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Project Report

**M6036 Ironbark Zinc Limited
Desktop Study Report
Citronen Zinc Project - Greenland**

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1 EXECUTIVE SUMMARY

1.1 Introduction

Bateman Engineering Pty Ltd (Bateman) has completed a Desktop Study (DS) for Ironbark Zinc Limited (Ironbark) to evaluate the Capital and Operating Costs for various zinc refining flowsheets to treat the ore sources that make up the Citronen Zinc Project.

The capital and operating costs for each of the four flowsheets evaluated in the study are shown in both to a level of +/-40% accuracy.

Table 1-1 Summary Capital and Operating Cost Estimates

	Imperial Smelting Process	Roast Leach Electrowinning	Pressure Leaching	Atmospheric Leaching
Capital Cost A\$'000,000	861	508.5	435	491
Operating Cost A\$/lb Zn produced	0.748	0.464	0.485	0.462

2 INTRODUCTION

2.1 Introduction and Project Background

Ironbark Zinc Limited (Ironbark) is the owner of the Citronen Zinc project located in Greenland. Ironbark are currently undertaking a Bankable Feasibility Study (BFS) for the development of a Zinc mine and processing facility at the Citronen Project site. The processing facility is based on a conventional open pit and underground mining operation and traditional crush, grind, flotation concentrator to produce zinc concentrate for shipment to third party smelters for refining.

As part of the Greenland government project approvals process, planned mining projects need to demonstrate that they have evaluated extracting the maximum value for the commodity being mined and processed in Greenland. For the Citronen Zinc Project this would be the production of zinc metal through a refinery process.

Bateman Engineering Pty Ltd (Bateman) has been engaged by Ironbark to assess the various zinc refining options available for the Citronen project and their subsequent capital and operating costs at a desktop study (DS) level.

2.2 Project Description

The current BFS being completed by Wardrop, Metso and MT Hojgaard is based on treating two ore zones, Beach and Discovery, at rates of 3.0 Mtpa to produce separate zinc and lead concentrates for off-site refining. A project life of 16 years is envisaged based on current known ore reserves. A preliminary concentrate production schedule on a year by year basis was provided by Ironbark from the BFS shown in Table 2-1. Based on this concentrate production rate the throughput and production rate for the refinery was selected by Bateman and Ironbark for the study as shown in Table 2-2.

Table 2-1 Concentrate Production Schedule

	Year	Zn Concentrate Production (dry)	Pb Concentrate Production (dry)	Contained Zinc	Contained Pb
2014	1	218,301	28,161	119,880.25	11,969
2015	2	251,185	31,000	137,938.13	13,175
2016	3	263,800	26,791	144,865.76	11,386
2017	4	264,429	30,166	145,211.22	12,821
2018	5	263,291	29,842	144,586.40	12,683
2019	6	253,809	26,791	139,379.43	11,386
2020	7	231,844	29,214	127,317.27	12,416
2021	8	227,675	29,518	125,027.95	12,545
2022	9	227,359	29,408	124,853.98	12,499
2023	10	221,529	28,887	121,652.51	12,277
2024	11	221,850	28,761	121,829.07	12,223
2025	12	223,289	30,432	122,619.26	12,934
2026	13	222,533	29,821	122,204.14	12,674
2027	14	220,278	29,651	120,965.63	12,602
2028	15	215,532	27,137	118,359.48	11,533
2029	16	165,543	14,340	90,908.20	6,094

Table 2-2 Refinery Production Schedule

Concentrate	Dry Concentrate tpa	% Zn	Contained Zinc tpa	% Pb	Contained Lead tpa	% Fe	% SiO ₂
Zinc	220,000	54.9	120,780	1.75	3,850	6.3	0.79
Lead	29,000	4.0	1,160	42.5	12,325	17.3	2.1

2.3 Site Location

The Citronen property is located in the Northern Region of Greenland on the coast of the Greenland Sea as shown in Figure 2-1. The property covers an area of approximately 36 km² and contains two main deposits 1 km apart, referred to as Discovery Zone and Beach Zone. Ironbark also retain 100% ownership of exploration licenses over 2,500 km² of prospective area in Greenland. The mineralisation is considered to be of a Sedex style zinc deposit starting at the surface, shallow, flat lying and adjacent to a deep water fjord. The lead zinc ore mineralisation of sphalerite and galena is mainly found in a primary zone within the Paleozoic Franklandian Basin rock types found in northern Greenland.



Figure 2-1 Site Location

2.4 Battery Limits

The battery limits relating to the scope of works undertaken by Bateman are detailed in Table 2-3.

Table 2-3 Battery Limits

Main Inputs	Stream	Limit
	Zinc Concentrate	Discharge from Zinc Concentrate thickener
	Raw Water	Outlet flange of a supply main
	Electric power	Medium voltage at a single point (to be confirmed)
	Reagents and consumables	Delivered by others to on-site storage
	Compressed air	Single point on compressed air main
	Instrument air	Single point on instrument air main
Main Outputs	Stream	Limit
	Treated Tailings	Tailings suitable for discharge to conventional tailings storage facility, at tails thickener underflow pump discharge.
	Gaseous discharges	Noxious gasses captured & treated
	Liquid effluents	Treated and suitable for disposal to conventional tailings storage facility, at liquid waste pump discharge
	Zinc Metal	Bundled zinc ingots and/or cathodes

2.5 Scope of Services

Bateman's scope of work included:

- Review possible zinc refining flowsheets
- List pros and cons for each refining flowsheet
- Develop order of magnitude Capital and Operating Costs (+/- 40%) for each refinery flowsheet option

Exclusions from the Bateman scope of services:

- Detailed design
- Preparation of detailed PFDs, P&IDs, GAs, Mass balances, Equipment lists
- Tailings disposal
- Development of financial model

2.6 Study Execution Methodology

2.6.1 Process Design

Ironbark provided Bateman with the concentrate production schedule for which it sought capital and operating costs for processing the zinc ore through a refining process.

The following process engineering activities were completed by Bateman:

- Review of available mineralogy, flotation metallurgical testwork and previous feasibility study reports;
- Evaluation of possible zinc refining flowsheets capable of treating the zinc concentrate planned to be produced at the Citronen Project;

Preliminary Process Flow Diagrams (PFDs) and Piping and Instrument Diagrams (P&IDs) were not developed as part of this study.

2.6.2 Engineering Design

No other engineering design was completed as part of this study.

2.6.3 Cost Estimate

The methodology employed in generating the capital cost estimate, including direct and indirect costs for the project, was based on discussions between zinc refining technology suppliers and Bateman, including Sherritt Technologies and Outotec, previous costs estimates from the Bateman cost database for similar projects and published information. A factored cost estimate was developed for each major circuit within the overall flowsheet. The factored estimates were calculated as a percentage of the estimated direct mechanical equipment cost, with the percentages applied based on prior experience for similar projects.

Operating costs were based on the consumption rates calculated for all relevant consumables and the unit transport and purchase prices obtained from the Bateman database for similar projects. Steady state power consumption and cost were calculated based on an estimated total power draw and local unit power cost. On this basis, preliminary capital and operating costs were generated within a +/-40% accuracy level.

2.6.4 Process Design Basis

A review of possible refining processes to convert the zinc concentrate at Citronen resulted in four (4) processes being selected for analysis. These four processes were:

- Imperial Smelting Process
- Roast Leach Electrowinning
- Pressure Leaching based on a two stage counter current leach
- Atmospheric Leaching based on Outotec Direct leaching technology

Each of the selected processes handles the treatment of lead (Pb) via different means. For all the cases investigated an intermediate product would be produced and would still require further processing to convert it into metal. Treating the lead concentrate through a dedicated refinery has been excluded from this study.

The physical parameters shown in Table 2-4 were used for sizing each of the flowsheets which ultimately were used as a basis to estimate the capital and operating costs.

Table 2-4 Design Basis

	Units	Imperial Smelting Process	Roast Leach Electrowinning	Pressure Leaching	Atmospheric Leaching
Concentrate Production	tpa	223,090	223,090	223,090	223,090
Zinc Concentrate Grade	% Zn	55%	55%	55%	55%
	%Pb	1.75%	1.75%	1.75%	1.75%
Zinc Metal Production	t/a	120,000	120,000	120,000	120,000
	Mlb Zn /a	265	265	265	265
Sulphuric Acid	tpa	175,000	185,000	n/a	0
Sulphur Production	tpa	n/a	n/a	non recoverable	50,000
Estimated Annual Power	MWh/a	451,765	532,530	514,000	468,000
Coal Consumption	tpa Carbon	112,000	n/a	n/a	n/a
Lime consumption	tpa	n/a	n/a	14,000	16,000
Estimated Total Manning requirements		285	285	350	330

Note: n/a – not applicable

3 REFINING PLANT OPTIONS

3.1 Selection Process

Bateman maintains an up to date database of the technical and commercial processes and technologies suitable for converting zinc concentrate into zinc metal. Bateman have used this knowledge and database as the basis of a preliminary evaluation to select processes that are suited to the Citronen concentrates to:

- produce zinc metal;
- recover lead as a by-product; and
- produce precious metal residues for further treatment off-site.

The approach to use commercialised processes reduces the technical risk and ensures that the delivery can be achieved within the timeframe set by Ironbark, without the need to invest time and money in proving new technology (unproven concepts).

Zinc smelting is the process of recovering and refining zinc metal out of zinc-containing feed material (sulphides or concentrates and oxides). Globally, two main zinc-smelting processes are in use, namely:

- Pyrometallurgical processes, such as the Imperial Smelting Process; and
- Hydrometallurgical or electrolytic processes, such as the Roast-Leach-Electrowinning and direct leaching process.

Since the development of the Roast-Leach-Electrowin (RLE) process in the early 1900's, the hydrometallurgical process route replaced the pyrometallurgical process as the preferred process. In 2008 around 86% of the world production utilised the electrolytic process. [Ref: *Ruonala M. et.al., Latest developments in Zinc Processing, PbZn 2010*]. The shift can largely be attributed to the improved energy-efficiency, higher recovery, ease of automation and hence higher productivity of the electrolytic process.

Two basic variants of electrolytic plants used for the recovery of zinc, are distinguished by the following characteristics:

- Oxidation of the sulphide minerals, sphalerite, by roasting and then leaching with sulphuric acid to produce a zinc sulphate solution, or
- Direct production of zinc sulphate solution by pressure or atmospheric leaching.

Some refining operations incorporate both operations on site, but the combination process have been excluded from this study for simplicity.

In the electrolytic process, aggressive leaching conditions are required to achieve high zinc recoveries and result in the leaching of iron. Iron is removed from the zinc bearing solution by precipitation. Typically ponded residues in the form of jarosite, goethite or hematite have been produced, but ever increasing environmental pressure is moving technology towards the production of an inert residue. A pyrometallurgical process utilising the Ausmelt Top Submerged Lancing (TSL) Technology or Waelz Kiln to produce a solid clinker as the residue is an

alternative method available for iron disposal. Such a method has been successfully incorporated into the traditional Roast Leach Electrowinning process. At Korea Zinc, the Neutral Leach residue is treated with TSL technology [Ref: Hughes S et.al., *Ausmelt technology for lead and zinc processing, International Symposium on Lead and Zinc Processing, SAIMM, 2008*].

Since the mid 1970s hydrometallurgy processes using sulphate and chloride as lixiviant, were developed as discussed in Sections 3.1.1 and 3.1.2.

3.1.1 Sulphate Leaching Processes/Technologies

The following processes/technologies are proven commercially for the production of zinc metal from concentrate:

- Agitated Tank Leaching;
- Biological leaching (tank and/or heap);
- Conventional Heap Leaching;
- Geocoat®/Geoleach®;
- Direct Leaching (Sherritt Pressure Leaching); and
- The Outotec Direct Leaching Process.

Bateman have considered each of these processes and selected the Direct Leaching (Sherritt Pressure Leaching and Outotec Direct Leach processes to evaluate as part of this study as these process have a proven history over 30 years at industrial scale and/or been installed at numerous operations throughout the world. The remaining processes were discounted as they were deemed not suitable for the ore types or climatic conditions at Citronen or only been proven at one or two operations at industrial scale as is the case with the Geocoat® process.

The following sulphate processes were not considered as suitable for treating the Citronen deposits as they have not been proven at industrial scale for Zinc recovery:

- Activox®
- Albion Leach

3.1.2 Chloride Leaching Processes/Technologies

The following processes/technologies have all been proposed for the production of zinc metal from concentrate:

- Intec Process
- Agitated Tank Leaching (ferric chloride)

Although both processes were tested on pilot plant scale, these chloride leaching processes have not been commercially proven. Based on the technical risk associated of using unproven technologies Bateman didn't consider chloride leaching technologies as being suitable to treat the Citronen deposits for this study.

4 FLOWSHEET DESCRIPTIONS

Four flowsheets were selected for evaluation during the study:

- Imperial Smelting Process (ISP)
- Roast Leach Electrowinning (Calcine Leaching)
- Direct Leaching at elevated pressure (Sherritt Pressure Leaching)
- Direct Leaching at atmospheric pressure (Outotec Technology)

A brief description of each flowsheet, the major unit operations and characteristic are provided below. Commentary regarding the commercialisation plus lists of the process major advantages and disadvantages are also included.

4.1 Imperial Smelting Process (ISP)

The Imperial Smelting Process Pty Company began development of a new smelting process for zinc in 1943 at Avonmouth in the UK. In 1947, an experimental Blast Furnace was built and trialled before the first two commercial furnaces were constructed in 1950 and 1952 respectively.

Currently about 8 -10% of the world's primary zinc is produced with ISP technology. The process co-produces lead bullion in addition to zinc metal using a mix of lead and zinc concentrates or complex lead-zinc concentrates as raw material. The ISP is similar to blast furnace processes used for steel production, except that it is operated with a hot top which prevents re-oxidation of zinc vapours. The process consists of two operations, namely:

- sintering; and
- blast furnace smelting of sintered lumps to extract lead and zinc simultaneously, before being refined into ingots.

The ISP is based on the reduction of lead and zinc into metal with carbon in a specially designed Imperial Smelting furnace. Pre-heated air is blown from below into the shaft furnace. The sinter is charged with pre-heated coke at the top of the furnace. Temperatures range from 1000°C at the top to 1500°C or more in the centre of the furnace. The coke is converted into carbon monoxide which provides the means to reduce lead and zinc oxides to metallic lead and zinc. Lead, which is below boiling point, flows from the bottom of the blast furnace carrying copper, silver and gold with it. Zinc evaporates and passes out of the furnace near the top along with other gases. To avoid re-oxidation, the zinc vapour is rapidly quenched and dissolved in a spray of molten lead in a condenser (lead splash condenser). By cooling the lead, crude zinc is released and separated. The lead is recycled to the condensing process.

The ISP is an energy-intensive process and thus became very expensive following the rise of energy prices in recent years. This and the lower production of bulk concentrates containing significant amounts of lead led to abandoning of the ISP. Today, Imperial Smelting furnaces are only in operation in China, India, Japan and Poland at 9 operations.

The closures during the 1970's and 90's were associated with the lack of raw materials whilst the closures of Imperial Smelting Furnaces (ISF) that occurred during 2003 and 2005 (6 furnaces) were associated with economic reasons (low zinc price, high coke prices, depreciation of US dollar and low treatment charges). The remaining ISF's survived in India and China mainly due to the presence of local indigenous concentrates, low cost of labour and growing domestic demand, whilst those in Japan have been able to survive through the minimisation of costs through use of cheaper secondary materials and improvements in technical efficiencies (plant utilisation).

The major difference between hydrometallurgical process and the Imperial Smelting process is that the first produce very pure zinc directly whereas the latter produces lower grades of zinc that still contains significant impurities that have to be removed by thermal refining in the refining stage.

The ISP consists of four main plant areas

- Acid Plant
- Sinter Plant
- Imperial Smelting Furnace
- Zinc Refinery

Minor metals are treated differently from site to site but the typical processes for the major impurities are:

- Cadmium: Pure Cd is produced in the Zinc refinery in the final stage of fractional distillation of ISF Zinc.
- Copper: Copper dross from the bullion floor is subjected to an ammonia leach followed by solvent extraction (SX) treatment to produce copper sulphate and finally, copper cathode via electrowinning or copper sulphate via crystallisation.
- Selenium: Selenium is removed from Hot Gas Precipitator (HGP) dust with a hot acid leach.
- Slag: ISF slag typically contains 6 to 8% Zn which is granulated and either stockpiled on site or sold. In recent years, a number of ISF's have installed a slag fuming plant after the forehearth to fume the zinc from the slag, which is then granulated. The low level of impurities and hence stability, means that slag can be used as a substitute for sand in concrete manufacturing.
- Note that the treatment of minor metals was not included in the capital estimate.

4.1.1 Process Description

A block flow diagram of the major unit operations for the ISP process is provided in Figure 4-1 with a description for the sintering and ISF unit operations outlined below.

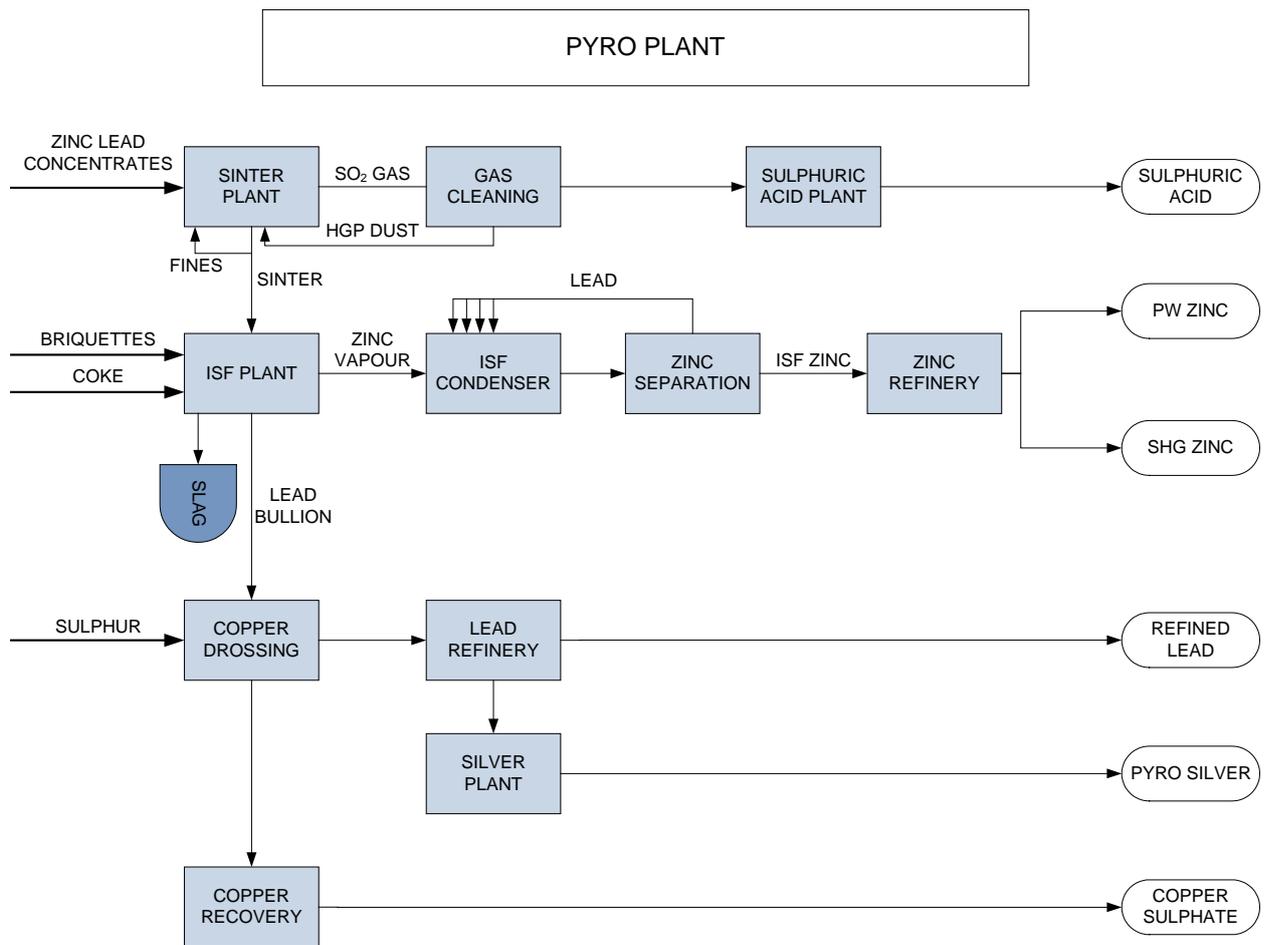


Figure 4-1 Imperial Smelting Process Blockflow Diagram

The Imperial Smelting Process (ISP), process starts with sintering, where bulk concentrates (a mixture of lead and zinc concentrates) are blended with oxidic secondaries and fluxes. The mixture is passed through the sinter plant to remove sulphur. Gas generated from the sintering process is cleaned before sulphuric acid production.

The de-sulphurised output of the sinter plant is fed into an Imperial Smelting Furnace (ISF). The ISF is designed to simultaneously produce molten lead and zinc by smelting agglomerated Sinter with preheated coke and preheated blast air. The chemical reaction between air and the coke produces carbon monoxide and generates heat to smelt the metallic oxide in the charge into the elemental metal. Molten lead is tapped from the bottom of the furnace and fed to a lead refinery. At the temperature of operation, zinc is vaporised. Zinc vapour passes through the condenser where it gets absorbed by lead. The lead is cooled to separate zinc and the molten zinc is passed through a double distillation column for further refining. Through this process, zinc is produced as a major product with cadmium as a by-product.

Waste gas leaving the condenser after zinc is condensed in lead pass through a gas cleaning system where it is cooled and particulate matter removed. The gas contains carbon monoxide of low calorific value. After cleaning, the gas is utilized in preheating the furnace blast air and in preheating the coke. Any remaining excess gases are passed through a boiler for steam generation which can be utilized in plant operation.

4.1.1.1 Sintering Process

Zinc and lead concentrates are produced as small particles which are unsuitable for direct treatment in a blast furnace. The goals of sintering are to:

- Remove sulphur from the concentrate;
- Produce SO₂ gas of consistent concentration for the acid plant;
- Produce quality sinter for smelting in the ISF; and
- Produce sufficient sinter to match furnace throughput.

The quality of the sinter product is critical for the ISF operation. An example of a typical sinter chemical quality specification is provided in Table 4-1.

Table 4-1 Typical Sintering Chemical Quality

	Typical Range	Reject Limit	Reason
Pb	18 -21%	23.5%	Causes sinter to soften
SiO ₂	4.3 to 4.8%	5.1%	Causes sinter to soften
Pb + SiO ₂		26.5%	Causes sinter to soften
CaO	1.0 – 1.2%	<0.95, >1.4	Improves softening temperature, and zinc recovery
Fe	14.3 – 14.8%	15.1%	Gives poor sinter structure
Cu	0.8 – 1.2%	3 consecutive assays ≥ 1.4%	Causes accretion in the ladles
S	<0.6%	1.1	Causes accretion in the ISF shaft and is an indicator that sinter has not sintered properly

4.1.1.2 Imperial Smelting Furnace Operation

The sintered product is charged into the ISF with the coke (A typical coke specification for the ISF in Table 4-2) either by a layering feeding consecutively. The charge enters the furnace via a double charge bell arrangement as the furnace needs to be sealed at all times to prevent gas and fume escape. Air is injected into the bells and the top of the furnace to maintain the charge are >1050°C and avoid accretion formation. Following each charge addition the furnace level is measured using dip rods to maintain a level of between 5.7m and 5.9m.

Table 4-2 Typical Coke Specification for ISF

Parameter	Target	Reason
Moisture	<5%	Reduce transport costs and the thermal load on the coke pre-heater
Ash Content	<12%	Minimise slag production (and zinc losses)
Volatiles	<1.5%	Poor quality coke, black stack at coke pre-heater
Fixed Carbon	>86.5%	Fuel content
Reactivity	<22%	Low reactivity to minimise loss in shaft – coke should burn in the hearth area
Strength Micron 40 Micron 10	>84% <5%	Does not break up – coke fines will be blown over into the condenser to form dross
Sizing 40 to 100mm 25 to 40mm -25mm -6mm	>78% <20% <2% Nil	Cannot be too small as it will cause a loss in shaft permeability

PbO is reduced at the top of the shaft and Pb flows to the bottom, whilst the coke burns at the bottom to produce carbon monoxide (CO) and heat. The CO reduced the ZnO formed in the sintering process into Zn vapour. Fuel to the furnace is controlled to maintain a C/Zn ratio typically between 0.65 and 0.74(w/w), but is adjusted based on the Zn reporting to the slag.

The majority of ISF operations practice batch tapping where they periodically open a taphole in a water-cooled copper block. As the liquid level increases in the furnace, the bustle main pressure increases. When the slag reaches the tuyers, the taphole is opened. Lead and slag are tapped into a forehearth, which is used as a settling tank to allow the lead and slag to separate. Slag in the overflow is granulated using high pressure water. The lead bullion is tapped into ladles and sent to the bullion floor for de-copperisation. The bullion contains approximately 95% Pb and minor metals Cu, Ag, Au.

Typically the zinc refinery will produce 2 grades of zinc in variety of shapes referred to as Prime Western Grade Zinc (PWG) which is used mainly for galvanizing and High Grade (HG) used for alloying, paint and medicinal purposes. Typical specifications of such products are shown in Table 4-3.

Table 4-3 Typical Zinc Product Specification from a ISP plant

Product	Zn %	Pb%	Cd ppm	Cu ppm	Fe ppm	Sn ppm	Al ppm
PWG Zinc	98.5	1.35	250	100	250		50
HG	99.95	0.01	75	20	100	10	50

4.2 Roast Leach Electrowinning Process

The electrolytic zinc smelting process can be divided into a number of generic sequential process steps.

1. Receipt of feed material (concentrates and secondary feed materials such as zinc oxides) and storage;
2. Roasting to oxidise zinc sulphide to sulphur dioxide and zinc oxide;
3. Acid production;
4. Leaching of zinc oxide in spent electrolyte;
5. Residue treatment;
6. Purification: removing impurities that could affect the quality of the electrolysis process (such as cadmium, copper, cobalt or nickel) from the leach solution;
7. Electrolysis or electrowinning to produce zinc cathodes; and
8. Melting and casting, typically using electrical induction furnaces and casting the molten zinc into ingots.

Additional steps can be added to the process transforming the pure zinc, typically 99.995% pure zinc known as Special High Grade (SHG), into various types of alloys or other marketable products.

4.2.1 Process Description

A generic RLE flowsheet is shown in Figure 4-2 and detailed below.

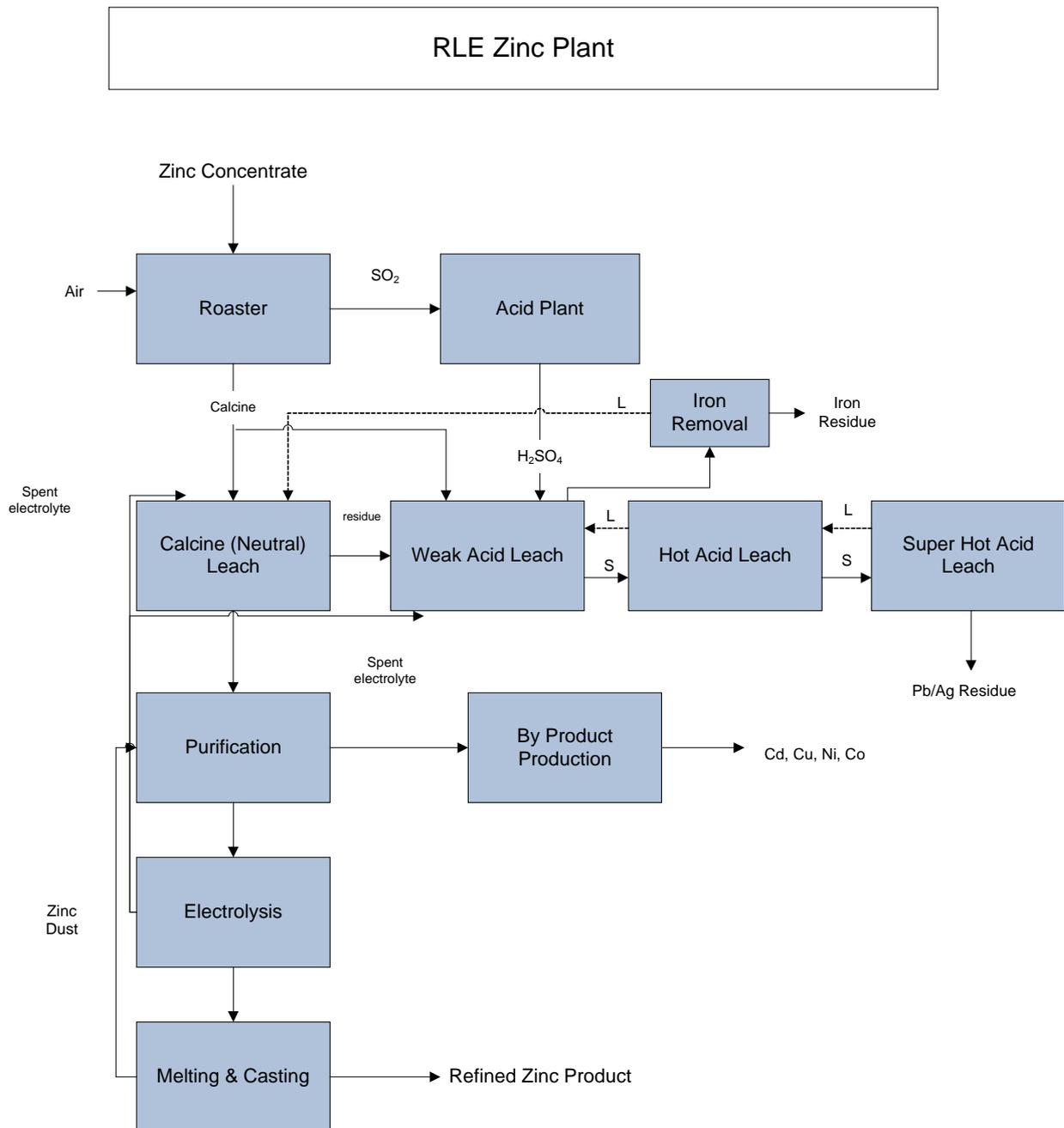
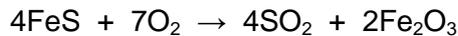
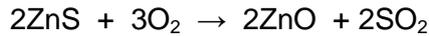


Figure 4-2 Generic Roast Leach Electrowinning Blockflow Diagram

Smelters use a mix of zinc-containing concentrates or secondary zinc material such as zinc oxides or scrap as feed to their roasting plant. Most zinc smelters use several sources of concentrates. These different materials are blended to obtain an optimal mix of feedstock for the roasting process.

4.2.1.1 Roasting and Acid Production

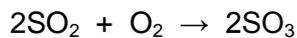
Through the roasting process, the zinc sulphides in the concentrates are converted into zinc oxide, known as calcine. A roaster operates at a temperature of approximately 950° C, generating enough energy to make the process autogenous. The chemical reactions taking place during the process are:



Approximately 90% of zinc in concentrates is oxidized to zinc oxide. At roasting temperatures of around 950°C approximately 10% of the zinc reacts with iron impurities within the zinc concentrates to form zinc ferrite or spinel, but that is dependent on the iron concentration in the feed.

Waste heat boilers remove the bulk of the calcine contained in the gas as well as recovering the heat in the form of steam that is used in the leaching plant and/or generation of electricity.

The by-product of roasting sulfur dioxide, is captured and further processed into sulfuric acid of between 92 and 98% strength through the double contact and adsorption process and sold as a by product. The reactions for conversion and absorption are shown below:



Three types of roasters have been used within the RLE process, including, multiple hearths, suspension and fluidized bed roasters. The main differences between these roasters are the temperature of operation, leading to varying rates of desulphurisation per unit type.

4.2.1.2 *Neutral leaching and Purification*

During the Neutral Leach stage 80 – 90% of the calcine is leached in recycled spent electrolyte from the electrowinning circuit. The pH is typically controlled at around 4 – 4.5 in the final leaching stage to precipitate iron. Zinc-ferrites don't leach under these conditions and the residue from the neutral leach is fed to the Residue Leach circuit.

The leach solution is sent to a purification circuit to remove impurities such as cadmium, copper, cobalt or nickel, which could also affect the electrolysis operation. These impurities are removed through cementation by adding zinc dust and other additives such as As or Se for Co removal and CuSO_4 for Cd removal. The resulting by-products are generally sold to third parties for further refining. The purified zinc sulphate solution is sent to the cell house for the electro-winning phase of the smelting process.

4.2.1.3 *Electrowinning and Smelting*

Zinc metal is extracted from the purified solution by means of electrolysis. An electric current is passed through the solution and the zinc is deposited on aluminium cathodes. At regular intervals, these cathodes are removed from the cells and the zinc deposit is stripped from the cathodes. Most of the Nyrstar smelters use a mechanised and automated process of stripping machines, which was initially developed at the Balen plant.

The zinc produced with the electrolysis process is typically of SHG grade containing 99.995% zinc.

The electrolysis phase uses ~ 3200 kWh/tonne zinc produced and the electrical energy contributes ~ 80% of the overall energy-cost in the smelting process. Typically about one third of total plant cash costs are for electricity. Hence, cell house productivity (and electrical current and energy efficiency in particular) is a crucial driver in overall plant efficiency.

Depending on the type of end-products produced, the zinc cathodes coming out of the electro-winning plant can undergo an additional transformation step in a foundry. Zinc cathodes are

melted in induction furnaces and cast into marketable products such as ingots. Other metals and alloy components may be added to produce zinc containing alloys used in die-casting or general galvanization applications.

4.2.1.4 Residue Treatment

The Neutral Leach residue is fed to the residue treatment circuit. Whilst zinc oxide can be leached with dilute sulphuric acid solutions, the leaching of zinc from ferrites requires more aggressive conditions. The residue treatment typically contains weak, hot and super hot acid leaching steps, where the super-hot acid leach is operated at 80 – 90°C with ~ 100 g/L H₂SO₄. High zinc leaching efficiencies can be achieved and the final residue (Pb/Ag residue) can be sold to third parties for further refining.

The dissolved iron is removed from the zinc sulphate solution as goethite, jarosite or hematite which is usually stored in ponds. The conditions for production and the composition of the final product are shown in Table 4-4.

Table 4-4 Iron residue conditions/Specification

		Jarosite M[Fe ₃ (SO ₄) ₂ (OH) ₆]	Goethite FeOOH	Hematite Fe ₂ O ₃
pH		1.8	2.8 - 3	
Temperature	°C			200
Pressure	Bar	Ambient	Ambient	10 - 15
Fe content	% w/w	25 - 30	40 - 45	55

The technology selection for iron removal will determine:

- the type of equipment required, e.g. and autoclave would be required for the production of hematite;
- the stability of the residue, e.g. if jarosite is produced a Jarofix process might be required where the residue is mixed with lime and cement to increase the long-term stability;
- the size of the residue pond based on the Fe content in the residue; and
- the overall Zn recovery in the process.

Ever increasing environmental pressure is moving technology towards the production of an inert residue. A pyrometallurgical process utilising the Ausmelt Top Submerged Lancing (TSL) Technology or Waelz Kiln to produce a solid clinker as the residue, is an alternative to the residue leach and iron removal. This was successfully incorporated into the traditional Roast Leach Electrowinning process at Korea Zinc. In this process, Fe reports to the slag and an upgraded Pb/Ag residue can be produced. Note that the pyrometallurgical process for residue treatment was not included in the capital and operating cost estimate.

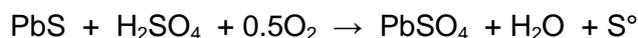
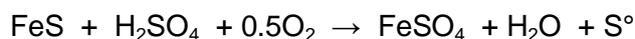
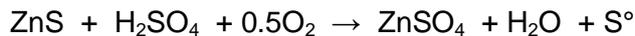
4.3 Pressure Leaching (Sherritt Zinc Pressure Leach Process)

The Direct Leaching of Zinc concentrate was developed and commercialised by Cominco and Sherritt-Gordon in Canada with the first plant coming on stream in 1981. Today there are 5 zinc refineries using the pressure leaching approach. The pressure leaching process was primarily developed to address the environmental problems associated with roasting of sulphide ore and the disposal of sulphuric acid. Generally the pressure leaching plants have a somewhat lower capital cost than conventional roast/leach plants, higher extraction efficiencies, produce a more acceptable residue as well as elemental sulphur which may have a potential commercial value.

The Sherritt Zinc Pressure Leach Process is inherently simple and is based upon the direct leaching of zinc sulphide bearing concentrates in zinc electrowinning spent electrolyte or other acid bearing solutions under oxygen pressure. Zinc sulphide and other metal sulphides react with sulphuric acid and oxygen to produce metal sulphates and elemental sulphur. An exception is pyrite (FeS₂) which is oxidized directly to sulphate without significant formation of elemental sulphur.

4.3.1 Process Chemistry

The zinc pressure leach process depends upon simple reactions in which zinc sulphide, pyrrhotite or iron in sphalerite, galena and chalcopyrite react with sulphuric acid and oxygen to yield metal sulphates and elemental sulphur.



These reactions are slow in the absence of a species which will facilitate oxygen transfer. One such species is dissolved iron. The net reaction, as shown above, is then the sum of two reactions.



Normally there is sufficient acid soluble iron in the concentrate to supply the needs of the leach.

Pyrite is present in most zinc concentrates and its behaviour in the pressure leach depends on a number of leaching parameters. Under normal zinc pressure leaching conditions, the sulphide content of pyrite is oxidized directly to sulphate, with little or no elemental sulphur production.

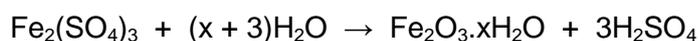


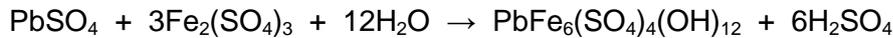
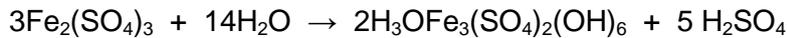
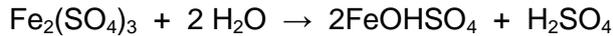
Usually, up to 5 % of the non-pyritic sulphide sulphur in the concentrate is oxidized to sulphate sulphur during the pressure leach.



Where *M* represents Zn, Pb, Cu or Fe.

Iron is removed from solution by hydrolysis reactions which can also remove sulphur from solution.





The precipitation of iron in the pressure leach autoclave may be minimized by maintaining a relatively high residual acid concentration in the pressure leach solution. It may be desirable to limit the precipitation of iron during the zinc extraction process if it is necessary to separate a high grade lead silver product from the pressure leach residue.

4.3.2 Pressure Leach Process

Zinc concentrate or bulk concentrate are leached in spent electrolyte under oxygen pressure to produce a zinc sulphate solution. The concentrate may be reground to improve both the rate of reaction and the overall zinc extraction. The particle size of the concentrate should be typically 90% passing 44 microns. Surface active agents, such as calcium ligno-sulphonate and quebracho, are added to inhibit coating of the un-reacted zinc sulphide particles by molten elemental sulphur and to control the particle size of the sulphur micro-pellets formed during discharge and cooling of the slurry.

Concentrate slurry is pumped to the first compartment of a multi-compartment autoclave. The autoclave is typically divided into five compartments with the first compartment being larger than the other compartments to optimize the thermal balance. Spent electrolyte is pumped to the first compartment and, optionally, to the second compartment to control both the thermal balance and the acid concentration profile. Each compartment is equipped with an agitator (more than one agitator in the first compartment). Oxygen is sparged to the base of each impeller.

Leaching is typically conducted at 150°C, under an oxygen partial pressure in the range 700 to 1500kPa, and with a retention time of one hour. The oxygen partial pressure is varied to control pyrite oxidation which is greater at higher pressures. Molten elemental sulphur has a minimum viscosity at the leaching temperature and leaching kinetics is acceptable. Under these conditions, the heat required to maintain the operating temperature is supplied by the heat of reaction with minimal solution preheating required.

Slurry is discharged from the autoclave under controlled conditions, to produce a residue amenable to the separation of elemental sulphur and un-reacted sulphides from gangue and oxidic species by flotation. The oxidic flotation tailings are either treated directly in a lead smelter or further upgraded to a product which can be treated in a lead smelter. The sulphidic concentrate is washed to remove salts, melted and filtered under pressure to recover an elemental sulphur by-product. The filter cake, containing un-reacted sulphides and entrained elemental sulphur, may be impounded or treated further for recovery of the contained metals, including precious metals, and sulphur.

Typical zinc extraction in the zinc pressure leach process is 98% or greater and the overall recovery of elemental sulphur is typically between 60 and 80%, depending on the pyrite content of the concentrate. A typical pressure leach zinc flowsheet is shown in Figure 4-3.

Two Stage Countercurrent Zinc Pressure Leach

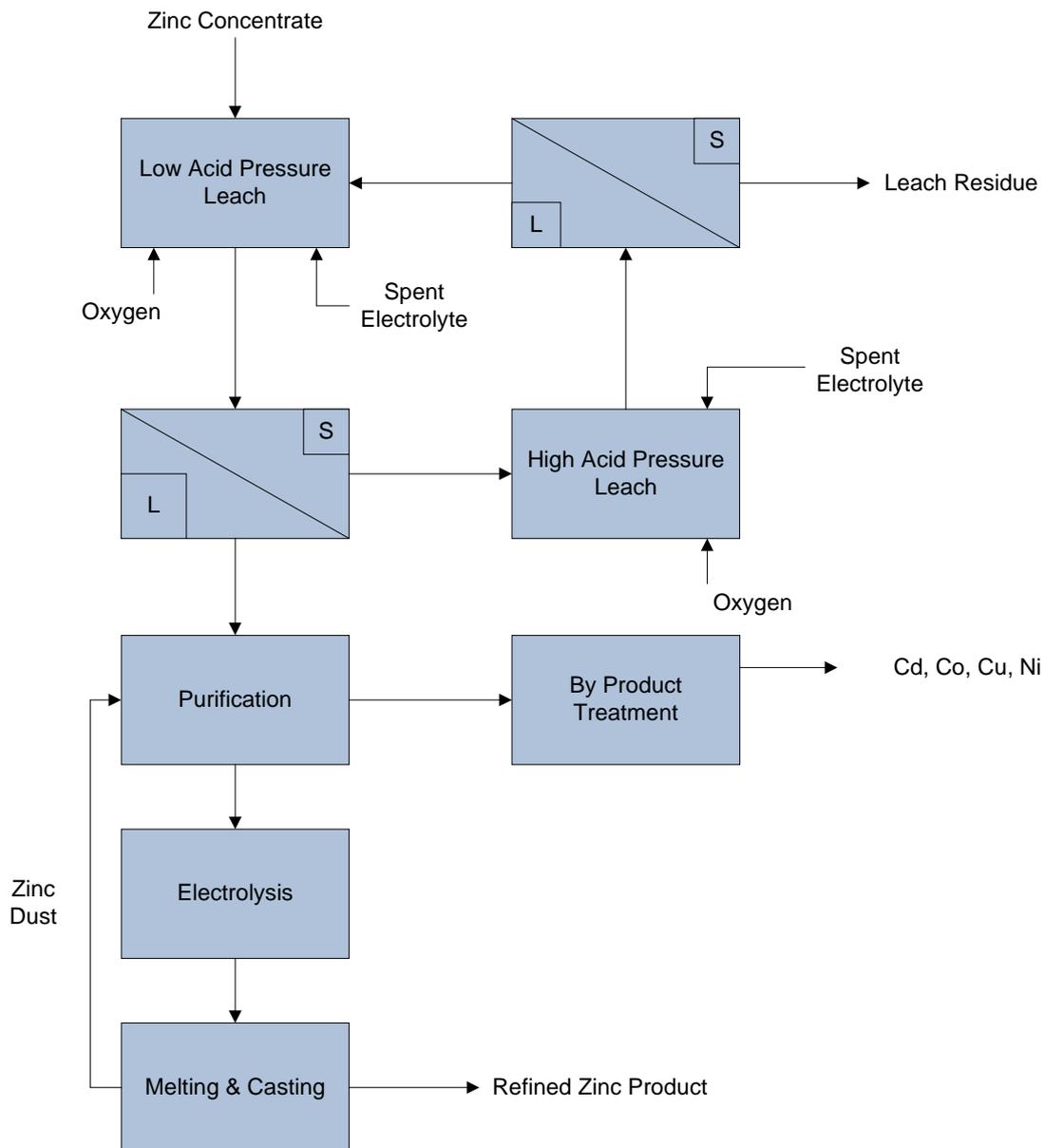


Figure 4-3 Pressure Leaching Blockflow Diagram

4.4 Direct Leaching – Atmospheric Leaching (Outotec Technology)

Outotec developed atmospheric direct leaching technology of zinc concentrates in the middle of 1990s. This technology has been practiced in industrial scale since 1998 in Kokkola, Finland and since 2004 in Odda, Norway at Outotec refining operations. Originally the technology was developed for Outokumpu’s internal use only, but is now marketed by Outotec as a

commercially available atmospheric direct leaching technology offering a way to produce zinc without generating sulphur dioxide formed in conventional roasting-leaching-electrowinning process, or requiring the use of high pressure and temperatures. The process has also been successfully installed and operated at Zhuzhou Zinc China.

4.4.1 Process Chemistry

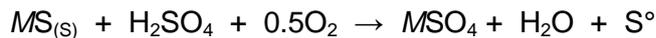
The direct leaching of zinc concentrates is based on the oxidation of zinc sulphide in an acidic environment. Several parallel and consecutive reactions take place simultaneously and also physical phenomena play an important role in the leaching system. However, the direct leaching of sulphide minerals can be written with the following simplified reaction equations. The reduction of ferric iron is accomplished by metal sulphides, which are leached:



Where $M = Zn, Fe, Cu, Co, Ni, Cd, Pb$, etc. Formed ferrous is oxidised by molecular oxygen back to the ferric form:



The reaction rate is catalyzed by copper. The overall reaction can be written as the following simplified equation:

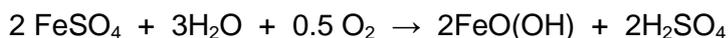


Zinc in the leach residues of neutral and weak acid leaching steps is mainly as zinc ferrites but also small amount of unreacted zinc sulphide is present. Leaching of zinc ferrites can be described with the following simplified equation:

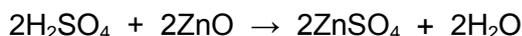


Leaching of ferrites requires temperature close to 100°C and acid concentration of above 30 g/L H_2SO_4 .

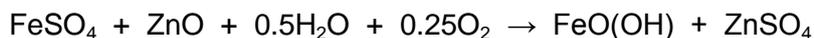
The most relevant reactions occurring in the precipitation process are described below. Oxidation of ferrous iron to the ferric iron, which is precipitated as goethite:



Neutralization of the formed sulphuric acid is made by zinc oxide:



The reaction system can be written as an overall reaction:



4.4.2 Atmospheric Leach Process

Zinc concentrate or bulk concentrate are leached in the presence of ferric sulphate at atmospheric pressure to produce a zinc sulphate solution. The concentrate may be reground to improve both the rate of reaction and the overall zinc extraction. The particle size of the concentrate can be as coarse as 90% passing 150 microns. As the leaching temperatures are under the melting point of sulphur there is no need for the use of surface active agents, such as ligno-sulphonates as required in the pressure leaching process.

Zinc concentrate is fed into various stages comprising of low acid leaching and hot acid leaching, either co- or counter-currently. OKTOP™ DDT reactors provide the essential dispersion of oxygen and slurry required for the related reactions.

Leaching is typically conducted at 90°C. The oxygen partial pressure is varied to control pyrite oxidation to ensure the ferric iron content, acidity and copper content are maintained at the desired levels.

A key benefit of the Outotec's direct leaching processes is that the production of sulphur dioxide and other harmful gaseous compounds is reduced, especially in comparison to the conventional pyrometallurgical methods. On the other hand, elemental sulphur, which requires a special storage area, is formed. Small amounts of compounds such as Se, Cd, As, Hg and Pb, hinder further usage of the sulphur residue without purification. The lead-rich fraction of leach residue is utilized in the existing lead smelter.

The amount of iron precipitate is dependent on the iron content in the concentrate and also on the iron removal methods.

Typical zinc extraction in the zinc pressure leach process is between 96% and 99% and the overall recovery of elemental sulphur is typically between 60 and 80%, depending on the pyrite content of the concentrate. A typical Outotec direct zinc leach flowsheet is shown in Figure 4-4.

Atmospheric Leach: Outotec Direct Leach Flowsheet

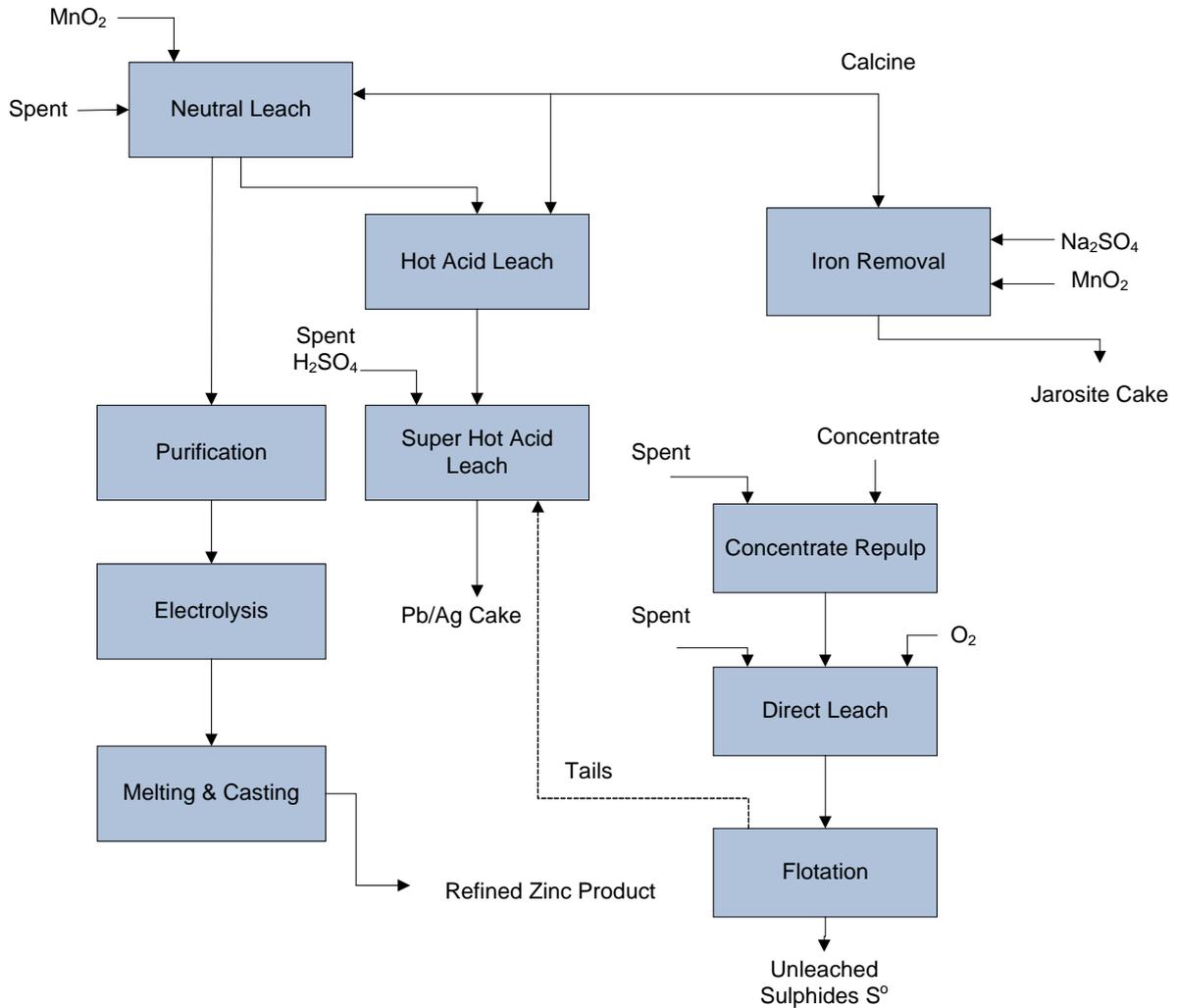


Figure 4-4 Outokumpu Direct Leaching Blockflow Diagram

5 CAPITAL COST ESTIMATE

5.1 Estimate Parameters

The estimate has been defined as a Desktop Study – Order of Magnitude class estimate. This corresponds to a Class 5 estimate in the Bateman guidelines and is deemed to have accuracy in the range of +/- 40%.

This estimate was produced to provide the client with sufficiently detailed options to assess the viability of a zinc refinery at the Citronen Project. The scope accuracy of these options is NOT to a level that allows the client to source project funding.

The estimates were prepared using the process design basis, input from Ironbark and the Bateman cost database. Vendor estimates were obtained for proprietary technologies from Outotec and Sherritt Technologies for the Atmospheric Leaching and Pressure Leaching flows.

5.2 Basis of Estimate

The estimate has been prepared in fourth quarter 2010 (Q4-2010) using Australian Dollars (A\$) to an accuracy of $\pm 40\%$. All vendor quotations (where applicable) were given as of that date with limited validity. The price at this base date must therefore be interpreted as escalatable, i.e. it is not a fixed price even if the project could have commenced at that point in time. The price estimate will be subject to the following categories of escalation:

- Price Escalation from the base date of end of Q4 2010 to the date of submission of this report.
- Price Escalation from the date of submission of this report to the commencement of the actual project.
- Price Escalation that will be incurred during the execution of the project.
- Demand inflation.

As a result of these factors, the reliability of the price for capital costs established at a point in time rapidly diminishes with time. The price must be revalidated on the basis of valid quotations within a short period of the actual project commencing. Even then, the reliability of the price will not be high as allowances must be made for price escalation that will be incurred during the execution of the project. In light of the foregoing, Bateman takes no responsibility for forecasting the Price Escalation effects and therefore the final as-built price of the project. Ironbark is responsible for making allowance for the different types of Price Escalation. Bateman has not reviewed these allowances and cannot be held responsible for the accuracy of the Client's prediction of Price Escalation effects.

5.3 Direct Cost Sourcing

The majority of the direct costs were derived from recently completed studies or from specific zinc refining technology providers. It can be generally said that the costs were based on single sourced budget quotes from suppliers sourcing a large proportion of the equipment (as a whole or in parts). The estimate strategy varied for each area of the study and is outlined in Table 5-1

Table 5-1 Flowsheet Estimate Design Basis

Flowsheet Option	Design	Cost Estimate
Imperial Smelting Plant	Based on traditional ISP flowsheet design	Based on data sourced in public domain regarding last ISF plant built in Rajasthan India, 1991
Roast Leach Electrowinning	Based on Bateman design	Estimate built up on a similar project feasibility study pricing using contemporary unit rates.
Pressure Leaching	Based on Sherritt Technology Design	Based on Sherritt Technology budget estimate and adjustments for site specific conditions
Atmospheric Leaching	Based on Outotec Design	Based on Outotec budget estimate and adjustments for site specific conditions

5.3.1 Battery Limits

Bateman closely observed the process plant battery limits to ensure duplication of costs with other areas did not occur. The following direct cost areas are not included in this report.

- Concentrate Storage.
- All power generation and reticulation facilities.
- All water supply costs.
- All tailings storage facility costs.

5.3.2 Infrastructure

No cost was included for infrastructure outside the plant area. including: Power generation, Borefields, Access roads, etc..

5.3.3 Bulk Earthworks

No costs were included for bulk earthworks, as these could not be determined due to the plant location not being established. All costs assume a clear level, accessible site.

5.4 Indirect Cost sourcing

5.4.1 Temporary Facility

An allowance to cover temporary works is included in each estimate. This works include:

- Construction offices
- Messing facility
- Concrete batch plant
- Generators
- Construction Services, such as air, power, water, etc.
- Sewage
- Cleaning and disposal services.

5.4.2 Engineering, Procurement and Construction Management (EPCM)

An allowance based on a percentage of direct costs has been included to cover EPCM costs.

5.4.3 Qualifications

No allowances have been made for the following items:

- Client costs.
- Infrastructure.
- Environmental studies, investigations, permits or liabilities.
- Escalation.
- Scope Changes.
- Costs of permits.
- Legal fees.
- Insurance.
- Land acquisition or rights of way and any resettlement Costs.
- Cost of decommissioning at end of project life.
- Force majeure events.
- Escalation.

5.4.4 Contingency

A contingency has been allowed for each flowsheet option. This has been calculated as a percentage of the direct cost. The purpose of the contingency is to make specific provision for undefined elements of costs within the defined project scope of work. The contingency is not intended to cover scope changes, force majeure, contractor's non-performance, escalation and other unforeseeable events.

5.5 Capital Cost Estimate

The total capital cost estimates for the four zinc refinery flowsheets evaluated as part of the desktop study are presented in Table 5-1. A break down estimate by major process area for each flowsheet option is presented in Table 5-2 to Table 5-5.

Table 5-2 Capital Cost Estimate

Imperial Smelting Process	Roast Leach Electrowinning	Pressure Leaching	Atmospheric Leaching
\$A 861,000,000	\$A 508,500,000	\$A 435,000,000	\$A 491,000,000

Table 5-3 Capital Cost Imperial Smelting Process Flowsheet

Direct Cost	\$A '000,000
Sinter Plant	123.5
Imperial Smelting Furnace	145
Refinery	145
Acid Plant	61
Utilities	140
Tank Storage	68
Total Direct Costs	682.5
Indirect Costs	
EPCM	71.5
Site Temporary Facilities	12
Technology Fees	15
Owners Cost	48
Total Indirect Costs	146.5
Contingency	32
Total A\$	861

Table 5-4 Capital Cost Roast Leach Electrowinning Flowsheet

<i>Direct Costs</i>	\$A '000,000
Raw & Finished Product handling	23
Roaster Plant	29
Sulphuric Acid Plant	34
Leaching and purification Plant	52
Zinc EW Plant	42
Melting plant and Cast House	16
Neutral Leach residue treatment	68
Utilities general facilities	77
Tank Storage	68
Total Direct Costs	409
<i>Indirect Costs</i>	
EPCM	40
Site Temporary Facilities	6.5
Technology Fees	8
Owners Cost	27
Total Indirect Costs	81.5
Contingency	18
Total A\$	508.5

Table 5-5 Capital Cost Pressure Leach Flowsheet

	\$A '000,000
Leaching	81
Purification	37
Refinery	122
Utilities	129
Total Direct Costs	370
Indirect Costs	65
Total Costs	435

Table 5-6 Capital Cost Atmospheric Leach Flowsheet

	\$A '000,000
Leaching	138
Purification	71
Refinery & Utilities	209
Total Direct Costs	417
Indirect Costs	74
Total Costs	491

6 OPERATING COST ESTIMATE

6.1 Basis of Estimate

The estimate has been prepared to a confidence level of +/-40%. All costs in the estimate are expressed in Australian Dollars (\$A) based on a base date as at the end of the 4th Quarter 2010.

6.2 Summary

The operating cost estimate was prepared using data from previous similar studies, preliminary estimated power draw for the refinery based on an annual power draw per tonne of zinc being produced, input from Ironbark and the Bateman reagent cost database. The operating costs were broken down into the following areas for calculation except for the Imperial Smelting Process flowsheet as the data sourced from an existing ISP was not presented in this format:

- Personnel;
- Power;
- Consumables and reagents; and
- Maintenance.

Exclusions from all of the Opex cost estimates include:

- administration costs,
- environmental services
- waste disposal services
- transportation costs of acid and/or sulphur products
- credit for acid and/or sulphur products
- Storage and/or production costs associated with lead/zinc concentrate production
- Residual treatment costs for processing lead concentrate and/or residues
- Fuels for light vehicles and or plant machinery

6.2.1 Power cost

The estimated power cost was based on a fuel consumption of \$A0.23 litres per kWh at a diesel price of \$A 0.64c/l as supplied by Ironbark.

6.2.2 Labour Cost

The labour costs were based on an estimated manpower requirement for each flowsheet, using an average cost of employment per employee of \$80,000pa. It has been assumed that some supervision and management of the zinc refinery can be provided from employees of the lead/zinc refinery.

The annual operating cost summary for the RLE, Pressure and Atmospheric Leaching flowsheet options are shown in Table 6-1 and Table 6-2.

Operating costs for the ISP flowsheet are presented in based on the major unit process areas as a cost breakdown by Opex type was not available in the sourced information.

Table 6-1 Annual Operating Cost Estimate

	Roast Leach Electrowinning	Pressure Leaching	Atmospheric Leaching
Personnel	26,400,000	28,000,000	22,800,000
Power	70,200,000	77,100,000	80,000,000
Consumables & Reagents	5,500,000	4,700,000	2,500,000
Maintenance	20,900,000	18,500,000	17,000,000
Total \$A/a	123,000,000	128,300,000	122,300,000

Table 6-2 Operating Cost per pound Zinc Produced

	Roast Leach Electrowinning	Pressure Leaching	Atmospheric Leaching
Personnel	0.100	0.106	0.086
Power	0.265	0.291	0.302
Consumables & Reagents	0.021	0.018	0.009
Maintenance	0.079	0.070	0.064
Total \$A/ lb Zn Produced	0.464	0.485	0.462

Table 6-3 Operating Cost Estimate for Imperial Smelting Process

	Annual Cost (\$A)	Annual Cost (\$A/lb Zn Produced)
Sintering Acid Plant	50,700,000	0.192
Imperial Smelting Furnace	87,800,000	0.332
Refinery	15,300,000	0.058
Administration and Engineering	44,200,000	0.167
Total	198,000,000	0.748



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