to the solution and a resultant loss of gold. With the addition of the high-activity carbon under CIL conditions a final extraction of 17% and 19% were achieved for the G209AS and G210AS AC samples respectively.

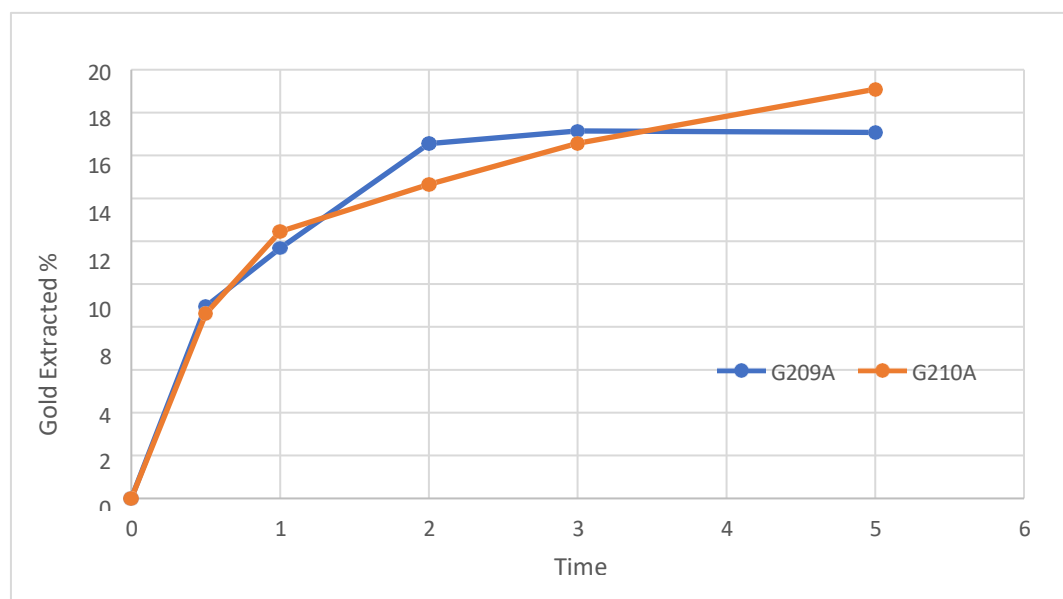


Figure 4-8: Gold Extraction Percentage Profile for Column Percolation Leach Experiment

4.2.4. Re-used Cyanide in Column Percolation Leaching

With an understanding of the characteristics of the Coarse material and the behaviour of the Coarse material under CIL conditions it was decided to try an alternative leaching as a comparison. Column percolation leaching was selected with the aim to increase the understanding of the Coarse Material's behavior in a continuous flow regime, facilitating a more comprehensive extraction efficiency, and the potential for process optimization using a reusable cyanide stream from the ERGO plant. This section presents a detailed investigation into the application of column percolation leaching on Coarse material. Percolation leaching can be advantageous for the leaching

of carbonaceous gold ore due to the continuous flow of the solution through the ore body. The continuous flow nature of percolation leaching facilitates the removal of the gold cyanide complex which prevents the accumulation of the gold cyanide complex within the solution in contact with the ore body (Heinen, et al., 1978). This can potentially decrease the contact time of gold cyanide with carbonaceous material in the ore body thereby avoiding the adsorption of the gold cyanide complex onto the carbonaceous material surface. Initial non-agitated leaching showed that there is a period of 30 min (between 0- 30 minutes) where the trend of the gold extraction to the solution is positive. The absorption of solution by the Coarse material was determined to be 0.27 L by calculating the difference in mass between wetted Coarse material and the filtered mass of the Coarse material. Therefore, a flow rate of 0.23 mL/s was selected as this allowed the damp ore body to become wetted with the cyanide effluent solution while accumulation of the solution in the column was avoided and allowed for a residence time of under 30 min.

The leaching solution was circulated through a bed of 1 kg of Coarse material. The leaching solution was replaced every day with fresh cyanide effluent from the Policeman column. The Policeman column cyanide effluent was collected from the sampler at the exit of the Policeman column and had an average free cyanide concentration of 396 mg/L with an average pH of 12 (Table 4-1).

Table 4-1: Free Cyanide Concentration and pH profile for the Column Percolation Leach

Sample Time (Days)	Leach Solution CN ⁻ Concentration at Day Start (mg/L)	Leach Solution CN ⁻ Concentration at Day End (mg/L)	Solution pH
1	400	160	12.05
2	375	170	11.93
3	395	195	12.00
4	415	330	12.07

The experiment was run for a duration of 4 days (96 hours). The cyanide effluent with a pH of 12.05 and free cyanide concentration of 400 mg/L was passed through the Watson Marlin Dosage pump at 75 rpm producing a stream of 0.23 mL/s onto 1 kg of unaltered Coarse material. The free cyanide concentration and pH of the cyanide effluent added each day can be seen in Table 4-2. Fluid flow only begins when the pores and small channels within the heap are saturated with the

leaching solution which then allows the cyanide effluent stream to seep through the heap (Bouffard & Dixon, 2001). 1.5 L of solution was required on the first day to mitigate the solution absorbed by the ore body with the remaining days requiring 1 L per day. The solution was placed in a mixing tank that included an agitator at 60 rpm. In this mixing tank, a sample of 25 grams of UDEC AC was added, at the beginning of each day to adsorb the gold in solution. This was done to avoid preg-robbing during subsequent recirculation of the leach solution. The purpose of agitation was twofold: first, to enhance the solubility of sodium hydroxide and maintain a pH level above 11, and second, to facilitate the adsorption of gold cyanide onto the surface of AC.

The extraction of gold is almost exclusively found on the surface of the AC as the cyanide gold complex in solution was absorbed by the AC in the feeder mixing tank. Figure 4-9 (Raw data: Appendix C, Table C-7) shows the cumulative amount of gold absorbed onto the AC at different times. At the end of day 1, there was a great deal of extraction of gold from the solution with 0.15 mg of gold found to be adsorbed onto the AC. This represents 14.7% of the gold present in the Coarse material sample. Over the course of the following three days, an additional 0.12 mg of gold was recovered from the Coarse material sample. A cumulative recovery of gold onto the carbon of 26.7 % was achieved over four days. Compared with the extraction of gold by a CIL leaching experiment the much simpler percolation leaching gave better extraction results, but, over a longer period.

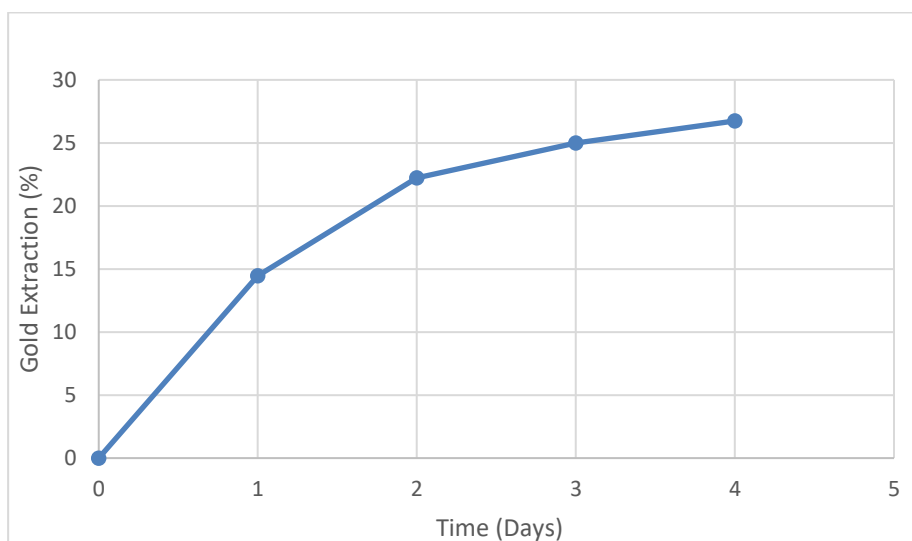


Figure 4-9: The Gold Extraction Profile for Column Percolation Leaching ($[\text{CN}]^- \sim 400 \text{ mg/L}$, $\text{pH} = 12$, Solution Aeration = 1 L/min)

Based on the provided data in Section 4.2 a process plant employing a column percolation leach has been designed. The percolation leach experiment provided a greater extraction of gold than under CIL conditions. A fundamental aspect of a CIL mining operation requires that the AC be recoverable from the slurry solution. Therefore, the processing of the Coarse material necessitates milling, incurring substantial costs attributed to both the significant capital investment and the fact that milling constitutes up to 50% of the total energy consumption in mining operations where this milling is utilized (Petrakis & Komnitsas, 2022). Therefore, the column percolation leaching design is favoured due to likely lower costs as well as higher extraction. A design and economic evaluation have been conducted in Chapter 5 to determine the profitability of reusing cyanide effluent in the treatment of highly carbonaceous gold ore in a column percolation leaching process.

Chapter 5: Economic Feasibility of Gold Cyanidation of the ERGO Plant

Coarse Material with Reused Cyanide Effluent

The cyanidation of highly carbonaceous low-grade gold ore (Coarse material) was shown to be possible, but the results showed low extraction rates of less than 30%. The Coarse material in question was discarded by the ERGO plant and cannot be processed in the CIL circuit feed due to its preg-robbing characteristics. As evidenced in the bottle-roll leach tests, addition of this carbonaceous ore to the CIL circuit would lead to severe preg-robbing. To prevent this, it is proposed that the Coarse material is processed separately to extract value from it. The proposed process makes use of cyanide effluent from the main ERGO CIL plant in a column percolation leach of the carbonaceous ore. This chapter reports on the economic analysis procedure and results of such a process. The analysis followed the framework set out in the textbook Chemical Engineering Design by Gavin Towler and Ray Sinnott (2013).

5.1. Process Design and Description

The process flow diagram is presented in Figure 5-1. The process equipment is denoted on the diagram. This is a batch percolation process with two batches processed simultaneously. The batch process has different steps which include the transportation of the Coarse material and loading into the leaching columns, the batch percolation leaching and the removal of effluent material to the tailings and the carbon to the carbon hopper.

5.1.1. Ore Transportation and Carbon Addition

At the start of the batch percolation process, a leach column is loaded with solid material. In order to transport the material from the material storage dumps the process requires the use of a conveyor belt often used in the field of gold mining. The conveyor belt is represented as B-101. The ore is loaded onto the conveyor using a tractor loader backhoe and is then transported 720 m to the leach columns C-101 and C-102 where it then goes up 2 elevators as part of the conveyor belt. Each column is loaded with 1000 tonnes of Coarse material which is allowed to settle in place.

5.1.2. Column Percolation Leaching Stage

Once the ore is settled within the leach columns, the percolation leaching of the ore can begin. The leaching solution is fed from a storage tank (CS-101 or CS-102) to the columns where it is sprayed onto the solid material and allowed to percolate through the beds circulating back to the storage tank.

Columns CS-101 and CS-102 are used interchangeably for storage of the cyanide containing effluent or for leach solution circulation, while columns C-101 and C-102 are used for leaching in parallel batch mode. When CS-101 is used as a circulation leach feed column, CS-102 is being filled and vice versa. The storage tanks are filled by pumping cyanide containing effluent from the policeman column 350 m away. The cyanide containing effluent is produced at roughly 250 m³ per day in the policeman column, with a free cyanide concentration of approximately 400 mg/L. For a 6-day process, 1 500 m³ of cyanide effluent is produced and available for storage. The batch leaching occurs for 6 days, and the circulation rate is such that it allows for 48 passes of the solution over the Coarse material in a day. Pump P-103 pumps the leach solution to a set of 6 wobbler sprinklers (S-101 and S-102) for each leach tank C-101 and C-102. Air begins to be pumped through the solution column by compressor CO-101. Only the circulation valves are open at the exits of C-101 and C-102, however, the circulating valve V-112 is closed as not to recycle the leaching solution to the non-circulating leach solution storage column. The filters F-101 and F-102 ensure there is no movement of ore out from the columns during the leaching stage. The total cyanide usage for the process is 730 g of free cyanide per ton of ore, 41% of which is provided using cyanide from the 1 500 m³ reused cyanide effluent solution from the policeman column tailings. Therefore, for every leach cycle, an extra 1 622 kg (Appendix E2) of sodium cyanide will need to be purchased.

5.1.3. Tailings Disposal Stage

Once the leaching cycle has ended after 6 days the spent ore and leach solution must be discarded to the tailings dam. Water is pumped from the on-site water storage facility to the circulating solution storage column. The effluent in the circulating solution storage column is concurrently pumped to the leach columns C-101 and C-102. The valves that control the circulation of the solution are closed while the effluent disposal valves are open. This bypasses the filters of the circulation stream allowing the spent ore to be pumped to the tailings distribution box along with the cyanide effluent from the circulating solution storage column.

5.1.4. Carbon Transport

When the solution storage columns CS-101 and CS-102 are empty activated carbon (AC) is inserted into filter drums within columns CS-101 and CS-102. AC remains within the solution

storage columns for 10 leach cycles to ensure adequate loading of gold prior to transport to carbon
hoppers consequently, carbon transport occurs every 20 weeks.

Column Percolation Leaching PFD

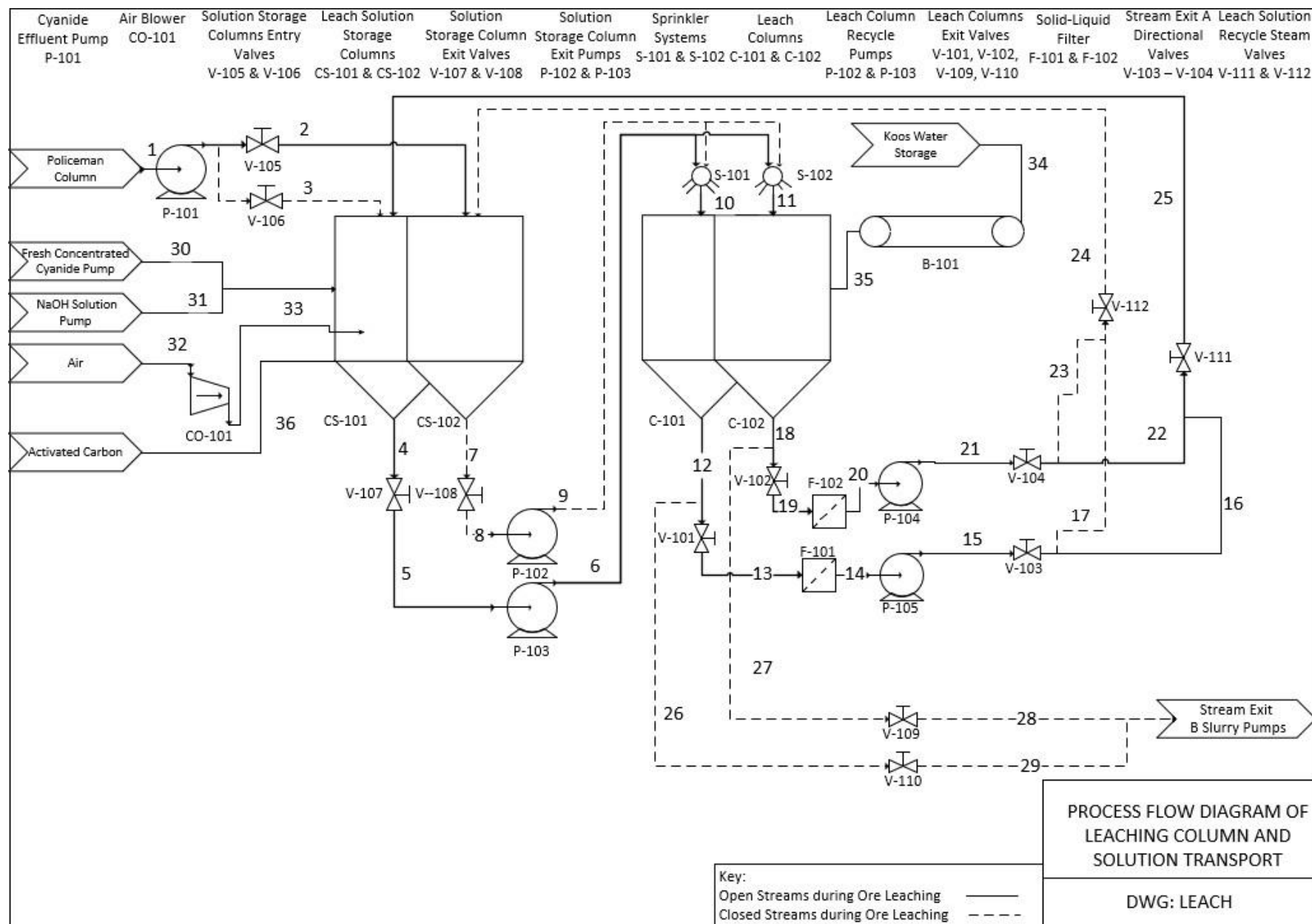


Figure 5-1: Process Flow Diagram for the Column Percolation Leaching Process

5.2. Equipment Specifications

The existing plant has 4 empty leach columns (1500 m³ active volume) available in the old CIP section which will be used as per the design description. These columns do however require retrofitting for usage as percolation columns and storage/circulation columns. While the possible implementation of the percolation leach process at the ERGO plant requires no significant capital investment into the leach and storage columns there will still need to be a significant capital investment into other areas such as pumps, valves, and conveyors.

5.2.1 Column Specifications and Alterations (Appendix D3)

Solution Storage Columns Alterations

The cyanide effluent storage columns require agitation to ensure that there is no crystallization of the NaOH found in the cyanide effluent while also improving the contact between the cyanide solution and the carbon in the column. The cyanide solution in the column will percolate through the ore and return to the storage column. Rather than having an adsorption column between the leach and storage columns, the adsorption will occur within the storage column. A scale top view diagram of the solution storage column is shown in Figure 5-2. The carbon will be kept within 4 filter drums along the walls of the storage column submerged within the cyanide solution. The filter drums use a 500 µm screen around the cylindrical wall and are 30 cm in diameter and 50 cm meter in height. To ensure that there is adequate carbon in the system there will be 15 kg of activated carbon in each of the 4 filter drums. The drums will be placed 90° from each other within the column. At the beginning of each leach cycle, the cyanide storage column will be empty. The final alteration is made to the bottom of the tank by installing an air pipeline to pump air into the solution tank (not shown in Figure 5-2).

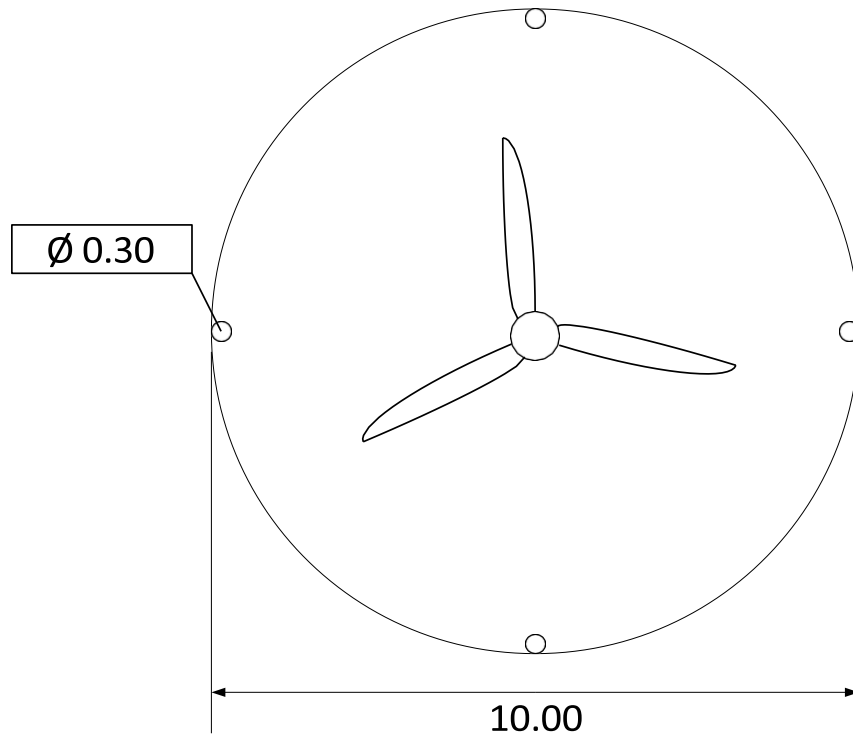


Figure 5-2: Top View Scale diagram of Cyanide Solution Storage Column

Leach Column Alterations

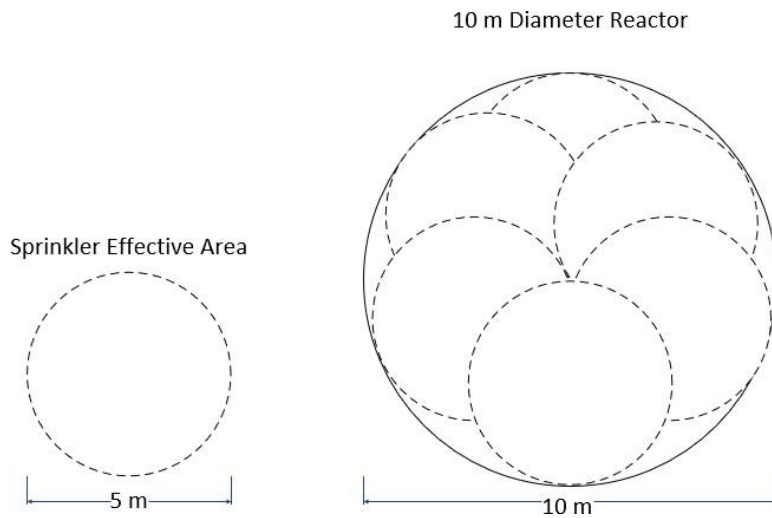


Figure 5-3: Top-Down Scale Diagram of Leach Column Sprinkler Layout

Due to the nature of the leaching process, there will be movement of ore on a much shorter basis than in other percolation leaching processes. Having to remove and re-install the commonly used heap leaching drip system after each leach will extend downtime between leaching cycles therefore, a sprinkler system only covering an area of 78 m² for 1 column will be used A sprinkler

with an effective area of 5 m in diameter leads to only 6 sprinkler heads being needed as displayed to scale in Figure 5-3.

Column Specifications

Table 5-1 lists the columns specification, function as well as alterations.

Table 5-1: List of Column Specifications and Alterations

Displayed Code	Column Task	Column Size (m³)	Additions/Alterations
C-101	Leach Column	1500	Sprinkler System
C-102	Leach Column	1500	Sprinkler System
CS-101	Solution Storage	1500	Agitator, Filter Drums
CS-102	Solution Storage	1500	Agitator, Filter Drums

5.2.2. Pumping Specifications (Appendix D1)

Table 5-2 provides specifications for the pumps, including the number of pumps required, the total power needed, efficiency, and the total rated power.

Table 5-2: List of Pump Requirements and Efficiencies.

Displayed Code	No. of Pumps	Total Power Required (kW)	Efficiency (%)	Total Power Rating (kW)
P-101	1	1.3	64	2
P-102	3	65	66	98
P-103	3	65	66	98
P-104	3	42	68	62
P-105	3	42	68	62
P-106	3	89	64	138
P-107	3	33	73	45
P-108	3	33	73	45

5.2.3. Filter Specifications (Appendix D2)

The filters are crucial to the process and the important design parameters include the material of construction, pore size, and the filter area shown in Table 5-3.

Table 5-3: List of Filters and Filter Specifications

Displayed Code	Material	Filter Pore Size (µm)	Filter Area (m ²)
F-101	316 Stainless Steel	25	10
F-102	316 Stainless Steel	25	10

5.2.4. Valve Specifications

The valves incorporated within the process serve as directional valves enabling precise control over the trajectory of the material leached within the process and are listed in Table 5-4.

Table 5-4: List of Valves and Valve Specifications

Displayed Code	Direction End Point	Matching Size Diameter (cm)
V-101	Stream Exit A to F-101	40
V-102	Stream Exit A to F-102	40
V-103	Leach Solution Storage Valves	20
V-104	Leach Solution Storage Valves	20
V-105	Leach Solution Column, CS 102	7.5
V-106	Leach Solution Column, CS 101	7.5
V-107	Leach Solution Pump, P-107	20
V-108	Leach Solution Pump, P-106	20
V-109	Slurry Pump, P-109	20
V-110	Slurry Pump, P-110	20
V-111	Leach Solution Column, CS-101	20
V-112	Leach Solution Column, CS 102	20
V-113	Tailings Distribution Box	20

5.2.5. Compressor Specifications

The compressor is designed to provide an air flow rate to ensure adequate aeration of the leaching solution with specifications found in Table 5-5.

Table 5-5: List of Air Compressor Specifications

Displayed Code	Description	Blower Capacity (m ³ /h)	Power Required (kW)
CO-101	Aeration Blower	500	27

5.2.6. Sprinkler System Specifications

Table 5-6: List of Sprinkler Specifications

Displayed Code	Associated Column	No. of Sprinkler Heads	Operating Pressure (Bar)
S-101	C-101	6	2
S-102	C-102	6	2

5.2.7. Conveyor Specification

Based on the coarse particle size and the use of percolation as the leaching mechanism the ore will be best transported by use of a conveyor belt (Götz & Wates, 2016). The conveyor belt is used for the transport of the ore with its specifications tabulated in Table 5-7.

Table 5-7: List of Conveyor Specifications

Displayed Code	Description	Conveyer Length (m)	Conveyor Width (m)
B-101	Ore Transport Conveyor	720	0.6

5.3. Economic Analysis of Column Percolation Leaching with Reused Cyanide Effluent at the ERGO Plant

This economic analysis delves into the financial feasibility of employing column percolation leaching with reused cyanide effluent for gold extraction. The cost, benefits, and potential returns associated with the plant will be examined to provide insight into the economic viability of the process.

5.3.1. Estimation of Capital Costs

The equipment costs were estimated using cost curves provided in Tables 7.2 and 7.4 in Chemical Engineering Design (2013) by Gavin Towler and Ray Sinnott and by previous cost estimates provided to DRDGOLD ERGO Brakpan. The total cost of each set of equipment is detailed in Table 5-8 below (Calculations: Appendix E1, Tables E1 – E-6).

Table 5-8: Equipment and Installed Cost for the Column Percolation Leaching Process

Set of Equipment	Cost (ZAR)	Installation Factor	Total Installed Cost (ZAR)
Pumps	4 534 936	4	18 139 744
Valves	132 170	2.5	330 424
Conveyor	19 872 701	2.5	49 681 752
Filters	6 445 182	2.5	16 112 955
Sprinklers	3 941	2.5	9 853
Compressor	363 111	2.5	907 778
Total	31 352 041	-	85 182 505

The largest capital cost relates to the cost of the movement of the ore and leaching solution with the conveyor and pumps forming 79% of the total installed capital cost. The use of the conveyor over hydro reclamation leads to no downtime due to the dewatering of the ore prior to its leaching.

5.3.2. Variable Costs

The variable cost of production will include the cost of raw materials and utilities. Effluent disposal is handled at a large scale by the current ERGO plant and will not be affected by the addition of 4 000 tons per week to the effluent discharge from the plant (M. Gerke, personal communication, October 17, 2023). Water is provided by onsite water storage.

Raw Materials

The Raw material costs are detailed in Table 5-9. Raw Materials consist of sodium cyanide and activated carbon needed for the process. The cyanide provided by using the policeman column effluent is not enough to provide complete leaching of the ore.

Table 5-9: Raw Material Costs of Column Percolation Leaching

Raw Material	Cost per Kilogram (ZAR/kg)	Kilograms per Year (kg)	Cost per Year (ZAR)
Sodium Cyanide	36	84 280	3 034 080
Activated Carbon	50	297	15 600

The use of the cyanide effluent stream provides a saving of R2 116 800 per year (Appendix E2). The effect of this is detailed later in the cyanide costing analysis in Section 5.3.6.

Utilities

The utilities needed for the process are limited to the electricity required to run the process equipment. Water is recycled through the process at the ERGO plant and the plant has a positive water balance therefore the cost of water is negligible to implementing the column percolation leach process at the ERGO plant. The cost and usage of the electrical utility is illustrated in Table 5-10 (Calculations: Appendix E2, Table E-7)

Table 5-10: Utility Cost of Column Percolation Leaching

Utility Expense	Cost per Unit (ZAR/Unit)	Units per Year (Unit)	Cost per Year (ZAR)
Electricity	0.98 (ZAR/kWh)	2 974 271 (kWh)	2 914 786

5.3.3. Fixed Costs

The fixed costs include the increase in labor and supervision, maintenance, and insurance. The fixed costs are listed in Table 5-11 (Calculations: Appendix E3). The land needed for the column percolation leaching process is part of a currently unused section of the plant. The need to pay for land is not necessary (including the property taxes thereof) as the process is an addition to land currently owned by the ERGO plant rather than an entirely new mine area acquisition.

Table 5-11: Yearly Fixed Costs Associated with Column Percolation Leaching

Fixed Cost	Cost per Year
Labor Costs	2 760 000
Maintenance	4 259 125
Insurance	851 825
Interest Payments	3 332 899

5.3.4. Revenues

The primary source of income generated through the column percolation leach process arises exclusively from the precious gold content contained within the Coarse material. With a volatile rand and the possibility of gold prices increasing or decreasing in the coming years determining the price of gold for an accurate estimate is challenging. Therefore, an estimate based on the current gold price as of October 2023 will be used. In October 2023 the average price of gold was above R 1 200 000 per kg (Gold Rate 24, 2023). Based on the current gold price the yearly revenue is

illustrated in Table 5-12.

Table 5-12: Revenue Generation of the Column Percolation Leach Process

Revenue Item	Price per kg of Gold (Gold Rate 24, 2023).	Gold Production per Year (kg)	Revenue per Year (ZAR/year)
Gold	1 200 000	35.1	42 120 000

5.3.5. Economic Evaluation of Column Percolation Leaching with Re-used Cyanide Effluent

Column Percolation Leach Cash Flow Diagram

The cashflow diagram presents a visual representation of the projected cumulative net cash flow throughout the entire lifespan of a project which provides a comprehensive and insightful financial forecast. The cash flow diagram does not include the discounted rate and enables the ability to estimate the project's net cash flow and understand when the investment is expected to generate positive returns. The Cash flow diagram for the column percolation leach process is shown in Figure 5-4.

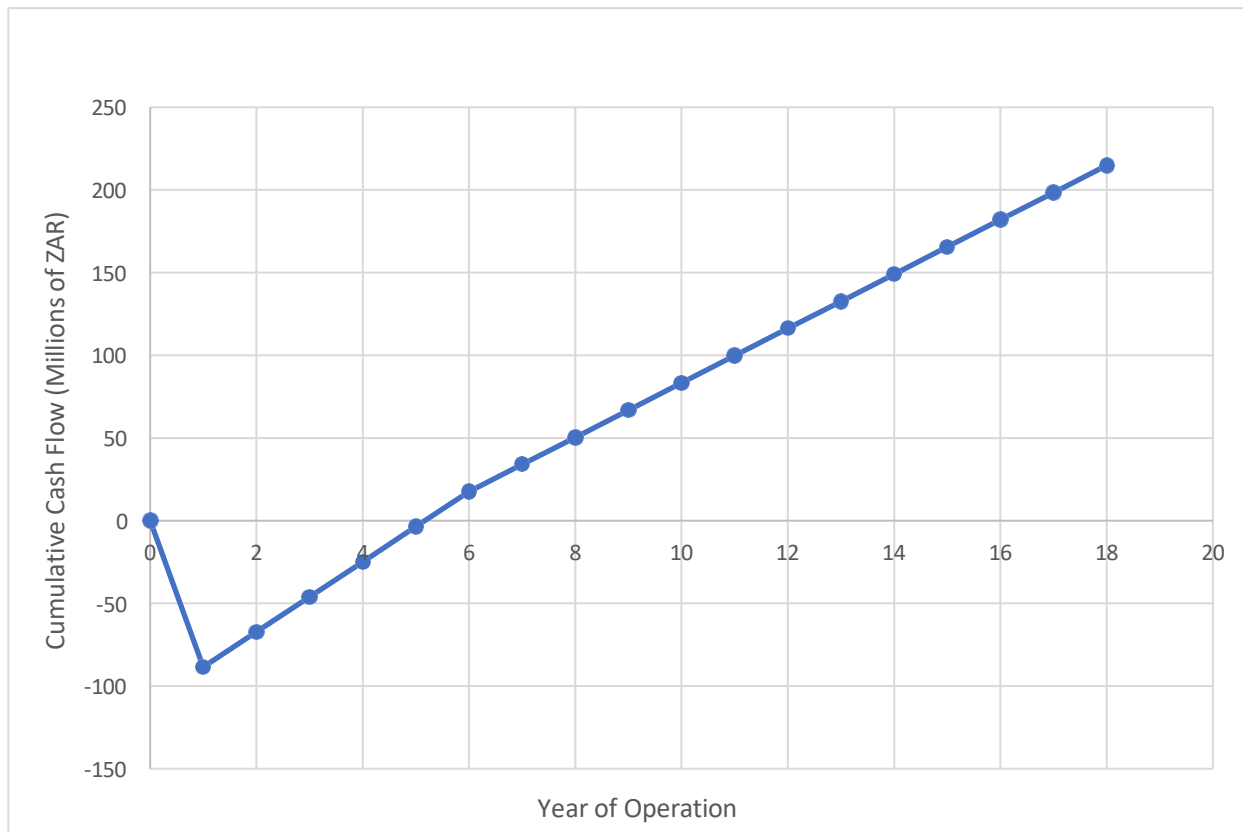


Figure 5-4: Cumulative Cash Flow Diagram for the Life of the Percolation Leach Process

The process cash flow financials reveal a consistently positive cash flow throughout its 18-year lifespan. At the project's conclusion, a cumulative cash flow of R240 million is anticipated.

Investment Options

Determining the preliminary investment attractiveness is done by use of economic measures such as the Payback Time, Return on Investment (ROI), and Net Present Value (NPV). To determine these Table 5-13 was drawn up indicating the yearly financial affairs of the Column Percolation Leaching Process. The table includes the Depreciation, Tax Payable, and net cash flow over a duration of 18 Years.

Table 5-13: Yearly Net Cash Flow for the Column Percolation Leach Process

Year	Gross Profit (ZAR)	Depreciation (ZAR)	Interest Payments (ZAR)	Taxable Income (ZAR)	Tax Payable (ZAR)	Cash Flow (ZAR)
0	0	0	0	0	0	0
1	0	0	3 332 899	-3 332 899	0	-88 515 405
2	28 285 327	17 036 501	3 332 899	7 915 926	2 216 459	22 735 968
3	28 285 327	17 036 501	3 332 899	7 915 926	2 216 459	22 735 968
4	28 285 327	17 036 501	3 332 899	7 915 926	2 216 459	22 735 968
5	28 285 327	17 036 501	3 332 899	7 915 926	2 216 459	22 735 968
6	28 285 327	17 036 501	3 332 899	7 915 926	2 216 459	22 735 968
7	28 285 327	0	3 332 899	24 952 427	6 986 680	17 965 748
8	28 285 327	0	3 332 899	24 952 427	6 986 680	17 965 748
9	28 285 327	0	3 332 899	24 952 427	6 986 680	17 965 748
10	28 285 327	0	3 332 899	24 952 427	6 986 680	17 965 748
11	28 285 327	0	3 332 899	24 952 427	6 986 680	17 965 748
12	28 285 327	0	3 332 899	24 952 427	6 986 680	17 965 748
13	28 285 327	0	3 332 899	24 952 427	6 986 680	17 965 748
14	28 285 327	0	3 332 899	24 952 427	6 986 680	17 965 748
15	28 285 327	0	3 332 899	24 952 427	6 986 680	17 965 748
16	28 285 327	0	3 332 899	24 952 427	6 986 680	17 965 748
17	28 285 327	0	3 332 899	24 952 427	6 986 680	17 965 748

18	28 285 327	0	3 332 899	24 952 427	6 986 680	17 965 748
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Table 5-13 lists the yearly net cash flow from year 0 to year 18 when all the available carbonaceous material has been leached. It can be said that there is a positive trend in the cumulative cash flow over the life of the plant. The first economic indicator on plant profitability is the payback period which is calculated as the total investment divided by the average annual cash flow. Using column 7 of Table 5-13 the average yearly net cash flow is R13 375 189. With a total capital investment of R85 182 505 the payback period is 6.4 years. The payback period gives insight into the possible time frame before a return on the investment into the process might become possible. This does not indicate the returns financiers/investors can expect from the process. Utilizing the ROI provides a more precise gauge in this context and quantifies the potential gains that investors can anticipate. The ROI is calculated by dividing the cumulative annual profit by the total investment multiplied by the plant life (Towler & Sinnott, 2013). The cumulative net profit equates to R240 753 408. This therefore equates to an ROI of 15.7% (Appendix E4).

The NPV serves as an alternative economic metric offering a definitive figure to determine whether an investment will meet the necessary return criteria. The NPV can yield either a negative, positive, or zero value when calculated. The investment proposal should only be considered if it has a positive NPV (L. Van Dyk, personal communication, 2021). The NPV is calculated by the formula Eq. 14 (Towler & Sinnott, 2013):

$$NPV = \sum_{n=1}^{n=t} \frac{C_n}{(1+i)^n} \quad (14)$$

Where C is the cash flow in year n , t is the project life in years and i is the cost of capital in percent.

The NPV was calculated with a Cost of Capital percentage of 18.4 % (Raw data: Appendix E4, Table E-8). The NPV at the end of the column percolation leach process is R1 881 877, a positive value, indicating that the proposal is acceptable for investment while a negative value would indicate the proposal is unacceptable for investment.

5.3.6. Financial Effect of Reused Cyanide Effluent

Earlier there was a reference to the annual expenses associated with cyanide usage and the yearly cost reductions resulting from the recycling of cyanide effluent. The reuse of cyanide effluent allows for a saving of R2 116 800 per year. This represents a saving of 13.3% of the yearly

operating expenses of the process. This saving decreases the share of the operating expenses that would be spent on sodium cyanide from 32.3% to 21.9% (Calculations: Appendix E5, Table E- 9). A 13% reduction in operating expenses is noteworthy with substantial implications on the cost of the plant. For comparison, the process financials will be scrutinized under the same indicators established in the section. Table 5-14 (Calculations: Appendix E5) provides a comparative representation of the financial indicators highlighting the differences between the scenarios. That which involved the reuse of cyanide effluent and that which relied solely on fresh cyanide solutions.

Table 5-14: Financial Indicator Comparison (Column Percolation Leach Process with and without Reused Cyanide Effluent)

Financial Indicator	With Reused Cyanide Effluent	Without Reused Cyanide Effluent
Payback Period	6.4 years	7.1 years
ROI	15.7 %	14.0%
NPV	R1 881 877	- R4 726 805

There is a decrease in the cash flow of the process due to the need to purchase fresh cyanide which affects all the financial indicators negatively. However, if one only looks at the payback period and the ROI the process may still seem viable. It is only when looking at the NPV that the effect of re-using unaltered cyanide effluent can have on the plant. The NPV goes from a positive to a negative when the cyanide costs need to be increased indicating that the process without reused cyanide is no longer an attractive investment option.

5.3.7. Effect of the Column Percolation Leach with Reused Cyanide Effluent on Profitability at the ERGO plant.

The ERGO plant was described both in an operational and financial capacity in Section 1.1. On a full-year scale, for the year ending on 30 June 2022, revenue at ERGO amounted to R3.7 billion and a profit after tax of R530.9 million (Appendix E6, Table E-10). Using the after-tax net cash flow from the Column Percolation Process as an addition to the ERGO plant profits the expected increase is 3.3 % for the remainder of the process after the payback period of 6.4 years based on the reporting figures for the financial year 1 July 2021 to 30 June 2022.

Chapter 6: Conclusions and Recommendations

The following bullet points summarize the key findings and recommendations about cyanide reuse at the ERGO plant, providing insight into the effect on the profitability of cyanide reuse in the leaching of carbonaceous ore:

- Cyanide recovery methods overwhelmingly require acidification and thereby HCN production as a process step. Due to safety concerns at the ERGO plant, cyanide recovery by these methods is not viable for implementation at the ERGO plant. However, cyanide from the Policeman column cyanide containing effluent stream can be reused as leaching solution/process water for leaching waste gold-bearing material.
- On-site heaps at the ERGO plant contain waste gold bearing material referred to as Coarse material. The Coarse material from the ERGO plant has a wide particle size distribution, and a high carbon content that causes preg-robbing during leaching. Experimental work demonstrated that leaching of the Coarse material exhibits a significant preg-robbing effect as was observed in initial leaching tests and the subsequent experiments involving agitation, aeration, and comminution. Aeration was the sole leaching modification that led to an enhancement in gold extraction to the solution, with no extraction observed in cases involving agitation and comminution.
- A traditional method of mitigating preg-robbing was experimentally tested by the addition of competing activated carbon during leaching as in CIL like conditions. The CIL simulation demonstrated improved gold extraction rates of 17% to 19% for different activity level AC samples.
- Column Percolation leaching using cyanide effluent was tested to determine the efficacy of a continuous percolating extraction process with in-tank adsorption and aeration. The results indicate a gold extraction of 26.74% at a liquid residence time of 28 min. When tested as a method to mitigate preg-robbing the column percolation process demonstrates an increase in gold extraction compared with the use of competing activated carbon under CIL-like conditions.
- A preliminary plant was designed based on a column percolation leach process. At a gold price of R 1 200 000/kg an estimated yearly revenue of R42 120 000 can be achieved. The

capital cost of the process is estimated at R85 182 505 with yearly operating costs at R13 834 673.

- An economic evaluation by Payback Period, ROI, and NPV (18.4 % discount rate) of the full-scale process indicates that the process is profitable with values of 6.4 years, 15.7%, and R1 881 877 respectively.
- If the extraction of gold does not utilise the reuse of cyanide effluent but uses only fresh cyanide (additional cost R2 116 800 per year) to generate revenue from the Coarse material, this result in a negative NPV of - R4 726 805, making it a non-profitable process.
- The ERGO plant can achieve an increase in profitability of 3.3% (based on profits for the financial year 1 July 2021 – 30 June 2022) by incorporating the reuse of cyanide in a column percolation leach process.

The following recommendations for future research with regards to reusing cyanide and cyanide recovery effect on the profitability of a gold mine.

- The column percolation leach process is relatively simple and the use of some preg robbing mitigation methods such as kerosene addition could be explored to determine if this leads to greater extraction of gold.
- There is potential to use thickeners or clarifiers on the CIL slurry cyanide-containing effluent and produce a larger volume of cyanide effluent for use in larger projects.
- During the design of the process, it was assumed that carbon loading will occur over a 10-week period to allow sufficient gold loading for elution. The 10-week period was calculated based on the loadings observed during the percolation experiments which ran over 4 days. To verify the carbon loading capacity, it is proposed to study the carbon loading characteristics over a period of 10 weeks to determine if other contaminant metals are being absorbed over multiple leach cycles.
- A 1 m³ per day circulating leach solution pilot plant based on the column percolation leach process should be designed and commissioned and undergo multiple leach cycles to determine the large-scale performance of the process before full-scale implementation.

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Appendices

Appendix A: The ERGO Plant Overview

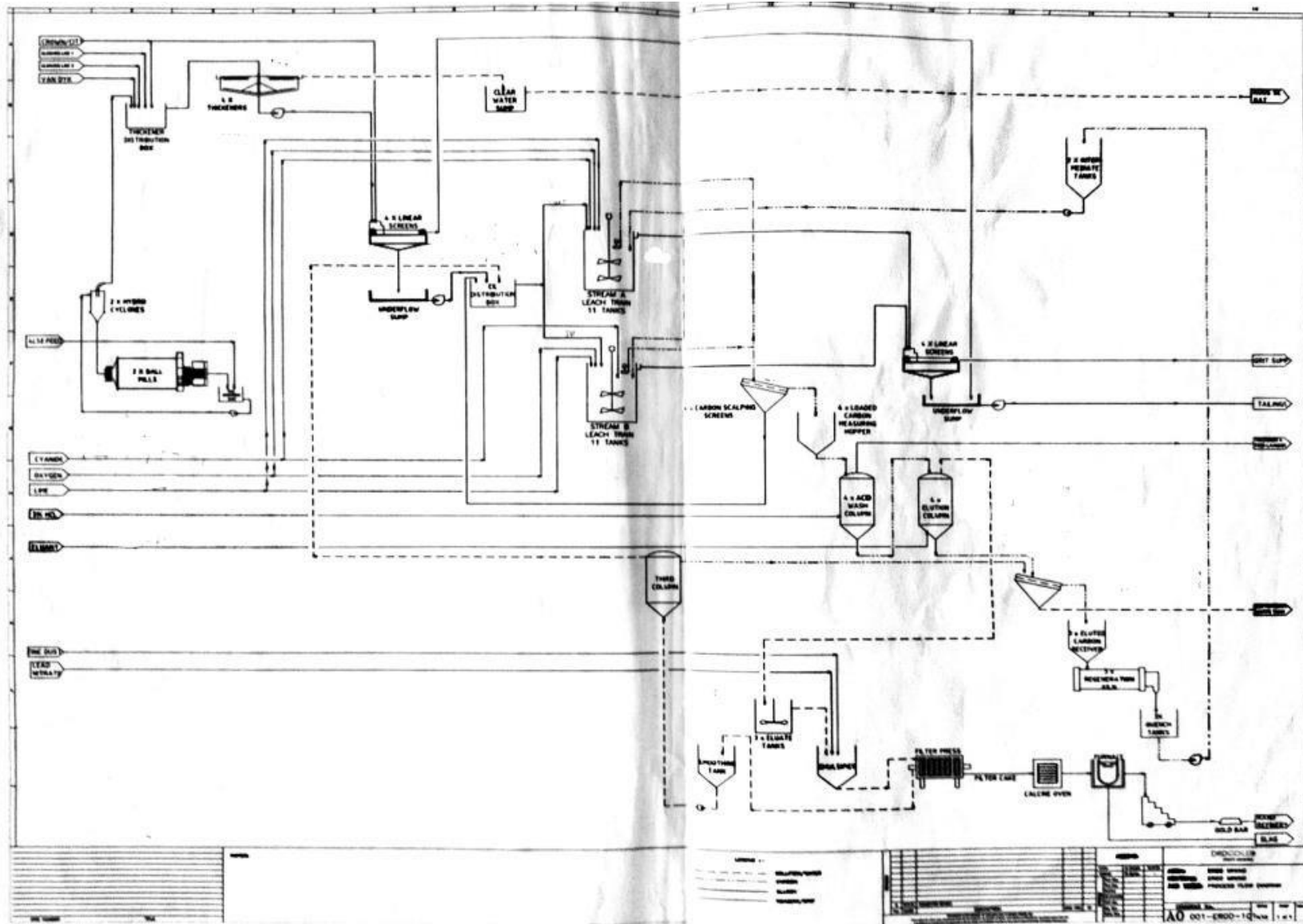


Figure A-1: The ERGO Plant Process Overview PFD

Table A-1 summarizes the overall reagent use at the ERGO Plant for the months July - December 2021 and is presented below. The data is extracted from the monthly ERGO Plant Reagents spreadsheets for the months July to December 2021.

Table A-1: Tonnage of Reagent Usage at the ERGO plant

	July	August	September	October	November	December	TOTAL
Cyanide	355	381	416	456	472	402	2482
Lime	2919	3420	3283	2486	2010	2620	16738
Oxygen	1101	1081	979	1742	1027	1200	7130
Caustic Soda	129	142	147	144	139	149	850
HCL	281	244	222	231	258	240	1476
SenFlocc	12.25	14.6	11.5	10.5	9.6	13.65	72.1
Carbon	51	43.8	54.6	54	48.6	60	312

Prices of each reagent were provided by conversation with Arshir Narain, ERGO Metallurgical Staff (2022). Figure A-2 provides an excerpt of page 7 from the Condensed Consolidated Unaudited interim Results for the six months ended 31 December 2021 for DRDGOLD LIMITED used in section 1.1.2.

8. OPERATING SEGMENTS (continued)

	Six months ended 31 Dec 2021 Unaudited			
	Ergo Rm	FWGR Rm	Corporate office and other reconciling items Rm	Total Rm
Revenue (external)	1 804.7	693.8	–	2 498.5
Cash operating costs	(1 471.8)	(208.4)	–	(1 680.2)
Movement in gold in process	20.0	(6.3)	–	13.7
Operating profit	352.9	479.1	–	832.0
Administration expenses and other costs	(3.9)	(1.2)	(76.9)	(82.0)
Interest income ¹	0.6	–	56.4	57.0
Dividends received	3.9	–	37.1	41.0
Interest expense ²	(1.9)	(0.2)	(2.2)	(4.3)
Current tax	(27.6)	(110.5)	(3.6)	(141.7)
Working profit before additions to property, plant and equipment	324.0	367.2	10.8	702.0
Additions to property, plant and equipment	(86.4)	(96.1)	(0.1)	(182.6)
Working profit after additions to property, plant and equipment	237.6	271.1	10.7	519.4

Figure A-2: Excerpt of Overview of DRDGOLD LIMITED Financial Position

Appendix B: Ore Characterization

Particle Size Distribution of Oversize Coarse material, Raw Values: 1. Sieve Analysis

Table B-1 Tabulated Data for the Particle Size Distribution of the Coarse material

Size Fraction (μm)	Sample 1	Sample 2	Sample 3	Average	Standard Deviation
5600+	278.60	196.20	278.80	251.20	19.52
5600- 4000+	14.10	35.70	22.30	24.03	1.23
4000- 2000+	21.60	45.80	30.40	32.60	1.56
2000- 1700+	6.30	10.40	8.30	8.33	0.02
1700- 1000+	15.80	26.00	16.70	19.50	1.98
1000- 850+	5.00	7.90	7.90	6.93	0.68
850- 300+	36.50	45.50	28.60	36.87	5.85
300-	121.30	135.10	109.80	122.07	8.67

Particle Size Distribution of Pulverized Coarse material, Raw Values: 1. Sieve Analysis

Table B-2: Tabulated Data for the Particle Size Distribution of Pulverized Coarse material

Size Fraction (μm)	Sample 1	Sample 2	Sample 3	Average	Standard Deviation
1000+	0	0	0	0.0	0.0
1000- 850+	1.9	2.3	2	2.1	0.2
850- 300+	31.5	28.3	29.8	29.9	1.6
300- 150+	221.6	201.6	211.5	211.6	10.0
150- 106+	211.1	243.1	227	227.1	16.0
106- 75+	66.1	50.7	58.3	58.4	7.7
75- 0+	4.4	3	3.6	3.7	0.7

X-Ray Diffraction Analysis Spectrum Analysis:

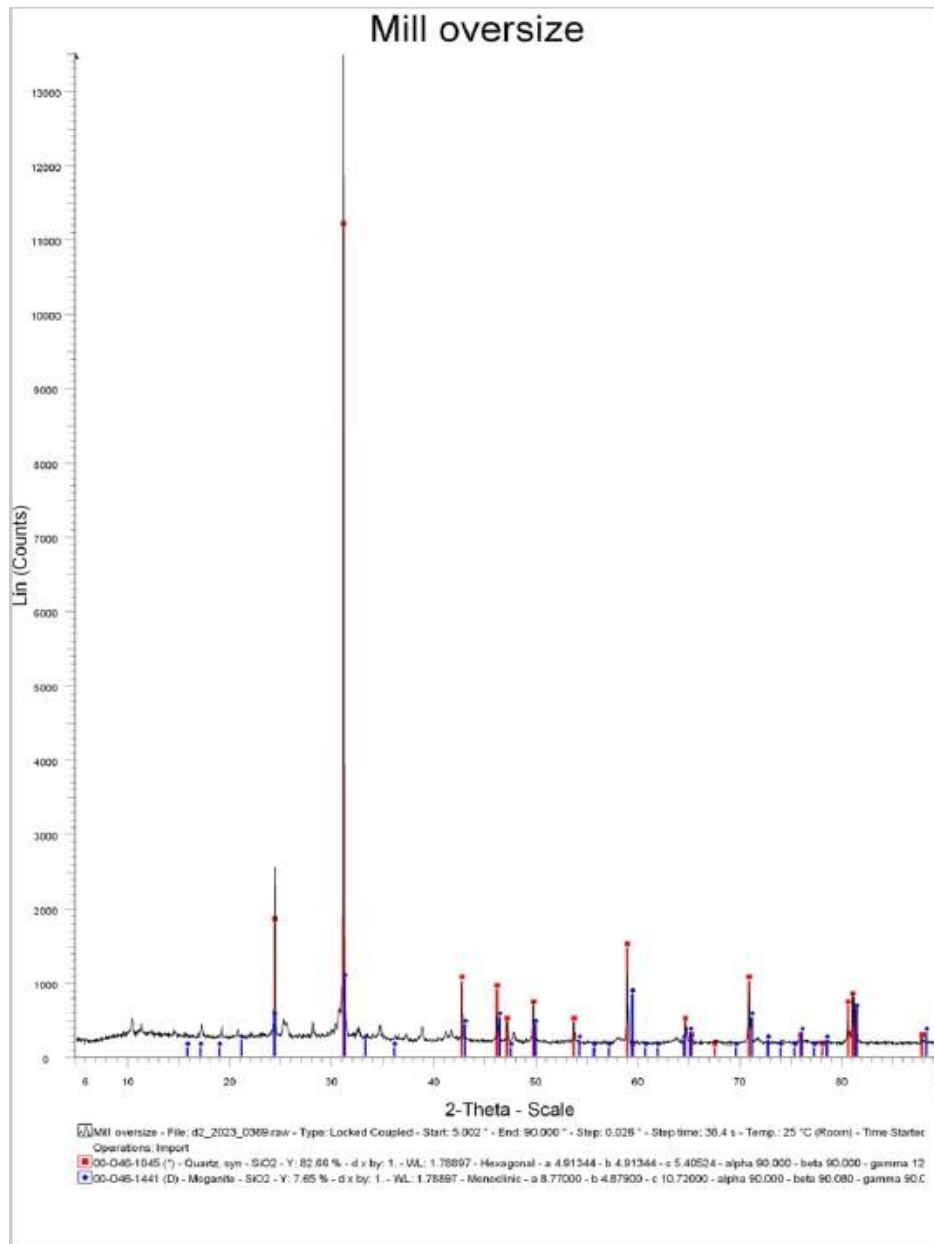


Figure B-1: X-ray Diffraction Spectrum Analysis for the Coarse material

Dense Media Separation:

Table B-3: Dense Media Separation at SG = 1.55

First DMS Analysis Parameters	
Ore Mass	504 grams

Specific gravity	1.55
Dried Floats Mass	19.0 grams
Floats %	3.8

Table B-4: Dense Media Separation of Size Separated Ore at SG = 1.65

Second DMS Analysis Parameters	
Ore Mass	492 grams
Specific gravity	1.65
+1 mm	298 grams
-1 mm	192 grams
Dried Floats Mass +1 mm	18.5 grams
Dred Floats Mass -1 mm	21.8 grams
Floats % +1 mm	6.2
Floats % -1 mm	11.3
Total Floats Percentage	8.1

Appendix C: Leaching Results and Discussion

24 Hour Results and Leach Conditions

Table C-1: Leach Conditions and Yield for 24 Hour Initial Leach

Sample time (h)	Gold Soln. Conc. (mg/L)	CN (mg/L)
0	0	520
0.25	0.009	275
0.5	0.0115	175
1	0.0072	140
2	0.0046	120
4	0.0022	110
24	0.00732	100

24 Hour Aerated Leach Results and Conditions:

Table C-2: Leach Conditions and Yield for 24 Hour Aerated Initial Leach

Sample time (hr)	Gold Soln. Conc. (mg/l)	CN (mg/l)
0	0	530
0.25	0.02	130
0.5	0.024	35
1	0.014	25
2	0.01	20
4	0.004	20
24	0.014	0

Bottle-roll Leaching.

Table C-3: Bottle Roll Leach of Unaltered Coarse Material

Sample time (hr)	Head CN- (mg/l)	Tailings CN- (mg/l)	Gold Soln. Conc. (mg/L)
0.5	500	367.5	0.005
1	505	255	0.004
2	495	222.5	0.002
3	500	187.5	0.002
5	510	177.5	0

Table C-4: Bottle Roll Leach of Pulverized Coarse Material

Sample time (hr)	Head CN- (mg/l)	Tailings CN- (mg/l)	Gold Soln. Conc. (mg/L)
0.5	500	292.5	0.000
1	510	252.5	0.000
2	500	230	0.000
3	500	212.5	0.000
5	495	207.5	0

CIL leaching with PICAGOLD G209A competing Activated Carbon

Table C-5: CIL Leach of Pulverized Coarse Material with added G209A Activated Carbon

Sample time (hr)	Head CN-	Tail CN-	Soln. Gold Conc. (mg/L)	Gold Conc. On Carbon (g/t)
0.5	500	265	0.001	1
1	495	225	0.001	1.2
2	495	170	0.000	2
3	510	185	0.000	2
5	500	165	0.001	2.3

CIL Leaching with PICAGOLD G210A competing Activated Carbon

Table C-6: Bottle Roll Leach of Pulverized Coarse Material with added G210A Activated Carbon

Sample time (h)	Head CN-	Tail CN-	Soln. Gold Conc. (mg/L)	Gold Conc. On Carbon (g/t)
0.5	495	285	0.001	1.1
1	500	225	0.003	1.8
2	500	170	0.001	1.9
3	500	165	0.001	2.3
5	510	155	0.001	2.5

Column Percolation Leach results Data

Table C-7: Column Percolation Process 4-day Leach Data

Sample time (h)	Gold Conc. Head Solution (mg/L)	Gold Conc. Tails Solution (mg/L)	Gold Conc. On Carbon (g/tonne of carbon)	Gold Mass on Carbon (mg)
0	0.006	-	-	-
1	0.003	0.002	6	0.15
2	0.002	0.003	3.1	0.0775
3	0.001	0.002	1.1	0.0275
4	-	0.002	0.7	0.0175

Appendix D: Equipment Specifications

Appendix D1: Determination of Pump Power and Efficiency

The formula for calculating the power required is in the equation below:

$$P_{kW} = \frac{Q\rho gH}{3.6 \cdot 10^6}$$

P_{kW} represents the power required of the pump in kW, Q is the flow rate in m^3/h , ρ is the density(kg/m^3) of the fluid to be pumped, g is the gravitational force in m/s^2 and H the differential head in m. Table D-1 represents the calculations for the power required to pump the specified fluid with gravitational constant at 9.81. The Friction Head was predominantly calculated using the piping sizes listed in column 3, matching size diameter, in table 6-3 as these matches the piping diameters of each stream/pipeline.

Table D-1: Pump Power Requirement Calculations

Pump	Power Required (kW)	Flow rate (m^3/h)	Density (kg/m^3)	Differential Head (m)	Discharge Pressure (Bar)
P-101	1.3	20.82	1000	23.36	1
P-102	65	540	1000	44.41	2
P-103	65	540	1000	44.41	2
P-104	41.95	540	1000	28.51	0.5
P-105	41.95	540	1000	28.51	0.5
P-106	88.56	500	1000	65.01	1
P-107	32.6	500	1400	17.09	1
P-108	32.6	500	1400	17.09	1

Efficiency was estimated using figure 20.14 from Chemical Engineering Design: Principles, Practice and Economics of Plant and Process Design (2013) by Gavin Towler and Ray Sinnott. An extract is displayed in figure D-6. For pumps with greater capacity than shown have been divided into pump stations to determine their pump efficiency by figure D-6

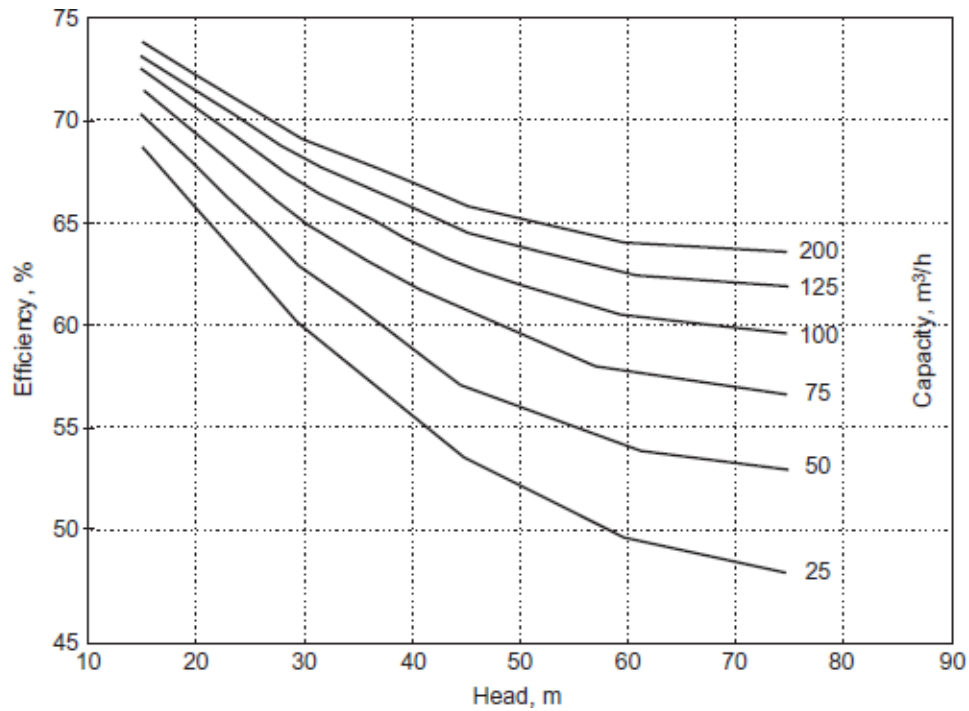


Figure D-1: Pump Efficiency Curves for Estimation of Pump Efficiency (Towler & Sinnott, 2013)

Appendix D2: Filter Sizing

Sizing the filters requires the use of a gravity(strainer) filter for dewatering and leach solution flow in the remining and leaching stages. The use of a steel plate filter with 25 µm pores would halt the ore while allowing the water/leach solution to flow freely out of the column and as a gravity filter suitable for a multitude of conditions (Towler & Sinnott, 2013). The filter works on displacement washing as the leach solution is passed through the ore bed.

Filter specification for the estimation of the filter is provided from the extract of Pail's PSS sintered metal powder media below (Purchas & Sutherland, 2002).

The best possible media grade would be media grade E as this will grade will be more than sufficient for halting ore particles greater than 25 µm. The permeability of media grade F to water is 1.13. With the permeability of the filter known it is possible to calculate the necessary filter area for the system by use of Darcy's Law:

$$Q = -\frac{kA}{\mu L} \Delta p$$

Where Q is the flow rate in m³/s, A is the Area in m², k is the permeability in m², µ is the dynamic viscosity in Pa.s and Δp is the differential pressure loss over the media.

Filter Area Calculation:

Dynamic Viscosity of water(25°C): 8.9×10^{-4} Pa.s

Length: 0.0016 m

Minimal Pressure Drop due to gravity filtration: 20 000 Pa.

Flow rate: $540 \text{ m}^3/\text{hour} = 0.15 \text{ m}^3/\text{s}$

Permeability = $1.13 / 1.013 \times 10^{12} = 1.11 \times 10^{-12}$

Area = 9.62 m^2

Table 7.8 Pall's PSS sintered metal powder media

Media grade	Form as produced	Micron removal rating				Nominal standard thickness (mm)	Permeability ^e		
		Liquid service ^a					to air	to water	
		$\beta=2$ (50%)	$\beta=10$ (90%)	$\beta=100$ (99%)	absolute ^c (100%)				
						Gas service ^b absolute ^c (100%)			
P05	Sheets	0.5	2	3	5	0.3	1.3	5.3	0.07
P09	Sheets	2	4	7	9	0.5	1.3	7.0	0.22
H	Sheets	5	7	9	13	1	1.6 ^d	17.7	0.26
F	Sheets	8	12	15	20	3	1.6 ^d	53	1.13
E	Sheets	15	22	25	35	11	1.6 ^d	98	3.7
D	Sheets	20	28	40	55	20	1.6 ^d	340	8.7
H150	Discs	—	6.5	9	15	—	2	77	—
H250	Discs	—	10.5	18	25	—	2	218	—
H550	Discs	—	33	54	55	—	2	285	—
S050	Cylinders	0.5	2	3	5	0.4	1.6 ^d	9	0.11
S100	Cylinders	4	7	8	10	0.8	1.6 ^d	28	0.28
S200	Cylinders	7	10	14	20	2.8	1.6 ^d	147	1.48
S350	Cylinders	13	17	24	35	11	1.6 ^d	442	5.90

^a Using AC dusts in water, efficiency measured by particle count.

^b At air flow velocities of 3–5 m/min.

^c Absolute ratings based on particle count.

^d 3.1 mm thick also available as standard.

^e l/dm²/min @ 10 mbar pressure drop.

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Figure D-2: Pall Corporation Woven Metal Sheets Specifications (Purchas & Sutherland, 2002)

Appendix D3: Column Alterations

Cyanide Containing effluent Storage Column:

Filter Drums as explained in 6.1.4.1, is a drum with an 800 μm screen around the cylindrical wall and are 3 dm in diameter and 5 dm meter in height. A filter of this size can be made on site. As part of the columns the previously installed agitator motors will be used in the solution storage column, while the removed agitator impellers placed back in position. The motors are rated at 50 kW and were used for slurry agitation for a prior leaching process. The leach solution is predominately water; thus, the previous agitator system will be more than sufficient. Wobbler

sprinkler's work in a multitude of pressures from 0.7 bar to 2 bar. Pumps P-106 and P-107 are designed to discharge to the sprinklers at pressure of 2 bar (Senninger, 2023).

Appendix E: Economic and Profitability Analysis

Appendix E1: Capital Expenditure Estimation

The capital cost of the Filters, Compressor and Conveyor were determined from Table 7.2 in Chemical Engineering Design (2013) by Gavin Towler and Ray Sinnott using a CEPCI of 532.9 for 2010 and 803.6 March 2023 together with the Cost Curve formula below. It was also assumed an exchange rate average of 19 rand to the dollar as of October 2023.

$$C_e = a + bS^n$$

Valves and pumps were estimated by ERGO cost estimation and vendor pricing (M. Gerke, 2023) while Sprinklers were determined by online resellers.

Capital cost of the pumps:

The average Rand to USD conversion was R19 per US dollar as of October 2023

Table E-1: Pumping Equipment Cost Estimation

Pump	Cost per Pump, 2019 (ZAR)	Cost per Pump, 2023 (ZAR)	Total Pump Cost, 2023 (ZAR)
P-101	68 290	90 333.90	90 333.90
P-102	160 000	211647.74	634 943.21
P-103	160 000	211647.74	634 943.21
P-104	160 000	211647.74	634 943.21
P-105	160 000	211647.74	634 943.21
P-106	160 000	211647.74	634 943.21
P-107	160 000	211647.74	634 943.21
P-108	160 000	211647.74	634 943.21
Total	-	-	4 534 936.37

Capital Cost of the filters:

Table E-2: Filter Equipment Cost Estimation

Filter	Filter Cost (USD)	Filter Cost (ZAR)
F-101	169 610	3 222 591
F-102	169 610	3 222 591

Vender Information was Available for all the size valves, but the information was from 2019. The CEPCI used for 2019 and March 2023 were 607.5 and 806.6 respectfully.

Capital Cost of the Valves:

Table E-3: Valve Cost Estimation

Valve	Cost (ZAR)
V-101	19 224.67
V-102	19 224.67
V-103	9 612.33
V-104	9 612.33
V-105	3 604.63
V-106	3 604.63
V-107	9 612.33
V-108	9 612.33
V-109	9 612.33
V-110	9 612.33
V-111	9 612.33
V-112	9 612.33
V-113	9 612.33

Capital Cost of the Compressor:

Table E-4: Blower Capital Cost Estimation

Compressor	Blower Cost, 2010 (USD)	Blower Cost, 2023 (USD)	Blower Cost, 2023 (ZAR)
CO-101	12673.39	19111.16	363 111.99

Capital Cost of the Sprinklers:

The sprinklers vendor information showed a cost of 27.46 AUD per sprinkler head (The Irrigation Shop, 2023). The exchange rate used as of October 2023 was 11.96 ZAR per AUD.

Table E-5: Sprinkler System Cost Estimation

Sprinkler System	Cost/Sprinkler Head (ZAR)	Total Cost (ZAR)
S-101	328	1 970.53
S-102	328	1 970.53

Capital Cost of the Conveyor:

Table E-6: Complete Conveyor Capital Cost Estimation

Conveyor B-101 Item	Total Cost, 2010 (USD)	Total Cost, 2023 (USD)	Total Cost, 2023 (ZAR)
Conveyor	607 600	916 246	17 408 669
Elevator	86 000	129 686	2 464 032
Total Conveyor Cost	693 600	1 045 932	19 872 701

Appendix E2: Variable Costs:

Raw Material Costs

The raw material costs are limited to sodium cyanide and activated carbon. The calculations for the usage listed in table 6-9 are detailed below.

Cyanide:

Cyanide Usage per ton of ore = 730 g CN⁻/tonne of ore x 2 000 tonnes = 1 460 000-gram CN⁻

Sodium Cyanide Usage per ton of ore = $1\,460\,000 \times 49/26 = 2\,751\,538.46$ -gram NaCN = 2 751.54 kg of NaCN per Leach.

At R36/kg, Cost per year = $2751.54 \times 52 \times 36 = \text{R}5\,150\,880$

CN⁻ provided by reused Cyanide = $400 \text{ g CN}^-/\text{m}^3 \times 1500 = 600$ kilograms CN-

Sodium Cyanide Provided by reused Cyanide = $600 \times 49/26 = 1130$ kg of NaCN

Saving provided = $36 \times 1130.76 \times 52 = \text{R}2\,116\,800$

Therefore, yearly Sodium Cyanide Cost = $5\,150\,880 - 2\,116\,800 = \text{R}3\,034\,080$

Activated Carbon

The leach solution storage columns CS-101 and CS-102 both carry 60 kg of carbon for a total of 120 kg. 120 kg lasts 21 weeks. Therefore, carbon usage over a year is equal to $120 \times 52/21 = 297.14$ kg per year.

Utilities

Electricity:

The cost of electricity afforded to ERGO Brakpan is R0.98 per kWh (M. Gerke, personal communication, October 25, 2023).

Table E-7: Electricity demands of Essential Equipment

Draw	Power (kW)	Usage per Week (hr)	kWh per Week	kWh per Year
Conveyor	554	24	13 302	691 679
P-101	2	144	288	14 976
P-102	98	72	7 056	366 912
P-103	98	72	7 056	366 912
P-104	62	72	4 464	232 128
P-105	62	72	4 464	232 128
P-106	138	10	1 380	71 760
P-107	45	10	450	23 400
P-108	45	10	450	23 400
Compressor	27	144	3 888	202 176

Agitators	100	144	14 400	748 800
Total	-	-	42 797	2 974 271

Appendix E3: Fixed Costs

Labor:

The operating labor is calculated at a cost of 1 operator at R230 000 per year (Talent.com, 2023).

The number of additional operators needed to be added to the process are calculated using the Figure 8.12 from Chemical Engineering Design by Gavin Towler and Ray Sinnott (2013). There are 3 new operating sections due to the column percolation leach. For every solid handling section an extra operator is required. Therefore 3 extra operators will be required each shift at ERGO. Accounting for a 4-shift rotation this is a total of 12 extra operators. At R230 000 on average per year this equates to a cost of R 2 760 000 per year. Since the process is a plant addition, plant supervision is already in place over the plant and should not add additional cost.

Maintenance and Insurance

The maintenance is estimated at 5% of the installed capital cost (Towler & Sinnott, 2013). This equates to R4 259 125 per year.

The Insurance is calculated as 1 percent of installed capital cost and is therefore R851 825 per year.

Interest:

Assuming a 50% split on loan financing and on ERGO cash investment the total annualized interest at 11.75% per year (Trading Economics, 2023) over the life of the plant (18 years) comes to: R59 992 190.82 of total interest. This equates to an annualized interest cost of R3 332 899 per year.

Appendix E4: Determination of Cash Flow

Gross Profit is the difference between yearly revenue and the yearly operating costs. Therefore, this equates to $R42\,120\,000 - 13\,834\,673 = 28\,285\,327$

Depreciation is determined by the straight-line method over 5 years with an assumption that the salvage value of all the equipment is 0 at the end of the project. Therefore, the yearly depreciation is calculated at $R85\,182\,505/5 = R\,17\,036\,501$ per year.

Taxable Income is calculated yearly at 28 % and depends on whether deductions from depreciation is available. Thereafter the cash flow for each year is calculated as the difference between the Gross Profit – Interest Payments – Tax. The cash flow for each year is shown in table 6-13(section 6.4.5.).

ROI calculation:

Cumulative Net Profit = 240 753 408

Capital Investment = 85 182 505

Plant life = 18 years

ROI = $240\,753\,408 / (85\,182\,505 \times 18) \times 100 = 15.7\%$

NPV calculations:

Table E-8: Net Present Value over the lifetime of the Column Percolation Leaching Process

Year	Discounted Cashflow (ZAR)	NPV (ZAR)
0	0	0
1	-88 515 405	-88 515 405
2	16 225 327	-72 290 078
3	13 706 718	-58 583 360
4	11 579 065	-47 004 295
5	9 781 681	-37 222 614
6	8 263 300	-28 959 315
7	5 516 014	-23 443 301
8	4 659 779	-18 783 521
9	3 936 456	-14 847 066
10	3 325 411	-11 521 654
11	2 809 218	-8 712 437
12	2 373 151	-6 339 285
13	2 004 774	-4 334 511
14	1 693 579	-2 640 933
15	1 430 690	-1 210 243

16	1 208 608	-1 635
17	1 020 999	1 019 364
18	862 513	1 881 877

The ERGO cash investment return is assumed at 25% (Towler & Sinnott, 2013) with the loan interest set at 11.75%. Therefore, using the cost of capital formula (Towler & Sinnott, 2013) the cost of capital with a 50% loan and 50% ERGO cash investment = $0.5 \times 0.1175 + 0.5 \times 0.25 = 0.184$

Appendix E5: Effect of Reused Cyanide

Yearly Operating Costs without Reused Cyanide = R15 951 473.09

Yearly Operating Costs with Reused Cyanide = R13 834 673.09

Yearly cost of cyanide with reused cyanide = R3 034 080

Yearly cost of cyanide without reused cyanide = R5 150 880

Cyanide Cost % with Reused Cyanide = $R3\,034\,080 / R13\,834\,673.09 = 21.9\%$

Cyanide Cost % without Reused Cyanide = $R5\,150\,880 / R15\,951\,473.09 = 32.3\%$

Table E-9: Yearly Net Cash Flow for the Column Percolation Leaching without Reused Cyanide Effluent

Year	Gross Profit (ZAR)	Depreciation (ZAR)	Interest Payments (ZAR)	Taxable Income (ZAR)	Tax Payable (ZAR)	Cash Flow (ZAR)
1	0	0	3 332 899	-3 332 899	0	-88 515 405
2	26 168 527	17 036 501	3 332 899	5 799 126	1 623 755	21 211 872
3	26 168 527	17 036 501	3 332 899	5 799 126	1 623 755	21 211 872
4	26 168 527	17 036 501	3 332 899	5 799 126	1 623 755	21 211 872
5	26 168 527	17 036 501	3 332 899	5 799 126	1 623 755	21 211 872
6	26 168 527	17 036 501	3 332 899	5 799 126	1 623 755	21 211 872
7	26 168 527	0	3 332 899	22 835 627	6 393 976	16 441 652

8	26 168 527	0	3 332 899	22 835 627	6 393 976	16 441 652
9	26 168 527	0	3 332 899	22 835 627	6 393 976	16 441 652
10	26 168 527	0	3 332 899	22 835 627	6 393 976	16 441 652
11	26 168 527	0	3 332 899	22 835 627	6 393 976	16 441 652
12	26 168 527	0	3 332 899	22 835 627	6 393 976	16 441 652
13	26 168 527	0	3 332 899	22 835 627	6 393 976	16 441 652
14	26 168 527	0	3 332 899	22 835 627	6 393 976	16 441 652
15	26 168 527	0	3 332 899	22 835 627	6 393 976	16 441 652
16	26 168 527	0	3 332 899	22 835 627	6 393 976	16 441 652
17	26 168 527	0	3 332 899	22 835 627	6 393 976	16 441 652
18	26 168 527	0	3 332 899	22 835 627	6 393 976	16 441 652

Appendix E6: Effect on Profitability at ERGO Brakpan

Table E-10: Possible Profitability Increase for ERGO Brakpan incorporating the Column Percolation Leach Process with Reused Cyanide.

ERGO June 2022 Profit (ZAR)	Additional Profit (ZAR)	Increase in Profit (%)
530 900 000	17 965 748	3.3

Adapted from Annual Financial Statements 30 June 2022 by DRDGOLD Ltd.