# EVALUATION OF PROCESSING OPTIONS FOR THE TREATMENT OF ZINC SULPHIDE CONCENTRATES AT SKORPION ZINC

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by

**Herman Franz Fuls**

Study Leader: Dr. J Petersen

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## ABSTRACT

Skorpion Zinc, an integrated zinc mine and refinery, located near Rosh Pinah in southern Namibia, produced its first metal in May 2003. The refinery has a production capacity of 150 000 tpa of special high grade (SHG) zinc (> 99.995 % Zn). The Skorpion zinc oxide resource will be depleted by end of 2015 with a possible extension to 2017. Extensive exploration drilling for additional zinc oxides was conducted, without success. With the abundant availability of zinc sulphide concentrates regionally (Black Mountain and Gamsberg in the Northern Cape or Rosh Pinah Zinc and new deposits in the Rosh Pinah region) the life of the operation may be extended by processing the zinc sulphide concentrates.

The leach kinetics of zinc sulphides is vastly slower than oxides, preventing the processing of zinc sulphides using the existing process, under current conditions. The challenge for Skorpion Zinc is to identify a suitable zinc sulphide treatment process that can be integrated with the existing plant. The selection of a treatment process option for the zinc sulphides will have to consider the feasibility, cost and benefits of converting the existing refinery. Various issues such as the impact on production, time to convert, capital investment and benefits will be considered.

The approach taken to identify a suitable process for the zinc sulphide concentrates included the following:

- Identifying various zinc sulphide processing options through an extensive literature search and elimination of the non-viable options. This step also considered the various zinc sulphide sources, inclusive of their contained impurities.
- Considered process flows and integration with the Skorpion Zinc refinery by the development of a block flow diagram for each option plus the construction of a mass & energy balance.
- Capital and operating cost estimates for the selected process options.
- Identification of a suitable process.

The primary criteria against which the process options were evaluated included:

a) the development of the process to commercial scale;

b) the ability to produce zinc by 2016-2017;

c) a favourable economic outcome (low capital and operating cost with positive returns);

d) a process that does not produce sulphuric acid. Due to the remote location of Skorpion Zinc, transport costs preclude the sulphuric acid from being competitively introduced into the market.

The pyrometallurgical processes and electrolytic (Roast-Leach-Electrowinning) process have been eliminated as they produce sulphuric acid as by-product. Hydrometallurgical processes in chloride media and processes involving other oxidants other than ferric in a sulphide media have been discarded based on a lack of development beyond the laboratory scale.

Hydrometallurgical processes that involve ferric leaching in sulphate media (pressure leach, atmospheric leach and the Albion process) are favoured and were further developed to establish their economic viability. This category also includes the stirred tank bioleach processes, which were rejected due to the diluted sulphuric acid produced and the, low zinc tenors which require the introduction of a high operating cost solvent extraction step. Bio heap leach processes, typically applied to low grade materials were also rejected based on the lack of industrial development and the short time frame available to have this process ready for production at Skorpion Zinc.

Mass and energy balances were developed for the Dynatec pressure leach, atmospheric leach and Albion processes – the information was used to establish an operating cost model plus used to evaluate the integration of each process with the existing Skorpion infrastructure. Capital costs were estimated by sizing and costing the mechanical equipment and factoring these to derive a total capital cost.

A phased approach where oxides and sulphides are processed in parallel was considered in order to evaluate the benefits of sulphuric acid production from the sulphide ore. This option is not financially viable as the savings in sulphuric acid plus the high processing cost of sulphide concentrates does not justify the capital expenditure of a parallel process, prior to the depletion of the oxide ore reserve.

The table below presents the relative operating and capital cost of the three most attractive options.

| *Cost as % of Option C total cost* | **OPTION C** | **OPTION D** | **OPTION F** |
|---|---|---|---|
| | **Atm Leach** | **Press Leach** | **Albion** |
| **Operating Cost** | **100.0** | **92.2** | **101.5** |
| **Capital Cost** | **100.0** | **122.4** | **101.0** |

The analysis indicates that the pressure leach has a lower operating cost to the other options. The operating cost comparison between these two options is highly sensitive to the process chemistry with regards to redox potential (ratio of ferric to ferrous), acidity of leach stages and the precipitation of iron during the leach process. The estimate of additional milling costs for the Albion process indicates it has a marginal increase in capital and operating costs and hence was discarded as an option.

This study concludes that ferric leaching of sphalerite in a sulphate medium under atmospheric or pressure leach conditions are both feasible and economically viable processes for Skorpion Zinc. These processes present advantages over all other zinc processing options when elemental sulphur is preferred as the final deportment for sulphide sulphur. Both these processes can be integrated into the existing refinery and will allow the processing of concentrates with a wide composition range.

In order to select the optimum process for Skorpion, it is recommended to undertake a bench scale atmospheric leach test and to engage with Outotec (technology supplier) in order to establish firmer numbers on the process parameters for the atmospheric leach. Reaction kinetics under the various leach conditions for atmospheric leach needs to be established.

Further test work on a laboratory scale is also recommended in order to establish the sensitivity of recovery and residence time on the particle size distribution for both processes. This should be conducted for a sphalerite that contains a high and low quantity of iron in the sphalerite mineral matrix. Information from the bench scale test work and an update of the economic models will provide sufficient accuracy for Skorpion to select either the pressure or the atmospheric leach process. A single option can then be considered and evaluated during the next phase of project development and implementation.

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# 1 INTRODUCTION

## 1.1 BACKGROUND AND FORMULATION OF RESEARCH PROBLEM

Skorpion Zinc, an integrated zinc mine and refinery, located near Rosh Pinah in southern Namibia (Figure 1), produced its first metal in May 2003. The refinery has a production capacity of 150 000 tpa of special high grade (SHG) zinc (> 99.995 % Zn). The ore body, a substantial oxide resource averaging 10.9 % Zn, was discovered in 1976. The main minerals are sauconite (a zinc-bearing clay mineral), smithsonite (zinc carbonate), and hemimorphite (zinc silicate). The elevated silica levels and the presence of halides and challenges to upgrade the zinc ore resulted in a delay in the development of a processing route that includes direct leaching of the ore, purification via a solvent extraction process and conventional electrowinning to produce the final zinc product on site.

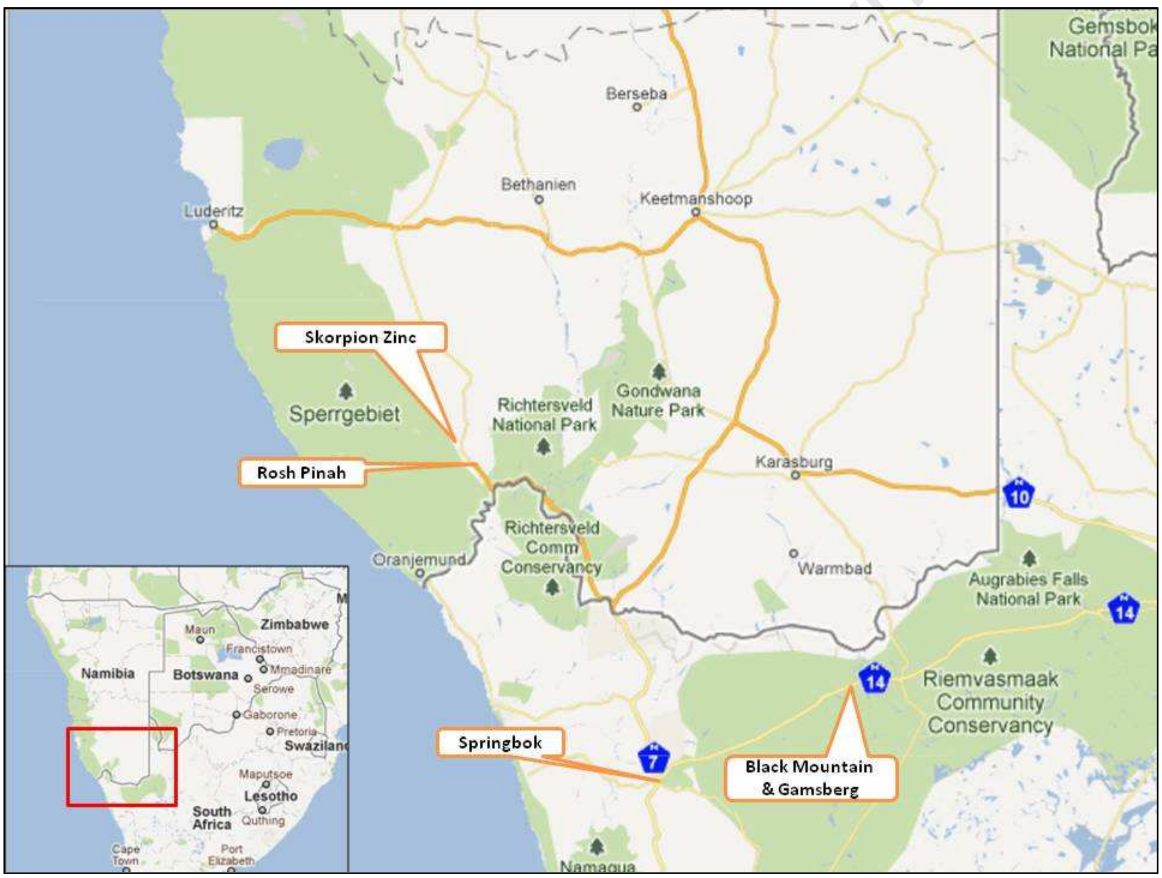



**Figure 1: Locality map of Skorpion Zinc**

The Skorpion zinc oxide resource will be depleted by 2016, with a possible extension to 2017. Extensive exploration drilling for additional zinc oxides was conducted without success. To extend the life of the operation, the following potential zinc oxide feed sources for Skorpion Zinc were considered:

i. The Tsumeb Smelter in the North of Namibia (owned by Weatherly International Plc. and in the process of sale to Dundee Precious Metals Inc.) have slag inventories of approximately 2 million tonnes, with an average zinc grade of 10% (200 000 tonnes of contained zinc). With a 90% zinc recovery under Skorpion Zinc Leach conditions, it is equivalent to just over one year of Skorpion's production. Due to the low zinc content, the transport cost outweighs the potential profit from this resource.

ii. Berg Aukas, a Zinc, Lead and Vanadium mine in the north of Namibia that was mothballed in 1978, and previously owned by Weatherly International Plc., was also considered as a feed source to Skorpion Zinc. At closure, the resource was estimated at 1.7 million tonnes of ore with a zinc grade of 17% and 5% lead, consisting of both oxide and sulphide zinc. The size of this resource does not support the capital required to re-commission the mine and upgrade the facility.

iii. Toll treatment of other zinc oxide deposits in the world was also considered. Upgrading of the ore to a zinc concentration >50% zinc is required to make toll-treatment economical. The transport cost of low grade material offsets any profits in the refining process. At present there are no economical upgraded zinc oxide resources available.

iv. Treatment of secondary zinc oxides, like Electric Arc Furnace (EAF) dust were considered. Again, to be economically viable, zinc concentration in the material of 50% is required. Typically EAF dust contains in the order of 20% to 30% Zinc. These oxides also contain high amounts of impurities that can to a limited extent be treated by Skorpion Zinc. The remote location of Skorpion from the industries producing the EAF dust makes this a less attractive option. This alternative has not been fully exploited and is the subject of another study.

Skorpion faces the option to either close the operation or consider the conversion of the Skorpion Zinc refinery to enable the processing of zinc sulphides. There is an abundant availability of zinc sulphide concentrates regionally, from Black Mountain and Gamsberg in the Northern Cape and/or Rosh Pinah Zinc and potential new deposits in the Rosh Pinah region, consideration was given to the conversion of the Skorpion Zinc refinery to enable the processing of zinc sulphides. Concentrates can also be imported via Luderitz harbour, approximately 400 km from Skorpion Zinc. The leach kinetics of zinc sulphides are vastly slower than oxides, preventing the processing of zinc sulphides using the existing process, under current conditions.

For sulphide ores, a concentrate containing zinc sulphide as sphalerite or marmatite is usually produced. The standard process for zinc sulphide concentrate treatment is the Roast-Leach-Electrowinning (RLE) process. This process is highly efficient at converting the zinc in concentrate to special high grade zinc metal by roasting to zinc oxide, acid leaching and purification, then electrolysis to zinc metal. The economics of this process are favourable when a market for sulphuric acid exists. There are a whole host of alternative hydrometallurgical processes for zinc. Sphalerite and marmatite are highly leachable in a variety of solutions including sulphuric and chloride media. As is often the case in hydrometallurgy, the sulphate route for zinc recovery has received the most commercial application with the Dynatec Zinc Pressure Leach, the Outokumpu Direct Leach and the Union Minière Direct Leach installed at a number of sites. These processes essentially avoid the roasting step of the RLE by directly dissolving zinc from the sulphide mineral followed by purification and electrolysis. These three processes have all been used to incrementally add to zinc production by being installed alongside existing RLE plants. The Dynatec Zinc Pressure Leach has been installed in one instance as a "stand alone" facility in Flin Flon Manitoba at Hudson Bay Mining and Smelting. In this case, the attraction was that sulphur was rejected as elemental sulphur to tails compared to the production of sulphuric acid.

The main challenge for Skorpion Zinc is to identify a suitable zinc sulphide treatment process for integration with the existing plant. The selection of a process option will have to consider the cost of implementation, i.e. the time and capital required to convert the existing refinery to process zinc sulphides, including the impact on production output. Consideration also has to be given to the various feed sources available, ensuring that Skorpion Zinc remains a low cost zinc producer in a volatile market and economic climate.

Once a suitable process option is identified, the related capital and operating cost will be used to establish the feasibility to proceed with the conversion of the refinery to process zinc concentrate or to close the Skorpion Zinc refinery. This decision to proceed or close will require an in depth investigation into the long term zinc concentrate supply and demand and will not be considered part of the scope of this research study.

The objective of this study is to identify a suitable process for refining zinc sulphide concentrates at Skorpion Zinc. The following questions and challenges will have to be addressed in order to reach a conclusion:

- Identify various zinc sulphide processing options.
- Consider process flows and integration with the Skorpion Zinc refinery.
- Capital cost of various process options.
- Operating cost of various process options.
- Various zinc sulphide sources and contained impurities.
- Impact of sulphide process integration on production schedules.

The approach taken to deal with the challenges and to achieve the objective is outlined in the following section.

## 1.2 STUDY PROCESS

Although most of the world's zinc (around 80%) is produced by the Roast Leach Electrowinning (RLE) process, a vast number of other zinc leach processes have either been commercially implemented or developed to pilot or laboratory scale. This section will describe the process followed to narrow the options down to a final one or two recommended processes.

In order to evaluate and eliminate or identify a suitable process, firstly a set of criteria for evaluation was established. This set of criteria portrays the company's strategy and vision.

The selected process must ultimately bring the company closer to its overall objective. Options that will not support the strategy and objectives can be eliminated during early stages of the project. There may be some processing options that are both feasible and viable with marginal differences. This developed set of selection criteria will simplify the selection process and ensure the optimum route is taken.

The second step was to compile a comprehensive list of the different process options – using a literature review. This entailed an exhaustive literature search inclusive of a very brief description of each option, which was sufficient to evaluate, eliminate or select each option against the defined selection criteria.

Following the elimination processes, a few potential process options remained. In order to select the optimum process, the integration with the current refinery was evaluated and the economic viability of each option considered. In order to evaluate the viability of integrating the process with the existing refinery, a mass and energy balance was compiled. This allowed the investigator to assess process flows, chemistry of various processes and the technical challenges of the options. This process flow development and technical evaluation is presented in chapter 5: 'Flowsheet development and mass & energy balances'.

In order to develop an economic assessment and business case against which to select the final option, a capital and operating cost estimate was required. The developed flowsheets were used to identify the major equipment requirements and capital cost estimate. The mass and energy balances were used to compile a comparative operating cost estimate for the various options.

Finally, the various options were evaluated and an optimum process selected. The final chapter summarises the study process and concludes with a motivation for the recommended option. The business case motivating further development and the recommend next phases of the study is included.

In summary the study process will be discussed under the following headings:

- Chapter 2: Selection criteria.
- Chapter 3: Literature review to identify zinc processing options.
- Chapter 4: Discussion of the Skorpion process and identification of the most promising options.
- Chapter 5: Flow sheet development and Mass & Energy balances for the selected processes.
- Chapter 6: Capital and operating cost estimates for the selected processes.
- Chapter 7: Recommendation of a suitable process for Skorpion Zinc.

# 2 SELECTION CRITERIA

A list of criteria was developed to assist with the selection of a suitable processing route for zinc sulphides. These criteria reflect the longer term vision and strategy of the company and the critical success factors for a suitable process. Development and approval of the criteria further ensures that there is common understanding between the sponsors, clients and other stakeholders on the deliverables and outcomes of the study. The criteria are first listed (Table 1) and the relevance of each item is then discussed in this chapter.

As some criteria will be more important than others on the final success, a weighting has been allocated to the various criteria. A weighting factor out of 100 was allocated to highlight the importance of the criteria relative to each other. An option is discarded if a 'fatal flaw' is identified.

| | **Description** | **Importance** | **Fatal flaw** |
|---|---|---|---|
| **1** | Operating Cost | Critical | If result in negative NPV |
| **2** | Capital cost | Critical | If result in negative NPV |
| **3** | Recovery | Critical | If result in negative NPV |
| **4** | Technical Risk | Critical | If process is not piloted |
| **5** | Safety, Health and Environmental (SHE) impact | Critical | |
| **6** | Sulphur deportment | Critical | If excess sulphuric acid is produced |
| **7** | Impact on Zn production during implementation | Important | |
| **8** | Schedule (time for implementation) | Important | If result in negative NPV |
| **9** | Flexibility to process various feeds | Important | |
| **10** | Energy Consumption | Important | |
| **11** | Water Consumptions | Important | |

**Table 1: Selection Criteria**

1. Operating Cost:

The continued existence of Skorpion Zinc depends on its profitability and ability to add value for its shareholders. Operating cost is a significant contributor to the profitability and therefore a low operating cost was a key factor for evaluation. To be competitive in the long term and ensure it will remain in business in a volatile economy, Skorpion Zinc aims to position itself in the lower half of cost producers. It is therefore essential that a low operating cost option was selected. If an option presents a negative NPV, the option was discarded.

2. Capital expenditure

A low capital cost is essential to ensure that the investment is paid back over a short period and that a high NPV is generated. This will reduce the risk of investment and enhance the profitability of the investment. A process that can use more of the existing equipment will result in lower capital expenditure and therefore be favoured.

3. Recovery:

Recovery is a strong driver of the profitability of the business, with a higher recovery positively impacting on the profitability. In general there is a trade off in business cases between capital expenditure and recovery – this was considered. Options with a higher recovery for lower capital expenditure were favoured with the outcome be reflected in the NPV.

4. Technical Risk

Technical risk refers to the certainty that the process will be able to deliver the design throughput and recovery at the estimated cost and capital expenditure. Processes that have been commercialised with a proven track record have a significantly higher certainty to deliver to design specifications when compared with a process that has only been demonstrated on the laboratory scale. A risk in capital cost, performance and scale-up exists for non-commercial processes. Historically, many projects that were developed to commercial scale, initially struggled with ramp-up and achievement of the design parameters (Nice, 2003). Due to the high technical risk of a new process not having been developed commercially, it has to be at least taken through both a continuous operating pilot and demonstration scale tests. This development is not only costly, but will add at least 3+ years of development before it can go through the final design and construction stages. This will result in missing the deadline required for the process to deliver zinc by the end of 2017. Processes not proven at commercial or demonstration scale will therefore not be considered and regarded as a 'fatal flaw' cases.

5. Safety, Health and Environmental (SHE) impact

Skorpion Zinc has committed itself to being a company conducting business sustainably. It will conduct business that aims to minimise the impact on the environment and with no harm in health and safety to all its stakeholders. Skorpion Zinc also has legal obligations to run an operation with zero effluent discharge. A newly selected process will have to conform to the current operating licenses and policies. A process is considered a fatal flaw if it results in residual effluent or tailings disposal that is not done to current Namibian legislation. A process with less environmental impact and safety risk is preferred above other processes.

6. Sulphur deportment

Sulphide mineral processes will eventually have sulphur deported in one or another form. Depending on the process, sulphur could be discarded in sulphuric acid, elemental sulphur or disposed with residue as jarosites or gypsum. Elemental sulphur can be disposed with the sulphide residue or recovered and sold or stockpiled for long periods. Processes that produce sulphuric acid as a by-product need to have an offset for the sulphuric acid. This offset can be a downstream use of acid or a local market for sulphuric acid. If no offset exists, it may result in stopping zinc production due to full sulphuric acid stock tanks. Sulphuric acid is produced as a by-product in many refineries and a sensitive balance in the sulphuric acid market exists. The demand is also highly integrated with the world sulphur market. Sulphur is produced as a by-product in the oil refining industry at an oversupply and the long term forecast is that that an oversupply in the sulphuric acid and sulphur market will remain (www.crugroup.com). Large consumers of sulphuric acid are typically the phosphate industry and other hydrometallurgical processes. Within South Africa there is a good balance between sulphuric acid supply and demand. The Uranium industry located near Swakopmund in Namibia is a net consumer of sulphuric acid. Skorpion zinc is very remotely located in the south of Namibia and will not be able to deliver sulphuric acid competitively to the market due to the transport costs involved. A zinc leach process at Skorpion that will produce sulphuric acid will therefore increase the business risk. At times when the sulphuric acid market is very constrained (oversupply), a significant additional operating cost will be added to the operation to either ship and find an offset for the sulphuric acid, or to neutralise and dispose the acid, should an acid producing process be selected. Having your primary business linked to another market, especially a by-product market, will be a disadvantage. A process producing sulphuric acid is therefore be considered as fatally flawed and not selected for Skorpion Zinc.

7. Impact on zinc production during implementation

A process that will have reduced impact on production during the conversion from oxides to sulphides is favoured. Integration of the new process may potentially result in a period where the production will be affected by the construction and integration of the new equipment with the existing process.

8. Schedule

The oxide resource will be depleted by the end of 2017. To ensure continuation of the business and avoid either mothballing costs or carrying the fixed operating cost of the Skorpion Refinery, it is essential for the selected process to deliver zinc output by the end of 2017.

9. Flexibility to various process feeds

Zinc sulphide concentrates for the Skorpion refinery may be obtained from various sources, including regional sources like Rosh Pinah, Black Mountain or from international suppliers of concentrates. The selected process must be flexible enough to competitively process regional and international concentrates thereby ensuring sustainability of the Skorpion Zinc operation.

10. Energy consumption

Energy supply in Southern Africa is highly constrained and costs are rising sharply (in the order of 25% per annum for 5 years from 2010). To remain competitive, a process with reduced energy consumption is favoured. Skorpion is committed to driving down its impact on the environment and reducing its carbon footprint.

11. Water consumption

Water is becoming an increasingly scarce resource. It is Skorpion's policy to drive down water consumption and further reduce its impact on the environment.

# 3 LITERATUE REVIEW

By the turn of the century, around 83% of the western world's zinc was produced using the Roast Leach Electrowinning process, 14% using the Imperial Smelting Process and the remaining 3% using other processes (Bucket et al, 1998). Although a few processes have historically been preferred, a fresh, unconstrained perspective was required when reviewing options for Skorpion Zinc. This is a unique application, requiring different drivers for the process selection. A literature search was conducted to identify all possible leach and production processes for zinc sulphides. Processes developed to piloting, demonstration and commercial scale are presented and briefly discussed in this chapter.

## 3.1 ROAST- LEACH- ELECTROWINNING

The **Roast-Leach-Electrowinning** (**RLE**) processing route required to produce zinc from a primary source (zinc sulphide concentrates) has its origins in 1916, when Anaconda Copper Mining Company and the Consolidated Mining and Smelting Company (today's Teck Cominco Ltd) started the operation of an electrolytic zinc plant (Filippou, 2004; Huggare et al, 1973). Today, about 65 zinc plants are using this process route (Svens et al, 2003). This process has favourable economics when a market for sulphuric acid exists. Figure 2 presents the process flow of the Roast Leach Electrowinning process.

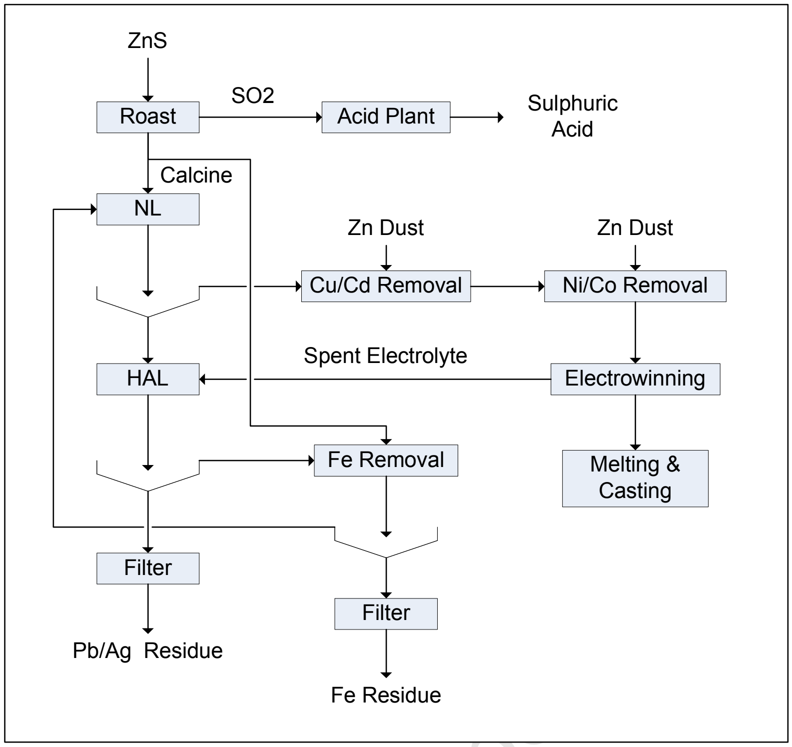



**Figure 2: RLE process Flow**

Fresh concentrate with a typical particle size distribution of 45% <53 μm and 50% between 53 and 210 μm is fed to the fluidized roasting process. The sulphides are oxidised at a temperature between 900°C and 1030°C (Sven et al, 2003; Huggare et al, 1973) in the roaster. Combustion air to the roaster serves as the carrier for the fluidized bed and provides oxygen for the reaction. The reaction is highly exothermic, producing excess steam that is used to heat up subsequent leach and iron precipitation processes. Sulphur dioxide is present at concentrations of around 10% in the gasses leaving the roaster. The sulphur dioxide gas is cleaned by an electrostatic precipitator and a wet gas cleaning stage to remove particulate matter and impurities like Hg, As, Sb and Se (Magoon et al, 1990). Sulphuric acid is then produced by conversion of the sulphur dioxide to sulphur trioxide over a catalyst and absorption in a circulating concentrated sulphuric acid stream. The sulphuric acid is sold off to other industries. Less than 100ppm $SO_2$ is released to the atmosphere in the off-gas from the sulphuric acid plant.

The calcine is leached in a two stage counter current neutral (pH 4-5) and hot acid (80-90°C) leach. A hot acid leach is required to dissolve the zinc ferrite ($ZnO.Fe_2O_3$) (Bhat et al, 1987) and Willemite ($Zn_2SiO_4$) formed in the roasting process (Sven et al, 2003; Claassen et al, 2002; Takala et al, 1999). Iron and other impurities also leach under these conditions. Spent electrolyte from the electrowinning plant is returned to the Hot Acid Leach. Solution from the Hot Acid Leach is sent to an Iron Removal step where Fe, Sb, Ge, Al and As is removed. Iron free solution is returned to the Neutral Leach where zinc oxide is dissolved. Solution from the Neutral Leach contains harmful metals (Cu, Co, Cd and Ni) which are removed in a two stage (cold and hot) zinc cementation process. Some operations like Zincor in South Africa apply more than one hot acid leach stage (Claassen et al, 2002).

Purified solution is sent to the Electrowinning plant where zinc is plated at current densities between 400 and 600 $A/m^2$ from solution. The purified solution typically has the following composition: ~140 g/l Zn, 4-10 mg/l Mn, 8-12 mg/l Mg, <10mg/l Fe, <0.01 mg/l Cu and Ni, <0.2mg/l Co and Cd. The zinc is stripped from the aluminium cathode starter sheets and smelted, ready for delivery to the market.

Several iron removal processes can be applied as discussed in section 3.13. Zinc recovery varies from 95% to 97%. Most of the zinc that is lost is via the iron removal stage where zinc oxide is used as a neutralising agent. Willimite and Zinc ferrite added in the last iron removal stages is not recovered. Soluble zinc losses also occur with the iron residue.

The chemistry of the RLE process is well known due to many years of research. The basics are presented below:

Roasting

$$ZnS(s)+1.5O_2\,(g) \rightarrow ZnO(s) + SO_2\,(g)$$

$$2\,FeS\,(s)+ZnS\,(s)+5O_2\,(g) \rightarrow ZnO.Fe_2O_3\,(s) + 3SO_2\,(g)$$

Mercury scrubbing (for environmental control)

$$Hg\,(g) + HgCl_2\,(aq) \rightarrow Hg_2Cl_2\,(s)$$

Production of concentrated Sulphuric Acid

$$SO_2\,(g) + 0.5O_2 + H_2O \rightarrow H_2SO_4$$

Leaching of the Roaster calcine with Spent Electrolyte

$$ZnO\,(s)+H_2SO_4(aq) \rightarrow ZnSO_4\,(aq)+ H_2O$$

The main advantages of the RLE process are:

- High recovery of zinc as Special High Grade Zinc for sale.
- Fixation of impurity elements in residues.

The main disadvantages of the process are:

- High grade concentrates are required for successful implementation of RLE, sometimes resulting in relatively high losses of zinc, depending on the iron removal process selected.
- The process produces sulphur dioxide which is used to manufacture acid. If acid cannot be sold locally, the value of acid may turn negative (due to the shipping cost), leading to an economic loss.
- The process is not well suited to complex concentrates containing Pb, Cu, Ag, and Au.

### 3.2 PYROMETALLURGICAL PROCESSES

The **Kivcet process** was developed in 1976 in Russia, for continuous smelting of mixed lead, zinc and copper concentrates and is described by Sannikov et al (1998) and Bartlett (1985). Currently, there are at least seven industrial operational units using the Kivcet technology. A schematic diagram of the Kivcet furnace is presented in Figure 3 below.

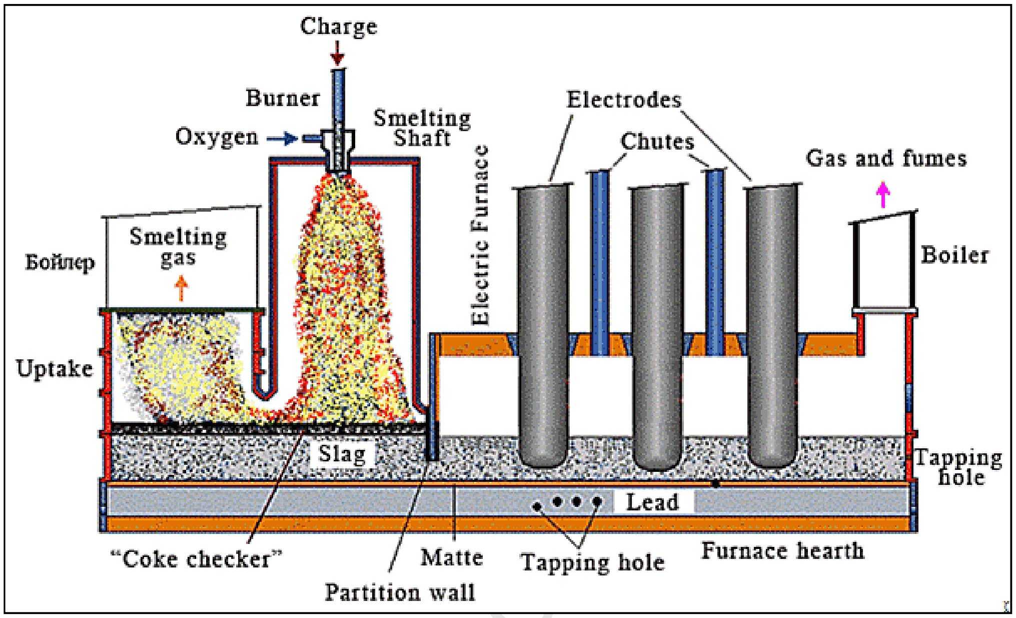



**Figure 3: Kivcet Furnace (source: VNIITVETMET 2011)**

The fine (<1mm) dry (0.5% moisture) feed is injected at the top of the smelting shaft along with oxygen. In the smelting shaft, the sulphur in the lead sulphide concentrate and the fine coal ignite instantly producing a hot, concentrated sulphur dioxide gas and the lead, zinc, iron, and other metals form metal oxides. The fluxing agents ($SiO_2$ and CaO) and the oxides form a semi-fused slag which falls to the bottom of the first compartment in the furnace along with the coarse coke. The coke collects as a surface layer, called a "coke checker", floating on top of the molten slag. When the metal oxides percolate through this layer of burning coke, it is reduced and the lead is converted to metal as bullion.

The bullion continues to settle through the molten slag layer beneath the coke checker. Together with the zinc-bearing iron slag, the bullion passes under a partition wall into a compartment, which is an electric furnace. This partition wall extends into the molten slag forcing the hot sulphur dioxide gas to pass through the waste heat boiler and on to the electrostatic precipitator rather than into the electric furnace compartment.

The larger second compartment serves primarily as a settling area where the heat from large graphite electrodes keeps the bullion-slag bath in a molten state. The lighter slag continues to float to the surface and the heavier lead bullion sinks to the bottom of the compartment. This separation enables them to be tapped separately from the furnace.

The slag containing the zinc is fed forward to a Slag Fuming Furnace where fine coal and air are injected into it. This injection generates more heat and causes the zinc to vaporise to form a mainly zinc oxide fume (also contains residual lead and silver, cadmium, indium and germanium), which is collected and further treated in the Oxide Leaching Plant in Zinc Operations to recover the zinc.

The **Ausmelt Top Submersed Lancing** technology has made high inroads in the zinc industry for the processing of zinc rich slag (slag fuming) and zinc leach residues (zinc ferrites) from the RLE process due to its energy efficiency.

Ausmelt Top Submerged Lancing (TSL) Technology has gained widespread commercial acceptance in the lead and zinc industries with twenty one furnaces in operation, producing lead and zinc (Hughes et al, 2008). This technology involves injection of fuel/air/oxygen into a molten slag bath. Solid feed materials are added via a feed port in the furnace roof. Conditions are created to effectively separate volatile species to fume, valuable non-volatile species to metal and low value non-volatiles to slag. Figure 4 displays a sketch of the Ausmelt TSL furnace.

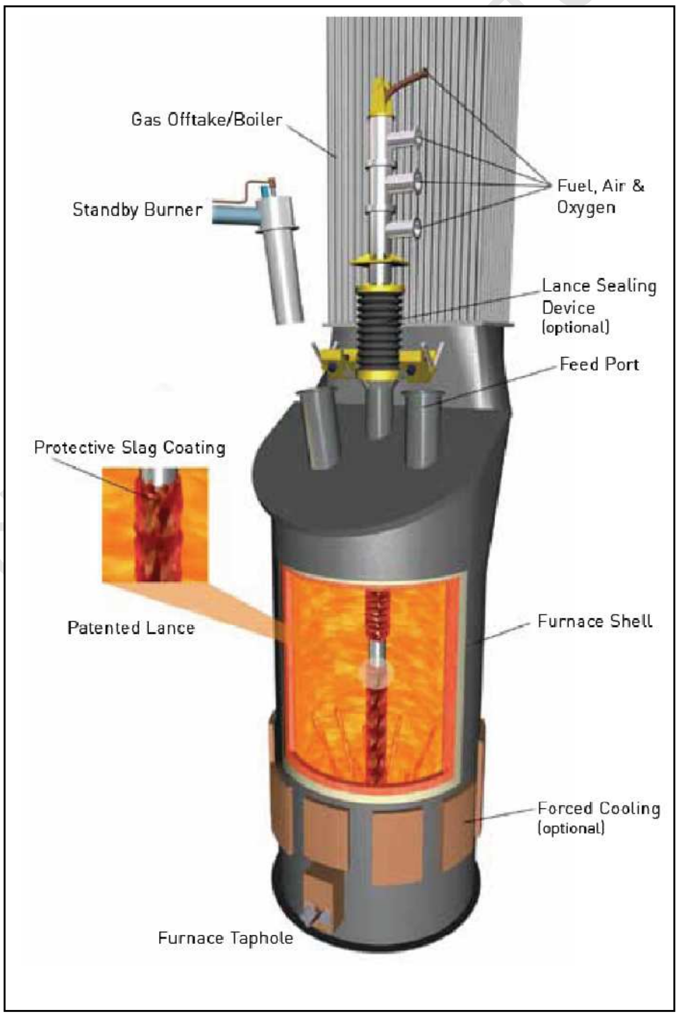



**Figure 4: Ausmelt TSL furnace (source: www.outotec.com)**

The Kivcet and Ausmelt TSL furnaces can be integrated into a flow sheet as presented in Figure 5 to produce zinc. The Kivcet furnace could also be replaced by an Ausmelt TSL furnace, having two Ausmelt TSL furnaces in series. In the first furnace, all sulphides will be oxidised with the molten zinc reporting to the slag. Volatile impurities are also driven off in the first furnace. The second furnace remains a step to fume the zinc, recovering a ZnO that needs to be leached and plated.

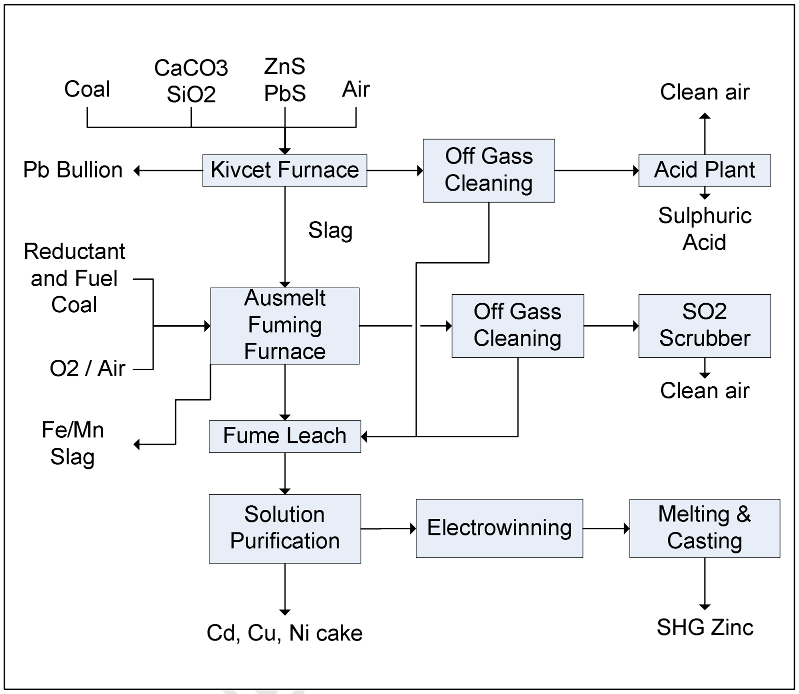



**Figure 5: Pyromet process flow including Kivcet & Ausmelt processes**

Chemistry of the process is as follow (Sannikov et al, 1998):

'Flash smelting':

$C(s) + O_2\,(g) \rightarrow CO_2\,(g)$

$C(s) + 0.5\ O_2\,(g) \rightarrow CO\,(g)$

$PbS + 1.5\ O_2\,(g) \rightarrow PbO + SO_2\,(g)$

$ZnS + 1.5\ O_2\,(g) \rightarrow ZnO + SO_2\,(g)$

Sulphates are also reduced by solid carbon at this stage according to:

$ZnSO_4 \ \ + C \rightarrow ZnO + SO_2 + CO$

$PbSO_4 \ \ + C \rightarrow PbO + SO_2 + CO$

The coke checker's function is to produce lead bullion and a ZnO and FeO bearing slag according to:

$PbO + CO\,(g) \rightarrow Pb + CO_2(g)$

$Fe_2O_3 + CO(g) \rightarrow 2FeO + CO_2(g)$

$C + CO2(g) \rightarrow 2\ CO(g)$

Zinc is then fumed off in the Ausmelt furnace and zinc oxides leached in the fume leach process. The benefit of this process is that zinc ferrite is not produced as in the roasting process, resulting in increased recovery. Iron is also trapped in an environmentally stable slag.

The **BUKA Zinc Process** was developed as a slight variation of the process described above (Buckett et al, 1998). It consists of mainly 3 processes:

a) smelting and fuming of sulphide ore or concentrate to produce a crude fume containing zinc oxide;

b) separation of a high grade zinc oxide and leaving a lead carbonate residue that can be converted to lead bullion;

c) zinc metal product using the electrolytic process. The smelting and fuming stage consists of two smelting stages in Ausmelt slag-bath furnaces.

Main advantages of the Kivcet / Ausmelt process:

- High degree of flexibility in terms of feed material quality.
- Iron locked up in slag as a stable and environmentally friendly method of iron disposal.
- High zinc recovery.

The main disadvantages of the process are:

- Primarily a process to produce lead.
- The process produces sulphur dioxide which is used to manufacture acid. If acid cannot be sold locally, the value of acid may turn negative (due to the shipping cost), leading to economic losses.
- A fuming furnace (Ausmelt) is required to produce ZnO for further processing, which will add further processing and capital cost.



### 3.3 IMPERIAL SMELTING FURNACE

The **Imperial Smelting Process** allows for simultaneous production of zinc and lead metals and is based on the reduction of zinc and lead into metal with carbon in a specially designed Imperial Smelting Furnace. An accurate process description is given on the website - http://www.williamhunter.co.uk/ZINC/isp.htm - and is quoted below:

"*An important thermal process for producing zinc is the Imperial Smelting Process and around 9% of the world's zinc is made by this process. The main difference between this and the other retort thermal processes is that the smelter produces lead as well as zinc, roughly one tonne of lead for every 2 tonnes of zinc. The roasting step in this case is carried out on a* *sinter machine, a long slowly-moving grate on which the feed material, which includes lead concentrates as well as zinc concentrates, and sometimes mixed lead-zinc concentrates, travels and through which air is blown to burn the sulphur. The reasons for using a sinter machine, apart from the burning of sulphur, are firstly that a normal zinc concentrate roasting furnace is not capable of handling significant amounts of lead and secondly that hard lumps of sintered charge are required for the blast furnace and a sinter machine is a better way to produce them than briquetting.*

*"The sulphur dioxide (SO2) gas produced from roasting concentrates on the sinter machine is then, after cleaning, further oxidised to sulphur trioxide (SO3), which is dissolved in strong sulphuric acid. The strong sulphuric acid is then diluted with water for reuse, the surplus representing production of sulphuric acid for sale. Some impurities in the original concentrate, e.g. mercury, pass with the SO2 gas and have to be removed in order not to contaminate the sulphuric acid. Sulphuric acid is a major by-product of zinc smelting, up to 2 tonnes being produced for every 1 tonne of zinc in some smelters.*

*"The charge to the blast furnace is lump sinter and coke, the coke burning in the lower part of the shaft and the heat from this and the carbon monoxide gas produced providing the means to reduce the zinc and lead oxides to metallic zinc and lead. The lead, which is below its boiling point, flows from the bottom of the blast furnace, carrying copper, silver and gold with it. The zinc vapour passes out of the furnace near the top and is rapidly quenched and dissolved into a spray of molten lead so that the zinc vapour has insufficient time to oxidise back to zinc oxide. Due to the special relationship between lead and zinc, by cooling the lead, crude zinc is released and is separated, and the lead returns to the "condensing" process for another cycle of dissolving and then releasing more zinc.*

*"A major difference between the thermal and the electrolytic processes for making zinc is that, whilst the latter produces very pure zinc directly because the removal of impurities has taken place before the reduction step, all the thermal processes, including the Imperial Smelting Process, produce a lower grade zinc that still contains significant impurities, in particular lead, cadmium, iron, copper and tin. Whilst some of these elements can be reduced to lower levels by simple means, and the zinc may then be useable for general galvanising purposes, to achieve the highest purity the zinc must be purified by distillation. Distillation can achieve as high purity as obtainable for electrolytic zinc, but the additional cost of treatment is very high, in particular for energy requirements. Many Imperial Smelting smelters sell as much zinc without further refining as possible in order to keep overall costs down."*

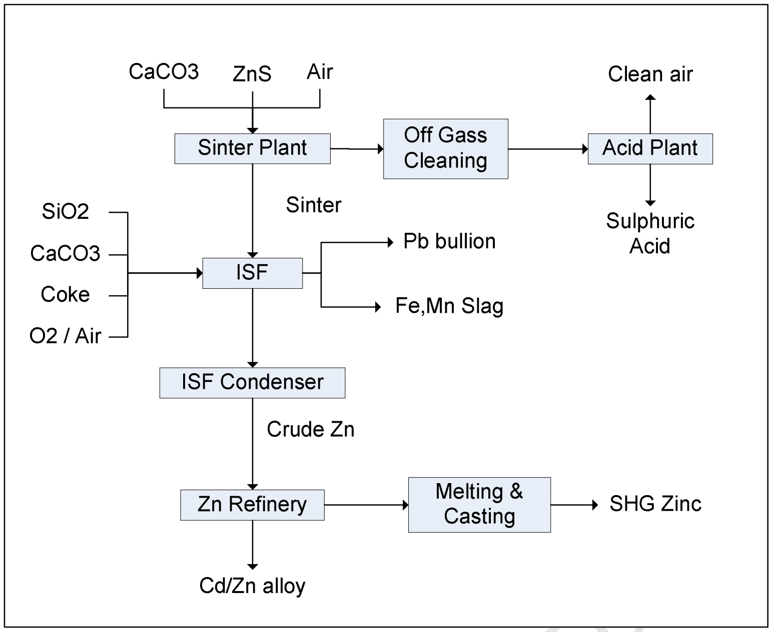



**Figure 6: Imperial Smelting Furnace process flow**

Main advantages of this process:

- Can handle relatively low feed concentrations.
- Iron locked up in slag as a stable and environmentally friendly method of iron disposal.

The main disadvantages of the process are:

- Very old technology to produce zinc. It has to a large extent been replaced with other process routes.
- The process produces sulphur dioxide which is used to manufacture acid. If acid cannot be sold locally, the value of acid may turn negative (due to the shipping cost), leading to economic losses.



### 3.4 DYNATEC PRESSURE LEACH

In the early 1950's the oxidising pressure leaching process was first applied to base metal sulphide concentrates, when a plant was constructed to leach a nickel sulphide concentrate at 85°C with an air pressure of 1MPa (Berezowsky et al, 1991). The **Dynatec** (previously Sherrit Gordon) **Pressure Leach** process was first used to treat zinc sulphide concentrates commercially in 1981 in the Cominco Zinc plant at Trail, British Columbia. The zinc pressure leaching plant included a single autoclave designed to treat 188 tpd of zinc concentrates (Ashman et al, 1990). It was an add-on to the existing Roast Leach Electrowinning (RLE) plant used to expand capacity. Following Cominco, the pressure leach has been added to existing RLE plants to expand capacity at Kidd Creek (now Falconbridge) and a third plant at Ruhr Zinc in 1991 (Buban et al, 2000). The first two-stage pressure leach for zinc sulphide concentrates was commissioned during 1993 as a replacement for the RLE process at Hudson Bay Mining and Smelting Co., Ltd. (Barth et al, 1998).

The two stage pressure leach process as illustrated in Figure 7 is discussed in this section with a brief comparison to the single stage process. Zinc concentrates are fed to a feed preparation where it is milled in close circuit to >95-98% passing 44 μm. (Doyle et al, 1978; Barth et al, 1998). Most of the zinc concentrates received from flotation plants are already ground to this size and can be pulped (~70% solids) and directly fed to the first autoclave.

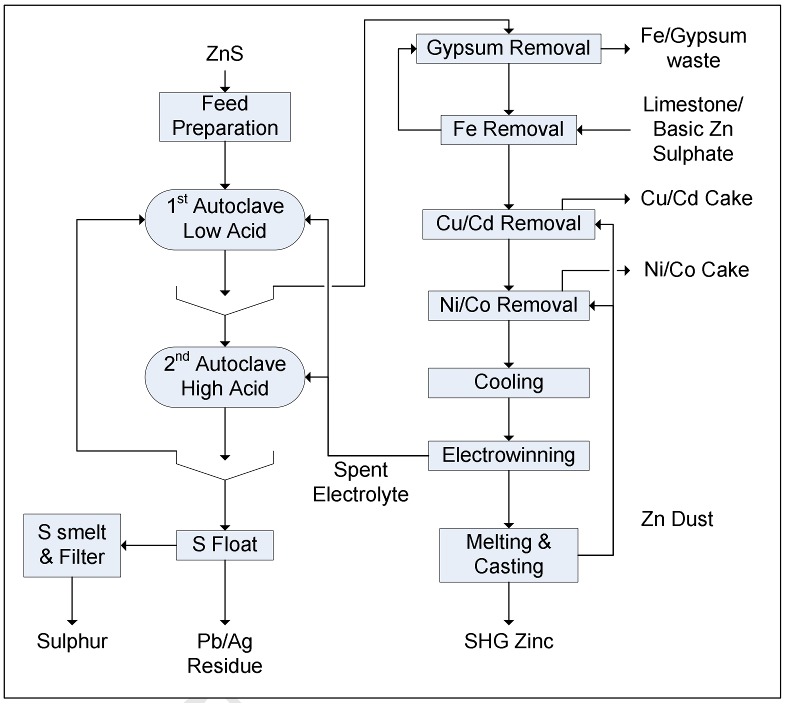



**Figure 7: Dynatec Pressure Leach process flow**

In the first stage autoclave, leaching is done at 10 to 11 bar total pressure and around 150°C. In this Low Acid Leach (LAL) stage, 75-85% of the zinc is leached, with the discharge containing around 5-8 g/l sulphuric acid. Unwashed, partially leached solids from the first stage are fed to the second stage High Acid Leach (HAL) autoclave to give an overall zinc extraction of ~98% (Berezowsky et al, 1991). Spent electrolyte is used to supply sulphuric acid for the leach process. The HAL autoclave operates at similar conditions to the LAL apart from the discharge acidity which is kept above 30 g/l. Oxygen is sparged into both the autoclaves, each with 4 compartments and 5 agitators (first compartment has two agitators) having a residence time of approximately 1 hour in each autoclave.

The first stage solution goes to a gypsum and iron removal step where the pH is raised by waste water treatment sludge (Basic Zinc Sulphate) and limestone. Ferrous is oxidised to ferric iron by sparging of oxygen through the atmospheric stirred tanks. Iron is precipitated as ferric hydroxide or goethite.

The iron free zinc sulphate solution is purified by a conventional zinc cementation process. Copper and cadmium is cemented from solution at temperatures around 70°C by the addition of zinc dust. Nickel and Cobalt is cemented with zinc and antimony or arsenic as catalyst at elevated temperatures ~90°C. The purified solution is cooled to 28-38°C before it is fed to the zinc electrowinning stages where SHG grade zinc is plated. Cathodes are melted and caste for dispatch to customers.

An optional sulphur and sulphide flotation process can be applied to the autoclave tailings to recover elemental sulphur. The lead/silver residue can be either disposed or sold to lead refineries for processing. The elemental sulphur is cleaned by a melting and filtration step to separate the floated sulphide minerals from the elemental sulphur.

The technical success of pressure leach processes is based on the use of surfactants for dispersion of molten elemental sulphur. Surfactants reduce the interfacial tension between sulphur and aqueous liquids (Owusu et al, 1995). Zinc pressure leach plants use Lignosulphonate and Quebracho as surfactants.

The process chemistry is described by Collins et al (2000) and Chalkley et al (1993) as follows:

Reactions during the pressure leach process:

$$ZnS + 0.5O_2 + H_2SO_4 \rightarrow ZnSO_4 + H_2O + S^0$$

$$PbS + 0.5O_2 + H_2SO_4 \rightarrow PbSO_4(s) + H_2O + S^0$$

This reaction is slow in the absence of an oxygen transfer agent and dissolved iron promotes the dissolution of zinc sulphide (generally known as ferric iron leach)

$$ZnS + Fe_2(SO_4)_3 \rightarrow ZnSO_4 + FeSO_4 + S^0$$

$$2FeSO_4 + 0.5O_2 + H_2SO_4 \rightarrow Fe_2(SO_4)_3 + H_2O$$

There is usually sufficient acid soluble iron in zinc concentrates to supply the needs of the leach. Pyrrhotite dissolves as follow:

$$FeS + 0.5O_2 + H_2SO_4 \rightarrow FeSO_4 + H_2O + S^0$$

The extent of pyrite ($FeS_2$) oxidation depends upon the conditions in leach. Under strongly oxidising conditions and at high temperatures (>180°C) oxidation of pyrite will result in sulphate generation:

$$2FeS_2 + 7.5O_2 + H_2O \rightarrow Fe_2(SO_4)_3 + H_2SO_4$$

$$FeS_2 + 7.5O_2 + 4H_2O \rightarrow Fe_2O_3 + 4H_2SO_4$$

With lower oxygen availability, lower temperature and higher acid concentration, oxidation of pyrite may result in some elemental sulphur production:

$$FeS_2 + 0.5O_2 + H_2SO_4 \rightarrow FeSO_4 + H_2O + 2\,S^0$$

Hydrolysis reactions remove iron from solution with the formation of hydrated ferric oxide, hydromium jarosite or plumbojarosite and result in the regeneration of some sulphuric acid:

$$Fe_2(SO_4)_3 + (x+3)H_2O \rightarrow Fe_2O_3.xH_2O + 3H_2SO_4$$

$3Fe_2(SO_4)_3 + 14H_2O \rightarrow 2H_3OFe_3(SO_4)_2(OH)_6 + 5H_2SO_4$

$PbSO_4(s) + 3Fe_2(SO_4)_3 + 12H_2O \rightarrow 2Pb0.5Fe_3(SO_4)_2(OH)_6 + 6H_2SO_4$

The two stage process is preferred for a stand-alone process as leaching conditions can be better controlled to minimise the free acid in the zinc rich solution from the leach step. This is essential to minimise the neutralisation costs. It further allows for conditions to be changed in order to control iron precipitation.

In the **Total Pressure Oxidation** process (220 to 230°C) all sulphide and elemental sulphur is oxidised to sulphuric acid (Harvey et al, 1993). The challenge in the application of total oxidation in zinc processing is to utilise the acid that is produced, or neutralise this acid at a high operating cost. On the positive side, all iron is oxidised to hematite (see also section 3.13.3), a stable and environmentally acceptable form of iron residue. Although pressure leaching at high temperatures is common in the nickel and copper industry, it is not commercially applied in the zinc processing industry, apart from the iron removal processes. The chemical reactions at high temperatures are as follows:

Sphalerite leaching: $ZnS + 2O_2 \rightarrow ZnSO_4$

Pyrrhotite leaching: $FeS + 2O_2 \rightarrow FeSO_4$

Pyrite leaching: $FeS_2 + 3.5O_2 + H_2O \rightarrow FeSO_4 + H_2SO_4$

Galena leaching: $PbS + 2O_2 \rightarrow PbSO_4(s)$

Ferrous oxidation: $4FeSO_4 + O_2 + 2H_2SO_4 \rightarrow 2Fe_2(SO_4)_3 + 2H_2O$

Plumbojarosite precipitation: $PbSO_4(s) + 3Fe_2(SO_4)_3 + 12H_2O$

$\rightarrow 2Pb0.5Fe_3(SO_4)_2(OH)_6 + 6H_2SO_4$

Ferric hydroxides formation: $2FeSO_4 + 0.5O_2 + 3H_2O = 2FeO.OH + 2H_2SO_4$

Hematite precipitation: $Fe_2(SO_4)_3 + 3H_2O \rightarrow Fe_2O_3 + 3H_2SO_4$

When pressure leaching is added as an extension to an existing RLE process plant, a single stage process is used. The required leaching can be obtained in a single stage autoclave, but the sulphuric acid content in the autoclave outlet solutions are relatively high (~20g/l). This solution is then further used in the hot acid leaches where the sulphuric acid is consumed. For a stand-alone zinc pressure leach, a high acid outlet will result in high costs to neutralise the acid. A two-stage counter-current process was therefore developed for Hudson Bay.

In this study, a stage refers to a single autoclave with several internal chambers in series or where atmospheric leach is discussed, it refers to a train of leach tanks in series. In a two-stage counter-current process, a solid liquid separation step separates the two stages (Figure 8).

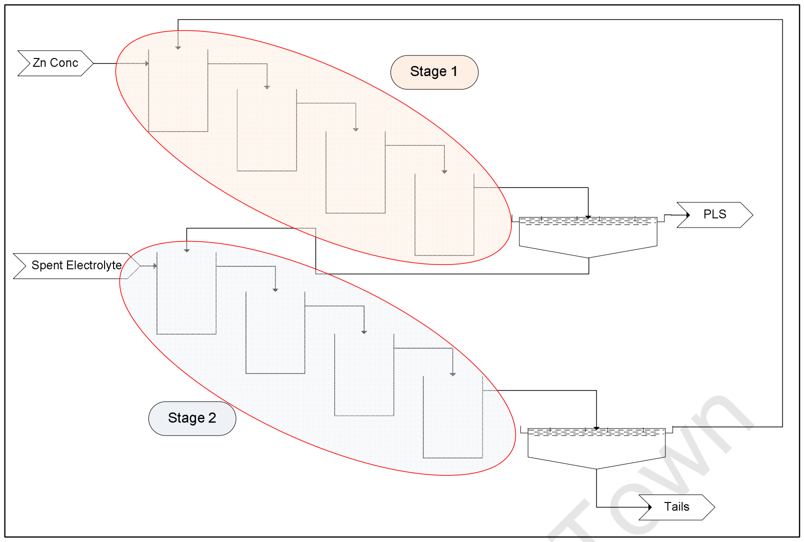



**Figure 8: Two stage counter-current flow**

Main advantages of the pressure leach process:

- Can handle concentrates with high iron and lead content.
- No sulphuric acid is produced in the 150° pressure leach.
- High zinc recovery.

The main disadvantages of the process are:

- Energy in converting elemental sulphur to acid is not harvested.
- Pure oxygen is required.
- Sulphuric acid is produced during the Total Pressure Oxidation (220°C) process.

## 3.5 ATMOSHPERIC LEACH

Union Minière developed and patented an atmospheric leach process to treat zinc sulphide concentrates mainly for integration with existing RLE plants where iron is rejected as goethite (Filippou, 2004). Outokumpu, now Outotec, has also worked extensively on the development of an atmospheric pressure leach process for zinc sulphides. Outotec's production scale applications, also as an extension to an existing RLE plants, have been implemented at Boliden's Kokkola (1998) and Odda Zinc (2004) plants. In 2008, Outotec installed a full scale atmospheric leach, stand-alone operation in China for the Zhuzhou Smelter Group.

**Union Minière** also developed a two stage variant (Figure 9) to their process used by the Korea Zinc company at their zinc refinery in Onsan, South Korea, as described by Filippou (2004) and Van Put et al (1999). The dissolution of zinc ferrite takes place mainly in the first stage and the oxidation of sphalerite takes place in the second stage. Oxygen is blown into all reactors except the last neutralisation stage. The concentration of acid and the ferric have to be well controlled in the two stage leaching. At an acid concentration below 10g/l zinc, dissolution slows down significantly and above 35g/l neutralisation costs become excessive. The ferric concentration is controlled between 0.1g/l and 2.0g/l. Union Minière patented a reactor specifically designed for atmospheric leach. The reactor is closed and equipped with a feed inlet, an oxygen inlet and overflow spillway and a draft-tube stirrer. The key to success is to disperse oxygen into the reaction mixture to re-oxidise the ferrous to ferric sulphate for zinc leaching (similar to the pressure leach chemistry).

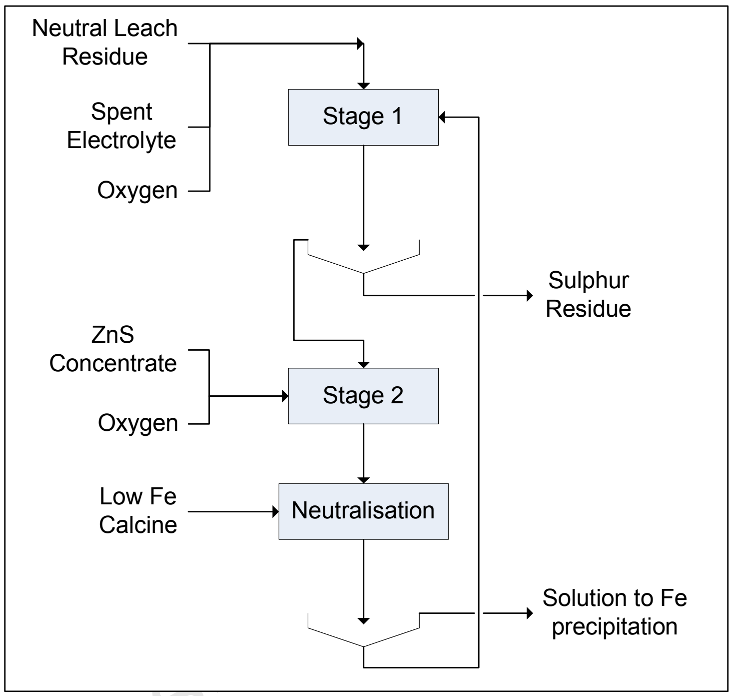



**Figure 9: Union Minière two stage process (Van Put et al, 1996)**

In the **Outotec** (previously Outokumpu) atmospheric leach process, leaching of the zinc ferrites and precipitation of jarosite takes place in the same step. Both jarosite and goethite iron removal steps can be applied with this process (see section 3.13). This is achieved by controlling the sulphuric acid concentration between 10 and 30g/l (Fugleberg et al, 1998). The Outotec stand-alone process flow sheet is very similar to the pressure leach process and presented in Figure 10.

The concentrates leached at Outotec's Kokkola plant is very fine, 15-25µm. Leaching is carried out at 100°C and a residence time of ~20 to 24hrs with recoveries of 97 to 98%. (Lahtinen et al, 2005). Outotec also patented a special 'OKTOP' reactor to efficiently suspend solids and improve gas dispersion in the slurry. The OKTOP reactor has a very large height to diameter ratio, to increase the pressure at the point of oxygen injection into the slurry.

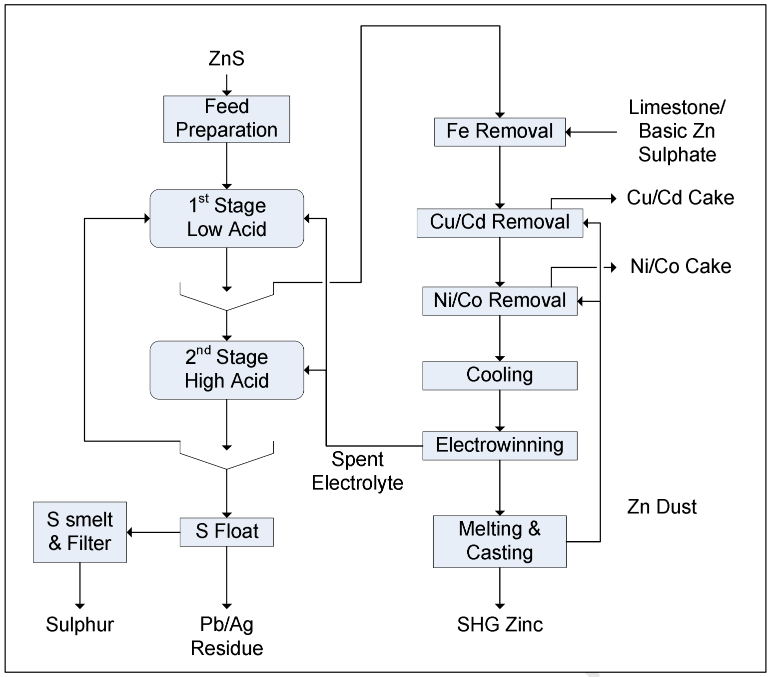



**Figure 10: Stand-alone Atmospheric Leach**

The processing chemistry for atmospheric leaching and pressure leaching is the same with two exceptions. Reaction kinetics greatly improve at elevated temperatures of 150°C, allowing for significantly lower retention time and sizing of pressure leach equipment. The other fundamental difference in process chemistry is that elemental sulphur produced is in the solid state in the atmospheric leach process and no problems are therefore experienced with sulphur coating of reactive mineral surfaces like the pressure leach process where sulphur is in the molten state.

Main advantages of the atmospheric leach process:

- Can handle concentrates with high iron and lead content.
- No sulphuric acid is produced.
- High zinc recovery.
- Relatively lower maintenance requirement on equipment compared to the pressure leach.

The main disadvantages of the process are:

- Energy in converting elemental sulphur to acid is not harvested.
- A fine grind process is required and leach recovery and kinetics will be highly dependent on particle size distribution.
- Pure oxygen is required

### 3.6 ALBION PROCESS

The **Albion** process was developed and patented in 1993 by MIM Holdings (now Xstrata) for the treatment of sulphide concentrates. The Albion Process was originally developed to treat refractory gold (Filippou, 2004; Hourn et al, 1996, 1999). The Albion Process for zinc concentrates is described on the website www.albionprocess.com. The key to the Albion process is the ultra-fine grinding of the concentrates, increasing the exposed surface of the ore particles which considerably improves the rate of the oxidative leaching compared to conventional leach reactors. The process is suitable for expansion of an existing RLE plant or as a stand-alone process as illustrated in Figure 11.

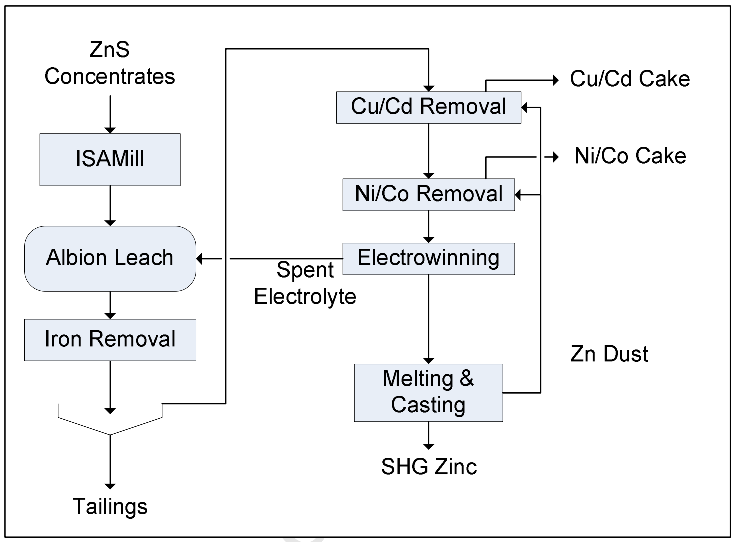



**Figure 11: Stand-alone Albion Process**

Fine grinding is achieved by using the ISAmill, grinding concentrates down to 80% < 20µm in general. The developers of the Albion process claim that the sulphur product layer passivates the mineral surface when the layer becomes >2-3 µm thick. By grinding to <10µm, the thickness will be less than 3µm thick when the entire mineral particle is consumed. However, the product layer does not passivate the surface, but rather reducing reaction rates towards the end of dissolution as diffusion through the product layer becomes rate limiting (Souza, 2007). The thickness of product layer is minimised by ultrafine grinding potentially giving the advantage. Fine grinding further potentially assists by increasing the surface area available for reaction.

Oxidative leaching is carried out in non-pressurised agitated tanks. Hourn et al (1996) describe an example of leaching of a milled (80%<3 µm) pulp at 90°C using a sulphuric acid ferric leach solution with oxygen sparging that achieves a zinc recovery of 97% over 8hrs. The pulp contained 10% solids; the initial sulphuric acid and ferric concentrations were 50g/l and 10g/l respectively. The developers claim that the leaching process can be conducted in ordinary leach reactors and the ‘specially designed’ reactors provided by Outotec might not be necessary. However oxygen supply will still be required to provide an oxidant and gas dispersion impellers are required.

Iron removal, solution purification and final zinc electrowinning are similar to those used in the pressure and atmospheric leach processes. The solution chemistry is the same as described for pressure leach, with the exception of elemental sulphur which remains in the solid state, as the reaction temperature is below the melting point of sulphur.

Main advantages of the Albion process:

- No sulphuric acid is produced.
- High zinc recovery.
- Relatively lower maintenance requirement on equipment (stirred tanks vs autoclave).

The main disadvantages of the process are:

- Energy in converting elemental sulphur to acid is not harvested.
- A fine grind process is required and leach recovery and kinetics will be highly sensitive to particle size distribution.
- This process is not proven on a commercial scale with zinc concentrates.
- Pure oxygen is required for the process

### 3.7 STIRRED TANK BIOLEACH

Bioleaching involves the microbial oxidation of elemental sulphur and ferrous to produce sulphuric acid and ferric iron that subsequently leaches sulphide minerals. Stirred tank bio-leach processes have been applied extensively in the precious metal ores (Olson et al, 2003) with the first application in 1986 to process a sulfidic gold concentrate. Percolation bioleach of copper sulphide ores is also extensively applied and will be discussed in paragraph 3.8. Lately, two plants have come into production for the recovery of cobaltiferous pyrite (Rawlings et al, 2003). Filippou (2004) described the bioleach processes developed by MIM Holdings (now Xstrata), Lulea University of Technology and the IBES or BRISA process more specifically for processing of zinc sulphide concentrates. The Stirred tank bioleach process has not been commercially applied to process zinc concentrates.

The role of the microbial organisms is to oxidize ferrous to ferric and elemental sulphur to sulphate according to the reactions below. The sphalerite dissolution follows the polysulphide mechanism (Rawlings et al, 2003; Rohwerder, 2003) where dissolution occurs through a combined attack by ferric ions and protons, with elemental sulphur as the main intermediate (Rawlings et al, 2003). Elemental sulphur can be oxidized to sulphate by sulphur oxidizing microbes, removing the sulphur product layer from the mineral surfaces. The reactions can be summarized as follows:

$$ZnS + Fe_2(SO_4)_3 \rightarrow ZnSO_4 + 2FeSO_4 + S^0$$

$$FeSO_4 + 0.5O_2 + H_2SO_4 \rightarrow Fe_2(SO_4)_3 + H_2O \text{ (Bacteria catalysed)}$$

$$2S^0 + 3O_2 + 2H_2O \rightarrow 2H_2SO_4 \text{ (Bacteria catalysed)}$$

$ZnS + 0.5O_2 + H_2SO_4 \rightarrow ZnSO4 + H_2O + S^0$

Efficient recovery of metals via bioleaching is dependent on the following factors (Boon et al, 1993):

- Optimum environment (Temperature, pH, nutrients) for the strain of micro-organism in use.
- Effective mass transfer of oxygen and carbon dioxide from the compressed air to the leach solution.
- Effective mass transfer between the lixiviant and the mineral surfaces.
- Effective oxidation of the sulphide mineral by ferric ions.
- Bacterial oxidation of the ferrous ion and sulphur compounds to form ferric ions and sulphuric acid respectively.
- The minimisation of jarosite and sulphur precipitation on the mineral surfaces which inhibit effective mass transfer.

A review of bioreactors (Rossi, 2001) used in bioleaching revealed that the literature mainly covered the performance analysis and design guidelines for stirred tank reactors, rather than having been designed as a result of an understanding of the bioleach processes. As the bioleach process involves reactions in solid, liquid and gas phases the design characteristics should incorporate the following (Rossi, 2001)

- Tank geometry: size, pulp residence time, aspect ratio, freeboard and number of tanks.
- Baffles: Number and geometry.
- Air sparger (if required) – type and position in reactor.
- Agitator: type, quantity, diameter of impeller, maximum impeller speed and power. The impeller must achieve the desired atmospheric oxygen and carbon dioxide mass transfer coefficients.

Experimental work conducted in stirred tank reactors has been limited to a 20% solids concentration. At higher solid concentrations, increased shear stresses due to the increased aeration and agitation, results in microbial cell damage (Rossi, 2001). In addition, at greater solid concentrations, the rate of oxygen consumption exceeds that of what is possible through mass transfer and hence gas-liquid mass transfer becomes rate limiting.

The MIM Holdings Pty. Ltd (now Xstrata) patented zinc metal production process from zinc concentrates is presented in Figure 12 (Steemson et al, 1994, 1997; Nilsson, 1996). The bioleach section of the **MIM Bioleach** process was piloted at 1300L scale. Leaching is done in stirred tank reactors at a pH of 1.6-1.7, solids concentration of 6-7% and 40-45°C, using a mixed bacterial population (Acidi*thiobacillus ferrooxidans, Leptospirillum ferrooxidans, Acidithiobacillus thiooxidans, Sulfobacillus* strains, *Acidithiobacillus caldus, Acidiphillium cryptum, Acidiphilium organovorum,* and other heterotrophic microorganisms). Residence time is around 72hrs. Reactors are aerated with a 2% (v/v) $CO_2$ gas as a source of carbon for autotrophic bacteria. Ammonium sulphate and mono-ammonium phosphate is added as nutrients. Bioleach recovery between 95% and 99% was obtained, depending on the concentrate (Steemson et al, 1997). Pregnant leach solutions containing 25-30g/l Zn, 3-4g/l Fe and other metals are purified by a solvent extraction process and SHG zinc is plated in the subsequent electrowinning process.

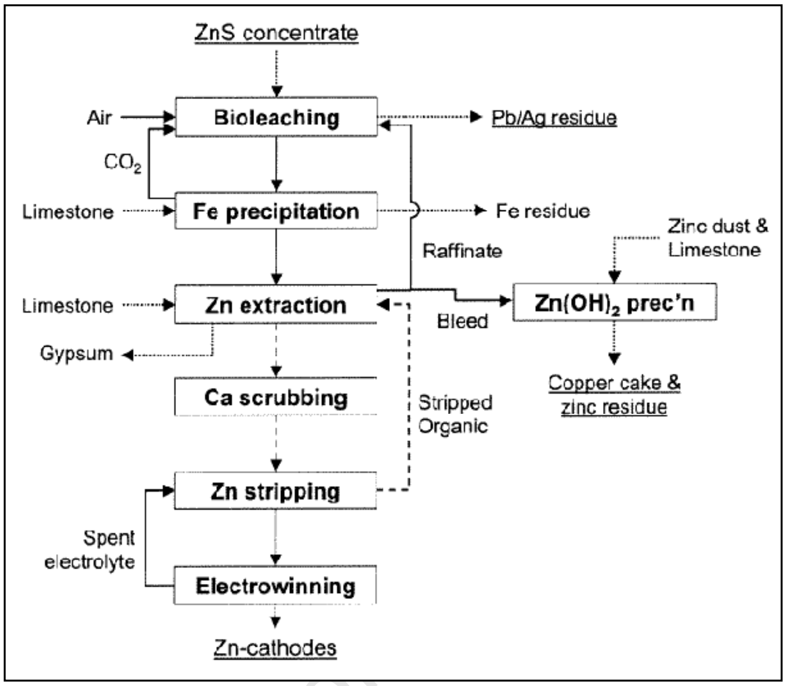



**Figure 12: MIM bioleach process (Filippou, 2004)**

Researchers from the **Lulea** University of Technology in Sweden (Sandstrom et al, 1997) developed a bioleaching process for leaching complex zinc sulphide ores, making use of extreme thermophilic archaea, which were shown to outperform the moderate thermophilic bacteria (Sandstrom et al, 1997). This is a fundamental improvement to the other bioleach processes. The extreme thermophiles (*Sulfolobus acidocaldarius* strain BC65) can tolerate temperatures of up to 65°C and higher pulp densities. Leach recoveries for zinc were in the range of 96-98%. To obtain a high zinc recovery with a low degree of pyrite oxidation, a fine particle size (20µm) was essential. Leaching is conducted over a period of 60hrs, solids concentration of 15% (mass) and temperatures of 65°C. High levels of As(III) may poison the thermophilic microorganisms, but this can be avoided by maintaining a high slurry redox potential in the leach. The Lulea process includes a zinc precipitation step after the leach (Figure 13). This precipitation/re-leach step could easily be replaced by a Solvent Extraction step as presented by the MIM process described before.

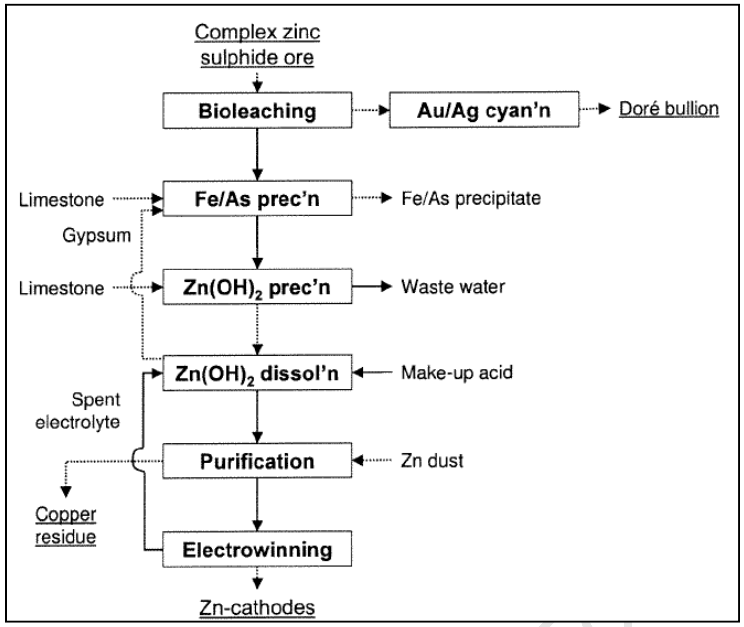



**Figure 13: Lulea Bioleach (Filippou, 2004)**

The **IBES** or **BRISA** (Indirect Bioleaching with Effects Separation) process (Filippou, 2004) uses an oxidative leach with ferric iron followed by the bacterial oxidation of ferrous to ferric for the treatment of complex zinc-copper sulphide concentrates or ores. Filippou (2004) reports that the IBES process leaches zinc at 75°, pH 1.25, 12g/l $Fe^{3+}$ for 5-10 hrs. Solids content ranges from 1 to 5% (w/w). The dissolved ferrous is reoxidised in a subsequent biooxidation stage (Figure 14). This is carried out in a 'supported bacterial film reactor' with Acidi*thiobacillus ferrooxidans* at a temperature of 30°C and pH of 1.25. The zinc rich solution is purified by SX and plated in an electrowinning step.

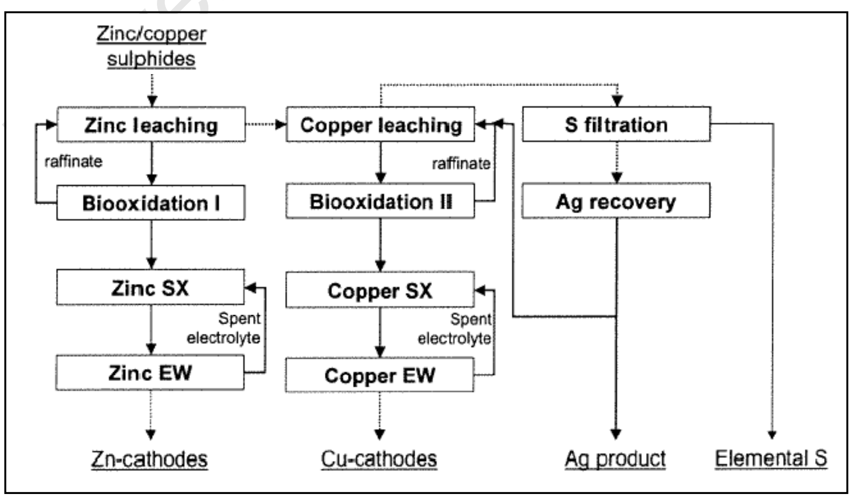



**Figure 14: IBES – BRISA bioleaching process (Filippou, 2004)**

Main advantages of the Bioleach process:

- Ability to treat low grade material.
- Claim of a lower capital cost requirement.

The main disadvantages of the process are:

- This process is not proven on a commercial scale using zinc concentrates.
- The IBES – BRISA process will probably be a challenge to scale up commercially as the engineering requirements to balance the heat between the oxidation and leaching steps will be complex.
- The low solids content of leach solutions and long residence times will increase the capital cost significantly.
- The operational discipline required to maintain optimum conditions for the bacteria.

## 3.8 PERCOLATION BIO LEACH

Several biooxidation - percolation leaching technologies have been developed and patented, including the Teck Cominco Hydrozinc, Geobiotic's Geocoat bioleach and BioHeap Limited's BioHeap process. The process chemistry is very similar to the stirred tank bioleach process chemistry with the fundamental difference in the method of contact between the various phases (gas/liquid/solid) which is taking place in a heap or dump and not in stirred tanks or pachuccas.

The **Teck Cominco HydroZinc** Process (Figure 15) was developed by the Canadian company for the recovery of zinc from low grade sulphide ores by heap bioleaching, solvent extraction and electrowinning (Filippou, 2004; Lizama et al, 2003). The novelty of the process was that heap bioleaching of low grade zinc sulphides were not previously reported and that electrowinning of zinc was performed with a manganese free electrolyte. This process was tested and developed to a 1 t/day demonstration plant at Kimberley, British Colombia. The ore was crushed to -12mm, agglomerated with acid mine drainage and stacked on a heap, 6m high.

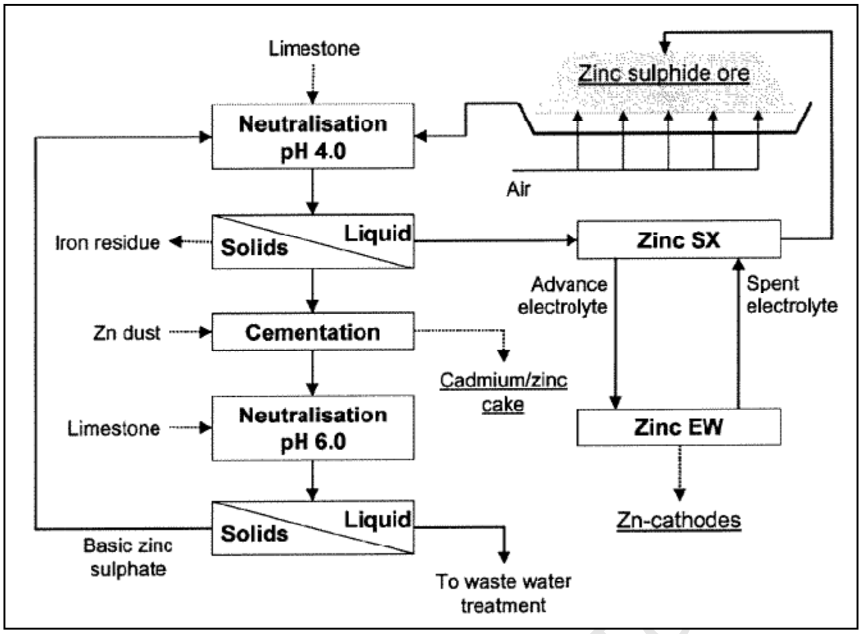



**Figure 15: Teck Cominco HydroZinc Process (Filippou, 2004)**

To prevent iron precipitation at the lower levels in the heap, it is essential to provide sufficient airflow (at least 5 l/min per $m^2$ of heap surface) to control redox potential and maintain a high acid concentration (15-30 g/l sulphuric acid). Pregnant leach solution (containing 10-15g/l zinc) is neutralized to pH 4 with limestone to precipitate ferric iron. Around 5% of the neutralized solution is bled to a cementation plant where Cu, Cd, Ni and Co is removed by zinc cementation. Other impurities (Mg, Mn, Cl, and F) are removed by a bleed stream to the waste water treatment plant. Zinc is extracted by solvent extraction with zinc being recovered by electrolysis. Electrowinning with manganese free electrolyte requires the use of lead anodes with elevated silver content (2%) or lead anodes alloyed with bismuth (1.7-1.9%) and silver (0.7-0.8%). Overall zinc recovery of this process is 80 to 85%. It must be noted that this is a recovery from the feed ore, and cannot be directly compared to the recovery of concentrates. During the flotation process, producing the concentrate around 10 to 15% of the zinc is lost. In the percolation leach, 40% iron was dissolved and around 50% of the reacted sulphide sulphur was oxidized to sulphuric acid which had to be neutralized and the other half precipitated as elemental sulphur.

The **Geobiotics** bioleach process for zinc is similar to the HydroZinc process. Process chemistry is the same as described in section 3.7. The main difference relates to the method that is used to coat a host rock with zinc sulphide concentrates and stacking of the coated material on the heap. This process is well described on the website: www.geobiotics.com. Zinc concentrates are applied to a substrate (inert coarse rock, 5-15mm) at a ratio between 1:5 and 1:10. The concentrate coating is quite stable and will not wash off the rock except for the first several layers of rock exposed to direct irrigation. The zinc is then heap leached and the solutions directed to iron control, solvent extraction and electrowinning.

The **BioHeap** process makes use of proprietary bacteria that are adapted and reproduced in a unit called the Bacterial Farm. According to Hunter et al (2004) the proprietary bacterial culture and process can operate at a wide range of temperatures without incurring the cost of bacterial oxidation cooling systems. Successful pilot trials have been demonstrated with respect to Ni, Cu and Co.

The main advantages of the percolation bioleach processes:

- Ability to treat low grade material – concentrates or mined ore.
- Lower capital investment.

The main disadvantages of the process are:

- This process is not proven on a commercial scale using zinc concentrates.
- Operational discipline required to maintain optimum conditions for the bacteria.
- Skorpion is not located in the immediate vicinity of a sulphide deposit. This will require long distance hauling of unconcentrated low grade ore, increasing the operating cost significantly.

## 3.9 DOWA PROCESS (LEACH IN MILLING)

Kanno et. al. (2002) from the **Dowa** Mining Company developed a 'leach in grind' process. Zinc concentrate is milled in a solution containing 5-15g/l ferric iron and not less than 40g/l sulphuric acid. The inventors claim that the leaching process is enhanced by stripping of the leach by-products from the mineral surfaces, thereby producing a fresh mineral surface for leaching. The ferrous in solution is regenerated (oxidised) by either processing the solution in pipe reactors under high pressure with the injection of oxygen or in stirred-tank reactors under atmospheric pressure. A recovery of 95% is achieved within 30 minutes. Concentrate size distribution is a median between 1 and 100 μm and 90% passing a size between 50 and 1000 μm.

Filippou (2004) noted that it is not known whether this process was developed beyond laboratory scale. It might be extremely costly to construct a mill to operate under these aggressive leaching conditions. Grinding media might also be very costly and ceramic media may need to be considered.

## 3.10 ALKALINE LEACH

The **CENIM-LNETI** process was developed by the Centro Nacional de Investigaciones Metalurgicas (CENIM), (Spain) and the Laboratorio Nacional de Engenharia e Tecnologia Industrial (LNETI) from Portugal (Filippou, 2004; Figueiredo et al, 1993, 1995). This process is based on oxidizing leaching in concentrate solutions of ammonium chloride using oxygen. The process involves the dissolution of the valuable elements (Cu, Zn, Pb and Ag), along with the production of ammonia and sulphate, which remains in solution and elemental sulphur which is discarded with the residue. The process flowsheet is presented in Figure 16.

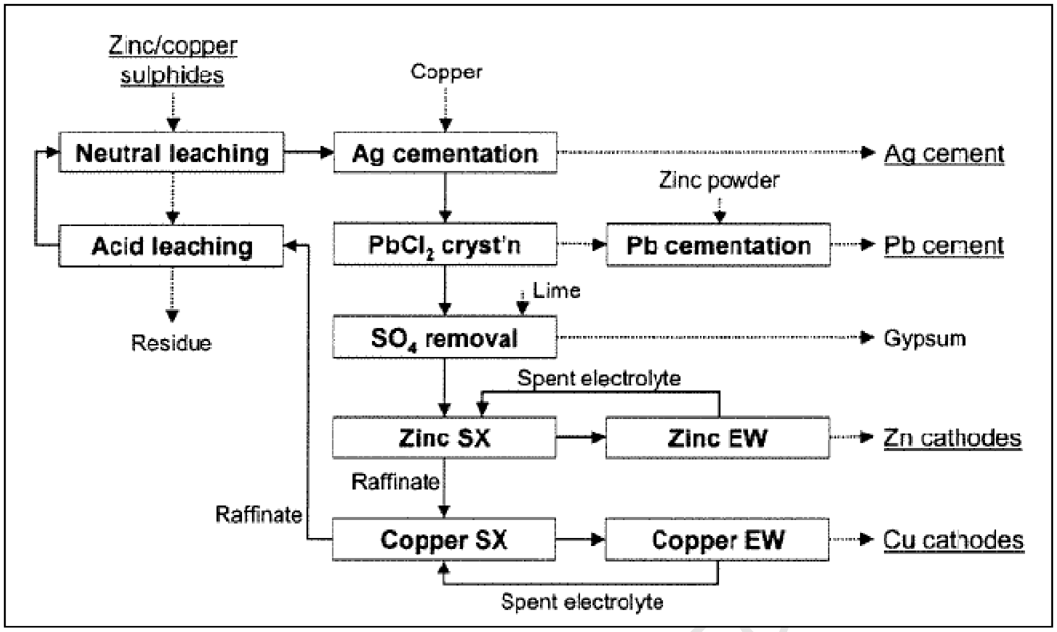



**Figure 16: The CENIM-LNETI process flow sheet (Filippou, 2004)**



Leaching is carried out in a 6 mol/l $NH_4Cl$ solution at 105°c and 1.5atm oxygen according to the following reaction:

$ZnS(s) + 2NH_4Cl(aq) + 1.50_2(g) \rightarrow Zn(NH_3)_2Cl_2(aq) + S(s) + H_20.$

Limpo et. al. (1992) stated that a part of the sulphide oxidises to sulphate while the rest oxidizes to elemental sulphur. Ammonia is formed with the production of elemental sulphur as follows:

$MeS + 2NH_4^+ + 1.5O_2 \rightarrow Me^{2+} + S + 2NH_3 + H_2O$

During the development of this process (Limpo et. al., 1992) it was shown that the effectiveness of leaching depends on the presence of cupric/cuprous ions in the leach solution. At least 1g/l $Cu^{2+}$ is required. The advantage of leaching in ammonium chloride solutions is that the pH remains constant at a pH of 6-7. The solution contains few dissolved impurities and pyrite does not get attacked, meaning that solution is almost free of iron. Figure 16 shows the use of cementation for silver, crystallization/cementation for lead and then also solvent extraction and electrowinning for each of copper and zinc.

Engitec Impianti S.P.A. of Milan, Italy developed the **EZINEX** (acronym derived from Engitec Zinc Extraction) process for the treatment of 10,000 t/year electric arc furnace (EAF) dust (Olper, 1998). In this process zinc oxide is leached with ammonium chloride where after zinc is recovered from purified solution by deposition onto titanium cathodes. Although this process has only been applied to a zinc oxide ore, the electrowinning could be combined with the CENIM-LNETI process to plate the zinc directly from the chloride medium, eliminating the need for the Solvent Extraction steps.

Main advantages of the alkali process:

- No iron residue to dispose of.
- It has the ability to process complex sulphide concentrates.

The main disadvantages of the process are:

- This process is not proven beyond laboratory scale.
- Current process using the SX step is complex.

### 3.11 OTHER CHLORIDE MEDIUM LEACH PROCESSES

Today's commercial zinc production processes rely heavily on the supply of zinc rich and impurity poor sulphide concentrates. The abundance of such non-renewable feedstock will eventually decline. Some complex sulphide ores are not amenable to flotation, for example where deposits contain finely disseminated Pb-Zn-Cu-Ag sulphides with high crystal intergrowth. For these 'difficult' or 'complex' sulphides, bioleaching and percolation leaching processes were developed. For processing of the complex sulphides several chloride medium leach processes were also developed, for example the Zinclor, INTEC, Noranda, Minemet Recherche and BHAS.

The **INTEC** Zinc Process (www.intec.com.au) consists of zinc leaching using a combination of Halex ($BrCl_2^-$) and oxygen. The zinc is leached to zinc chloride. Copper, silver, lead and iron are removed from solution prior to electrolysis. The removal of Mn and Mg with lime is also accomplished. Zn is plated at a current density of 500A/$m^2$ on the cathode from the chloride medium in a diaphragm cell with Halex formed at the anode. The attraction of the Halex system with bromide versus the straight chloride system is that a chloride remains in solution, rather than the evolution of chlorine gas at the anode.

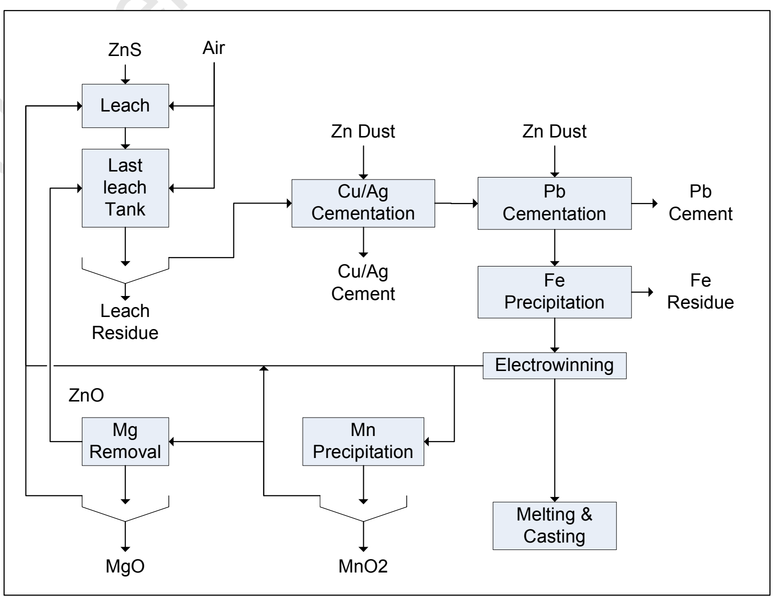



**Figure 17: INTEC Process**

An oxidative chloride/sulphate leaching process was developed by The Broken Hill Associated Smelters (**BHAS**) to process a complex sulphide ore (Figure 18) (Ricketts et al, 1989). Leaching recoveries of copper and zinc greater than 95% is achieved within 6 hours of a concentrate size distribution of 90% passing 37 µm. A two-stage leaching process is recommended to prevent the formation of jarosite during leaching and to present a low acid feed to downstream processes (SX). Leaching is done in a 15 – 30 g/l chloride solution at 100°C. Copper is extracted by an oxime solvent extraction with subsequent electroplating. Copper free solution is then sent to a goethite iron removal stage where neutralisation is done by the addition of ZnO. Zinc is removed from the iron free solution with solvent extraction, followed by conventional electrowinning.

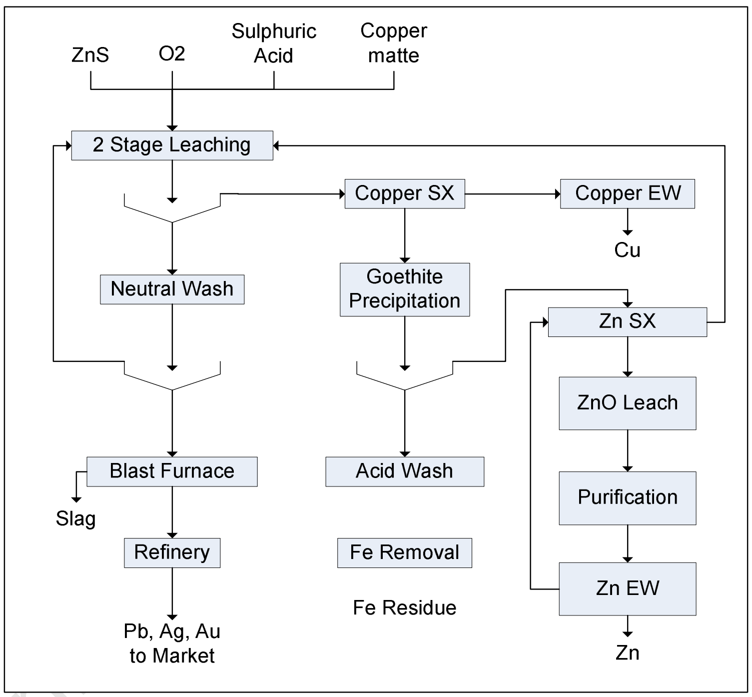



**Figure 18: BHAS flowsheet (Ricketts et al, 1989)**

The **Noranda** Chloride process (Allen et al, 2001) is a modification of earlier work by Van Weert and van Sandwijk (1999) where ZnS is dissolved with hydrochloric acid, forming $ZnCl_2$. In the Noranda process, zinc is precipitated as $4Zn(OH)_4.ZnCl_2$ by neutralising the zinc chloride solution at 80°C with MgO. This precipitate is converted to ZnO by destabilisation with milk of lime ($Ca(OH)_2$) at 95°C and pH of 9.5. This oxide can then be dissolved in spent electrolyte and recovered by conventional electrolysis.

Demarthe and Georgeaux (1978) developed the **Minemet Recherche** process in the late seventies. It consists of a selective leach of a complex sulphide ore (Zn, Pb, and Cu) by cupric chloride solution at moderate temperature (50-100°C). Pyrite is not dissolved and filtered out with elemental sulphur. Lead chloride is crystallized by simple cooling of the leach solution. The lead could then be recovered by hydrogen reduction after smelting or by iron cementation. Copper and zinc is recovered by solvent extraction, similar to the BHAS process described above.

Main advantages of the chloride medium process:

- No iron residue to dispose of.
- High recovery.
- It has the ability to process complex sulphide concentrates.

The main disadvantages of the process are:

- This process is not proven on a commercial scale.
- Current process with the SX step is complex.
- Capital costs are likely to be extremely high due to exotic material required for construction.

## 3.12 OTHER OXIDATIVE LEACH PROCESS IN SULPHIDE MEDIA

The processes discussed in the previous sections, 3.1 to 3.9, involve the leaching of sphalerite under oxidising conditions in an acidic ferric solution. Important parameters of these leaching processes are temperature and ferric ion concentration with rapid dissolution rates occurring near the solution boiling temperature and higher. Regeneration of ferrous to ferric is critical for a sustainable leach process. This regeneration can be effected by (Ferron, 2000):

- Using oxygen under pressure (Pressure leaching process) or at atmospheric pressure (Atmospheric leach).
- Using bacteria in the presence of oxygen (bioleach).
- Using chlorine gas.
- Using mixture of $SO_2$ and $O_2$.
- Using strong oxidizers such as chlorate, $MnO_2$, hydrogen peroxide or permanganate.

The methods of oxidative leach not discussed in previous sections are briefly discussed below:

Adams et al (1981) describe the leaching of metal sulphide concentrates at atmospheric pressure using $SO_2/O_2$ mixtures. His laboratory test work indicates that **$SO_2/O_2$** mixtures of appropriate composition rapidly leach a variety of zinc concentrates at atmospheric pressure in aqueous solution containing sulphuric acid and ferrous iron. Their work indicates that an oxygen to sulphur dioxide ratio of 1:1.5 is required for the reaction to occur. The oxygen consumption will therefore be fairly high if the gas is not recycled. This process has not been commercialised and the main challenge still remains the scale-up of this process due to the large airflow volumes required.

Filippou et al (2004) cited work by many others regarding the use of manganese dioxide as an oxidant for zinc sulphide. **$MnO_2$** is a very effective oxidant at low pH and can directly leach zinc from zinc sulphides as follows:

$$MnO_2 + ZnS + 2H_2SO_4 \rightarrow ZnSO_4 + MnSO_4 + S^0$$

Electrowinning of zinc from solutions containing high levels of Mn is virtually impossible. A method to separate the Mn from the zinc rich solution needs to be found for this process to be successful.

The leaching of sphalerite concentrate by means of hydrogen peroxide (**$H_2O_2$**) as an oxidant in sulphuric acid solutions was examined by Pecina et al (2008) and Aydogan (2006). Hydrogen peroxide is a strong oxidizer which forms water during the oxidation of sulphides. An increase in sulphuric acid and hydrogen peroxide concentration, increasing temperature and decreasing particle size increased the leaching rate. Test work indicates a recovery rate between 60% and 80%. With this low recovery and high cost of hydrogen peroxide it is unlikely that this process will be economical.

## 3.13 IRON REMOVAL FROM ZINC RICH SOLUTIONS

Zinc sulphide concentrates contain iron in the form of pyrite; pyrrhotite or it exists as co-ion in sphalerite minerals. In the pyrometallurgical processes the iron is locked up in the slag in a very stable form that can be disposed and stored without a negative impact on the environment (risk of leaching of iron and associated heavy metals). On the other hand, in most of the hydrometallurgical processes discussed in this chapter, iron is dissolved to some extend and needs to be removed from the solutions in the processing of zinc. Removal of iron further assists solution purification as it contributes to the removal of other impurities like arsenic, antimony, manganese and some rare earths. Iron can be precipitated as jarosite, goethite or hematite. The pH temperature stability regions for precipitation of the various precipitates are presented in Figure 19 (Claassen, 2003). Of these, the hematite is the most stable form which can be sold or stored without special precautions. The jarosite and goethite products are less stable and contain heavy metals that are easily released into the environment, resulting in the requirement of strict and costly containment systems. Some of the hydrometallurgical processes are constrained into producing a specific iron residue where others leave a choice of the iron removal process. In all cases, it is necessary to separate the iron precipitate in as coarse a crystalline form as possible to achieve high settling and filtration rates and to reduce solution entrainment. Generally this requires iron to be precipitated from a low iron concentration in solution in order to minimise nucleation. To promote crystal growth, the addition of seed material in the form of recycle precipitate can assist. Each of the iron removal processes for the hydrometallurgical routes will be briefly discussed with its advantages and disadvantages as a recommendation for the final process will have to consider the residues produced.

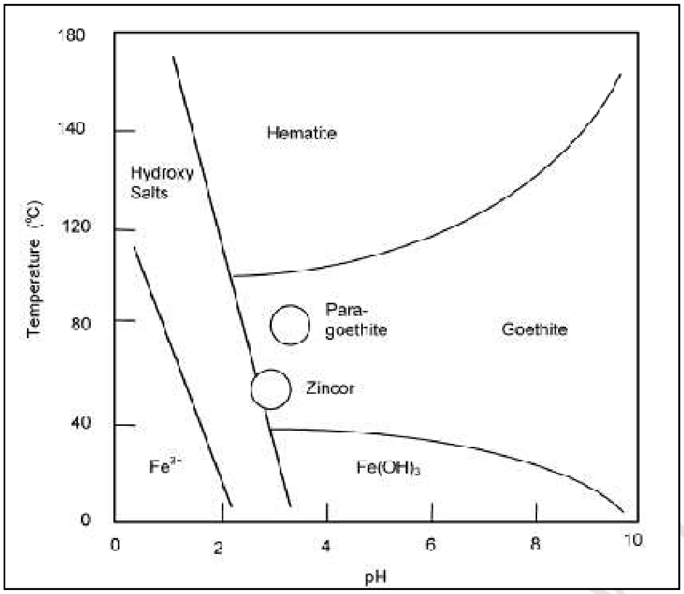



**Figure 19: Temperature – pH stability regions for the precipitation of iron (hydroxy salts include the jarosites) from0.5M ferric sulphate solutions. (Claassen, 2003)**

### 3.13.1 Jarosite

Jarosite is a complex basic iron sulphate represented as $R_2Fe_6(OH)_{12}(SO_4)_4$ where R may be any of the ions $H_3O^+$, $K^+$, $NH_4^+$ , $Na^+$, $Ag^+$, or $R_2$ can be $Pb^{2+.}$ The stability of the jarosite compound roughly follows the above series with potassium jarosite being the most stable (Doyle et al, 1989). It can be precipitated at lower iron levels and at higher acidities (Sinclair, 2005). The lead form or plumbojarosite, generally requires elevated temperatures to form and usually only appears in autoclave residues. The process is very dependent on temperature, pH and contact time (Ismael et al, 2003). Increasing the temperature from 70°C to 100°C increases the rate of precipitation considerably. The ideal conditions for jarosite formation under atmospheric pressure are temperatures close to 100°C, pH 1.5 to 1.8 (Pelino et al, 1996), vigorous agitation and the presence of seed material (jarosite). Jarosite is formed according to the reaction below where $NH_4^+$ can be substituted for $H_3O^+$, $K^+$, $Na^+$, $Ag^+$, $Pb^{2+}$. Lead sulphate is insoluble and plumbojarosite formation is therefore unfavourable at temperatures of 90-100°C and typically forms in pressure leaching at high temperatures from 145°C to 155°C (Doyle, 1989). The reaction below presents the chemistry of jarosite formation, where $NH_4^+$ can be substituted by the cations listed above.

$$3Fe_2(SO_4)_3 + (NH_4)_2SO_4 + 12H_2O = (NH_4)_2Fe_6(OH)_{12}(SO_4)_4 + 6H_2SO_4$$

This reaction liberates acid which has to be neutralised. In the conventional RLE process, zinc oxide is used for neutralisation, resulting in a loss of zinc as the zinc ferrite is not recovered in the neutralisation process. Process improvements have been made over the years to reduce the zinc losses, heavy metals and sulphate content of the process, but not to a point where this residue can be land filled without special and costly measures (impermeable ponds, control of pond waters and capping at closure).

One important consequence of the jarosite process is that it provides an outlet for sulphate from the closed solution circuit. In the RLE process, there is some addition from the sulphate content of calcine but this is insufficient to balance the loss and sulphuric acid must be added to maintain the circuit sulphate balance.

In certain circumstances in the pressure leach and atmospheric leach processes where the concentrates contain high amount of lead and silver, it is undesirable to precipitate lead or silver jarosites, but rather as sulphates as the jarosites are not recovered in a subsequent flotation step. Conditions at 150°C where jarosite is not precipitated was investigated by De Nys (1990) and presented in Figure 20.

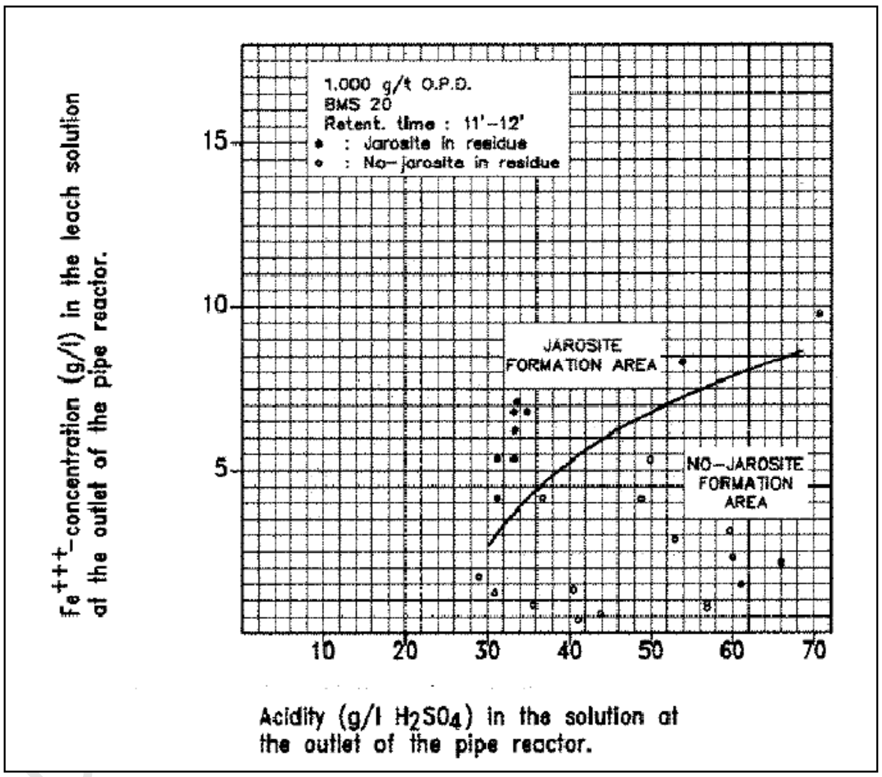



**Figure 20: Relationship between ferric, sulphuric acid and jarosite formation at 150°C (De Nys, 1990)**

### 3.13.2 Goethite

The phase diagram for the system Fe-SO4- H2O is shown in Figure 21. This indicates that at high ferric iron concentrations the stable equilibrium solid phase is ‘hydronium jarosite’. At ferric iron concentrations above 2 g/l and less than 12 g/l the stable phase is an amorphous basic sulphate, and below 2 g/l ferric the stable solid phase will be hydrated ferric oxide or goethite (Sinclair, 2005). The highly basic sulphate tends to be gelatinous and difficult to settle and filter (Loan et al, 2006). Goethite is relatively crystalline and has good settling and filtration properties. Therefore, for a goethite process the ferric iron concentration in solution needs to be maintained below 2 g/l during precipitation, ideally less than 1g/l. This can be accomplished by reducing all ferric ions to the ferrous state (known as the Vieille Montagne process) and keeping oxidation rates low or by adding the concentrated ferric solution to the precipitation process at the same rate as the goethite is precipitated by diluting into a high volume tank (Ismael, 2003). The latter process is known as the Para-goethite process.

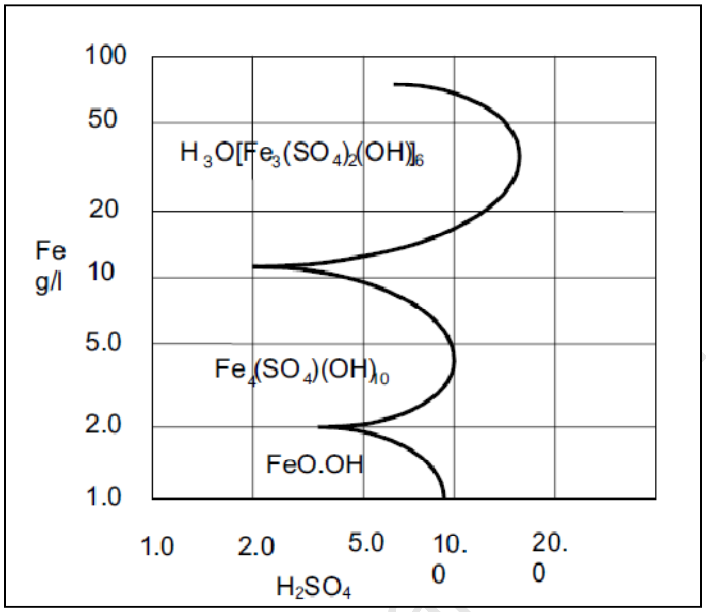



**Figure 21: Fe – SO4 – H2O system at 100°C (Sinclair, 2005)**

Goethite is precipitated at 80-90°C, pH of 2-3 and residence times of 3-5 hrs according to the reaction below. The reaction indicates that, like jarosite, acid is formed during the precipitation process that needs to be neutralised. In the RLE process, where neutralisation is done by calcine, a high amount of zinc ferrite is lost with the goethite residue. Para-goethite also contains significant levels of zinc from non-calcine origin, potentially from entrainment in amorphous structures and high adsorption of zinc onto ferrihydrite (Loan et al, 2006). The goethite residue is commonly washed to reduce zinc losses. Zinc losses in the goethite process are generally higher than the jarosite process. According to the equation below, iron is present in the precipitate at around 63%. Typically residue from the goethite process contains 40-43% iron due to the entrainment of other sulphate salts, ferrite residue and an uncertain degree of hydration. Jarosite residue contains around 30% iron, significantly lower than goethite (Sinclair, 2005). Therefore the advantage of the goethite process is to deliver a lower volume of waste.

$2FeSO_4 + 0.5O_2 + 3H_2O = 2FeO.OH + 2H_2SO_4$

The advantages of the Paragoethite Process (inexpensive to implement, ease of operation, lowest capital, low sulphate residue) are offset by the poor liquid-solid separation characteristics, high neutralizing agent requirements and higher zinc losses (Loan et al, 2006).

Disposal of goethite waste presents a serious environmental problem due to the presence of heavy metal impurities. Like jarosites, special and costly landfill requirements exist, with the advantage of lower volume for goethite residues.

### 3.13.3 Hematite

The phase diagram for the system Fe-SO4- H2O, (Figure 22 ) changes significantly at elevated temperatures, with much less hydration. Iron may be precipitated from solutions containing higher acid concentrations, and at lower final ferric iron concentrations hematite ($Fe_2O_3$) rather than goethite (FeO.OH) is the stable phase (Sinclair, 2005).

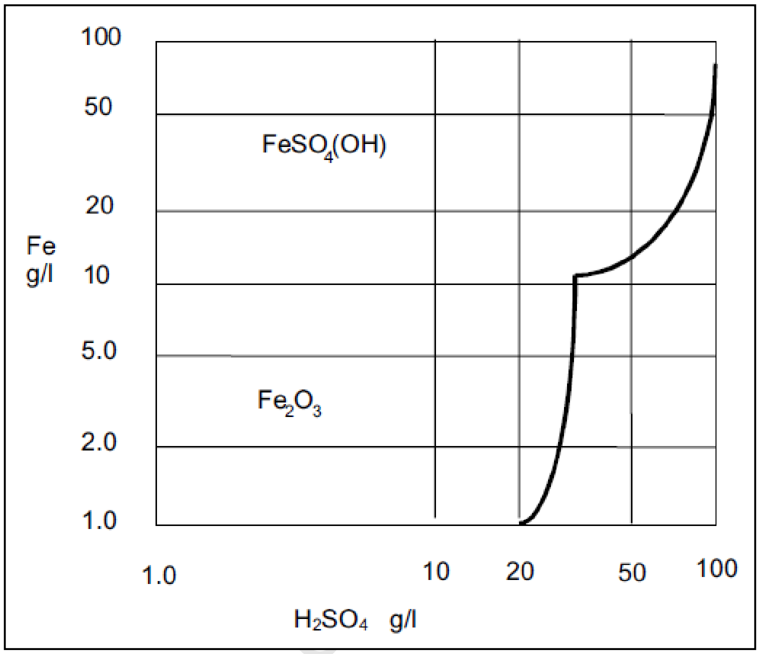



**Figure 22: Fe – $SO_4$ – $H_2O$ system at 175°C (Sinclair, 2005)**

In the hematite process, ferrous is oxidized and precipitated as hematite under oxidizing conditions at temperatures higher than 185°C and by injection of oxygen into the autoclave according to the reaction below. Acid is produced by the reaction and needs to be maintained below 56g/l to prevent the formation of $FeOHSO_4$ (Ismael et al, 2003). This places a practical limit on the iron concentration in the feed. Due to the high temperatures of this process, high capital and operating costs result. The advantage of this process is significantly reduced zinc losses as no calcine (zinc ferrites) are required for a neutralisation step (the precipitate forms even at high acid concentrations). Recovery up to 99% zinc is easily achieved. The further advantage is that a significantly lower volume of residue is produced, containing around 66% iron. This residue is free of heavy metals and can be sold to steel producers or disposed without adverse environmental consequences. This process is commercially applied at Akita Zinc and Ruhr Zinc, with the latter company no longer operating.

The various iron removal processes are summarised and compared in Table 2. Values are based on a 100t zinc concentrate feed (Sinclair, 2005).

| Process | Jarosite | VM-goethite | Paragoethite | Hematite |
|---|---|---|---|---|
| Iron Residue<br>Fe content | 29.0% | 40.0% | 34.0% | 57.0% |
| Zn content | 3.5% | 8.5% | 13.0% | 1.0% |
| Pb content | 1.9% | 1.9% | 2.2% | 0% |
| Quantity of Fe residue | 22.5 t | 16.2 t | 19.2 t | 11.2 t |
| Zinc loss in Fe residue | 1.51% | 2.65% | 4.79% | 0.21% |
| Quantity of sec leach residue | 6.0 t | 6.5 t | 6.0 t | 8.0 t |
| Zinc loss in sec leach residue | 0.58% | 0.63% | 0.58% | 0.77% |
| Overall zinc recovery | 97.9% | 96.7% | 94.6% | 99.0% |

**Table 2: Comparison of various iron removal processes (Sinclair, 2005)**

## 3.14 SUMMARY

The processes available for the processing of zinc concentrates have been presented. These are at various stages of development and summarised in Table 3. RLE dominated the industry for quite some time with the Sherrit pressure leach been introduced in the early eighties. The Umicor and Outokumpo atmospheric leach and the MIM Albion processes started to compete with the RLE and pressure leach processes. The current commercial favourites, RLE, zinc pressure leach, the Umicor and Outotec atmospheric leach processes are clearly robust processes for treating clean, high grade zinc concentrates. However, with the depletion of low impurity zinc sulphide deposits, complex sulphide deposits will become more important. Many of the processes discussed are potentially very effective for dealing with low grade or complex sulphide concentrates and will attract attention in future research.

Table 3: Summary of zinc sulphide processes

| Process | Technology | Feed Type | Leach Temp °C | Leach Pressure bar | Residence Time hrs | Leach Recovery % | Feed PSD | Oxidant | Media | Fe Disposal | Status | S product | Env. Impact | Comments |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| **RLE** | Pyro/hydro | Clean Concentrates | 65-90 | Atm | | 95 | | Oxidised Feed | Sulphate | Jarosite; Goethite; Hematite | Comm | Sulphuric Acid | A,W,S | 80% of world Zn production |
| **Kivcet/Ausmelt** | Pyromet | Lead/Zinc concentrates | 65-90 | Atm | | 99 | | Oxidised Feed | Sulphate | Slag | Comm | Sulphuric Acid | A | Mainly developed for lead concentrates |
| **Imperial Smelting Furnace** | Pyromet | Lead/Zinc concentrates | 65-90 | Atm | | 95 | | Oxidised Feed | Sulphate | Slag | Comm | Sulphuric Acid | A | Old process |
| **Dynatec Pressure** | Press Leach | Zinc Concentrates | 150 | 12 | 2 | 99 | | Oxygen | Sulphate | Jarosite; Goethite; Hematite | Comm | S | W,S | Proven technology; Avoid sulphuric acid production |
| **Total Pressure oxidation** | Press Leach | Zinc Concentrates | 220 | | 1 | ~99 | | Oxygen | Sulphate | Hematite | Comm | Sulphuric Acid | W,S | Used widely on Cu, Ni and Au, not in zinc production; Acid producing |
| **Union Miniere** | Atmospheric | Zinc Concentrates | 90 | Atm | 11 | ~95 | P90: 45µm | Oxygen | Sulphate | Goethite | Comm | S | W,S | No acid; PSD critical; Special reactors |
| **Outokumpu** | Atmospheric | Zinc Concentrates | 100 | Atm | 20 | ~98 | P90: 45µm | Oxygen | Sulphate | Jarosite | Comm | S | W,S | No acid; PSD critical; Special reactors |
| **Albion** | Fine Grind | Complex Sulphides | 90 | Atm | 8 | ~97 | P90: 10-20µm | Oxygen | Sulphate | Goethite | Pilot | S | W,S | Ultra fine grind |
| **MIM bioleach** | Bioleach | Complex Sulphides | 45 | Atm | 72 | 95-98 | P80: 35µm | Air | Sulphate | Goethite | Pilot | Gypsum | W,S | Not commercialised |
| **Lulea** | Bioleach | Complex Sulphides | 65 | Atm | 60 | | P80: 20µm | | Sulphate | Hematite | Pilot | Gypsum | W,S | High As poison bacteria; Process has difficulty to process impurities. |
| **IBES-BRISA** | Bioleach | Complex Sulphides | 75 | Atm | 10 | 60-90 | | | Sulphate | Goethite | Pilot | S | W,S | Sequencial zinc then copper leach. |
| **Hydrozinc** | Percolation Bioleach | Ore | 35-60 | Atm | 740 days | 80-85 | Crush & Agglomerate | Air | Sulphate | Goethite | Pilot | Gypsum | W,S | Relatively low recovery |
| **Geobiotics** | Percolation Bioleach | Complex Sulphides | 75 | Atm | | | P90: 45µm | Air | Sulphate | Goethite | | Gypsum | W,S | Not commercialised or demonstrated for zinc concentrates |
| **BioHeap** | Percolation Bioleach | Ore | | | | | | | | | | | | |
| **DOWA** | Leach in mil | Zinc Concentrates | 80-95 | Atm | 1 | 95 | P50: 1-100µm | Oxygen | Sulphate | | Lab | S | W,S | Challenge to construct mill for this leach environment. |
| **CENIM-LNETI** | Alkaline | Complex Sulphides | 105 | Pressure | 3 | 95 | P90: 45µm | Oxygen | Ammonium Chloride | Goethite | Lab | S | W,S | Not commercialised |
| **EZINEX** | Alkaline | EAF Dust | 70-80 | Atm | 1 | 90 | Fine | Oxidised Feed | Ammonium Chloride | | Comm | n.a | | Zn leach strongly dependent on Zn/Fe ratio; 90% at Zn:Fe of 3:1. |
| **INTEC** | Chloride Medium | Complex Sulphides | 85 | Atm | | 90 | P80: 40µm | | Chloride | Hematite | Lab | S | | |
| **BHAS** | Chloride Medium | Complex Sulphides | 100 | Atm | 6 | 95 | P90: 37µm | | Chloride | Goethite | Lab | n.a | | |
| **Noranda** | Chloride Medium | Complex Sulphides | 80 | Atm | | | | | Chloride | | N | n.a | | |
| **Minemet Recherce** | Chloride Medium | Ore; Complex sulphides | 50-100 | Atm | 2-4 | 90-99 | P80: 28-500 micron | Cupric | Chloride | Goethite | Lab | n.a | | |
| **SO2/O2** | Other | Zinc Concentrates | 80-100 | Atm | | | | SO2/O2 | Sulphate | Jarosite; Goethite | Lab | S | | |
| **MnO2** | Other | Zinc Concentrates | 60 | Atm | 5 | 60 | | MnO2 | Sulphate | Jarosite; Goethite; | Lab | S | | |
| **H2O2** | Other | Zinc Concentrates | 60 | Atm | 4 | 60-80 | P90: 38µm | H2O2 | Sulphate | | Lab | | | |

**Environmental Impact:** **A**: possible air pollution; **W**: possible water pollution; **S**: possible soil pollution

# 4 IDENTIFICATION OF SUITABLE OPTIONS

Various options to process zinc concentrates were discussed in the previous chapter. The multitude of options needs to be reduced to a suitable option for Skorpion Zinc. This chapter will briefly describe the Skorpion Zinc process, as a basis for creating a better understanding of the discussion on the integration of the potential options to the zinc oxide process. A screening of all the process options against the set criteria (chapter 2) is discussed in the second half of this chapter. The objective was to eliminate as many processes as possible resulting in the minimum number of possible processes remaining for further flow sheet development and costing, in the remaining part of this study.

## 4.1 SKORPION ZINC PROCESS DESRIPTION

The description of the Skorpion process flow should be read in conjunction with the overall process flow block diagram presented in Figure 23.

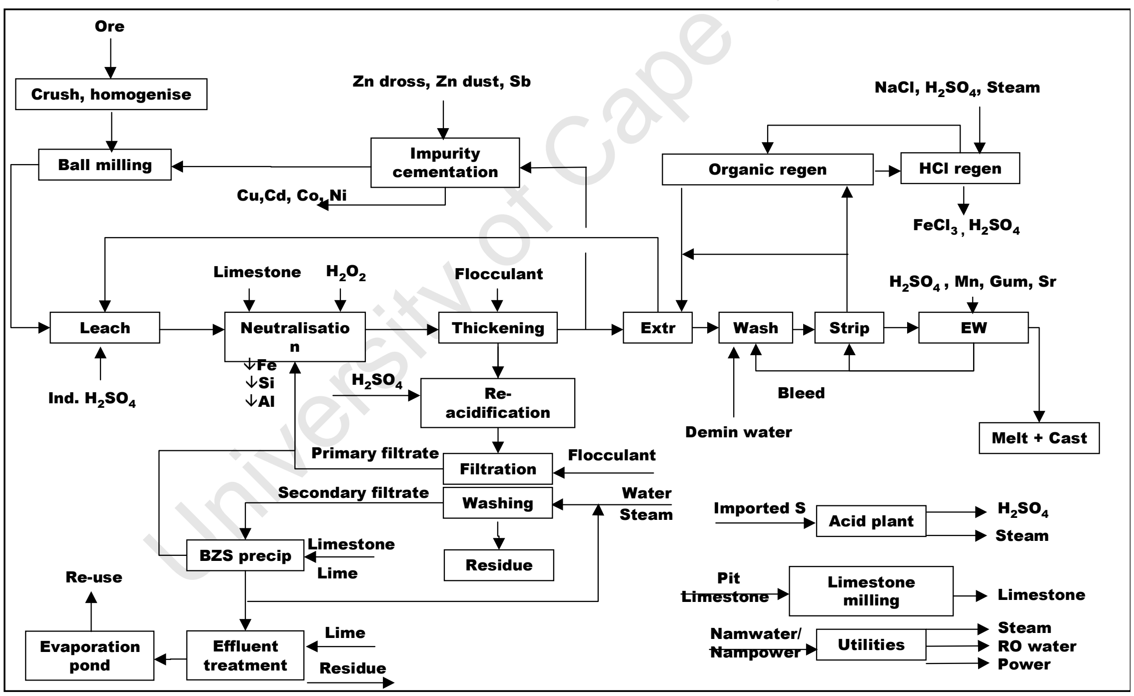





**Figure 23: Skorpion Zinc flowsheet**

### 4.1.1 Comminution and leaching

Oxide ore (hemimorphite, sauconite, willemite and smithsonite), containing zinc and silicon at averages of 25.9% and 10.6% respectively, is crushed, homogenised and milled in a closed circuit ball mill to 80% passing 200 µm. During leaching, silicates are leached to primarily form monomeric silicic acid ($Si(OH)_4$), which is polymerised partially to form colloidal silica. Colloidal silica can combine to form either open structures, within which water is bound (silica gel), or crystalline coagulates, depending on process conditions. In order to avoid the formation of gelatinous silica which causes settling and filtration problems, it is necessary to control pH, temperature and residence time. The operating conditions in the leach process are pH 1.8-2.0 (5g/l free sulphuric acid), temperature of 50°C and a residence time of 2 hours.

### 4.1.2 Neutralisation

In this step conditions are manipulated so as to yield filterable silica colloids through controlled coagulation. pH is increased to 4.3 at a temperature of 50°C at a residence time of around 4.5 hours.

Iron is present in Skorpion ore at levels of around 2%, primarily as ferric iron oxides. The sulphuric acid leach stage dissolves some of this iron, which must be removed from solution prior to solvent extraction, as zinc solvent extraction with D2EHPA is not selective for zinc over iron. Co-extraction of iron with zinc will yield high levels of iron in the loaded organic, which would be co-stripped with zinc into the loaded electrolyte. Iron levels in loaded electrolyte in excess of 5-10mg/l are known to reduce zinc current efficiency. The neutralisation stage, incorporated to deal with silica, is also effective for ferric iron removal to levels below 1-2mg/l by precipitation of ferric oxides and oxy-hydroxides, such as goethite.

Aluminium is also removed during the neutralisation stage as aluminium hydroxides, and aluminium fluoride ($AlF_3$) depending on fluorine levels in solution.

The neutralised slurry is thickened to produce a pregnant leach solution containing 35g/l Zn and an underflow pulp, containing leach residue and precipitates, formed during neutralisation. The thickener underflow also contains minor quantities of basic zinc sulphate ($(ZnO)_3ZnSO_4$), which precipitates during neutralisation. This basic zinc sulphate is re-dissolved in a re-acidification step at a pH value of 3.8. The residue from re-acidification is filtered and washed by belt filtration with two steps of counter current washing. The residue is deposited as a filter cake.

### 4.1.3 Bleed and effluent treatment

Copper, cobalt, nickel and cadmium are not co-extracted with zinc during solvent extraction and are recycled with aqueous raffinate to the leach circuit. In order to control impurity levels and the overall water balance, primary filtrate is bled from the primary belt filtration circuit and combined with the secondary filtrate from the belt filters. Zinc is recovered from the bleed solution and secondary filtrate by conventional precipitation of basic zinc sulphate by neutralisation to a pH of 6 with limestone and lime. The precipitated basic zinc sulphate is separated from the impurity containing bleed solution by thickening and filtration. The overflow solution from the basic zinc sulphate is re-used for filter cake washing, with the remainder sent to the effluent treatment plant where the remaining impurity elements are precipitated at a pH of 10, with lime. This bleed stream also provides an exit for elements like Mn, Mg, Cl, Na and K. The final effluent slurry is filtered to produce a clean liquid for re-use or discharge to evaporation ponds. The solid cake is blended with the residue.

During the precipitation of basic zinc sulphate, copper is completely and nickel, cobalt and cadmium partly precipitated and therefore recycled to the neutralisation step. The bleed treatment system does therefore not provide an outlet for these elements (particularly copper) and dedicated copper, cadmium, nickel and cobalt removal is required to prevent build-up in the circuit. Copper and cadmium removal is done through cementation with metallic zinc dust. Nickel and cobalt are removed using zinc dust and antimony tartrate.

### 4.1.4 Solvent extraction

Zinc extraction takes place in three stages in which organic raffinate flows counter current to pregnant liquor. During extraction a delta zinc of ~22g/l is achieved in the aqueous stream, yielding an aqueous raffinate stream at 13g/l zinc and 30g/l sulphuric acid.

During the extraction stage, phase separation does not always reach completion and minor amounts of aqueous phase remains entrained in the loaded organic and vice-versa. Entrained organic phase is removed from the aqueous raffinate through activated carbon filters, before returning the raffinate to the leach circuit. The aqueous phase entrained in the loaded organic must be removed to prevent the transfer of impurities to the electrowinning circuit. Also, minor amounts of calcium are extracted with zinc, which must be removed to prevent the formation of gypsum in the electrowinning circuit. This is done by a three-stage washing process. The washing process uses a mixture of demineralised water and spent electrolyte (from electrowinning). The washing process is both physical and chemical and removes entrained aqueous phase as well as calcium as $CaSO_4.2H_2O$ (gypsum).

After washing, the loaded organic is brought into contact with spent electrolyte from electrowinning to strip zinc from the organic, producing loaded electrolyte and organic raffinate. As with raffinate, entrained and dissolved organic are removed from the loaded electrolyte by activated carbon columns. The loaded electrolyte is then pumped to the electrowinning cellhouse. The stripped organic (organic raffinate) is returned to the extraction stage, completing the organic cycle.

Iron is co-extracted with zinc, even at low levels of iron in the pregnant leach solution, and since iron is particularly detrimental to electrowinning, it is removed from the organic raffinate in a regeneration step, using 6M hydrochloric acid. The $FeCl_3$ containing HCl solution from this step is sent to an HCl regeneration step, where sulphuric acid and salt are added to produce regenerated HCl. This regeneration step involves a high amount of concentrated sulphuric acid to be added.

### 4.1.5 Electrowinning, melting and casting

Conventional zinc electrowinning is used to produce special high grade (99,995% Zn) cathodes. As electrolyte is circulated in close loop with the Solvent Extraction circuit, small amounts of co-extracted impurities are building up in the electrolyte. To ensure that impurity limits remain below the threshold values, a small amount of electrolyte is bleed off to scrubbing settlers SX, where it is used as scrub liquor. The lost electrolyte is replaced with demineralised water and chemically pure sulphuric acid. Finally, special high grade ingots (25kg) and Jumbos (1000kg) are produced in a melting and casting step.

### 4.1.6 Sulphuric acid plant

A sulphur burning acid plant is included in the overall process to produce industrial grade and chemically pure sulphuric acid for use in leaching and electrowinning, respectively. The sulphuric acid plant also produces 1.05t of saturated steam per ton of acid production, which is used for process heating.

### 4.1.7 Limestone circuit

Limestone is in abundance in the resource and is mined as waste from the pit. The limestone is crushed and milled to 90% minus 45 μm, before addition to the process.

## 4.2 IDENTIFICATION AND DISCUSSION OF SUITABLE OPTIONS

This section will describe the screening of all process options listed in Table 3 and compare it to the selection criteria discussed in Chapter 2. Options with a clear disadvantage and containing a ‘fatal flaw’ characteristic are eliminated with sound arguments and discussion. Options that indicate potential for further development are discussed and motivated.

It is imperative to consider the potential feed sources for this refinery when the different process options are evaluated. Due to economies of scale and to achieve the lowest possible operating cost the production capacity of 150,000 tonnes per annum of the existing electrowinning unit needs to be utilised or exceeded. For the purpose of this exercise an increase of capacity above the current design capacity will not be considered, as the objective of the study is to select a suitable process, in other words it is a trade-off study. An amount of ~300,000 tonnes of zinc concentrate at zinc grade of 50% needs to be sourced for the refinery at this current capacity. Zinc concentrates could be sourced regionally, like Rosh Pinah Zinc, Gamsberg, potential new zinc deposits, or sourced from international markets. Figure 24 presents a typical composition of zinc concentrates in the industry, sourced from the author’s database. The Rosh Pinah values are presented separately as an indication of what composition can be expected from the regional concentrates. It must be noted that the iron concentration in the Rosh Pinah concentrate is lower compared to the industry values, as this concentrate is produced for processing in a RLE process and the iron is therefore kept low by manipulation of the flotation process, at the expense of zinc recovery in the flotation process. It must also be noted that the manganese is also slightly higher, originating from the mineralogy of the Rosh Pinah deposit.

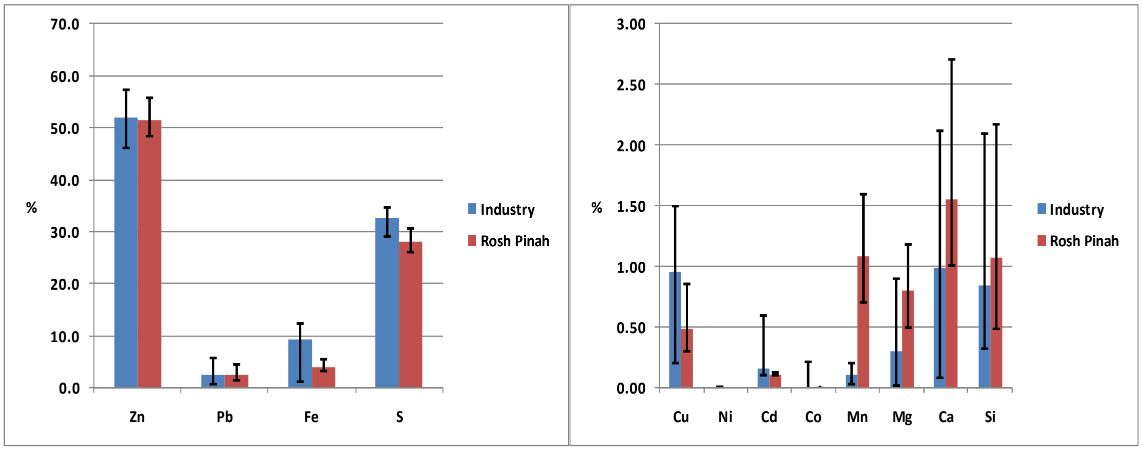



**Figure 24: Typical concentrate composition**

For this study, the selected process should be able to process a concentrate in the range of the industry min and max values presented above and with Mn, Mg, Ca and Si in the range of the Rosh Pinah concentrates. This will allow flexibility to process a wide range of concentrates.

### 4.2.1 Elimination of "other oxidative leaching" processes ($MnO_2$, $H_2O_2$, $SO_2/O_2$)

Oxidative leach processes with $MnO_2$, $H_2O_2$, $SO_2/O_2$ have all only been tested on a scale practical in the laboratory. Zinc recovery from the hydrogen peroxide and pyrolusite processes are low and together with high reagent costs these processes are highly unlikely to deliver positive returns on investment. The sulphur dioxide/oxygen system has received widespread attention in the last decade with very promising results on laboratory scale. Scale-up of this process remains a challenge due to the high gas flow requirements. These processes have not been proven on a commercial scale and will require significant development work to take them through piloting and demonstration scales, adding at least 3 years to the time for a commercial scale plant to be constructed and operating. These options are therefore discarded due to the low potential financial benefits, high technical risk and the lack of commercial development.



### 4.2.2 Elimination of chloride medium options

The challenge to process complex sulphide ore economically has driven the development of the various chloride based processes (INTEC, BHAS, Noranda, Minemet Recherché). These developments were further driven by the need to achieve residue free production processes to reduce the impact on the environment and to reduce the high disposal and closure costs associated with the conventional zinc processes. The **Noranda** process could not deliver a high quality zinc product and recovery is low and was therefore never developed past the laboratory scale. Though the inventors claim reduced operating and capital cost for the **INTEC** process with environmental benefits, it has not been developed beyond a locked-cycle batch processing pilot plant and therefore regarded as a fatal flaw.

The same concern is valid for the **BHAS** and **Minemet Recherché** processes too. Construction material is also very costly in a leach environment in chloride media at high temperature (80-100°C) and is a clear disadvantage to sulphate medium processes. These processes might have some advantages when complex sulphides have to be processed to maximise recovery of all valuable elements. Skorpion Zinc will process a typical zinc concentrate and therefore will not be considering any of these processes.

### 4.2.3 Elimination of alkaline leach options

Though the **EZINEX** process was developed commercially, it is applied to zinc oxides and not zinc sulphide concentrates and therefore has to be discarded. It is interesting though that the EZINEX process has commercialised the direct electrowinning from ammonium chloride solutions. This could potentially be used with the **CENIM-LNETI** process by eliminating the solvent extraction step, consequently reducing operating and capital costs of the CENIM-LNETI process. The latter process was also developed to treat complex sulphides and the leaching taking place at a pH of 6-7 (Figueiredo et al, 1993) makes it more attractive with regards to construction materials. Even though it will be able to process the zinc concentrate, a lack of development of this process beyond laboratory scale makes it unappealing in the Skorpion Zinc case.

### 4.2.4 Elimination of the Dowa process

The high recovery and short residence time of the **Dowa** process makes this process attractive, but like the previous processes this has not been developed on a larger or commercial scale. As mentioned before, construction of an industrial scale mill to withstand the harsh leaching conditions (sulphuric acid and high ferric concentrations) still needs to be proven and poses a high technical and financial risk. This process option is therefore discarded.

### 4.2.5 Elimination of heap bioleach

There are two options to consider when the bio-heap leach process is considered. The first is the '**Geocoat**' process from Geobiotics where the zinc sulphide concentrate is coated on an inert substrate before stacked and leached. Although it has not been commercialised to process zinc concentrates, it was applied at Agnes (South Africa) to leach refractory gold concentrates. This mine is not operating anymore. Some other projects are also under evaluation but not implemented yet. Harvey et al (2002) reported a 95% recovery from concentrates on a laboratory scale. This recovery is expected to reduce on a commercial scale due to the inherent challenges of solution distribution and percolation in heaps. It will be hard for this option, processing high grade concentrates, to compete with alternative options with high recovery and slightly higher capital investment. This option is discarded based on the limited commercial implementation and the long lead time associated with heap leach piloting and demonstration test work.

The second heap bioleach option (Teck Cominco's **HydroZinc** process) is to process the mined ore directly. This option therefore, cannot be compared directly to the other options discussed in this chapter as it does not process a concentrate. This option is therefore compared to the crushing, milling, flotation and leaching capital and operating costs. This process can only be considered as an option for Skorpion Zinc if a complete new mine is to be developed, where capital investment for a milling and concentrator plant can be offset with the capital of the heap leach and where the mine is located close to Skorpion. At present, potential feed sources are not close to Skorpion Zinc and trucking of a low grade whole ore will not be viable due to the high transportation cost - therefore this option is eliminated.

### 4.2.6 Elimination of the Bioleach process

Although the stirred tank bioleach process has not been commercialised in zinc processing, it is commercially applied in the gold industry and can therefore not merely be rejected. The role of the bacteria in the leaching process is to oxidise the ferrous to ferric for the leach of sphalerite. Sulphur oxidising bacteria, also present, oxidises the product sulphur layer formed on the sphalerite surfaces. The downside of the bioleach process is the long residence time required. The sphalerite dissolution is surface reaction rate controlled (Harvey et al, 1993; Souza et al, 2007) and the kinetics is strongly driven by reaction temperatures. This explains the long residence times required for bio-leach (typically at temperatures of 45-65°C compared to the atmospheric leach process at temperatures of 95-100°C. It further explains the reduction in residence time required when extreme thermophilic organisms are used allowing operation at temperatures of 65°C compared to the MIM Bioleach process that operates at ~45°C with moderate thermophilic bacteria. In the order of 3-5 days of residence time is required to leach a zinc concentrate with a particle size distribution of 90% passing 45micron. At 15% solids concentration and the long residence time of the bioleach process the reactor volume for the sphalerite leach is very high - roughly 3-4 times higher than the atmospheric leach. Work by Sandstrom et al (1997a) indicated that residence time can be shortened by fine grinding (90% passing 20μm). This compares to a feed size distribution of the Albion process, still with a much longer residence time than the Albion process due to the low leach temperatures. Xstrata (Hourn et al, 2005) conducted an economic trade-off study for a refractory gold ore where the Albion, Pressure Leach and Bacterial oxidation options were compared. The bacterial oxidation process presented the highest capital and operating cost.

Pyrite, the most noble sulphide mineral in the electrochemical series, also starts to oxidise under the conditions of the bioleach. All pyrrhotite is dissolved and some pyrite, resulting in a total iron dissolution between 30 and 90%, depending on the mineralogy, temperature, redox potential and leach conditions. At 65°C and redox potentials to achieve high sphalerite dissolution, between 60 and 90% iron is dissolved, depending on the residence times (Sandstrom, 1997a). This is significantly higher than the atmospheric, pressure and Albion leach processes, putting it in a less advantaged position compared to the other processes, as a higher amount of iron has to be removed from solution (most likely as a jarosite due to the high iron content and relatively lower acidic conditions – jarosite is the least preferred iron residue).

This process has a further disadvantage compared to the pressure, atmospheric and Albion processes due to the oxidation of sulphur to sulphate. The additional sulphate is removed from solution by jarosite formation and the remaining sulphates have to be precipitated.

Zinc tenor from the bio-leach process (~30g/l) is not suitable for direct electrowinning which requires zinc tenors in solution around 150-160g/l. A solvent extraction step is therefore required to upgrade the solution to become suitable for electrowinning. Although no capital cost will be required if this process is chosen for Skorpion Zinc, as an SX plant is already installed, it has higher sulphuric acid and neutralisation costs compared to conventional purification processes and hence is at a disadvantage compared to the pressure and atmospheric leach processes.

As this process has a disadvantage relative to the pressure, atmospheric and Albion leach processes, it will not be further considered as an option for Skorpion Zinc. It has an application in the gold industry where the objective is to dissolve the pyrite and other sulphide gangue to liberate the gold.

### 4.2.7 Elimination of Roast Leach Electrowinning (RLE) and Pyrometallurgical Processes.

The Imperial Smelting Furnace process (**ISF**) has been developed in 1943 (Sinclair, 2005) with only a few operating plants in the world. Zinc from this process needs to be refined to produce high quality (SHG) zinc and recovery is relatively low at 95%. The **Kivcet** process has been developed to process a high lead concentrate and it has not been proven to process zinc concentrates with low lead content. The **Ausmelt TSL** process could be applied to zinc and lead concentrates at high recovery (99%). The advantages of the ISF, Kivcet and Ausmelt processes are that iron is locked up in a compact and stable form in the slag which could be disposed without environmental risk and the processes can treat lower grade concentrates. The **RLE** process is well developed and widely applied in the industry, however, it has challenges with respect to residue disposal as an unstable jarosite or goethite is produced that needs special containment for disposal.

The disadvantage of a process producing sulphuric acid as by-product has been discussed previously (Chapter 2). A major downside of the Roast Leach Electrowinning process and the pyrometallurgical processes is the sulphuric acid that is produced. These processes are rejected based on this weakness.



### 4.2.8 Motivation for selection of the Dynatec pressure leach process

The **Dynatec** pressure leach process is a mature technology with successful implementation at various sites in the gold, platinum and base metals industries. It is also commercialised as a stand-alone two-stage pressure leach zinc process at Hudson Bay, reducing the technology risk and the requirement for extensive piloting and demonstration scale test work. The two-stage pressure leach (stand-alone) option will be applicable to Skorpion. The two stage process has significantly lower free acid in the PLS solution compared to the single stage processes developed as an add-on to other processes where the residual sulphuric acid is consumed.

One of the primary advantages of this process is that sulphur deports to elemental sulphur which can be recovered or discarded with the residue. This is a critical consideration for

Skorpion as discussed in chapter 2. This process further has high zinc recovery with a lower capital investment compared to the RLE process, and has more flexibility to treat concentrates with iron, silica and lead content. It can tolerate a higher amount of pyrite and hence could positively impact zinc recovery at the concentrators. Silicates are essentially inert to the zinc pressure leach process and concentrates with high silica content are readily treated.

The downside of this process is a higher energy cost (compared to the RLE process) as the energy to convert elemental sulphur to sulphuric acid is not harvested. It further requires pure oxygen for the process, at high cost. Although a proven technology, operating at high pressure increases safety risks and requires a higher level of attention during operation and maintenance. No fatal flaws associated with this process option are identified so far. Further development is required to establish the economics of this process.

#### 4.2.9 Motivation for selection of an Atmospheric leach process

The atmospheric leach processes were initially developed as add-on processes with a two-stage process patented by Union Minière. The single stage process has been commercially developed and applied at various sites. The key to success of the atmospheric leach processes are the specially designed reactors to enhance the air/oxygen dispersion in the slurry. The technology can be considered mature with low technical risk for implementation. This process has a high recovery with the primary advantage of producing elemental sulphur and not sulphuric acid. It has the perceived advantage over the pressure leach to have lower capital and maintenance cost due to operation at atmospheric pressures. This process also requires oxygen supply and has a net energy requirement, similar to the pressure leach process. As it operates below the melting point of sulphur, it has the advantage over pressure leach by not requiring a surfactant (such as Lignosulphonate or Quebracho). There are no fatal flaws associated with the atmospheric leach process and the economics compared to other potential options for this process will have to be evaluated.

#### 4.2.10 Motivation for selection of the Albion process.

The Albion process depends on a find grind to achieve fast reaction kinetics to achieve reasonable economics. Although this process has not been proven commercially on refractory minerals or zinc concentrates, aspects of the Albion process are being applied in industry. Fine milling technology is being applied successfully in the minerals processing industry and atmospheric leach processes for zinc sulphides have been proven to be successful. The first commercial scale project (Certej project in Romania to process a gold refractory ore – www.albionprocess.com) is due for commissioning towards the end 2011. Sphalerite is a sulphide mineral that leaches relatively easily compared to other sulphide minerals, which enables it to be leached under atmospheric conditions, as proven by Outotec and Union Minière. Test work has also proven that kinetics are enhanced by a finer particle size distribution. The solution chemistry is very similar to the pressure and atmospheric leach processes. The decisive question to be answered is whether the additional capital and operating costs for fine grinding can be offset by the reduced capital cost and operating costs of atmospheric leach by speeding up kinetics. There is no fundamental flaw but also no clear disadvantage to this process. Further development is required to establish true capital and operating costs more accurately.

#### 4.2.11 Motivation for selection of the Total Pressure Oxidation leach process

The **Total Pressure Oxidation** (**TPOX**) process is conducted at higher temperatures (220° to 230°C) and pressures than the conventional zinc pressure leach process. This results in oxidation of the sulphide sulphur and elemental sulphur to sulphuric acid in solution. Even though it is acid producing, this process presents a unique opportunity for integration with the Skorpion Zinc process. The zinc oxide leach process is an acid consuming process. Imported sulphur is burned in a sulphur burning acid plant to produce acid for the leach process. The conversion of the refinery from a zinc oxide processing plant to a sulphide plant could be phased in. Initially the oxide processing rate could be reduced and sulphides processed through a total pressure oxidation process at a rate to balance the acid requirement of the oxide leach while maintaining the annual zinc production at 150,000 tonnes. This will reduce the operating cost for the oxide ore. Once all the oxide ore is depleted, a second autoclave can be installed and reverting to a two stage moderate temperature (150°C) pressure oxidation process for sulphide processing.

Although the total pressure oxidation process has not been commercially implemented in zinc processing, it is not a unique process. Gold, copper and nickel pressure leach processes have been commercially operated at temperatures of 230°C (40bar) for many years. Technology has been developed and proven to be reliable at this temperature.

To summarise, the favoured processes for Skorpion Zinc are the ones not producing sulphuric acid as a by-product and which have been developed on a commercial scale. The pyrometallurgical process and electrolytic (RLE) processes have been eliminated as they produce sulphuric acid as a by-product. Hydrometallurgical processes in chloride media and processes involving other oxidants other than ferric in a sulphide media have been discarded based on a lack of development past the laboratory scale. Hydrometallurgical processes that involve ferric leaching in sulphate media (pressure leach, atmospheric leach and the Albion process) are favoured and will have to be considered further to establish economic viability. This category also includes stirred tank bioleach processes which have been rejected due to the diluted sulphuric acid produced and its lack of commercial application in zinc. Bio heap leach processes, typically applied to low grade materials are also rejected based on a lack of industrial development and a lack of time available to have this process ready for production at Skorpion Zinc. The Dynatec pressure leach, atmospheric leach, Albion process and the total pressure oxidation option is further developed in the following chapters.

# 5 FLOWSHEET DEVELOPMENT AND MASS & ENERGY BALANCES

From the literature analysis of various zinc production processes the Dynatec pressure leach, Atmospheric leach and the Albion process options were selected as potential options to be considered for Skorpion Zinc as they do not present any fatal flaws and compare favourably to other processes. No clear advantages or disadvantages have been identified when comparing these processes, therefore requiring further development to determine the best fit to the current infrastructure and establish the operating and capital cost models required to select one of the processes.

A phased implementation will be investigated, where one quarter of Skorpion's zinc is produced from sulphides in parallel with the oxide ore, harnessing the benefit of sulphur/sulphuric acid produced from the sulphide concentrate to reduce the processing cost of the oxide ore. When the oxide ore is depleted, the sulphide process can be expanded to produce 100% of the zinc from zinc sulphide concentrates. The parallel processing of oxides and sulphides will be referred to as 'Phase I' in this study. To simplify phase I of the study, only two process options were evaluated. Firstly the Total Pressure Oxidation process (TPOX) will be considered as it presents a unique opportunity to convert sulphide sulphur to sulphuric acid. Secondly the pressure leach process at 150°C, producing elemental sulphur, will be evaluated. This will demonstrate the viability of the process route producing and harvesting elemental sulphur for conversion in the sulphur burning acid plant. The pressure leach process option can be substituted by atmospheric leach or the Albion process in Phase I.

Phase II will refer to the process options to be selected when all zinc oxide is depleted and the refinery will have to produce all the zinc from a sulphide concentrate feed. The process options to be traded off in this phase will be the pressure leach, atmospheric leach and the Albion process.

Although the commercial pressure and atmospheric acid leach processes do not require a solvent extraction stage to upgrade the zinc, there could be a potential benefit from the solvent extraction plant as it can act as an additional buffer to prevent impurities from entering the zinc electrowinning stage. As a solvent extraction circuit is installed as part of the oxide process, no additional capital will be required. The impact of incorporating the SX with the proposed processes was evaluated by incorporating it as part of the pressure leach process for comparison. The pressure leach can be substituted with any of the atmospheric leach or Albion processes.

In summary: the process options considered and evaluated in this chapter are listed below.

Phase I: Co-treatment of sulphide concentrates and oxide ore

Option A: 1/5 Sulphides; **Pressure leach** & Sulphur Recovery

Option B: 1/5 Sulphides; **Total Pressure Oxidation** (TPOX)

Phase II: Processing only zinc sulphide concentrate

Option C: 100% Sulphides, **Atmospheric** Leach, Sulphur Recovery, Iron removal, Solution Purification and EW

Option D: 100% Sulphides, **Pressure** Leach, Sulphur Recovery, iron Removal, Solution Purification and EW

Option E: 100% Sulphides, **Pressure** Leach, Sulphur Recovery, Iron Removal, **Solvent Extraction** and EW

Option F: 100% Sulphides, **Albion** process, Sulphur Recovery, Iron Removal, Solution purification and EW.

## 5.3 MAJOR ASSUMPTIONS FOR MASS & ENERGY BALANCES AND FLOWSHEET DEVELOPMENT

The process selection must cater for a wide range of feed compositions as presented in Table 4**Error! Reference source not found.**. In order to develop a mass and energy balance and cost estimate for the trade-off study, a feed composition as presented in Table 4 is assumed. Though composition of concentrates varies from mine to mine and it will have an effect on the final capital and operating cost, the variance in the feed composition will not make a material difference on the process trade-off study.

| Mineral Composition | % | Elemental Composition | % |
|---|---|---|---|
| ZnS | 76.6 | Zn | 51.4 |
| FeS | 6.9 | Fe | 4.4 |
| CuFeS2 | 1.4 | Cu | 0.5 |
| CdS | 0.1 | Cd | 0.1 |
| MnS | 1.7 | Mn | 1.1 |
| SiO2 | 12.3 | Si | 5.7 |
| Al2O3 | 1.0 | Al | 0.5 |
| | | S | 28.9 |
| | | O | 6.6 |

**Table 4: Composition of concentrates for mass & energy balances**

The production rate of Skorpion Zinc is limited to 150,000 tpa, with the electrowinning circuit being the bottleneck. The mass & energy balances were developed to produce this quantity of final zinc. For the trade-off study between options, expansion of capacity beyond the 150,000 tpa will not be considered.

It is further assumed that the particle size distribution of the zinc concentrates is 90% passing 45 micron. This means concentrates can be processed in the various processes without further grinding, apart from the Albion process which requires a size distribution of 80% <20 micron.

## 5.4 OPTION A: PHASE I – MODERATE TEMP. (150°C) PRESSURE LEACH

### 5.4.1 Process description

In option A (Figure 25), the target zinc production is achieved by the processing of zinc oxide ore and sulphide concentrate in parallel. The ratio of zinc sulphides to oxides is driven by acid production in the sulphide process and acid consumption by the zinc oxide ore. The dotted (red) lines in Figure 25 present the sulphide processing circuit and expansions required to the oxide process. Zinc sulphide concentrate and raffinate from the SX are fed into a single stage autoclave operating at 150°C with a residence time of 1 hr.

The autoclave will be oversized - sufficient for the first stage leach in the two stage leach process for phase II when 150,000 tpa zinc is produced from sulphides. This reduces the overall capital expenditure for phase II. A single stage autoclave can be tolerated in this circuit, because sulphuric acid in the zinc rich liquor from the autoclave will be utilized in the oxide leach circuit.

At the lower operating temperature of 150°C, the sulphur from the reacted sulphide minerals is deported as elemental sulphur reporting to the discharge slurry. The autoclave discharge is thickened and the underflow is treated in the sulphur recovery circuit which consists of a sulphur flotation step where unreacted sulphides and sulphur are floated off from gangue residue. The sulphur and sulphide minerals are then separated by a melting and filtration step where clean sulphur is recovered. Around 80% of the elemental sulphur in the autoclave residue is recovered (Chalkley et al, 1993). This sulphur could be burned in the Sulphuric Acid plant to produce acid, offsetting the cost of imported sulphur.

The thickener overflow, rich in zinc (74g/l), dissolved impurities (5g/l Fe; 1g/l Cu) and sulphuric acid (24g/l) is sent to the existing oxide leach plant where the sulphuric acid is consumed by leaching the oxide ore. The sulphuric acid in this stream limits the amount of zinc that can be produced from sulphides to ~ 30 ktpa. If more sulphides are treated, the free acid tenors in the oxide leach circuit will rise above the 5g/l free acid target (even when concentrated acid addition to the oxide leach is stopped), negatively impacting on the impurity leach in the oxide circuit and increasing the cost of neutralisation.

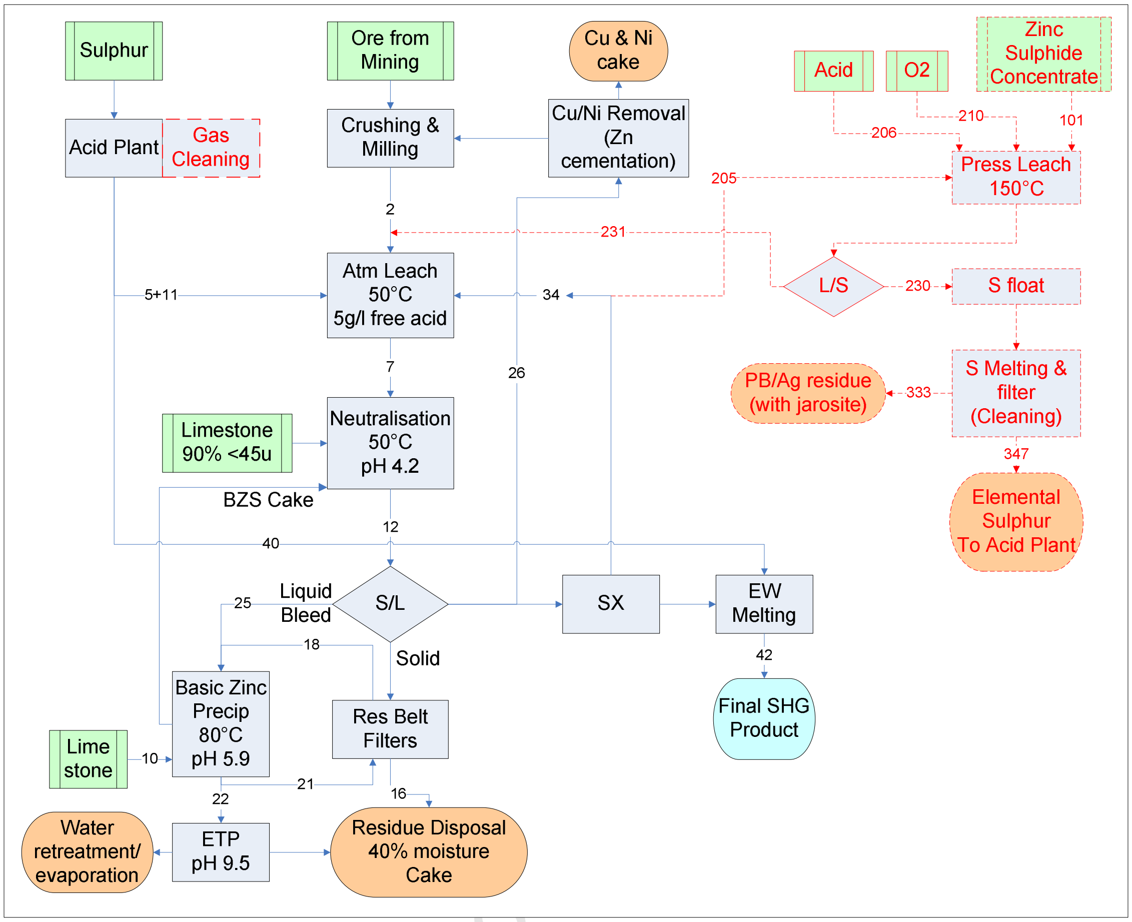



**Figure 25: Phase 1 – Option A: Pressure leach in parallel with oxide process**

(*Numbers on the diagram represent stream labels*)

### 5.4.2 Chemical reactions and extents

The following chemical reactions and extent have been used in the mass and energy balance to calculate the conversion in the autoclave:

| | |
|---|---|
| $ZnS + 0.5O_2 + H_2SO_4 \rightarrow ZnSO_4 + H_2O + S^0$ | 98% |
| $S^0 + 1.5O_2 + H_2O \rightarrow H_2SO_4$ | 5% |
| $FeS + 0.5O_2 + H_2SO_4 \rightarrow FeSO_4 + H_2O + S^0$ | 98% |
| $CdS + 0.5O_2 + H_2SO_4 \rightarrow CdSO_4 + H_2O + S^0$ | 98% |
| $MnS + 0.5O_2 + H_2SO_4 \rightarrow MnSO_4 + H_2O + S^0$ | 98% |
| $CuFeS_2 + 2.5O_2 + H_2SO_4 \rightarrow FeSO_4 + CuSO_4 + H_2O + S^0$ | 93% |
| $2FeSO4 + 0.5O2 + H2SO4 \rightarrow Fe2(SO4)3 + H2O$ | varied |
| $Fe2(SO4)3 + 3H2O \rightarrow Fe2O3 + 3H2SO4$ | varied |
| $3Fe2(SO4)3 + 14H2O \rightarrow 2H3OFe3(SO4)2(OH)6 + 5H2SO4$ | varied |

Typically below 180°C around 5% of elemental sulphur is oxidised to sulphate (Hourn et al, 1996).

The extents of the last three iron reactions were varied in the model to give a total iron concentration of 5g/l and ferrous concentration of 1.7-2.0g/l (Doyle et al, 1978; Zhang et al, 2010). The iron content in the zinc concentrate was assumed to be 4%. It has been reported by Ashman et al (1990) that a sample from the Red Dog mine, containing 4% iron, contains too little soluble iron to assist with the ferric leach of the zinc sulphides. Though the concentrate composition was selected as 4%, it was modelled to be all present as the soluble mineral pyrrhotite (FeS) and the mass balance indicates that there is sufficient iron for the zinc dissolution to take place. Iron is present in the zinc concentrates in the form of pyrrhotite, pyrite and marmatite (iron rich sphalerite). In the concentrator process some activated pyrite and pyrrhotite is floated with the sphalerite providing a source of iron for the leach reaction. Producing a concentrate with high iron content can easily be obtained and also benefit the concentrator operators as it typically results in a higher zinc recovery.

### 5.4.3 Fit to Skorpion Process

The Mass and Energy balance is presented in Appendix 1, with some of the critical parameters summarised in Table 5. The mass balance indicates that the feed rate to the oxide plant can be reduced by 20% so that roughly 30,000 tonnes of contained zinc can be fed via the zinc sulphide concentrates. The mass balance was conducted such that the zinc tenor in the PLS stream remains at 32g/l (current operating condition), by varying the circulating aqueous stream through the leach and SX circuits. This higher solution flow through the leach and neutralisation circuits results in a solids density in the neutralisation and leach circuit that is below the minimum required density to keep the solids in suspension. This could easily be rectified by modification to the agitators or operating the plant at slightly higher zinc tenors in the PLS solution, slowing down the circulating solution and increase the solids percentage. This will result in a slightly longer residence time of oxides in the leach and neutralisation circuits (roughly 20% more) with improvement of the neutralisation efficiency. Temperature in the leach and neutralisation circuits is controlled by heating the raffinate from the solvent extraction circuit. Hot solution from the pressure leach circuit will reduce the heating requirement of the raffinate circuit to a minor extent. It will not impact on the temperature control with the oxide leach temperatures remaining at 50°C.

| Stream No | | | Skorpion Design | Option A: 150°C PL |
|---|---|---|---|---|
| | **Leach and Neut** | | | |
| 1 | Ore feed to leach | t/h | 190.1 | 122.0 |
| 1 | Zn feed to leach | t/h | 17.1 | 13.6 |
| 7 | Leach % Solids | % | 10-23 | 7.3 |
| 12 | Neut % Solids | % | 10-23 | 8.8 |
| | **PLS tenor** | | | |
| 23 | Tenor | g/l | 28-40 | 32.4 |
| 23 | MaxVolume | $m^3/h$ | 1,100 | 904 |
| | **Raffinate** | | | |
| 31 | Tenor | g/l | 10-25 | 12.5 |
| 205+31 | MaxVolume | $m^3/h$ | 1,250 | 994 |
| | **BZS feed** | | | |
| 25+21 | Volume | $m^3/h$ | 240 | 163 |
| | **Cu/Cd Removal** | | | |
| 26 | Metal tenor | g/l | 0.9 | 0.5 |
| 26 | Volume | $m^3/h$ | 350 | 166 |

**Table 5: Option A: critical parameters**

The volume of PLS to the solvent extraction plant will reduce below the current operating flow rate without any negative impact on SX. Excess solution in the circuit (from gland seal water, electrowinning bleed, the wash stages in the SX plant and flocculent make-up) exit the circuit via the 40% moisture content in the tailings circuit and the bleed from the Basic Zinc Sulphate (BZS) plant (stream 22). This amount of bleed solution is limited by the design capacity of 240m³/h of the BZS plant. The solution balance indicates that the feed to the BZS plant will be 163 m³/h, which is well within the design limit.

An oxygen plant will be constructed to supply oxygen for the atmospheric leach and iron precipitation processes.

The amount of dissolved copper and cadmium that can be removed by the purification circuit (zinc cementation) is driven by the volumetric flow rate to the cementation circuit and the metal tenors. The mass balance indicates that there is adequate design capacity to remove all copper and cadmium that is dissolved from the zinc concentrates.

The movement of rare earth elements that could be present in the sulphide concentrate is not modelled due to the low quantities present. Impurity elements like germanium, arsenic, antimony; thallium is also removed in the zinc cementation circuit and precipitation in the autoclave (Doyle et al, 1978). Mercury, arsenic and selenium however find their way out of the circuit dissolved in the elemental sulphur. When this elemental sulphur is burned in the sulphur burning acid plant, it contaminates the final sulphuric acid product (Ashman et al, 1990). Sulphur contamination from the pressure leach process is assumed to be the same as the Cominco operations and presented below. It is compared to the sulphur specification for Skorpion Zinc (Table 6). Blending of the sulphur with new sulphur sourced externally will reduce the contamination, but it will still be just above the specification. The actual numbers will depend on the concentration of the impurity elements in the sulphide concentrates. For the purpose of this study it has been assumed that the sulphur contamination will be above the specified limits.

| | | Skorpion S Specification | Cominco typical |
|---|---|---|---|
| **Sulphur** | % (w/w) | 99.9 | 99.7 |
| **Ash** | % (w/w) | 0.03 | 0.03 |
| **Arsenic** | ppm | 1 | 10 |
| **Selenium** | ppm | 1 | 5 |
| **Tellurium** | ppm | 1 | 1 |
| **Mercury** | ppm | 0.3 | |

**Table 6: Comparison of Skorpion sulphur specification to Cominco typical values**

The stringent sulphur specification for Skorpion is because 25% of the sulphuric acid produced in the sulphur burning acid plant is directly transferred to the electrowinning circuit to make up electrolyte lost via the electrolyte bleed from the electrowinning circuit (Figure 23). This acid needs to be free of elements deleterious to the electrowinning process, like selenium, nickel, germanium etc. which can be tolerated up to 10 parts per billion in electrolyte. The potential for high concentrations of selenium and mercury in the sulphur will find their way to the acid produced. The impact on the design will be a gas scrubbing system to be installed to scrub the $SO_2$ gas stream prior to the catalytic converter in the acid plant to remove the contaminated elements (Hultbom, 2003). This will significantly increase the capital cost of this option.

This is a viable process option with no fatal flaw. However it requires a number of additional unit processes to recover and re-use the harvested sulphur and deal with trace contaminants.

## 5.5 OPTION B: PHASE I - TOTAL PRESSURE OXIDATION

### 5.5.1 Process description

The total pressure oxidation process is a pressure leach at high temperature (220°C), resulting in oxidation of all sulphide sulphur to sulphate and the iron to hematite. This process, in parallel with the oxide ore process, is illustrated in Figure 26, with the dotted (red) lines presenting new equipment and flows.

Zinc concentrate is pulped with heated raffinate from the solvent extraction circuit and fed to the autoclave. Concentrated acid is also dosed to ensure the acidity in the first autoclave compartment is achieved. Slurry (2% solids) from the autoclave, containing dissolved zinc (80g/l), copper (1.2g/l) and impurities (Mn 6.7g/l), hematite and excess free acid (44 g/l) is pumped at a rate of 61m³/h to the oxide leach.

Free acid from the pressure leach is used to dissolve the zinc oxide at pH of 1.8 – 2.0 and final free acid content of 5g/l. The hematite and precipitated lead will remain in a solid form and exit the circuit with the oxide leach residue. As with the moderate temperature pressure leach process, the amount of sulphides processed is limited by the free acid requirement in the oxide leach and any excess acid will increase the neutralisation cost.

The novelty of this process is that all sulphide sulphur is converted to acid and utilized in the oxide leach process without any requirement to recover sulphur or cleaning of sulphur dioxide gas streams. It is a relatively simple and straightforward process circuit. The autoclave will be oversized - sufficient for the first stage leach in the two stage leach process for phase II when 150,000 tpa zinc is produced from sulphides. This reduces the overall capital expenditure for phase II. The downside is that this autoclave will be fairly costly as it has to be designed for high temperatures and pressure for phase I, which is not required for phase II.

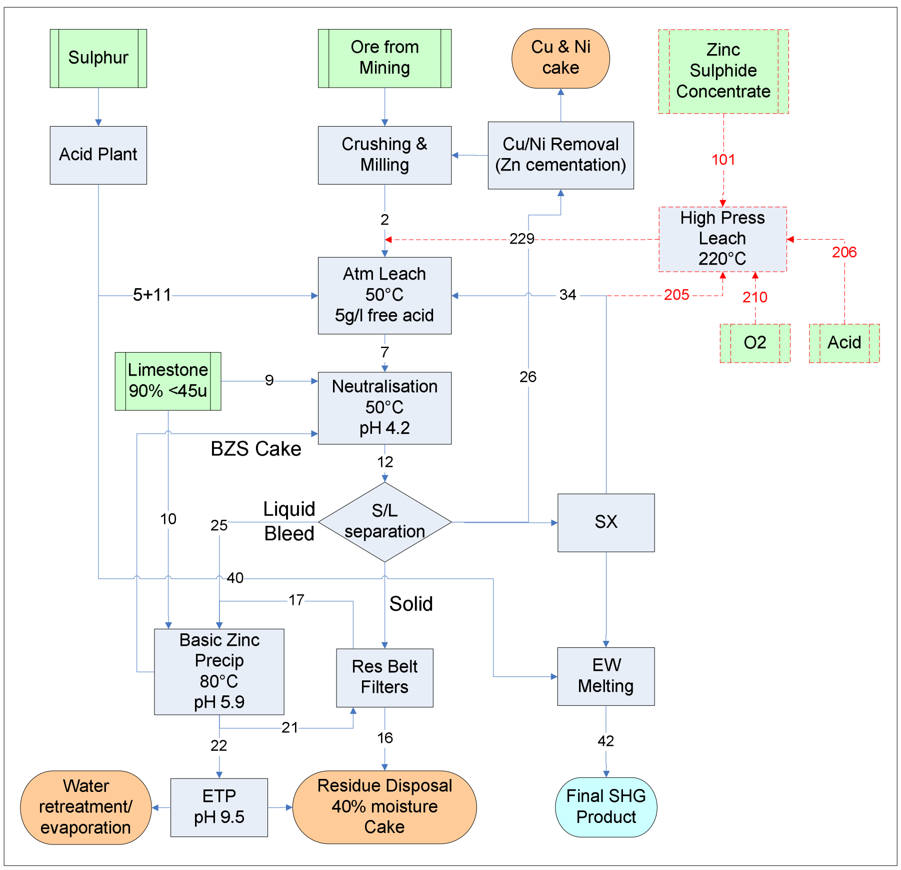



**Figure 26: Option B – Total pressure oxidation**

### 5.5.2 Chemical reactions and extents

The same chemical reactions as in Option A were considered and used to model the conversion in the autoclave. The extent of each reaction is presented below. The major difference to the 150°C pressure leach is the oxidation of all elemental sulphur to sulphate (sulphuric acid) and the conversion of most of the iron to a hematite precipitate. When pyrite is present, it will also be oxidised.

$ZnS + 0.5O_2 + H_2SO_4 \rightarrow ZnSO_4 + H_2O + S^0$ 98%

| | |
|---|---|
| $S^0 + 1.5O_2 + H_2O \rightarrow H_2SO_4$ | 98% |
| $FeS + 0.5O_2 + H_2SO_4 \rightarrow FeSO_4 + H_2O + S^0$ | 98% |
| $CdS + 0.5O_2 + H_2SO_4 \rightarrow CdSO_4 + H_2O + S^0$ | 98% |
| $MnS + 0.5O_2 + H_2SO_4 \rightarrow MnSO_4 + H_2O + S^0$ | 98% |
| $CuFeS_2 + 2.5O_2 + H_2SO_4 \rightarrow FeSO_4 + CuSO_4 + H_2O + S^0$ | 98% |
| $2FeSO_4 + 0.5O_2 + H_2SO_4 \rightarrow Fe_2(SO_4)_3 + H_2O$ | 100% |
| $Fe_2(SO_4)_3 + 3H_2O \rightarrow Fe_2O_3 + 3H_2SO_4$ | varied |

### 5.5.3 Fit to Skorpion Process

This circuit could be integrated with the oxide refinery with minimal implications. Table 7 presents some critical parameters from the mass and energy balance with the details in Appendix 2. This process option has a very similar impact on the oxide leach process to option A. The PLS zinc tenors will have to be increased slightly to maintain the minimum percentage solids in the leach and neutralisation circuits. No modifications or expansions are required for the zinc cementation circuit, BZS and tailings disposal.

An oxygen plant will be constructed to supply oxygen for the atmospheric leach and iron precipitation processes.

| Stream No | | | Skorpion Design | Option B: TPOX |
|---|---|---|---|---|
| | **Leach and Neut** | | | |
| 1 | Ore feed to leach | t/h | 190.1 | 129.0 |
| 1 | Zn feed to leach | t/h | 17.1 | 14.3 |
| 7 | Leach % Solids | % | 10-23 | 7.8 |
| 12 | Neut % Solids | % | 10-23 | 8.9 |
| | **PLS tenor** | | | |
| 23 | Tenor | g/l | 28-40 | 32.4 |
| 23 | MaxVolume | m³/h | 1,100 | 904 |
| | **Raffinate** | | | |
| 31 | Tenor | g/l | 10-25 | 12.5 |
| 205+31 | MaxVolume | m³/h | 1,250 | 994 |
| | **BZS feed** | | | |
| 25+21 | Volume | m³/h | 240 | 152.0 |
| | **Cu/Cd Removal** | | | |
| 26 | Metal tenor | g/l | 0.9 | 0.5 |
| 26 | Volume | m³/h | 350 | 175 |

**Table 7: Option B critical parameters**

Due to the lower concentrated acid consumption, the sulphur burning acid plant will be reduced to around 20-30% of designed capacity. 40% of design capacity is the minimum capacity the acid plant can be operated at. The turn-down to this minimum capacity can be achieved by the installation of baffles in the converter and modifications to the heat exchangers in the circuit. The further disadvantage of this process is that only low pressure steam (4-7 bar) will be produced and some energy needs to be vented to the atmosphere from the flash cascade at the autoclave discharge.

## 5.6 OPTION C: PHASE II - ATMOSPHERIC LEACH

### 5.6.1 Process description

The process flow for option C is presented in Figure 27. Zinc sulphide concentrate is fed to a two stage counter current atmospheric leach circuit at a rate of 34t/h. The two stage process will be applied in order to reduce the residual free acid concentration in the PLS (reducing neutralisation cost). PLS solution from the Low Acid Leach (LAL) contains ~10g/l free acid and total soluble iron of ~8g/l by careful control of acid and oxygen addition. Solids from the first stage are transferred to the High Acid Leach (HAL) stage where it is subjected to a high concentration sulphuric acid leach for another 15hrs at close to 100°C. Reaction enthalpy is sufficient to maintain temperature without any external heating required (Lahtinen et al, 2005). Solution from the HAL contains 13-14g/l Fe, ~48g/l $H_2SO_4$, ~134 g/l zinc. Solids concentration is around 9% in both leach stages and oxygen is sparged into the leach trains to oxidise ferrous to ferric. The solids from the HAL contain elemental sulphur, some unreacted sulphides (pyrite and chalcopyrite) and insoluble sulphates (lead sulphate). A total residence time of 30 hrs has been used for this process (even though 24 hrs are recommended by Lahtinen, 2005). This is to ensure the high zinc recovery of 98% is achieved at a concentrate feed particle size distribution of 90% passing 45 micron.

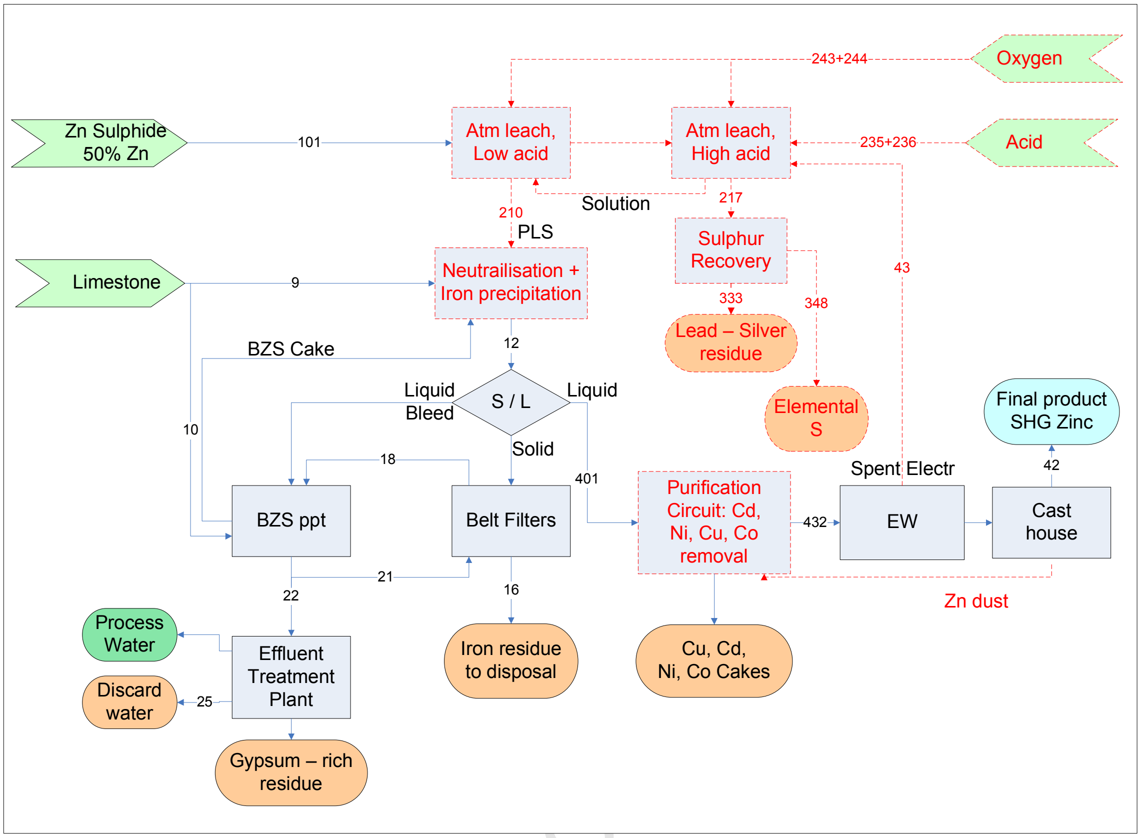



**Figure 27: Option C - Atmospheric leach of zinc concentrates**

The HAL solution is thickened and can be further processed to recover the elemental sulphur from flotation and melting processes (as discussed in section 5.4.1). The residue which is high in lead and silver can be discarded or sold, if the jarosite content is low.

The PLS solution, with zinc tenors suitable for electrowinning (156g/l) is sent to sequential purification steps. Iron is firstly removed in an iron precipitation process. Any type of iron removal (jarosite, goethite or hematite) can be applied (Haakana et al, 2008). It is assumed for this study that a goethite removal step will be used due to the availability of Basic Zinc Sulphate (BZS) from the BZS process and limestone as neutralisation reagents. Limestone is present in the oxide resource and can be mined from and milled to 90% passing 45 µm. Iron will be precipitated at a pH between 3 and 4 with oxygen sparging to oxidise the ferrous to ferric. This process typically takes place at 60-90°C. The iron content in solution is reduced to 6-10 mg/l, suitable for electrowinning. Solid precipitated iron residue is filtered and washed with a zinc free solution, provided by the BZS plant.

Solution from the iron removal process is further purified by a cold and hot conventional zinc cementation process to remove Cu, Cd, Ni, Co and rare earths, before being processed in the zinc electrowinning plant, where zinc is plated.

### 5.6.2 Chemical reactions and extents

The same chemical reactions in Option A were considered and have been used to model the leaching in the two stage atmospheric leach process. The overall extent of each reaction is presented below.

| | |
|---|---|
| $ZnS + 0.5O_2 + H_2SO_4 \rightarrow ZnSO_4 + H_2O + S^0$ | 98% |
| $S^0 + 1.5O_2 + H_2O \rightarrow H_2SO_4$ | 5% |
| $FeS + 0.5O_2 + H_2SO_4 \rightarrow FeSO_4 + H_2O + S^0$ | 98% |
| $CdS + 0.5O_2 + H_2SO_4 \rightarrow CdSO_4 + H_2O + S^0$ | 98% |
| $MnS + 0.5O_2 + H_2SO_4 \rightarrow MnSO_4 + H_2O + S^0$ | 98% |
| $CuFeS_2 + 2.5O_2 + H_2SO_4 \rightarrow FeSO_4 + CuSO_4 + H_2O + S^0$ | 85% |
| $2FeSO_4 + 0.5O_2 + H_2SO_4 \rightarrow Fe_2(SO_4)_3 + H_2O$ | varied |
| $Fe_2(SO_4)_3 + 4H_2O = 2FeO.OH + 3H_2SO_4$ | varied |
| $3Fe_2(SO_4)_3 + 14H_2O \rightarrow 2H_3OFe_3(SO_4)_2(OH)_6 + 5H_2SO_4$ | 0% |

The extent of the iron reactions are varied compared to the targeted ferric and total iron concentrations in solution within each stage of the leach. Although it has been reported that iron can be precipitated as jarosite in the atmospheric leach (van Put et al, 1999; Lahtinen, 2005) under certain circumstances, jarosite formation is more favourable at higher temperatures. It has been assumed that no hydronium jarosite is formed at atmospheric conditions (last reaction above).

### 5.6.3 Fit to existing infrastructure

New equipment is represented by the dotted (red) lines in Figure 27 with a detailed mass & energy balance appended (Appendix 3). The oxide crushing and milling circuit will become obsolete and will be replaced with a concentrates receiving and re-pulping section. As specially designed reactors are required for the concentrate atmospheric leach process, the existing agitated leach tanks cannot be utilised and will be replaced. A new sulphur recovery and cleaning process will be installed.

An oxygen plant will be constructed to supply oxygen for the atmospheric leach and iron precipitation processes.

The sulphuric acid consumption will be 4.6 t/h, which is about 10% of the design capacity of the existing sulphur burning acid plant - well below its turn-down capacity. This consumption equates to 3-4 acid tankers per day, and it is assumed that it will be sourced and trucked in from Gauteng. A trade-off study to construct a low capacity acid plant is recommended for a later stage. Offloading and storage facilities do exist.

Iron removal, especially the goethite removal described in literature, is typically conducted at temperatures between 60°C and 95°C. This is primarily to achieve high reaction rates and reduces equipment sizing. In most of the processes the feed solution is already at high temperatures (from the leach processes) and no additional heating is required. In the proposed atmospheric leach circuit the LAL solution feeding the precipitation circuit will be around 70-78°C. The existing neutralisation equipment is designed for a temperature up to 55°C and therefore not be suitable for the neutralisation process. The reaction kinetics of the goethite iron removal process at 55°C need to be better understood to assess if the existing equipment could provide sufficient residence time for this process at lower temperatures, before a final conclusion on existing equipment can be made. The second challenge will be the high cost (capital and operating) required for cooling the solution down to 55°C to use the existing equipment and then heating it back up to 85°C for the following hot zinc cementation circuit. For the purpose of this study it was therefore assumed that the existing equipment will not be used, and the cost of a new iron removal circuit will be included.

The mass balance indicates that the flow rate of iron free PLS to the purification circuit is 177m$^3$/h with a copper tenor of 0.9 g/l. The existing copper cementation circuit is designed for a flow rate of ~350m$^3$/h and solution tenor of 0.88 g/l. This indicates that the existing copper removal circuit is sufficient. However, the existing hot purification circuit (for Ni and Co removal) is sized for 70m$^3$/h and will have to be upgraded to process the targeted 177 m$^3$/h.

The iron residue to be disposed of is estimated to be 22 dry tonnes per hour with significantly lower filtration fluxes than the current residue. The existing residue filtration is modular and able to process 195t/h material. The existing system will be able to process the projected iron residue. The BZS circuit will provide wash solution for the iron residue as well as the lead-silver residue. The projected required feed rate to the BZS circuit is around 61 m$^3$/h, which is within the turn-down ratio of the BZS plant. No capital is therefore required for modification of the BZS circuit.

Iron and the lead-silver residue may contain heavy metals that could be mobilised during high rainfall years. For the purpose of this study it has been assumed that all residues will be disposed using lined facilities. The selection of the goethite process above the jarosite will require a smaller footprint and reduce the disposal capital requirements.

## 5.7 OPTION D: PHASE II - DYNATEC PRESSURE LEACH

### 5.7.1 Process description

The Dynatec pressure leach (150°C) process flow sheet (Figure 28) is very similar to the atmospheric leach process flow. The leach process will also be conducted in a two-stage counter-current flow to achieve a lower final acid concentration of 8g/l and 2g/l dissolved iron. Zinc tenor in the LAL discharge slurry is 160g/l, sufficient for electrowinning without upgrading. The HAL solution contains 30g/l acid and 15g/l dissolved iron (Barth et al, 1998). The process chemistry of pressure leach and atmospheric leach is very similar, with the main difference in reaction kinetics and stability regions of jarosites. At elevated temperatures, the reaction kinetics are significantly faster (Harvey et al, 1993; Butinelli et al, 1992). At 150°C an overall residence time of around 2 hours (~ 1hr per stage) is sufficient to achieve 98%

dissolution of zinc sulphides. The downside is that special equipment (autoclave) is required to operate at elevated temperatures to keep water in the liquid state. This equipment requires high capital and skills to maintain. The other downside of high pressure leaching is scaling that takes place in the autoclaves, resulting in bi-annual shutdowns to clean the autoclaves. To achieve high availabilities, typically a third standby autoclave is constructed and piped in to be used either as LAL or HAL stage. This further increases the capital cost. The advantage though is a smaller footprint and associated civil and structural costs.

As with the atmospheric leach process, the PLS will go through an iron removal step and purification circuit to remove metal impurities before zinc is plated in an electrowinning step. Residue from the HAL stage can go through sulphur recovery and cleaning steps with a lead-silver residue that can be disposed of. Elemental sulphur could be stored separately or sold to interested buyers.

Heat required for the pressure leach process and iron removal step is supplied by the exothermic reactions.

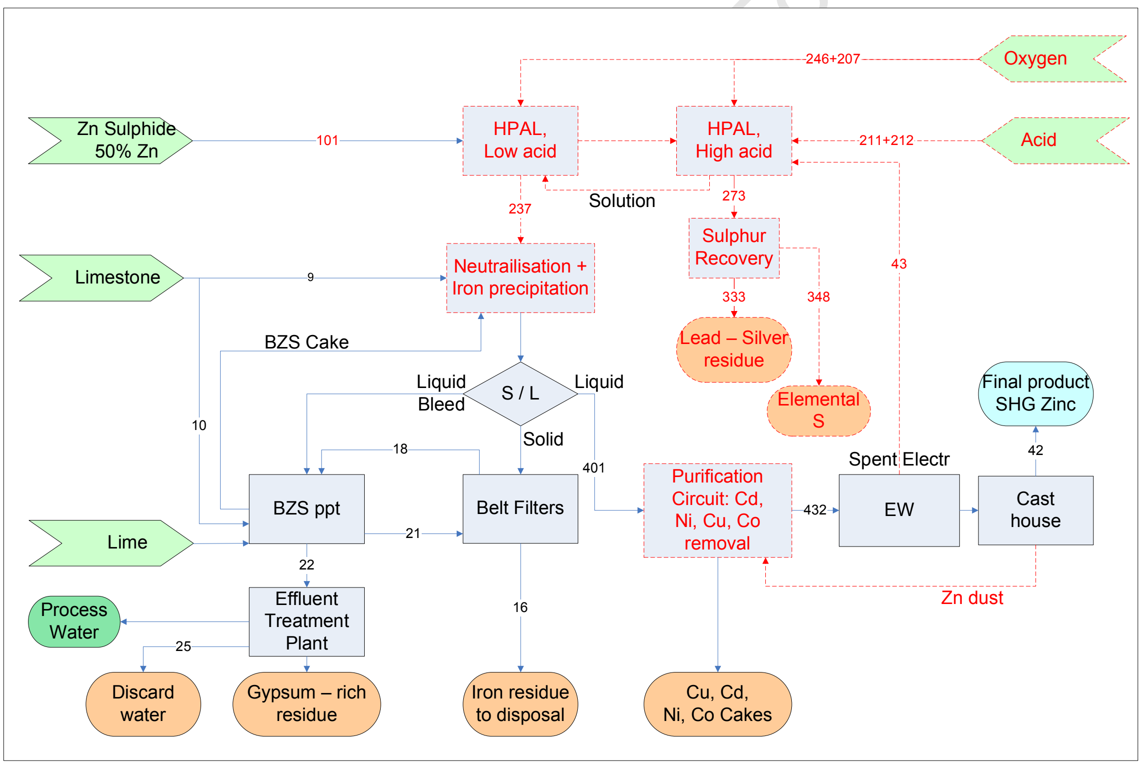



**Figure 28: Option D - 150°C Pressure leach of zinc concentrates**

### 5.7.2 Chemical reactions and extents

The chemical reactions for the pressure leach at 150°C are similar to the atmospheric leach reactions discussed previously. The only exception is the allowance for some jarosite to form in the pressure leach reaction. The stability of jarosites increases with increasing temperatures (Claassen, 2004; Doyle et al, 1989). The last reaction below was varied to precipitate around 20% of the iron (Krysa, 1995) as a jarosite.

$ZnS + 0.5O_2 + H_2SO_4 \rightarrow ZnSO_4 + H_2O + S^0$ *98%*

$S^0 + 1.5O_2 + H_2O \rightarrow H_2SO_4$ *5%*

$FeS + 0.5O_2 + H_2SO_4 \rightarrow FeSO_4 + H_2O + S^0$ *98%*

$CdS + 0.5O_2 + H_2SO_4 \rightarrow CdSO_4 + H_2O + S^0$ *98%*

$MnS + 0.5O_2 + H_2SO_4 \rightarrow MnSO_4 + H_2O + S^0$ *98%*

$CuFeS_2 + 2.5O_2 + H_2SO_4 \rightarrow FeSO_4 + CuSO_4 + H_2O + S^0$ *85%*

$2FeSO_4 + 0.5O_2 + H_2SO_4 \rightarrow Fe_2(SO_4)_3 + H_2O$ *varied*

$Fe_2(SO_4)_3 + 3H_2O \rightarrow Fe_2O_3 + 3H_2SO_4$ *varied*

$3Fe_2(SO_4)_3 + 14H_2O \rightarrow 2H_3OFe_3(SO_4)_2(OH)_6 + 5H_2SO_4$ *varied*

### 5.7.3 Fit to existing infrastructure

The mass balance of the pressure leach (Appendix 4) is very similar to the atmospheric leach process. Table 8 below compares some selected parameters. Apart from the capital costs for the leach section, the remaining equipment requirements will be similar to those discussed for the atmospheric leach in section 5.6.3.

| Stream No | | | Skorpion Design | Option C: Atm L | Option D: Pres L |
|---|---|---|---|---|---|
| | **Acid consumption** | | | | |
| 204/211 | Sulphuric acid | t/h | 44.0 | 4.6 | 2.1 |
| | **PLS** | | | | |
| 210/237 | Zn tenor | t/h | | 160 | 160 |
| | **Iron Removal** | | | | |
| 210/237 | Leach Residue Fe tenor | g/l | | 8 | 2 |
| | Fe Removal temp | °C | 55 | 70-78 | 70-78 |
| | **BZS feed** | | | | |
| 21+22+307 | Volume | $m^3/h$ | 240 | 64 | 38 |
| | **Cu Removal** | | | | |
| 210/237 | Metal tenor | g/l | 0.9 | 0.9 | 1.0 |
| 210/237 | Volume | $m^3/h$ | 350 | 177 | 144 |
| | **Tailings** | | | | |
| 16+304 | Iron Removal and HAL Solids | dry t/h | 195 | 36 | 25 |

**Table 8: Selected parameters from option C & D**

## 5.8 OPTION E: PHASE II - PRESSURE LEACH COMBINED WITH SX

### 5.8.1 Process description

The leach section of this process is the same pressure leach process described in section 5.7.1, with the difference that spent electrolyte is not returned to the autoclave, but rather the raffinate from the extraction circuit. PLS from the leach section will go through an iron removal step and residue through a wash step before disposed on a lined tailings facility. A more detailed mass balance is given in Appendix 5.

As in the Skorpion Oxide process, Cu, Cd, Ni, Co will build up in the circuit and a removal step needs to be introduced. This can be achieved by either via a bleed circuit or by processing the total PLS stream, depending on the allowable copper tenor in PLS. When a certain threshold copper tenor in PLS is reached, there is some carry-over to electrolyte via entrainment. For this study, it was assumed that the threshold copper concentration is 0.9g/l. In this case, the full PLS stream will have to be processed through the impurity removal process.

Solids from the pressure leach circuit can go through an optional sulphur recovery step as described previously and a lead-silver residue will be discarded with the iron residue.

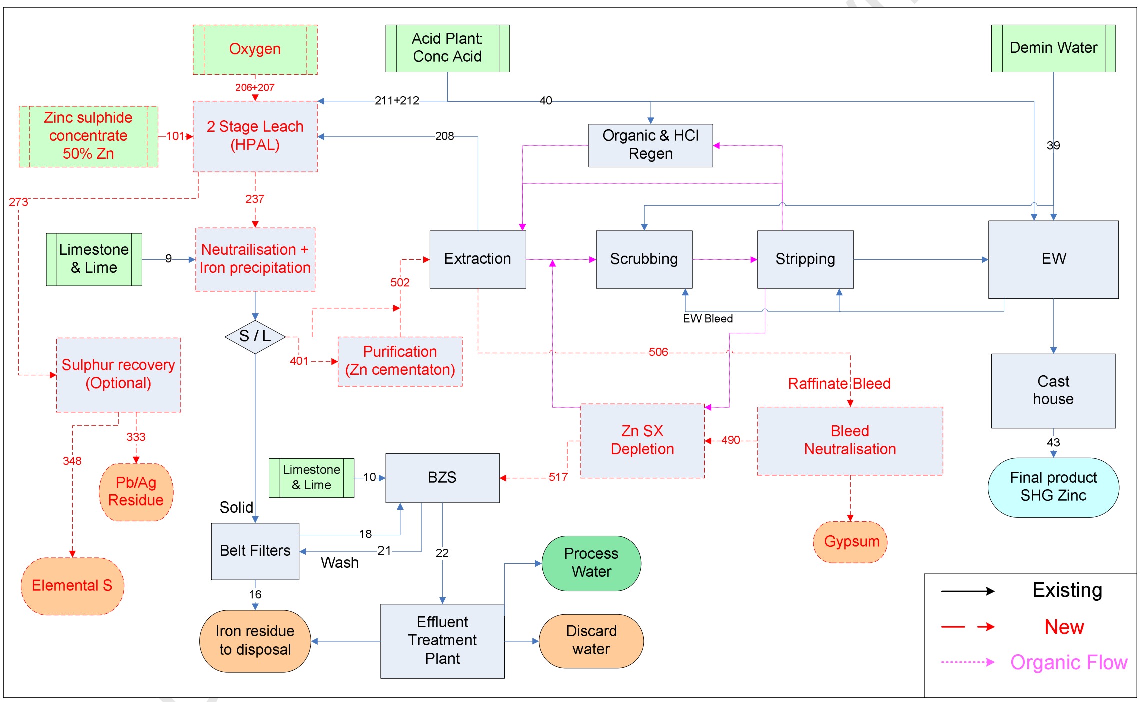



**Figure 29: Option E – Pressure leach with SX of zinc concentrates**

Zinc rich solution (145g/l Zn) from the iron removal/purification circuit is extracted in the solvent extraction, leaving a raffinate with a zinc tenor of ~55.4g/l. The expected percentage extraction is scaled up linearly from the current operating circuit. Isotherms for extraction at elevated temperatures and PLS zinc tenors will have to be developed if a more accurate estimate is required. For the purpose of this study the linear scale-up will be sufficient. Zinc is stripped from the loaded electrolyte and recovered in a conventional zinc electroplating process. As described in section 4.1.5, electrolyte has to be bled from the electrowinning process to prevent a build-up of impurities in the closed circuit electrolyte solution. Lost electrolyte is replaced with demineralised water and chemically pure sulphuric acid.

Excess water and sulphuric acid present in the circuit, mainly from the electrolyte bleed, scrubbing stages and the acid requirement for the organic regeneration necessitates a liquid bleed from the process. This is done by scavenging zinc from a raffinate bleed stream in a zinc depletion SX and a final polishing step in the BZS circuit. From here the solution is bled from the circuit. Excess sulphuric acid is removed by a neutralisation step before the zinc SX depletion step. Acid consumption could potentially be reduced by 50% by using alternative technology (like resin or evaporative regeneration) to regenerate the hydrochloric acid used to remove iron from the stripped organic. The acid requirement to replace bled electrolyte will remain.

### 5.8.2 Chemical reactions and extents

The same chemical reactions and extents as for the Pressure Leach option (section 5.7.2) will be applicable. The organic load and stripping reactions are presented below:

$Zn^{2+}(aq) + jk\ 2RH(org) \rightarrow R_2Zn(org) + nH^+(org)$

$R_2Zn(org) + nH^+(org) \rightarrow Zn^{2+}\ (aq) + 2RH(org)$

### 5.8.3 Fit to existing infrastructure

Table 9 below compares some of the selected parameters of options D and E with the Skorpion process design. Some of the existing infrastructure - residue disposal, BZS, Tailings and the solvent extraction circuit can be utilised. The PLS temperature from the iron removal and purification steps will be significantly higher than the operating temperature (~33°C) of the SX. Operating the SX at elevated temperatures may be feasible if alternative diluents with a higher flashpoint can be sourced. The challenge is that construction materials will have to be upgraded as the existing equipment will not be suitable for elevated temperatures. This will require extensive downtime to modify the existing equipment and increase the capital cost. For the purpose of this study it was decided to allow for cooling of the PLS in the capital and operating cost estimates.

| Stream No | | | Skorpion Design | Option D: Press L | Option E: Pres L & SX |
|---|---|---|---|---|---|
| | **Acid consumption** | | | | |
| 211+212+40 | Sulphuric acid | t/h | 44.0 | 2.1 | 8.0 |
| | **PLS** | | | | |
| 237 | Zn tenor | t/h | | 160 | 156 |
| | **Iron Removal** | | | | |
| 237 | Leach Residue Fe tenor | g/l | | 2 | 2 |
| | Fe Removal temp | °C | 55 | 70-78 | 70-78 |
| | **BZS feed** | | | | |
| 21+22+307 | Volume | $m^3/h$ | 240 | 38 | 35 |
| | **Cu Removal** | | | | |
| 237 | Metal tenor | g/l | 0.9 | 1.0 | 0.9 |
| 237 | Volume | $m^3/h$ | 350 | 144 | 165 |
| | **Tailings** | | | | |
| 16+333+348+348 | Iron Removal and HAL Solids | dry t/h | 195 | 25 | 59 |

**Table 9: Selected parameters from option D & E**

Acid consumption of this circuit at 8 t/h is roughly 18% of the design capacity of the existing acid plant. This is significantly below the turn down ratio of 30-40%. Significant modifications to the acid plant will have to be done in order to produce sulphuric acid. It was assumed for this study that the acid will be sourced from the Gauteng region and trucked in. It will however require quite a large number of trucks (~6 per day).

The only benefit from this circuit is to provide an additional buffer for impurities, allowing more consistent production of SHG zinc. This will however add much more complexity to the design and operations of the plant. The high sulphuric acid consumption and associated neutralisation and tailings disposal costs make this option very unattractive.

## 5.9 OPTION F: PHASE II – ALBION LEACH

### 5.9.1 Process description

The process flow of the Albion process is presented in Figure 30. It is very similar to the atmospheric leach process (option C) with the difference of a milling step before the leach. Received concentrates (90% passing 45μm) are milled to ~90% passing 20μm. The objective is to improve kinetics and reduce residence time in the leaching process, hence reducing the capital required. Leach reactors will also require less power input compared to the Outotec reactors as a more conventional stirred tank reactor with hydrofoil impellers with lower power input can be used. Oxygen is sparged into the tanks using a supersonic oxygen injection lance in combination with the hydrofoil impellers to achieve the required mass transfer and leaching rate.

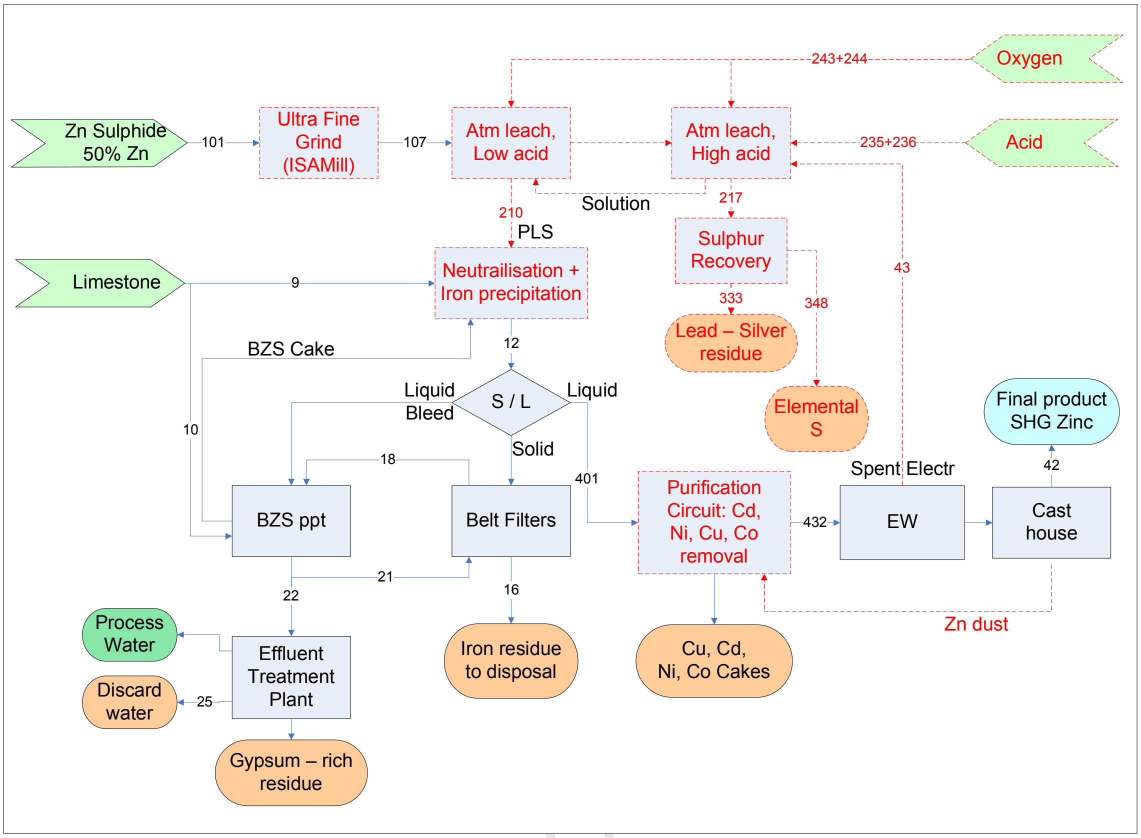



**Figure 30: Option F – Albion process flow diagram**

### 5.9.2 Chemical reactions and extents

The leaching process for the Albion process is very similar to the atmospheric leach process and the same reactions and rates (section 5.6.2) have been assumed for the Albion process. The only difference is the required residence time that is reduced to 24hrs to achieve the same reaction extents.

### 5.9.3 Fit to existing infrastructure

When considering the fit with existing infrastructure and equipment, the same arguments as for the Atmospheric leach (option C) prevail for the Albion leach (option F). The existing oxide ore ball mill is not suitable for the ultra-fine grinding and will have to be replaced with the ISAmill technology provided by Xstrata Technology.

## 6 CAPITAL AND OPERATING COST ESTIMATES

The final and most important step in the selection process is to determine and compare the financial viability of each option. The process selected will determine the long-term competitiveness of Skorpion in the industry. The best option will be a low capital expenditure and low production cost model which will position Skorpion in the lower half of cost producers. Once the plant is constructed, it is extremely costly to change the process for a lower cost option. Careful consideration is therefore imperative at this point in the project lifecycle. Determining the financial parameters requires the development of an operating and capital cost model for each option. The process followed and results are presented and discussed in this chapter.

## 6.1 OPERATING COST

The operating costs are discussed and compared for each of the two phases separately.

### 6.1.1 Phase 1 – Parallel processing of oxides and sulphides

During Phase 1 oxide ore and sulphide concentrates will be processed in parallel, roughly at a zinc feed ratio of 4:1 oxides to sulphides. The reagent consumptions determined from the Mass and Energy balances of option A and B are presented in Table 10 and compared to the current oxide operation.

| | | Oxide Process | OPTION A 150° PO | OPTION B 220° PO | |
|---|---|---|---|---|---|
| **Reagents** | **Units** | **Flow** | **Flow** | **Flow** | **Stream number** |
| **Sulphuric Acid** | t/hr | 27.2 | 22.1 | 13.4 | 5+11+32+40+206 |
| **Limestone** | t/hr | 36.3 | 31.5 | 27.7 | 9+10 |
| **Hydrated lime (CaOH2)** | t/hr | 0.8 | 1.1 | 0.9 | 19+318 |
| **Oxygen** | t/hr | | 1.8 | 5.6 | 210 |
| **Organic diluent** | $m^3$/a | 1,593 | 1,593 | 1,593 | |
| **Extractant** | $m^3$/a | 103 | 103 | 103 | |
| **Zn Dust** | t/hr | 0.1 | 0.1 | 0.1 | 27 |
| **Flocculant (total)** | kg/hr | 41.0 | 34.2 | 45.1 | |
| **Lignosulphonate** | kg/hr | | | 21.4 | |
| **Flotation Reagents** | kg/hr | | 0.1 | | |
| **Water** | $m^3$/h | 71.0 | 71.0 | 71.0 | 3+6+13+15+33+39+610 |
| **Demineralised Water** | $m^3$/h | 66.0 | 66.0 | 66.0 | 46 |
| **Steam** | t/hr | - | 4.3 | 5.3 | Calculation |

**Table 10: Reagent consumptions for options A & B**

The values indicate a healthy 50% reduction in sulphuric acid consumption for option B (Total pressure oxidation process) compared to the existing oxide process. Option A indicates a reduced acid consumption, but not with the same benefit as option B. This is due to the fact that not all sulphide sulphur is converted to elemental sulphur (chalcopyrite dissolution is only 85%) in option A and more significantly the fact that only 80% of the elemental sulphur is recovered during sulphur flotation and melting process.

Limestone consumption for options A and B is also reduced as result of a lower volume of solution that needs to be treated in the Basic Zinc Sulphate process. This volume is driven by the solution balance in the circuit and the tonnes of final residue that need to be washed. Option B has the lowest limestone consumption.

The pressure oxidation circuits require injection of oxygen as oxidant for the ferric/ferrous leaching couple to leach the sphalerite and to oxidise the elemental sulphur to sulphate (sulphuric acid). The total pressure oxidation has a higher oxygen consumption driven by the additional oxidation of the elemental sulphur.

Total water consumption and other reagent consumptions remains very similar between the various options, with flotation reagents required for sulphur flotation in option A and lignosulphonate for option B.

The mass and energy balances indicated a steam consumption of 24t/h and 19t/h for option A and B respectively. Option B's consumption is lower due to more energy harvested in the autoclave from the oxidation of elemental sulphur. The steam consumption presented in the mass and energy balance (Table 10) has been offset by steam available from the sulphur burning acid plant for each option. The net steam consumption of option B is slightly higher than option A due to a higher amount of 'low quality' steam produced in the autoclave which is vented to atmosphere after heat recovery is maximised.

The operating costs of options A & B were calculated and compared to the operating cost of the existing oxide refinery (each of the options with a total zinc production of 150,000 tonnes per annum). The cost presented is for the refinery section of Skorpion Zinc only. The mining and comminution operating costs have been excluded from this calculation and the data presented.

The reagent cost is calculated using actual prices of reagents consumed at Skorpion Zinc. Lignosulphonate, flotation reagents and oxygen prices were obtained from recent quotes from suppliers. The oxygen price is based on an 'over the fence' supply by a contractor. The limestone cost as a percentage of the total cost is relatively low. Limestone is mined from the oxide resource as waste and only the actual crushing and milling cost included in the analysis below. Consumables consist mainly of anodes and cathodes for the electrowinning plant and filter cloths for the refinery. The fixed cost portion includes actual labour, management, administration, logistics and maintenance costs. The new sulphide process will require additional labour, which is included in the 'labour variance' line. Option B has more unit operations and therefore the higher labour cost. The maintenance cost for options A and B was factored in (2% of direct capital cost).

| *Cost as % of total Oxide Process cost* | Oxide Process | OPTION A | OPTION B |
|---|---|---|---|
| Sulphuric Acid | 9.8 | 7.9 | 4.8 |
| Limestone | 1.6 | 1.4 | 1.2 |
| Hydrated lime (CaOH2) | 0.8 | 1.0 | 0.8 |
| Oxygen | - | 0.5 | 1.5 |
| SX reagents | 1.4 | 1.4 | 1.4 |
| Zn Dust | 2.0 | 2.0 | 2.0 |
| Water | 0.4 | 0.4 | 0.4 |
| Demineralised Water | 0.6 | 0.6 | 0.6 |
| Steam | - | 0.5 | 0.6 |
| Labour variance | - | 0.5 | 0.2 |
| Maintenance variance | - | 1.9 | 0.4 |
| Electricity | 16.0 | 16.0 | 16.0 |
| Reagents other | 3.8 | 3.7 | 4.0 |
| Consumables | 5.9 | 5.9 | 5.9 |
| Fixed Cost Incl Labour & Maintenance | 57.8 | 57.8 | 57.8 |
| Total | 100.0 | 101.5 | 97.6 |

**Table 11: Phase I operating cost comparison**

It is clear from the data presented in Table 11 that there is a marginal improvement in operating costs for option B and very similar costs for option A. The savings made in sulphuric acid is mainly offset by oxygen costs, additional labour requirements and the net steam consumption. This operating cost is sensitive to the input price of sulphur. The costs above are based on a sulphur price of US$ 43/t sulphur, currently the long term outlook. In the short term, there is variation in the sulphur price of up to US$ 200/t S. The sensitivity of potential savings is presented in Figure 31. It indicates that when the sulphur price increases to US$200 / t S, the saving is more attractive at 8%.

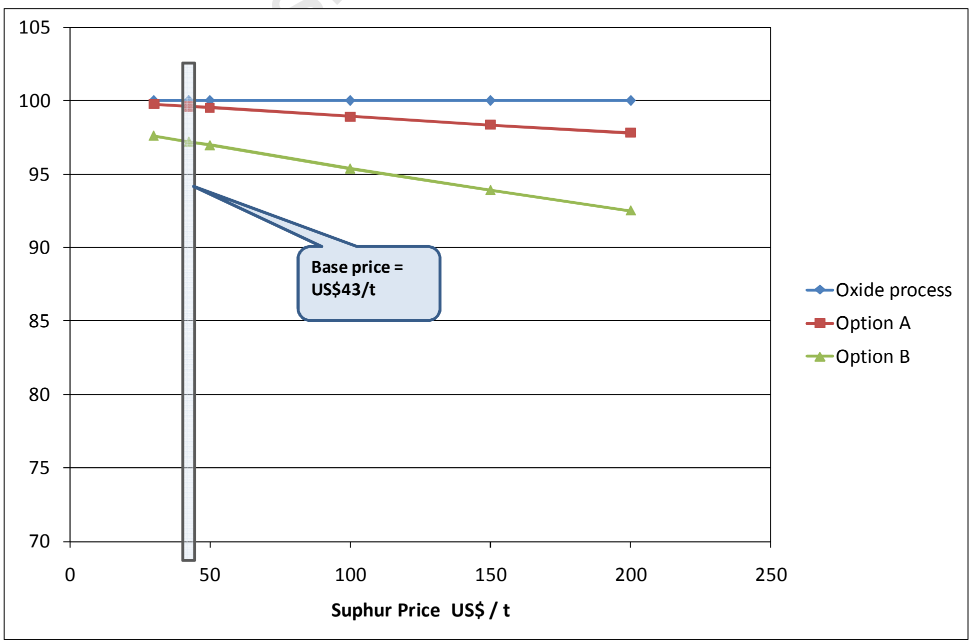



**Figure 31: Effect of sulphur price on operating cost saving – Phase 1**

From the analysis of the operating costs it can be concluded that option B is a more favourable process compared to option A. The long term projected price is around US$ 50-100 per tonne of sulphur. In this range there is no more than a 3-5% saving in the operational costs compared to the parallel processing of oxides and sulphides. This is a very small margin that will not meet the hurdle rate of a 10% saving required to justify the capital expenditure.

### 6.1.2 Phase II – Operating cost for processing of sulphide concentrates only.

This option considers the conversion of the refinery to process sulphide concentrates when the oxide resource is depleted. This comprises options C, D, E and F. The reagent consumptions of the various options are presented in Table 12.

| | | OPTION C Atmospheric L | OPTION D Pressure L | OPTION E Pres L & SX | OPTION F Albion | |
|---|---|---|---|---|---|---|
| **Reagents** | **Units** | **Flow** | **Flow** | **Flow** | **Flow** | **Stream number** |
| **Sulphuric Acid** | t/hr | 4.6 | 2.1 | 12.7 | 4.6 | 235+236 / 212+242 / 32+40+212+242 |
| **Limestone** | t/hr | 12.5 | 6.4 | 30.9 | 12.5 | 9+10 / 9+10+481 |
| **Hydrated lime (CaOH2)** | t/hr | 0.6 | 0.5 | 0.5 | 0.6 | 19+318 |
| **Oxygen** | t/hr | 7.6 | 7.6 | 7.6 | 7.6 | 243+244 / 207 + 246 |
| **Organic diluent** | $m^3/a$ | - | - | 1,593 | - | |
| **Extractant** | $m^3/a$ | - | - | 103 | - | |
| **Zn Dust** | t/hr | - | - | - | - | |
| **Flocculant (total)** | kg/hr | 15.4 | 14.5 | 22.1 | 15.4 | |
| **Lignosulphonate** | kg/hr | - | 79.5 | 80.7 | - | |
| **Flotation Reagents** | kg/hr | 0.5 | 0.5 | 0.5 | 0.5 | |
| **Water** | $m^3/h$ | 73.7 | 26.3 | 26.3 | 73.7 | Calculated |
| **Demineralised Water** | $m^3/h$ | - | - | 107.6 | - | 522+39 |
| **Steam** | t/hr | 9.8 | 7.1 | 9.9 | 9.8 | Calculated |
| **Grinding media** | kg/t | - | - | - | 0.5 | Calculated |

**Table 12: Reagent consumptions for phase II options**

Sulphuric acid consumption for the atmospheric leach (option C) is twice the consumption of the pressure leach (option D). This is a direct result of the leach conditions selected for the two processes – specifically the outlet acid, ferrous and ferric concentrations. Reaction extent for both processes was the same with the exception of jarosite, which was allowed to form in the pressure leach process, also impacting on the acid consumption. Jarosite precipitation liberates acid and reduces the overall acid consumption. The process outlet tenors for sulphuric acid, ferric and ferrous chosen were sourced from literature. Extensive research and development were conducted and published on the pressure leach process since it was introduced in the 1980's. The atmospheric leach process was only commercially introduced in the late 1990's with comparatively less research and actual data published. Lathtinen et al (2005) reports that the conditions in the atmospheric leach can be manipulated to achieve a wide range of acid tenor and redox potential in the atmospheric leach process. This can very well be achieved by the variation of the oxygen injection and slurry density in the leach. However, for the purpose of this study, parameter selection was based on published data. The mass and energy balance indicates that reagent consumption is very sensitive to the actual leaching kinetics and conditions and the form of iron precipitation.

The acid consumption for option E (pressure leach with SX) is six times the acid consumption for option D, a major disadvantage of this process option. This is primarily a result of the water requirement for the washing stages of SX and the electrolyte bleed that consumes acid. Demineralised water is used to top up the water losses in the SX and EW circuits of option E – further increasing the cost of this option.

The limestone requirement for option E follows the high acid consumption, as the refinery bleed stream (to balance Mn tenors and total solution) has a higher acid tenor, which has to be neutralised. Likewise, the higher limestone requirement for option C compared to option D is due to higher sulphuric acid tenor in the solution to the neutralisation section.

Oxygen consumption for all the options in the mass balance was assumed to be the same. The number reported in Table 12 is based on the stoichiometric requirement. In reality, the oxygen conversion efficiency will differ between the atmospheric and pressure leach. Although Lathtinen et al (2005) mentioned that oxygen consumption in the pressure leach is higher than in the atmospheric leach, it is expected that the oxygen will have lower efficiency due to the ambient operating pressure and the higher volumes of the atmospheric reactors. It has been assumed that the oxygen conversion rate for the atmospheric leach will be 87% of the conversion rate in the pressure leach reactors. This has been accounted for in the figures presented in Table 13.

Flocculant consumption for option E is higher than option C and D due to the additional raffinate bleed neutralisation section.

Although it is reported in literature that both the atmospheric and pressure leach processes operate autothermally due to the heat released by the exothermic reactions, there is a steam requirement in the sulphur melting process, the hot purification circuit and to pre-heat solutions to the leach sections. Option E (with SX) has an additional heat requirement due to the cooling of PLS before the SX circuit that results in a heating requirement of the raffinate before it is fed back to the pressure leach circuit. The variation in steam requirement between option D and E is a result of the reaction extents and solution chemistry differences.

The calculated operating costs for each option are presented in Table 13. Again, it is expressed as a percentage of the total cost of option C. Around 55% of the costs are locked in fixed costs which include maintenance, labour, management, administration and support services. These were calculated/derived using actual operating costs at Skorpion Zinc.

For option F, the grinding cost presented includes energy consumption (assuming 25kWh/t energy requirement), grinding media cost and maintenance cost. The maintenance cost includes wear items like the shell, disks and other wear items.

Zinc dust consumption is indicated as zero. Zinc dust will be produced internally and circulated back to the purification circuit, with no requirement to buy in zinc dust.

| *Cost as % of option C operating cost* | OPTION C Atmospheric L | OPTION D Pressure L | OPTION E Pres L & SX | OPTION F Albion |
|---|---|---|---|---|
| **Sulphuric Acid** | 10.0 | 4.6 | 27.7 | 10.0 |
| **Limestone** | 4.7 | 2.4 | 11.6 | 4.7 |
| **Hydrated lime (CaOH2)** | 0.7 | 0.6 | 0.6 | 0.7 |
| **Oxygen** | 2.8 | 2.4 | 2.4 | 2.8 |
| **SX reagents** | - | - | 1.7 | - |
| **Zn Dust** | - | - | - | - |
| **Water** | 0.6 | 0.2 | 0.2 | 0.6 |
| **Demineralised Water** | - | - | 1.1 | - |
| **Steam** | 1.4 | 1.0 | 1.5 | 1.4 |
| **Electricity** | 18.8 | 18.8 | 18.8 | 18.8 |
| **Reagents other** | 0.4 | 0.7 | 0.9 | 0.4 |
| **Consumables (Anodes & Cathodes)** | 4.3 | 4.3 | 4.3 | 4.3 |
| **Labour variance** | - | - | 0.6 | 0.3 |
| **Maintenance cost** | 9.0 | 9.9 | 12.9 | 9.0 |
| **Fine Grind variance** | - | - | - | 1.2 |
| **Fixed Cost Incl Labour, admin, overheads** | 47.2 | 47.2 | 47.2 | 47.2 |
| **Total** | **100.0** | **92.2** | **131.7** | **101.5** |

**Table 13: Phase II operating cost comparison**

The major contributors to the variable costs are electricity, sulphuric acid, limestone and consumables (anodes and cathodes). The reagent cost estimate is based on consumption rates presented and recent quotations from suppliers.

Option D has the lowest operating cost, roughly 8% below that of atmospheric leach (option C). This is primarily driven by the assumptions made for the leaching conditions and iron precipitation. Maintenance cost is expected to be lower for the atmospheric leach equipment as specialised components and brick lining are not required for the atmospheric reactors. This is reflected in the factored maintenance cost estimate (2% of capital cost).

## 6.2 CAPITAL COST

### 6.2.1 Process followed to establish the capital cost

Finally the capital cost required for each process option was established. Process Flow Diagrams were developed from the block flow diagrams for each unit process included in the options investigated and a mechanical equipment list was compiled. Sizing of all mechanical equipment was done according to the mass and energy balance calculations. Construction materials were specified based on the temperature and process conditions. The mechanical equipment cost was determined from HATCH Africa's database of recently (post 2001) completed projects.

The direct capital costs were then factored from the mechanical equipment cost estimate and include mechanical fabrication and installation, electrical equipment, control and instrumentation, earthworks, civil works, piping and structural steel. Indirect costs were factored from the direct costs and make allowance for project management and detailed design costs and include the following: EPCM and owners team cost, first fills, commissioning costs, freight and insurance cost. Accuracy of the capital cost estimate is ± 30%.

The atmospheric leach circuit (option C) cost estimate includes 12 reactors (7 for LAL and 5 for HAL), each with 5m diameter and 25m high to achieve the 30hrs residence time. The atmospheric leach tanks are equipped with 90kW and 75kW motors for the LAL and HAL respectively. Some uncertainty exists on the mechanical costs for the atmospheric leach reactors. The reactors of Outotec and Union Minière are both specialised and proprietary equipment and actual costs are not available in the open domain. Costs were estimated based on the reactor dimensions and assumptions on the construction materials. It is known that the agitators require a high power input and the aforementioned power requirements were assumed. One spare reactor was included for each of the HAL and LAL stages.

For the capital cost estimate of the pressure leach (option D and E) circuits, three carbon steel brick lined autoclaves were included - two operational and a third one as standby. The standby autoclave can be tied into the circuit into the place of any one of the HAL or LAL autoclaves. The standby autoclave is to ensure a plant availability of 98% is achieved despite the maintenance stops required for brick lining and de-scaling. The size required to achieve a retention time of 60 min per autoclave is estimated at 4.5m diameter and 22.5 m long, 4 compartments with 5 agitators (45kW each).

The capital cost for the Albion process (option F) included an ISAmill for the fine grind duty and atmospheric leach tanks. The M3000 model ISAmill with 1.1 MW drive was selected for this process. It was sized based on a conservative power consumption estimate of 35kWh/t to grind from $P_{90}$ of 45μm to 20 μm. A percentage runtime of 85% was assumed based on actual experience within AngloAmerican Platinum. The capital estimate for the Albion atmospheric leach circuit was based on a total of 11 reactors (including 2 stand-by reactors) and power duty of 72kW and 60kW drives for the LAL and HAL respectively. A conventional reactor with length to diameter ratio of 1:1 was assumed for the costing.

The zinc concentrate handling, iron removal, hot and cold purification circuits were all assumed to have the same duty for options C-F. Table 14 presents the duty of all the various unit operations for the capital cost estimates. The sulphur recovery circuit for phase I, option A was sized for treatment of 300,000 tonnes of concentrate to be processed during phase II of the project. This was done to minimise overall capital expenditure should a phased approach be proven viable.

| *Design Duty: unit processes* | | OPTION A | OPTION B | OPTION C | OPTION D | OPTION E | OPTION F |
|---|---|---|---|---|---|---|---|
| | | Pres Leach | TPOX | Atm Leach | Press Leach | Press. L & SX | Albion |
| Oxide leach | dry t/h ore | 122 | 128 | - | - | - | - |
| Leach | dry t/h concentrates | 9 | 8 | 34 | 34 | 34 | 34 |
| Sulphur Recovery | t/h solids | 16 | - | 16 | 16 | 16 | 16 |
| Purification | $m^3$/h solution | - | - | 177 | 177 | 195 | 177 |
| Bleed neutralisation | $m^3$/h solution | - | - | - | - | 23 | - |
| Fe removal | $m^3$/h slurry | 1249 | 1237 | 145 | 145 | 165 | 145 |
| SX | t/h Zn | 17 | 17 | - | - | 17 | - |

**Table 14: Design duty for unit processes**

### 6.2.2 Capital cost summary

The capital cost estimate for Phase 1 (options A & B) is presented in Table 15. It is expressed as a percentage of the total cost of option B. The total cost of option A is almost 5 times higher than option B. This is due to the additional unit processes required to recover and re-use the elemental sulphur. A significant amount of capital is required to install a $SO_2$ gas cleaning process to remove the Se and Hg. The advantage of a simplified process step (single autoclave) on the capital requirements is clearly evident from the capital cost estimate.

| *Cost as % of Option B total cost* | OPTION B | OPTION A |
|---|---|---|
| | 220℃ Pres L | 150℃ Pres L |
| ZnS concentrate solids handling | 8 | 8 |
| Pressure leach | 58 | 55 |
| Sulphur flotation | | 42 |
| Sulphur melting | | 29 |
| Reagent supply and Acid Plant Modifications | | 190 |
| **DIRECT COSTS** | **66** | **324** |
| **INDIRECT COSTS** | **34** | **164** |
| **TOTAL COSTS** | **100** | **488** |

**Table 15: Capital costs for options A & B**

The phase II options' capital is expressed as a percentage of the total cost of option C for discussion purposes and presented in Table 16. The total capital cost for option D is around 22% more expensive than option C. This is a result of the higher capital cost of the autoclaves and auxiliary equipment compared to the atmospheric leach reactors. Integration of the solvent extraction circuit to the conventional process route (option D) has a further disadvantage of capital cost, around 41% higher than the atmospheric leach option. This is owing to the additional equipment required to neutralise the bleed stream and additional zinc depletion SX circuit. Capital required for the Albion process (option F) is very similar to that of the atmospheric leach process. The additional capital for the ISAmill is offset by a 20% reduction in the residence time and hence cost of the leach reactors. The marginal reduction in agitation power also contributes to a reduction in capital cost.



| *Cost as % of Option C total cost* | OPTION C | OPTION D | OPTION E | OPTION F |
|---|---|---|---|---|
| | Atm Leach | Press Leach | Press. L & SX | Albion |
| ZnS concentrate solids handling | 1 | 1 | 1 | 5 |
| Leach | 16 | 28 | 28 | 14 |
| Sulphur flotation | 9 | 9 | 9 | 9 |
| Sulphur melting | 7 | 7 | 7 | 7 |
| Purification | 18 | 18 | 20 | 18 |
| Bleed neutralisation | | | 8 | |
| Fe removal | 10 | 10 | 12 | 10 |
| SX | | | 5 | |
| Reagent make-up and supply | 5 | 6 | 5 | 5 |
| **DIRECT COSTS** | **67** | **79** | **94** | **67** |
| **INDIRECT COSTS** | **33** | **43** | **46** | **34** |
| **TOTAL COSTS** | **100** | **122** | **141** | **101** |

**Table 16: Capital costs for phase II options (C, D, E, and F)**

Buban et al (2000) compared the direct and atmospheric leach to each other. In their estimate, the pressure leach capital cost is between 1.08 and 1.25 times higher than the atmospheric leach. This study estimates the pressure leach capital cost at 1.75 times the atmospheric leach cost (leach sections only), higher than the estimate by Buban et al. They compared the cost of a single stage atmospheric and pressure leach process, whereas this study investigates a two stage leach processes and includes a standby autoclave, which Buban et al have not considered. The assumptions made on the cost estimate of the proprietary equipment supplied by Outotec must also be noted with the potential impact on the results.

## 7 CONCLUSION AND RECOMMENDATION OF A SUITABLE PROCESS

The electrolytic zinc process (Roast Leach Electrowinning) dominated the zinc production towards the end of the 20th century, compared with the other pyrometallurgical processes. As illustrated in Figure 32, this is primarily due to low operating cost and the high purity of the final product produced. Acid production from the RLE process and a reduction in concentrate quality (higher iron, copper and silica content) drove the development of alternative hydrometallurgical processes towards the turn of the century. The primary criteria for a suitable process for Skorpion Zinc includes favourable economics (low capital investment and operating cost with a high zinc recovery), a process that does not produce sulphuric acid as by-product and does not possesses high investment and technical risk.

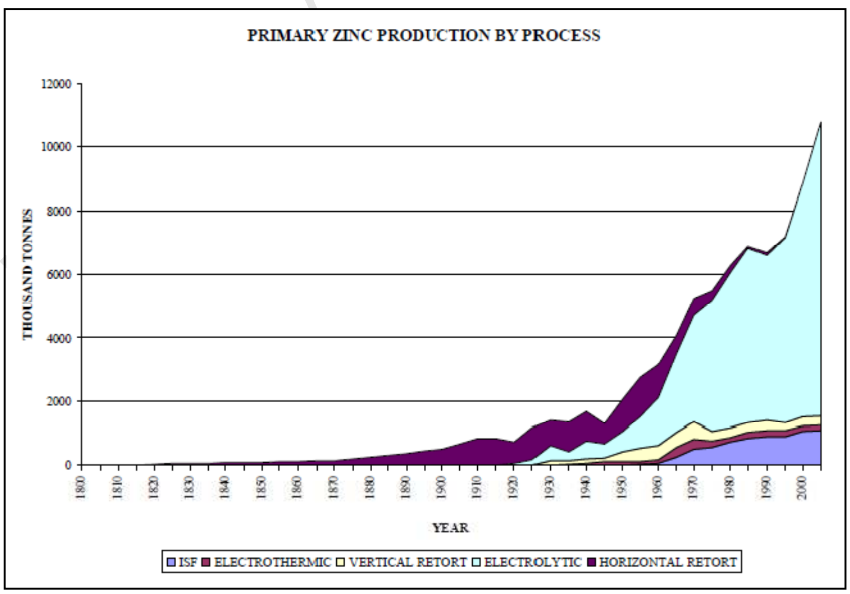



**Figure 32: Primary zinc production by process (Brook Hunt, 2004)**

A review of all zinc producing processes compared with the requirements of Skorpion Zinc firstly eliminated all the pyrometallurgical processes and the RLE process that produces sulphuric acid as by-product. The alkaline, chloride medium and oxidative leach processes in sulphide medium (excluding ferric leach) are eliminated due to the low development status – none of these processes have been commercialised and would require additional development work past the laboratory stage, piloting and demonstration scale. This will add additional technical risk to the project and extend past the deadline to have the processing capacity installed by the end of the oxide ore resource life. The bio-leach processes have been discarded as they produce a dilute acid and the PLS tenors from the bioleach is not suitable for direct electrowinning. A Solvent extraction step would be required to upgrade the zinc tenors, increasing the operating cost significantly.

Ferric leaching in sulphate media is the hydrometallurgical process that has gained popularity over the past ~20-30 years with expansion of zinc production capacity in existing plants and in the context of the development of new deposits. The enhanced kinetics of dissolving sphalerite with ferric ions in sulphate solutions above oxidation directly with oxygen is well known. During the sphalerite leaching in ferric sulphate media the four main rate processes taking place are (Baldwin et al, 1995) a) oxygen mass transfer from the gas phase to the leach liquor b) dissolution of the mineral sulphide c) oxidation of ferrous to ferric and d) precipitation of ferric as jarosite or iron oxides. The leaching process follows a shrinking core model with surface reaction control initially and diffusion layer control later in the process (Harvey et al, 1993; Souza et al, 2007). The kinetics are primarily influenced by temperature, particle size, ferric concentration and oxygen concentration, (Dreisinger et al, 1990; Harvey et al, 1993; Baldwin et al, 1995; Demopoulos et al, 1999; Souza et al, 2007).

The various ferric leach processes developed aimed to benefit from thes various parameters that drive the reaction kinetics. Temperature is one of the primary drivers for the reaction kinetics with an exponential increase in reaction rates with increasing temperatures from 60-150°C or beyond, also evident from the decrease in residence time from around 72-100hrs to 1-2hrs within this temperature range. Pressure leaching takes advantage of operating at temperatures where all the sphalerite will dissolve with a minimum oxidation of elemental sulphur. This however comes at a substantial capital expense required for the autoclaves. Outotec and Union Minière have designed reactors, especially to ensure sufficient oxygen supply to the liquid phase so that the sphalerite dissolution remains surface reaction rate limiting at atmospheric pressures. The Albion process takes advantage of increased available surface area to enhance the reaction rates, leaching in atmospheric conditions without the need for specially designed reactors. The increased surface area most likely is an advantage during later stages of the leach reaction where the reaction becomes product layer diffusion limiting. However, the Albion process suffers from the additional cost incurred by having to achieve a fine grind.

The dissolution of sphalerite is also influenced by the amount of iron present in the sphalerite matrix (Crundwell, 1988). The dissolution rate increases with the increase of iron content (first order to the concentration of iron in the solid matrix). This is attributed to the reduction in activation energy in the rate limiting charge transfer step on the surface of the minerals. Iron content in the sphalerite matrix could vary considerably for different ore deposits as illustrated in Table 17.

| Sphalerite | Fe in (Zn,Fe)S % |
|---|---|
| Black Mountain | 8.02 |
| Rosh Pinah | 1.79 |
| Gamsberg | 8.62 |

**Table 17: Iron content in sphalerite (Zn,Fe)S mineral phase**

Even though there is a high variation, it will change the capital cost estimates of each option by the same ratio. Operating costs will remain the same for each of the options. This will therefore have an immaterial effect on the final process selection, especially within the accuracy of the capital cost estimate.

When considering a phased implementation process, this study has indicated that at the long term forecasted sulphur price (US$ 50-100 / t sulphur), the potential saving is not sufficient to justify the capital expenditure. Saving on the acid consumption of the oxide process is offset by oxygen consumption and the maintenance cost. The phased approach will therefore not be further considered.

The option to consider inclusion of the solvent extraction circuit (option E) to reduce risk of 'off spec' zinc production cannot be justified. There is no benefit/reduction in the capital cost if the SX is included and more importantly, it increases operating cost by ~40% against the lowest operating cost option (option D). This additional operating cost is not justified by the reduction in risk, at all.

The options on the table for Skorpion Zinc are therefore the atmospheric leach, pressure leach and the Albion processes. The capital and operating costs for the three options are presented in Table 18, expressed as a percentage of the atmospheric leach option.

| *Cost as % of Option C total cost* | **OPTION C** | **OPTION D** | **OPTION F** |
|---|---|---|---|
| | **Atm Leach** | **Press Leach** | **Albion** |
| **Operating Cost** | **100.0** | **92.2** | **101.5** |
| **Capital Cost** | **100.0** | **122.4** | **101.0** |

**Table 18: Cost comparison of Options C, D, and F**

The analysis indicates that the pressure leach has a lower operating cost compared to the other options. The operating cost comparison between these options is highly sensitive to the process chemistry with regards to redox potential (ratio of ferric to ferrous), acidity of leach stages and the precipitation of iron during the leach process. The estimate of additional milling costs for the Albion process indicates it has a marginal increase in overall costs and is not preferred above the atmospheric leach option. The pressure leach has a disadvantage due to the increased capital costs.

Options C and D compare equally favourably against the criteria set in chapter 2, with option F at some disadvantage. All three processes have high zinc recovery which is very dependent on the variability of concentrate size distribution. The atmospheric leach and pressure leach are preferred over the Albion leach process as the latter has not been applied commercially. Even though the ultrafine grind is well established commercially (~96 mills operational worldwide), leaching of a fine grind in conventional stirred tank reactors has not been applied commercially on zinc concentrates and will increase the risk. This will require further development work. The Albion process will therefore not be considered further. However, the need for a fine particle size distribution might be required by either the pressure leach or atmospheric leach processes depending on the mineralogy and reaction kinetics. In such a case the ISAmill technology could be considered. It is expected for the atmospheric leach process to be more sensitive to the particle size distribution than the pressure leach due to leaching under milder temperature.

If the economics are considered for the two process options C and D, at a zinc price of US$2000/t zinc produced and a life of 20 years, the Net Present Value of both options are equal. However, the uncertainty on the assumptions made for the leach conditions of the atmospheric leach process has a much higher impact on the NPV of the two options than variations in economic conditions (price, escalation and resource life). Whilst both options are attractive and suitable for conversion of the Skorpion Zinc refinery to process sulphide concentrates, a higher degree of uncertainty exists on operating and capital costs for the atmospheric leach process.

The pressure leach option (D) appears to be more attractive due to the higher number of standalone processes commercialised and a lower sensitivity to the size distribution of the zinc concentrates. However, the atmospheric leach option cannot be discarded based on the findings of this study. The atmospheric leach process' operating cost is highly sensitive to the leach parameters (residual ferric, ferrous and acid concentration) selected for the mass and energy balances. Atmospheric leach is relatively new and little information on the process is published and unpublished variations could exist. In addition, some uncertainty exists on the concentrate particle size distribution required for the atmospheric leach process, as conflicting data is reported in literature. Comparative test work by Buban et al (2000) indicates that a finer grind is required to obtain high recovery under atmospheric leach conditions, whilst Lahtinen et al (2005) reports that a concentrate particle size of ~90% passing 44 μm is sufficient to obtain a high recovery within 24-30 hrs.

This study therefore concludes that ferric leaching of sphalerite in a sulphate medium under atmospheric or pressure leach conditions are both suitable for Skorpion Zinc. These processes present advantages over all other zinc processing options when elemental sulphur is preferred as the final deportment for sulphide sulphur. Both these processes can be integrated into the existing refinery and will be able to process concentrates with a wide composition range.

The project development time to commercialisation and first zinc production will be similar for both processes. Pilot scale test work will have to be conducted for each process to obtain accurate design parameters for the specific blend of concentrates to be processed. Detailed design and construction time will also be similar for the two options. It is estimated that a feasibility study, detailed engineering and construction process will be in the order of 4 years. This makes the target delivery date of 2015 very optimistic. With potential extension of the oxide resource to 2016, the conversion of the refinery to process zinc sulphide concentrate will be achievable.

In order to select one process above the other, it is recommended to conduct a bench scale atmospheric leach test work and to engage with Outotec (technology supplier) to establish more firm numbers on process parameters for the atmospheric leach. Reaction kinetics under the various leach conditions for atmospheric leach needs to be established. Further test work on a laboratory scale is also recommended to establish the sensitivity of recovery and residence time on the particle size distribution of both processes. This is to be conducted for a sphalerite that contains high and low quantity of iron in the sphalerite mineral matrix. Information from the bench scale test work and an update of the economic models will have sufficient accuracy, enabling Skorpion to select either the pressure or the atmospheric leach process. A single option can then be considered in the next phase of project development and implementation.

Sufficient capital and operating cost data have been generated as input to a separate marketing and financial study to evaluate the option to proceed with the conversion of Skorpion to process zinc concentrate or to close the refinery.

# 8 LIST OF REFERENCES


1. Adams, R.W. and Matthew, I.G., "Leaching of Metal Sulphide Concentrates at Atmospheric Pressure Using SO2/O2 Mixtures", Proc. Australasian Institute of Mining and Metallurgy, 280, 41-53, 1981.
2. Allen, C., Kondos, P., Payant, S., Van Weert, G., and Van Sandwijk, A., "Production of Zinc Oxide from Complex Sulfide Concentrates Using Chloride Processing", PCT World Patent WO 01/25497, April 12, 2001.
3. Ashman, D. W. and Jankola, W. A., "Recent Experience with Zinc Pressure Leaching at Cominco", pp. 253-275, in : Lead-Zinc '90, Eds. :T. S. Mackey and R. D. Prengaman, TMS, Warrendale, PA, 1990.
4. Aydogan, S., "Dissolution kinetics of sphalerite with hydrogen peroxide in sulphuric acid medium", Chemical Engineering Journal, vol 123, pp. 65-70, 2006.
5. Baldwin, S. A.; Demopoulos, G. P. and Papangelakis, V. G., "Mathematical Modeling of the Zinc Pressure leach Process", Metall. Trans. Vol. 26B , pp. 1035-1047, October 1995.
6. Barth, T.R; Hairu, A.T.C and Meier, T.P. "The operation ofhte HBM&S zinc pressure leach plant", Zinc and Lead Processing, The Metallurgical Society of CIM, editors J. E. Dutrizac, J. A. Gonzalez, G. L. Bolton and P. Hancock, 811-823, 1998.
7. Bartlett, R. W., "Economic Trade-offs in Increasing Revenue from a Massive Complex Sulfide Deposit-Caribou", pp. 783-799, in : Complex Sulfides, Processing of Ores, Concentrates and Byproducts, Eds. : A. D. Zunkel, R. S. Boorman, A. E. Morris and R. J. Wesely, TMS-AIME, Warrendale, PA, 1985.
8. Berezowsky, R. M., Collins, M. J., Kerfoot, D. G. and Torres N., "The Commercial Status of Pressure Leaching Technology, JOM, pp. 9-15, February 1991.
9. Bhat, K. I. and Natarajan, K. A., "Recovery of Zinc from Leach Residues - Problems and Developments", Trans. Ind. Inst. Met., Vol. 40, No. 4, pp. 361-372, August 1987.
10. Boon, M., Heijnen, J.J, "Mechanisms and rate limiting steps in bioleaching of sphalerite, chalcopyrite and pyrite with thiobacillus ferrooxidans", Biohydrometallurgical Technologies, The Minerals, Metals and Materials Society, pp. 217-235, 1993.
11. Buban, K.R., Collins, M.J., Masters, I.M., Trytten, L.C., "Comparison of direct pressure leaching with atmospheric leaching of zinc concentrates", Lead-Zinc 2000, The Minerals, Metals and Materials Society, pp 727-738, October 22-25, 2000.
12. Buckett, G.A. and Sinclair, R.J., "The BUKA Zinc Process: A 21st Century Standard", Proceedings of Zinc and Lead Processing, Eds. J.E. Dutrizac et al, CIM, 579-596, 1998.
13. Buttinelli, D., Lavecchia, R., Pochetti, F., Geveci, A., Guresin, N., Topkaya, Y., "Leaching by ferric sulphate of raw and concentrated copper-zinc complex sulphide ores", International Journal of Mineral Processing, vol. 36, pp 245-257, 1992.
14. Chalkley, M.E., Ozberk, E., Vardill, W.D., "The treatment of bulk concentrates by the Sherritt zinc pressure leach process", Minerals Engineering, vol6, No. 8-10, pp. 937-948, 1993.
15. Claassen, J.O. Rennie, J., Van Niekerk, W.H., Meyer, E.H.O., Howard, G., Vegter, N.M., " Implementation of Belt Filtration to improve soluble zinc recovery", The South African Institure of Minign and Metallurgy, Conference Recovery of base metals in Southern Africa, 16-18 July 2002.
16. Collins, M.J., Stiksma, J., Buban, K.R., Masters, I.M., "Pressure acid leaching of zinc and copper concentrates by Dynatec", Proceedings of Sessions and Symposia sponsored by the Extraction and Processing Division of The Minerals, Metals & Materials Society held during the 2000 TMS annual meeting in Nashville, Tennessee, p597-605, March 12-16, 2000.

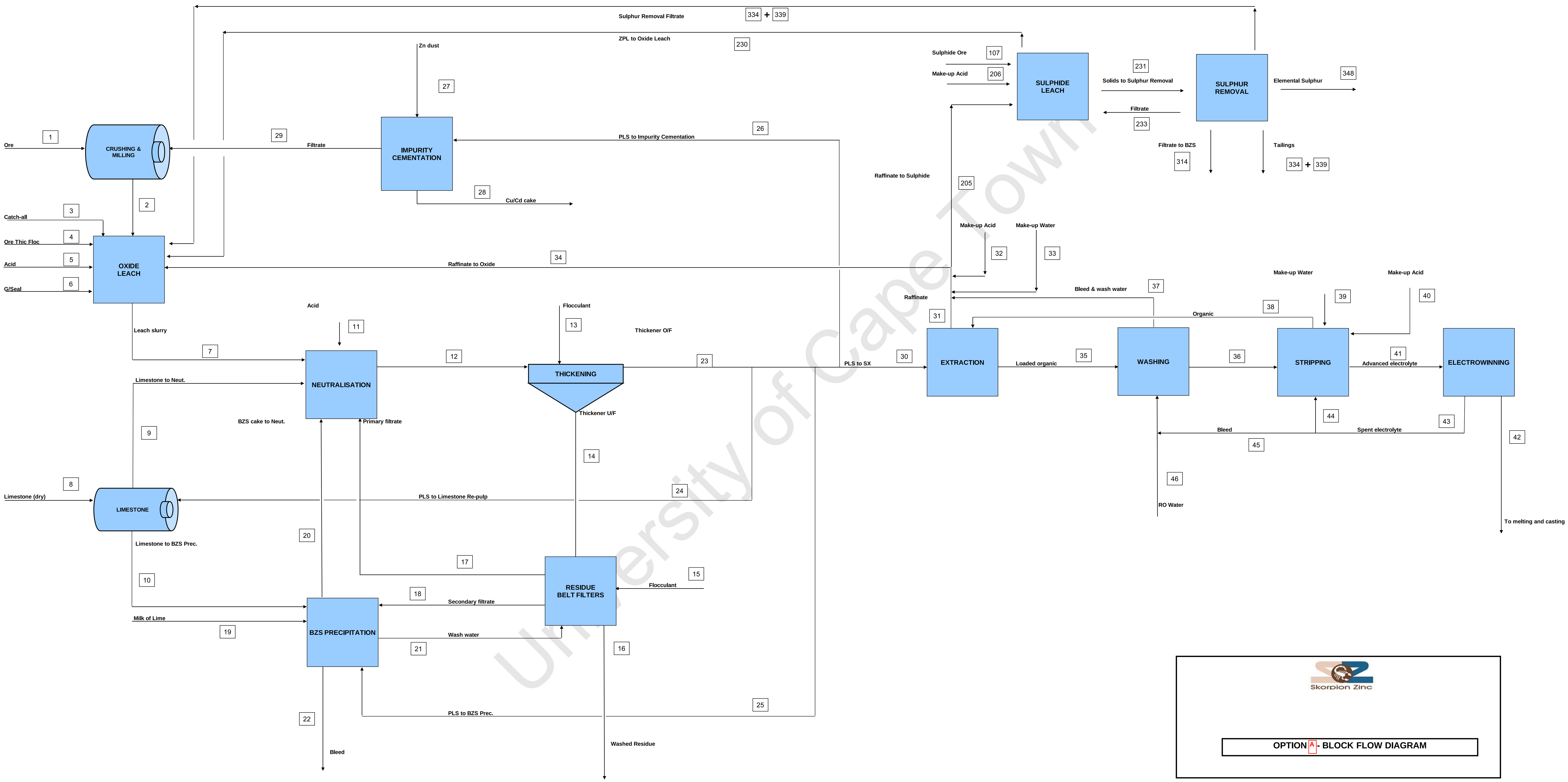
17. Crundwell, F.K., "Effect of iron impurity in zinc sulphide concentrates on the rate of dissolution", AIChE Journal, 34, 1128-1134, 1988.
18. De Nys, T., Terwinghe, F., "Integration of direct leaching of zinc based sulphidic concentrates and the Veille-Montagne goethite process", Lead-Zinc '90, The Minerals Metals & Materials Society, pp335-350,1990.
19. Demarthe, J. M., Georgeaux, A., "Hydrometallurgical Treatment of Complex Sulfides", pp. 113-120, in : Complex Metallurgy '78, Ed. : M. J. Jones, IMM. London, 1978.
20. Demopoulos, G.P., Baldwin, S.A., "Stoichemometric and kinetic effects on the pressure leaching of zinc concentrates", The Minerals, Metals & Materials Society, 1999 TMS annual meeting, 28Feb – 4 March 1999.
21. Doyle, B.N., Masters, I.M., Webster, I.C., Veltman, H., "Acid pressure leaching of zinc concentrates with elemental sulphur as by-product", Proceedings of the Eleventh Commonwealth Mining and Metallurgical Congress, Hong Kong, p645-653, 1978.
22. Doyle, F. M.; Arauco, H. and Feng, L. M., "Iron Removal During Oxidative, Acid Pressure Leaching of a Zinc Sulfide Concentrate", Intl. J. Min. Process., vol. 25,pp. 241-260, 1989.
23. Dreisinger, D. B.; Peters, E.; Talaba, M.; DeGraaf, K. B.; Owusu, G. and Swiniarski, R., "The Kinetics of the Sherritt Gordon Zinc Pressure Leach Process", pp. 313-333, in : Lead-Zinc '90, Eds. :T. S. Mackey and R. D. Prengaman, TMS, Warrendale, PA, 1990.
24. Ferron, C.J., "Atmospheric leaching of zinc sulphide concentrates using regenerated ferric sulphate solutions", Lead-Zinc 2000, TMS Fall Extraction and Process Metallurgy Meeting, Pittsburgh, USA, eds Dutrizac, J.E., Gonzalez, J.A., Henke, D.M., James, S.E. and Siegmund, H.A.J, 711-726, 2000.
25. Figueiredo, J.M., Novais, A.Q., Limpo Gill, J., Amer, S., " THe CENIM-LNETI Process for zinc recovery from complex sulphides", Zinc & Lead '95, Mining and Materials Processing, Japan, pp. 151-160, 1995.
26. Figueiredo, J.M., Novais, A.Q., Limpo Gill,J., Amer, S., "The CENIM-LNETI Process for Zinc Recovery from Complex Sulfides", Proceedings of Zinc '93, AUSIMM, 333-339, 1993.
27. Filippou, D., "Innovative hydrometallurgical processes for the primary processing of zinc", Mineral Processing and Extractive Metallurgy Review, 25, 205-252, 2004.
28. Fugleberg, S. and Jarvinen, A., Method for Leaching Zinc Concentrate in Atmospheric Conditions. PCT World Patent WO98/06879, February 19, 1998.
29. Haakana, T., Saxen, B., Lehtinen, L., Takala, H., "Outotec direct leaching application in China", The Southern African Institute of Mining and Metallurgy, Lead and Zinc 2008, pp. 69-84, 2008.
30. Habashi, F., "A short history of hydrometallurgy", Hydrometallurgy, 79, 15-22, 2005.
31. Harvey, T. J., Yen, W. T. and Paterson, J. G., "A Kinetic Investigation into the Pressure Oxidation of Sphalerite from a Complex Concentrate", Minerals Engineering, Vol. 6, No. 8-10, pp. 949-967, 1993.
32. Harvey, T.J., Van der Merwe, W., Afewu, K., " The application of the GeoBiotics GEOCOAT Biooxidation technology for the treatment of Sphalerite at Kumba Resources' Rosh Pinah Mine", Bio & Hydrometallurgy '02, Cape Town, 13-15 March 2002.
33. Hourn, M.M., Rohner, P., Bartsch, P., Ngoviky, K., "Benefits of using the Albion process for a North Queensland project and a case study of capital and operating cost benefits versus bacterial oxidation and pressure oxidation", Albion Process, www.albionprocess.com/publications.html, 29 August 2005.
34. Hourn, M.M., Turner, D.w., Holzberger, I.R., "Atmospheric mineral leaching process", PCT World Patent WO 96/29439, September 26, 1996.
35. Hourn, M.M., Turner, D.w., Holzberger, I.R., "Atmospheric mineral leaching process", US Patent 5,993,635 November 30, 1999.
36. http://www.albionprocess.com.

37. http://www.crugroup.com
38. http://www.intec.com.au
39. http://www.williamhunter.co.uk/ZINC/isp.htm
40. Huggare, T.L., Ojanen, A., Kuivala, A., "How zinc concentrates are processed at the Outokumpu zinc plant in Kokkola", International symposium on hydrometallurgy, Page 770 – 805, 1973.
41. Hughes, S., Reuter, M.A., Baxter, R., Kaye, A., "Ausmelt technology for lead and zinc processing", The Southern African Institute of Mining and Metallurgy, Lead and Zinc , p147-162, 2008.
42. Hultbom, K.G., "Industrially proven methods for mercury removal from gases", EPD Congress 2003, The Minerals, Metals & Materials Society, 2003.
43. Hunter, C.J., Williams, T.L., Purkiss, S.A.R., Cheung, W. , Connors, E., Gilder, R.D., "Bacterial oxidation of sulphide ores and concentrates", United States, US2004/0206208 A1, 2004.
44. Ismael, M.R.C., Carvalho, J.M.R., "Iron recovery from sulphate leach liquors in zinc hydrometallurgy", Minerals Engineering, vol. 16, pp.31-39, 2003.
45. Kanno, M., Watanabe, Y., Saruta, K., Narumi, A., "Method and Apparatus for Leaching Zinc Concentrates", European Patent Application EP 1245686 A2, January 29, 2002.
46. Krysa, B.D., " Zinc pressure leach at HBM", Hydrometallurgy, vol. 39, pp. 71-77, 1995.
47. Lahtinen, M., Svens, K. and Lehtinen, L., "Hematite versus jarosite precipitation in zinc production", Iron Control Technologies, Eds J. E. Dutrizac and P. A. Riveros, Montreal, Canada, 2006.
48. Lahtinen, M., Takala, H., "Atmospheric zinc concentrate leaching technology of Outokumpu", Lead & Zinc '05, Proceedings of the International Symposium on Lead and Zinc Processing, Kyoto, Japan, pp. 803-816, October 17-19, 2005.
49. Limpo, J.L., Luis, A., Gomez, C., "Reactions During the Oxygen Leaching of Metallic Sulphides in the CENIM – LNETI Process." Hydrometallurgy, 28, 163-178, 1992.
50. Lizama, H.M., Harlamovs, J.R., Belange, S., Brienne, S.H.R., "The Teck Cominco HydroZinc Process", Proceedings of Hydrometallurgy 2003, Eds. C.A. Young et al, TMS (Warrendale), 1503-1516, 2003.
51. Loan, M., Newman, O.M.G., Cooper, R.M.G, Farrow, J.B., Parkinson, G.M., "Defining the Paragoethite process for iron removal in zinc hydrometallurgy", Hydrometallurgy, vol. 81, pp. 104-129, 2006.
52. Magoon, D.H., Babcock, A.R., "Equipment and processes for roasting Zinc concentrates", Lead-Zinc '90, The Mineals, Metals & Materials Society, 1990, p 389 – 412.
53. Martin, M.T., Jankola, W.A., "Cominco's Trail zinc pressure leach operation", CIM Extractive Metallurgy, CIM Bulletin, vol.78, no.876, April 1985.
54. Nice, R.W., "Ramp up times Why are they never properly considered", Proceedings: Alta 2003, 2003.
55. Nilsson, L., Petersson, S. and Sandstrom, A, "A New Process for Zinc Recovery from Bacterial Leach Solutions", Scand. J. Metallurgy, 25, 161-171, 1996.
56. Olper, M., "The EZINEX process – Five years of development from bench scale to a commercial plant", Zinc and Lead Processing, Eds. J. E. Dutrizac, J. A. Gonzalez, G. L. Bolton and P. Hancock, The Metallurgical Society of CIM, 545-560, 1998.
57. Owusu, G.; Dreisinger, D. B. and Peters, E., "Interfacial Effects of Surface-Active Agents Under Zinc Pressure Leach Conditions", Metall. Trans. Vol. 26B , pp. 5-12, February 1995.
58. Pecina, T., Franco, T., Castillo, P., Orrantia, E., "Leaching of a zinc concentrate in $H_2SO_4$ solutions containing $H_2O_2$ and complexing agents.
59. Pelino, M., Cantalini, C., Abbruzzese, C., Plescia,P., "Treatment and recycling of goethite waste arising from the hydrometallurgy of zinc", Hydrometallurgy, vol. 40, pp25-35, 1996.

60. Ricketts, N. J.; Johnson, R. D. and Smith, G. D. J., “Hydrometallurgical Treatment of Cu-Pb-Zn Sulfide Concentrates”, pp. 705-720, in : Extractive Metallurgy ’89, IMM, London, 1989.
61. Rossi, G., “The design of bioreactors”, Hydrometalurgy, vol. 59, pp. 217-231, 2001.
62. Sandstrom, A., and Petersson, S., “Bioleaching of a Complex Sulphide Ore with Moderate Thermophilic and Extreme Thermophilic Microorganisms”, Hydrometallurgy, 46, 181-190, 1997a.
63. Sandstrom, A., Sundkvist, J.-E., and Petersson, S., “Biooxidation of a Complex Zinc Sulphide Ore: A Study Performed in Continuous Bench- and Pilot Scale”, Proc. Int. Biohydrometallurgy Symposium, IBS97-BIOMINE97, Australia Mineral Foundation, pp. M.1.1.1-M.1.1.11, 1997b.
64. Sannikov, Y.I., Liamina, M.A., Shumskij, V.A., Grinin, Y.A., Radashin, M.V., “A physical and chemical description of the Kivcet lead flash smelting process”, CIM Bulletin, Vol 91, No 1022, July/August 1998.
65. Sinclair, R.J., “The extractive metallurgy of zinc”, The Australasian Institute of Mining and Metallurgy Spectrum Series, vol. 13, 2005.
66. Souza, A.D., Pina, P.B., Leão, V.A., Silva, C.A., Siqueira, P.F., “The leaching kinetics of a zinc sulphide concentrate in acid ferric sulphate”. Hydrometallurgy, vol. 89, pp. 72-81, 2007.
67. Steemson, M.L, Sheehan, G.J., Winborne, D.A. and Wong, F.S., “An Integrated Bioleach/Solvent Extraction Process for Zinc Metal Production from Zinc Concentrates”, PCD World Patent WO 94/28184, May 24, 1994.
68. Steemson, M.L., Wong, F.S., Goebel, B., “The Integration of Zinc Bioleaching with Solvent Extraction for the Production of Zinc Metal from Zinc Concentrates”, Proceedings International Biohydrometallurgy Symposium. IBS 97 – BIOMINE97 –“Biotechnology Comes of Age”, Australian Mineral Foundation, pp. M1.4.1-M1.4.10, 1997.
69. Svens,K., Kerstiens, B. and Runkel, K., “Recent experiences with modern zinc processing technology”, Erzmetall, 56, 94-103, 2003.
70. Takala, H., “Leaching Zinc Concentrates at Outokumpu Kokkola Plant”, Erzmetall, 52, 37-42.
71. Van Put, J.W., Terwinghe, F.M.I.G. and De Nys, T.S.A., “Process for the Extraction of Zinc from Sulfide Concentrates, U.S. Patent 5, 858, 315,January 12, 1999.
72. Van Weert, G. and van Sandwijk, T., “Routes to residue-free zinc refining”, Journal of the Minerals Metals & Materials Society, 51, 12, 26-28, 1999.
73. VNIITSVETMET, The Eastern Mining and Metallurgical Research Institute for Non Ferrous Metals, 2011, http://vcm.ukg.kz/en/working4.htm#1
74. Zhang, H, Wang,X, Wang Xuming., Jian, S., Zhao, Q., “ A novel combined flowsheet of beneficiation and acid leaching under high pressure for complex lead-zinc ores”, Advanced materials research, vol.92, pp. 13-21, 2010.

# APPENDICES

## APPENDIX 1: OPTION A MASS & ENERGY BALANCE



Sulphur Removal Filtrate
334 + 339
ZPL to Oxide Leach
230
Zn dust
27
Sulphide Ore
107
Make-up Acid
206
SULPHIDE LEACH
231
Solids to Sulphur Removal
Filtrate
233
SULPHUR REMOVAL
Elemental Sulphur
348
Filtrate to BZS
314
Tailings
334 + 339
Ore
1
CRUSHING & MILLING
29
Filtrate
IMPURITY CEMENTATION
PLS to Impurity Cementation
26
28
Cu/Cd cake
Raffinate to Sulphide
205
2
3
Catch-all
4
Ore Thic Floc
5
Acid
6
G/Seal
OXIDE LEACH
34
Raffinate to Oxide
Make-up Acid
32
Make-up Water
33
Raffinate
Bleed & wash water
37
Make-up Water
39
Make-up Acid
40
38
Organic
31
Acid
11
Flocculant
13
Thickener O/F
Leach slurry
7
12
THICKENING
23
PLS to SX
30
EXTRACTION
Loaded organic
35
WASHING
36
STRIPPING
41
Advanced electrolyte
ELECTROWINNING
Limestone to Neut.
NEUTRALISATION
BZS cake to Neut.
Primary filtrate
Thickener U/F
44
43
Bleed
Spent electrolyte
45
9
14
42
46
8
Limestone (dry)
LIMESTONE
PLS to Limestone Re-pulp
24
RO Water
To melting and casting
Limestone to BZS Prec.
20
10
17
15
Flocculant
RESIDUE BELT FILTERS
18
Secondary filtrate
Milk of Lime
19
BZS PRECIPITATION
Wash water
21
16
22
25
PLS to BZS Prec.
Bleed
Washed Residue
Skorpion Zinc
OPTION A - BLOCK FLOW DIAGRAM

| Stream No. | | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 | 11 | 12 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | Dry Oxide Ore | Oxide Ore Slurry | Catch All Water | Ore Thickener Flocc | Make up Acid to Oxide | GSW to Oxide | Oxide Leach Discharge | Dry Limestone | Limestone to Neutralisation | Limestone to BZS | Acid to Neutralisation | Neut. Discharge |
| Flow | t/h | 122.1 | 300.5 | 12.0 | 3.3 | 1.0 | 10.0 | 1383.5 | 31.5 | 73.2 | 35.3 | 3.0 | 1714.8 |
| Flow | m3/h | 44.2 | 210.5 | 12.0 | 3.3 | 0.5 | 10.0 | 1248.5 | 11.7 | 56.2 | 27.1 | 1.6 | 1517.4 |
| Temperature | C | 25.0 | 40.0 | 25.0 | 25.0 | 25.0 | 25.0 | 50.0 | 25.0 | 40.6 | 40.6 | 25.0 | 49.7 |
| Solids | wt % | 96.0 | 39.0 | 0.0 | 0.0 | 0.0 | 0.0 | 7.3 | 100.0 | 29.0 | 29.0 | 0.0 | 8.8 |
| Solids | t/h | 117.2 | 117.2 | 0.0 | 0.0 | 0.0 | 0.0 | 101.1 | 31.5 | 21.2 | 10.2 | 0.0 | 150.5 |
| Aq. Liquids | t/h | 4.9 | 183.3 | 12.0 | 3.3 | 1.0 | 10.0 | 1282.4 | 0.0 | 51.9 | 25.1 | 3.0 | 1564.3 |
| Gases | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Org. Liquids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Zn | g/L | 0.0 | 32.0 | 0.0 | 0.0 | 0.0 | 0.0 | 28.5 | 0.0 | 32.4 | 32.4 | 0.0 | 32.4 |
| Aq. Liq.: aH2SO4 | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 1850.1 | 0.0 | 5.0 | 0.0 | 0.0 | 0.0 | 1850.1 | 0.0 |
| Aq. Liquids: Fe | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.4 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | wt % | 11.6 | 11.6 | 0.0 | 0.0 | 0.0 | 0.0 | 1.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.7 |
| Solids: Fe | wt % | 2.2 | 2.2 | 0.0 | 0.0 | 0.0 | 0.0 | 2.4 | 0.0 | 0.0 | 0.0 | 0.0 | 1.9 |
| Solids: S | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1.2 | 0.0 | 0.0 | 0.0 | 0.0 | 4.5 |
| Solids: Zn | t/h | 13.6 | 13.6 | 0.0 | 0.0 | 0.0 | 0.0 | 1.1 | 0.0 | 0.0 | 0.0 | 0.0 | 1.1 |
| Solids: Fe | t/h | 2.6 | 2.6 | 0.0 | 0.0 | 0.0 | 0.0 | 2.4 | 0.0 | 0.0 | 0.0 | 0.0 | 2.9 |
| Solids: S | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1.2 | 0.0 | 0.0 | 0.0 | 0.0 | 6.7 |
| Aq. Liq: Zn | t/h | 0.0 | 5.5 | 0.0 | 0.0 | 0.0 | 0.0 | 34.5 | 0.0 | 1.6 | 0.8 | 0.0 | 47.3 |
| Aq. Liq: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 1.0 | 0.0 | 6.0 | 0.0 | 0.0 | 0.0 | 2.9 | 0.0 |
| Aq. Liq: SO4[2-] | t/h | 0.0 | 9.2 | 0.0 | 0.0 | 1.0 | 0.0 | 73.6 | 0.0 | 2.7 | 1.3 | 2.9 | 80.9 |
| Aq. Liquids: SO4[2-] | g/L | 0.0 | 54.0 | 0.0 | 0.0 | 1812.1 | 0.0 | 60.8 | 0.0 | 55.5 | 55.5 | 1812.1 | 55.4 |
| Aq. Liquids: Mn | g/L | 0.0 | 3.9 | 0.0 | 0.0 | 0.0 | 0.0 | 4.0 | 0.0 | 4.0 | 4.0 | 0.0 | 4.0 |
| Aq. Liq: Mn | t/h | 0.0 | 0.7 | 0.0 | 0.0 | 0.0 | 0.0 | 4.8 | 0.0 | 0.2 | 0.1 | 0.0 | 5.8 |
| Aq. Liquids: Cu | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.4 | 0.0 | 0.5 | 0.5 | 0.0 | 0.5 |
| Enthalpy | kJ/kg | -13273.5 | -14311.4 | -15877.1 | -15877.1 | -8486.6 | -15877.1 | -14954.4 | -12831.3 | -14386.5 | -14386.5 | -8486.6 | -14888.5 |

| Stream No. | | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 | 22 | 23 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | Flocc. to Neut. Thickener | Neut. Thickener U/F | Flocc. to Neut. BF | Neut. BF Cake | Neut. BF Prim. Filtrate | Neut. BF Sec. Filtrate | Milk of Lime to BZS | BZS Cake | BZS Wash Water | BZS Bleed | PLS |
| Flow | t/h | 4.3 | 442.6 | 3.0 | 247.3 | 194.8 | 120.6 | 3.4 | 45.3 | 117.2 | 49.2 | 1276.5 |
| Flow | m3/h | 4.3 | 329.2 | 3.0 | 156.2 | 181.4 | 116.1 | 2.9 | 31.3 | 120.4 | 50.6 | 1187.7 |
| Temperature | C | 25.0 | 41.9 | 25.0 | 55.5 | 41.8 | 55.5 | 25.0 | 80.0 | 70.0 | 70.0 | 41 .95 |
| Solids | wt % | 0.0 | 34.0 | 0.0 | 60.8 | 0.0 | 0.0 | 25.0 | 60.0 | 0.0 | 0.0 | 0.0 |
| Solids | t/h | 0.0 | 150.5 | 0.0 | 150.5 | 0.0 | 0.0 | 0.8 | 27.2 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids | t/h | 4.3 | 292.1 | 3.0 | 96.8 | 194.8 | 120.6 | 2.5 | 18.1 | 117.2 | 49.2 | 1276.5 |
| Gases | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Org. Liquids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Zn | g/L | 0.0 | 32.4 | 0.0 | 0.5 | 32.0 | 25.9 | 0.0 | 0.5 | 0.5 | 0.5 | 32.4 |
| Aq. Liq.: aH2SO4 | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Fe | g/L | 0.0 | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.1 | 0.1 | 0.1 | 0.0 |
| Solids: Zn | wt % | 0.0 | 0.7 | 0.0 | 0.7 | 0.0 | 0.0 | 0.0 | 19.9 | 0.0 | 0.0 | 0.0 |
| Solids: Fe | wt % | 0.0 | 1.9 | 0.0 | 1.9 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | wt % | 0.0 | 4.5 | 0.0 | 4.5 | 0.0 | 0.0 | 0.0 | 10.2 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | t/h | 0.0 | 1.1 | 0.0 | 1.1 | 0.0 | 0.0 | 0.0 | 5.4 | 0.0 | 0.0 | 0.0 |
| Solids: Fe | t/h | 0.0 | 2.9 | 0.0 | 2.9 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | t/h | 0.0 | 6.7 | 0.0 | 6.7 | 0.0 | 0.0 | 0.0 | 2.8 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: Zn | t/h | 0.0 | 8.8 | 0.0 | 0.1 | 5.8 | 3.0 | 0.0 | 0.0 | 0.1 | 0.0 | 38.5 |
| Aq. Liq: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: SO4[2-] | t/h | 0.0 | 15.1 | 0.0 | 0.8 | 9.9 | 5.3 | 0.0 | 0.2 | 1.0 | 0.4 | 65.9 |
| Aq. Liquids: SO4[2-] | g/L | 0.0 | 55.4 | 0.0 | 8.0 | 54.8 | 45.5 | 0.0 | 7.8 | 7.9 | 7.9 | 55.4 |
| Aq. Liquids: Mn | g/L | 0.0 | 4.0 | 0.0 | 4.0 | 3.9 | 3.9 | 0.0 | 3.9 | 3.9 | 3.9 | 4.0 |
| Aq. Liq: Mn | t/h | 0.0 | 1.1 | 0.0 | 0.4 | 0.7 | 0.5 | 0.0 | 0.1 | 0.5 | 0.2 | 4.7 |
| Aq. Liquids: Cu | g/L | 0.0 | 0.5 | 0.0 | 0.0 | 0.5 | 0.4 | 0.0 | 0.0 | 0.0 | 0.0 | 0.5 |
| Enthalpy | kJ/kg | -15877.1 | -14618.8 | -15877.1 | -14536.0 | -15030.4 | -15090.2 | -15234.3 | -12438.0 | -15582.0 | -15582.0 | -15021.8 |

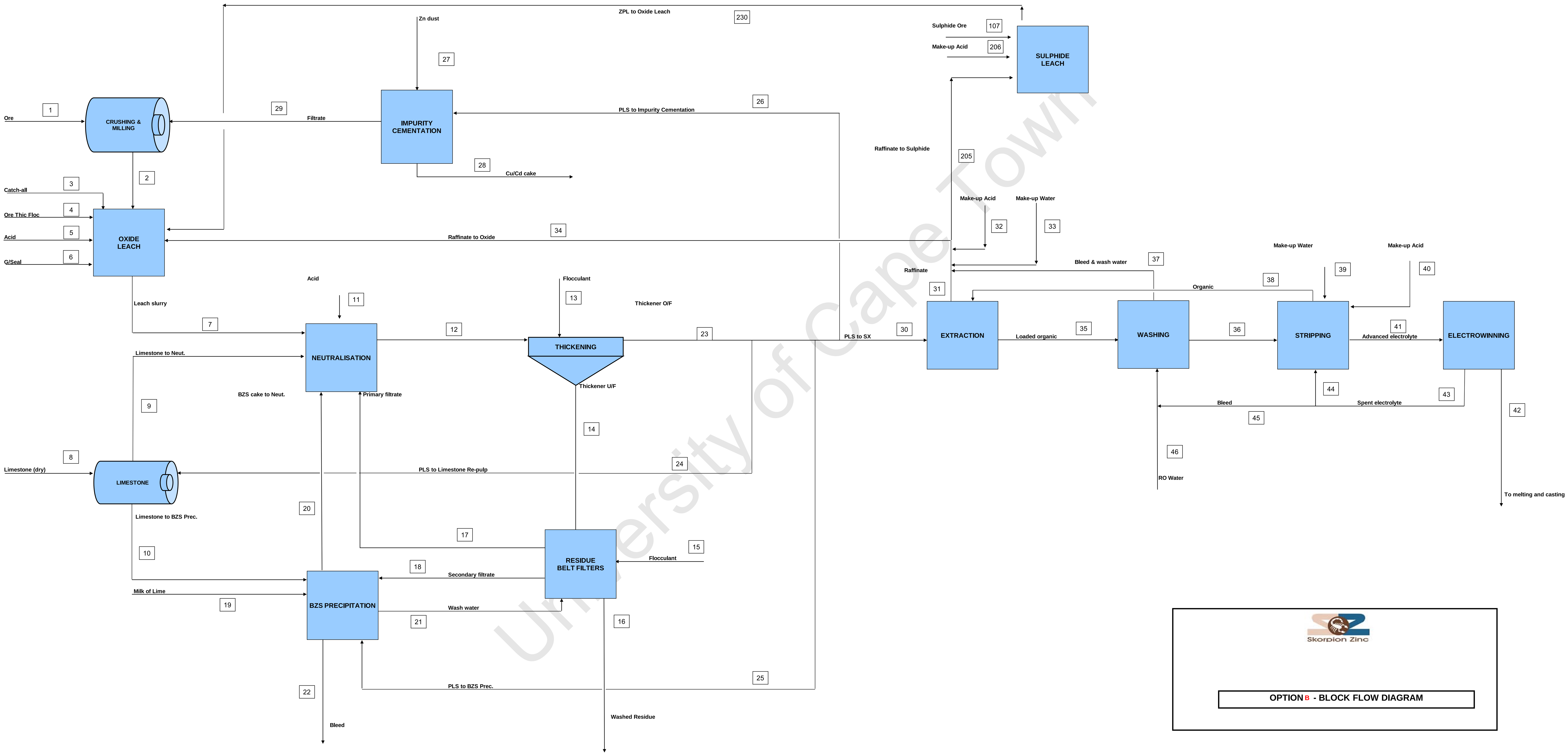
| Stream No. | | 24 | 25 | 26 | 27 | 28 | 29 | 30 | 31 | 32 | 33 | 34 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | PLS to Limestone | PLS to BZS | PLSto purification | Zn Dust | CulCd Cake | Purification Filtrate | PLS to SX | Raffinate | Raff Regen. Acid | Raff Make-up Water | Raffinate to Oxide |
| Flow | t/h | 77.1 | 49.9 | 178.5 | 0.1 | 0.2 | 178.4 | 971 .10 | 954.0 | 6.4 | 0.2 1 | 970.2 |
| Flow | m3/h | 71.7 | 46.4 | 166.1 | 0.0 | 0.1 | 166.3 | 903.6 | 912.3 | 3.4 | 0.2 | 924 .03 |
| Temperature | C | 41 .95 | 42.0 | 41 .95 | 25.0 | 42.4 | 42.4 | 42.0 | 43.7 | 25.0 | 25.0 | 40.0 |
| Solids | wt % | 0.0 | 0.0 | 0.0 | 100.0 | 60.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids | t/h | 0.0 | 0.0 | 0.0 | 0.1 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids | t/h | 77.1 | 49.9 | 178.5 | 0.0 | 0.1 | 178.4 | 971 .10 | 954.0 | 6.4 | 0.2 | 970.2 |
| Gases | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Org. Liquids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Zn | g/L | 32.4 | 32.4 | 32.4 | 0.0 | 33.0 | 33.0 | 32.4 | 12.5 | 0.0 | 0.0 | 12.2 |
| Aq. Liq.: aH2SO4 | g/L | 0.0 | 0.0 | 1.0 | 0.0 | 0.0 | 0.0 | 0.0 | 29.4 | 1850.1 | 0.0 | 36.4 |
| Aq. Liquids: Fe | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | wt % | 0.0 | 0.0 | 0.0 | 100.0 | 15.6 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Fe | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | t/h | 0.0 | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: Zn | t/h | 2.3 | 1.5 | 5.4 | 0.0 | 0.0 | 5.5 | 29.3 | 11.4 | 0.0 | 0.0 | 11 .25 |
| Aq. Liq: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 26.8 | 6.3 | 0.0 | 33.6 |
| Aq. Liq: SO4[2-] | t/h | 4.0 | 2.6 | 9.2 | 0.0 | 0.0 | 9.2 | 50.1 | 50.1 | 6.2 | 0.0 | 56.0 |
| Aq. Liquids: SO4[2-] | g/L | 55.4 | 55.4 | 55.4 | 0.0 | 55.5 | 55.5 | 55.4 | 54.9 | 1812.1 | 0.0 | 60.6 |
| Aq. Liquids: Mn | g/L | 4.0 | 4.0 | 4.0 | 0.0 | 4.0 | 4.0 | 4.0 | 4.0 | 0.0 | 0.0 | 3.6 |
| Aq. Liq: Mn | t/h | 0.3 | 0.2 | 0.7 | 0.0 | 0.0 | 0.7 | 3.6 | 3.6 | 0.0 | 0.0 | 3.3 |
| Aq. Liquids: Cu | g/L | 0.5 | 0.5 | 0.5 | 0.0 | 0.1 | 0.1 | 0.5 | 0.5 | 0.0 | 0.0 | 0.5 |
| Enthalpy | kJ/kg | -15021.8 | -15021.8 | -15021 .80 | 0.0 | -6004.7 | -15021 .60 | -15021.8 | -15236.3 | -8486.6 | -15877.1 | -15223.9 |

| Stream No. | | 35 | 36 | 37 | 38 | 39 | 40 | 41 | 42 | 43 | 44 | 45 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | Loaded Organic | LO to Strip | EW Bleed & Wash Water | Stripped Organic | SX Make-up Water | SX Make-up Acid | Advance Electrolyte | Plated Zn | Spent Electrolyte | Spent Electrolyte to Strip | Spent Electrolyte Bleed |
| Flow | t/h | 1161.4 | 1161.4 | 88.0 | 1144.1 | 21.7 | 4.4 | 301.6 | 17.1 | 280.1 | 284.2 | 21 .90 |
| Flow | m3/h | 1450.4 | 1450.4 | 8503.0 | 1430.0 | 21.7 | 2.3 | 230.2 | 2.4 | 232.1 | 237.5 | 18.3 |
| Temperature | C | 43.7 | 41.9 | 41.9 | 42.0 | 25.0 | 25.0 | 42.0 | 40.0 | 40.0 | 41.5 | 41.5 |
| Solids | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 100.0 | 0.0 | 0.0 | 0.0 |
| Solids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 17.1 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids | t/h | 0.0 | 0.0 | 88.0 | 0.0 | 21.7 | 4.4 | 301 .55 | 0.0 | 280.1 | 284.2 | 21 .90 |
| Gases | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Org. Liquids | t/h | 1161.4 | 1161.4 | 0.0 | 1144.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Zn | g/L | 0.0 | 0.0 | 8.8 | 0.0 | 0.0 | 0.0 | 119.8 | 0.0 | 45.1 | 40.9 | 40.9 |
| Aq. Liq.: aH2SO4 | g/L | 0.0 | . 0.00 | 37.7 | 0.0 | 0.0 | 1850.1 | 64.1 | 0.0 | 174.3 | 175.0 | 175.0 |
| Aq. Liquids: Fe | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 100.0 | 0.0 | 0.0 | 0.0 |
| Solids: Fe | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 17.1 3 | 0.0 | 0.0 | 0.0 |
| Solids: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: Zn | t/h | 0.0 | 0.0 | 0.8 | 0.0 | 0.0 | 0.0 | 27.6 | 0.0 | 10.5 | 9.7 | 0.8 |
| Aq. Liq: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | t/h | 0.0 | 0.0 | 3.2 | 0.0 | 0.0 | 4.3 | 14.8 | 0.0 | 40.5 | 41 .57 | 3.2 |
| Aq. Liq: SO4[2-] | t/h | 0.0 | 0.0 | 4.2 | 0.0 | 0.0 | 4.2 | 55.0 | 0.0 | 55.0 | 55.0 | 4.2 |
| Aq. Liquids: SO4[2-] | g/L | 0.0 | 0.0 | 49.8 | 0.0 | 0.0 | 1812.1 | 238.8 | 0.0 | 236.9 | 231.4 | 231.4 |
| Aq. Liquids: Mn | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: Mn | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Cu | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Enthalpy | kJ/kg | -2782.0 | -2785.9 | -15366.8 | -2820.2 | -15877.1 | -8486.6 | -13394.9 | 5.8 | -13980.6 | -14035.6 | -14035.6 |

| Stream No. | | 46 | 101 | 107 | 205 | 206 | 210 | 230 | 231 | 314 | 333 | 334 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | RO Water to scrub | Sulphide Conc Feed | Sulphide Conc. Slurry | Raffinate to PAL | Acid Make-up to PAL | Oxygen to PAL | PAL Discharge Thickener U/F | PAL Discharge Thickener OIF | BF wash Filtrate to BZS | Sulphur Flotation Tails | Sulphur Tails Thickener Overflow |
| Flow | t/h | 66.1 | 9.2 | 14.1 | 73.3 | 7.3 | 1.8 | 11.7 | 83.0 | 5.7 | 5.1 | 4.6 |
| Flow | m3/h | 66.3 | 2.4 | 7.1 | 69.8 | 3.9 | 1338.7 | 8.3 | 71.2 | 5.1 | 3.9 | 4.7 |
| Temperature | C | 25.0 | 25.0 | 37. 02 | 40.0 | 25.0 | 25.0 | 88.3 | 88.3 | 42.7 | 30.2 | 30.2 |
| Solids | wt % | 0.0 | 100.0 | 65.0 | 0.0 | 0.0 | 0.0 | 35.0 | 0.0 | 1.0 | 40.0 | 0.0 |
| Solids | t/h | 0.0 | 9.2 | 9.2 | 0.0 | 0.0 | 0.0 | 4.1 | 0.0 | 0.1 | 2.0 | 0.0 |
| Aq. Liquids | t/h | 66. 10 | 0.0 | 4.9 | 73.3 | 7.3 | 0.0 | 7.6 | 83.0 | 5.6 | 4.6 | 0.0 |
| Gases | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1.8 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Org. Liquids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Zn | g/L | 0.0 | 0.0 | 12.1 8 | 12.2 | 0.0 | 0.0 | 74.1 | 74.1 | 47.6 | 0.1 | 0.1 |
| Aq. Liq.: aH2SO4 | g/L | 0.0 | 0.0 | 36.4 | 36.4 | 1850.1 | 0.0 | 24.7 | 24.7 | 15.9 | 0.0 | 0.0 |
| Aq. Liquids: Fe | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 5.0 | 5.0 | 3.2 | 0.0 | 0.0 |
| Solids: Zn | wt % | 0.0 | 51.4 | 51.4 | 0.0 | 0.0 | 0.0 | 2.3 | 2.3 | 2.3 | 0.8 | 0.8 |
| Solids: Fe | wt % | 0.0 | 4.5 | 4.5 | 0.0 | 0.0 | 0.0 | 1.1 | 1.1 | 1.1 | 1.7 | 1.7 |
| Solids: S | wt % | 0.0 | 28.7 | 28.7 | 0.0 | 0.0 | 0.0 | 64.6 | 64.6 | 64.6 | 2306.0 | 2306.0 |
| Solids: Zn | t/h | 0.0 | 4.7 | 4.7 | 0.0 | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Fe | t/h | 0.0 | 0.4 | 0.4 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | t/h | 0.0 | 2.6 | 2.6 | 0.0 | 0.0 | 0.0 | 2.6 | 0.0 | 0.0 | 0.5 | 0.0 |
| Aq. Liq: Zn | t/h | 0.0 | 0.0 | 0.1 | 0.9 | 0.0 | 0.0 | 0.5 | 5.3 | 0.2 | 0.0 | 0.0 |
| Aq. Liq: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | | 0.0 | 0.4 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | t/h | 0.0 | 0.0 | 0.2 | 2.5 | 7.1 | 0.0 | 0.2 | 1.8 | 0.1 | 0.0 | 0.0 |
| Aq. Liq: SO4[2-] | t/h | 0.0 | 0.0 | 0.3 | 4.2 | 7.0 | 0.0 | 1.0 | 11 .05 | 0.5 | 0.0 | 0.0 |
| Aq. Liquids: SO4[2-] | g/L | 0.0 | 0.0 | 60.7 | 60.6 | 1812.1 | 0.0 | 155.2 | 155.2 | 99.6 | 0.1 | 0.1 |
| Aq. Liquids: Mn | g/L | 0.0 | 0.0 | 3.6 | 3.6 | 0.0 | 0.0 | 4.9 | 4.9 | 3.2 | 0.0 | 0.0 |
| Aq. Liq: Mn | t/h | 0.0 | 0.0 | 0.0 | 0.3 | 0.0 | 0.0 | 0.0 | 0.4 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Cu | g/L | 0.0 | 0.0 | 0.5 | 0.5 | 0.0 | 0.0 | 1.0 | 1.0 | 0.7 | 0.0 | 0.0 |
| Enthalpy | kJ/kg | -15877.1 | -3831 .38 | -7818.8 | -15223.9 | -8486.6 | 0.0 | -10698.7 | -13819.4 | -14472.2 | -13969.8 | -15853.5 |

| Stream No. | | 347 | 348 | 601 | 603 | 604 | 605 | 606 | 607 | 608 | 611 |
|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | Sulphur r Pressure Filter Cake | Elemental Sulphur | Total Process Water | Total Oxygen | Total Steam | Total Sulphuric Acid | Total Flotation Reagents | Milk of Lime | Total Flocculant | Total GSW |
| Flow | t/h | 0.2 | 2.0 | 45.0 | 1.8 | 24.2 | 22.1 | 0.0 | 5.6 | 11 .82 | 29.9 |
| Flow | m3/h | 0.1 | 1.0 | 45.1 | 1338.7 | 4702.4 | 11 .71 | 0.0 | 4.8 | 11.9 | 30.0 |
| Temperature | C | 140.0 | 140.0 | 25.0 | 25.0 | 179.9 | 25.0 | 25.0 | 25.0 | 25.0 | 25.0 |
| Solids | wt % | 98.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 25.0 | 0.0 | 0.0 |
| Solids | t/h | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1.4 | 0.0 | 0.0 |
| Aq. Liquids | t/h | 0.0 | 2.0 | 45.0 | 0.0 | 0.0 | 22.1 | 0.0 | 4.2 | 11 .82 | 29.9 |
| Gases | t/h | 0.0 | 0.0 | 0.0 | 1.8 | 24.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Org. Liquids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Zn | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1850.1 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Fe | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | wt % | 46.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Fe | wt % | 5.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | wt % | 38.4 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | t/h | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | t/h | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: Zn | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 21 .67 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: SO4[2-] | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 21.2 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: SO4[2-] | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1812.1 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Mn | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: Mn | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Cu | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Enthalpy | kJ/kg | -2565.7 | 157.9 | -15877.1 | 0.0 | -13200 .93 | -8486.6 | -15877.1 | -15234.3 | -15877.1 | -15877.1 |

## APPENDIX 2: OPTION B MASS & ENERGY BALANCE


ZPL to Oxide Leach
230
Zn dust
27
Sulphide Ore
107
Make-up Acid
206
SULPHIDE LEACH
Ore
1
CRUSHING & MILLING
29
Filtrate
IMPURITY CEMENTATION
PLS to Impurity Cementation
26
28
Cu/Cd cake
Raffinate to Sulphide
205
2
3
Catch-all
4
Ore Thic Floc
5
Acid
6
G/Seal
OXIDE LEACH
Raffinate to Oxide
34
Make-up Acid
Make-up Water
32
33
Raffinate
Bleed & wash water
37
Make-up Water
Make-up Acid
39
40
Organic
38
31
Leach slurry
7
Acid
11
Flocculant
13
Thickener O/F
NEUTRALISATION
12
THICKENING
23
PLS to SX
30
EXTRACTION
Loaded organic
35
WASHING
36
STRIPPING
41
Advanced electrolyte
ELECTROWINNING
Limestone to Neut.
BZS cake to Neut.
Primary filtrate
Thickener U/F
9
14
44
43
Bleed
Spent electrolyte
45
42
46
Limestone (dry)
8
LIMESTONE
PLS to Limestone Re-pulp
24
RO Water
To melting and casting
Limestone to BZS Prec.
20
17
15
Flocculant
10
RESIDUE BELT FILTERS
18
Secondary filtrate
Milk of Lime
19
BZS PRECIPITATION
Wash water
21
16
22
PLS to BZS Prec.
25
Bleed
Washed Residue
Skorpion Zinc
OPTION B - BLOCK FLOW DIAGRAM

| Stream No. | | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 | 11 | 12 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | Dry Oxide Ore | Oxide Ore Slurry | Catch All Water | Ore Thickener Flocc | Make up Acid to Oxide leach | GSWto Oxide | Oxide Leach slurry Discharge | Dry Limestone | Limestone to Neutralisati on | Limestone to BZS | Acid to Neutralisa tion | Neut. Discharg e slurry |
| Flow | t/h | 128.6 | 316.5 | 11 .96 | 3.5 | 1.0 | 10.0 | 1385.9 | 27.7 | 66.3 | 29.4 | 1.6 | 1703 05 |
| Flow | m3/h | 46.6 | 221.6 | 12.0 | 3.5 | 0.5 | 1003.0 | 1237.4 | 10.3 | 50.8 | 22.6 | 0.8 | 1502.7 |
| Temperature | C | 25.0 | 40.0 | 25.0 | 25.0 | 25.0 | 25.0 | 50.0 | 25.0 | 40.6 | 40.6 | 25.0 | 49.3 |
| Solids | wt % | 96.0 | 39. 00 | 0.0 | 0.0 | 0.0 | 0.0 | 7.8 | 100.0 | 29.0 | 29.0 | 0.0 | 8.9 |
| Solids | t/h | 123.5 | 123.5 | 0.0 | 0.0 | 0.0 | 0.0 | 108.3 | 27.7 | 19.2 | 8.5 | 0.0 | 151 .60 |
| Aq. Liquids | t/h | 5.1 | 193.1 | 11 .96 | 3.5 | 1.0 | 10.0 | 1277.6 | 0.0 | 47.1 | 20.9 | 1.6 | 1551 .45 |
| Gases | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Org. Liquids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Zn | g/L | 0.0 | 32.0 | 0.0 | 0.0 | 0.0 | 0.0 | 29.2 | 0.0 | 32.4 | 32.4 | 0.0 | 32.4 |
| Aq. Liq.: aH2SO4 | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 1850. 12 | 0.0 | 5.0 | 0.0 | 0.0 | 0.0 | 1850.1 | 0.0 |
| Aq. Liquids: Fe | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | wt % | 11.6 | 11.6 | 0.0 | 0.0 | 0.0 | 0.0 | 1.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.8 |
| Solids: Fe | wt % | 2.2 | 2.2 | 0.0 | 0.0 | 0.0 | 0.0 | 2.6 | 0.0 | 0.0 | 0.0 | 0.0 | 2.0 |
| Solids: S | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1.3 | 0.0 | 0.0 | 0.0 | 0.0 | 4.1 |
| Solids: Zn | t/h | 14.3 | 14.3 | 0.0 | 0.0 | 0.0 | 0.0 | 1.2 | 0.0 | 0.0 | 0.0 | 0.0 | 1.2 |
| Solids: Fe | t/h | 2.7 | 2.7 | 0.0 | 0.0 | 0.0 | 0.0 | 2.8 | 0.0 | 0.0 | 0.0 | 0.0 | 3.1 |
| Solids: S | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1.4 | 0.0 | 0.0 | 0.0 | 0.0 | 6.2 |
| Aq. Liq: Zn | t/h | 0.0 | 5.7 | 0.0 | 0.0 | 0.0 | 0.0 | 35.0 | 0.0 | 1.4 | 0.6 | 0.0 | 46.8 |
| Aq. Liq: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 1.0 | 0.0 | 6.0 | 0.0 | 0.0 | 0.0 | 1.5 | 0.0 |
| Aq. Liq: SO4[2-] | t/h | 0.0 | 9.9 | 0.0 | 0.0 | 1.0 | 0.0 | 75.2 | 0.0 | 2.5 | 1.1 | 1.5 | 81.9 |
| Aq. Liquids: SO4[2-] | g/L | 0.0 | 55.2 | 0.0 | 0.0 | 1812.1 | 0.0 | 62.8 | 0.0 | 56.7 | 56.7 | 1812.1 | 56.7 |
| Aq. Liquids: Mn | g/L | 0.0 | 4.6 | 0.0 | 0.0 | 0.0 | 0.0 | 4.8 | 0.0 | 4.7 | 4.7 | 0.0 | 4.7 |
| Aq. Liq: Mn | t/h | 0.0 | 0.8 | 0.0 | 0.0 | 0.0 | 0.0 | 5.7 | 0.0 | 0.2 | 0.1 | 0.0 | 6.8 |
| Aq. Liquids: Cu | g/L | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.4 | 0.0 | 0.5 | 0.5 | 0.0 | 0.5 |
| Enthalpy | kJ/kg | -13273.5 | -14303 .54 | -15877.1 | -15877.1 | -8486.6 | -15877.1 | -14929.7 | -12831.3 | -14377.1 | -14377.1 | -8486.6 | -14883.5 |

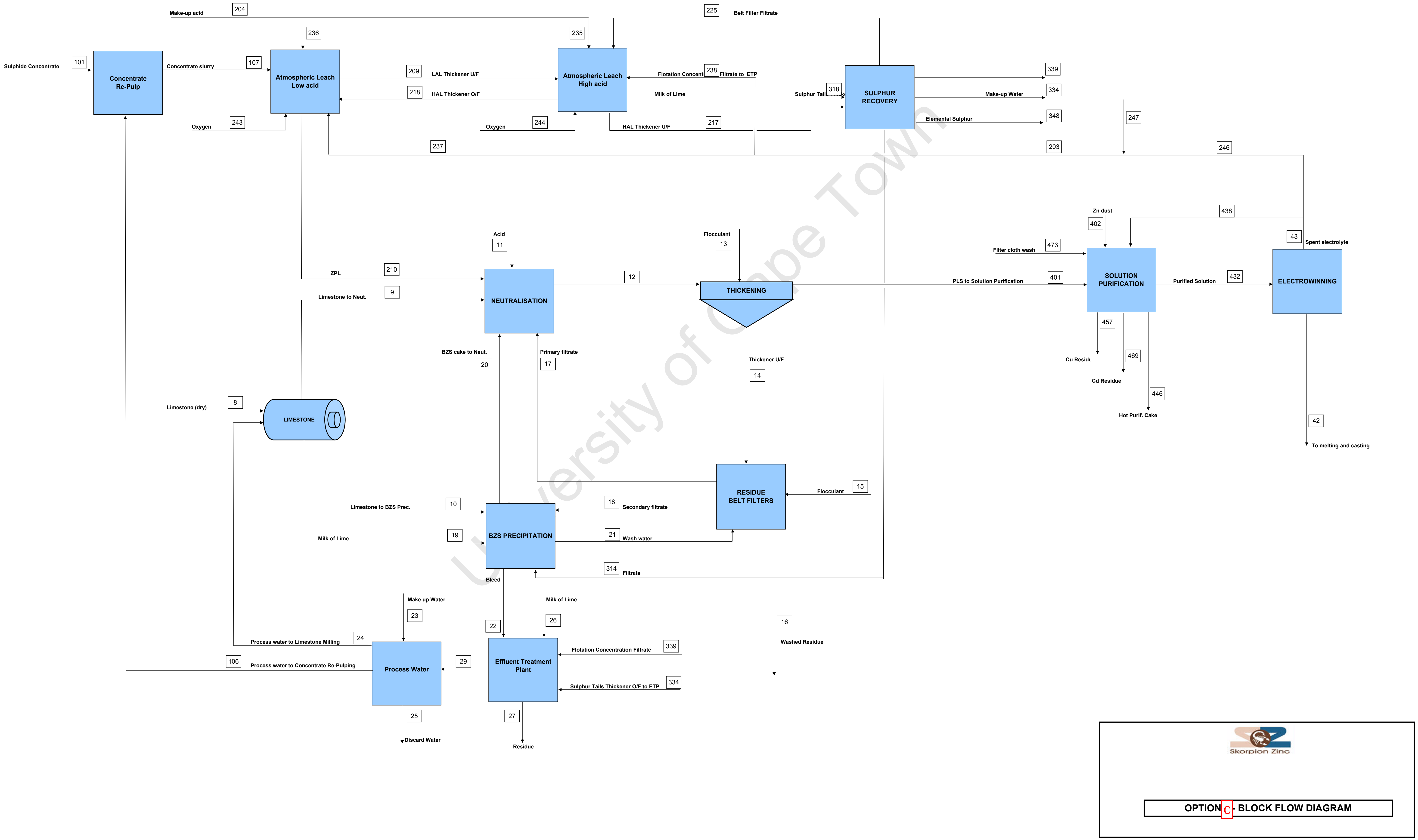
| Stream No. | | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 | 22 | 23 | 24 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| **Stream Name** | | Flocc. to Neut. Thickener | Neut. Thickener U/F | Flocc. to Neut. BF | Neut. BF Cake | Neut. BF Sec. Filtrate | Neut. BF Prim. Filtrate | Milk of Lime to BZS | BZS Cake | BZS Wash Water | BZS Bleed | PLS | PLS to Limeston e |
| **Flow** | **t/h** | 4.3 | 445.9 | 3.0 | 250.6 | 122.0 | 196.2 | 2.7 | 37.7 | 119.8 | 26.4 | 1261.5 | 67.9 |
| **Flow** | **m3/h** | 4.3 | 330.9 | 3.0 | 156.9 1 | 115.5 | 182.4 | 2.3 | 25.8 | 121 .28 | 26.7 | 1171.7 | 63.1 |
| **Temperature** | **C** | 25.0 | 42.0 | 25.0 | 55.7 | 55.7 | 41 .78 | 25.0 | 80.0 | 70.0 | 70.0 | 42.0 | 42.0 |
| **Solids** | **wt %** | 0.0 | 34.0 | 0.0 | 60.5 | 0.0 | 0.0 | 25.0 | 60.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids** | **t/h** | 0.0 | 151 .60 | 0.0 | 151.6 | 0.0 | 0.0 | 0.7 | 22.6 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids** | **t/h** | 4.3 | 294.3 | 3.0 | 99.0 | 122.0 | 196.2 | 2.0 | 15.1 | 119.8 | 26.4 | 1261.5 | 67.9 |
| **Gases** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Org. Liquids** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Zn** | **g/L** | 0.0 | 32.4 | 0.0 | 0.5 | 26.2 | 32.0 | 0.0 | 0.5 | 0.5 | 0.5 | 32.4 | 32.4 |
| **Aq. Liq.: aH2SO4** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Fe** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Zn** | **wt %** | 0.0 | 0.8 | 0.0 | 0.8 | 0.0 | 0.0 | 0.0 | 20.1 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Fe** | **wt %** | 0.0 | 2.0 | 0.0 | 2.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **wt %** | 0.0 | 4.1 | 0.0 | 4.1 | 0.0 | 0.0 | 0.0 | 10.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Zn** | **t/h** | 0.0 | 1.2 | 0.0 | 1.2 | 0.0 | 0.0 | 0.0 | 4.5 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Fe** | **t/h** | 0.0 | 3.1 | 0.0 | 3.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **t/h** | 0.0 | 6.2 | 0.0 | 6.2 | 0.0 | 0.0 | 0.0 | 2.3 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq: Zn** | **t/h** | 0.0 | 8.9 | 0.0 | 0.1 | 3.0 | 5.8 | 0.0 | 0.0 | 0.1 | 0.0 | 38.0 | 2.0 |
| **Aq. Liq: Fe** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq.: aH2SO4** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq: SO4[2-]** | **t/h** | 0.0 | 15.5 | 0.0 | 0.9 | 5.5 | 10.2 | 0.0 | 0.1 | 1.1 | 0.2 | 66.4 | 3.6 |
| **Aq. Liquids: SO4[2-]** | **g/L** | 0.0 | 56.7 | 0.0 | 9.2 | 47.3 | 56.1 | 0.0 | 9.1 | 9.1 | 9.1 | 56.7 | 56.7 |
| **Aq. Liquids: Mn** | **g/L** | 0.0 | 4.7 | 0.0 | 4.8 | 4.7 | 4.7 | 0.0 | 4.7 | 4.8 | 4.8 | 4.7 | 4.7 |
| **Aq. Liq: Mn** | **t/h** | 0.0 | 1.3 | 0.0 | 0.5 | 0.5 | 0.9 | 0.0 | 0.1 | 0.6 | 0.1 | 5.6 | 0.3 |
| **Aq. Liquids: Cu** | **g/L** | 0.0 | 0.5 | 0.0 | 0.0 | 0.4 | 0.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.5 | 0.5 |
| **Enthalpy** | **kJ/kg** | -15877.1 | -14636.3 | -15877.1 | -14582.9 | -15078.4 | -15017.3 | -15234.3 | -12432 .64 | -15567.2 | -15567.2 | -15008.5 | -15008.5 |

| Stream No. | | 25 | 26 | 27 | 28 | 29 | 30 | 31 | 32 | 33 | 34 | 35 | 36 | 37 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | PLS to BZS | PLS to Ore | Zn Dust | Cu/Cd Cake | Purification Filtrate | PLS to SX | Raffinate | Raff Regen. Acid | Raff Make-up Water | Raffinate to Oxide leach | Loaded Organic | LO to Strip | EW Bleed & Wash Water |
| Flow | t/h | 32.5 | 188. 00 | 0.1 | 0.2 | 187.9 | 973.0 | 955.7 | 6.4 | 0.0 | 971.5 | 1161.4 | 1161.4 | 88.0 |
| Flow | m3/h | 30.2 | 174.6 | 0.0 | 0.1 | 175.1 | 903.8 | 912.3 | 3.4 | 0.0 | 923.7 | 1450.4 | 1450.4 | 85.0 |
| Temperature | C | 41 .95 | 41 .95 | 25.0 | 42.3 | 42.3 | 42.0 | 43.7 | 25.0 | 25.0 | 40.0 | 43.7 | 41.9 | 41.9 |
| Solids | wt % | 0.0 | 0.0 | 100.0 | 60.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids | t/h | 0.0 | 0.0 | 0.1 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids | t/h | 32.5 | 188.0 | 0.0 | 0.1 | 187.9 | 973.0 | 955.7 | 6.4 | 0.0 | 971.5 | 0.0 | 0.0 | 88.0 |
| Gases | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Org. Liquids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1161.4 | 1161.4 | 0.0 |
| Aq. Liquids: Zn | g/L | 32.4 | 32.4 | 0.0 | 32.9 | 32.9 | 32.4 | 12.5 | 0.0 | 0.0 | 12.2 | 0.0 | 0.0 | 8.8 |
| Aq. Liq.: aH2SO4 | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 29.4 | 1850.1 | 0.0 | 36.4 | 0.0 | 0.0 | 37.7 |
| Aq. Liquids: Fe | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | wt % | 0.0 | 0.0 | 100.0 | 14.9 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Fe | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | t/h | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: Zn | t/h | 1.0 | 5.7 | 0.0 | 0.0 | 5.7 | 29.3 | 11.4 | 0.0 | 0.0 | 11 .24 | 0.0 | 0.0 | 0.8 |
| Aq. Liq: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 26.8 | 6.3 | 0.0 | 33.6 | 0.0 | 0.0 | 3.2 |
| Aq. Liq: SO4[2-] | t/h | 1.7 | 9.9 | 0.0 | 0.0 | 9.9 | 51.2 | 51 .21 | 6.2 | 0.0 | 57.0 | 0.0 | 0.0 | 4.2 |
| Aq. Liquids: SO4[2-] | g/L | 56.7 | 56.7 | 0.0 | 56.8 | 56.8 | 56.7 | 56.1 | 1812.1 | 0.0 | 61.7 | 0.0 | 0.0 | 49.8 |
| Aq. Liquids: Mn | g/L | 4.7 | 4.7 | 0.0 | 4.7 | 4.7 | 4.7 | 4.7 | 0.0 | 0.0 | 4.3 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: Mn | t/h | 0.1 | 0.8 | 0.0 | 0.0 | 0.8 | 4.3 | 4.3 | 0.0 | 0.0 | 4.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Cu | g/L | 0.5 | 0.5 | 0.0 | 0.1 | 0.1 | 0.5 | 0.5 | 0.0 | 0.0 | 0.4 | 0.0 | 0.0 | 0.0 |
| Enthalpy | kJ/kg | -15008 .52 | -15008.5 | 0.0 | -5999.4 | -15008.3 | -15008.5 | -15222.2 | -8486.6 | -15877.1 | -15211.1 | -2781 .98 | -2785.9 | -15366.9 |

| Stream No. | | 38 | 39 | 40 | 41 | 42 | 43 | 44 | 45 | 46 | 101 | 107 | 205 | 206 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | Stripped Organic | SX Make-up Water | SX Make-up Acid | Advance Electrolyte | Plated Zn | Spent Electrolyte | Spent Electrolyte to Strip | Spent Electrolyte Bleed | RO Water to EW | Sulphide Conc. Feed | Sulphide Conc. Slurry | Raffinate to PAL | Make-up Acid |
| Flow | t/h | 1144.1 | 21.7 | 4.4 | 299.1 | 17.1 | 277.6 | 281.8 | 21.9 | 66.1 | 7.8 | 12.0 | 74.5 | 0.0 |
| Flow | m3/h | 1430.0 | 21.7 | 2.3 | 228.2 | 2.4 | 230.1 | 235.5 | 18.3 | 66.3 | 2.1 | 6.1 | 70.8 | 0.0 |
| Temperature | C | 42.1 | 25.0 | 25.0 | 42.1 | 40.0 | 40.0 | 41.5 | 41.5 | 25.0 | 25.0 | 37.0 | 40.0 | 0.0 |
| Solids | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 100.0 | 0.0 | 0.0 | 0.0 | 0.0 | 100.0 | 65.0 | 0.0 | 0.0 |
| Solids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 17.1 | 0.0 | 0.0 | 0.0 | 0.0 | 7.8 | 7.8 | 0.0 | 0.0 |
| Aq. Liquids | t/h | 0.0 | 21.7 | 4.4 | 299.1 | 0.0 | 277.6 | 281.8 | 21.9 | 66.1 | 0.0 | 4.2 | 74.5 | 0.0 |
| Gases | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Org. Liquids | t/h | 1144.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Zn | g/L | 0.0 | 0.0 | 0.0 | 120.4 7 | 0.0 | 45.1 | 40.8 | 40.8 | 0.0 | 0.0 | 12.2 | 12.2 | 0.0 |
| Aq. Liq.: aH2SO4 | g/L | 0.0 | 0.0 | 1850.1 | 63.2 | 0.0 | 174.3 | 175.0 | 175.0 | 0.0 | 0.0 | 36.4 | 36.4 | 0.0 |
| Aq. Liquids: Fe | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 100.0 | 0.0 | 0.0 | 0.0 | 0.0 | 51.4 | 51.4 | 0.0 | 0.0 |
| Solids: Fe | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 4.5 | 4.5 | 0.0 | 0.0 |
| Solids: S | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 28.7 | 28.7 | 0.0 | 0.0 |
| Solids: Zn | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 17.1 | 0.0 | 0.0 | 0.0 | 0.0 | 4.0 | 4.0 | 0.0 | 0.0 |
| Solids: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.4 | 0.4 | 0.0 | 0.0 |
| Solids: S | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 2.2 | 2.2 | 0.0 | 0.0 |
| Aq. Liq: Zn | t/h | 0.0 | 0.0 | 0.0 | 27.5 | 0.0 | 10.4 | 9.6 | 0.8 | 0.0 | 0.0 | 0.1 | 0.9 | 0.0 |
| Aq. Liq: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | t/h | 0.0 | 0.0 | 4.3 | 14.4 | 0.0 | 40.1 | 41.2 | 3.2 | 0.0 | 0.0 | 0.2 | 2.6 | 0.0 |
| Aq. Liq: SO4[2-] | t/h | 0.0 | 0.0 | 4.2 | 54.5 | 0.0 | 54.5 | 54.5 | 4.2 | 0.0 | 0.0 | 0.3 | 4.4 | 0.0 |
| Aq. Liquids: SO4[2-] | g/L | 0.0 | 0.0 | 1812.1 | 238.8 | 0.0 | 236.9 | 231.4 | 231.4 | 0.0 | 0.0 | 61.8 | 61 .72 | 0.0 |
| Aq. Liquids: Mn | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 4.3 | 4.3 | 0.0 |
| Aq. Liq: Mn | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.3 | 0.0 |
| Aq. Liquids: Cu | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.4 | 0.4 | 0.0 |
| Enthalpy | kJ/kg | -2820.2 | -15877.1 | -8486.6 | -13390.2 | 5.8 | -13980 .61 | -14036.0 | -14036.0 | -15877.1 | -3831 .38 | -7814.3 | -15211.1 | 0.0 |

| Stream No. | | 21 0 | 229 | 601 | 603 | 604 | 605 | 607 | 608 | 610 |
|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | Oxygen to PAL | PAL Di scharge slurry | Total Process Water | Total Oxygen | Total Steam | Total Sulphuric Acid | Total Milk of Lime | Flocculant | Total GSW |
| Flow | t/h | 5.6 | 73.3 | 33.6 | 5.6 | 19.4 | 13.4 | 3.5 | 10.8 | 29.9 |
| Flow | m3/h | 4273.5 | 60.9 | 33.7 | 4273.5 | 3768.4 | 7.1 | 3.0 | 10.8 | 30.0 |
| Temperature | C | 25.0 | 101 .75 | 25. 00 | 25.0 | 179.9 | 25.0 | 25.0 | 25.0 | 25.0 |
| Solids | wt % | 0.0 | 2.1 | 0.0 | 0.0 | 0.0 | 0.0 | 25.0 | 0.0 | 0.0 |
| Solids | t/h | 0.0 | 1.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.9 | 0.0 | 0.0 |
| Aq. Liquids | t/h | 0.0 | 71.7 | 33.6 | 0.0 | 0.0 | 13.4 | 2.6 | 10.8 | 29.9 |
| Gases | t/h | 5.6 | 0.0 | 0.0 | 5.6 | 19.4 | 0.0 | 0.0 | 0.0 | 0.0 |
| Org. Liquids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Zn | g/L | 0.0 | 80.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | g/L | 0.0 | 44.7 | 0.0 | 0.0 | 0.0 | 1850.1 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Fe | g/L | 0.0 | 1.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | wt % | 0.0 | 5.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Fe | wt % | 0.0 | 18.8 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | wt % | 0.0 | 5.7 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | t/h | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Fe | t/h | 0.0 | 0.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | t/h | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: Zn | t/h | 0.0 | 4.9 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: Fe | t/h | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | t/h | 0.0 | 2.7 | 0.0 | 0.0 | 0.0 | 13.2 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: SO4[2-] | t/h | 0.0 | 11 .07 | 0.0 | 0.0 | 0.0 | 12.9 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: SO4[2-] | g/L | 0.0 | 183.3 | 0.0 | 0.0 | 0.0 | 1812.1 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Mn | g/L | 0.0 | 6.7 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: Mn | t/h | 0.0 | 0.4 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Cu | g/L | 0.0 | 1.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Enthalpy | kJ/kg | 0.0 | -13560 .71 | -15877. 10 | 0.0 | -13200 .93 | -8486.6 | -15234.3 | -15877.1 | -15877.1 |

## APPENDIX 3: OPTION C MASS & ENERGY BALANCE


Sulphide Concentrate
101
Concentrate Re-Pulp
Concentrate slurry
107
Make-up acid
204
236
Atmospheric Leach Low acid
Oxygen
243
209
LAL Thickener U/F
218
HAL Thickener O/F
235
Atmospheric Leach High acid
Oxygen
244
225
Belt Filter Filtrate
Flotation Concentrate
238
Filtrate to ETP
Milk of Lime
HAL Thickener U/F
217
Sulphur Tails
318
SULPHUR RECOVERY
339
334
Make-up Water
348
Elemental Sulphur
237
203
247
246
ZPL
210
Limestone to Neut.
9
Acid
11
NEUTRALISATION
12
Flocculant
13
THICKENING
PLS to Solution Purification
401
Zn dust
402
438
Filter cloth wash
473
SOLUTION PURIFICATION
Purified Solution
432
43
Spent electrolyte
ELECTROWINNING
457
Cu Residue
469
Cd Residue
446
Hot Purif. Cake
42
To melting and casting
BZS cake to Neut.
20
Primary filtrate
17
Thickener U/F
14
Limestone (dry)
8
LIMESTONE
RESIDUE BELT FILTERS
Flocculant
15
Limestone to BZS Prec.
10
18
Secondary filtrate
Milk of Lime
19
BZS PRECIPITATION
21
Wash water
314
Filtrate
Bleed
16
Washed Residue
Make up Water
23
Milk of Lime
26
22
Process water to Limestone Milling
24
106
Process water to Concentrate Re-Pulping
Process Water
29
Effluent Treatment Plant
Flotation Concentration Filtrate
339
Sulphur Tails Thickener O/F to ETP
334
25
Discard Water
27
Residue
Skorpion Zinc
OPTION C - BLOCK FLOW DIAGRAM

| Stream No. | | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | Limestone (dry) | Limestone to Neutralisation | Limestone to BZS | Acid to Neutralisation | 2nd Stage Fe Rem. Discharge | Flocc. 10 Neut. Thickener | Neul. Thickener U/F | Flocc. to NeutBF | . Neul. BF Cake | Neul. BF PrimFiltrate . | Neut. BF Sec Fi ltrale | Milk of Lime to BZS Precipitati | BZS Cake |
| Flow | t/h | 12.5 | 17.7 | 25.5 | 3.4 | 295.1 | 0.6 | 65.0 | 0.4 | 33.7 | 28.6 | 20.4 | 101.0 | 31 .37 |
| Flow | m3/h | 4.7 | 14.6 | 21.1 | 1.8 | 218.6 | 0.6 | 41.9 | 0.4 | 20.8 | 22.1 | 17.2 | 0.9 | 21 .55 |
| Temperature | C | 25.0 | 48.6 | 48.6 | 25.0 | 78.1 | 25.0 | 77.9 | 25.0 | 78.7 | 77.2 | 78.7 | 25.0 | 80.0 |
| Pressure | kPa | 101 .33 | 101 .33 | 101.3 | 101 .33 | 101.3 | 101 .33 | 101.3 | 101 .33 | 101.3 | 101 .33 | 101 .33 | .33 101.33 | 101 .33 |
| Density | kg/m3 | 2681.6 | 1210.2 | 1210.2 | 1,911,40 | 1350.1 | 997.0 | 1552.1 | 997.0 | 1617.9 | 1.297.59 | 1.185.93 | 1.157.62 | 1456.2 |
| Solids | wt % | 100.0 | 29.0 | 29.0 | 0.0 | 7,49 | 0.0 | 34.0 | 0.0 | 65.7 | 0.0 | 0.0 | 25.0 | 60.0 |
| Solids | t/h | 12.5 | 5.1 | 7,40 | 0.0 | 22.1 | 0.0 | 22.1 | 0.0 | 22. 10 | 0.0 | 0.0 | 0.3 | 18.8 |
| Aq. Liquids | t/h | 0.0 | 12.6 | 18.1 | 3.4 | 273.0 | 0.6 | 42.9 | 0.4 | 11.6 | 28.6 | 20.4 | 0.8 | 12.6 |
| Gases | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Org. Liquids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Zn | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 141.0 | 0.0 | 140.6 | 0.0 | O.SO | 138.7 | 91 .99 | 0.0 | O.SO |
| Aq. Liq.: aH2SO4 | g/L | 0.0 | 0.0 | 0.0 | 1,91 1,40 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Fe | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0. 13 | 0.0 | 0.1 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 18.0 |
| Solids: Fe | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 6.2 | 0.0 | 6.2 | 0.0 | 6.2 | 0.0 | 0.0 | 0.0 | 1.0 |
| Solids: Cu | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 |
| Solids: Cd | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 14.2 | 0.0 | 14.2 | 0.0 | 14.2 | 0.0 | 0.0 | 0.0 | 11.7 |
| Solids: Zn | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 3.4 |
| Solids: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 1.4 | 0.0 | 1.4 | 0.0 | 1.4 | 0.0 | 0.0 | 0.0 | 0.2 |
| Solids: Cu | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Cd | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 3.1 | 0.0 | 3.1 | 0.0 | 3.1 | 0.0 | 0.0 | 0.0 | 2.2 |
| Aq. Liquids: Zn | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 29.6 | 0.0 | 4.6 | 0.0 | 0.0 | 3.1 | 1.6 | 0.0 | 0.0 |
| Aq. Liq: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | t/h | 0.0 | 0.0 | 0.0 | 3.4 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: SO4[2-] | t/h | 0.0 | 0.0 | 0.0 | 3.4 | 45.0 | 0.0 | 7.1 | 0.0 | 0.1 | 4.7 | 2,42 | 0.0 | 0.1 |
| Aq. Liquids: SO4[2-] | g/L | 0.0 | 0.0 | 0.0 | 1,872. 11 | 214,41 | 0.0 | 2 13.79 | 0.0 | 4.1 | 211.0 | 141.1 | 0.0 | 4.1 |
| Aq. Liquids: Mn | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 3.3 | 0.0 | 3.3 | 0.0 | 1.9 | 3.3 | 2.8 | 0.0 | 1.9 |
| Aq. Liquids: Cu | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.9 | 0.0 | 0.9 | 0.0 | 0.0 | 0.9 | 0.6 | 0.0 | 0.0 |
| Aq. Liquids: Cd | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0.2 | 0.0 | 0.1 | 0.2 | 0.2 | 0.0 | 0.1 |
| Aq. Liquids: Al | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.1 |
| Enthalpy | kJ/kg | 12 ,831.34 | -14918.0 | -14918.0 | -8295.6 | 1 3,045.50 | -15877.1 | -12624.9 | -15877.1 | 12 ,945.93 | -13198.7 | -13656.2 | 1 5,234.32 | 1 2,471.27 |

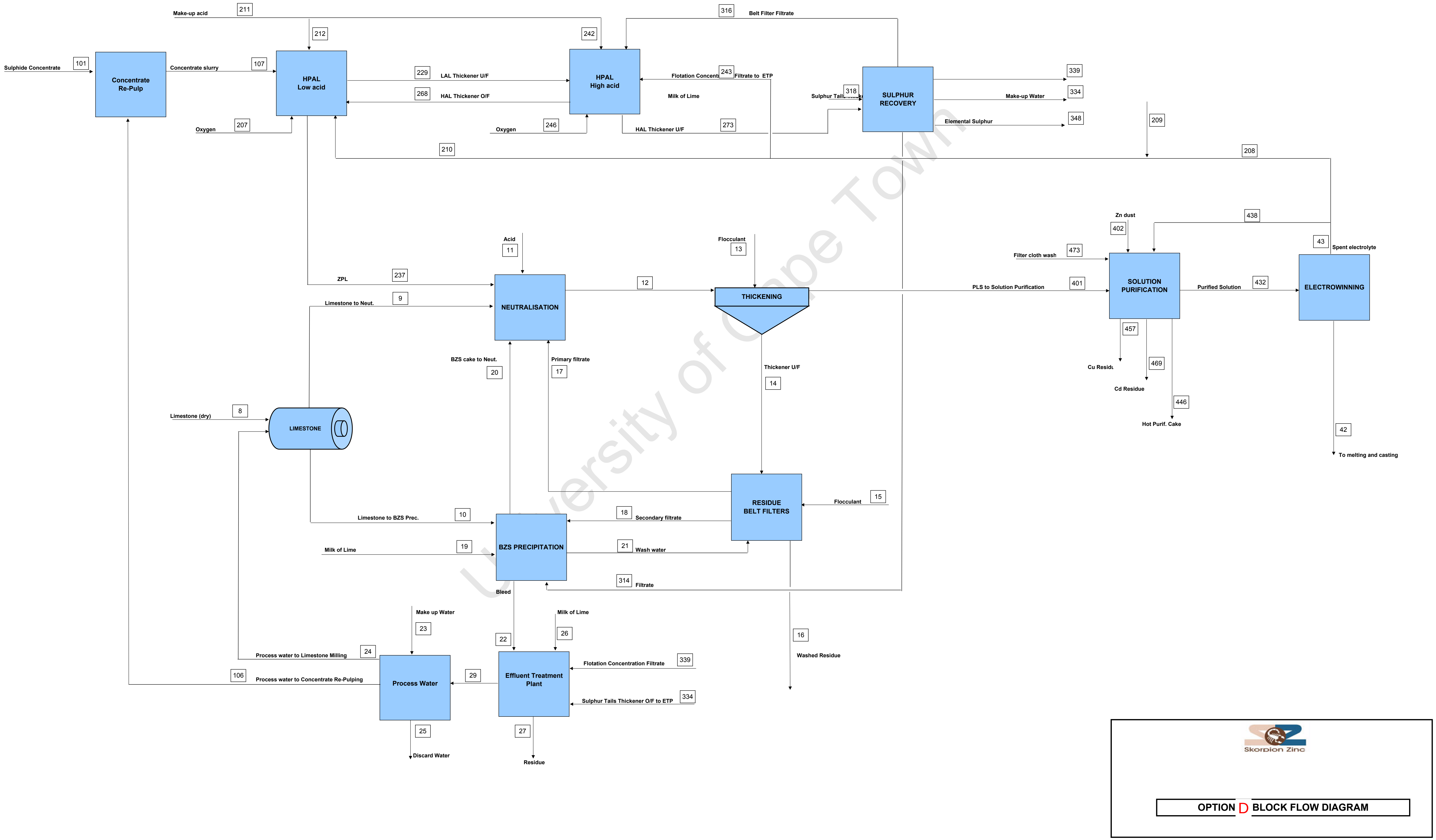
| Stream No. | | 21 | 22 | 23 | 24 | 25 | 26 | 27 | 29 | 42 | 43 | 101 | 106 | 107 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | BZS Wash Walerlo Residue BF | BZS Bleed | Process Dislr. Make up Water | Process Wateto Limestone Milt r | Discard Wat er | MOL to ETP | ETP Residue | Water from ETP | Plaled Zn | Spent Electrolyte fromEW | Zin c Sulphide Concentra te | Process Water 10 Concenlral e Repulping | Zinc Sulphide Concentra te Slurry |
| Flow | t/h | 17.2 | 21 .00 | 10.1 | 30.7 | 0.0 | 0.3 | 0.7 | 39.0 | 17.1 | 182.4 | 3403.0 | 18.3 | 52.4 |
| Flow | m3/h | 17.7 | 21.6 | 10.1 | 31 .09 | 0.0 | 0.3 | 0.6 | 39.6 | 2,40 | 148.6 | 9.0 | 18.6 | 27.5 |
| Temperature | C | 80.0 | 80.0 | 25.0 | 50.6 | 50.6 | 25.0 | 57.2 | 57.2 | 38.0 | 38.0 | 25.0 | 50.6 | 45.7 |
| Pressure | kPa | 101.3 | 101.3 | 101 .33 | 101 .33 | 101 .33 | 101 .33 | 101 .33 | 101.3 | 101 .33 | | 101 .33 | 101 .33 | 101.3 |
| Density | kg/m3 | 970.6 | 970.6 | 997.0 | 987.8 | 987.8 | 1157.6 | 1305.3 | 964 .80 | 7140.0 | 1227.1 | 3777.0 | 987.8 | 1902.3 |
| Solids | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 25.0 | 40.0 | 0.0 | 100.0 | 0.0 | 100.0 | 0.0 | 65.0 |
| Solids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 | 0.3 | 0.0 | 17.1 | 0.0 | 3403.0 | 0.0 | 34.0 |
| Aq. Liquids | t/h | 17.2 | 2100.0 | 10.1 | 30.7 | 0.0 | 0.3 | 0.4 | 39.0 | 0.0 | 182.4 | 0.0 | 18.3 | 18.3 |
| Gases | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Org. Liquids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Zn | g/L | O.SO | O.SO | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 51 .76 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 176,49 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Fe | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 5.3 | 0.0 | 100.0 | 0.0 | 51.4 | 0.0 | 51.4 |
| Solids: Fe | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0. 16 | 0.0 | 0.0 | 0.0 | 4.5 | 0.0 | 4.5 |
| Solids: Cu | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.5 | 0.0 | 0.5 |
| Solids: Cd | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.8 | 0.0 | 0.0 | 0.0 | 0. 11 | 0.0 | 0. 11 |
| Solids: S | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 12.4 | 0.0 | 0.0 | 0.0 | 28.7 | 0.0 | 28.7 |
| Solids: Zn | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 17.1 | 0.0 | 17.5 | 0.0 | 17.5 |
| Solids: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1.5 | 0.0 | 1.5 |
| Solids: Cu | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0. 17 |
| Solids: Cd | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 9.8 | 0.0 | 9.8 |
| Aq. Liquids: Zn | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 7.7 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 26.2 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: SO4[2-] | t/h | 0.1 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 37,49 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: SO4[2-] | g/L | 4. 10 | 4.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 252.3 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Mn | g/L | 1.9 | 1.9 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1.9 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Cu | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Cd | g/L | 0.1 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Al | g/L | 0.1 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Enthalpy | kJ/kg | -15587.4 | 1 5,587.35 | 15,877. 10 | -15770.3 | -15770.3 | 1 5,234.32 | 13.606. 39 | 1 5,742.63 | 5.1 | 1 3,944.88 | -3831.4 | -15770.3 | -8010.0 |

| Stream No. | | 203 | 204 | 209 | 210 | 217 | 218 | 235 | 236 | 237 | 238 | 239 | 243 | 244 | 245 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | Spent Eleclrolyte 10 Atm Leach | Acid Make-up 10 Aim Leach | LAL Slurry Thickener Underflow | LAL Slurry Thickener Overflow | HA L Slurry Thickener Underflow | HAL Slurry Thickener Overflow | Acid Make-up 10 HAL | Acid Make-up 10 LAL | LAL Spenl Eleclrolyte | HAL Spenl Eleclrolyte | Condilioned eleclrolyte 10 HA L | Oxygen 10 LAL | Oxygen 10 HAL | Flocculant to Atm Leach thickeners |
| **Flow** | **t/h** | 181.7 | 4.6 | 63.7 | 196.5 | 51 .14 | 130.3 | 1.6 | 3.0 | 83.6 | 98.1 | 99.7 | 4.3 | 3.3 | 7.3 |
| **Flow** | **m3/h** | 148.3 | 2.4 | 38.0 | 144.5 | 33.6 | 96.7 | 0.8 | 1.6 | 68.2 | 80.1 | 80.8 | 3284.5 | 2502.1 | 7.4 |
| **Temperature** | **C** | 37.9 | 25.0 | 78.8 | 78.8 | 70.1 | 70.1 | 25.0 | 25.0 | 37.9 | 37.9 | 41.3 | 25.0 | 25.0 | 25.0 |
| **Pressure** | **kPa** | 101 .33 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101 .33 | 101.3 | 101 .33 | 101.3 | 101.3 | 101 .33 | 101.3 | 101 .33 |
| **Density** | **kg/m3** | 1225.3 | 1911.4 | 1675.0 | 1359.7 | 1521.5 | 1347.3 | 1911.4 | 1911.4 | 1225.3 | 1.225.33 | 1233.9 | 1.3 | 1.3 | 997.0 |
| **Solids** | **wt %** | 0.0 | 0.0 | 35.0 | 0.1 | 28.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids** | **t/h** | 0.0 | 0.0 | 22.3 | 0.1 | 14.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids** | **t/h** | 181 .68 | 4.6 | 41.4 | 196.4 | 36.8 | 130.3 | 1.6 | 3.0 | 83.6 | 98.1 | 99.7 | 0.0 | 0.0 | 7.3 |
| **Gases** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 4.3 | 3.3 | 0.0 |
| **Org. Liquids** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Zn** | **g/L** | 51.4 | 0.0 | 180.0 | 160.0 | 133.8 | 133.8 | 0.0 | 0.0 | 51 .36 | 51.4 | 50.9 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq.: aH2SO4** | **g/L** | 175.1 | 1911.4 | 10.2 | 10.2 | 48.4 | 48.4 | 1,91 1.40 | 1911.4 | 175.1 | 175.1 | 193.4 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Fe** | **g/L** | 0.0 | 0.0 | 8.2 | 8.2 | 13.2 | 13.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Zn** | **wt %** | 0.0 | 0.0 | 26.6 | 26.6 | 1.0 | 1.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Fe** | **wt %** | 0.0 | 0.0 | 6.1 1 | 6.1 1 | 1.1 | 1.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cu** | **wt %** | 0.0 | 0.0 | 0.3 | 0.3 | 0.1 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cd** | **wt %** | 0.0 | 0.0 | 0.1 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **wt %** | 0.0 | 0.0 | 43.4 | 43.4 | 64.6 | 64.6 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Zn** | **t/h** | 0.0 | 0.0 | 5.9 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Fe** | **t/h** | 0.0 | 0.0 | 1.4 | 0.0 | 0. 16 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cu** | **t/h** | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cd** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **t/h** | 0.0 | 0.0 | 9.7 | 0.0 | 9.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Zn** | **t/h** | 7.6 | 0.0 | 4.9 | 23. 11 | 3.9 | 12.9 | 0.0 | 0.0 | 3.5 | 4. 11 | 4.1 1 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq: Fe** | **t/h** | 0.0 | 0.0 | 0.3 | 118.0 | 0.4 | 1.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq.: aH2SO4** | **t/h** | 26.0 | 4.6 | 0.3 | 1.5 | 1.3 | 4.7 | 1.6 | 3.0 | 11.9 | 14.0 | 15.6 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq: SO4[2-]** | **t/h** | 37. 12 | 4.5 | 7.9 | 37.5 | 6.9 | 24.4 5 | 1.6 | 2.9 | 17.1 | 2004.0 | 21.6 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: SO4[2-]** | **g/L** | 250.3 | 1872.1 | 259.3 | 259.3 | 252.8 | 252.8 | 1.872. 11 | 1,872. 11 | 250.3 | 250.3 | 267.5 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Mn** | **g/L** | 1.9 | 0.0 | 4. 14 | 4.1 | 3.7 | 3.7 | 0.0 | 0.0 | 1.9 | 1.9 | 1.9 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Cu** | **g/L** | 0.0 | 0.0 | 1.0 | 1.0 | 0.9 | 0.9 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Cd** | **g/L** | 0.0 | 0.0 | 0.2 | 0.2 | 0.2 | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Al** | **g/L** | 0.0 | 0.0 | 0.2 | 0.2 | 0.3 | 0.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Enthalpy** | **kJ/kg** | -13857.6 | -8295.6 | -9807.4 | 1 2,709.07 | 1 0,667.20 | 1 2,838.77 | -8295.6 | -8295.6 | 1 3,857.64 | 1 3,857.64 | 1 3,767.84 | 0.0 | 0.0 | 1 5,877.10 |

| Stream No. | | 246 | 247 | 314 | 318 | 333 | 334 | 339 | 348 | 401 | 402 | 432 | 438 | 446 | 457 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | Spenl Electrolyte from EWto Leach | SE Make-up water | Sulphur Removal Filtrate to BZS | Milk of Lime to Sulphur Recover | Sulphur Flotation Tail s | Sulphur Tails Thickener Overflow | Flotalion Concentration Fillrate | Elemental Sulphur | Fe Removal Thickener O/F | Zinc dust to Solution Purificatio n | Purffied Solution to EW | Spenl Electrolyte to Solulion Purfficatio n | Hot Purificatio n Cake | Cu Cake |
| Flow | t/h | 180.5 | 1.1 | 25.5 | 1.0 | 13.9 | 15.5 | 2.8 | 7.2 | 230.7 | 0.4 | 229.2 | 1.8 | 2.6 | 0.4 |
| Flow | m3/h | 147.1 | 1.1 | 20.6 | 0.7 | 10.7 | 15.9 | 2.9 | 3.5 | 177.3 | 0.1 | 172.0 | 1.5 | 1.5 | 0. 19 |
| Temperature | C | 38.0 | 25.0 | 44.0 | 25.0 | 31.2 | 31.1 9 | 31 .31 | 140.0 | 77.9 | 25.0 | 30.0 | 38.0 | 76.8 | 67.4 |
| Pressure | kPa | 101.3 | 101.3 | 101 .33 | 101 .33 | 101.3 | 101 .33 | 101.3 | 101 .33 | 101 .33 | 101.3 | 101 .33 | 101 .33 | 101 .33 | 101 .33 |
| Density | kg/m3 | 1227.1 | 997.0 | 1240.1 | 1157.6 | 1293.9 | 979.6 | 979.6 | 2070.0 | 1301.2 | 7140.0 | 1332.5 | 1227.1 | 1.690.26 | 2,241 .35 |
| Solids | wt % | 0.0 | 0.0 | 1.0 | 25.0 | 40.0 | 0.0 | 0.0 | 0.0 | 0.0 | 100.0 | 0.0 | 0.0 | 70.0 | 70.0 |
| Solids | t/h | 0.0 | 0.0 | 0.3 | 0.3 | 5.6 | 0.0 | 0.0 | 0.0 | 0.0 | 0.4 | 0.0 | 0.0 | 1.8 | 0.3 |
| Aq. Liquids | t/h | 180.5 | 1.1 | 25.2 | 0.8 | 8.3 | 15.5 | 2.8 | 7.2 | 230.7 | 0.0 | 229.2 | 1.8 | 0.8 | 0.1 |
| Gases | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Org. Liquids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Zn | g/L | 51.8 | 0.0 | 89.3 | 0.0 | 0.3 | 0.3 | 0.3 | 0.1 | 140.6 | 0.0 | 144.3 | 51 .76 | 131 .84 | 82.8 |
| Aq. Liq.: aH2SO4 | g/L | 176.5 | 0.0 | 32.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 | 176.5 | 0.0 | 0.0 |
| Aq. Liquids: Fe | g/L | 0.0 | 0.0 | 8.8 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | wt % | 0.0 | 0.0 | 1.0 | 0.0 | 0.4 | 0.4 | 0.0 | 0.0 | 0.0 | 10002.0 | 0.0 | 0.0 | 22.7 | 16.2 |
| Solids: Fe | wt % | 0.0 | 0.0 | 1.1 | 0.0 | 2.0 | 2.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Cu | wt % | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 52.1 |
| Solids: Cd | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1.1 |
| Solids: S | wt % | 0.0 | 0.0 | 64.6 | 0.0 | 28.6 | 28.6 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 11 .15 | 4.4 |
| Solids: Zn | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.4 | 0.0 | 0.0 | 0.4 | 0.1 |
| Solids: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0. 11 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Cu | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 |
| Solids: Cd | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | t/h | 0.0 | 0.0 | 0.2 | 0.0 | 1.6 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 |
| Aq. Liquids: Zn | t/h | 7.6 | 0.0 | 1.8 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 24.9 | 0.0 | 24.8 | 0.1 | 0.1 | 0.0 |
| Aq. Liq: Fe | t/h | 0.0 | 0.0 | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | t/h | 26.0 | 0.0 | 0.7 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.3 | 0.0 | 0.0 |
| Aq. Liq: SO4[2-] | t/h | 37.1 | 0.0 | 3.5 | 0.0 | 0. 00 | 0.0 | 0.0 | 0.0 | 37.9 | 0.0 | 37.5 | 0.4 | 0.1 | 0.0 |
| Aq. Liquids: SO4[2-] | g/L | 252.3 | 0.0 | 166.7 | 0.0 | 0.5 | 0.5 | 0.5 | 0.1 | 213.8 | 0.0 | 218.0 | 252.3 | 199.2 | 131 .12 |
| Aq. Liquids: Mn | g/L | 1.9 | 0.0 | 2.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 3.3 | 0.0 | 3.4 | 1.9 | 3.1 | 1.1 |
| Aq. Liquids: Cu | g/L | 0.0 | 0.0 | 0.6 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.9 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Cd | g/L | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 | 9.0 |
| Aq. Liquids: Al | g/L | 0.0 | 0.0 | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Enthalpy | kJ/kg | -13844.9 | 1 5,877.10 | 1 3,607.69 | -15234.3 | 1 3,705.21 | 1 5,843.77 | -15843.2 | 157.2 | -13171.7 | 0.0 | 1 3,319.67 | -13844.9 | 11 ,440.25 | -7.130.87 |

| Stream No. | | 469 | 473 | 601 | 603 | 604 | 605 | 606 | 607 | 608 | 611 |
|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | Cd Cake | Air 10 Purification Cooling Tow | Tolal Process Waler | TolalOxyge n | Tolal Sleam | T olal Sulphuri Acid | c Tolal Fioialion Reagenls | T olal Milk of Lime | Tolal Flocculanl | Tolal GSW |
| Flow | t/h | 0.1 | 530.8 | 62.4 | 7.6 | 9.8 | 8.0 | 0.1 | 2.3 | 9.7 | 19.9 |
| Flow | m3/h | 0.0 | 450102.6 | 62.6 | 5786.6 | 1909.3 | 4.2 | 0.1 | 2.0 | 9.7 | 20.0 |
| Temperature | C | 69.2 | 25.0 | 25.0 | 25.0 | 179.9 | 25.0 | 25.0 | 25.0 | 25.0 | 25.0 |
| Pressure | kPa | 101.3 | 101.3 | 101.3 | 101 .33 | 1000.0 | 101 .33 | 101 .33 | 101 .33 | 101 .33 | 101 .33 |
| Density | kg/m3 | 2270.6 | 1.2 | 997.0 | 1.3 | 5.2 | 1911.4 | 997.0 | 1157.6 | 997.0 | 997.0 |
| Solids | wt % | 70.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 25.0 | 0.0 | 0.0 |
| Solids | t/h | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.6 | 0.0 | 0.0 |
| Aq. Liquids | t/h | 0.0 | 0.0 | 62.4 | 0.0 | 0.0 | 8.0 | 0.1 | 1.8 | 9.7 | 19.9 |
| Gases | t/h | 0.0 | 530.8 | 0.0 | 7.6 | 9.8 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Org. Liquids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Zn | g/L | 88.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1911.4 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Fe | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | wt % | 8.7 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Fe | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Cu | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Cd | wt % | 61.6 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | wt % | 4.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Cu | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Cd | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Zn | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 8.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: SO4[2-] | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 7.9 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: SO4[2-] | g/L | 131.5 | 0.0 | 0.0 | 0.0 | 0.0 | 1872.1 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Mn | g/L | 1.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Cu | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Cd | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Al | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Enthalpy | kJ/kg | -7061.6 | 0.0 | -15877.1 | 0.0 | -13200.9 | -8295.6 | -15877.1 | -1 5234.32 | -1 5877.10 | -15877.1 |

## APPENDIX 4: OPTION D MASS & ENERGY BALANCE



Sulphide Concentrate
101
Concentrate Re-Pulp
Concentrate slurry
107
Make-up acid
211
212
HPAL Low acid
Oxygen
207
229
LAL Thickener U/F
268
HAL Thickener O/F
242
316
Belt Filter Filtrate
HPAL High acid
Oxygen
246
Flotation Concent
243
Filtrate to ETP
Milk of Lime
HAL Thickener U/F
273
Sulphur Tails
318
SULPHUR RECOVERY
339
Make-up Water
334
Elemental Sulphur
348
209
210
208
Acid
11
Flocculant
13
ZPL
237
Limestone to Neut.
9
NEUTRALISATION
12
THICKENING
Zn dust
402
438
43
Spent electrolyte
Filter cloth wash
473
PLS to Solution Purification
401
SOLUTION PURIFICATION
Purified Solution
432
ELECTROWINNING
457
Cu Residu
469
Cd Residue
446
Hot Purif. Cake
42
To melting and casting
BZS cake to Neut.
20
Primary filtrate
17
Thickener U/F
14
Limestone (dry)
8
LIMESTONE
RESIDUE BELT FILTERS
15
Flocculant
Limestone to BZS Prec.
10
18
Secondary filtrate
Milk of Lime
19
BZS PRECIPITATION
21
Wash water
314
Filtrate
Bleed
Make up Water
23
Milk of Lime
26
22
16
Washed Residue
Process water to Limestone Milling
24
106
Process water to Concentrate Re-Pulping
Process Water
29
Effluent Treatment Plant
Flotation Concentration Filtrate
339
Sulphur Tails Thickener O/F to ETP
334
25
Discard Water
27
Residue
Skorpion Zinc
OPTION D BLOCK FLOW DIAGRAM

| Stream No. | | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | Limestone (dry) | Limestone to Neutralisat ion | Limeston e to BZS | Acid to Neutralisat i on | 2nd Stage Fe Rem. Discharge | Flocc. to Neut. Thickener | Neut. Thi ck ener u/F | Flocc. to Neut. BF | Neu!. BF Cake | Neut. SF Prim. Filtrate | Neu!. BF Sec. Filtrate | Milk of Lime to BZS Precipitati on | BZS Cake |
| Flow | t/h | 6.4 | 5.1 | 16.9 | 2.0 | 255.2 | 0.3 | 32.9 | 0.2 | 17.0 | 14.5 | 10.3 | 0.7 | 21 .20 |
| Flow | m3/h | 2.4 | 4.2 | 14.0 | 1.1 | 190.7 | 0.3 | 21 .13 | 0.2 | 10.6 | 11 .06 | 8.7 | 0.6 | 14.6 |
| Temp erature | C | 25.0 | 52.5 | 52.5 | 25.0 | 77.6 | 25.0 | 77.5 | 25.0 | 78.5 | 76.6 | 78.5 | 25.0 | 80.0 |
| Press ure | kPa | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 |
| Density | kg/m3 | 2.681 .64 | 1208.3 | 1 208.33 | 1911.4 | 1.338.57 | 997.0 | 1.555.26 | 997.0 | 1.607.92 | 1 307.28 | 1190.9 | 1157.6 | 1.458.96 |
| Solids | wt % | 100.0 | 29.0 | 29.0 | 0.0 | 4.4 | 0.0 | 34.0 | 0.0 | 65.8 | 0.0 | 0.0 | 25.0 | 60.0 |
| Solids | t/h | 6.4 | 1.5 | 4.9 | 0.0 | 11 .17 | 0.0 | 11 .17 | 0.0 | 11 .17 | 0.0 | 0.0 | 0.2 | 12.1 |
| AQ. liquids | t/h | 0.0 | 3.6 | 12.0 | 2.0 | 244.1 | 0.3 | 21 .69 | 0.2 | 5.8 | 14.5 | 10.3 | 0.5 | 8.5 |
| Gases | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| OrQ. liquids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| AQ. liquids: Zn | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 144.9 | 0.0 | 144.7 | 0.0 | 0.5 | 142.8 | 94.0 | 0.0 | 0.5 |
| AQ. LiQ.: aH2S04 | g/L | 0.0 | 0.0 | 0.0 | 1911.4 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| AQ. liquids: Fe | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 | 0.0 | 0.1 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 18.0 |
| Solids: Fe | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 4.1 | 0.0 | 4.1 | 0.0 | 4.1 | 0.0 | 0.0 | 0.0 | 1.3 |
| Solids: Cu | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 |
| Solids: Cd | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 15.2 | 0.0 | 15.2 | 0.0 | 15.2 | 0.0 | 0.0 | 0.0 | 11 .75 |
| Solids : Zn | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 2.3 |
| Solids: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.5 | 0.0 | 0.5 | 0.0 | 0.5 | 0.0 | 0.0 | 0.0 | 0.2 |
| Solids : Cu | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Cd | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 1.7 | 0.0 | 1.7 | 0.0 | 1.7 | 0.0 | 0.0 | 0.0 | 1.5 |
| AQ. Liquids : Zn | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 27.0 | 0.0 | 2.4 | 0.0 | 0.0 | 1.6 | 0.8 | 0.0 | 0.0 |
| AQ. liquids: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| AQ. Liquids: aH2SO4 | t/h | 0.0 | 0.0 | 0.0 | 2.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| AQ. Liquids: SO4 2- | t/h | 0.0 | 0.0 | 0.0 | 2.0 | 41.1 | 0.0 | 3.7 | 0.0 | 0.0 | 2.4 | 1.3 | 0.0 | 0.0 |
| AQ. Liquids: SO4 2- | g/L | 0.0 | 0.0 | 0.0 | 1872.1 | 220.7 | 0.0 | 220.3 | 0.0 | 3.9 | 217.5 | 144.3 | 0.0 | 3.9 |
| AQ . Liquids: Mn | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 3.6 | 0.0 | 3.6 | 0.0 | 1.2 | 3.5 | 2.9 | 0.0 | 1.8 |
| AQ . Liquids: Cu | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.9 | 0.0 | 0.9 | 0.0 | 0.0 | 0.9 | 0.6 | 0.0 | 0.0 |
| AQ. liquids: Cd | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0.2 | 0.0 | 0.1 | 0.2 | 0.2 | 0.0 | 0.1 |
| AQ. Liquids: Al giL | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.1 |
| Enthalpy | kJ/kg | -12831.3 | -14 905.66 | -14.905.66 | -8295.6 | -13.052.60 | -15.877.10 | -12.648.04 | -1 5.877.10 | -13.055.69 | -13144.0 | -13823.8 | -15234.3 | -12.439.57 |

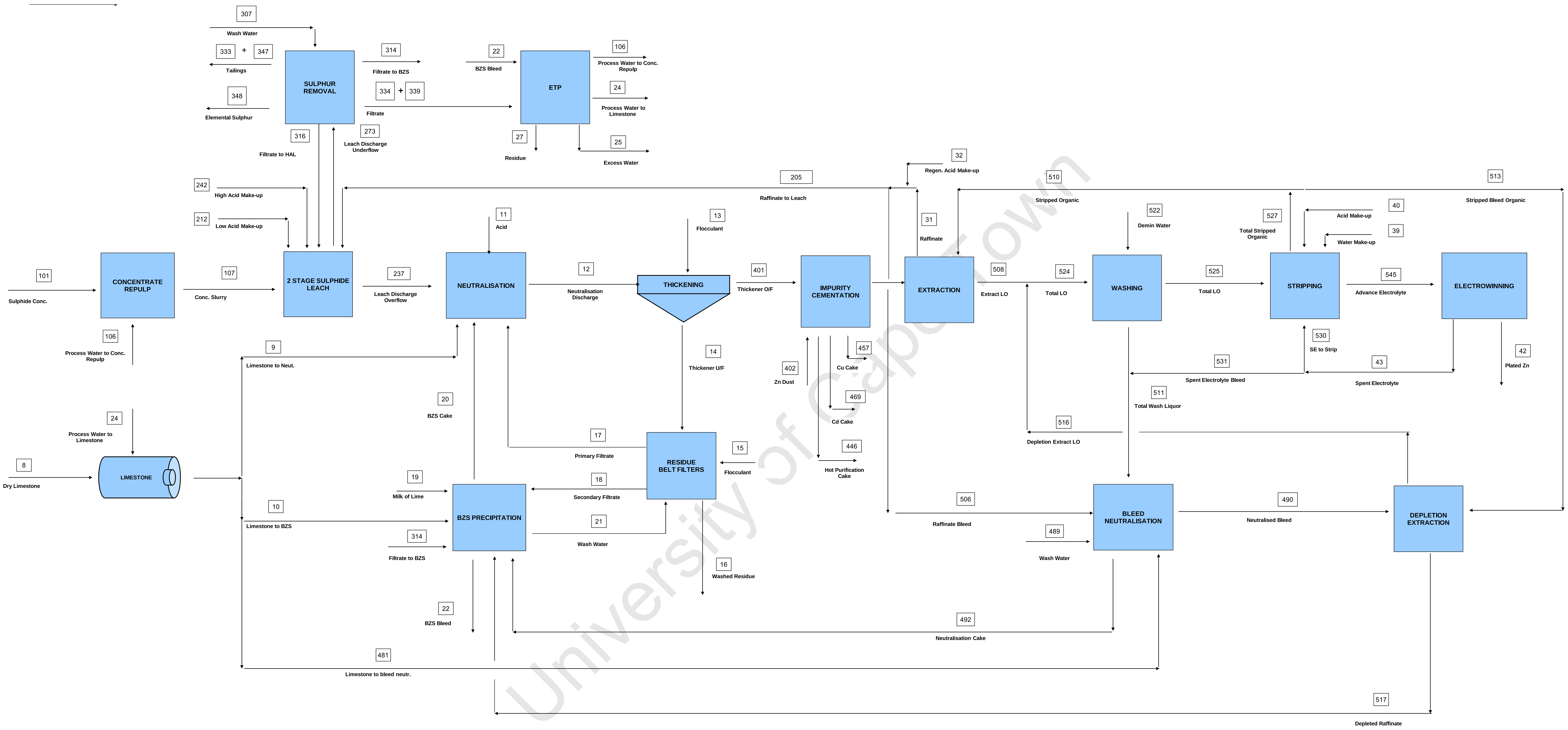
| Stream No. | | 21 | 22 | 23 | 24 | 25 | 26 | 27 | 28 | 29 | 42 | 43 | 101 | 106 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| **Stream Name** | | BZS Wash Water to Neutralisat ion BF | BZS Bleed | Make up Water | Proc ess Water to Limestone Mill | Discard Waler | MOL to ETP | ETP Residue | Air to ETP | Water from ETP | Plated Zn | Spent El ectroly1 e fromEW | Zinc Sulphide Concentrat e | Process Wale to Concentrat e Repulping |
| **Flow** | **t/h** | 8.7 | 19.3 | 0.0 | 15.6 | 5.0 | 0.3 | 0.6 | 0.0 | 38.9 | 17.1 | 174.8 | 34.0 | 18.3 |
| **Flow** | **m3/h** | 8.9 | 19.8 | 0.0 | 15.8 | 5.1 | 0.2 | 0.5 | 0.1 | 39.4 | 2.4 | 141.8 | 9.0 | 18.6 |
| **Temp erature** | **C** | 80.0 | 80.0 | 25.0 | 54.7 | 54.7 | 25.0 | 54.7 | 25.0 | 54.7 | 38.0 | 38.0 | 25.0 | 54.7 |
| **Press ure** | **kPa** | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 |
| **Density** | **kg/m3** | 970.5 | 970.5 | 997.0 | 985.8 | 985.8 | 1 157.62 | 1306.5 | 1.2 | 985.8 | 7.140.00 | 1.232.61 | 3.776.97 | 985.8 |
| **Solids** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 25.0 | 40.0 | 0.0 | 0.0 | 100.0 | 0.0 | 100.0 | 0.0 |
| **Solids** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 | 0.3 | 0.0 | 0.0 | 17.1 | 0.0 | 34.0 | 0.0 |
| **AQ. liquids** | **t/h** | 8.7 | 19.3 | 0.0 | 15.6 | 5.0 | 0.2 | 0.4 | 0.0 | 38.9 | 0.0 | 174.8 | 0.0 | 18.3 |
| **Gases** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **OrQ. liquids** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ. liquids: Zn** | **g/L** | 0.5 | 0.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 51 .76 | 0.0 | 0.0 |
| **AQ. LiQ.: aH2S04** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 185.2 | 0.0 | 0.0 |
| **AQ. liquids: Fe** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Zn** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 4.1 | 0.0 | 0.0 | 100.0 | 0.0 | 51.4 | 0.0 |
| **Solids: Fe** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 4.5 | 0.0 |
| **Solids: Cu** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.5 | 0.0 |
| **Solids: Cd** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.8 | .0.00 | 0.0 | 0.0 | 0.0 | 0.1 | 0.0 |
| **Solids: S** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 12.4 | 0.0 | 0.0 | 0.0 | 0.0 | 28.7 | 0.0 |
| **Solids : Zn** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 17.1 | 0.0 | 17.5 | 0.0 |
| **Solids: Fe** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1.5 | 0.0 |
| **Solids : Cu** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 |
| **Solids: Cd** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 9.8 | 0.0 |
| **AQ. Liquids : Zn** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 7.3 | 0.0 | 0.0 |
| **AQ. liquids: Fe** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ. Liquids: aH2SO4** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 26.3 | 0.0 | 0.0 |
| **AQ. Liquids: SO4 2-** | **t/h** | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 37.0 | 0.0 | 0.0 |
| **AQ. Liquids: SO4 2-** | **g/L** | 3.9 | 3.9 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 261.0 | 0.0 | 0.0 |
| **AQ . Liquids: Mn** | **g/L** | 1.8 | 1.8 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 2.0 | 0.0 | 0.0 |
| **AQ . Liquids: Cu** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ. liquids: Cd** | **g/L** | 0.1 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ. Liquids: Al giL** | **g/L** | 0.1 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Enthalpy** | **kJ/kg** | -15.589.27 | -1 5.589.27 | -15.877.10 | -15752.9 | -15752.9 | -15.234.32 | -13.629.65 | 0.0 | -15752.9 | 5.1 | -13803.3 | -3.831.38 | -15.752.92 |

| Stream No. | | 107 | 207 | 208 | 209 | 210 | 211 | 212 | 229 | 237 | 242 | 243 | 246 | 268 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| **Stream Name** | | Zinc Sulphide Concentrat e Slurry | Oxygen to LAL | Spent Electrolyte fromEW | Make-up wate to PAL | LAL Spent Electrolyte | Acid Make-up to PAL | Acid Make-up to LAL | LAL Atm . Let-down Vent | LAL Slurry Thickener Overflow | Acid Make-up to HAL | HAL Spent Electrolyte | Oxygen to HAL | HAL Slurry Thickener Overflow |
| **Flow** | **t/h** | 52.3 | 4.3 | 173.0 | 10.5 | 84.4 | 2.1 | 2.0 | 6.0 | 193.4 | 0.1 | 99.1 | 3.3 | 138.2 |
| **Flow** | **m3/h** | 27.5 | 3329.5 | 140.4 | 10.5 | 69.4 | 1.1 | 1.1 | 10 161 .87 | 144.4 | 0.0 | 81.5 | 179.2 | 103.8 |
| **Temp erature** | **C** | 49.1 | 25.0 | 38.0 | 25.0 | 37.2 | 25.0 | 25.0 | 102.9 | 78.4 | 25.0 | 37. 18 | 25.0 | 69.7 |
| **Press ure** | **kPa** | 101.3 | 101 .33 | 101 .33 | 101 .33 | 101 .33 | 101 .33 | 101.3 | 101 .33 | 101.3 | 101 .33 | 101 .33 | 1412.0 | 101.3 |
| **Density** | **kg/m3** | 1900.4 | 1.3 | 1232.6 | 997.0 | 1216.3 | 1,91 1.40 | 1,911 .40 | 0.6 | 1339.9 | 1.911.40 | 1216.3 | 18.2 | 1331.5 |
| **Solids** | **wt %** | 65.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids** | **t/h** | 34.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ. liquids** | **t/h** | 18.3 | 0.0 | 173.0 | 10.5 | 84.4 | 2.1 | 2.0 | 0.0 | 193.4 | 0.1 | 99.1 | 0.0 | 138.2 |
| **Gases** | **t/h** | 0.0 | 4.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 6.0 | 0.0 | 0.0 | 0.0 | 3.3 | 0.0 |
| **OrQ. liquids** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ. liquids: Zn** | **g/L** | 0.0 | 0.0 | 51 .76 | 0.0 | 48.2 | 0.0 | 0.0 | 0.0 | 160.0 | 0.0 | 48.2 | 0.0 | 132.1 |
| **AQ. LiQ.: aH2S04** | **g/L** | 0.0 | 0.0 | 185. 19 | 0.0 | 172.3 | 1,91 1.40 | 1911.4 | 0.0 | 8.2 | 1911.4 | 172.3 | 0.0 | 29.1 |
| **AQ. liquids: Fe** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 2.1 | 0.0 | 0.0 | 0.0 | 13.2 |
| **Solids: Zn** | **wt %** | 51.4 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 24.0 | 0.0 | 0.0 | 0.0 | 0.9 |
| **Solids: Fe** | **wt %** | 4.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 10.2 | 0.0 | 0.0 | 0.0 | 6.8 |
| **Solids: Cu** | **wt %** | 0.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.3 | 0.0 | 0.0 | 0.0 | 0.1 |
| **Solids: Cd** | **wt %** | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **wt %** | 28.7 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 39.9 | 0.0 | 0.0 | 0.0 | 57.8 |
| **Solids : Zn** | **t/h** | 17.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Fe** | **t/h** | 1.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids : Cu** | **t/h** | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cd** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **t/h** | 9.8 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ. Liquids : Zn** | **t/h** | 0.0 | 0.0 | 7.3 | 0.0 | 3.3 | 0.0 | 0.0 | 0.0 | 23.1 | 0.0 | 3.9 | 0.0 | 13.7 |
| **AQ. liquids: Fe** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.3 | 0.0 | 0.0 | 0.0 | 1.4 |
| **AQ. Liquids: aH2SO4** | **t/h** | 0.0 | 0.0 | 26.0 | 0.0 | 11 .96 | 2. 13 | 2.0 | 0.0 | 1.1 8 | 0.1 | 14.0 | 0.0 | 3.0 |
| **AQ. Liquids: SO4 2-** | **t/h** | 0.0 | 0.0 | 36.6 | 0.0 | 16.9 | 2.1 | 2.0 | 0.0 | 36.9 | 0.1 | 19.8 | 0.0 | 24.0 |
| **AQ. Liquids: SO4 2-** | **g/L** | 0.0 | 0.0 | 261.0 | 0.0 | 242.8 | 1872.1 | 1872.1 | 0.0 | 254.0 | 1.872.11 | 242.8 | 0.0 | 231.5 |
| **AQ . Liquids: Mn** | **g/L** | 0.0 | 0.0 | 2.0 | 0.0 | 1.9 | 0.0 | 0.0 | 0.0 | 4.2 | 0.0 | 1.9 | 0.0 | 3.7 |
| **AQ . Liquids: Cu** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1.0 | 0.0 | 0.0 | 0.0 | 0.9 |
| **AQ. liquids: Cd** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0.0 | 0.0 | 0.2 |
| **AQ. Liquids: Al giL** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.3 | 0.0 | 0.0 | 0.0 | 0.3 |
| **Enthalpy** | **kJ/kg** | -8003.9 | 0.0 | -1 3803.32 | -1 5877.10 | -1 3,921.60 | -8295.6 | -8295.6 | -1 3296.98 | -12810.6 | -8295.6 | -1 3921.60 | 0.0 | -12932.2 |

| Stream No. | | 273 | 314 | 316 | 318 | 334 | 339 | 348 | 401 | 402 | 432 | 438 | 446 | 457 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| **Stream Name** | | HAL Slurry Thickener Underflow | Sulphur Removal Filtrate to BZS | Sulphur Removal Filtrate to HAL thickener | Milk of Lime to Sulphur Recover | Sulphur Tails Thickener Overflow | Flotation Concentrati on Filrate | Elemental Sulphur | Fe Removal Thickener O/F | Zinc dust to Solution Purificat | Purified Solution to EW | Spent Electrolyte to Solution Purification | Hot Purification Cake | Cu Cake |
| **Flow** | **t/h** | 46.3 | 22.6 | 22.6 | 1.0 | 86.3 | 81.0 | 6.9 | 222.7 | 0.4 | 221.7 | 1.8 | 2.6 | 0.4 |
| **Flow** | **m3/h** | 29.4 | 18.7 | 18.7 | 0.9 | 88.4 | 83.1 | 3.3 | 169.8 | 0.1 | 165.3 | 1.4 | 1.5 | 0.2 |
| **Temp erature** | **C** | 69.7 | 40.9 | 40.9 | 25.0 | 41.0 | 41.3 | 140.0 | 77.2 | 25.0 | 30.0 | 38.0 | 76.7 | 68.0 |
| **Press ure** | **kPa** | 101.3 | 101.3 | 87.7 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 |
| **Density** | **kg/m3** | 1574.5 | 1208.5 | 1208.5 | 1157.6 | 975.7 | 975.5 | 2070.0 | 1311.3 | 7140.0 | 1341.3 | 1232.6 | 1694.6 | 2243.6 |
| **Solids** | **wt %** | 35.0 | 1.0 | 1.0 | 25.0 | 0.0 | 0.0 | 0.0 | 0.0 | 100.0 | 0.0 | 0.0 | 70.0 | 70.0 |
| **Solids** | **t/h** | 16.2 | 0.2 | 0.2 | 0.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.4 | 0.0 | 0.0 | 1.8 | 0.3 |
| **AQ. liquids** | **t/h** | 30.1 | 22.4 | 22.4 | 0.8 | 86.3 | 81.0 | 6.9 | 222.7 | 0.0 | 221.7 | 1.8 | 0.8 | 0.1 |
| **Gases** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **OrQ. liquids** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ. liquids: Zn** | **g/L** | 132.1 | 80.1 | 80.1 | 0.0 | 0.0 | 0.0 | 0.0 | 144.7 | 0.0 | 148.1 | 51.8 | 135.4 | 83.6 |
| **AQ. LiQ.: aH2S04** | **g/L** | 29.1 | 17.6 | 17.6 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 185.2 | 0.0 | 0.0 |
| **AQ. liquids: Fe** | **g/L** | 13.2 | 8.0 | 8.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Zn** | **wt %** | 0.9 | 0.9 | 0.9 | 0.0 | 0.3 | 0.0 | 0.0 | 0.0 | 100.0 | 0.0 | 0.0 | 22.7 | 16.2 |
| **Solids: Fe** | **wt %** | 6.8 | 6.8 | 6.8 | 0.0 | 11.8 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cu** | **wt %** | 0.1 | 0.1 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 52.0 |
| **Solids: Cd** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1.1 |
| **Solids: S** | **wt %** | 57.8 | 57.8 | 57.8 | 0.0 | 24.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 11.2 | 4.4 |
| **Solids : Zn** | **t/h** | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.4 | 0.0 | 0.0 | 0.4 | 0.0 |
| **Solids: Fe** | **t/h** | 1.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids : Cu** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 |
| **Solids: Cd** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **t/h** | 9.4 | 0.1 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 |
| **AQ. Liquids : Zn** | **t/h** | 3.0 | 1.5 | 1.5 | 0.0 | 0.0 | 0.0 | 0.0 | 24.6 | 0.0 | 24.5 | 0.1 | 0.1 | 0.0 |
| **AQ. liquids: Fe** | **t/h** | 0.3 | 0.1 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ. Liquids: aH2SO4** | **t/h** | 0.7 | 0.3 | 0.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.3 | 0.0 | 0.0 |
| **AQ. Liquids: SO4 2-** | **t/h** | 5.2 | 2.6 | 2.6 | 0.0 | 0.0 | 0.0 | 0.0 | 37.4 | 0.0 | 37.0 | 0.4 | 0.1 | 0.0 |
| **AQ. Liquids: SO4 2-** | **g/L** | 231.5 | 140.3 | 140.3 | 0.0 | 0.1 | 0.1 | 0.0 | 220.4 | 0.0 | 223.9 | 261.0 | 204.7 | 132.8 |
| **AQ . Liquids: Mn** | **g/L** | 3.7 | 2.2 | 2.2 | 0.0 | 0.0 | 0.0 | 0.0 | 3.6 | 0.0 | 3.7 | 2.0 | 3.3 | 1.1 |
| **AQ . Liquids: Cu** | **g/L** | 0.9 | 0.5 | 0.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.9 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ. liquids: Cd** | **g/L** | 0.2 | 0.1 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 | 9.5 |
| **AQ. Liquids: Al giL** | **g/L** | 0.3 | 0.2 | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Enthalpy** | **kJ/kg** | -10261.8 | -13844.2 | -13844.2 | -15234.3 | -15808.8 | -15808.2 | 158.0 | -13117.0 | 0.0 | -13268.0 | -13803.3 | -11425.7 | -7121.1 |

| Stream No. | | 469 | 473 | 601 | 603 | 604 | 605 | 606 | 607 | 608 | 611 |
|---|---|---|---|---|---|---|---|---|---|---|---|
| **Stream Name** | | Cd Cake | Air to Purification Cooling Tower | Total Process Water | Total Oxygen | Total Steam | Total Sulphuric Acid | Total Flotation Reagents | Total Milk of Lime | Total Flocculant | Total GSW |
| **Flow** | **t/h** | 0.1 | 508.8 | 22.5 | 7.6 | 7.1 | 4.1 | 0.1 | 2.0 | 10.3 | 19.9 |
| **Flow** | **m3/h** | 0.0 | 431485.6 | 22.6 | 418.2 | 1379.9 | 2.2 | 0.1 | 1.7 | 10.3 | 20.0 |
| **Temp erature** | **C** | 69.9 | 25.0 | 25.0 | 25.0 | 179.9 | 25.0 | 25.0 | 25.0 | 25.0 | 25.0 |
| **Press ure** | **kPa** | 101.3 | 101.3 | 101.3 | 1412.0 | 1000.0 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 |
| **Density** | **kg/m3** | 2273.1 | 1.2 | 997.0 | 18.2 | 5.1 | 1911.4 | 997.0 | 1157.6 | 997.0 | 997.0 |
| **Solids** | **wt %** | 70.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 25.0 | 0.0 | 0.0 |
| **Solids** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.5 | 0.0 | 0.0 |
| **AQ. liquids** | **t/h** | 0.0 | 0.0 | 22.5 | 0.0 | 0.0 | 4.1 | 0.1 | 1.5 | 10.3 | 19.9 |
| **Gases** | **t/h** | 0.0 | 508.8 | 0.0 | 7.6 | 7.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **OrQ. liquids** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ. liquids: Zn** | **g/L** | 89.4 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ. LiQ.: aH2S04** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1911.4 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ. liquids: Fe** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Zn** | **wt %** | 8.7 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Fe** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cu** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cd** | **wt %** | 61.6 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **wt %** | 4.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids : Zn** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Fe** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids : Cu** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cd** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ. Liquids : Zn** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ. liquids: Fe** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ. Liquids: aH2SO4** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 4.1 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ. Liquids: SO4 2-** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 4.1 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ. Liquids: SO4 2-** | **g/L** | 133.3 | 0.0 | 0.0 | 0.0 | 0.0 | 1872.1 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ . Liquids: Mn** | **g/L** | 1.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ . Liquids: Cu** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ. liquids: Cd** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **AQ. Liquids: Al giL** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Enthalpy** | **kJ/kg** | -7055.0 | 0.0 | -15877.1 | 0.0 | -13200.9 | -8295.6 | -15877.1 | -15234.3 | -15877.1 | -15877.1 |

## Appendix 5: Option E mass & energy balance




Skorpion Zinc

OPTION E - BLOCK FLOW DIAGRAM

| Stream No. | | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | Total Dry Limestone | Limestone to Neutralisation | Limestone to BZS | Acid to Neutralisation | 2nd Stage Fe Rem. Discharge | Flocc. to Neut. Thickener | Neut. Thickener U/F | Flocc. to Neut. BF | Neut. BF Cake | Neut. BF Prim. Filtrate | Neut. BF Sec. Filtrate | Milk of Lime to BZS | BZS Cake |
| **Flow** | **t/h** | 30.9 | 5.4 | 35.3 | 3.4 | 314.0 | 0.6 | 57.9 | 0.4 | 29.9 | 25.5 | 18.4 | 0.9 | 37.8 |
| **Flow** | **m3/h** | 11.5 | 4.5 | 29.4 | 1.8 | 231.3 | 0.6 | 37.2 | 0.4 | 18.6 | 19.4 | 15.3 | 0.8 | 25.5 |
| **Temperature** | **C** | 25.0 | 63.1 | 63.1 | 25.0 | 78.1 | 25.0 | 77.9 | 25.0 | 73.5 | 77.3 | 73.5 | 25.0 | 47.8 |
| **Pressure** | **kPa** | 101.3 | 87.7 | 87.7 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 |
| **Density** | **kg/m3** | 2681.6 | 1202.6 | 1202.6 | 1887.9 | 1357.5 | 997.0 | 1557.9 | 997.0 | 1604.0 | 1314.3 | 1196.4 | 1157.6 | 1480.4 |
| **Solids** | **wt %** | 100.0 | 29.0 | 29.0 | 0.0 | 6.3 | 0.0 | 34.0 | 0.0 | 66.0 | 0.0 | 0.0 | 25.0 | 60.0 |
| **Solids** | **t/h** | 30.9 | 1.6 | 10.2 | 0.0 | 19.7 | 0.0 | 19.7 | 0.0 | 19.7 | 0.0 | 0.0 | 0.2 | 22.7 |
| **Aq. Liquids** | **t/h** | 0.0 | 3.8 | 25.1 | 3.4 | 294.3 | 0.6 | 38.2 | 0.4 | 10.2 | 25.5 | 18.4 | 0.7 | 15.1 |
| **Gases** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Org. Liquids** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Zn** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 142.6 | 0.0 | 142.2 | 0.0 | 0.5 | 140.4 | 91.6 | 0.0 | 0.5 |
| **Aq. Liq.: aH2SO4** | **g/L** | 0.0 | 0.0 | 0.0 | 1850.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Fe** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Zn** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 14.4 |
| **Solids: Fe** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 2.6 | 0.0 | 2.6 | 0.0 | 2.6 | 0.0 | 0.0 | 0.0 | 0.7 |
| **Solids: Cu** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 |
| **Solids: Cd** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 15.3 | 0.0 | 15.3 | 0.0 | 15.3 | 0.0 | 0.0 | 0.0 | 12.2 |
| **Solids: Zn** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 3.3 |
| **Solids: Fe** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.5 | 0.0 | 0.5 | 0.0 | 0.5 | 0.0 | 0.0 | 0.0 | 0.2 |
| **Solids: Cu** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cd** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 3.0 | 0.0 | 3.0 | 0.0 | 3.0 | 0.0 | 0.0 | 0.0 | 2.8 |
| **Aq. Liquids: Zn** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 31.8 | 0.0 | 4.1 | 0.0 | 0.0 | 2.7 | 1.4 | 0.0 | 0.0 |
| **Aq. Liq: Fe** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq.: aH2SO4** | **t/h** | 0.0 | 0.0 | 0.0 | 3.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq: SO4[2-]** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: SO4[2-]** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Mn** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 8.5 | 0.0 | 8.5 | 0.0 | 1.8 | 8.4 | 6.1 | 0.0 | 1.9 |
| **Aq. Liquids: Cu** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.8 | 0.0 | 0.8 | 0.0 | 0.0 | 0.8 | 0.5 | 0.0 | 0.0 |
| **Aq. Liquids: Cd** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0.2 | 0.0 | 0.0 | 0.2 | 0.1 | 0.0 | 0.0 |
| **Aq. Liquids: Al** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Enthalpy** | **kJ/kg** | -12831.3 | -14871.4 | -14871.4 | -8486.6 | -13011.7 | -15877.1 | -12695.6 | -15877.1 | -13190.2 | -13117.1 | -13836.6 | -15234.3 | -12784.3 |

| Stream No. | | 21 | 22 | 23 | 24 | 25 | 27 | 31 | 32 | 33 | 39 | 40 | 42 | 43 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | BZS Wash Water | BZS Bleed to ETP | PLS | Process Water to Limestone | Excess ETP Water | ETP Residue | Extraction Raffinate | Raff Regen. Acid | Raff Make-up Water | SX Make-up Water | SX Make-upAcid | Plated Zn | Spent Electrolyte to SX |
| Flow | t/h | 15.4 | 184.8 | 256.6 | 75.6 | 106.3 | 5.2 | 240.1 | 6.4 | 2.3 | 47.5 | 4.9 | 17.3 | 274.2 |
| Flow | m3/h | 15.8 | 189.3 | 194.7 | 77.2 | 108.5 | 4.0 | 196.0 | 3.4 | 2.3 | 47.6 | 2.6 | 2.4 | 221.3 |
| Temperature | C | 70.0 | 70.0 | 77.9 | 66.2 | 66.2 | 66.2 | 34.5 | 25.0 | 25.0 | 25.0 | 25.0 | 38.0 | 38.0 |
| Pressure | kPa | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 |
| Density | kg/m3 | 975.9 | 975.9 | 1318.2 | 979.8 | 979.8 | 1301.6 | 1225.2 | 1887.9 | 997.0 | 997.0 | 1887.9 | 7140.0 | 1239.3 |
| Solids | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 40.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 100.0 | 0.0 |
| Solids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 2.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 17.3 | 0.0 |
| Aq. Liquids | t/h | 15.4 | 184.8 | 256.6 | 75.6 | 106.3 | 3.1 | 240.1 | 6.4 | 2.3 | 47.5 | 4.9 | 0.0 | 274.2 |
| Gases | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Org. Liquids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Zn | g/L | 0.5 | 0.5 | 142.2 | 0.0 | 0.0 | 0.0 | 56.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 52.3 |
| Aq. Liq.: aH2SO4 | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 126.7 | 1850.1 | 0.0 | 0.0 | 1850.1 | 0.0 | 198.7 |
| Aq. Liquids: Fe | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 4.4 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 100.0 | 0.0 |
| Solids: Fe | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Cu | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Cd | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 12.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 17.3 | 0.0 |
| Solids: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Cu | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Cd | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Zn | t/h | 0.0 | 0.1 | 27.7 | 0.0 | 0.0 | 0.0 | 11.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 11.6 |
| Aq. Liq: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 24.8 | 6.3 | 0.0 | 0.0 | 4.8 | 0.0 | 44.0 |
| Aq. Liq: SO4[2-] | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: SO4[2-] | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Mn | g/L | 1.8 | 1.8 | 8.5 | 0.0 | 0.0 | 0.0 | 8.4 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Cu | g/L | 0.0 | 0.0 | 0.8 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Cd | g/L | 0.0 | 0.0 | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Al | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Enthalpy | kJ/kg | -15635.1 | -15635.1 | -13089.2 | -15704.6 | -15704.6 | -13537.3 | -13926.9 | -8486.6 | -15877.1 | -15877.1 | -8486.6 | 5.1 | -13766.7 |

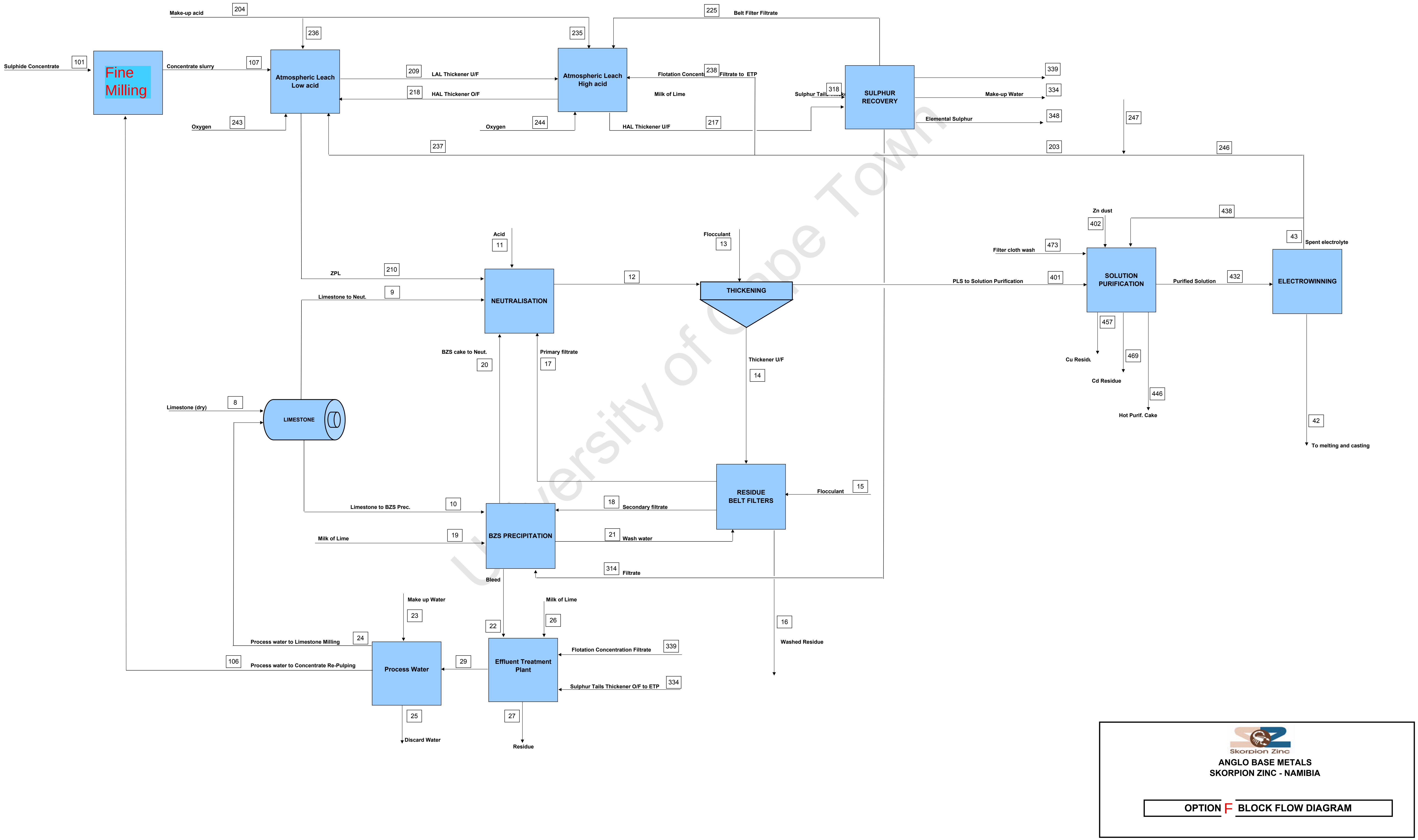
| Stream No. | | 101 | 106 | 107 | 206 | 207 | 208 | 211 | 212 | 237 | 242 | 273 | 307 | 314 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| **Stream Name** | | Zinc Sulphide Concentrate Feed | Water to Concentrate Re-Pulping | Zinc Sulphide Concentrate Slurry | Oxygen to PAL | Oxygen to LAL | Raffinate to PAL | Acid Make-up to PAL | Acid Make-up to LAL | LAL Slurry Thickener Overflow | Acid Make-upto HAL | HAL Slurry Thickener Underflow | Wash Water to PAL Discharge BF | Sulphur Removal Filtrate to BZS |
| **Flow** | **t/h** | 34.5 | 18.6 | 53.1 | 7.7 | 4.4 | 216.9 | 1.4 | 1.1 | 223.1 | 0.3 | 46.8 | 24.5 | 23.3 |
| **Flow** | **m3/h** | 9.1 | 19.0 | 28.0 | 5914.1 | 3379.6 | 175.1 | 0.8 | 0.6 | 165.8 | 0.2 | 29.6 | 24.6 | 19.3 |
| **Temperature** | **C** | 25.0 | 66.2 | 58.4 | 25.0 | 25.0 | 40.0 | 25.0 | 25.0 | 79.6 | 25.0 | 71.1 | 25.0 | 41.6 |
| **Pressure** | **kPa** | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 87.7 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 87.7 | 101.3 |
| **Density** | **kg/m3** | 3777.0 | 979.8 | 1894.7 | 1.3 | 1.3 | 1239.1 | 1887.9 | 1887.9 | 1345.8 | 1887.9 | 1581.8 | 997.0 | 1208.4 |
| **Solids** | **wt %** | 100.0 | 0.0 | 65.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 35.0 | 0.0 | 1.0 |
| **Solids** | **t/h** | 34.5 | 0.0 | 34.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 16.4 | 0.0 | 0.2 |
| **Aq. Liquids** | **t/h** | 0.0 | 18.6 | 18.6 | 0.0 | 0.0 | 216.9 | 1.4 | 1.1 | 223.1 | 0.3 | 30.5 | 24.5 | 23.0 |
| **Gases** | **t/h** | 0.0 | 0.0 | 0.0 | 7.7 | 4.4 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Org. Liquids** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Zn** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 55.4 | 0.0 | 0.0 | 155.8 | 0.0 | 129.2 | 0.0 | 76.5 |
| **Aq. Liq.: aH2SO4** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 156.5 | 1850.1 | 1850.1 | 8.2 | 1850.1 | 29.1 | 0.0 | 17.3 |
| **Aq. Liquids: Fe** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 2.1 | 0.0 | 13.2 | 0.0 | 7.8 |
| **Solids: Zn** | **wt %** | 51.4 | 0.0 | 51.4 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 23.8 | 0.0 | 0.9 | 0.0 | 0.9 |
| **Solids: Fe** | **wt %** | 4.5 | 0.0 | 4.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 10.6 | 0.0 | 6.5 | 0.0 | 6.5 |
| **Solids: Cu** | **wt %** | 0.5 | 0.0 | 0.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.3 | 0.0 | 0.1 | 0.0 | 0.1 |
| **Solids: Cd** | **wt %** | 0.1 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **wt %** | 28.7 | 0.0 | 28.7 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 39.6 | 0.0 | 58.0 | 0.0 | 58.0 |
| **Solids: Zn** | **t/h** | 17.7 | 0.0 | 17.7 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 |
| **Solids: Fe** | **t/h** | 1.6 | 0.0 | 1.6 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1.1 | 0.0 | 0.0 |
| **Solids: Cu** | **t/h** | 0.2 | 0.0 | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cd** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **t/h** | 9.9 | 0.0 | 9.9 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 9.5 | 0.0 | 0.1 |
| **Aq. Liquids: Zn** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 9.7 | 0.0 | 0.0 | 25.8 | 0.0 | 2.9 | 0.0 | 1.5 |
| **Aq. Liq: Fe** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.3 | 0.0 | 0.3 | 0.0 | 0.2 |
| **Aq. Liq.: aH2SO4** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 27.4 | 1.4 | 1.1 | 1.4 | 0.3 | 0.7 | 0.0 | 0.3 |
| **Aq. Liq: SO4[2-]** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: SO4[2-]** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Mn** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 8.3 | 0.0 | 0.0 | 10.3 | 0.0 | 9.9 | 0.0 | 5.9 |
| **Aq. Liquids: Cu** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.9 | 0.0 | 0.8 | 0.0 | 0.5 |
| **Aq. Liquids: Cd** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0.2 | 0.0 | 0.1 |
| **Aq. Liquids: Al** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0.3 | 0.0 | 0.2 |
| **Enthalpy** | **kJ/kg** | -3831.4 | -15704.6 | -7987.0 | 0.0 | 0.0 | -13783.8 | -8486.6 | -8486.6 | -12791.4 | -8486.6 | -10234.6 | -15877.1 | -13849.4 |

| Stream No. | | 316 | 333 | 334 | 339 | 347 | 348 | 401 | 402 | 446 | 457 | 469 | 481 | 489 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| **Stream Name** | | Sulphur Removal Filtrate to HAL | Sulphur Flotation Tails | Sulphur Tails Thickener Overflow | Flotation Concentration Filtrate | Sulphur Pressure Filter Cake | Elemental Sulphur | Fe Removal Thickener O/F | Zinc dust to Solution Purification | Hot Purification Cake | Cu Cake | Cd Cake | Limestone to Bleed Neutr. | Wash Water to Bleed BF |
| **Flow** | **t/h** | 23.3 | 21.2 | 16.0 | 2.6 | 0.3 | 7.4 | 256.6 | 0.4 | 2.6 | 0.4 | 0.1 | 65.9 | 19.4 |
| **Flow** | **m3/h** | 19.3 | 16.2 | 16.3 | 2.6 | 0.1 | 3.6 | 194.7 | 0.1 | 1.6 | 0.2 | 0.0 | 54.8 | 19.4 |
| **Temperature** | **C** | 41.6 | 30.3 | 30.3 | 30.5 | 140.0 | 140.0 | 77.9 | 25.0 | 77.1 | 68.8 | 70.7 | 63.1 | 25.0 |
| **Pressure** | **kPa** | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 87.7 | 101.3 |
| **Density** | **kg/m3** | 1208.4 | 1313.8 | 979.5 | 979.4 | 3186.0 | 2070.0 | 1318.2 | 7140.0 | 1699.4 | 2243.6 | 2275.6 | 1202.6 | 997.0 |
| **Solids** | **wt %** | 1.0 | 40.0 | 0.0 | 0.0 | 98.0 | 0.0 | 0.0 | 100.0 | 70.0 | 70.0 | 70.0 | 29.0 | 0.0 |
| **Solids** | **t/h** | 0.2 | 8.5 | 0.0 | 0.0 | 0.3 | 0.0 | 0.0 | 0.4 | 1.8 | 0.3 | 0.1 | 19.1 | 0.0 |
| **Aq. Liquids** | **t/h** | 23.0 | 12.7 | 16.0 | 2.6 | 0.0 | 7.4 | 256.6 | 0.0 | 0.8 | 0.1 | 0.0 | 46.8 | 19.4 |
| **Gases** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Org. Liquids** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Zn** | **g/L** | 76.5 | 0.1 | 0.1 | 0.1 | 0.0 | 0.0 | 142.2 | 0.0 | 134.2 | 84.6 | 90.5 | 0.0 | 0.0 |
| **Aq. Liq.: aH2SO4** | **g/L** | 17.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Fe** | **g/L** | 7.8 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Zn** | **wt %** | 0.9 | 0.3 | 0.3 | 0.0 | 33.8 | 0.0 | 0.0 | 100.0 | 22.7 | 16.2 | 8.7 | 0.0 | 0.0 |
| **Solids: Fe** | **wt %** | 6.5 | 12.0 | 12.0 | 0.0 | 7.4 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cu** | **wt %** | 0.1 | 0.0 | 0.0 | 0.0 | 2.9 | 0.0 | 0.0 | 0.0 | 0.0 | 52.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cd** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 1.1 | 61.7 | 0.0 | 0.0 |
| **Solids: S** | **wt %** | 58.0 | 20.6 | 20.6 | 0.0 | 41.5 | 0.0 | 0.0 | 0.0 | 11.2 | 4.4 | 4.3 | 0.0 | 0.0 |
| **Solids: Zn** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 | 0.4 | 0.4 | 0.1 | 0.0 | 0.0 | 0.0 |
| **Solids: Fe** | **t/h** | 0.0 | 1.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cu** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0.0 | 0.0 |
| **Solids: Cd** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **t/h** | 0.1 | 1.8 | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Zn** | **t/h** | 1.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 27.7 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq: Fe** | **t/h** | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq.: aH2SO4** | **t/h** | 0.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq: SO4[2-]** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: SO4[2-]** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Mn** | **g/L** | 5.9 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 8.5 | 0.0 | 8.0 | 0.5 | 0.5 | 0.0 | 0.0 |
| **Aq. Liquids: Cu** | **g/L** | 0.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.8 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Cd** | **g/L** | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0.0 | 9.6 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Al** | **g/L** | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Enthalpy** | **kJ/kg** | -13849.4 | -13616.7 | -15852.2 | -15851.4 | -3152.1 | 157.9 | | | | -7118.9 | -7040.7 | -14871.4 | -15877.1 |

| Stream No. | | 490 | 492 | 502 | 506 | 508 | 510 | 511 | 513 | 516 | 517 | 522 | 524 | 527 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| **Stream Name** | | Neutralised Raff. to Depletion | Bleed BF Cake to BZS | Cooled Purified PLS | Raffinate to Bleed | Loaded Extraction Organic | Stripped Organic toExtraction | EW Bleed & Wash Water | Stripped Organic to Depletion | Loaded Depletion Organic | Depletion Raffinate to BZS | Demin Water to SX | Total Loaded Organic | Stripped Organic |
| **Flow** | **t/h** | 145.2 | 30.3 | 256.1 | 29.6 | 988.7 | 972.7 | 82.3 | 95.1 | 96.7 | 143.7 | 60.2 | 1085.4 | 1067.8 |
| **Flow** | **m3/h** | 142.1 | 15.7 | 190.0 | 23.9 | 1234.6 | 1215.8 | 78.9 | 118.9 | 120.7 | 142.6 | 60.4 | 1355.3 | 1334.6 |
| **Temperature** | **C** | 48.5 | 48.3 | 29.9 | 40.0 | 34.5 | 36.1 | 34.6 | 36.1 | 45.1 | 45.1 | 25.0 | 35.5 | 36.1 |
| **Pressure** | **kPa** | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 87.7 |
| **Density** | **kg/m3** | 1022.2 | 1928.4 | 1348.0 | 1239.1 | 800.8 | 800.1 | 1042.9 | 800.1 | 800.8 | 1007.7 | 997.0 | 800.8 | 800.1 |
| **Solids** | **wt %** | 0.0 | 80.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids** | **t/h** | 0.0 | 24.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids** | **t/h** | 145.2 | 6.1 | 256.1 | 29.6 | 0.0 | 0.0 | 82.3 | 0.0 | 0.0 | 143.7 | 60.2 | 0.0 | 0.0 |
| **Gases** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Org. Liquids** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 988.7 | 972.7 | 0.0 | 95.1 | 96.7 | 0.0 | 0.0 | 1085.4 | 1067.8 |
| **Aq. Liquids: Zn** | **g/L** | 14.2 | 14.1 | 145.1 | 55.4 | 0.0 | 0.0 | 9.9 | 0.0 | 0.0 | 2.8 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq.: aH2SO4** | **g/L** | 0.1 | 0.1 | 0.0 | 156.5 | 0.0 | 0.0 | 41.8 | 0.0 | 0.0 | 17.1 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Fe** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Zn** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Fe** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cu** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cd** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **wt %** | 0.0 | 9.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Zn** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Fe** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cu** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cd** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **t/h** | 0.0 | 2.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Zn** | **t/h** | 2.0 | 0.1 | 27.6 | 1.3 | 0.0 | 0.0 | 0.8 | 0.0 | 0.0 | 0.4 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq: Fe** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq.: aH2SO4** | **t/h** | 0.0 | 0.0 | 0.0 | 3.7 | 0.0 | 0.0 | 3.3 | 0.0 | 0.0 | 2.4 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq: SO4[2-]** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: SO4[2-]** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Mn** | **g/L** | 1.3 | 1.3 | 8.7 | 8.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1.3 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Cu** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Cd** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Al** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Enthalpy** | **kJ/kg** | -15433.5 | -13077.6 | -13245.6 | -13783.8 | -2798.2 | -2833.0 | -15348.1 | -2833.0 | -2775.8 | -15584.3 | -15877.1 | -2796.2 | -2833.0 |

| Stream No. | | 531 | 545 | 601 | 603 | 604 | 605 | 606 | 607 | 608 | 611 |
|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | Spent Electrolyte to Scrub | Advance Electrolyte to EW | Total Process Water | Total Oxygen | Total Steam | Total Sulphuric Acid | Total Flotation Reagents | Total Milk of Lime | Total Flocculant | Total GSW |
| Flow | t/h | 22.1 | 322.2 | 126.3 | 7.7 | 14.6 | 16.2 | 0.1 | 4.2 | 12.3 | 19.9 |
| Flow | m3/h | 18.3 | 244.9 | 126.7 | 5933.0 | 2831.6 | 8.6 | 0.1 | 3.7 | 12.3 | 20.0 |
| Temperature | C | 38.8 | 36.1 | 25.0 | 25.0 | 179.9 | 25.0 | 25.0 | 25.0 | 25.0 | 25.0 |
| Pressure | kPa | 101.3 | 101.3 | 101.3 | 101.3 | 1000.0 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 |
| Density | kg/m3 | 1205.0 | 1315.3 | 997.0 | 1.3 | 5.2 | 1887.9 | 997.0 | 1157.6 | 997.0 | 997.0 |
| Solids | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 25.0 | 0.0 | 0.0 |
| Solids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1.1 | 0.0 | 0.0 |
| Aq. Liquids | t/h | 22.1 | 322.2 | 126.3 | 0.0 | 0.0 | 16.2 | 0.1 | 3.2 | 12.3 | 19.9 |
| Gases | t/h | 0.0 | 0.0 | 0.0 | 7.7 | 14.6 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Org. Liquids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Zn | g/L | 42.7 | 118.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | g/L | 180.0 | 74.5 | 0.0 | 0.0 | 0.0 | 1850.1 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Fe | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Fe | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Cu | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Cd | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Cu | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Cd | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Zn | t/h | 0.8 | 29.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | t/h | 3.3 | 18.3 | 0.0 | 0.0 | 0.0 | 15.9 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: SO4[2-] | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: SO4[2-] | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Mn | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Cu | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Cd | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Al | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Enthalpy | kJ/kg | -13993.8 | -13395.6 | -15877.1 | 0.0 | -13200.9 | -8486.6 | -15877.1 | -15234.3 | -15877.1 | -15877.1 |

## APPENDIX 6: OPTION F MASS & ENERGY BALANCE


Sulphide Concentrate
101
Fine Milling
Concentrate slurry
107
Make-up acid
204
236
Atmospheric Leach Low acid
209
LAL Thickener U/F
218
HAL Thickener O/F
Oxygen
243
237
235
Atmospheric Leach High acid
Oxygen
244
HAL Thickener U/F
217
225
Belt Filter Filtrate
Flotation Concentrate
238
Filtrate to ETP
Milk of Lime
Sulphur Tails
318
SULPHUR RECOVERY
339
Make-up Water
334
Elemental Sulphur
348
247
203
246
Acid
11
ZPL
210
Limestone to Neut.
9
NEUTRALISATION
12
Flocculant
13
THICKENING
PLS to Solution Purification
401
Zn dust
402
438
Filter cloth wash
473
43
Spent electrolyte
SOLUTION PURIFICATION
Purified Solution
432
ELECTROWINNING
457
Cu Residue
469
Cd Residue
446
Hot Purif. Cake
42
To melting and casting
BZS cake to Neut.
20
Primary filtrate
17
Thickener U/F
14
Limestone (dry)
8
LIMESTONE
RESIDUE BELT FILTERS
15
Flocculant
Limestone to BZS Prec.
10
18
Secondary filtrate
Milk of Lime
19
BZS PRECIPITATION
21
Wash water
314
Filtrate
Bleed
Make up Water
23
Milk of Lime
26
22
16
Washed Residue
Process water to Limestone Milling
24
339
Flotation Concentration Filtrate
106
Process water to Concentrate Re-Pulping
Process Water
29
Effluent Treatment Plant
334
Sulphur Tails Thickener O/F to ETP
25
Discard Water
27
Residue
Skorpion Zinc
ANGLO BASE METALS
SKORPION ZINC - NAMIBIA
OPTION F BLOCK FLOW DIAGRAM

| Stream No. | | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | Total Dry Limestone | Limestone to Neutralisation | Limestone to BZS | Acid to Neutralisation | 2nd Stage Fe Rem. Discharge | Flocc. to Neut. Thickener | Neut. Thickener U/F | Flocc. to Neut. BF | Neut. BF Cake | Neut. BF Prim. Filtrate | Neut. BF Sec. Filtrate | Milk of Lime to BZS | BZS Cake |
| Flow | t/h | 30.9 | 5.4 | 35.3 | 3.4 | 314.0 | 0.6 | 57.9 | 0.4 | 29.9 | 25.5 | 18.4 | 0.9 | 37.8 |
| Flow | m3/h | 11.5 | 4.5 | 29.4 | 1.8 | 231.3 | 0.6 | 37.2 | 0.4 | 18.6 | 19.4 | 15.3 | 0.8 | 25.5 |
| Temperature | C | 25.0 | 63.1 | 63.1 | 25.0 | 78.1 | 25.0 | 77.9 | 25.0 | 73.5 | 77.3 | 73.5 | 25.0 | 47.8 |
| Pressure | kPa | 101.3 | 87.7 | 87.7 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 |
| Density | kg/m3 | 2681.6 | 1202.6 | 1202.6 | 1887.9 | 1357.5 | 997.0 | 1557.9 | 997.0 | 1604.0 | 1314.3 | 1196.4 | 1157.6 | 1480.4 |
| Solids | wt % | 100.0 | 29.0 | 29.0 | 0.0 | 6.3 | 0.0 | 34.0 | 0.0 | 66.0 | 0.0 | 0.0 | 25.0 | 60.0 |
| Solids | t/h | 30.9 | 1.6 | 10.2 | 0.0 | 19.7 | 0.0 | 19.7 | 0.0 | 19.7 | 0.0 | 0.0 | 0.2 | 22.7 |
| Aq. Liquids | t/h | 0.0 | 3.8 | 25.1 | 3.4 | 294.3 | 0.6 | 38.2 | 0.4 | 10.2 | 25.5 | 18.4 | 0.7 | 15.1 |
| Gases | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Org. Liquids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Zn | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 142.6 | 0.0 | 142.2 | 0.0 | 0.5 | 140.4 | 91.6 | 0.0 | 0.5 |
| Aq. Liq.: aH2SO4 | g/L | 0.0 | 0.0 | 0.0 | 1850.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Fe | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 14.4 |
| Solids: Fe | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 2.6 | 0.0 | 2.6 | 0.0 | 2.6 | 0.0 | 0.0 | 0.0 | 0.7 |
| Solids: Cu | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 |
| Solids: Cd | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 15.3 | 0.0 | 15.3 | 0.0 | 15.3 | 0.0 | 0.0 | 0.0 | 12.2 |
| Solids: Zn | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 3.3 |
| Solids: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.5 | 0.0 | 0.5 | 0.0 | 0.5 | 0.0 | 0.0 | 0.0 | 0.2 |
| Solids: Cu | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Cd | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 3.0 | 0.0 | 3.0 | 0.0 | 3.0 | 0.0 | 0.0 | 0.0 | 2.8 |
| Aq. Liquids: Zn | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 31.8 | 0.0 | 4.1 | 0.0 | 0.0 | 2.7 | 1.4 | 0.0 | 0.0 |
| Aq. Liq: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | t/h | 0.0 | 0.0 | 0.0 | 3.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: SO4[2-] | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: SO4[2-] | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Mn | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 8.5 | 0.0 | 8.5 | 0.0 | 1.8 | 8.4 | 6.1 | 0.0 | 1.9 |
| Aq. Liquids: Cu | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.8 | 0.0 | 0.8 | 0.0 | 0.0 | 0.8 | 0.5 | 0.0 | 0.0 |
| Aq. Liquids: Cd | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0.2 | 0.0 | 0.0 | 0.2 | 0.1 | 0.0 | 0.0 |
| Aq. Liquids: Al | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Enthalpy | kJ/kg | -12831.3 | -14871.4 | -14871.4 | -8486.6 | -13011.7 | -15877.1 | -12695.6 | -15877.1 | -13190.2 | -13117.1 | -13836.6 | -15234.3 | -12784.3 |

| Stream No. | | 21 | 22 | 23 | 24 | 25 | 27 | 31 | 32 | 33 | 39 | 40 | 42 | 43 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| **Stream Name** | | BZS Wash Water | BZS Bleed to ETP | PLS | Process Water to Limestone | Excess ETP Water | ETP Residue | Extraction Raffinate | Raff Regen. Acid | Raff Make-up Water | SX Make-up Water | SX Make-upAcid | Plated Zn | Spent Electrolyte to SX |
| **Flow** | **t/h** | 15.4 | 184.8 | 256.6 | 75.6 | 106.3 | 5.2 | 240.1 | 6.4 | 2.3 | 47.5 | 4.9 | 17.3 | 274.2 |
| **Flow** | **m3/h** | 15.8 | 189.3 | 194.7 | 77.2 | 108.5 | 4.0 | 196.0 | 3.4 | 2.3 | 47.6 | 2.6 | 2.4 | 221.3 |
| **Temperature** | **C** | 70.0 | 70.0 | 77.9 | 66.2 | 66.2 | 66.2 | 34.5 | 25.0 | 25.0 | 25.0 | 25.0 | 38.0 | 38.0 |
| **Pressure** | **kPa** | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 |
| **Density** | **kg/m3** | 975.9 | 975.9 | 1318.2 | 979.8 | 979.8 | 1301.6 | 1225.2 | 1887.9 | 997.0 | 997.0 | 1887.9 | 7140.0 | 1239.3 |
| **Solids** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 40.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 100.0 | 0.0 |
| **Solids** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 2.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 17.3 | 0.0 |
| **Aq. Liquids** | **t/h** | 15.4 | 184.8 | 256.6 | 75.6 | 106.3 | 3.1 | 240.1 | 6.4 | 2.3 | 47.5 | 4.9 | 0.0 | 274.2 |
| **Gases** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Org. Liquids** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Zn** | **g/L** | 0.5 | 0.5 | 142.2 | 0.0 | 0.0 | 0.0 | 56.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 52.3 |
| **Aq. Liq.: aH2SO4** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 126.7 | 1850.1 | 0.0 | 0.0 | 1850.1 | 0.0 | 198.7 |
| **Aq. Liquids: Fe** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Zn** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 4.4 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 100.0 | 0.0 |
| **Solids: Fe** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cu** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cd** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 12.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Zn** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 17.3 | 0.0 |
| **Solids: Fe** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cu** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cd** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Zn** | **t/h** | 0.0 | 0.1 | 27.7 | 0.0 | 0.0 | 0.0 | 11.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 11.6 |
| **Aq. Liq: Fe** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq.: aH2SO4** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 24.8 | 6.3 | 0.0 | 0.0 | 4.8 | 0.0 | 44.0 |
| **Aq. Liq: SO4[2-]** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: SO4[2-]** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Mn** | **g/L** | 1.8 | 1.8 | 8.5 | 0.0 | 0.0 | 0.0 | 8.4 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Cu** | **g/L** | 0.0 | 0.0 | 0.8 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Cd** | **g/L** | 0.0 | 0.0 | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Al** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Enthalpy** | **kJ/kg** | -15635.1 | -15635.1 | -13089.2 | -15704.6 | -15704.6 | -13537.3 | -13926.9 | -8486.6 | -15877.1 | -15877.1 | -8486.6 | 5.1 | -13766.7 |

| Stream No. | | 101 | 106 | 107 | 206 | 207 | 208 | 211 | 212 | 237 | 242 | 273 | 307 | 314 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| **Stream Name** | | Zinc Sulphide Concentrate Feed | Water to Concentrate Re-Pulping | Zinc Sulphide Concentrate Slurry | Oxygen to PAL | Oxygen to LAL | Raffinate to PAL | Acid Make-up to PAL | Acid Make-up to LAL | LAL Slurry Thickener Overflow | Acid Make-upto HAL | HAL Slurry Thickener Underflow | Wash Water to PAL Discharge BF | Sulphur Removal Filtrate to BZS |
| **Flow** | **t/h** | 34.5 | 18.6 | 53.1 | 7.7 | 4.4 | 216.9 | 1.4 | 1.1 | 223.1 | 0.3 | 46.8 | 24.5 | 23.3 |
| **Flow** | **m3/h** | 9.1 | 19.0 | 28.0 | 5914.1 | 3379.6 | 175.1 | 0.8 | 0.6 | 165.8 | 0.2 | 29.6 | 24.6 | 19.3 |
| **Temperature** | **C** | 25.0 | 66.2 | 58.4 | 25.0 | 25.0 | 40.0 | 25.0 | 25.0 | 79.6 | 25.0 | 71.1 | 25.0 | 41.6 |
| **Pressure** | **kPa** | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 87.7 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 87.7 | 101.3 |
| **Density** | **kg/m3** | 3777.0 | 979.8 | 1894.7 | 1.3 | 1.3 | 1239.1 | 1887.9 | 1887.9 | 1345.8 | 1887.9 | 1581.8 | 997.0 | 1208.4 |
| **Solids** | **wt %** | 100.0 | 0.0 | 65.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 35.0 | 0.0 | 1.0 |
| **Solids** | **t/h** | 34.5 | 0.0 | 34.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 16.4 | 0.0 | 0.2 |
| **Aq. Liquids** | **t/h** | 0.0 | 18.6 | 18.6 | 0.0 | 0.0 | 216.9 | 1.4 | 1.1 | 223.1 | 0.3 | 30.5 | 24.5 | 23.0 |
| **Gases** | **t/h** | 0.0 | 0.0 | 0.0 | 7.7 | 4.4 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Org. Liquids** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Zn** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 55.4 | 0.0 | 0.0 | 155.8 | 0.0 | 129.2 | 0.0 | 76.5 |
| **Aq. Liq.: aH2SO4** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 156.5 | 1850.1 | 1850.1 | 8.2 | 1850.1 | 29.1 | 0.0 | 17.3 |
| **Aq. Liquids: Fe** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 2.1 | 0.0 | 13.2 | 0.0 | 7.8 |
| **Solids: Zn** | **wt %** | 51.4 | 0.0 | 51.4 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 23.8 | 0.0 | 0.9 | 0.0 | 0.9 |
| **Solids: Fe** | **wt %** | 4.5 | 0.0 | 4.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 10.6 | 0.0 | 6.5 | 0.0 | 6.5 |
| **Solids: Cu** | **wt %** | 0.5 | 0.0 | 0.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.3 | 0.0 | 0.1 | 0.0 | 0.1 |
| **Solids: Cd** | **wt %** | 0.1 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **wt %** | 28.7 | 0.0 | 28.7 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 39.6 | 0.0 | 58.0 | 0.0 | 58.0 |
| **Solids: Zn** | **t/h** | 17.7 | 0.0 | 17.7 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 |
| **Solids: Fe** | **t/h** | 1.6 | 0.0 | 1.6 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1.1 | 0.0 | 0.0 |
| **Solids: Cu** | **t/h** | 0.2 | 0.0 | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cd** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **t/h** | 9.9 | 0.0 | 9.9 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 9.5 | 0.0 | 0.1 |
| **Aq. Liquids: Zn** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 9.7 | 0.0 | 0.0 | 25.8 | 0.0 | 2.9 | 0.0 | 1.5 |
| **Aq. Liq: Fe** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.3 | 0.0 | 0.3 | 0.0 | 0.2 |
| **Aq. Liq.: aH2SO4** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 27.4 | 1.4 | 1.1 | 1.4 | 0.3 | 0.7 | 0.0 | 0.3 |
| **Aq. Liq: SO4[2-]** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: SO4[2-]** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Mn** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 8.3 | 0.0 | 0.0 | 10.3 | 0.0 | 9.9 | 0.0 | 5.9 |
| **Aq. Liquids: Cu** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.9 | 0.0 | 0.8 | 0.0 | 0.5 |
| **Aq. Liquids: Cd** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0.2 | 0.0 | 0.1 |
| **Aq. Liquids: Al** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0.3 | 0.0 | 0.2 |
| **Enthalpy** | **kJ/kg** | -3831.4 | -15704.6 | -7987.0 | 0.0 | 0.0 | -13783.8 | -8486.6 | -8486.6 | -12791.4 | -8486.6 | -10234.6 | -15877.1 | -13849.4 |

| Stream No. | | 316 | 333 | 334 | 339 | 347 | 348 | 401 | 402 | 446 | 457 | 469 | 481 | 489 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| **Stream Name** | | Sulphur Removal Filtrate to HAL | Sulphur Flotation Tails | Sulphur Tails Thickener Overflow | Flotation Concentration Filtrate | Sulphur Pressure Filter Cake | Elemental Sulphur | Fe Removal Thickener O/F | Zinc dust to Solution Purification | Hot Purification Cake | Cu Cake | Cd Cake | Limestone to Bleed Neutr. | Wash Water to Bleed BF |
| **Flow** | **t/h** | 23.3 | 21.2 | 16.0 | 2.6 | 0.3 | 7.4 | 256.6 | 0.4 | 2.6 | 0.4 | 0.1 | 65.9 | 19.4 |
| **Flow** | **m3/h** | 19.3 | 16.2 | 16.3 | 2.6 | 0.1 | 3.6 | 194.7 | 0.1 | 1.6 | 0.2 | 0.0 | 54.8 | 19.4 |
| **Temperature** | **C** | 41.6 | 30.3 | 30.3 | 30.5 | 140.0 | 140.0 | 77.9 | 25.0 | 77.1 | 68.8 | 70.7 | 63.1 | 25.0 |
| **Pressure** | **kPa** | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 87.7 | 101.3 |
| **Density** | **kg/m3** | 1208.4 | 1313.8 | 979.5 | 979.4 | 3186.0 | 2070.0 | 1318.2 | 7140.0 | 1699.4 | 2243.6 | 2275.6 | 1202.6 | 997.0 |
| **Solids** | **wt %** | 1.0 | 40.0 | 0.0 | 0.0 | 98.0 | 0.0 | 0.0 | 100.0 | 70.0 | 70.0 | 70.0 | 29.0 | 0.0 |
| **Solids** | **t/h** | 0.2 | 8.5 | 0.0 | 0.0 | 0.3 | 0.0 | 0.0 | 0.4 | 1.8 | 0.3 | 0.1 | 19.1 | 0.0 |
| **Aq. Liquids** | **t/h** | 23.0 | 12.7 | 16.0 | 2.6 | 0.0 | 7.4 | 256.6 | 0.0 | 0.8 | 0.1 | 0.0 | 46.8 | 19.4 |
| **Gases** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Org. Liquids** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Zn** | **g/L** | 76.5 | 0.1 | 0.1 | 0.1 | 0.0 | 0.0 | 142.2 | 0.0 | 134.2 | 84.6 | 90.5 | 0.0 | 0.0 |
| **Aq. Liq.: aH2SO4** | **g/L** | 17.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Fe** | **g/L** | 7.8 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Zn** | **wt %** | 0.9 | 0.3 | 0.3 | 0.0 | 33.8 | 0.0 | 0.0 | 100.0 | 22.7 | 16.2 | 8.7 | 0.0 | 0.0 |
| **Solids: Fe** | **wt %** | 6.5 | 12.0 | 12.0 | 0.0 | 7.4 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cu** | **wt %** | 0.1 | 0.0 | 0.0 | 0.0 | 2.9 | 0.0 | 0.0 | 0.0 | 0.0 | 52.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cd** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 1.1 | 61.7 | 0.0 | 0.0 |
| **Solids: S** | **wt %** | 58.0 | 20.6 | 20.6 | 0.0 | 41.5 | 0.0 | 0.0 | 0.0 | 11.2 | 4.4 | 4.3 | 0.0 | 0.0 |
| **Solids: Zn** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 | 0.4 | 0.4 | 0.1 | 0.0 | 0.0 | 0.0 |
| **Solids: Fe** | **t/h** | 0.0 | 1.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cu** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0.0 | 0.0 |
| **Solids: Cd** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **t/h** | 0.1 | 1.8 | 0.0 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Zn** | **t/h** | 1.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 27.7 | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq: Fe** | **t/h** | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq.: aH2SO4** | **t/h** | 0.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq: SO4[2-]** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: SO4[2-]** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Mn** | **g/L** | 5.9 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 8.5 | 0.0 | 8.0 | 0.5 | 0.5 | 0.0 | 0.0 |
| **Aq. Liquids: Cu** | **g/L** | 0.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.8 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Cd** | **g/L** | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0.0 | 9.6 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Al** | **g/L** | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Enthalpy** | **kJ/kg** | -13849.4 | -13616.7 | -15852.2 | -15851.4 | -3152.1 | 157.9 | | | | -7118.9 | -7040.7 | -14871.4 | -15877.1 |

| Stream No. | | 490 | 492 | 502 | 506 | 508 | 510 | 511 | 513 | 516 | 517 | 522 | 524 | 527 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | Neutralised Raff. to Depletion | Bleed BF Cake to BZS | Cooled Purified PLS | Raffinate to Bleed | Loaded Extraction Organic | Stripped Organic toExtraction | EW Bleed & Wash Water | Stripped Organic to Depletion | Loaded Depletion Organic | Depletion Raffinate to BZS | Demin Water to SX | Total Loaded Organic | Stripped Organic |
| Flow | t/h | 145.2 | 30.3 | 256.1 | 29.6 | 988.7 | 972.7 | 82.3 | 95.1 | 96.7 | 143.7 | 60.2 | 1085.4 | 1067.8 |
| Flow | m3/h | 142.1 | 15.7 | 190.0 | 23.9 | 1234.6 | 1215.8 | 78.9 | 118.9 | 120.7 | 142.6 | 60.4 | 1355.3 | 1334.6 |
| Temperature | C | 48.5 | 48.3 | 29.9 | 40.0 | 34.5 | 36.1 | 34.6 | 36.1 | 45.1 | 45.1 | 25.0 | 35.5 | 36.1 |
| Pressure | kPa | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 | 87.7 |
| Density | kg/m3 | 1022.2 | 1928.4 | 1348.0 | 1239.1 | 800.8 | 800.1 | 1042.9 | 800.1 | 800.8 | 1007.7 | 997.0 | 800.8 | 800.1 |
| Solids | wt % | 0.0 | 80.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids | t/h | 0.0 | 24.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids | t/h | 145.2 | 6.1 | 256.1 | 29.6 | 0.0 | 0.0 | 82.3 | 0.0 | 0.0 | 143.7 | 60.2 | 0.0 | 0.0 |
| Gases | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Org. Liquids | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 988.7 | 972.7 | 0.0 | 95.1 | 96.7 | 0.0 | 0.0 | 1085.4 | 1067.8 |
| Aq. Liquids: Zn | g/L | 14.2 | 14.1 | 145.1 | 55.4 | 0.0 | 0.0 | 9.9 | 0.0 | 0.0 | 2.8 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | g/L | 0.1 | 0.1 | 0.0 | 156.5 | 0.0 | 0.0 | 41.8 | 0.0 | 0.0 | 17.1 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Fe | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Fe | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Cu | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Cd | wt % | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | wt % | 0.0 | 9.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Zn | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Cu | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: Cd | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Solids: S | t/h | 0.0 | 2.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Zn | t/h | 2.0 | 0.1 | 27.6 | 1.3 | 0.0 | 0.0 | 0.8 | 0.0 | 0.0 | 0.4 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: Fe | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liq.: aH2SO4 | t/h | 0.0 | 0.0 | 0.0 | 3.7 | 0.0 | 0.0 | 3.3 | 0.0 | 0.0 | 2.4 | 0.0 | 0.0 | 0.0 |
| Aq. Liq: SO4[2-] | t/h | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: SO4[2-] | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Mn | g/L | 1.3 | 1.3 | 8.7 | 8.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1.3 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Cu | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Cd | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Aq. Liquids: Al | g/L | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Enthalpy | kJ/kg | -15433.5 | -13077.6 | -13245.6 | -13783.8 | -2798.2 | -2833.0 | -15348.1 | -2833.0 | -2775.8 | -15584.3 | -15877.1 | -2796.2 | -2833.0 |

| Stream No. | | 531 | 545 | 601 | 603 | 604 | 605 | 606 | 607 | 608 | 611 |
|---|---|---|---|---|---|---|---|---|---|---|---|
| **Stream Name** | | Spent Electrolyte to Scrub | Advance Electrolyte to EW | Total Process Water | Total Oxygen | Total Steam | Total Sulphuric Acid | Total Flotation Reagents | Total Milk of Lime | Total Flocculant | Total GSW |
| **Flow** | **t/h** | 22.1 | 322.2 | 126.3 | 7.7 | 14.6 | 16.2 | 0.1 | 4.2 | 12.3 | 19.9 |
| **Flow** | **m3/h** | 18.3 | 244.9 | 126.7 | 5933.0 | 2831.6 | 8.6 | 0.1 | 3.7 | 12.3 | 20.0 |
| **Temperature** | **C** | 38.8 | 36.1 | 25.0 | 25.0 | 179.9 | 25.0 | 25.0 | 25.0 | 25.0 | 25.0 |
| **Pressure** | **kPa** | 101.3 | 101.3 | 101.3 | 101.3 | 1000.0 | 101.3 | 101.3 | 101.3 | 101.3 | 101.3 |
| **Density** | **kg/m3** | 1205.0 | 1315.3 | 997.0 | 1.3 | 5.2 | 1887.9 | 997.0 | 1157.6 | 997.0 | 997.0 |
| **Solids** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 25.0 | 0.0 | 0.0 |
| **Solids** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 1.1 | 0.0 | 0.0 |
| **Aq. Liquids** | **t/h** | 22.1 | 322.2 | 126.3 | 0.0 | 0.0 | 16.2 | 0.1 | 3.2 | 12.3 | 19.9 |
| **Gases** | **t/h** | 0.0 | 0.0 | 0.0 | 7.7 | 14.6 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Org. Liquids** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Zn** | **g/L** | 42.7 | 118.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq.: aH2SO4** | **g/L** | 180.0 | 74.5 | 0.0 | 0.0 | 0.0 | 1850.1 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Fe** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Zn** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Fe** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cu** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cd** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **wt %** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Zn** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Fe** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cu** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: Cd** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Solids: S** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Zn** | **t/h** | 0.8 | 29.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq: Fe** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq.: aH2SO4** | **t/h** | 3.3 | 18.3 | 0.0 | 0.0 | 0.0 | 15.9 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liq: SO4[2-]** | **t/h** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: SO4[2-]** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Mn** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Cu** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Cd** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Aq. Liquids: Al** | **g/L** | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| **Enthalpy** | **kJ/kg** | -13993.8 | -13395.6 | -15877.1 | 0.0 | -13200.9 | -8486.6 | -15877.1 | -15234.3 | -15877.1 | -15877.1 |