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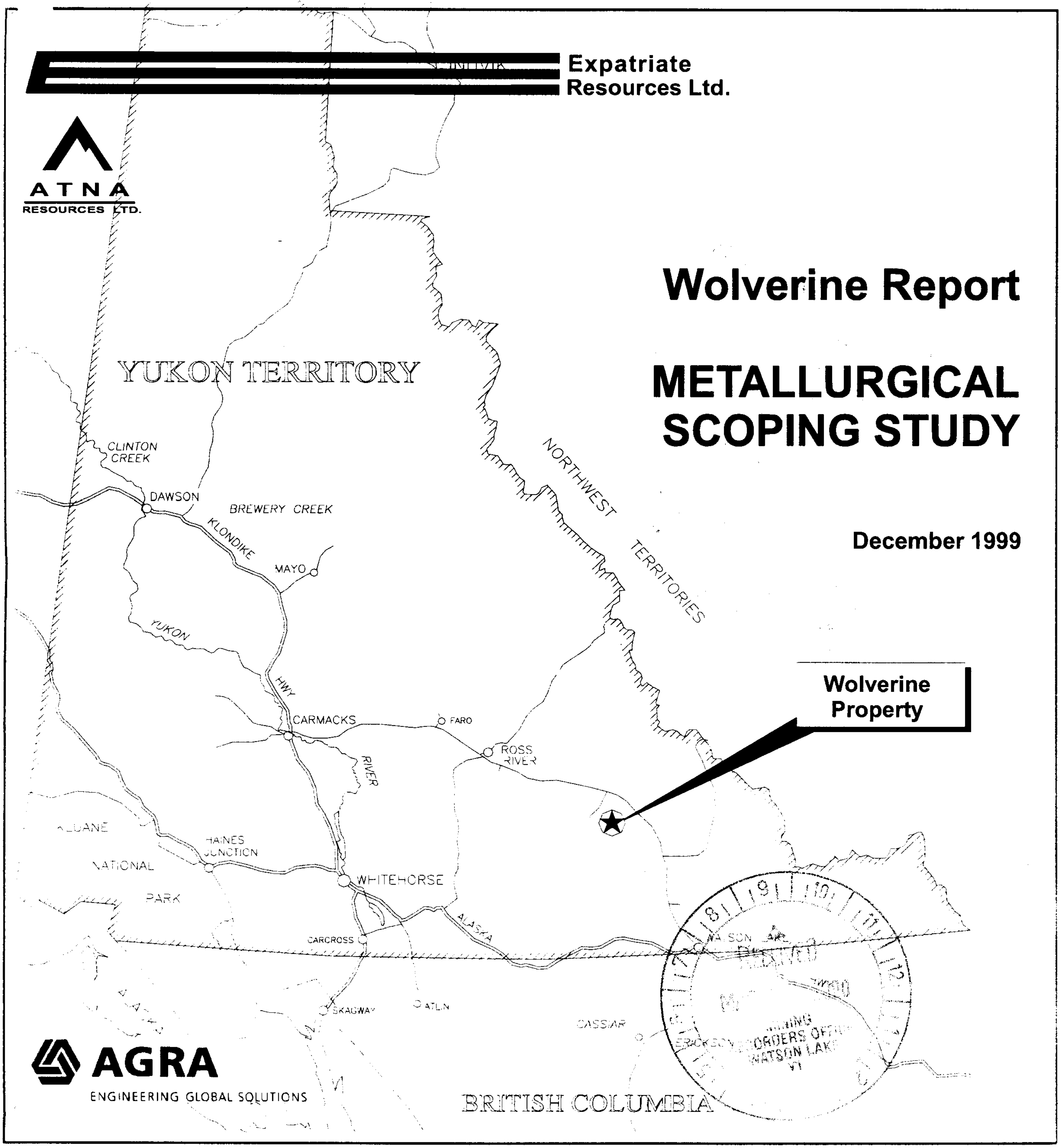
Expatriate Resources Ltd.

ATNA RESOURCES LTD.

Wolverine Report

METALLURGICAL SCOPING STUDY

December 1999



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This report has been examined by
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under Section 53 (4) Yukon Quartz
Mining Act and is allowed as
representation work in the amount
of $ 110,675.00.

for Regional Manager, Exploration and
Geological Services for Commissioner
of Yukon Territory.

AGRA
ENGINEERING GLOBAL SOLUTIONS

AGRA Simons Limited
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January 10, 2000

Expatriate Resources Ltd.
Suite 1016 - 510 West Hastings Street
Vancouver, British Columbia
Canada V6B 1L8

Attention: Mr. Brad Marchant
Project Manager

Dear Brad:

Re: Wolverine Project Scoping Study

Enclosed please find six copies of the final draft of the study entitled:

Metallurgical Scoping Study

We appreciate your confidence in entrusting this study to AGRA Simons Limited, and we look forward to providing further assistance as required.

Yours truly,

for
Lynton Gormely, Ph.D., P.Eng.
Principal Process Engineer

Encl.

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CERTIFICATE OF QUALIFICATIONS

I, Terry L. Tucker, P.Geo., residing at 1541 Mahon Avenue of the City of North Vancouver, in the Province of British Columbia,

HEREBY CERTIFY THAT:

1. I am a graduate of the University of Alberta (BSc. Geology, 1989),
2. I am a registered Professional Geoscientist with the Association of Professional Engineers and Geoscientists of British Columbia (registration number 20108),
3. I was engaged full time as a mineral exploration geologist for Westmin Resources Ltd. between 1992 to 1997, and have been a consulting geologist from 1997 to present,
4. Between 1992 to 1997 while an employee of Westmin Resources Ltd., I participated in the discovery and acquisition of the Wolverine property and with my colleagues jointly planned and implemented the advanced exploration effort at that property; my most recent site visit was in 2000,
5. I have been contracted by Expatriate Resources Ltd. to prepare an independent technical report on the Wolverine property and am qualified to do so,
6. I do own shares and hold stock options in Expatriate Resources Ltd. or its affiliates,
7. I have reviewed all available exploration information referenced herein and based my recommendations upon the exploration information,
8. I am unaware of any material fact or material change with respect to the technical matter of this report that might cause the technical report to be inaccurate or misleading,

SIGNED at Vancouver, British Columbia this 2nd day of March 2001.

Terry L. Tucker, P.Geo.
Exploration Manager

T. L. TUCKER

IMPORTANT NOTICE

This report was prepared exclusively for Expatriate Resources/Atna Resources Ltd. by AGRA Simons Limited (AGRA Simons). The quality of information, conclusions and estimates contained herein is consistent with the level of effort involved in AGRA Simons' services and based on: i) information available at the time of preparation, ii) data supplied by outside sources, and iii) the assumptions, conditions and qualifications set forth in this report. This report is intended to be used by Expatriate Resources/Atna Resources Ltd. only, subject to the terms and conditions of its contract with AGRA Simons. Any other use of, or reliance on, this report by any third party is at that party's sole risk.

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Refer to "Expatriate Resources Ltd. 138 kV Transmission Line Prefeasibility Study", prepared by Ian Hayward International Ltd., November 1999. (Not bound with this report).

Appendix C

Basis for Bio-Leach Process letter from Pacific Ore Technologies.

Appendix D

Refer to "Wolverine Joint Venture Preliminary Metallurgical Test Work Results", prepared by International Metallurgical and Environmental Inc., September 1999. (Not bound with this report).

1.0 TERMS OF REFERENCE

1.1 PROJECT DESCRIPTION

The Wolverine and Kudz Ze Kayah deposits are volcanogenic massive sulfide deposits located in the Finlayson Lake District of the Yukon Territory (see Figure 1.2). The Wolverine deposit is under development by a joint venture including Expatriate Resources Ltd. and Atna Resources Ltd., while the Kudz Ze Kayah deposit is being developed by Cominco. While significant reserves have been established for both deposits through the initial exploration activities and the excellent potential for additional discoveries, development of these resources has been hampered by technical and economic factors.

The metallurgical testwork conducted on the deposits has demonstrated that separate copper, lead and zinc concentrates can be produced at concentrate grades and recoveries that are typical for the industry. The concentrates produced from the Wolverine deposit contain selenium concentrations which are about ten times those found in more typical deposits. This selenium content limits the ability of smelters to accept the concentrates as well as resulting in significant smelter penalties. Kudz Ze Kayah concentrates also contain elevated selenium levels but not to the same extent as Wolverine. It has been determined for both deposits that the use of physical mineral processing procedures will not result in lowering of the selenium content of the concentrates. Secondary pyrometallurgical or hydrometallurgical processes will therefore be required to produce marketable products. Arsenic and antimony are also present in the concentrates and their presence further affects the marketability of the products.

Due to the combination of losses during the concentrating process, smelter charges and deductions as well as transportation charges, approximately two thirds of the gross metal value in the ore is lost. In evaluating process alternatives to deal with the selenium issue, consideration must also be given to the opportunity for capturing more of the gross metal value of the ore.

It is assumed that the two deposits could most logically be developed as a single project. In order to address the metallurgical and economic issues for such a project, Expatriate and Atna have assembled a study team to evaluate available technology for achieving this objective. Preliminary cost estimates have been completed for nine process alternatives which were judged to have sufficient technical merit for the project. The results are presented in this report.

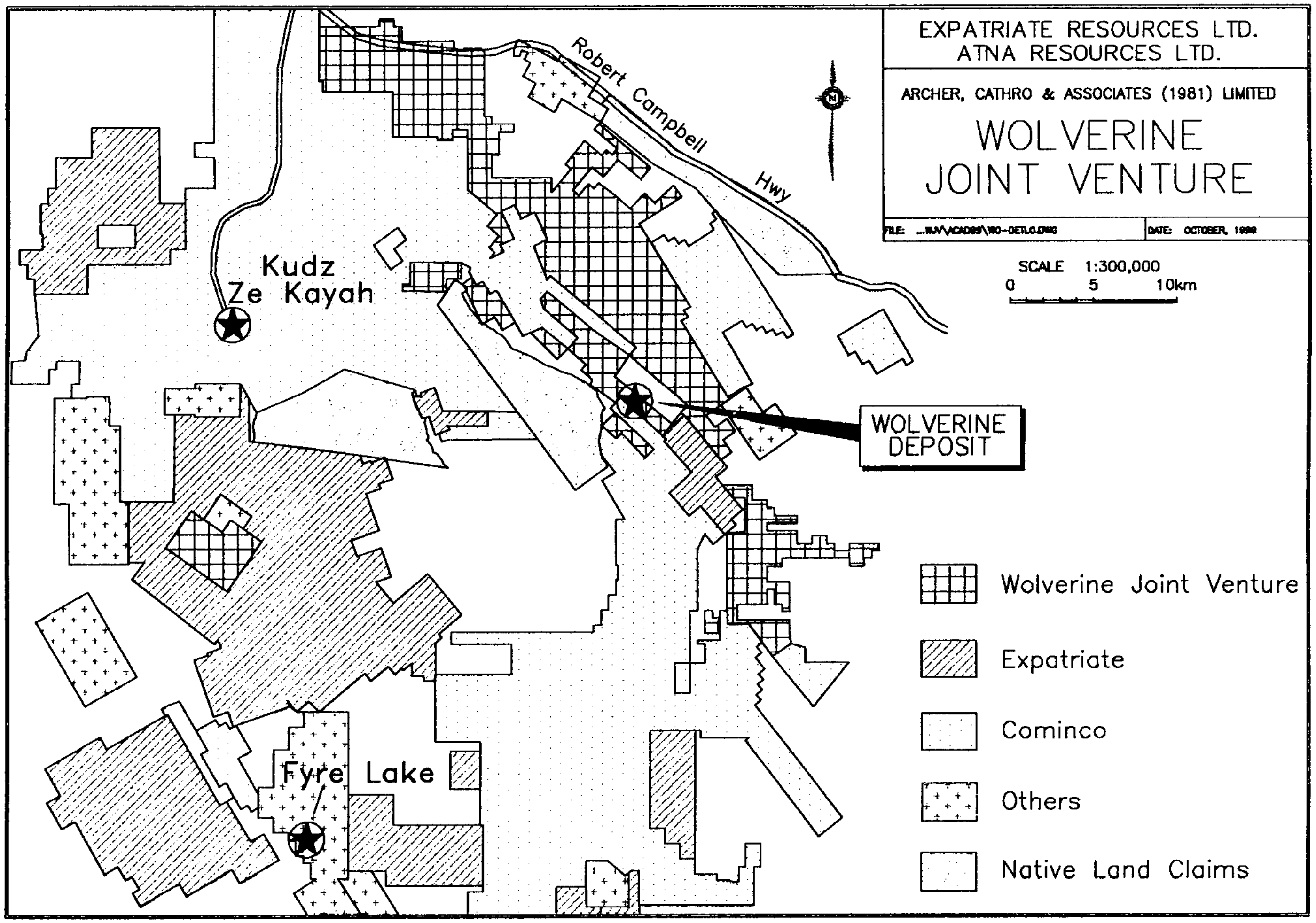

Figure 1.1 Wolverine Joint Venture

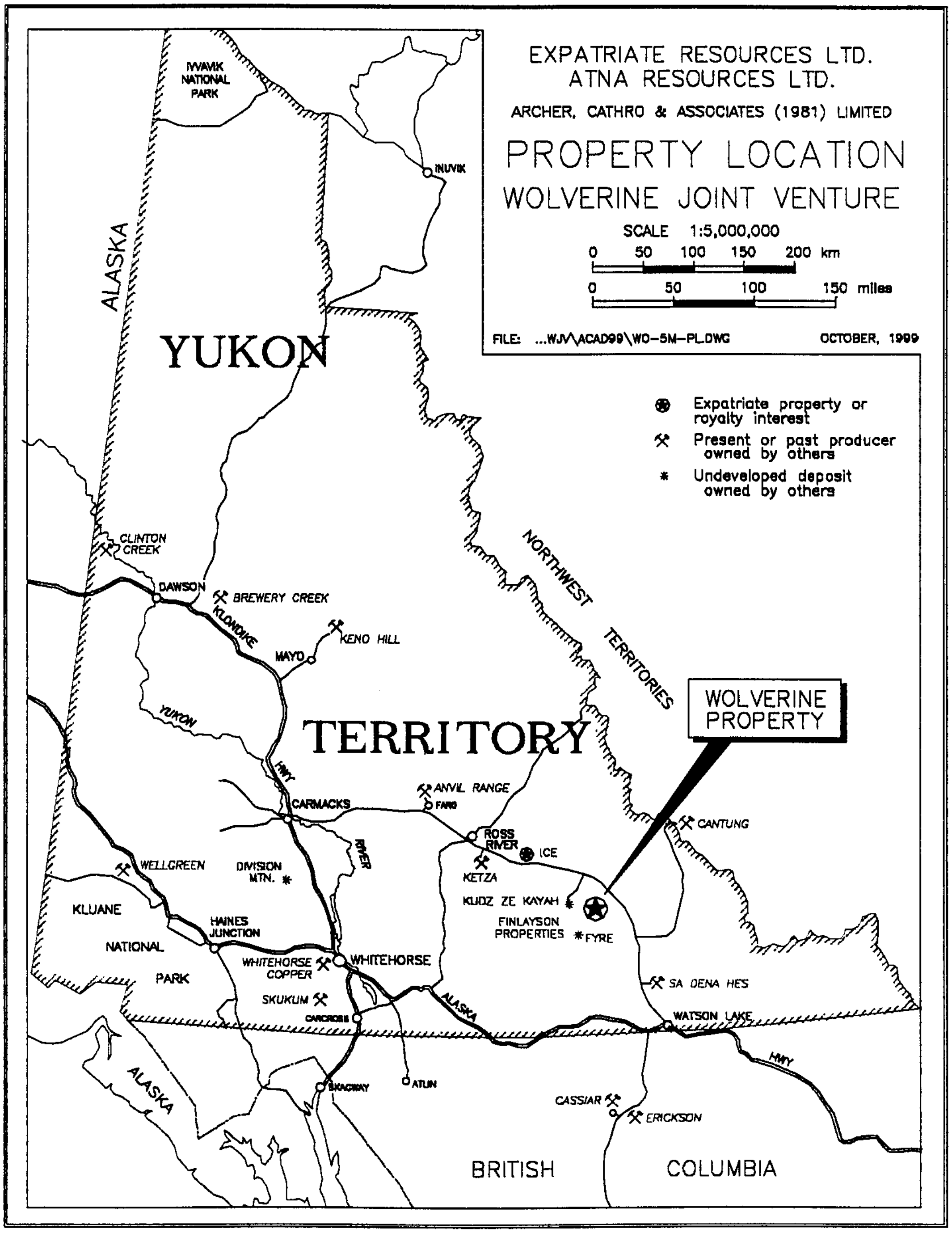

Figure 1.2 Property Location

1.2 PROJECT DESIGN CRITERIA AND ASSUMPTIONS

The objective of the Metallurgical Scoping Study was to provide process direction to treat a combined mill feed from the Wolverine and Kudz Ze Kayah deposits. The result of the study should provide viable process alternatives, if any, to continue to prefeasbility level testwork and engineering.

The accuracy of process and cost estimates for this study is ±30% on cost estimates and all data employed was either extracted from previous work or calculated. No new testwork was conducted for this study.

Concurrent with the metallurgical study, an investigation was initiated to better define the potential for flotation concentrate marketing, with and without further treatment at the site. Mr. Jack Butterfield is coordinating the marketing study. This study is in progress and will be reported under separate cover.

For the purpose of costing exercises the total production rate from the two deposits was taken as 3000 tpd. The resource base used for the two deposits was:

Deposit	**Wolverine**	**KZK**
Tonnes	6,200,000	10,000,000
Grade		
Zn (%)	10.13	5.90
Pb (%)	1.24	1.50
Cu (%)	1.06	0.90
Ag (g/t)	297	133
Au (g/t)	1.41	1.33

Process concepts were evaluated based on the processing of combined mill feed from the two deposits at a ratio of 5/8 Kudz Ze Kayah (KZK) and 3/8 Wolverine. At this ratio, the combined tonnage is 16.2 million tonnes and the head grade is:

Zn	7.49 %
Pb	1.40 %
Cu	0.96 %
Ag	194.40 g/t
Au	1.34 g/t

Wolverine reserves were based on 20% expected dilution at zero grade and 80% mining recovery from the Wolverine underground deposits and open pit production from KZK, providing a combined feed for a 15 year mine life.

Three alternatives for the production of concentrates were considered, in addition to the possibility of treating the whole ore:

- Selective Cu, Pb and Zn concentrates
- Bulk Cu/Pb and separate Zn concentrate
- Bulk Cu/Pb/Zn concentrate

Since the metallurgical testwork has progressed further on the KZK deposit than on Wolverine, the predictions for concentrate grades and recoveries, based on the above head grade, were taken from the KZK testwork and adjusted to account for higher grade Wolverine ore and higher copper and silver recovery from Wolverine testwork. The concentrate grades and recoveries for the selective and bulk concentrate options are summarized in Tables 1.1 through 1.3.

Table 1.1 Annual Material Balance for Selective Concentrates (3,000 tpd)

	Tonnage	Assays, %, g/t					Distribution, %				
Product	**(tpa)**	**Cu**	**Pb**	**Zn**	**Au**	**Ag**	**Cu**	**Pb**	**Zn**	**Au**	**Ag**
Cu conc	33,281	26.5	0.9	5	20.5	2,853	83.9	2.0	2.0	46.5	44.6
Pb conc	15,594	0.2	52.2	5.4	1.6	1,160	0.3	53.1	1.0	1.7	8.5
Zn conc	140,103	0.5	1.5	52.1	1.5	350	6.7	13.7	89.0	14.3	23.0
Tailing	906,021	0.11	0.53	0.72	0.6	56	9.1	31.2	7.9	37.5	23.9

Table 1.2 Balance for Zn Concentrate plus Bulk Cu/Pb Concentrate

	Tonnage	Assays, %, g/t					Distribution, %				
Product	**(tpa)**	**Cu**	**Pb**	**Zn**	**Au**	**Ag**	**Cu**	**Pb**	**Zn**	**Au**	**Ag**
Cu/Pb	74,460	12	16.5	6.0	12.8	2,001	85.0	80.0	5.4	65.0	70.0
Zn conc	138,529	0.5	1.5	52.1	1.0	150	6.6	13.6	88.0	9.4	9.8
Tailing	882,011	0.10	0.11	0.61	0.4	49	8.4	6.4	6.6	25.6	20.2

Table 1.3 Balance for Production of Bulk Concentrate

	Tonnage	Assays, %, g/t					Distribution, %				
Product	**(tpa)**	**Cu**	**Pb**	**Zn**	**Au**	**Ag**	**Cu**	**Pb**	**Zn**	**Au**	**Ag**
Conc	225,000	4.3	6.4	34.3	5.0	757	93.0	94.0	94.0	76.0	80.0
Tailing	870,000	0.08	0.11	0.57	0.4	49	7.0	6.0	6.0	24.0	20.0

Three parameters were evaluated for each process option; technical viability, economic merit and environmental issues. Preference was given to processes which are in commercial operation but at the same time, alternate processes which have the potential to solve particular technical problems or which may offer substantial savings in costs were also considered. Following a wide ranging discussion, during which many process alternatives were considered, nine basic process alternatives were selected for further evaluation.

Assumptions used in this study include:

Metal Prices:	Copper (US$/lb.)	0.80
	Zinc (US$/lb.)	0.55
	Lead (US$/lb.)	0.30
	Gold (US$/oz)	300
	Silver (US$/oz)	5.50
Power Cost:	US$/kWh	0.056
Capital Construction (year):		2002
Construction Indirects:		50%
Personnel		Fly In/Out

For the economic evaluation it has been assumed that electric power would be available for purchase at a cost of 5.6 cents US per kwh. While this source of power does not currently exist in the Yukon it has been accepted that the projected demand for the current project over a period of 15 years or more would be sufficient for a third party to build a coal fired power plant in the region.

Since many of the processes involve neutralization of acid, the availability of limestone was given particular attention as part of infrastructure considerations. Detailed engineering information on sources of limestone in the area was not developed, and so the cost of this important process consumable is an important subject for sensitivity evaluation.

While it is recognized that the concentrates contain significant concentrations of arsenic and antimony, no special provisions have been made for their removal in the process flowsheets since impurity control was considered to be a basic concern for all processes. Similarly no special provision was made in the costing for impurity control beyond basic neutralization as the level of costing being conducted did not warrant greater detail.

It has been assumed that the disposal of lead sulphate formed from the treatment of concentrates containing galena will be acceptable and that elemental sulphur and/or gypsum containing selenium is also acceptable for disposal.

1.3 DATABASE DESCRIPTION

This study is characterized as an "Order of Magnitude" phase document. This level of study is largely used for pre-screening of projects prior to significant activity on the property, or to compare or short-list basic development alternatives. Estimates are prepared during these studies largely by factoring previous project cost data according to the respective capacities of the projects. Such estimates are often referred to as ± 30% accurate, prior to application of contingency. Contingencies typically range from 15 to 30% for order of magnitude estimates. Deliverables are limited to a brief scope and project description, highly qualified geologic reserve indications, no metallurgical testwork, a general location drawing, and possibly a conceptual or typical flowsheet.

The capital and operating costs which are presented in this study are for hypothetical mine and process plants considered to have possible application for the Wolverine property. The reserves to justify the assumptions used here have not yet been developed. No mining study has been completed to provide a basis for mining costs which are excluded. There are no test results to provide a basis for the concentrate processing option estimates from a combined mill feed.

The process cost estimates given in this study are based on sufficient engineering to develop preliminary heat and material flows for the major process operations. These were used with the team's professional judgement to estimate approximate sizes and costs from experience and historical data. Where information was available on a process unit cost, it was factored and modified to fit the current situation. Detailed equipment lists were not developed, and no quotes were obtained (either for equipment or for consumables) for this study. We have attempted to bring all costs to the same basis so that comparisons will be valid; however, the estimates were developed individually by a team of professionals who are not all AGRA Simons employees, so that discrepancies in methodology and basis may exist.

A very important component of the cost structure is electrical power. Both the energy cost and the capital costs for delivering the power to the site were provided by the client. AGRA Simons has not reviewed the basis for these costs and cannot express an opinion on their validity.

The division of responsibilities for the work reported here is generally as follows:

Dr. Morris Beattie

- Testwork review, descriptive text and material balances for production of the flotation concentrates
- Investigation of the potential for intensive cyanidation

Dr. David Dreisinger

- Flowsheets, design criteria, descriptive text, heat and mass balances, sizing and costing for bulk concentrate, copper/lead concentrate, and zinc concentrate processing by total pressure oxidation or mid-temperature pressure oxidation, zinc and copper solvent extraction and electrowinning, and gold and silver recovery by CIP.
- Flowsheet, descriptive text and mass balance for zinc concentrate roasting

Brad Marchant

- Study Manager
- Flowsheet, design criteria, and descriptive text for the bioleaching of bulk concentrate

Usibelli Coal Mine Inc.

- Power costs for a 50MW coal fired power plant

Ian Hayward International Ltd.

- Power transmission costs from the coal fired plant to the mine site, including substations and upgrades to existing lines.

International Metallurgical and Environmental

- Flotation report

Advanced Mineral Technology Laboratory

- Report on flotation and mineralogical analysis

AGRA Simons

- Infrastructure descriptive text and costs except power supply
- Flotation concentrator capital and operating cost estimate
- Bioleaching heat and mass balances for bulk concentrate and Cu-Pb concentrate, sizing and costing for capital and operating cost estimates
- Zinc concentrate roasting sizing and costing for capital and operating cost estimates
- Coordination of the cost estimates
- Assembly of the study report.

1.4 SAMPLE AVAILABILITY

During the course of the diamond drilling of the Wolverine Deposit, Westmin Resources Limited retained all coarse rejects from assay samples of the ore intersections and had Chemex Labs store them in pails with the air removed and nitrogen added. As of November 1999, approximately 584 kg of sample remains and are available from drill cores of the 1996 and 1997 drilling. From this material it should be possible to composite representative samples from the Wolverine deposit.

Note that all previous Wolverine testwork was conducted on similar drill core reject material. Sample availability for KZK is not known.

1.5 GENERAL INFRASTRUCTURE

The general infrastructure includes necessary facilities to support the development and operation of the process plant.

Capital costs developed are based on previous work done in the same region of the north.

A summary of the infrastructure capital cost is given in Table 1.4 below. Detailed breakdowns are given in Section 2.5.

Table 1.4 Summary of Infrastructure Capital Costs

	Process Option								
	I	II	III	IV	V	VI	VII	VIII	IX
Capital Costs (US$1000)	76,885	75,295	75,993	75,526	75,993	69,484	69,484	56,958	54,958

2.0 PROCESS ALTERNATIVES

Concentrate Marketing

Based on the predicted grades of zinc and copper concentrates from the combined mill feed it was decided that additional concentrate marketing was warranted. The combined mill feed can take advantage of higher metal grades from Wolverine and lower potential penalty elements (Se and Sb) from Kudz Ze Kayah. There is no market for a combined copper-lead concentrate. The lead concentrate was not included in the marketing study. In addition, alternatives for processing at site, or at the smelter, to remove some of the potential penalty elements have been addressed in the marketing study, including:

- Roasting the zinc concentrate and ship the calcine to the smelter.
- Bioleaching the copper concentrate to remove diluents associated with various sulpho-salt and lead minerals.

The marketing study has initially been restricted to smelters in Japan, due to favorable shipping costs and capacity for selenium and antimony recovery/removal, see Table 2.1 and 2.2. A comparison of the analysis of zinc concentrate versus a hypothetical zinc calcine product is included in Table 2.3 and a hypothetical comparison of a copper concentrate before and after bioleaching is included in Table 2.4. These are the possible products that are being studied in the current market analysis.

Table 2.1 Concentrate Values – FOB Minesite (US$)

Zinc price	0.55 US$/lb
Silver price	5.75 US$/oz
Gold price	325 US$/oz

					(US$/t)	
Zinc Concentrate				**Trail**	**Noranda**	**Japan**
Payables:						
	Assay	%Payable				
Zn (%)	52	0.85		536	536	536
Ag (g/t)	150	0.65		7	7	7
Au (g/t)	1	0.65		0	0	0
			Subtotal	**543**	**543**	**543**
Treatment Charge:				**185**	**185**	**185**
penalties				0	0	0
Shipping:						
Inland - Skagway				47	47	47
Loading				15	15	15
Ocean				20	60	25
Off-load				6	6	0
Inland - Smelter				20	10	0
			Subtotal	**108**	**138**	**87**
FOB Value at Minesite				**250**	**220**	**271**

Table 2.2 Concentrate Values - FOB Minesite (US$)

Copper price	0.8	US$/lb
Silver price	5.75	US$/oz
Gold price	325	US$/oz

						(US$/t)	
Copper Concentrate					**Trail**	**Noranda**	**Japan**
	Payables:						
		Assay	Payable				
	Cu (%)	25	95%		419	419	419
	Ag (g/t)	2800	95%		476	476	476
	Au (g/t)	16	95%		144	144	144
				Subtotal	**1039**	**1039**	**1039**
	Treatment Charge:				**95**	**95**	**95**
		penalties			0	0	0
	Shipping:						
		Inland - Skagway			47	47	47
		Loading			15	15	15
		Ocean			20	60	25
		Off-load			6	6	0
		Inland - Smelter			20	10	0
				Subtotal	**108**	**138**	**87**
	FOB Value at Minesite				**836**	**806**	**857**

Table 2.3 Zinc Concentrate Options - Market Study

	Concentrate Options					
Assay (%)	**As Is**	**Leached I**	**Leached 2**	**Calcine**	**Calcine Leach**	**Comments**
Cu	0.5			0.58		
Pb	1.5			1.73		
Zn	52.1			59.7		
Au(g/t)	1.0			1.155		
Ag(g/t)	150			173.3		
Fe	10.1			11.7		
S	32.1			0.74		
Se	0.086			0.00199		
As	0.132			0.00305		
Sb	0.03			0.00104		
Hg(g/t)	0.013			0		
Carbon	0.33			0		
Ni	<0.01			<0.01		
Bi	<0.01			<0.01		
Cd	0.403			0.465		
Co	<0.01			<0.01		
Cr	<0.01			<0.01		
Sn	<0.01			<0.01		
F	0.01			0.00231		
Cl	0.0042			0.000097		
Te	<0.01			<0.01		
Ba	NA			NA		
Ge	<0.01			<0.01		
SiO2	0.07			0.081		
MgO	0.03			0.035		
CaO	NA			NA		
Insol	2.91			3.36		
Mn	0.01			0.012		
Al2O3	<0.01			<0.01		
Oxygen				21.55		
Total	**100**			**100**		
Tonnes/a	**140,000**			**121,000**		

Table 2.4 Copper Concentrate Options - Market Study

	Concentrate Options				
Assay (%)	**As Is**	**Leached**	**Calcine**	**Calcine Leach**	**Comments**
Cu	24.7	26.0			
Pb	2.03	2.1			
Zn	4.26	4.5			
Au(g/t)	16.2	17.1			
Ag(g/t)	2794	2941			
Fe	28.75	26.5			
S	35.3	32.0			
Se	0.18	0.08			
As	0.13	0.03			
Sb	0.31	0.10			
Hg(g/t)	20	20			
Carbon	NA	NA			
Ni	<0.01	<0.01			
Bi	0.1	0.1			
Cd	0.02	0.02			
Co	<0.01	<0.01			
Cr	0.03	0.03			
Sn	0.01	0.01			
F	0.0075	0.0075			
Cl	0.005	0.005			
Te	<0.01	<0.01			
Ba	NA	NA			
Ge	<0.01	<0.01			
SiO2	0.1125	0.12			
MgO	<0.01	<0.01			
CaO	NA	NA			
Insol	3.25	3.4			
Mn	<0.01	<0.01			
Al2O3	<0.01	<0.01			
Oxygen					
Total	**100**	**100**			
Tonnes/a	**33,000**	**31,500**			

The marketing study is in progress and will be reported under separate cover.

Metallurgical Processing at Site

The primary objective of this study is to evaluate potential metallurgical processing routes that maximize payable metals from the two deposits. Most of the processing alternatives studied included hydrometallurgial processing at site to produce refined copper metal and zinc metal products as well as silver/gold doré product. The lead was not considered for further processing.

Description of Process Alternatives

Process alternatives selected for this scoping study were based on a constant mine plan from a combination of open pit mining (1875 tpd) and underground mining (1,125 tpd) from two distinct deposits to make up 3,000 tpd mill feed. The ore is processed by conventional crushing and grinding and then further processed in one of two optional flotation plants:

- Bulk flotation of the copper, lead and zinc minerals
- Selective flotation of a copper-lead concentrate and a separate zinc concentrate

The flotation plant would be operated to maximize metal recovery at the expense of concentrate grade in each case.

The flotation concentrates would then be processed in one of the following flowsheet alternatives for metal recovery:

Option1:	Bulk concentrate with total sulphide oxidation in an autoclave followed by separate Cu, Zn, Ag/Au recovery
Option 2:	Bulk concentrate with low temperature oxidation (Dynatec Process) in an autoclave followed by separate Au, Zn and Ag/Au recovery.
Option 3:	Bulk concentrate oxidation in a two stage bioleach circuit followed by separate Cu, Zn and Ag/Au recovery.
Option 4:	Zinc concentrate with total sulphide oxidation in an autoclave; Cu-Pb concentrate oxidation in a low temperature autoclave (Dynatec Process) with separate Cu, Zn and Ag/Au recovery.
Option 5:	Zinc concentrate with total suphide oxidation in an autoclave: Cu-Pb concentrate with total sulphide oxidation in an autoclave with separate Cu, Zn and Ag/Au recovery.

Option 6: Zinc concentrate roasting and shipping to a smelter; Cu-Pb oxidation in a low temperature autoclave (Dynatec Process) followed by separate Cu, Zn and Ag/Au recovery.

Option 7: Zinc concentrate roasting and shipping to a smelter; Cu-Pb concentrate total sulphide oxidation in an autoclave followed by separate Cu, Zn and Ag/Au recovery.

Option 8: Zinc concentrate roasting and shipping to a smelter; partial oxidation of the Cu-Pb concentrate in a single stage bioleach followed by Ag/Au recovery only by conventional cyanidation.

Option 9: Zinc concentrate roasting and shipping to a smelter; partial oxidation of a copper flotation concentrate in a single stage bioleach to remove mineral sulpho-salts followed by shipping to a smelter.

None of the process alternatives included lead recovery.

A detailed description of each flowsheet is included in the report was well as metallurgical and energy balance estimates. Each flowsheet alternative included all process steps, metal recovery steps and waste treatment.

2.1 CRUSH, GRIND, FLOTATION

2.1.1 Bulk Flotation

2.1.1.1 Process Description

Wolverine is a classic volcanogenic massive sulphide deposit hosted in felsic volcanic and argillaceous sedimentary rocks. The sulphide mineralogy is typical of such deposits with copper occurring predominantly as chalcopyrite, lead as galena and lead sulphosalts and zinc as sphalerite. There is an abundance of pyrite associated with the sulphides in the deposit so that the average feed will contain in the order of 35% sulphide sulphur. Pyrrhotite, arsenopyrite, marcasite and sulphosalts such as tetrahedrite are present in small amounts. The tetrahedrite is the major carrier of silver in Wolverine, containing about 19% (w/w) silver. The sphalerite is iron-bearing, with an average iron content of 6.7% by weight. An unusual feature of Wolverine is the very high content of selenium resulting from the substitution of this element for sulphur in the sulphide minerals. Any processing strategies developed for the deposit will have to address the issue of selenium deportment through the process.

Flotation testwork has been conducted on samples of drill core from Wolverine by International Metallurgical and Environmental Inc. (IME) of Kelowna, BC and by Advanced Mineral Technology Laboratory (Amtel) of London, Ontario. The results of those test programs have been used as the basis for predicting the flotation requirements and results.

Cominco's Kudz Ze Kayah deposit is mineralogically similar to Wolverine. Although this deposit also contains elevated selenium concentrations, the concentrations are not as high as those for Wolverine. Metallurgical testwork on Kudz Ze Kayah has been conducted by several commercial laboratories as well as by Cominco. The assumed flotation response for this deposit is based on information supplied by Cominco.

2.1.1.2 Flowsheet

The metallurgical testwork conducted by Amtel used composite material derived from drilling conducted during 1997. The work was directed primarily at the production of selective copper, lead and zinc concentrates. The predicted metallurgical results based on this work are summarized in Table 2.5. The study concluded that a moderately coarse primary grind of 70% minus 200 mesh would be adequate for the rougher flotation separation, followed by fine regrinding to 80% passing 25 microns before cleaning. Since it has subsequently been determined that there is no merit in producing a separate lead concentrate and most of the tests did not produce a bulk copper-lead concentrate prior to separation, this test program has limited use for the present study. The study does confirm that reasonably high recoveries of the metals of interest can be expected and that most of the gold and silver which is recovered will report to the copper-lead product. The testwork also confirms that a good grade zinc concentrate can be produced at a recovery near 90%.

Table 2.5 Summary of Predicted Metallurgy Based on Amtel Testwork

		Assays, %, g/t					% Distribution				
Product	**WT. %**	**Cu**	**Pb**	**Zn**	**Au**	**Ag**	**Cu**	**Pb**	**Zn**	**Au**	**Ag**
Cu Conc.	4	21.8	3.9	3.2	7.8	4,068	75	11	1	20	54
Pb Conc.	2	2.9	23.2	8.1	16.3	2,199	6	40	2	26	18
Zn Conc.	18	0.4	1.0	51.0	1.4	146	6	13	87	17	9
Pb Cl Sc Tail	4	0.8	1.8	9.5	1.2	267	3	6	4	4	4
Zn Comb Tail	72	0.2	0.6	0.9	0.7	60	10	30	6	33	15
Final Tail	**76**						**13**	**36**	**10**	**37**	**19**
Head		**1.1**	**1.4**	**10.4**	**1.5**	**290**					

The IME test program was conducted on a composite sample of assay rejects from a 1995 drill program. The head assay for this sample was comparable to that of the composite used by Amtel in terms of base metals and was somewhat higher on gold and silver content as follows:

Cu, %	1.06
Pb, %	1.43
Zn, %	11.6
Au, g/t	2.13
Ag, g/t	454

Gravity concentration testwork was conducted by IME and it was determined that up to 15% of the gold could be recovered to a gravity concentrate, although visual observations indicated that the gold being recovered was very fine. Although this level of recovery was judged to be too low to justify further development of a gravity concentration circuit, the results do provide an indication that at least a portion of the gold is present in its free state and may lend itself to cyanidation of a concentrate.

The test program was directed at both concurrent copper-lead followed by zinc flotation and sequential copper, lead and zinc flotation. For the present study, only the work involving a combined copper-lead product has been considered. Both batch tests and locked cycle tests were conducted in this program. For the prediction of expected metallurgical performance in the present study, the results of batch test number 106 were used and a prediction was made of the expected commercial results. This test had been continued through copper-lead separation and the bulk Cu-Pb concentrate grade was calculated from the various test products. The test procedure included a copper-lead scavenger which produced a product having a grade similar to the head grade. This product has been treated as new feed and the results normalized to 100% in order to simplify the balance. The normalized results for Test 106 are summarized in Table 2.6 and the predicted metallurgy derived from this test is summarized in Table 2.7. A locked cycle test was conducted using a similar procedure as for Test 106 but did not achieve comparable results in any of the cycles. This test was conducted on a separate composite prepared from the same assay rejects. It is postulated that the two composites were not identical or the samples may have experienced some superficial oxidation to account for the difference in response. Additional flotation work is clearly required to confirm that the projected results are achievable.

Table 2.7 includes data for a bulk Cu-Pb-Zn concentrate obtained by simply combining the Cu-Pb and zinc concentrates. It is recognized that the recovery for a bulk concentrate should be somewhat greater than obtained simply by adding the other products together but there are no data available to support the magnitude of such an increase. The reagents used for test 106 are summarized in Table 2.8. The reagent additions have not been optimized.

Table 2.6 Results of IME Flotation Test 106

		Assays, %, g/t					% Distribution				
Product	**WT. %**	**Cu**	**Pb**	**Zn**	**Au**	**Ag**	**Cu**	**Pb**	**Zn**	**Au**	**Ag**
Cu/Pb Conc.	7.9	12.1	12.2	9.3	14.8	4,509	85.5	68.5	6.4	55.9	79.8
Cu/Pb 2nd Cl T	1.7	1.25	2.5	11	9.5	690	1.9	3.0	1.6	7.7	2.6
Cu/Pb 1st Cl T	6.1	0.54	1.6	11.3	2.7	335	3.0	6.9	6.0	7.9	4.6
Cu/Pb Ro Con	15.8	6.42	7.0	10.3	9.5	2,474	90.4	78.4	13.9	5.1	4.7
Zn Conc.	14.3	0.16	0.27	53.8	0.7	75	2.0	2.7	66.3	4.8	2.4
Zn 3rd Cl T	2.5	0.28	0.53	39.5	0.9	120	0.6	0.9	8.6	1.1	0.7
Zn 2nd Cl T	2.1	0.31	0.72	21.3	2.4	130	0.6	1.1	3.9	2.4	0.6
Zn 1st Cl T	5.9	0.2	0.64	5.8	1.2	110	1.1	2.7	3.0	3.4	1.5
Final Tail	**59.3**	**0.1**	**0.34**	**0.84**	**0.6**	**60**	**5.3**	**14.2**	**4.3**	**16.9**	**7.9**
Head		**1.12**	**1.42**	**11.6**	**2.1**	**449**					

Table 2.7 Predicted Flotation Results for Wolverine

		Assays, %, g/t					% Distribution				
Product	**WT. %**	**Cu**	**Pb**	**Zn**	**Au**	**Ag**	**Cu**	**Pb**	**Zn**	**Au**	**Ag**
Bulk Cu/Pb and Zn Concentrates											
Cu/Pb Conc.	7.9	12	11.8	9.3	11.5	3,205	89.7	75.3	7.3	64.6	85.5
Zn Conc.	17.3	0.4	1	51	1.0	150	6.5	13.9	87.0	12.3	8.7
Tailing	74.8	0.05	0.18	0.78	0.4	23	3.8	10.8	5.7	23.1	5.8
Bulk Cu/Pb/Zn Concentrates											
Zn Conc.	20.5	5.0	5.4	46.5	5.3	1,362	96.2	89.2	94.3	76.9	94.2
Final Tail	**79.5**	**0.05**	**0.17**	**0.73**	**0.4**	**22**	**3.8**	**10.8**	**5.7**	**23.1**	**5.8**
Head		**1.06**	**1.2**	**10.1**	**1.41**	**297**					

Table 2.8 Reagent Additions for Flotation of Cu-Pb and Zn Concentrates from Wolverine

Mineral	Wt. %
$ZnSO_4$	75
Na_2SO_3	1,485
A303	110
3418A	14
Lime	2,080
$CuSO_4$	450
PAX	110
MIBC	81
$Na_2S_2O_5$	1,865
NaCN	100

A metallurgical prediction for the combined treatment of Wolverine and Kudz Ze Kayah (KZK) was previously provided by Cominco, based on the testwork they had conducted on KZK. A revised prediction for the treatment of a combined mill feed treating 3/8 part by weight Wolverine and 5/8 parts KZK is included as Table 2.9. This revised prediction takes into account the higher copper and silver extraction obtained for the Wolverine material.

Table 2.9 Predicted Flotation Results for 3/8 Wolverine, 5/8 KZK Feed

		Assays, %, g/t					% Distribution				
Product	**WT. %**	**Cu**	**Pb**	**Zn**	**Au**	**Ag**	**Cu**	**Pb**	**Zn**	**Au**	**Ag**
Bulk Cu/Pb and Zn Concentrates											
Cu/Pb Conc.	6.9	12	15.8	6.0	12.5	2,122	86.8	78.2	5.6	64.9	75.8
Zn Conc.	12.7	0.5	1.5	51.7	1.0	150	6.6	13.6	87.6	9.5	9.8
Tailing	80.4	0.08	0.14	0.64	0.4	35	6.6	8.2	6.8	25.6	14.4
Bulk Cu/Pb/Zn Concentrates											
Zn Conc.	20.5	4.4	6.4	34.3	5.0	810	93.4	94.0	94.0	76.0	85.6
Final Tail	**79.5**	**0.08**	**0.11**	**0.57**	**0.4**	**35**	**6.6**	**6.0**	**6.0**	**24.0**	**14.4**
Head		**0.96**	**1.4**	**7.49**	**1.34**	**194**					

A bond work index determination was made by IME for Wolverine using a classification size of 200 mesh and gave a result of 10.4 (metric).

The impurity concentrations obtained during testwork for the two deposits are summarized in Table 2.10. Since these concentrations are for concentrates having different major metals contents than those predicted for the project, these impurity values can be used as an indication only of typical impurity contents to be experienced.

Table 2.10 Impurity Contents of Selected Test Products

	Cu Conc		**Pb Conc**		**Zn Conc**	
•	**W**	**KZK**	**W**	**KZK**	**W**	**KZK**
Pb	3.9	0.9	23.2	52.2	1.0	1.6
Zn	3.2	4.9	8.1	5.4	51.0	52.1
Cu	21.8	26.5	3.9	0.2	0.4	0.6
Fe		28		14.7	9.7	10.4
Ni		<.01		0.01		<.01
Bi		0.09		0.08		<.005
Cd		0.02		0.02		0.35
Co		<.01		<.01		<.01
Cr		.03		<.01		<.01
As	0.042	0.18	0.049	.27	0.052	0.18
Sb	0.78	0.03	0.58	.21	0.030	0.03
Sn		0.01		<.01		<.01
F	0.02		0.02		0.01	
Cl	0.007		0.002		0.004	
S	35.8		35.1		35.2	30.3
Hg	15	23	22	20	120	130
Ag	4,040	2,047	2,199	665	146	76
Au	7.76	21.3	16.3	1.5	1.43	2.5
Se	0.41	0.04	1.65	0.25	0.18	0.03
Te		<.01		<.01		<.01
Ba						
Ge		<.01		.01		<.01
C						
SiO_2		0.12		0.1		0.07
MgO		<.01		.01		0.03
CaO						
Insol	3.25		4.08			
Mn		<.01		<.01		0.01
Al_2O_3		<.01		<.01		<0.01

* All values are % except Hg =ppm and Au, Ag are g/t

2.1.1.3 Basic Design Criteria

A schematic flowsheet for the flotation circuit is shown in Figure 2.1. The feed to flotation is crushed prior to being ground to 70% minus 200 mesh. The required grind has yet to be optimized as some of the testwork has indicated that improved overall results may be achieved with a finer primary grind. Grinding is carried out at natural pH with additions of zinc sulphate and sodium sulphite for zinc and pyrite depression. The copper and lead are activated with additions of A303 and 3418A collectors prior to bulk copper-lead flotation. The bulk copper-lead concentrate is cleaned twice with additions of zinc sulphate and A303 to produce a concentrate for dewatering.

The tailing from the copper-lead rougher is conditioned in two stages with lime to pH11.5, copper sulphate and PAX to activate the zinc mineral while depressing pyrite. The zinc rougher concentrate is reground to 80% passing 30 microns before being cleaned twice with lime and copper sulphate addition to produce a final concentrate for dewatering. The final concentrate grade is limited by the presence of simple pyrite-sphalerite middlings. Finer regrinding can be expected to improve the zinc concentrate grade with minimum loss in recovery. Preliminary testwork has indicated that it may be possible to divert a part of the zinc rougher concentrate directly to the second cleaning stage thereby minimizing regrinding costs. This alternative will have to be explored more fully during subsequent testing before it can be included in the flowsheet.

Flotation retention times of 12 minutes for the Cu/Pb rougher and 6 minutes for the zinc rougher were used for the batch testwork. The flotation time for each of the stages of cleaning was 4 minutes.

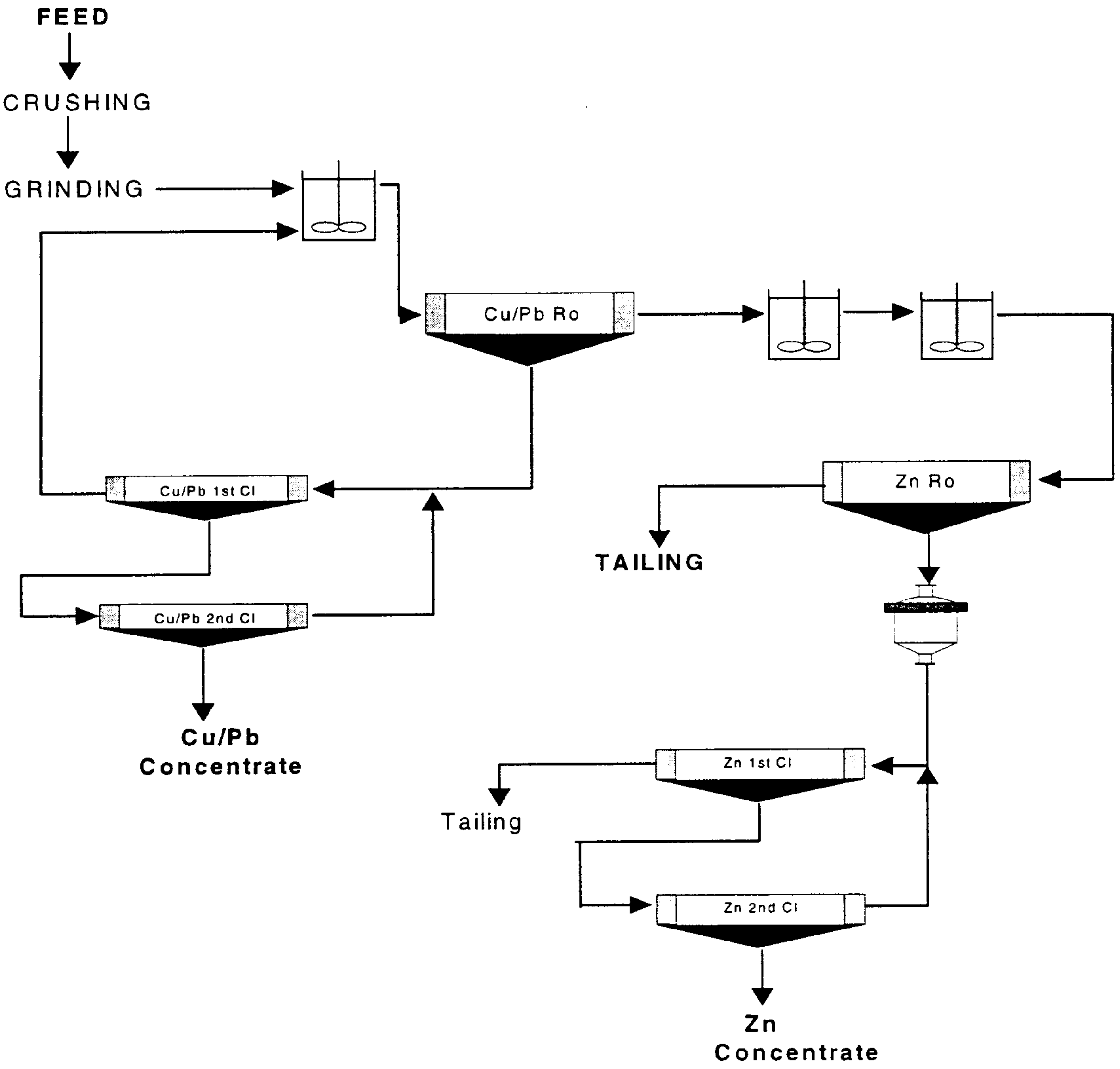

Figure 2.1 Flotation Flowsheet for Cu/Pb and Zn Concentrates

2.1.1.4 Capital and Operating Costs

Capital Costs

Capital costs for the mill and flotation plants handling 3,000 tonnes/day ore are given in Table 2.11 below.

Table 2.11 Mill/Flotation Capital Costs

	Bulk Concentrate (US$1000)	Two Concentrates (US$1000)
Primary crusher and concentrator buildings	4,700	5,200
Buildings at Wolverine (shop, warehouse, dry, office)	2,200	2,200
Equipment	9,930	10,480
Piping, E & I	4,200	4,400
Total Direct Cost	**21,030**	**22,280**
Indirects (50%)	10,515	11,140
Contingency (20%)	6,309	6,684
Total Cost (US$1000)	**37,854**	**40,104**

Operating Costs

Operating costs for the mill and flotation plants are given in Table 2.12 below.

Table 2.12 Mill/Flotation Operating Costs

	Bulk Concentrate		Two Concentrates	
	No.	(US$1000)	No.	(US$1000)
Operating labour	33	1,405.9	34.5	1,470.7
Maintenance labour	23.5	1,156.2	23.5	1,156.2
Subtotal labour	**2,562.1**		**2,626.9**	
Fringe benefits @ 35%	896.7		919.4	
Total labour	**3,458.1**		**3,546.3**	
Operating supplies	1,936.0		2,146.2	
Maintenance supplies	1,010.0		1,112.0	
Total supplies	**2,946.0**		**3,258.0**	
Power cost	1,456.0		1,736.0	
Total cost (US$1000)	**7,860.1**		**8,540.3**	
Unit cost ($/t ore)	**7.18**		**7.80**	

2.1.1.5 Environmental Considerations

The flotation processes treat a massive sulphide feed, and so the minerals rejected to tails may be expected to have a high sulphur content with little alkalinity. The tails will therefore be subject to oxidation if simultaneously wet and exposed to air. To prevent formation of acid rock drainage, it will be necessary to ensure that the tailings are permanently placed under water where access to air is prevented. It is anticipated that selenium in the tails will behave like sulphur: if access to air is prevented, it should mobilize only very slowly if at all. Thus, closure will need to address the permanence of water cover for the tailings deposit.

2.1.1.6 Testwork Required for Prefeasibility

All previous testwork conducted for the project was directed at the production of selective copper, lead and zinc concentrates. While two locked cycle tests were included in the test program, these did not achieve stable results and more importantly did not produce the same results in the first cycle as had been obtained in previous batch tests. Further locked cycle tests should only be conducted once consistent batch test results are achieved. Additional testwork is required to optimize the conditions for the production of a bulk copper-lead concentrate with a primary objective to maximize precious metals recovery. Regrinding of the Cu/Pb cleaner tails prior to returning them to the rougher circuit will have to be considered fully during this testwork.

If the results of the scoping study indicate the preferred route for secondary processing to be the treatment of a bulk Cu/Pb/Zn concentrate, testwork will have to be conducted on this alternative since this has not been done to date. Both batch and locked cycle testing of this option will be required although the program is expected to be less comprehensive than for the two concentrate alternative since less stages of reagent optimization are required.

A side benefit of the flotation testing will be the production of concentrates for testing of the secondary processing alternatives. As much as possible, the production of concentrates for subsequent testing should be done in a manner that will confirm conditions and requirements for the flotation separation.

There presently exists about 580 kg of drill core rejects at Chemex, mostly stored under nitrogen. This material is considered to be suitable for the additional flotation testwork that is required. In addition, some samples of feed and test products remain at International Metallurgical.

2.2 Bulk Concentrate Alternatives

The bulk concentrate represents a flotation concentrate of Cu/Pb/Zn with associated gold and silver content. The concentrate was analyzed and divided into an assumed elemental and mineralogical distribution.

Table 2.13 Elemental Distribution

	Tonnage	Assays (%, g/t)						
Product	**(tpa)**	**Cu**	**Pb**	**Zn**	**Au**	**Ag**	**S(total)**	**Fe(total)**
Cu/Pb/Zn Conc	225,000	4.4	6.4	34.3	5	810	35	16.70

Table 2.14 Mineralogical Distribution

Mineral	Wt. %
ZnS	51.12
FeS	5.43
PbS	7.39
$CuFeS_2$	12.71
FeS_2	20.15
Au	0.00050
Ag	0.0810
SiO_2	3.20
Total	**100.00**

2.2.1 Total Oxidation Autoclave/CIP

2.2.1.1 Process Description

The total pressure oxidation process is designed to totally oxidize all sulfide sulfur to sulfate sulfur, extract copper and zinc into solution, precipitate lead as lead sulfate and liberate the precious metals (Au, Ag) for subsequent cyanidation recovery.

The process concept uses a direct autoclave leach of the reground concentrate at 220°C followed by flashing, slurry cooling, acid neutralization and solid liquid separation. The recovered residue is sent to cyanidation for precious metal recovery after a lime boil treatment. The recovered autoclave leach liquor would then be directed to copper solvent extraction and electrowinning, neutralization and iron removal and finally zinc solvent extraction and electrowinning. A bleed stream would be taken out for neutralization and recycle/disposal.

Process Chemistry

The oxygen pressure leaching of reground concentrate will take place at 220°C. The main products of leaching will be dissolved base metals, lead sulfate precipitate and hematite. It is expected that gold will be liberated in the product solids. Silver will likely form argentojarosite and require a hot lime boil to decompose to the cyanide soluble silver hydroxide species. The autoclave discharge is expected to contain some free acid and therefore will require neutralization prior to direct solvent extraction of copper. Copper will be extracted using a conventional oxime solvent extractant followed by stripping in spent electrolyte and electrowinning of LME Grade 1 copper cathode. The copper raffinate will go to iron removal and neutralization to prepare an iron free feed for zinc solvent extraction. Zinc will then be recovered using D2EHPA as an extractant. Zinc is stripped with spent electrolyte followed by acid sulfate electrowinning of Special High Grade zinc. The zinc raffinate is recycled to leaching (providing an acid source for leaching) except for a bleed stream to maintain water balance.

Leaching

Sphalerite Leaching: $ZnS + 0.5O_2 + H_2SO_4 \rightarrow ZnSO_4 + S + H_2O$
Chalcopyrite Leaching: $CuFeS_2 + O_2 + 2H_2SO_4 \rightarrow CuSO_4 + FeSO_4 + S + H_2O$
Galena Leaching: $PbS + 0.5O_2 + H_2SO_4 \rightarrow PbSO_4(s) + S + H_2O$
Pyrrhotite/Marmatite Leaching: $FeS + 0.5O_2 + H_2SO_4 \rightarrow FeSO_4 + S + H_2O$
Pyrite Leaching: $FeS_2 + 0.5O_2 + H_2SO_4 \rightarrow FeSO_4 + 2S + H_2O$
Ferrous Sulfate Oxidation: $FeSO_4 + 0.25O_2 + 0.5H_2SO_4 \rightarrow 0.5Fe_2(SO_4)_3 + 0.5H_2O$
Elemental Sulfur Oxidation: $S + 1.5O_2 + H_2O \rightarrow H_2SO_4$
Hematite Precipitation: $Fe_2(SO_4)_3 + 3H_2O \rightarrow Fe_2O_3 + 3H_2SO_4$

Lime Boil

Argentojarosite Decomposition:

$$AgFe_3(SO_4)_2(OH)_6 + 2Ca(OH)_2 \rightarrow AgOH + 2.5Fe_2O_3 + 2CaSO_4.2H_2O + 0.5H_2O$$

Cyanidation

Silver Leaching: $AgOH + 2NaCN \rightarrow NaAg(CN)_2 + NaOH$
Gold Leaching: $Au + 0.25O_2 + 2NaCN + 0.5H_2O \rightarrow NaAu(CN)_2 + NaOH$

Copper Solvent Extraction and Electrowinning

Copper Extraction: $CuSO_4 + 2\overline{HR} \rightarrow \overline{CuR_2} + H_2SO_4$

Copper Stripping: $\overline{CuR_2} + H_2SO_4 \rightarrow CuSO_4 + 2\overline{HR}$

Copper Electrowinning: $CuSO_4 + H_2O \rightarrow Cu + H_2SO_4 + 0.5O_2$

Neutralization and Iron Oxyhydrolysis

Acid Neutralization: $CaO + H_2SO_4 + H_2O \rightarrow CaSO_4.2H_2O$

Iron Precipitation: $Fe_2(SO_4)_3 + 3CaO + 7H_2O \rightarrow 2FeO(OH) + 3CaSO_4.2H_2O$

Zinc Solvent Extraction and Electrowinning

Zinc Extraction: $ZnSO_4 + 2\overline{HR} \rightarrow \overline{ZnR_2} + H_2SO_4$

Zinc Stripping: $\overline{ZnR_2} + H_2SO_4 \rightarrow ZnSO_4 + 2\overline{HR}$

Zinc Electrowinning: $ZnSO_4 + H_2O \rightarrow Zn + H_2SO_4 + 0.5O_2$

Selenium Deportment

In this flowsheet, the selenium will largely deport to the leach residue as a S-Se product or as unreacted sulfide minerals (principally pyrite). Any selenium that is oxidized in the leach will end up as either calcium precipitate (calcium selenite/selenate), lead precipitate (lead selenite/selenate) or will be incorporated in iron residues (eg. jarosites or basic ferric sulfates). Selenium will not enter the copper or zinc electrowinning circuits due to the solvent extraction barrier to its transfer. This should ensure high purity final metal products.

2.2.1.2 Flowsheet

The flowsheet is shown in Figure 2.2. The products are copper metal, zinc metal, gold and silver (impure) and various residues.

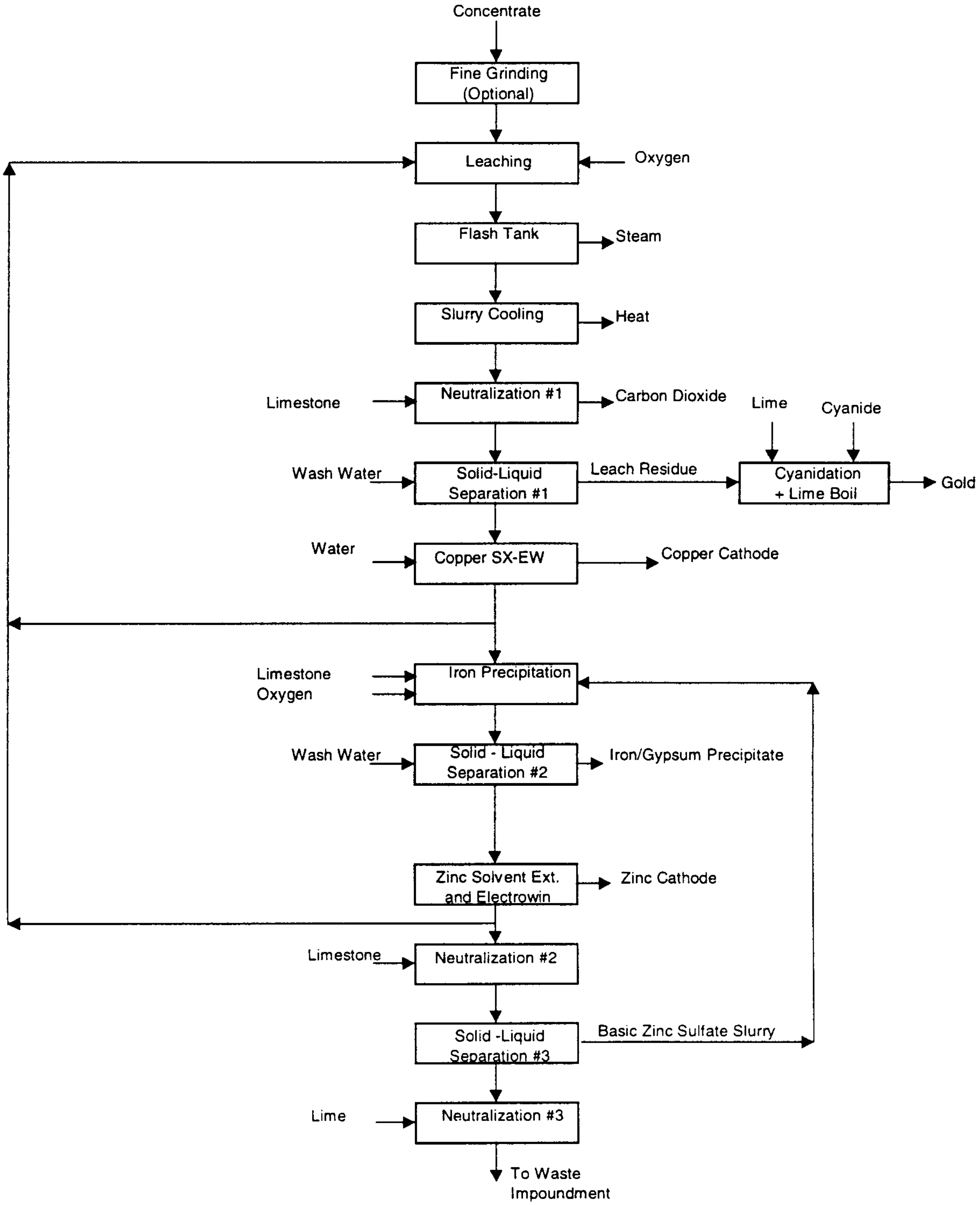

Figure 2.2 Total Pressure Oxidation Process Applied To Bulk Zn/Cu/Pb Concentrate

2.2.1.3 Basic Design Criteria

Design criteria used for total oxidation of bulk concentrate (Process Option I) is given in Table 2.15 below.

Table 2.15 Total Oxidation of Bulk Concentrate

Mass Treated		**tonnes/yr**	**225,000**
Analysis	Copper	%	4.4
	Zinc	%	34.3
	Gold	g/tonne	5
	Silver	g/tonne	810
Products Recovered	Copper	tonnes/yr	9,749
	Zinc	tonnes/yr	75,157
	Gold	oz/yr	34,365
	Silver	oz/yr	5,567,122
Overall Recovery	Copper	%	98.48
	Zinc	%	97.38
	Gold	%	95
	Silver	%	95

2.2.1.4 Capital and Operating Costs

Capital Cost

The capital cost for the total oxidation of a bulk concentrate is tabulated in Table 2.16 below.

Table 2.16 Capital Cost for Total Oxidation of a Bulk Concentrate

	Capital Cost (US$1000)
Process plant	128,445
Limestone	7,580
S° disposal	-
Effluent treatment	540
Subtotal	**136,565**
Indirects (50%)	68,283
O_2 plant	11,000
Subtotal	**215,848**
Contingency (20%)	43,169
Total	**259,017**

Operating Costs

The operating cost for the total oxidation of a bulk concentrate is tabulated in Table 2.17 below. A full breakdown is provided in Appendix A.

Table 2.17 Operating Cost for Total Oxidation of a Bulk Concentrate

	Operating Cost (US$1000)
Process plant	45,346
O_2 plant	4,901
Total	**50,247**
$/t concentrate feed	**223.32**

2.2.1.5 Environmental Considerations

Virtually all selenium will be oxidized to selenite/selenate by the autoclave process. This selenium will deport in a number of ways. First, some selenium will precipitate as a lead product during autoclave leaching. This product will likely have the same stability as lead sulfate and may well be the best way to dispose of selenium. The remaining selenium that has been oxidized will be precipitated as calcium selenite/selenate during neutralization and will become part of the solid waste in tails. The stability of this product is not known relative to the stability of gypsum.

2.2.1.6 Testwork Required for Prefeasibility

Leaching testwork will be required to demonstrate effective oxidation of the various minerals and to determine the deportment of selenium in the product and waste streams. Leaching will have to be studied as a function of time, temperature and oxygen pressure. The behaviour of selenium in acid neutralization should also be quantified. The oxidation products will need to be subjected to washing, lime boil and cyanidation testing. Key parameters will be % excess of lime for lime boil, temperature and time for lime boil and cyanidation conditions (time, pH and cyanide addition).

It will be necessary for prefeasibility to conduct a limited amount of testwork on copper and zinc SX-EW. This should consist in the preparation of standard McCabe – Thiele diagrams for copper and zinc extraction. It will not be necessary to perform any electrowinning of copper or zinc for this study.

2.2.2 Low Temperature Autoclave - Dynatec Process/CIP

2.2.2.1 Process Description

The Dynatec process uses a direct autoclave leach of the reground concentrate at 150°C with addition of low grade coal (10-50 kg/tonne addition) as a surfactant. In contrast to the total pressure oxidation process, some sulfur and selenium is left as elemental products in the residue from leaching. The autoclave discharge is flashed to atmospheric pressure, cooled via slurry coolers and then neutralized prior to copper SX-EW. The washed residue is sent to lime boil – cyanidation. The use of a lime boil for a product containing elemental sulfur has not been tested. It is believed that the sulfur will react with excess lime to produce a lot of soluble thiosalts and sulfides. If this happens, the lime boil may have to be augmented by flotation and/or hot sulfur filtration. This has not been designed into the flowsheet at this point due to the uncertainty of this result.

The recovered autoclave leach liquor would then be directed to copper solvent extraction and electrowinning, neutralization and iron removal and finally zinc extraction using zinc solvent extraction and electrowinning. A bleed stream would be taken out for neutralization and recycle/disposal.

Process Chemistry

The oxygen pressure leaching of reground concentrate will take place at 150°C. The main products of leaching will be dissolved base metals, lead sulfate precipitate, elemental sulfur, hematite and unreacted pyrite. It is expected that gold will be liberated in the product solids. Silver will likely form argentojarosite and require a hot lime boil to decompose to the cyanide soluble silver hydroxide species. The autoclave discharge is expected to be low in acid, suitable for direct solvent extraction of copper after neutralization. Copper will be extracted using a conventional oxime solvent extractant followed by stripping in spent electrolyte and electrowinning of LME Grade 1 copper cathode. The copper raffinate will go to iron removal and neutralization to prepare an iron free feed for zinc solvent extraction. Zinc will then be recovered using D2EHPA as an extractant. Zinc is stripped with spent electrolyte followed by acid sulfate electrowinning of special high grade zinc. The zinc raffinate is recycled to leaching (providing an acid source for leaching) except for a bleed stream to maintain water balance.

Leaching

Sphalerite Leaching: $ZnS + 0.5O_2 + H_2SO_4 \rightarrow ZnSO_4 + S + H_2O$

Chalcopyrite Leaching: $CuFeS_2 + O_2 + 2H_2SO_4 \rightarrow CuSO_4 + FeSO_4 + S + H_2O$

Galena Leaching: $PbS + 0.5O_2 + H_2SO_4 \rightarrow PbSO_4(s) + S + H_2O$

Pyrrhotite/Marmatite Leaching: $FeS + 0.5O_2 + H_2SO_4 \rightarrow FeSO_4 + S + H_2O$

Pyrite Leaching: $FeS_2 + 0.5O_2 + H_2SO_4 \rightarrow FeSO_4 + 2S + H_2O$

Ferrous Sulfate Oxidation: $FeSO_4 + 0.25O_2 + 0.5H_2SO_4 \rightarrow 0.5Fe_2(SO_4)_3 + 0.5H_2O$

Elemental Sulfur Oxidation: $S + 1.5O_2 + H_2O \rightarrow H_2SO_4$

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

Lime Boil

Argentojarosite Decomposition:

$$AgFe_3(SO_4)_2(OH)_6 + 2Ca(OH)_2 \rightarrow AgOH + 2.5Fe_2O_3 + 2CaSO_4.2H_2O + 0.5H_2O$$

Cyanidation

Silver Leaching: $AgOH + 2NaCN \rightarrow NaAg(CN)_2 + NaOH$

Gold Leaching: $Au + 0.25O_2 + 2NaCN + 0.5H_2O \rightarrow NaAu(CN)_2 + NaOH$

Copper Solvent Extraction and Electrowinning

Copper Extraction: $CuSO_4 + 2\overline{HR} \rightarrow \overline{CuR_2} + H_2SO_4$

Copper Stripping: $\overline{CuR_2} + H_2SO_4 \rightarrow CuSO_4 + 2\overline{HR}$

Copper Electrowinning: $CuSO_4 + H_2O \rightarrow Cu + H_2SO_4 + 0.5O_2$

Neutralization and Iron Oxyhydrolysis

Acid Neutralization: $CaO + H_2SO_4 + H_2O \rightarrow CaSO_4.2H_2O$

Iron Precipitation: $Fe_2(SO_4)_3 + 3CaO + 7H_2O \rightarrow 2FeO(OH) + 3CaSO_4.2H_2O$

Zinc Solvent Extraction and Electrowinning

Zinc Extraction: $ZnSO_4 + 2\overline{HR} \rightarrow \overline{ZnR_2} + H_2SO_4$

Zinc Stripping: $\overline{ZnR_2} + H_2SO_4 \rightarrow ZnSO_4 + 2\overline{HR}$

Zinc Electrowinning: $ZnSO_4 + H_2O \rightarrow Zn + H_2SO_4 + 0.5O_2$

2.2.2.2 Flowsheet

The flowsheet is shown in Figure 2.3. The products are copper metal, zinc metal, gold and silver (impure) and various residues.

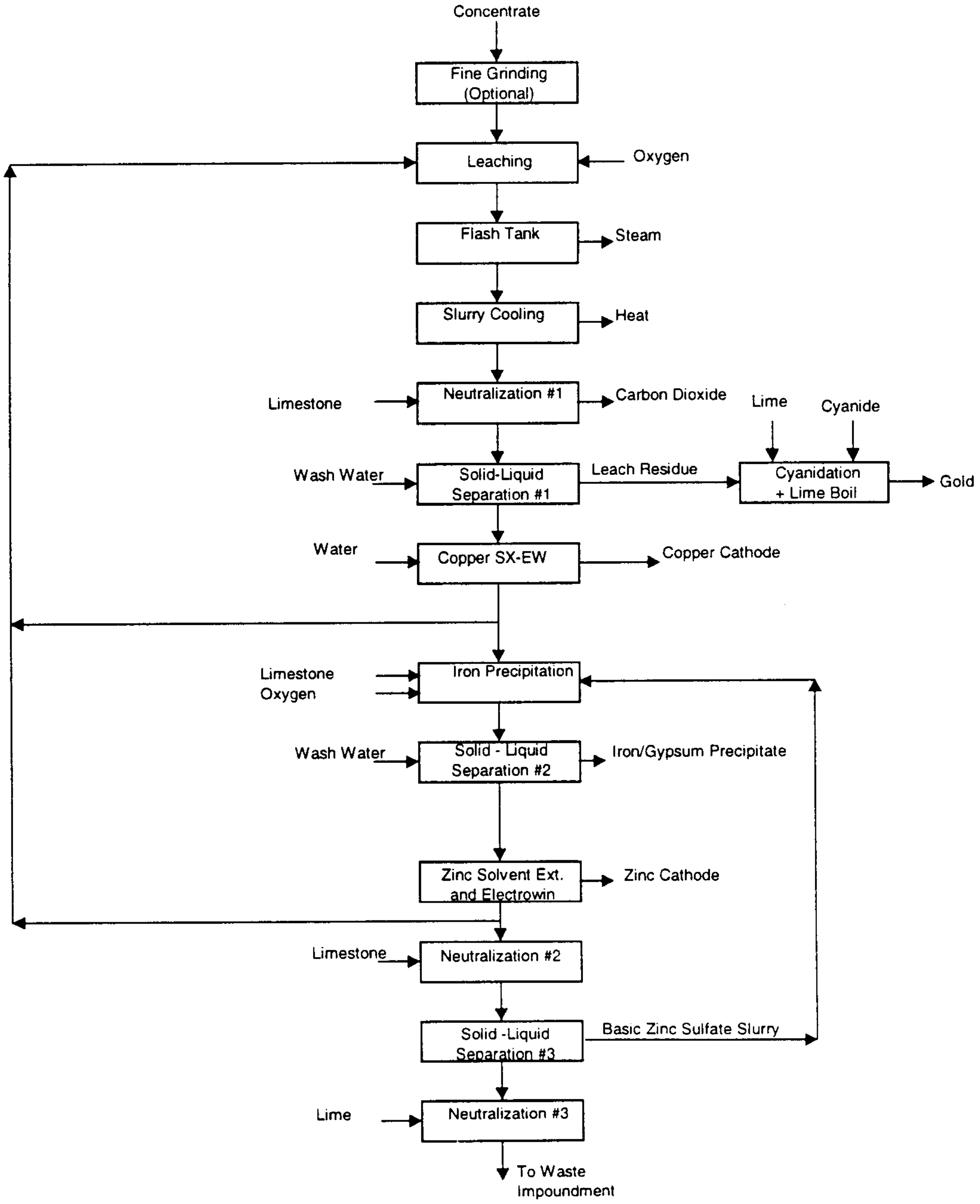

Figure 2.3 **Flowsheet for Dynatec Process for Treatment of Bulk Zn/Cu/Pb Concentrate**

2.2.2.3 Basic Design Criteria

Design criteria used for a low temperature process, such as provided by Dynatec, on bulk concentrate (Process Option II) is given in Table 2.18 below.

Table 2.18 Low Temperature Process - Option II

Mass Treated		**tonnes/yr**	**225,000**
Analysis	Copper	%	4.4
	Zinc	%	34.3
	Gold	g/tonne	5
	Silver	g/tonne	810
Products Recovered	Copper	tonnes/yr	9,455
	Zinc	tonnes/yr	74,471
	Gold	oz/yr	34,365
	Silver	oz/yr	5,274,116
Overall Recovery	Copper	%	95.50
	Zinc	%	96.50
	Gold	%	95
	Silver	%	90

2.2.2.4 Capital and Operating Costs

Capital Cost

The capital cost for low temperature processing of bulk concentrate is tabulated in Table 2.19 below. A full breakdown is provided in Appendix A.

Table 2.19 Capital Cost for Low Temperature Processing of Bulk Concentrate

	Capital Cost (US$1000)
Process plant	115,602
Limestone	4,066
S° disposal	1,000
Effluent treatment	323
Subtotal	**120,991**
Indirects (50%)	60,496
O_2 plant	13,015
Subtotal	**194,501**
Contingency (20%)	38,900
Total	**233,401**

Operating Cost

The operating cost for the low temperature processing of bulk concentrate is tabulated in Table 2.20 below. A full breakdown is provided in Appendix A.

Table 2.20 Operating Cost for Total Oxidation of a Bulk Concentrate

	Operating Cost (US$1000)
Process plant	49,883
O_2 plant	2,153
Total	**52,036**
$/t concentrate feed	**231.28**

2.2.2.5 Environmental Considerations

In this flowsheet, the selenium will largely deport to the leach residue as a S-Se product or as unreacted sulfide minerals (principally pyrite). Any selenium that is oxidized in the leach will end up as either calcium precipitate (calcium selenite/selenate), lead precipitate (lead selenite/selenate) or will be incorporated in iron residues (eg. jarosites or basic ferric sulfates). Selenium will not enter the copper electrowinning circuits due to the solvent extraction barrier to its transfer. This should ensure high purity final copper metal product. Similarly, the use of solvent extraction – electrowinning for zinc recovery should prevent any contamination of the final zinc cathode product.

It is possible that selenium will react with cyanide in the same way that elemental sulfur will react to form thiocyanate. The analogous species of selenium would be $SeCN^-$. This species is known to be difficult to treat. It is more stable with respect to oxidation than SCN^-.

2.2.2.6 Testwork Required for Prefeasibility

Leaching testwork will be required to demonstrate effective oxidation of the various minerals and to determine the deportment of selenium in the product and waste streams. Leaching will have to be studied as a function of time, temperature and oxygen pressure. The behaviour of selenium in acid neutralization should also be quantified. The oxidation products will need to be subjected to washing, lime boil and cyanidation testing. Key parameters will be % excess of lime for lime boil, temperature and time for lime boil and cyanidation conditions (time, pH and cyanide addition).

It will be necessary for prefeasibility to conduct a limited amount of testwork on copper and zinc SX-EW. This should consist in the preparation of standard McCabe – Thiele diagrams for copper and zinc extraction. It will not be necessary to perform any electrowinning of copper or zinc for this study.

2.2.3 Bioleach

2.2.3.1 Process Description

The process includes a pretreatment step for pyrrhotite removal, using recycled leachate, followed by a solid liquid separation step to dispose of the solution to lime or limestone neutralization. The solids from the pretreatment stage are pumped to the first stage direct biological leach of the bulk concentrate at 35° C, using a mixed culture of mesophilic thiobacilli and ferrobacilli, to oxidize the sphalerite and dissolve the zinc as zinc sulphate and precipitate lead as lead sulphate. Some pyrite is leached in this stage to provide sufficient ferric iron in solution for rapid sphalerite dissolution. The tetrahedrite will also leach in this stage, resulting in some copper and antimony in solution.

The first bioleach stage is followed by solid liquid separation. The solution reports to limestone neutralization for iron and copper/antimony removal, prior to zinc refining, while the solids report to a second stage bioleach. The second stage bioleach is conducted at approximately 70° C using a mixed culture of thermophilic thiobacilli to further oxidize the sulphides, primarily chalcopyrite for copper recovery. The second bioleach stage is followed by solid liquid separation, including a wash. The recovered bioleach liquor would then be directed to copper solvent extraction and electrowinning, neutralization with limestone for iron removal and finally zinc solvent extraction and electrowinning. A bleed stream would be taken out for use in the pretreatment step and finally to neutralization and disposal with the mill tailings.

The washed residue from the second stage bioleach is directed to a limestone neutralization step followed by cyanidation/CIP of the residue for gold and silver recovery by conventional methods. The cyanidation tailings are subjected to an oxidative cyanide destruction step prior to disposal with the mill tailings.

The commercial basis for this leaching process is the biological oxidation process applied to gold ores or concentrates and pyrite – cobalt concentrates as developed by Gencor (now Billiton), Bactech and others and practiced at Sao Bento, Wiluna, Harbour Lights, Ashanti, Kasese (cobalt). However, there are no commercial processes oxidizing chalcopyrite concentrates for copper recovery. Further comments are given under "Testwork Required for Prefeasibility".

Process Chemistry

The biological leaching of bulk concentrate will take place at 35° C, in the first stage, and 70° C in the second stage, at a pH of 1.3- 1.8. The main products of leaching will be dissolved base metals, lead sulfate precipitate, elemental sulfur, ferric precipitates and unreacted pyrite. It is expected that gold and silver will be liberated in the second stage bioleach residue. Silver will likely form argentojarosite and require a hot lime boil to decompose to the cyanide soluble silver hydroxide species. The leach discharge is expected to be low in acid, suitable for direct solvent extraction of copper. Copper will be extracted using a conventional oxime solvent extractant followed by stripping in spent electrolyte and electrowinning of LME grade 1 copper cathode. The copper raffinate will go to iron removal and neutralization, using lime or limestone, to prepare an iron free feed for zinc solvent extraction. Zinc will then be recovered using D2EHPA as an extractant. Zinc is stripped with spent electrolyte followed by acid sulfate electrowinning of Special High Grade zinc. The zinc raffinate is recycled to leaching (providing an acid source for leaching) except for a bleed stream to maintain water balance.

Leaching

Sphalerite Leaching: $ZnS + 0.5O_2 + H_2SO_4 \rightarrow ZnSO_4 + S'' + H_2O$

Chalcopyrite Leaching: $CuFeS_2 + O_2 + 2H_2SO_4 \rightarrow CuSO_4 + FeSO_4 + 2S'' + 2H_2O$

Galena Leaching: $PbS + 0.5O_2 + H_2SO_4 \rightarrow PbSO_4(s) + S'' + H_2O$

Pyrrhotite/Marmatite Leaching: $FeS + 0.5O_2 + H_2SO_4 \rightarrow FeSO_4 + S'' + H_2O$

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

Ferrous Sulfate Oxidation: $FeSO_4 + 0.25O_2 + 0.5H_2SO_4 \rightarrow 0.5Fe_2(SO_4)_3 + 0.5H_2O$

Elemental Sulfur Oxidation: $S'' + 1.5O_2 + H_2O \rightarrow H_2SO_4$

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

Lime Boil

Argentojarosite Decomposition:

$$AgFe_3(SO_4)_2(OH)_6 + 2Ca(OH)_2 \rightarrow AgOH + 2.5Fe_2O_3 + 2CaSO_4.2H_2O + 0.5H_2O$$

Cyanidation

Silver Leaching: $AgOH + 2NaCN \rightarrow NaAg(CN)_2 + NaOH$

Gold Leaching: $Au + 0.25O_2 + 2NaCN + 0.5H_2O \rightarrow NaAu(CN)_2 + NaOH$

Copper Solvent Extraction and Electrowinning

Copper Extraction: $CuSO_4 + 2\overline{HR} \rightarrow \overline{CuR_2} + H_2SO_4$

Copper Stripping: $\overline{CuR_2} + H_2SO_4 \rightarrow CuSO_4 + 2\overline{HR}$

Copper Electrowinning: $CuSO_4 + H_2O \rightarrow Cu + H_2SO_4 + 0.5O_2$

Neutralization and Iron Oxyhydrolysis

Acid Neutralization: $CaO + H_2SO_4 + H_2O \rightarrow CaSO_4.2H_2O$

Iron Precipitation: $Fe_2(SO_4)_3 + 3CaO + 7H_2O \rightarrow 2FeO(OH) + 3CaSO_4.2H_2O$

Zinc Solvent Extraction and Electrowinning

Zinc Extraction: $ZnSO_4 + 2\overline{HR} \rightarrow \overline{ZnR_2} + H_2SO_4$

Zinc Stripping: $\overline{ZnR_2} + H_2SO_4 \rightarrow ZnSO_4 + 2\overline{HR}$

Zinc Electrowinning: $ZnSO_4 + H_2O \rightarrow Zn + H_2SO_4 + 0.5O_2$

2.2.3.2 Flowsheet

As shown in Figure 2.4, the selenium will largely deport to the leach residue as a S-Se product or as unreacted sulfide minerals (principally pyrite). Any selenium that is oxidized in the leach will end up as either calcium precipitate (calcium selenite/selenate), lead precipitate (lead selenite/selenate) or will be incorporated in iron residues (eg. jarosites or basic ferric sulfates). Selenium will not enter the copper or zinc electrowinning circuits due to the solvent extraction barrier to its transfer. This should ensure high purity final metal products.

Antimony and minor arsenic will report to the bioleachate phase and then be neutralized and precipitated in the iron removal step as ferric antimonate and ferric arsenate for disposal with the mill tailings.

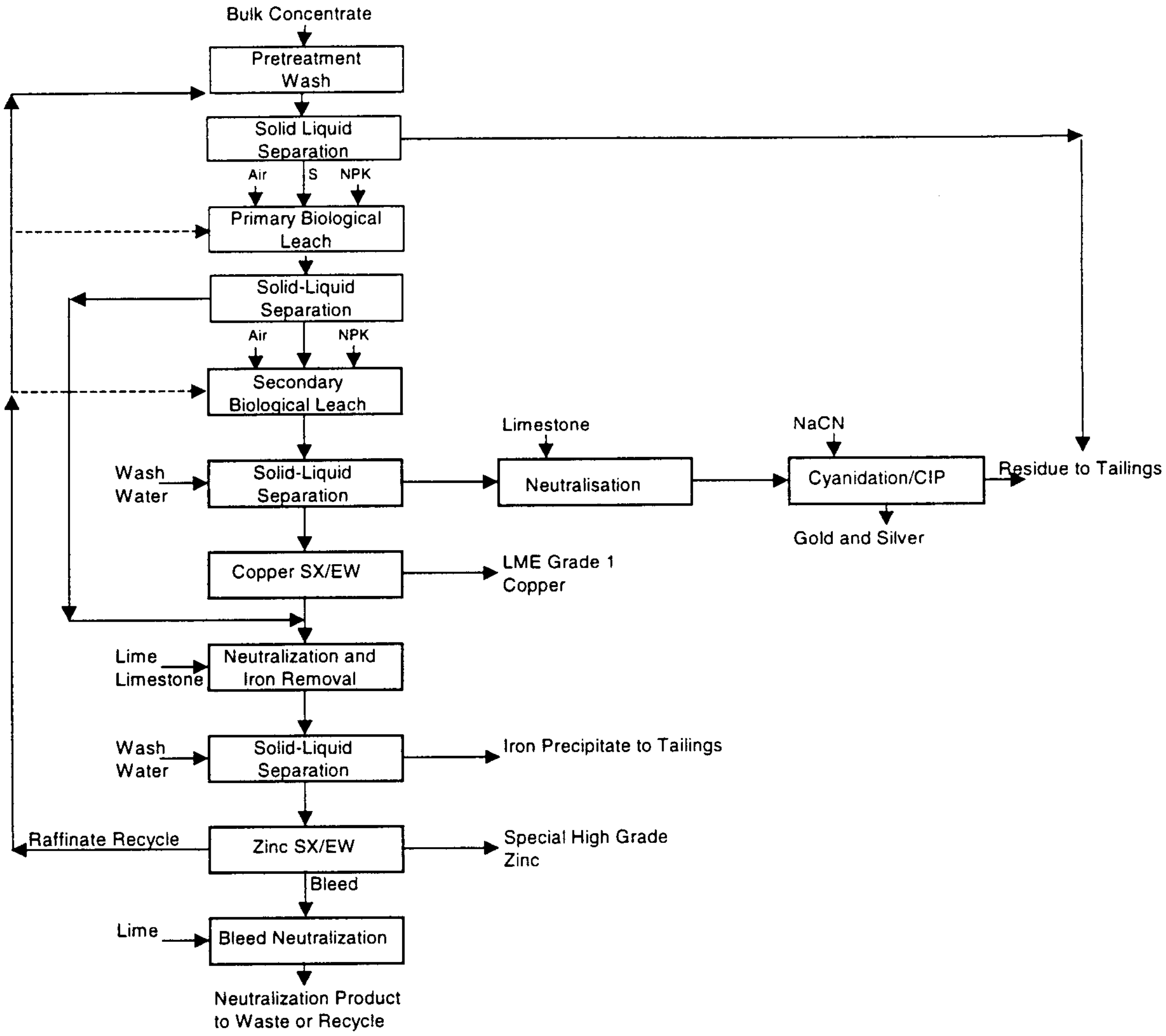

Figure 2.4 **Flowsheet for Biological Treatment of a Bulk Concentrate**

2.2.3.3 Basic Design Criteria

Design criteria used for bioleaching of a bulk concentrate (Process Option III) is given in Table 2.21 below.

Table 2.21 Bioleaching Process - Option III

Mass Treated		**tonnes/yr**	**225,000**
Analysis	Copper	%	4.4
	Zinc	%	34.3
	Gold	g/tonne	5
	Silver	g/tonne	810
Products Recovered	Copper	tonnes/yr	8,415
	Zinc	tonnes/yr	73,316
	Gold	oz/yr	34,365
	Silver	oz/yr	5,274,116
Overall Recovery	Copper	%	85
	Zinc	%	95
	Gold	%	95
	Silver	%	90

2.2.3.4 Capital and Operating Costs

Capital Cost

The capital cost for bioleaching of a bulk concentrate is tabulated in Table 2.22 below. A full breakdown is provided in Appendix A.

Table 2.22 Capital Cost for Bioleaching of a Bulk Concentrate

	Capital Cost (US$1000)
Process plant	102,092
Limestone	6,918
S° disposal	-
Effluent treatment	540
Subtotal	**109,550**
Indirects (50%)	54,775
O_2 plant	-
Subtotal	**164,325**
Contingency (20%)	32,865
Total	**197,190**

Operating Cost

The operating cost for bioleaching of a bulk concentrate is tabulated in Table 2.23 below. A full breakdown is provided in Appendix A.

Table 2.23 Operating Cost for Bioleaching of a Bulk Concentrate

	Operating Cost (US$1000)
Process plant	51,921
Total	**51,921**
$/t concentrate feed	**230.76**

2.2.3.5 Environmental Considerations

The most important environmental issues will relate to aqueous discharges and storage of tailings. The major residue stream will issue from the cyanidation plant. It will have been treated for cyanide destruction, most likely by an SO_2-air process operating on the final pulp. Most of its acid generating capabilities will have been destroyed as a result of the oxidation reactions; the elemental sulfur will have been largely removed by flotation. However, it is likely that this sulfur would be reintroduced with the tailings stream for discharge. There may be a need to add additional limestone to supply alkalinity that will ensure that no acid can be generated in the tailings impoundment in the future. The impoundment may require a lining to control leakage of the dissolved components remaining in the tailings water. The deposited solids will probably require a permanent water cover to reduce access to oxygen. If these measures are followed, the environment should be adequately protected from the potential effects of acid drainage and residual cyanide compounds.

The selenium will oxidize during the first stage bioleach and most will report with the lead as a lead sulphate precipitate in the bioleach residue. This product will have similar stability to lead sulphate. Some of the selenium will remain in the bioleachate solution, as selenite/selenate species, and will be precipitated as calcium selenite/selenate during neutralization of the leachate and ultimately report to the solid tailings waste storage. The stability of the selenium in this product is not known.

The minor amounts of arsenic and antimony present in the concentrate will also oxidize during the first stage bioleach and report to the leachate product. The leachate product will contain significant dissolved ferric iron (as sulphate) and, following neutralization with limestone, the arsenic and antimony will precipitate as calcium ferric arsenate and calcium ferric antimonate and ultimately report to the solid tailings storage. These precipitate products are relatively stable, however, not as stable as the complimentary As-Sb precipitates produced in an autoclave.

The overall water balance for this project has not been analyzed as yet. If zero discharge cannot be maintained, then a further treatment process may be required on the effluent discharge stream.

2.2.3.6 Testwork Required for Prefeasibility

In our judgement, the major process risk in this option relates to the thermophilic secondary bioleach stage. There are no commercial processes successfully oxidizing chalcopyrite concentrates for copper production. This technology has historically suffered from poor copper recovery, in the order of 50-60%, while the associated sulfide sulfur is oxidized to sulfate and must be reacted with lime or limestone for its disposal.

The design criteria used here are based on assertions by Pacific Ore Technologies that they have a thermophilic process that can achieve high copper recoveries while stopping the sulfur oxidation at elemental sulfur, thereby saving costs due to limestone. These claims have not been supported by any technical reports or published scientific papers, so the processing economics presented in this section are speculative.

Accordingly, testing needs to be conducted to verify the design criteria for the entire process route, but especially the second stage bioleach. At the prefeasibility level, the testing need only be in small scale batch vessels, to demonstrate amenability and obtain initial estimates of the extractions, reagent consumptions, optimum conditions, and range of products that may be obtained. The oxidation products will need to be subjected to washing, lime boil, flotation, and cyanidation testing. Because of its importance in this project, the behaviour of selenium at all stages should be determined.

Based on the results of the batch scoping tests, a continuous laboratory scale bioleach circuit is operated to determine bioleach operating conditions and control criteria, retention time, metal oxidation rates in a continuous circuit and internal recycle requirements. The products from the continuous circuit can then be used to determine solid/liquid separation and washing design criteria, neutralization requirements, detailed solution analyses for downstream metal recovery (Cu and Zn) and Ag/Au leaching/recovery information.

The final deportment of selenium, arsenic and antimony will be determined during continuous circuit testing. Initial stability testing could be completed using the final precipitate products.

It will be necessary as part of the prefeasibility testing to conduct a limited amount of testwork on copper and zinc solvent extraction. The work can be limited to batch shake-out tests showing the relevant equilibria and deportment of impurities. Electrowinning tests for zinc or copper are not required for this level of study.

2.3 ZINC CONCENTRATE ALTERNATIVES

The zinc concentrate represents a flotation concentrate of Zn with minor associated gold and silver content. The concentrate was analyzed and divided into an assumed elemental and mineralogical distribution.

Table 2.24 Elemental Distribution

	Tonnage	Assays (%, g/t)						
Product	**(tpa)**	**Cu**	**Pb**	**Zn**	**Au**	**Ag**	**S (total)**	**Fe(total)**
Zn Conc	138,966	0.5	1.5	51.7	1	150	35	10.80

Table 2.25 Mineralogical Distribution

Mineral	**Wt. %**
ZnS	77.05
FeS	8.18
PbS	1.73
$CuFeS_2$	1.44
FeS_2	11.09
Au	0.00010
Ag	0.0150
SiO_2	0.50
Total	**100.00**

2.3.1 Dynatec (Sherritt) Pressure Leach/CIP

2.3.1.1 Process Description

The Dynatec (Sherritt) Zinc Pressure Leach process uses a direct autoclave leach of the reground zinc concentrate at 150°C with addition of low grade coal (10-50 kg/tonne addition) as a surfactant. Other surfactants such as lignin sulfonate and Quebracho can also be used instead of coal. In contrast to the total pressure oxidation process, some sulfur and selenium is left as elemental products in the residue from leaching.

The autoclave discharge is flashed to atmospheric pressure, cooled via slurry coolers and then neutralized prior to copper SX-EW. The washed residue is sent to lime boil – cyanidation. The use of a lime boil for a product containing elemental sulfur has not been tested. It is believed that the sulfur will react with excess lime to produce a lot of soluble thiosalts and sulfides. If this happens, the lime boil may have to be augmented by flotation and/or hot sulfur filtration. This has not been designed into the flowsheet at this point due to the uncertainty of this result.

The recovered autoclave leach liquor would then be directed to copper solvent extraction and electrowinning, neutralization and iron removal and finally zinc extraction using zinc solvent extraction and electrowinning. A bleed stream would be taken out for neutralization and recycle/disposal.

Process Chemistry

The oxygen pressure leaching of reground concentrate will take place at 150° C. The main products of leaching will be dissolved base metals, lead sulfate precipitate, elemental sulfur, hematite and unreacted pyrite. It is expected that gold will be liberated in the product solids. Silver will likely form argentojarosite and require a hot lime boil to decompose to the cyanide soluble silver hydroxide species. The autoclave discharge is expected to be low in acid, suitable for direct solvent extraction of copper after neutralization. Copper will be extracted using a conventional oxime solvent extractant followed by stripping in spent electrolyte and electrowinning of LME grade 1 copper cathode. The copper raffinate will go to iron removal and neutralization to prepare an iron free feed for zinc solvent extraction. The zinc solution will then be purified of copper, cadmium, cobalt and nickel content by zinc dust cementation. Zinc will be directly electrowon from the purified solution as Special High Grade zinc. The zinc spent electrolyte is recycled to leaching (providing an acid source for leaching) except for a bleed stream to maintain water and impurity balance.

Leaching

Sphalerite Leaching:	$ZnS + 0.5O_2 + H_2SO_4 \rightarrow ZnSO_4 + S + H_2O$
Chalcopyrite Leaching:	$CuFeS_2 + O_2 + 2H_2SO_4 \rightarrow CuSO_4 + FeSO_4 + S + H_2O$
Galena Leaching:	$PbS + 0.5O_2 + H_2SO_4 \rightarrow PbSO_4(s) + S + H_2O$
Pyrrhotite/Marmatite Leaching:	$FeS + 0.5O_2 + H_2SO_4 \rightarrow FeSO_4 + S + H_2O$
Pyrite Leaching:	$FeS_2 + 0.5O_2 + H_2SO_4 \rightarrow FeSO_4 + 2S + H_2O$
Ferrous Sulfate Oxidation:	$FeSO_4 + 0.25O_2 + 0.5H_2SO_4 \rightarrow 0.5Fe_2(SO_4)_3 + 0.5H_2O$
Elemental Sulfur Oxidation:	$S + 1.5O_2 + H_2O \rightarrow H_2SO_4$

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

Lime Boil

Argentojarosite Decomposition:

$$AgFe_3(SO_4)_2(OH)_6 + 2Ca(OH)_2 \rightarrow AgOH + 2.5Fe_2O_3 + 2CaSO_4.2H_2O + 0.5H_2O$$

Cyanidation

Silver Leaching: $AgOH + 2NaCN \rightarrow NaAg(CN)_2 + NaOH$

Gold Leaching: $Au + 0.25O_2 + 2NaCN + 0.5H_2O \rightarrow NaAu(CN)_2 + NaOH$

Copper Solvent Extraction and Electrowinning

Copper Extraction: $CuSO_4 + 2\overline{HR} \rightarrow \overline{CuR_2} + H_2SO_4$

Copper Stripping: $\overline{CuR_2} + H_2SO_4 \rightarrow CuSO_4 + 2\overline{HR}$

Copper Electrowinning: $CuSO_4 + H_2O \rightarrow Cu + H_2SO_4 + 0.5O_2$

Neutralization and Iron Oxyhydrolysis

Acid Neutralization: $CaO + H_2SO_4 + H_2O \rightarrow CaSO_4.2H_2O$

Iron Precipitation: $Fe_2(SO_4)_3 + 3CaO + 7H_2O \rightarrow 2FeO(OH) + 3CaSO_4.2H_2O$

Zinc Purification and Electrowinning

Zinc Dust Cementation: $MSO_4 + Zn \rightarrow M + ZnSO_4$, M = Cu, Cd, Ni, Co

Zinc Electrowinning: $ZnSO_4 + H_2O \rightarrow Zn + H_2SO_4 + 0.5O_2$

2.3.1.2 Flowsheet

The flowsheet is shown in Figure 2.5. The products are copper metal, zinc metal, gold and silver (impure) and various residues.

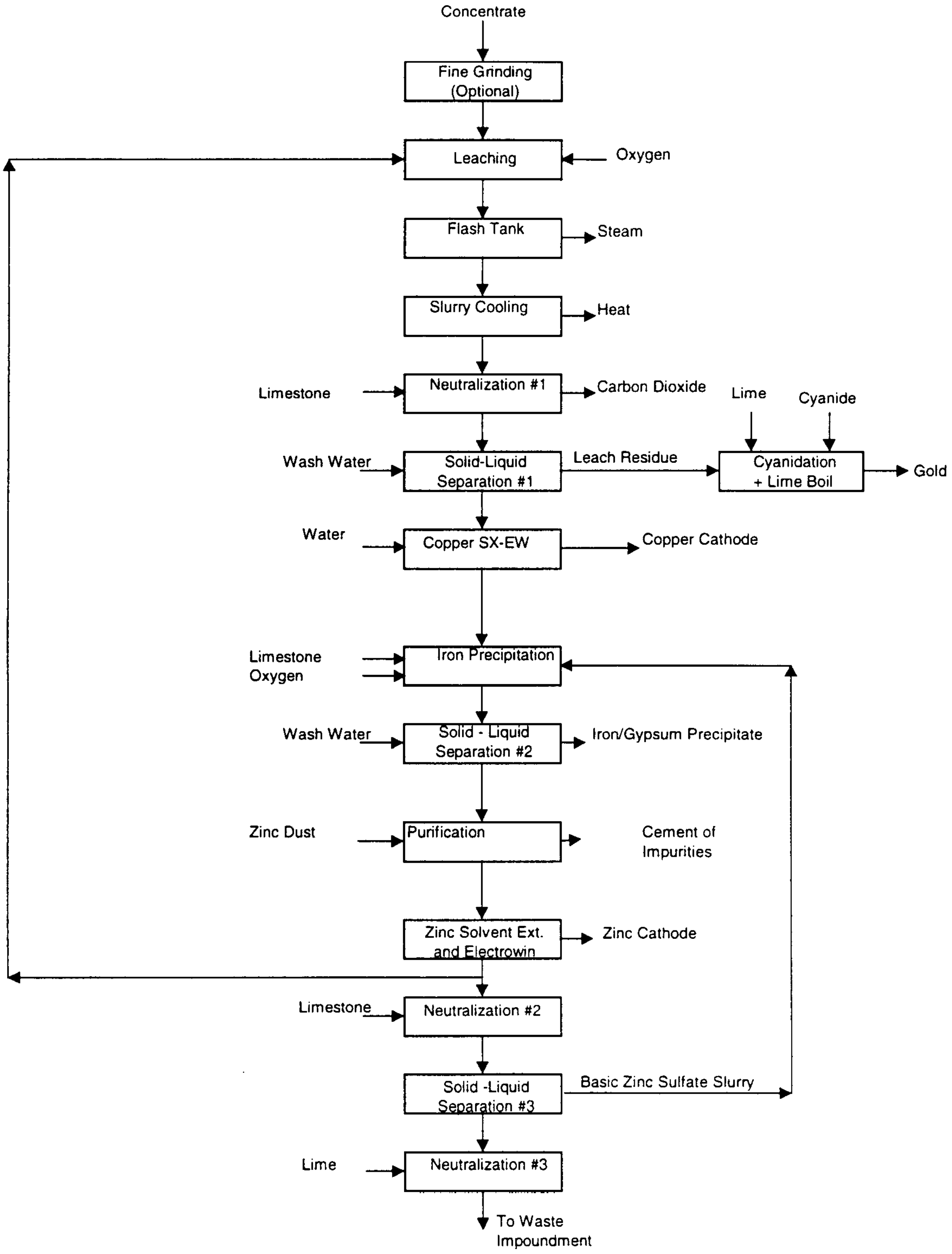

Figure 2.5 Flowsheet for Dynatec Process for Treatment of Zn Concentrate

2.3.1.3 Basic Design Criteria

Design criteria used for conventional pressure leaching of zinc concentrates (Process Options IV and V) is given in Table 2.26 below.

Table 2.26 Conventional Pressure Leaching of Zinc Concentrates - Options IV and V

Mass Treated		**tonnes/yr**	**138,966**
Analysis	Copper	%	0.5
	Zinc	%	51.3
	Gold	g/tonne	1
	Silver	g/tonne	150
Products Recovered	Copper	tonnes/yr	648
	Zinc	tonnes/yr	67,714
	Gold	oz/yr	4,245
	Silver	oz/yr	603,229
Overall Recovery	Copper	%	93.20
	Zinc	%	94.98
	Gold	%	95
	Silver	%	90

2.3.1.4 Capital and Operating Costs

Capital Cost

The capital cost for conventional pressure leaching of a zinc concentrate in Option IV is tabulated in Table 2.27 below. A full breakdown is provided in Appendix A. Note that capital for limestone mining/preparation, as well as allowances for sulphur in the tailings and effluent treatment for the complete Option IV plant is included.

Table 2.27 Capital Cost for Conventional Pressure Leaching of a Zinc Concentrate - Option IV

	Capital Cost (US$1000)
Process plant	87,780
Limestone	3,124
S° disposal	957
Effluent treatment	310
Subtotal	**92,171**
Indirects (50%)	46,085
O_2 plant	8,469
Subtotal	**146,725**
Contingency (20%)	29,345
Total	**176,070**

The capital cost for processing zinc concentrate in Option V is tabulated in Table 2.28 below. Allowances for limestone mining/preparation, sulphur in the tailings and effluent treatment for the complete Option V plant is excluded here.

Table 2.28 Capital Cost for Conventional Pressure Leaching of a Zinc Concentrate - Option V

	Capital Cost (US$1000)
Process plant	87,780
Limestone	-
S° disposal	-
Effluent treatment	-
Subtotal	**87,780**
Indirects (50%)	43,890
O_2 plant	8,469
Subtotal	**140,139**
Contingency (20%)	28,027
Total	**168,166**

Operating Cost

The operating cost for conventional pressure leaching of a zinc concentrate in Option IV is tabulated in Table 2.29 below. A full breakdown is provided in Appendix A. Note that the operating cost is different for each process plant due to a different cost of limestone in each option.

Table 2.29 Operating Cost for Conventional Pressure Leaching of a Zinc Concentrate

	Option IV	**Option V**
	Operating Cost (US$1000)	**Operating Cost (US$1000)**
Process plant	27,321	27,052
O_2 plant	1,052	1,052
Total	**28,373**	**28,104**
$/t concentrate feed	**204.17**	**202.24**

2.3.1.5 Environmental Considerations

In this flowsheet, the selenium will largely deport to the leach residue as a S-Se product or as unreacted sulfide minerals (principally pyrite). Any selenium that is oxidized in the leach will end up as either calcium precipitate (calcium selenite/selenate), lead precipitate (lead selenite/selenate) or will be incorporated in iron residues (eg. jarosites or basic ferric sulfates). Selenium will not enter the copper electrowinning circuits due to the solvent extraction barrier to its transfer. This should ensure high purity final copper metal product. There is some question about how small amounts of selenium will behave in zinc electrowinning. There is no solvent extraction barrier to selenium in this case (although this could be introduced if warranted). Selenium may co-deposit with zinc and contaminate the final deposit. This will need experimental investigation.

It is possible that selenium will react with cyanide in the same way that elemental sulfur will react to form thiocyanate. The analogous species of selenium would be $SeCN^-$. This species is known to be difficult to treat. It is more stable with respect to oxidation than SCN^-.

2.3.1.6 Testwork Required for Prefeasibility

Leaching testwork will be required to demonstrate effective oxidation of the various minerals and to determine the deportment of selenium in the product and waste streams. Leaching will have to be studied as a function of time, temperature and oxygen pressure. The behaviour of selenium in acid neutralization should also be quantified. The oxidation products will need to be subjected to washing, lime boil and cyanidation testing. Key parameters will be % excess of lime for lime boil, temperature and time for lime boil and cyanidation conditions (time, pH and cyanide addition).

It will be necessary for prefeasibility to conduct a limited amount of testwork on copper and zinc SX-EW. This should consist in the preparation of standard McCabe – Thiele diagrams for copper and zinc extraction. It will not be necessary to perform any electrowinning of copper or zinc for this study.

2.3.2 Roast, Ship Calcine

2.3.2.1 Process Description

The roasting and shipping of calcine is a potential “intermediate” process alternative. The roasting and shipping of calcine will allow for a selenium rejection through roasting, allowing a low selenium product to be shipped to a smelter.

The sulfur in the concentrate is roasted to produce sulfur dioxide while the selenium becomes selenium dioxide. Three routes are possible to treat the roaster off gas. First, the gas can be cleaned of its selenium content and then forwarded to a conventional acid plant for acid manufacture. The scrubbing of selenium from sulfur dioxide is possible due to the relative nobility of selenium. Sulfur dioxide is a reductant for selenium dioxide. This is probably undesirable at the minesite but might be feasible if the concentrate could be trans-shipped to a suitable site where an acid market would exist. A second possibility is to reduce the sulfur/selenium oxides in the gas phase to elemental sulfur /selenium for disposal. A suitable reductant for such a treatment would be hydrogen sulfide from a sour gas well field. If this option were chosen, shipping to a site where hydrogen sulfide is relatively free would be desirable. Third, the gas stream can be scrubbed of selenium and sulfur dioxide content with water, with the water subsequently being neutralized with limestone. This latter option has been assumed for the roasting options in this study.

Process Concept

The Roast and Calcine Shipping process is basically the same as the front end of the Roast – Leach – Electrowin (RLE) process. The roaster calcine is produced at site and then shipped to a zinc plant for leaching, purification and electrowinning. The roasting step produces a selenium dioxide – sulfur dioxide off gas that is treated at site for removal/fixation of elemental selenium or calcium selenite/sulfite precipitates. Precious metal recovery from the zinc calcine shipped to a zinc plant will be necessary. For example, at Cominco, the zinc plant leach residues are forwarded to the lead smelter for treatment. Gold and silver are recovered in the lead refining process.

Process Chemistry

The roasting step takes place in air at temperatures of about 850°C. The main products of roasting will be metal oxides including zinc, copper and iron oxides. Zinc oxides and iron oxides tend to combine in the roaster product as a zinc ferrite material. The zinc oxide – zinc ferrite calcine is shipped to a zinc refinery.

Roasting

$$ZnS(s) + 3/2O_2(g) \rightarrow ZnO(s) + SO_2(g)$$
$$2FeS(s) + ZnS(s) + 5O_2(g) \rightarrow ZnFe_2O_4(s) + 3SO_2(g)$$
$$PbS(s) + 3/2O_2(g) \rightarrow PbO(s) + SO_2(g)$$
$$MSe(s) + 3/2O_2(g) \rightarrow MO(s) + SeO_2(g), \text{ M = Zn, Cu, Pb, Fe}$$

Selenium Scrubbing and Reduction with Sulfur Dioxide

$$SeO_2(g) + H_2O \rightarrow H_2SeO_3(aq)$$
$$H_2SeO_3(aq) + 2SO_2(g) + H_2O \rightarrow Se + 2H_2SO_4$$

Sulfur Dioxide and Selenium Dioxide Scrubbing with Lime or Limestone

$$SO_2(g) + 2H_2O + CaO \rightarrow CaSO_3.2H_2O(s)$$
$$SeO_2(g) + 2H_2O + CaO \rightarrow CaSeO_3.2H_2O(s)$$

2.3.2.2 Flowsheet

The flowsheet is shown in Figure 2.6 along with the anticipated calcine analysis for the roasted concentrate product, Table 2.30. Note that the weight loss is relatively minor – the concentrate weight at 138,966 tonnes per annum is replaced by calcine production of 120,294 tonnes per annum.

Table 2.30 Zinc Concentrate Analysis

Species	Wolverine	KZK	3/8 W + 5/8 KZK	Calcine	% Retained in Calcine	Fate
Cu	0.4	0.5	0.5	0.58	100	CuO
Pb	1	1.5	1.5	1.73	100	PbO
Zn	51	52.1	51.7	59.73	100	ZnO
Au	1	1	1	1.155	100	Au
Ag	150	150	150	173.3	100	Ag
Fe	9.7	10.4	10.138	11.71	100	Fe_2O_3 (or $ZnFe_2O_4$)
S	35.2	30.3	32.138	0.743	2	SO_2 (g)
Se	0.18	0.03	0.086	0.00199	2	SeO_2 (g)
As	0.052	0.18	0.132	0.00305	2	As_2O_3 (g)
Sb	0.03	0.03	0.030	0.00104	3	Sb_2O_3 (g)
Hg	0.012	0.013	0.013	0	0	Hg (g)
Carbon	0.33	NA	0.33	0	0	CO_2
Ni	NA	<0.01	NA	NA	100	
Bi	NA	<0.005	NA	NA	100	
Cd	0.49	0.35	0.403	0.465	100	CdO
Co	NA	<0.01	NA	NA	100	
Cr	NA	<0.01	NA	NA	100	
Sn	NA	<0.01	NA	NA	100	
F	0.01	NA	0.01	0.00231	20	CaF_2 or MgF_2 will stay in calcine
Cl	0.0042	NA	0.0042	0.000097	2	MCl_2(g)
Te	NA	<0.01	NA	NA	1	
Ba	NA	NA	NA	NA	100	
Ge	NA	<0.01	NA	NA	20	
SiO2	NA	0.07	0.07	0.081	100	Inert
MgO	NA	0.03	0.03	0.035	100	$MgSO_4$ or $MgSeO_4$ or MgF_2
CaO	NA	NA	NA	NA	100	$CaSO_4$ or $CaSeO_4$ or CaF_2
Insol	1.59	4.49	2.91	3.36	100	
Mn	NA	0.01	0.01	0.012	100	MnO
Al_2O_3	NA	<0.01	NA	NA	100	
Oxygen				21.55		
Total	**100.00**	**100.00**	**100.00**	**100.00**		

Insol is assumed to balance to 100%

Zinc Concentrate Production: 138,966 tonnes per annum

Zinc Calcine Production: 120,294 tonnes per annum

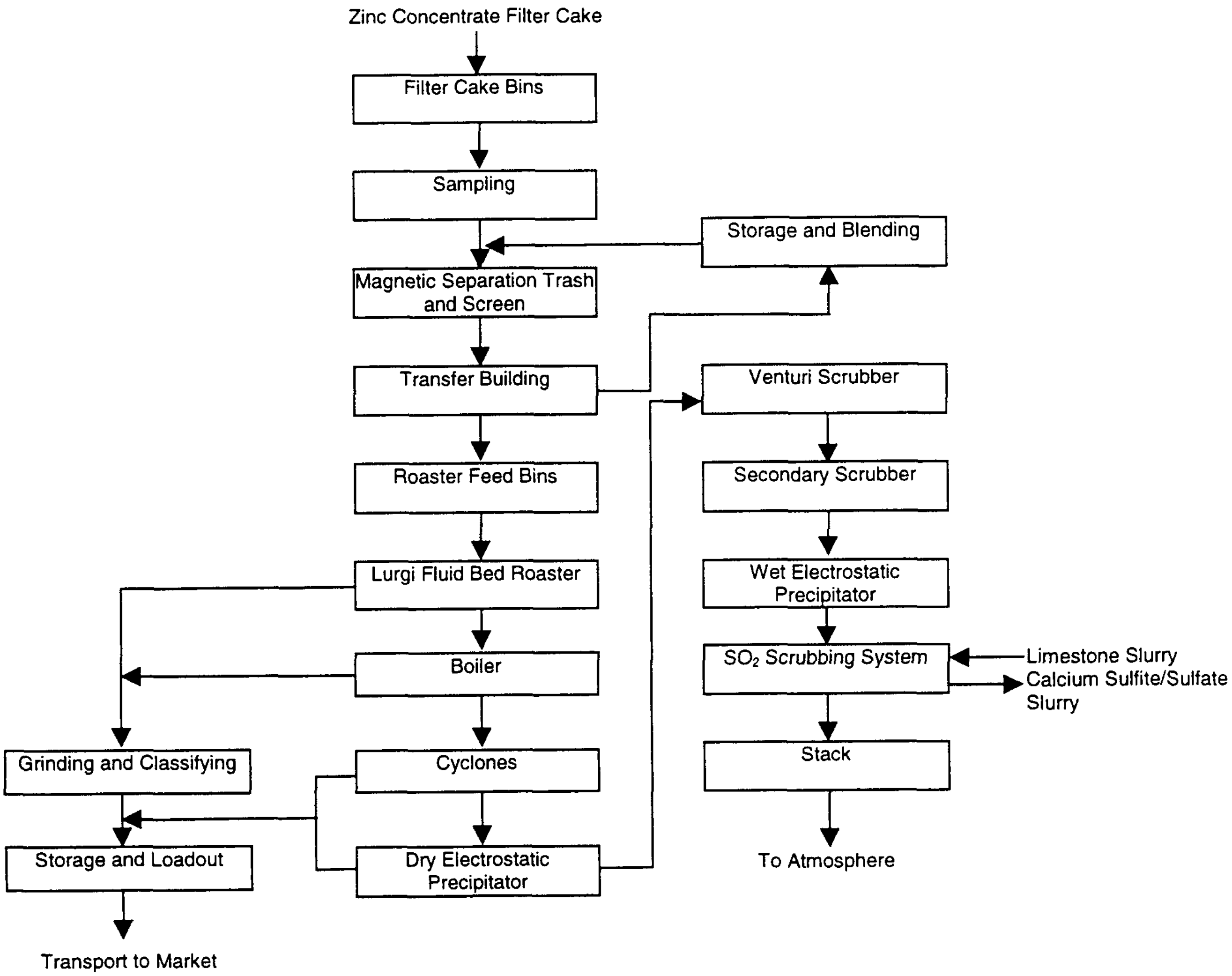

Figure 2.6 **Roaster Flowsheet**

2.3.2.3 Basic Design Criteria

Design criteria used for roasting of the zinc concentrates (Process Options VI and VII) is given in Table 2.31 below. The sulphur dioxide is scrubbed with limestone slurry and the slurry disposed into a tailings pond. The metal recoveries shown indicate the smelter credit.

Table 2.31 Roasting of the Zinc Concentrates - Options VI and VII

Mass Treated		**tonnes/yr**	**138,966**
Analysis	Copper	%	0.5
	Zinc	%	51.3
	Gold	g/tonne	1
	Silver	g/tonne	150
Products Recovered	Zinc Calcine	tonnes/yr	120,294
Overall Recovery	Copper	%	0
	Zinc	%	85
	Gold	%	65
	Silver	%	65

2.3.2.4 Capital Costs and Operating Costs

Capital Cost

The capital cost for zinc roasting and scrubbing of a zinc concentrate is tabulated in Table 2.32 below. These costs apply to both Process Options VI , VII and VIII.

Table 2.32 Capital Cost for Zinc Roasting and Scrubbing of Zinc Concentrate - Options VI ,VII and VIII

	Capital Cost (US$1000)
Process plant	24,000
Limestone	5,100
S° disposal	-
Effluent treatment	310
Subtotal	**29,400**
Indirects (50%)	14,700
O_2 plant	-
Subtotal	**44,100**
Contingency (20%)	2,900
Total	**47,000**

Operating Cost

The estimated operating cost for each of the three options is shown in the following table. The differences are due to the cost of limestone, which varies depending on the total annual requirement when the roaster is combined with the processes for treating copper/lead concentrate. The operating cost for option 8 assumes a higher power cost of $US 0.10 per kWh associated with the use of diesel power in this case.

Table 2.33 Operating Costs for Roasting of a Conventional Zinc Concentrate

	Option VI Operating Cost ($/tonne feed)	Option VII Operating Cost ($/tonne feed)	Option VIII Operating Cost ($/tonne feed)
Roaster	21.00	21.00	24.00
Gas cleaning and scrubbing	11.00	11.00	11.00
Limestone for neutralization	19.00	16.00	21.00
Total	**51.00**	**48.00**	**56.00**

2.3.2.5 Environmental Considerations

In this flowsheet, the selenium will largely deport to the off gas from roasting. The wet scrubbing of the off gas will result in an elemental selenium precipitate. Selenious acid is produced by wet scrubbing and further reaction with sulfur dioxide (the dominant species in the off gas) results in homogeneous reduction of selenium to the elemental form. Residual selenium may be scrubbed and precipitated using a slurry of quick lime (limestone may work for this application but it is unclear if the pH is high enough for effective capture and precipitation.

A small amount of selenium will likely remain in the roaster calcine either as unreacted metal selenide species (short circuiting through roaster) or as metal selenites/selenates formed during roasting. Experimental work is required to determine the eventual deportment of this selenium.

2.3.2.6 Testwork Required for Prefeasibility

The testwork required for prefeasibility is to perform proof or principle roasting tests under variety of temperature conditions to determine the degree of selenium rejection as a function of the intensity of the roasting condition.

The reduction of selenium from the gas phase should be demonstrated by wet scrubbing or neutralization testing.

The type of work required to validate this process route should probably be undertaken by a group such as Lurgi or Torco with the experience to take the results at a small scale and be able to extrapolate to a reasonable design basis.

2.4 Copper-Lead Alternatives

The copper-lead concentrate represents a floatation concentrate of copper-lead with significant associated gold and silver content. The concentrate was analyzed and divided into an assumed elemental and mineralogical distribution.

Table 2.34 Elemental Distribution

	Tonnage	Assays (%, g/t)						
Product	**(tpa)**	**Cu**	**Pb**	**Zn**	**Au**	**Ag**	**S (total)**	**Fe(total)**
Cu/Pb Conc	76,037	12	15.8	6	12.5	2,122	35	26.09

Table 2.35 Mineralogical Distribution

Mineral	Wt. %
ZnS	8.94
FeS	0.95
PbS	18.24
$CuFeS_2$	34.65
FeS_2	32.10
Au	0.00125
Ag	0.2122
SiO_2	5.11
Total	**100.00**

2.4.1 Total Oxidation Autoclave/CIP

2.4.1.1 Process Description

The total pressure oxidation process is designed to totally oxidize all sulfide sulfur to sulfate sulfur, extract copper and zinc into solution, precipitate lead as lead sulfate and liberate the precious metals (Au, Ag) for subsequent cyanidation recovery.

The process concept uses a direct autoclave leach of the reground concentrate at 220°C followed by flashing, slurry cooling, acid neutralization and solid liquid separation. The recovered residue is sent to cyanidation for precious metal recovery after a lime boil treatment. The recovered autoclave leach liquor would then be directed to copper solvent extraction and electrowinning, neutralization and iron removal and finally zinc solvent extraction and electrowinning. A bleed stream would be taken out for neutralization and recycle/disposal.

Process Chemistry

The oxygen pressure leaching of reground concentrate will take place at 220°C. The main products of leaching will be dissolved base metals, lead sulfate precipitate and hematite. It is expected that gold will be liberated in the product solids. Silver will likely form argentojarosite and require a hot lime boil to decompose to the cyanide soluble silver hydroxide species. The autoclave discharge is expected to contain some free acid and therefore will require neutralization prior to direct solvent extraction of copper. Copper will be extracted using a conventional oxime solvent extractant followed by stripping in spent electrolyte and electrowinning of LME grade 1 copper cathode. The copper raffinate will go to iron removal and neutralization to prepare an iron free feed for zinc solvent extraction. Zinc will then be recovered using D2EHPA as an extractant. Zinc is stripped with spent electrolyte followed by acid sulfate electrowinning of Special High Grade zinc. The zinc raffinate is recycled to leaching (providing an acid source for leaching) except for a bleed stream to maintain water balance.

Leaching

Sphalerite Leaching: $ZnS + 0.5O_2 + H_2SO_4 \rightarrow ZnSO_4 + S + H_2O$

Chalcopyrite Leaching: $CuFeS_2 + O_2 + 2H_2SO_4 \rightarrow CuSO_4 + FeSO_4 + S + H_2O$

Galena Leaching: $PbS + 0.5O_2 + H_2SO_4 \rightarrow PbSO_4(s) + S + H_2O$

Pyrrhotite/Marmatite Leaching: $FeS + 0.5O_2 + H_2SO_4 \rightarrow FeSO_4 + S + H_2O$

Pyrite Leaching: $FeS_2 + 0.5O_2 + H_2SO_4 \rightarrow FeSO_4 + 2S + H_2O$

Ferrous Sulfate Oxidation: $FeSO_4 + 0.25O_2 + 0.5H_2SO_4 \rightarrow 0.5Fe_2(SO_4)_3 + 0.5H_2O$

Elemental Sulfur Oxidation: $S + 1.5O_2 + H_2O \rightarrow H_2SO_4$

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

Lime Boil

Argentojarosite Decomposition:

$$AgFe_3(SO_4)_2(OH)_6 + 2Ca(OH)_2 \rightarrow AgOH + 2.5Fe_2O_3 + 2CaSO_4.2H_2O + 0.5H_2O$$

Cyanidation

Silver Leaching: $AgOH + 2NaCN \rightarrow NaAg(CN)_2 + NaOH$

Gold Leaching: $Au + 0.25O_2 + 2NaCN + 0.5H_2O \rightarrow NaAu(CN)_2 + NaOH$

Copper Solvent Extraction and Electrowinning

Copper Extraction: $CuSO_4 + 2\overline{HR} \rightarrow \overline{CuR_2} + H_2SO_4$

Copper Stripping: $\overline{CuR_2} + H_2SO_4 \rightarrow CuSO_4 + 2\overline{HR}$

Copper Electrowinning: $CuSO_4 + H_2O \rightarrow Cu + H_2SO_4 + 0.5O_2$

Neutralization and Iron Oxyhydrolysis

Acid Neutralization: $CaO + H_2SO_4 + H_2O \rightarrow CaSO_4.2H_2O$

Iron Precipitation: $Fe_2(SO_4)_3 + 3CaO + 7H_2O \rightarrow 2FeO(OH) + 3CaSO_4.2H_2O$

Zinc Solvent Extraction and Electrowinning

Zinc Extraction: $ZnSO_4 + 2\overline{HR} \rightarrow \overline{ZnR_2} + H_2SO_4$

Zinc Stripping: $\overline{ZnR_2} + H_2SO_4 \rightarrow ZnSO_4 + 2\overline{HR}$

Zinc Electrowinning: $ZnSO_4 + H_2O \rightarrow Zn + H_2SO_4 + 0.5O_2$

Selenium Deportment

In this flowsheet, the selenium will largely deport to the leach residue as a S-Se product or as unreacted sulfide minerals (principally pyrite). Any selenium that is oxidized in the leach will end up as either calcium precipitate (calcium selenite/selenate), lead precipitate (lead selenite/selenate) or will be incorporated in iron residues (eg. jarosites or basic ferric sulfates). Selenium will not enter the copper or zinc electrowinning circuits due to the solvent extraction barrier to its transfer. This should ensure high purity final metal products.

2.4.1.2 Flowsheet

The flowsheet is shown in Figure 2.7. The products are copper metal, zinc metal, gold and silver (impure) and various residues.

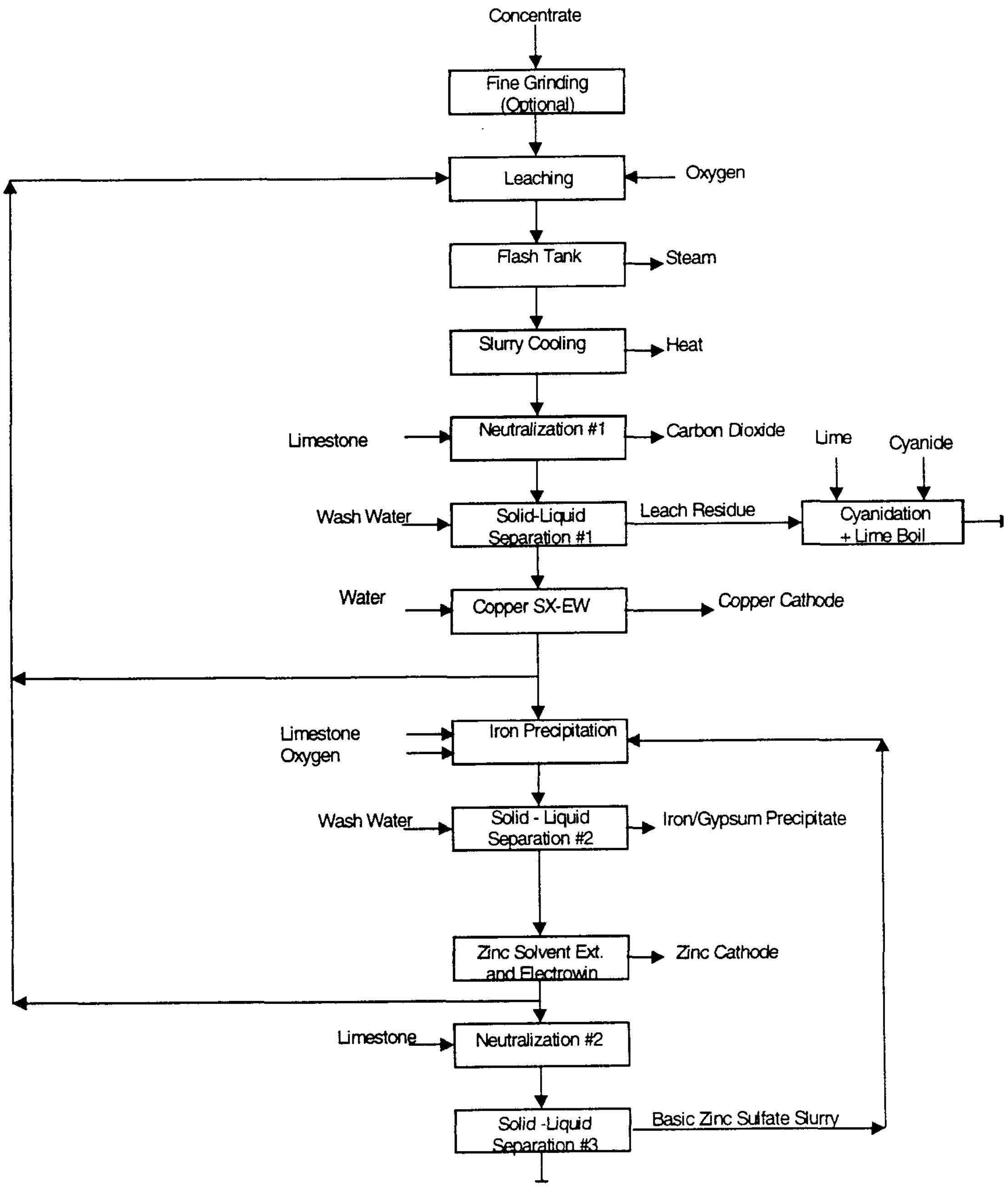

Figure 2.7 The Total Pressure Oxidation Process Applied To Cu/Pb Concentrate

2.4.1.3 Basic Design Criteria

Design criteria used for total oxidation of a Cu-Pb concentrate (Process Options V and VII) is given in Table 2.36 below.

Table 2.36 Total Oxidation of a Cu-Pb Concentrate - Options V and VII

Mass Treated		**tonnes/yr**	**76,037**
Analysis	Copper	%	12
	Zinc	%	6
	Gold	g/tonne	12.5
	Silver	g/tonne	2,122
Products Recovered	Copper	tonnes/yr	8,987
	Zinc	tonnes/yr	4,400
	Gold	oz/yr	29,033
	Silver	oz/yr	4,928,713
Overall Recovery	Copper	%	98.49
	Zinc	%	96.43
	Gold	%	95
	Silver	%	95

2.4.1.4 Capital and Operating Costs

Capital Cost

The capital cost for the total oxidation of a Cu-Pb concentrate (Process Options V and VII) is given in Table 2.37 below. A full breakdown is provided in Appendix A. Note that capital also has been included in both options for limestone mining/preparation, sulphurs in tailings and effluent treatment. This capital covers the complete process plant in each of these options.

Table 2.37 Capital Cost for Total Oxidation of Cu-Pb Concentrate - Options V and VII

	Option V	Option VII
	Capital Cost (US$1000)	**Capital Cost (US$1000)**
Process plant	56,338	56,338
Limestone	4,826	6,918
S° disposal	653	-
Effluent treatment	400	330
Subtotal	**62,217**	**63,586**
Indirects (50%)	31,109	31,793
O_2 plant	11,000	11,000
Subtotal	**104,326**	**106,379**
Contingency (20%)	20,865	21,276
Total	**125,191**	**127,655**

Operating Cost

The operating cost for total oxidation of a Cu-Pb concentrate for both process options is tabulated in Table 2.38 below. A full breakdown is provided in Appendix A. Note that the operating cost is different for each process plant due to a different cost of limestone in each option.

Table 2.38 Operating Cost for Total Oxidation of a Cu-Pb Concentrate

	Option V	Option VII
	Operating Cost (US$1000)	**Operating Cost (US$1000)**
Process plant	10,911	10,947
O_2 plant	1,652	1,652
Total	**12,563**	**12,599**
$/t concentrate feed	**165.23**	**165.70**

2.4.1.5 Environmental Considerations

Virtually all selenium will be oxidized to selenite/selenate by the autoclave process. This selenium will deport in a number of ways. First, some selenium will precipitate as a lead product during autoclave leaching. This product will likely have the same stability as lead sulfate and may well be the best way to dispose of selenium. The remaining selenium that has been oxidized will be precipitated as calcium selenite/selenate during neutralization and will become part of the solid waste in tails. The stability of this product is not known relative to the stability of gypsum.

2.4.1.6 Testwork Required for Prefeasibility

Leaching testwork will be required to demonstrate effective oxidation of the various minerals and to determine the deportment of selenium in the product and waste streams. Leaching will have to be studied as a function of time, temperature and oxygen pressure. The behaviour of selenium in acid neutralization should also be quantified. The oxidation products will need to be subjected to washing, lime boil and cyanidation testing. Key parameters will be % excess of lime for lime boil, temperature and time for lime boil and cyanidation conditions (time, pH and cyanide addition).

It will be necessary for prefeasibility to conduct a limited amount of testwork on copper and zinc SX-EW. This should consist in the preparation of standard McCabe – Thiele diagrams for copper and zinc extraction. It will not be necessary to perform any electrowinning of copper or zinc for this study.

2.4.2 Low Temperature Autoclave/Refining and PM's - Dynatec Process/CIP

2.4.2.1 Process Description

The Dynatec process uses a direct autoclave leach of the reground concentrate at 150°C with addition of low grade coal (10-50 kg/tonne addition) as a surfactant. In contrast to the total pressure oxidation process, some sulfur and selenium is left as elemental products in the residue from leaching. The autoclave discharge is flashed to atmospheric pressure, cooled via slurry coolers and then neutralized prior to copper SX-EW. The washed residue is sent to lime boil – cyanidation. The use of a lime boil for a product containing elemental sulfur has not been tested. It is believed that the sulfur will react with excess lime to produce a lot of soluble thiosalts and sulfides. If this happens, the lime boil may have to be augmented by flotation and/or hot sulfur filtration. This has not been designed into the flowsheet at this point due to the uncertainty of this result.

The recovered autoclave leach liquor would then be directed to copper solvent extraction and electrowinning, neutralization and iron removal and finally zinc extraction using zinc solvent extraction and electrowinning. A bleed stream would be taken out for neutralization and recycle/disposal.

Process Chemistry

The oxygen pressure leaching of reground concentrate will take place at 150°C. The main products of leaching will be dissolved base metals, lead sulfate precipitate, elemental sulfur, hematite and unreacted pyrite. It is expected that gold will be liberated in the product solids. Silver will likely form argentojarosite and require a hot lime boil to decompose to the cyanide soluble silver hydroxide species. The autoclave discharge is expected to be low in acid, suitable for direct solvent extraction of copper after neutralization. Copper will be extracted using a conventional oxime solvent extractant followed by stripping in spent electrolyte and electrowinning of LME grade 1 copper cathode. The copper raffinate will go to iron removal and neutralization to prepare an iron free feed for zinc solvent extraction. Zinc will then be recovered using D2EHPA as an extractant. Zinc is stripped with spent electrolyte followed by acid sulfate electrowinning of Special High Grade zinc. The zinc raffinate is recycled to leaching (providing an acid source for leaching) except for a bleed stream to maintain water balance.

Leaching

Sphalerite Leaching: $ZnS + 0.5O_2 + H_2SO_4 \rightarrow ZnSO_4 + S + H_2O$
Chalcopyrite Leaching: $CuFeS_2 + O_2 + 2H_2SO_4 \rightarrow CuSO_4 + FeSO_4 + S + H_2O$
Galena Leaching: $PbS + 0.5O_2 + H_2SO_4 \rightarrow PbSO_4(s) + S + H_2O$
Pyrrhotite/Marmatite Leaching: $FeS + 0.5O_2 + H_2SO_4 \rightarrow FeSO_4 + S + H_2O$
Pyrite Leaching: $FeS_2 + 0.5O_2 + H_2SO_4 \rightarrow FeSO_4 + 2S + H_2O$
Ferrous Sulfate Oxidation: $FeSO_4 + 0.25O_2 + 0.5H_2SO_4 \rightarrow 0.5Fe_2(SO_4)_3 + 0.5H_2O$
Elemental Sulfur Oxidation: $S + 1.5O_2 + H_2O \rightarrow H_2SO_4$
Hematite Precipitation: $Fe_2(SO_4)_3 + 3H_2O \rightarrow Fe_2O_3 + 3H_2SO_4$

Lime Boil

Argentojarosite Decomposition:

$$AgFe_3(SO_4)_2(OH)_6 + 2Ca(OH)_2 \rightarrow AgOH + 2.5Fe_2O_3 + 2CaSO_4.2H_2O + 0.5H_2O$$

Cyanidation

Silver leaching: $AgOH + 2NaCN \rightarrow NaAg(CN)_2 + NaOH$

Gold Leaching: $Au + 0.25O_2 + 2NaCN + 0.5H_2O \rightarrow NaAu(CN)_2 + NaOH$

Copper Solvent Extraction and Electrowinning

Copper Extraction: $CuSO_4 + 2\overline{HR} \rightarrow \overline{CuR_2} + H_2SO_4$

Copper Stripping: $\overline{CuR_2} + H_2SO_4 \rightarrow CuSO_4 + 2\overline{HR}$

Copper Electrowinning: $CuSO_4 + H_2O \rightarrow Cu + H_2SO_4 + 0.5O_2$

Neutralization and Iron Oxyhydrolysis

Acid Neutralization: $CaO + H_2SO_4 + H_2O \rightarrow CaSO_4.2H_2O$

Iron Precipitation: $Fe_2(SO_4)_3 + 3CaO + 7H_2O \rightarrow 2FeO(OH) + 3CaSO_4.2H_2O$

Zinc Solvent Extraction and Electrowinning

Zinc Extraction: $ZnSO_4 + 2\overline{HR} \rightarrow \overline{ZnR_2} + H_2SO_4$

Zinc Stripping: $\overline{ZnR_2} + H_2SO_4 \rightarrow ZnSO_4 + 2\overline{HR}$

Zinc Electrowinning: $ZnSO_4 + H_2O \rightarrow Zn + H_2SO_4 + 0.5O_2$

2.4.2.2 Flowsheet

The flowsheet is shown in Figure 2.8. The products are copper metal, zinc metal, gold and silver (impure) and various residues.

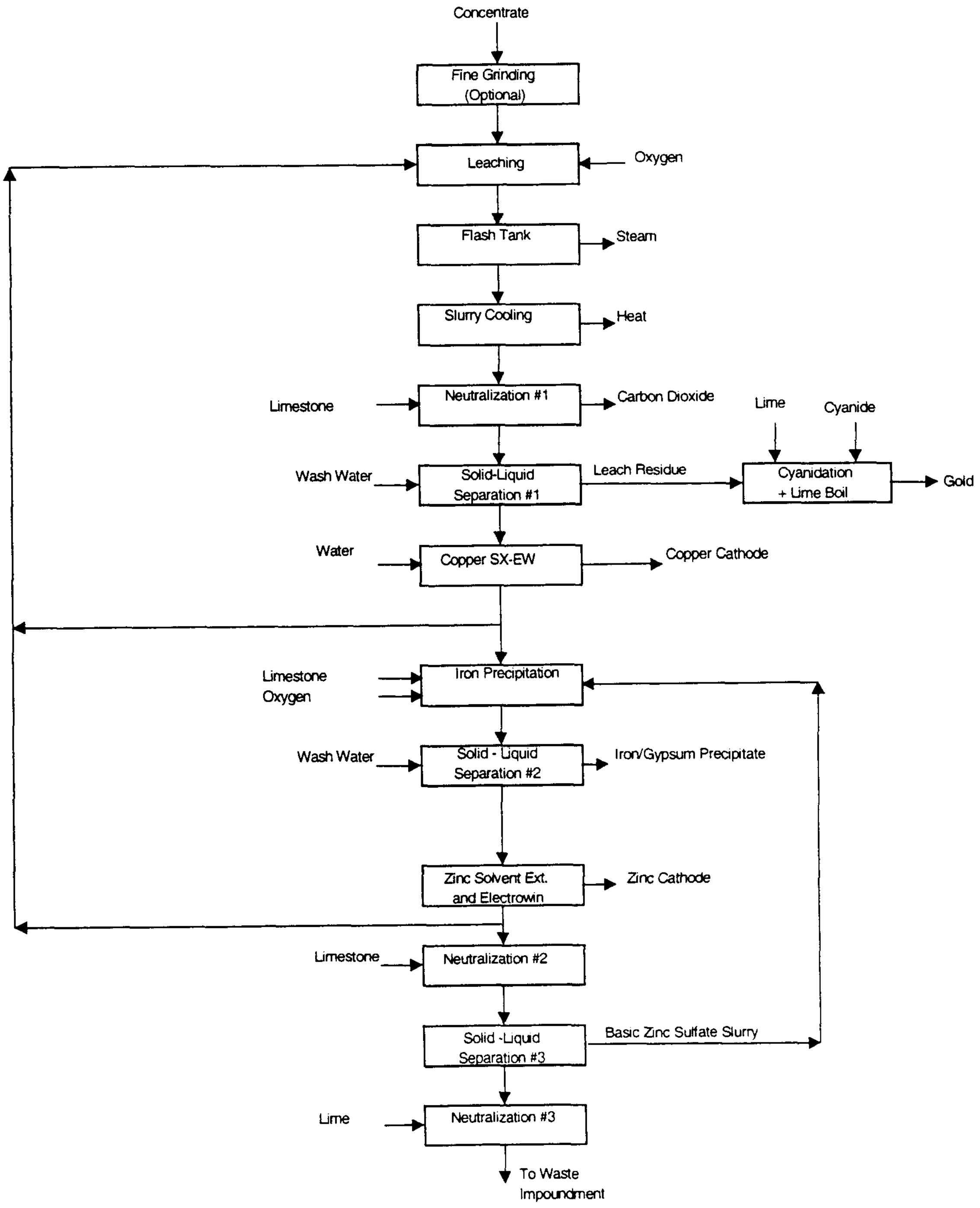

Figure 2.8 Flowsheet for Dynatec Process for Treatment of Cu/Pb Concentrate

2.4.2.3 Basic Design Criteria

Design criteria used for total oxidation of a Cu-Pb concentrate (Process Options IV and VI) is given in Table 2.39 below.

Table 2.39 *Dynatec Pressure Oxidation of a Cu-Pb Concentrate - Options IV and VI*

Mass Treated		tonnes/yr	76,037
Analysis	Copper	%	12
	Zinc	%	6
	Gold	g/tonne	12.5
	Silver	g/tonne	2,122
Products Recovered	Copper	tonnes/yr	8,713
	Zinc	tonnes/yr	4,350
	Gold	oz/yr	29,033
	Silver	oz/yr	4,669,307
Overall Recovery	Copper	%	95.49
	Zinc	%	95.35
	Gold	%	95
	Silver	%	90

2.4.2.4 Capital and Operating Costs

Capital Cost

The capital cost for the low temperature leaching of a Cu-Pb concentrate (Process Option IV) is tabulated in Table 2.40 below. Allowances for limestone mining/preparation, sulphur in tailings and effluent treatment in Option IV have been included elsewhere.

Table 2.40 Capital Cost for Dynatec Leach of Cu-Pb Concentrate - Option IV

	Capital Cost (US$1000)
Process plant	48,950
Limestone	-
S° disposal	-
Effluent treatment	-
Subtotal	**48,950**
Indirects (50%)	24,475
O_2 plant	6,136
Subtotal	**79,561**
Contingency (20%)	15,912
Total	**95,473**

The capital cost for processing Cu-Pb concentrate in Option VI is given in Table 2.41 below.

Table 2.41 Capital Cost for Dynatec Leach of Cu-Pb Concentrate - Option VI

	Capital Cost (US$1000)
Process plant	48,950
Limestone	5,744*
S° disposal	609*
Effluent treatment	226*
Subtotal	**55,529**
Indirects (50%)	27,764
O_2 plant	6,136
Subtotal	**89,430**
Contingency (20%)	17,885
Total	**107,315**

*Note that capital for limestone mining/preparation, as well as allowances for sulphur in tailings and effluent treatment for the complete Option VI plant is included.

Operating Cost

The operating cost for low temperature leaching of a Cu-Pb concentrate with Process Options IV and VI is given in Table 2.42 below. A full breakdown is provided in Appendix A. The operating cost is different for each process plant due to a different cost of limestone.

Table 2.42 Operating Cost for Low Temperature Leaching of a Cu-Pb Concentrate - Options IV and VI

	Option IV	Option VI
	Operating Cost (US$1000)	**Operating Cost (US$1000)**
Process plant	12,677	12,563
O_2 plant	615	615
Total	**13,292**	**13,178**
$/t concentrate feed	**174.81**	**173.32**

2.4.2.5 Environmental Considerations

In this flowsheet, the selenium will largely deport to the leach residue as a S-Se product or as unreacted sulfide minerals (principally pyrite). Any selenium that is oxidized in the leach will end up as either calcium precipitate (calcium selenite/selenate), lead precipitate (lead selenite/selenate) or will be incorporated in iron residues (eg. jarosites or basic ferric sulfates). Selenium will not enter the copper electrowinning circuits due to the solvent extraction barrier to its transfer. This should ensure high purity final copper metal product. Similarly, the use of solvent extraction – electrowinning for zinc recovery should prevent any contamination of the final zinc cathode product.

It is possible that selenium will react with cyanide in the same way that elemental sulfur will react to form thiocyanate. The analogous species of selenium would be $SeCN^-$. This species is known to be difficult to treat. It is more stable with respect to oxidation than SCN^-.

2.4.2.6 Testwork Required for Prefeasibility

Leaching testwork will be required to demonstrate effective oxidation of the various minerals and to determine the deportment of selenium in the product and waste streams. Leaching will have to be studied as a function of time, temperature and oxygen pressure. The behaviour of selenium in acid neutralization should also be quantified. The oxidation products will need to be subjected to washing, lime boil and cyanidation testing. Key parameters will be % excess of lime for lime boil, temperature and time for lime boil and cyanidation conditions (time, pH and cyanide addition).

It will be necessary for prefeasibility to conduct a limited amount of testwork on copper and zinc SX-EW. This should consist in the preparation of standard McCabe – Thiele diagrams for copper and zinc extraction. It will not be necessary to perform any electrowinning of copper or zinc for this study.

2.4.3 Intensive Cyanidation for PM's Only

2.4.3.1 Process Description

A two stage intensive cyanidation test was conducted by International Metallurgical and Environmental Inc (Report dated September 23, 1999) on a sample of cleaned bulk copper-lead concentrate. The composition of the concentrate was as follows:

Au, g/t	9.99
Ag, g/t	2,996
Cu, %	9.6
Pb, %	9.8
Zn, %	9.7

Prior to cyanidation the concentrate was reground to a P_{80} of approximately 30 microns. During the initial 24 hours the cyanide concentration was maintained at 10 gpl, after which time the slurry was filtered and fresh solution added. An addition of 5 gpl lead nitrate was made to each stage of the test. The cyanide concentration was maintained at 5 gpl during the remaining 48 hours of the test.

While about 85% of the gold dissolved during the initial 24 hours, less than 10% of the silver or copper was dissolved. Negligible additional values dissolved during the subsequent 48 hours. Intensive cyanidation therefore does not appear to be a viable process option for this product.

No further estimate work was done since preliminary process tests indicated that leaching of the precious metals was poor.

2.4.4 Partial Biooxidation for Precious Metal Liberation

2.4.4.1 Process Description

The partial oxidation process is designed to oxidize a portion of the sulfide sulfur to sulfate sulfur and liberate a portion of the precious metals (Au, Ag) for subsequent cyanidation recovery. The process will extract some copper and zinc into solution, precipitate some lead as lead sulfate. However, the focus of the option is to liberate the precious metals for cyanidation; any copper an zinc leached will be precipitated by hydrolysis with limestone and lime and deposited in the tailings pond.

The conditions have not been established for the biooxidation process and its technical feasibility remains to be demonstrated. In concept, the desired results could be achieved by a single stage mesophilic bioleach together with reduced residence time.

The process used as the basis for this estimate uses a conventional mesophilic bioleach of the reground concentrate at 35°C. Air requirements and heat generation are based on the following extents of oxidation:

Pyrite	10%
Pyrrhotite	100%
Sphalerite	80%
Chalcopyrite	30%
Galena	50%

The bioleach slurry is thickened and neutralised. The recovered bioleach liquor is directed to neutralization for precipitation of dissolved copper, zinc, and iron. The resulting neutral slurry goes to tails. Elemental sulfur is removed by flotation. After a lime boil treatment, the bioresidue joins the sulfide flotation tails. The combined residue and tails are sent to cyanidation for precious metal recovery.

Process Chemistry

The partial bioleaching of reground concentrate will take place at 35°C. The main products of leaching will be partiallly oxidized sulfides, dissolved base metals, lead sulfate precipitate and hematite. It is expected that gold will be at least partially liberated in the product solids. Silver that is released by oxidation will likely form argentojarosite and require a hot lime boil to decompose to the cyanide soluble silver hydroxide species. Dissolved copper and zinc will precipitate by reaction with calcium hydroxide slurry.

Leaching

Sphalerite Leaching: $ZnS + 0.5O_2 + H_2SO_4 \rightarrow ZnSO_4 + S + H_2O$
Chalcopyrite Leaching: $CuFeS_2 + O_2 + 2H_2SO_4 \rightarrow CuSO_4 + FeSO_4 + S + H_2O$
Galena Leaching: $PbS + 0.5O_2 + H_2SO_4 \rightarrow PbSO_4(s) + S + H_2O$
Pyrrhotite/Marmatite Leaching: $FeS + 0.5O_2 + H_2SO_4 \rightarrow FeSO_4 + S + H_2O$
Pyrite Leaching: $FeS_2 + 0.5O_2 + H_2SO_4 \rightarrow FeSO_4 + 2S + H_2O$
Ferrous Sulfate Oxidation: $FeSO_4 + 0.25O_2 + 0.5H_2SO_4 \rightarrow 0.5Fe_2(SO_4)_3 + 0.5H_2O$
Elemental Sulfur Oxidation: $S + 1.5O_2 + H_2O \rightarrow H_2SO_4$
Hematite Precipitation: $Fe_2(SO_4)_3 + 3H_2O \rightarrow Fe_2O_3 + 3H_2SO_4$

Leachate Neutralization

Acid Neutralization: $CaO + H_2SO_4 + H_2O \rightarrow CaSO_4.2H_2O$
Iron Precipitation: $Fe_2(SO_4)_3 + 3CaO + 7H_2O \rightarrow 2FeO(OH) + 3CaSO_4.2H_2O$

Copper Precipitation

$$3CuSO_4 + 2CaCO_3 + 6H_2O \rightarrow CuSO_4 \cdot 2Cu(OH)_2 + 2CaSO_4 \cdot 2H_2O + 2CO_2$$

Zinc Precipitation

$$3ZnSO_4 + 2CaCO_3 + 6H_2O \rightarrow ZnSO_4 \cdot 2Zn(OH)_2 + 2CaSO_4 \cdot 2H_2O + 2CO_2$$

Lime Boil

Argentojarosite Decomposition:
$$AgFe_3(SO_4)_2(OH)_6 + 2Ca(OH)_2 \rightarrow AgOH + 2.5Fe_2O_3 + 2CaSO_4.2H_2O + 0.5H_2O$$

Cyanidation

Silver Leaching: $AgOH + 2NaCN \rightarrow NaAg(CN)_2 + NaOH$
Gold Leaching: $Au + 0.25O_2 + 2NaCN + 0.5H_2O \rightarrow NaAu(CN)_2 + NaOH$

Selenium Deportment

In this flowsheet, the selenium will largely deport to the leach residue as a S-Se product or as unreacted sulfide minerals (principally pyrite). Any selenium that is oxidized in the leach will end up as either calcium precipitate (calcium selenite/selenate), lead precipitate (lead selenite/selenate) or will be incorporated in iron residues (eg. jarosites or basic ferric sulfates).

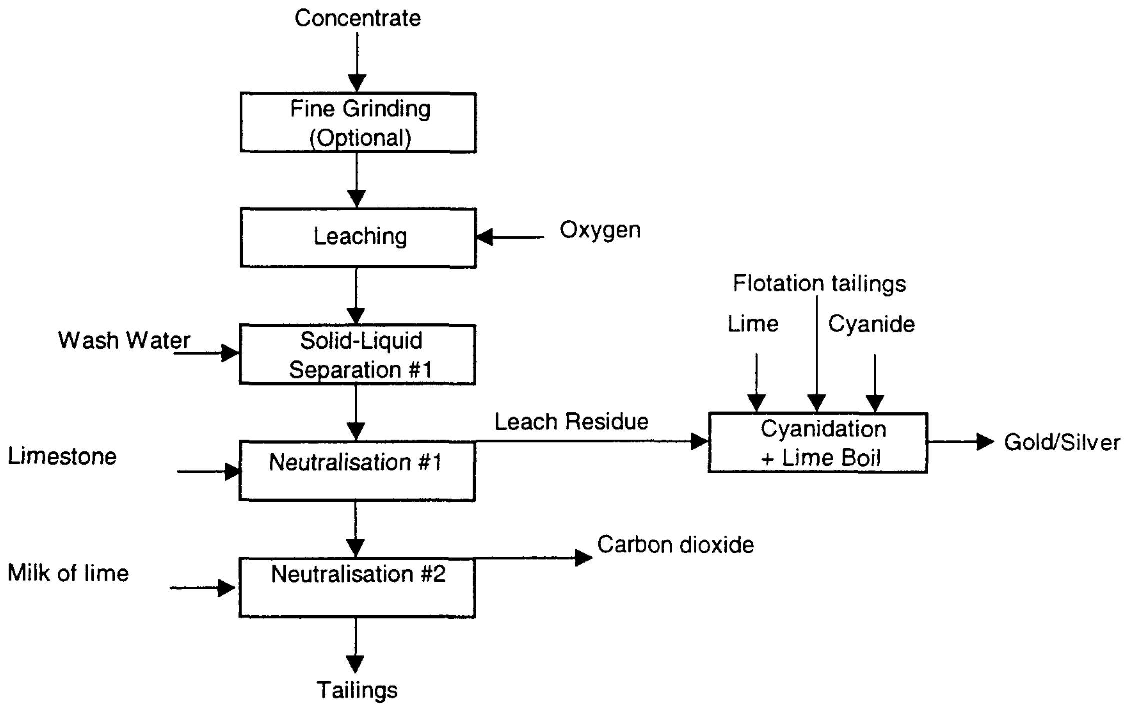

Figure 2.9 Partial Biooxidation Applied to a Cu/Pb Concentrate

2.4.4.2 Flowsheet

The flowsheet for the bioleach/CIP circuit is shown in Figure 2.9. The products are zinc calcine, gold and silver (impure), and various residues.

2.4.4.3 Basic Design Criteria

Design criteria used for the partial biooxidation of a Cu-Pb concentrate (Process Option VIII) are shown in Table 2.43 below:

Table 2.43 Partial Pressure Oxidation of Cu-Pb Concentrate

Mass Treated		tonnes/yr	76,037
Analysis	Copper	%	12
	Zinc	%	6
	Gold	g/tonne	12.5
	Silver	g/tonne	2,122
Products Recovered	Copper	tonnes/yr	0
	Zinc	tonnes/yr	0
	Gold	oz/yr	27,505
	Silver	oz/yr	4,669,307
Overall Recovery	Copper	%	0
	Zinc	%	0
	Gold	%	95
	Silver	%	95

2.4.4.4 Capital and Operating Costs

Capital Cost

The capital cost for the partial oxidation of a Cu-Pb concentrate is given in Table 2.44 below. A full breakdown is provided in Appendix A. Note that capital also has been included for limestone mining/preparation, sulphur in tailings and effluent treatment. This capital covers the complete process plant in each of these options.

Table 2.44 Capital Cost for Partial Biooxidation of Cu-Pb Concentrate – Option VIII

	Capital Cost (US$1000)
Process plant	10,500
S° disposal	0
Effluent treatment	0
Subtotal	**10,500**
Indirects (50%)	5,250
Subtotal	**15,750**
Contingency (20%)	3,150
Total	**18,900**

Operating Cost

The operating cost for partial bio-oxidation of a Cu-Pb concentrate is tabulated in Table 2.45 below. A full breakdown is provided in Appendix A.

Table 2.45 Operating Cost for Partial Biooxidation of a Cu-Pb Concentrate

	Operating Cost (US$1000)
Process plant including cyanidation	6,300
Total	**6,300**
$/t concentrate feed	**83.00**

2.4.4.5 Environmental Considerations

The most important environmental issues will relate to aqueous discharges and storage of tailings. The major residue stream will issue from the cyanidation plant. It will have been treated for cyanide destruction, most likely by an SO_2-air process operating on the final pulp. Some of its acid generating capabilities will have been destroyed as a result of the oxidation reactions; the elemental sulfur will have been largely removed by flotation. However, it is likely that this sulfur would be reintroduced with the tailings stream for discharge. There may be a need to add additional limestone to supply alkalinity that will ensure that no acid can be generated in the tailings impoundment in the future. The impoundment may require a lining to control leakage of the dissolved components remaining in the tailings water. The deposited solids will probably require a permanent water cover to reduce access to oxygen. If these measures are followed, the environment should be adequately protected from the potential effects of acid drainage and residual cyanide compounds.

Selenium may be oxidized to selenite/selenate, either directly by oxygen or indirectly through the action of ferric iron. This selenium will deport in a number of ways. Some will precipitate with lead during the oxidation reactions. The remaining oxidized selenium will precipitate with calcium derived from limestone or lime in the neutralisation processes. The stability of these products is not known relative to the stability of gypsum.

The overall water balance for this project has not been analyzed as yet. If zero discharge cannot be maintained, then a further treatment process may be required on the effluent discharge stream.

2.4.4.6 Testwork Required for Prefeasibility

The design criteria that were used for this option represent a best guess at how such a leach might proceed, but must be regarded as speculative. Issues to be resolved include the relation of gold and silver extraction to sulfide oxidation, the extent to which the various reactions proceed in an arrested oxidation situation, and particularly, the deportment of the sulfur between sulfides, elemental sulfur and sulfate. This latter information is essential to determine the actual heat balance and oxygen demand, both of which are of economic significance.

Leaching testwork will be required to demonstrate effective oxidation of the various minerals and to determine the deportment of selenium in the product and waste streams. Leaching will have to be studied as a function of time, temperature and oxygen pressure. The behavior of selenium in acid neutralization should also be quantified. The oxidation products will need to be subjected to washing, lime boil and cyanidation testing. Key parameters will be % excess of lime for lime boil, temperature and time for lime boil, and cyanidation conditions (time, pH and cyanide addition).

2.4.5 Bioleach for Smelter Preparation

2.4.5.1 Process Description

The process for concentrate processing is the same as described in section 2.4.4 above except the washed bioleach residue is filtered and shipped to a smelter for copper, silver and gold recovery. The cyanidation/CIP process circuit is omitted.

2.4.5.2 Flowsheet

The flowsheet is the same as shown in section 2.4.4 above except the cyanidation/CIP circuit is not included.

2.4.5.3 Basic Design Criteria

The design criteria is the same as that described in Section 2.4.4 above.

2.4.5.4 Capital and Operating Costs

The capital costs are similar to the capital costs shown in Section 2.4.4 above but have been reduced by approximately US$4.5 million by removing the cyanidation/CIP and associated waste treatment circuits as well as a slight reduction in infrastructure costs.

The operating costs for the copper concentrate process circuit have been reduced by approximately $50 per tonne compared with the process described in Section 2.4.4 above by elimination of the cyanidation/CIP and associated waste treatment costs. Smelting charges are not included in the operating cost comparison.

2.4.5.5 Environmental Considerations

The potential environmental impact at site is similar to that described in Section 2.4.4 above except there is no cyanide process risk and the associated disposal of dissolved cyanide species in the tailings pond. There is no cyanide destruction circuit.

Also, the copper concentrate is shipped off site, reducing the overall tailings pond metal loading and potential impact.

2.4.5.6 Testwork Required for Prefeasibility

Similar to the testwork described in Section 2.4.4 above except the testing of cyanidation conditions for copper concentrate processing and cyanide destruction is eliminated from the scope.

2.5 INFRASTRUCTURE

2.5.1 Summary of Infrastructure Costs

The infrastructure direct capital costs and the general/administration operating costs are given below for the nine process plant options.

Table 2.46 Infrastructure and General/Administration Costs

	Cost (US$1000)								
	Process Option								
	I	II	III	IV	V	VI	VII	VIII	IX
Infrastructure Total Capital Cost	76,885	75,295	75,993	75,526	75,993	69,484	69,484	66,984	66,984
G & A Expense and Labour	11,177	10,619	11,177	10,729	11,177	7,343	7,343	7,343	7,343
G & A $/tonne ore	10.21	9.70	10.21	9.80	10.21	6.71	6.71	6.71	6.71

Capital and operating costs for limestone mining and preparation at each different capacity are as shown in Table 2.47.

Table 2.47 Limestone Mining and Preparation Capital and Operating Costs

Production (t/y)	**271,763**	**72,845**	**226,935**	**37,982**	**107,093**	**155,296**	**224,406**	**140,000**	**140,000**
Process Option	**I**	**II**	**III**	**IV**	**V**	**VI**	**VII**	**VIII**	**IX**
Mining/Preparation Capital Cost (US$1000)	7,580	4,066	6,918	3,124	4,826	5,744	6,918	5,091	5,091
Mining (US$/tonne)	9.65	10.08	11.17	17.24	10.10	13.77	11.27	14.36	14.36
Processing (US$/tonne)	3.75	6.14	4.17	9.19	4.86	5.03	4.20	5.95	5.95
Total (US$1000)	**13.40**	**16.22**	**15.34**	**26.43**	**14.96**	**18.80**	**15.47**	**20.31**	**20.31**

For this study it is assumed that the products from the operation will be trucked to the port of Skagway for transhipment to an ocean freighter to market.

It is assumed that the calcine product will be shipped in bulk, similar to concentrate. The zinc and copper will be cast into ingots with the ingots being strapped onto pallets. The shipping costs are estimated as indicated in Table 2.48 below.

Table 2.48 Product Shipping Costs

	US$/Tonne	
	Calcine	**Zn & Cu Metal**
Palletization and Strapping	-	3
Trucking Minesite to Skagway	38	45
Skagway Terminal Charges	9	15
Ocean Freight	35	45
Total Transportation	**82**	**108**

2.5.2 Assumptions for Capital and Operating Cost Estimates

1. The plant will be adjacent to the KZK deposit.
2. A new road will be constructed between the KZK and Wolverine deposits – an estimated distance of 26 km.
3. All the power will be generated at the KZK site. A high tension power line will follow the road to the Wolverine deposit.
4. All manpower will be accommodated and fed from the KZK site – no allowance at Wolverine for food and accommodation.
5. The limestone deposit will be located within 10 km of the KZK site. Assume an extra 5 km of road needs to be built.
6. Limestone will be trucked to the KZK site for crushing and grinding (nominal 80% - 500 μm).
7. Busses will be required to transport the workers to the Wolverine and limestone sites.
8. The airstrip will be located adjacent to the Robert Campbell Highway and Finlayson Lake.
9. Calcine and copper and zinc metal will be transported to Skagway for export by ocean freighter.

2.5.3 Infrastructure Capital Cost Estimate

The infrastructure capital costs for the seven process options are summarized in Table 2.49. Details for each cost are discussed in the sections that follow.

Table 2.49 Infrastructure Capital Cost

	Capital Cost (1,000 US$)								
	Process Option								
	I	II	III	IV	V	VI	VII	VIII	IX
Site Access Road	8,300	8,300	8,300	8,300	8,300	8,300	8,300	8,300	6,500
Site Preparation	600	600	600	600	600	600	600	600	600
Airstrip	500	500	500	500	500	500	500	500	500
Fresh Water System	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000
Process Water System	500	500	500	500	500	500	500	500	500
Firewater System	150	150	150	150	150	150	150	150	150
Sewage Collection & Treatment	290	280	290	280	290	150	150	150	150
Main Sub Station	6,000	5,300	5,300	5,300	5,300	4,000	4,000	4,000	4,000
Power Line (Local)	1,040	1,040	1,040	1,040	1,040	1,040	1,040	1,040	1,040
Site Mobile Equipment	2,721	2,586	2,721	2,586	2,721	2,288	2,288	2,288	2,288
Communications	500	500	500	500	500	500	500	500	500
Camp Facilities	6,435	6,149	6,435	6,292	6,435	4,290	4,290	4,290	3,755
Office/Warehouse/ Shop	3,500	3,500	3,500	3,500	3,500	3,500	3,500	3,500	3,500
Cold Storage Warehouse	300	300	300	300	300	300	300	300	300
Assay Office	500	500	500	500	500	500	500	500	500
Infrastructure Direct Costs	32,336	31,355	31,786	31,498	31,786	27,768	27,618	27,618	25,283
Indirect Costs (35%)	11,318	10,974	11,125	11,024	11,125	9,719	9,719	9,719	9,719
Subtotal	**43,654**	**42,329**	**42,911**	**42,522**	**42,911**	**37,487**	**37,487**	**37,487**	**37,487**
Contingency (20%)	8,731	8,466	8,582	8,504	8,582	7,497	7,497	7,497	7,497
Power Line (Hayward)	24,500	24,500	24,500	24,500	24,500	24,500	24,500	0	0
Diesel Power								16,000	14,000
Total Infrastructure Cost	**76,885**	**75,295**	**75,993**	**75,526**	**75,993**	**69,484**	**69,484**	**56,958**	**54,958**

Indirect costs have been taken as 35% of the direct infrastructure costs and include the following items:

- temporary facilities
- construction plants
- construction equipment and consumables
- EPCM
- Owner's cost.

Excluded costs are:

- financing charges
- land acquisition and interest during construction
- permits, licenses and approvals
- ongoing capital requirements after construction
- cost escalation beyond date of estimate.

2.5.3.1 Site Access Road

This consists of upgrading the existing road from Robert Campbell Highway to KZK and constructing a new road 26 km from KZK to Wolverine. This needs to be a good two-lane all-weather highway as ore needs to be trucked between the two sites. The estimated road costs (US$1,000) are:

Upgrade road to KZK	300
Road KZK to Wolverine	8,000
	8,300

2.5.3.2 Site Preparation

This work is only being carried out at the KZK site and has been assumed to include all the non-mining aspects of the operation.

2.5.3.3 Airstrip

The 1,500 m long airstrip will be located at the junction of the KZK access road with the Robert Campbell Highway. This will be a gravel all-weather strip with a locator beacon and landing lights only. A heated trailer for traffic control and a waiting room (with diesel generator) is also provided. The airstrip is suitable for twin engine turbo prop commuter planes.

2.5.3.4 Fresh Water System

Fresh water will be supplied from wells adjacent to the site for the various processes, including potable water.

2.5.3.5 Process Water System

This covers internal recirculation of fresh water within the various processes.

2.5.3.6 Firewater System

This will distribute firewater around the plant and camp site with wall hydrants at strategic locations.

2.5.3.7 Sewage Collection and Treatment

The cost of this will depend on the manpower involved at the site.

2.5.3.8 Power Line (Hayward)

The total estimated capital from Ian Hayward International to construct a new 138 kVA transmission line (175 km) and upgrade an existing line (280 km) is US$24.5 million. Details are given in Appendix B.

2.5.3.9 Main Sub Station

The sub station will be adjacent to the plant and allow for the distribution of power to the various parts of the plant.

2.5.3.10 Power Line (Local)

The cost assumed for a high tension line (66 kv) from KZK to Wolverine (26 km) is US$40,000/km.

2.5.3.11 Site Mobile Equipment

The cost of the non-mining mobile equipment required for the operation is detailed in Table 2.50.

Table 2.50 Site Mobile Equipment

		Capital Cost (US$1000)																	
Process Option		I		II		III		IV		V		VI		VII		VIII		IX	
Equipment	**Unit Cost**	**No.**	**$**	**No.**	**$**	**No.**	**$**	**No.**	**$**	**No.**	**$**	**No.**	**$**	**No.**	**$**	**No.**	**$**	**No.**	**$**
Pickups	20	15	300	14	280	15	300	14	280	15	300	12	240	12	240	12	240	12	240
Ambulance	75	1	75	1	75	1	75	1	75	1	75	1	75	1	75	1	75	1	75
Front End Loader	175	2	350	2	350	2	360	2	350	2	350	2	350	2	350	2	350	2	350
Mobile Crane	150	1	150	1	150	1	150	1	150	1	150	1	150	1	150	1	150	1	150
5 Tonne HCAB	45	2	90	2	90	2	90	2	90	2	90	1	45	1	45	1	45	1	45
Tandem Axel Truck	80	2	160	2	160	2	160	2	160	2	160	1	80	1	80	1	80	1	80
Backhoe	90	1	90	1	90	1	90	1	90	1	90	1	90	1	90	1	90	1	90
Bobcat	18	2	36	2	36	2	36	2	36	2	36	1	18	1	18	1	18	1	18
Forklift 2 Tonne	25	2	50	2	50	2	50	2	50	2	50	2	50	2	50	2	50	2	50
Forklift 5 Tonne	40	2	80	2	80	2	80	2	80	2	80	2	80	2	80	2	80	2	80
Bus	115	6	690	5	575	6	690	5	575	6	690	4	460	4	460	4	460	4	460
Grader	200	2	400	2	400	2	400	2	400	2	400	2	400	2	400	2	400	2	400
Dozer	250	1	250	1	250	1	250	1	250	1	250	1	250	1	250	1	250	1	250
Total			**2,721**		**2,586**		**2,721**		**2,586**		**2,721**		**2,288**		**2,288**		**2,288**		**2,288**

This list does not include any equipment associated with the mining operation including:

	Unit Cost (US$1000)
Pickups	20
Lube Truck	125
Mechanics Truck	90
Tire Handler	35
Mine Rescue Truck	50

2.5.3.12 Communications

Telephone communication by satellite service for voice, fax and data will be required. Two-way radio networks will be provided for the various work areas.

2.5.3.13 Limestone Quarry

The quarry is assumed to be within 10 km of KZK with only 5 km of new road being required. A stripping ratio of 1:1 is assumed with the mine trucks hauling the limestone to the KZK plant site. Drilling and blasting is carried out on a contract basis.

Excluded from the limestone quarry costs in Table 2.51 are land acquisition, permitting and environmental requirements for this operation.

Table 2.51 Limestone Quarry Costs - 272,000 t/y Option

	US$1000
Access Road	1,000
Pre-Production	150
Loader	400
Haulage Trucks (4)	1,600
Miscellaneous	50
Total	**3,200**

Access road and pre-production costs for the limestone quarry are considered to be identical for all process plant options. Equipment costs have been reduced using the 0.6 power rule for the other limestone production rates.

2.5.3.14 Limestone Preparation

Equipment required in the limestone preparation area will include:

- Hopper
- Feeder
- Jaw crusher
- Conveyor (2)
- Storage bin
- Feeder
- Conveyor
- Grinding mill

- Pump and pump box
- Cyclone
- Agitated storage tank.

All equipment must be inside a building. The cost for this equipment to be installed in a building, for a production rate of 272,000 t/y, is as follows:

Table 2.52 Limestone Equipment Costs (US$1,000) - 272,000 t/y Option

	Unit Cost
Crushing and Storage	1,750
Milling and Slurry Storage	2,630
Total	**4,380**

Limestone preparation capital costs have been adjusted for the other limestone production rates using the 0.6 power rule.

2.5.3.15 Tailings Disposal – General

A flat rate of US$5 million direct cost has been used for each option. Indirects and contingency included give a final cost of US$8.1 million. The cost is slightly higher than was estimated for KZK but the material was thought to be potentially less reactive than with the current chemically treated residue.

2.5.3.16 Sulphur Disposal

An allowance based on the proposed tonnage to be disposed of has been made.

2.5.3.17 Effluent Treatment

A water treatment plant (H.D.S. Type) to handle the effluent from each option has been included in the cost estimate.

2.5.3.18 Camp Facilities

The crew sizes have been adjusted to take into account the limestone quarry and preparation crew. It has been assumed that each employee will have their own room. The size and cost of the camp for each process option is given in Table 2.53.

Table 2.53 Camp Facility Size/Cost

Option	Crew Size	Limestone Crew	Total Crew	Camp Size	Cost (US$1,000)
I	582	52	634	650	6,435
II	582	-	582	600	6,149
III	582	52	634	650	6,435
IV	597	-	597	610	6,292
V	597	36	633	650	6,435
VI	247	46	293	310	4,290
VII	247	52	299	310	4,290
VIII	215	42	257	265	3,755
IX	215	42	257	265	3,755

Costs have been factored, where required, using the 0.6 power rule.

2.5.3.19 Office/Warehouse/Shop

This complex will house the following:

- General Manager
- Accounting and Payroll
- Personnel
- Training
- Safety and First Aid
- Purchasing and Expediting
- Communications
- Warehouse
- Electrical/Instrumentation Shop
- Mechanical/Welding/Machine Shop
- Conference Room.

No allowance has been made for the mine engineering and operations group, mine dry or mine equipment maintenance shop.

The assumption has been made that the requirement for all options will be identical.

2.5.3.20 Cold Storage Warehouse

An across the board allowance has been made for all options to have a covered but unheated warehouse with a concrete floor and racking of items that do not require heated storage.

2.5.3.21 Assay Office

An equal allowance has been made for all options to provide the required quality control. This cost includes the building and equipment.

2.5.4 Operating Cost Estimate

2.5.4.1 Limestone Mining

The limestone mine operating costs for all production levels are given in Table 2.54. It is assumed that the stripping ratio is 1:1 and the quarry is within 10 km of the KZK plant.

2.5.4.2 Limestone Preparation

The operating costs for all limestone production levels are given in Table 2.55. It is assumed the limestone is crushed and ground to ~80% - 500 micron and that the slurry is stored at 50% solids in an agitated tank.

2.5.4.3 Overall Limestone Cost

The total limestone operating cost for all production levels is summarized in Table 2.56.

2.5.4.4 General and Administrative Expenses

Tables 2.57, 2.58 and 2.59 provide details of the G & A labour and expenses for the seven process options.

Table 2.54 Limestone Mining Operating Cost

Process Option	**I**		**II**		**III**		**IV**		**V**		**VI**		**VII**		**VIII**		**IX**	
Limestone Rate (t/y)	**271,763**		**72,845**		**226,935**		**37,982**		**107,093**		**155,296**		**224,406**		**140,000**		**140,000**	
Operating Labour	**No.**	**US $1000**	**No.**	**US $1000**	**No.**	**US $1000**	**No.**	**US $1000**	**No.**	**US $1000**	**No.**	**US $1000**	**No.**	**US $1000**	**No.**	**US $1000**	**No.**	**US $1000**
Truck Driver	10	359	2	72	10	359	2	72	3	108	8	287	10	359	5	215	5	215
Loader/Operator	5	220	2	88	5	220	2	88	3	132	5	220	5	220	5	220	5	220
Mechanic	1	48	0.5	24	1	48	0.3	16	0.5	24	1	48	1	48	1	48	1	48
Total	**16**	**627**	**4.5**	**184**	**16**	**627**	**4.3**	**176**	**6.5**	**264**	**14**	**555**	**16**	**627**	12	**483**	12	**483**
Fringe Benefits @ 35%		220	-	64	-	220	-	62	-	92	-	194	-	220		169		169
Total Labour (US$1000)		**847**		**248**		**847**		**238**		**356**		**749**		**847**		**652**		**652**
Operating Supplies **(US$1000)**		868		240		868		240		360		760		868		760		760
Repair Supplies **(US$1000)**		365		101		365		101		151		319		365		319		319
Total Cost (US$1000)		**2,080**		**589**		**2,080**		**579**		**867**		**1,828**		**2,080**		**1,731**		**1,731**
Subtotal Unit Cost (US$/t)		**7.65**		**8.08**		**9.17**		**15.24**		**8.10**		**11.77**		**9.27**		**12.36**		**12.36**
Drilling & Blasting (US$/t)		2.00		2.00		2.00		2.00		2.00		2.00		2.00		2.00		2.00
Total Unit Cost (US$/t)		**9.65**		**10.08**		**11.17**		**17.24**		**10.10**		**13.77**		**11.27**		**14.36**		**14.36**

Table 2.55 Limestone Preparation Operating Cost

Process Option	I		II		III		IV		V		VI		VII		VIII		IX	
Limestone Rate (t/y)	**271,763**		**72,845**		**226,935**		**37,982**		**107,093**		**155,296**		**224,406**		**140,000**		**140,000**	
Operating Labour	**No.**	**US $1000**	**No.**	**US $1000**	**No.**	**US $1000**	**No.**	**US $1000**	**No.**	**US $1000**	**No.**	**US $1000**	**No.**	**US $1000**	**No.**	**US $1000**	**No.**	**US $1000**
Crusher Operator	3	108	2	72	3	108	2	72	2	72	3	108	3	108	3	**108**	3	**108**
Grinding Operator	5	198	2	79	5	198	2	79	2	79	3	144	5	198	3	**144**	3	**144**
Mechanic	2	96	1	48	2	96	0.5	24	1	48	2	96	2	96	2	**96**	2	**96**
Supervisor	0.5	27	0.3	16	0.5	27	0.2	11	0.3	16	0.5	27	0.5	27	0.5	**27**	0.5	**27**
Total	**10.5**	**429**	**5.3**	**215**	**10.5**	**429**	**4.7**	**186**	**5.3**	**215**	**8.5**	**375**	**10.5**	**429**	**8.5**	375	**8.5**	375
Fringe Benefits @ 35%		150		75		150		65		75		131		150		**131**		**131**
Total Labour (US$1000)		**579**		**290**		**579**		**251**		**290**		**506**		**579**		506		506
Operating Supplies (US$1000)		91		24		76		13		36		52		75		**47**		**47**
Maintenance Supplies (US$1000)		136		51		113		34		75		84		112		**70**		**70**
Power @ 0.056 (US$/kWh)		**213**		**82**		**178**		**51**		**120**		**139**		**176**		110		110
Total Cost (US$1000)		**1,019**		**447**		**946**		**349**		**521**		**781**		**942**		833		833
Total Unit Cost (US$/t)		3.75		6.14		4.17		9.19		4.86		5.03		4.20		**5.95**		**5.95**

Table 2.56 Total Limestone Operating Cost - Mining and Preparation

Process Option	I	II	III	IV	V	VI	VII	VIII	IX
Limestone Rate (t/y)	**271,763**	**72,845**	**226,935**	**37,982**	**107,093**	**155,296**	**224,406**	**140,000**	**140,000**
Unit Cost (US$/t)									
Mining	9.65	10.08	11.17	17.24	10.10	13.77	11.27	14.36	14.36
Preparation	3.75	6.14	4.17	9.19	4.86	5.03	4.20	5.95	5.95
Total (US$/t)	**13.40**	**16.22**	**15.34**	**26.43**	**14.96**	**18.80**	**15.47**	**20.31**	**20.31**

Table 2.57 General and Administrative Expenses

	Annual Expenses (1,000 x US$)								
Process Option	**I**	**II**	**III**	**IV**	**V**	**VI**	**VII**	**VIII**	**IX**
Administration									
Communications	30	30	30	30	30	30	30	30	30
Travel Expense (not commute)	20	20	20	20	20	20	20	20	20
Insurance	500	500	500	500	500	500	500	500	500
Office Supplies	15	15	15	15	15	15	15	15	15
Computer Operation	30	30	30	30	30	30	30	30	30
Legal and Audit	35	35	35	35	35	35	35	35	35
Property Taxes	400	400	400	400	400	400	400	400	400
Misc. Outside Consulting	50	50	50	50	50	50	50	50	50
Community and Government Relations	15	15	15	15	15	15	15	15	15
Miscellaneous	35	35	35	35	35	35	35	35	35
Sub-total	**1,130**	**1,130**	**1,130**	**1,130**	**1,130**	**1,130**	**1,130**	**1,130**	**1,130**
Camp									
Camp Costs @ $8,000/Employee	5,200	4,800	5,200	4,880	5,200	2,480	2,480	2,120	2,120
Transportation									
Bus Operation	50	42	50	42	50	33	33	29	29
Airstrip and Security	250	250	250	250	250	200	200	200	200
Road Maintenance	70	70	70	70	70	70	70	70	70
Commuter Airline @ $3,000/Employee	1,950	1,800	1,950	1,830	1,950	930	930	795	795
Sub-Total	**2,320**	**2,162**	**2,320**	**2,192**	**2,320**	**1,233**	**1,233**	**1,094**	**1,094**
Safety and Personnel									
First Aid Supplies	10	10	10	10	10	10	10	10	10
Mine Rescue Supplies	10	10	10	10	10	10	10	10	10
Fire Prevention	10	10	10	10	10	10	10	10	10
Training	100	100	100	100	100	100	100	100	100
Medical Examinations	35	35	35	35	35	18	18	18	18
Recruitment	25	25	25	25	25	15	15	15	15
Safety Awards and Misc.	50	50	50	50	50	50	50	50	50
Sub-Total	**240**	**240**	**240**	**240**	**240**	**213**	**213**	**213**	**213**
Environmental Control									
Outside Consultants	50	50	50	50	50	50	50	50	50
Outside Testing	50	50	50	50	50	50	50	50	50
Sub-Total	**100**	**100**	**100**	**100**	**100**	**100**	**100**	**100**	**100**
Surface Crew									
Miscellaneous Supplies	50	50	50	50	50	50	50	50	50
Total G & A Expenses	**9,040**	**8,482**	**9,040**	**8,592**	**9,040**	**5,206**	**5,206**	**4,707**	**4,707**

Table 2.58 General and Administrative Operating Labour

	No.	Annual Salary*	Total (US$1000)
Staff			
General Manager	1	85,000	85
Secretary	2	31,000	62
Personnel Superintendent	1	69,000	69
Personnel Officer	1	55,000	55
Native Coordinator	1	55,000	55
Purchasing Agent	1	60,000	60
Buyer	3	40,000	120
Warehouseman	6	32,300	194
Safety & Training Officer	2	52,000	104
First Aid	2	48,000	96
Comptroller	1	60,000	60
Accountant	2	45,000	90
Payroll Clerk	2	28,000	56
Accounting Clerk	2	32,000	64
Systems Coordinator	1	45,000	45
Surface Foreman	2	50,000	100
Mobile Equipment Foreman	1	50,000	50
Subtotal	**31**	-	**1,365**
Hourly			
Lead Hand	2	48,000	96
Truck Driver	2	39,500	79
Bus Driver	4	39,500	158
Equipment Operator	4	43,200	173
Labourer	2	28,500	57
Janitor	2	25,000	50
Lead Hand Mechanic	1	50,000	50
Journeyman Mechanic	1	48,000	48
Apprentice	2	30,500	61
Subtotal	**20**	-	**772**
Total	**51**	-	**2,137**

*** Salary includes 35% fringe benefits**

Table 2.59 Total General and Administrative Expenses and Labour

	Annual G & A Costs (US$1000)							
Process Option	**I**	**II**	**III**	**IV**	**V**	**VI**	**VII**	**VIII & IX**
G & A Annual Expenses	9,040	8,482	9,040	8,592	9,040	5,206	5,206	4,707
G & A Labour	2,137	2,137	2,137	2,137	2,137	2,137	2,137	2,137
Total G & A Cost	**11,177**	**10,619**	**11,177**	**10,729**	**11,177**	**7,343**	**7,343**	**6,844**
G & A Unit Cost ($/t)	**10.21**	**9.70**	**10.21**	**9.80**	**10.21**	**6.71**	**6.71**	**6.25**

3.0 COST SUMMARY

The capital and operating costs for each process alternative were modularized for process comparative purposes. AGRA Simons reviewed the cost basis and provided "fill-in" costing as required to allow comparative cost analysis.

The results of the cost analysis are summarized in Table 3.1, overleaf.

3.1 Process Discussion

Based on the results of the comparative cost estimates, none of the hydrometallurgical processing alternatives that produce a refined metal product at site, Cases 1-5, warrant further investigation. In addition, Cases 6 and 7, where the zinc concentrate is roasted and shipped to smelter and a Cu-Pb concentrate is further processed at site for copper, silver and gold refining, do not warrant further work, due to comparative capital risk. It is unlikely that without a change in scope, further engineering and estimating at this time will improve the costing of these alternatives to the point where prefeasibility work is warranted.

Cases 8 and 9 show the probable processing direction to pursue if the zinc and copper concentrates cannot be marketed as produced. Both process flowsheets employ conventional process technology. There is lower capital risk for both flowsheets. Also, both flowsheets require significantly less power and therefore a dedicated diesel power plant is included in the infrastructure rather than the coal fired power plant and power transmission lines included for Cases 1 - 7.

Case 9 is more attractive than Case 8 due to the incremental value of copper payables. However, the merits of Case 9 are dependent on marketing a concentrate that contains significant potential penalty elements (Se and Sb), the market for which has yet to be determined. Note that the treatment charge assumed for the copper concentrate in Case 9 is $185/t, which is double the normal treatment charge for copper concentrates due to expected penalties.

There is additional upside to Cases 8 and 9, not included in this study, which includes: higher concentrate grade to the smelter, lower shipping costs and potential lower treatment charges at the smelter for the calcine product, all of which will off-set, in part, the incremental cost of roasting.

It should be noted that Cases 8 and 9 will require significant cost allowances, both capital and operating, to ensure acceptable waste treatment and storage is included. The roaster, a conventional fluidized bed, will include standard gas scrubbing for mercury, selenium and arsenic collection. The bioleach step will solubilize arsenic, antimony and selenium in a sulphate solution, which will require standard neutralisation and precipitation with limestone for storage with the mill tailings. Long term stability testing of the waste products will be required for permitting purposes.

Table 3.1 Cost Summary

	PROCESS OPTION								
Capital Costs (US$ X 1,000)	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**
Concentrator	37,854	37,854	37,854	40,104	40,104	40,104	40,104	40,104	40,104
Tailings	8,100	8,100	8,100	8,100	8,100	8,100	8,100	8,100	8,100
Process Plant	259,018	233,722	197,190	271,543	293,357	150,515	170,855	66,489	64,000
Infrastructure	76,885	75,295	75,993	75,526	75,993	69,484	69,484	57,000	55,000
Total	381,857	354,971	319,137	395,273	417,554	268,203	288,543	171,693	167,204
Operating Costs									
Concentrator, $/tonne	7.18	7.18	7.18	7.8	7.8	7.8	7.8	7.8	7.8
Zn Process, $/tonne conc.				204.17	202.24	51	48	56	56
Cu Process, $/tonne conc.				174.81	165.23	173.32	165.7	80	28
Bulk Conc, $/tonne conc.	223.32	231.28	230.76						
G&A, $/tonne	10.21	9.7	10.21	9.8	10.21	6.71	6.71	6.25	6.25
Production									
Zinc (t/a)	77,095	77,095	77,095	76,408	76,408	61,069	61,069	61,069	61,069
Copper (t/a)	9,818	9,818	9,818	9,819	9,819	8,987	8,987	-	8,668
Gold (ounces)	35,742	35,742	35,742	34,976	34,976	31,891	31,891	27,505	31,891
Silver (ounces)	5,840,225	5,840,225	5,840,225	5,839,708	5,839,708	5,347,000	5,347,000	4,669,307	5,347,000

3.2 Sensitivity Analysis

The project will be most sensitive to metal pricing and operating costs. The mining, milling and infrastructure costs are the most important operating costs to optimize in the prefeasibility study, due to the potential impact on project economics.

The processing rate will not change the economics significantly, within reasonable expectations of the reserve base, if the mill feed grade does not change. Optimization of the mine plan to provide somewhat higher grade material in the first 3-4 years will be important for project economics.

4.0 CONCLUSIONS AND RECOMMENDATIONS

4.1 Conclusions

The process selected for further development is described in Case 8 and Case 9, subject to the marketing study currently in progress for zinc calcine and bioleached copper concentrate smelting. The focus of future work should be on reduction of capital cost by 10 - 20% combined with a reduction in operating cost of approximately 20% and a detailed analysis of mine and infrastructure costs. These are realistic expectations based on the scoping level results.

From the results of the cost estimates, there is a reasonable expectation that there are at least two viable process alternatives to treat a combined Wolverine/Kudz Ze Kayah mill feed. If the zinc and copper concentrates cannot be sold to a smelter as produced, Cases 8 and 9 show potential for positive project economics. Additional costing detail and market value of the calcine will be required to provide a more definitive assessment of both of the process options.

4.2 Recommendations

4.2.1 Process Selection

The most favourable process alternative to develop the Wolverine and Kudz Ze Kayah deposits is direct shipping of zinc and copper concentrates, subject to smelter treatment charges and long term contracts. A co-venture with a smelter, where additional smelter capacity for selenium handling is included in the project capital, is the most likely development alternative.

However, Cases 8 and 9 offer process alternatives at site that show potential for favourable economics, within the sensitivity limits of a scoping study. It is recommended that, following the response from smelters on treatment of products from Case 8 and Case 9, prefeasibility level work is initiated to define Cases 8 and 9 metallurgy and waste product stability. Concurrently, a more rigorous concentrate marketing study should be completed, if warranted.

4.2.2 Prefeasibility Study

Terms of Reference

The prefeasibility study will be based on new cost estimates at an accuracy of +/- 25%. Mine plans and scheduling will be developed using the existing geological and drilling database from each deposit. Processing will be based on new prefeasibility level metallurgical testwork using existing samples.

The scope will include open pit and underground mine planning, metallurgical flowsheet and mass balance development for a single process alternative, general arrangement drawings for quantity estimating, capital and operating cost estimates, financial and sensitivity analysis and recommendations and costing for future drilling and metallurgical testwork to complete a feasibility study.

Mining

The project will be sensitive to mining costs. Therefore, additional engineering is required to develop mine planning to a point where reasonable production estimates and scheduling can be completed. The existing database for each deposit should allow a prefeasibility estimate of mine costs.

Infrastructure

The project is highly sensitive to infrastructure capital and operating costs. There are no previous infrastructure costs developed to an engineering level that would allow a prefeasibility estimate. New costs must be developed for the infrastructure.

Processing

A single flowsheet will be required for the prefeasibility study. The flowsheet selected will depend on the initial response from smelters, currently in progress. Metallurgical testing will be contingent on the flowsheet selected. However, crush/grind and flotation work can proceed in the interim, subject to sample availability from the KZK deposit. Some of the testwork can be initiated using only Wolverine samples, however, for prefeasibility level analysis, samples of the KZK deposit will be required.

Integral in the processing testwork is definition and stability testing of the process waste products.

Estimate of Incremental Costs

The incremental cost to complete a prefeasibility level report using the Scoping Study as the basis has been estimated by AGRA Simons at 1,850 manhours totalling CDN$220,000.

4.2.3 Metallurgical Testwork Programme

Description of Scope

The metallurgical test programme to support a prefeasibility study will include the following, subject to the results of the marketing study:

- Crush/grind work index
- Differential flotation, batch and locked cycle testing
- Batch roasting tests on the zinc concentrate
- Bioleach testing of a bulk Cu-Pb concentrate or a copper concentrate, batch and continuous testing
- Solid/liquid separation testing on metallurgical and waste products
- Stability testing on all waste products, static and kinetic testing

The testing should monitor the deportment of Cu, Pb, Zn, Au, Ag, Fe, S (and sulphate), Se, As and Sb. All concentrate products scheduled for smelter analysis should include a whole rock analysis and multi-element assay for smelter diluents.

Sample Requirements

Minimum sample requirements for the metallurgical testwork include:

1. 100 kg Wolverine “average grade” sample from drill core or crushed sample stored in N2.
2. 200 kg of Kudz Ze Kayah “average grade” sample from drill core or crushed sample stored in N2.
3. 25 kg from each deposit of material from the first 3 years of production.

Schedule

Following is an approximate schedule for planning purposes:

January 00:	Completion of initial marketing study – decision point Initiate sample collection and preparation
February 00:	Crushing/Grinding indexes Initial batch flotation testing
March 00:	Batch flotation followed by locked cycle Roast scoping testwork Bioleach batch initiation
April 00:	Final flotation locked cycle work Continuous bioleach Final roasting testing Solid/liquid separation testing Static stability testing on waste products
May 00:	Market study on actual products Continuation of continuous bioleach Long term kinetic stability testing (6-12 months) Cyanidation testing
June 00:	Completion of bioleach and cyanidation testing

The mining and infrastructure development as well as flowsheet development and costing will be ongoing during the metallurgical testwork.

Estimate of Costs

The costs to complete the metallurgical testwork as outlined will be approximately:

Crush/grind	$4,000
Flotation	$35,000
Roasting	$15,000
Bioleaching	$20,000
Solid/liquid separation	$3,000
Stability testing	$21,000
Microprobe analysis	$6,000
Assays	$11,000
Subtotal	**$115,000**
Management and reporting	$16,000
Total	**$131,000**

Cost Summary - Total Oxidation of Bulk Concentrate

Area	**Option I Installed Cost** (US$)	**Option I Operating Cost** ($/tonne feed)
Fine Grinding	811,152	0.99
Autoclave (not including Oxygen Plant)	33,000,000	11.62
Slurry Cooling	4,000,000	0.65
Neutralization #1	1,606,384	13.42
Solid-Liquid Separation #1	2,665,259	1.44
Copper SX-EW	21,570,402	13.36
Lime Boil	1,373,893	6.44
Leach/CIP	4,143,272	9.39
Gold Refinery	5,379,084	6.95
Iron Precipitation	2,195,801	2.53
Solid-Liquid Separation #2	971,211	0.40
Zn SX-EW	49,288,339	131.69
Neutralization #2	570,002	2.30
Solid-Liquid Separation #3	870,260	0.35
Sub Total	128,445,061	201.54
Limestone(Quarry&Prep)	7,580,000	(in above costs)
Sulphur Disposal	0	
Effluent Treatment	540,000	
Indirect Costs	68,282,531	
Oxygen Plant	11,000,000	21.78
Total	215,847,592	223.32 $/mt feed
Contingency(20%)	43,169,518	
TOTAL	259,017,110	
	Total Op. Cost	50,246,741 $/annum

Cost Summary - Dynatec Leach of Bulk Concentrate

Area	**Option II Installed Cost** (US$)	**Option II Operating Cost** ($/tonne feed)
Fine Grinding	682,558	0.87
Autoclave (not including Oxygen Plant)	29,000,000	10.83
Slurry Cooling	4,000,000	0.65
Neutralization #1	1,372,167	1.97
Solid-Liquid Separation #1	1,552,363	0.71
Copper SX-EW	19,042,073	12.96
Lime Boil	792,846	13.92
Leach/CIP	2,391,000	36.82
Gold Refinery	5,207,385	6.62
Iron Precipitation	1,968,864	1.78
Solid-Liquid Separation #2	1,003,896	0.41
Zn SX-EW	47,009,148	130.48
Neutralization #2	588,827	3.27
Solid-Liquid Separation #3	990,393	0.40
Sub Total	115,601,519	221.70
Limestone(Quarry&Prep)	4,066,000	(in above costs)
Sulphur Disposal	1,000,000	
Effluent Treatment	323,000	
Indirect Costs	60,495,260	
Oxygen Plant	13,015,355	9.57
Total	194,501,134	231.28 $/mt feed
Contingency(20%)	38,900,227	
TOTAL	233,401,361	
Total Op. Cost		52,036,886 $/annum

Cost Summary - Bioleaching of Bulk Concentrate

Area	Option III Installed Cost (US$1000)	Option III Operating Cost ($/tonne feed)
Regrind and Feed Prep	1,915	2.31
Primary bioleachand L/S sep'n	14,041	24.36
Secondary bioleachand L/S sep'n	4,944	6.47
Neutralization and Boil	466	35.01
Sulphur Flotation	634	1.22
Zn SX-EW	49,288	131.69
Cu SX-EW	21,570	13.36
Leach/CIP	4,143	9.39
Gold Refinery	5,379	6.95
Sub Total	102,092	230.76
Limestone(Quarry&Prep)	6,918	(in above costs)
Sulphur Disposal	0	
Effluent Treatment	540	
Indirect Costs	54,775	
Total	164,325	
Contingency(20%)	32,865	
TOTAL	197,190	
	Total Op. Cost	51,921,000 $/annum

Cost Summary - Dynatec Zinc Concentrate Pressure Leach

Area	Option IV Installed Cost (US$)	Option V Installed Cost (US$)	Option IV Operating Cost ($/tonne feed)	Option V Operating Cost ($/tonne feed)
Fine Grinding	682,558	682,558	1.10	1.10
Autoclave (not including Oxygen Plant)	29,000,000	29,000,000	13.09	13.09
Slurry Cooling	4,000,000	4,000,000	1.06	1.06
Neutralization #1	1,018,802	1,018,802	2.69	1.80
Solid-Liquid Separation #1	1,014,387	1,014,387	0.67	0.67
Copper SX-EW	5,000,000	5,000,000	1.85	1.85
Lime Boil	523,083	523,083	9.02	9.02
Leach/CIP	1,577,471	1,577,471	33.75	33.75
Gold Refinery	1,417,813	1,417,813	1.24	1.24
Iron Precipitation	1,547,438	1,547,438	1.30	1.30
Solid-Liquid Separation #2	397,881	397,881	0.22	0.22
Zn SX-EW	40,628,378	40,628,378	127.70	127.70
Neutralization #2	199,189	199,189	2.42	1.37
Solid-Liquid Separation #3	772,659	772,659	0.48	0.48
Sub Total	87,779,660	87,779,660	196.60	194.67
Limestone(Quarry&Prep)	3,124,000	0		(in above costs)
Sulphur Disposal	957,000	0		
Effluent Treatment	310,000	0		
Indirect Costs	46,085,330	43,889,830		
Oxygen Plant	8,469,085	8,469,085	7.57	7.57
Total	146,725,074	140,138,574	204.17	202.24 $/mt feed
Contingency(20%)	29,345,015	28,027,715		
TOTAL	176,070,089	168,166,289		
		Total Op. Cost $/annum	28,373,234	28,104,209

Cost Summary - Roast of Zinc Concentrate

Area	Options VI & VII Installed Cost ($)	Option VI Operating Cost ($/tonne feed)	Option VII Operating Cost ($/tonne feed)	
Roasting	15,000,000	21.00	21.00	
Gas Cleaning and Scrubbing	9,000,000	11.00	11.00	
Limestone for Neutralization		19.00	16.00	
Sub Total	24,000,000	51.00	48.00	
Limestone(Quarry&Prep)	0			
Sulphur Disposal	0			
Effluent Treatment	0			
Indirect Costs	12,000,000			
Contingency	2,400,000			
Total	38,000,000	51.00	48.00	
	Total Op. Cost	7,100,000	6,700,000	$/annum
		0.045	0.043	per lb Zn

Cost Summary - Roast of Zinc Concentrate

Area	Options VIII & IX Installed Cost ($)	Option VIII & IX Operating Cost ($/tonne feed)	
Roasting	15,000,000	24.00	
Gas Cleaning and Scrubbing	9,000,000	11.00	
Limestone for Neutralization		21.00	
Sub Total	24,000,000	56.00	
Limestone(Quarry&Prep)	5,097,393		
Sulphur Disposal	0		
Effluent Treatment	310,000		
Indirect Costs	14,703,697		
Contingency	2,940,739		
Total	47,000,000	56.00	
	Total Op. Cost	7,800,000	$/annum
		0.050	per lb Zn

Cost Summary - Dynatec Leach of Cu-Pb Concentrate

Area	Option IV Installed Cost (US$)	Option VI Installed Cost (US$)	Option IV Operating Cost ($/tonne feed)	Option VI Operating Cost ($/tonne feed)
Fine Grinding	682,558	682,558	1.59	1.59
Autoclave (not including Oxygen Plant)	14,500,000	14,500,000	14.01	14.01
Slurry Cooling	2,000,000	2,000,000	0.97	0.97
Neutralization #1	616,904	616,904	1.68	1.45
Solid-Liquid Separation #1	944,935	944,935	1.13	1.13
Copper SX-EW	12,123,921	12,123,921	35.34	35.34
Lime Boil	468,881	468,881	22.71	22.71
Leach/CIP	1,385,490	1,385,490	43.52	43.52
Gold Refinery	4,840,401	4,840,401	17.26	17.26
Iron Precipitation	493,820	493,820	1.84	1.47
Solid-Liquid Separation #2	431,058	431,058	0.44	0.44
Zn SX-EW	9,759,586	9,759,586	22.55	22.55
Neutralization #2	313,895	313,895	3.29	2.40
Solid-Liquid Separation #3	388,405	388,405	0.39	0.39
Sub Total	48,949,854	48,949,854	166.72	165.23
Limestone(Quarry&Prep)	0	5,744,000	(in above costs)	(in above costs)
Sulphur Disposal	0	609,000		
Effluent Treatment	0	226,000		
Indirect Costs	24,474,927	27,764,427		
Oxygen Plant	6,135,864	6,135,864	8.09	8.09
Total	79,560,645	89,429,145	174.81	173.32
Contingency(20%)	15,912,129	17,885,829		
TOTAL	95,472,774	107,314,974		
		Total Op. Cost $/annum	13,291,774	13,178,363

Cost Summary - Total Oxidation of Cu-Pb Concentrate

Area	**Option V Installed Cost** ($)	**Option VII Installed Cost** ($)	**Option V Operating Cost** ($/tonne feed)	**Option VII Operating Cost** ($/tonne feed)
Fine Grinding	535,161	535,161	1.23	1.23
Autoclave (not including Oxygen Plant)	16,500,000	16,500,000	15.05	15.05
Slurry Cooling	2,000,000	2,000,000	0.97	0.97
Neutralization #1	799,999	799,999	15.13	15.61
Solid-Liquid Separation #1	1,518,118	1,518,118	2.05	2.05
Copper SX-EW	14,000,000	14,000,000	36.45	36.45
Lime Boil	710,690	710,690	10.97	10.97
Leach/CIP	2,004,923	2,004,923	16.34	16.34
Gold Refinery	5,000,000	5,000,000	18.11	18.11
Iron Precipitation	680,543	680,543	2.22	2.27
Solid-Liquid Separation #2	437,092	437,092	0.45	0.45
Zn SX-EW	11,457,079	11,457,079	22.81	22.81
Neutralization #2	306,605	306,605	1.34	1.28
Solid-Liquid Separation #3	388,200	388,200	0.39	0.39
Sub Total	56,338,410	56,338,410	143.50	143.97
Limestone(Quarry&Prep)	4,826,000	6,918,000	(in above costs)	(in above costs)
Sulphur Disposal	653,000	0		
Effluent Treatment	400,000	330,000		
Indirect Costs	31,108,705	31,793,205		
Oxygen Plant	11,000,000	11,000,000	21.73	21.73
Total	104,326,114	106,379,614	165.23	165.70
Contingency(20%)	20,865,223	21,275,923		
TOTAL	125,191,337	127,655,537		
		Total Op. Cost $/annum	12,563,834	12,599,694

Cost Summary - Partial Biooxidation of Cu-Pb Concentrate

Mass Treated	tonnes/yr	76037

Analysis

Copper	%	12
Zinc	%	6
Gold	g/tonne	12.5
Silver	g/tonne	2122

Products Recovered

Copper	tonnes/yr	0
Zinc	tonnes/yr	0
Gold	oz/y	27,505
Silver	oz/y	4,669,307

Overall Recovery

Copper	%	0.00
Zinc	%	0.00
Gold	%	90
Silver	%	90

Cost Summary

Area		Installed Cost ($)	Operating Cost ($/tonne feed)
Bioleach plant		1,435,574	28.00
Lime boil		468,881	3.11
Leach/CIP		3,953,776	19.20
Gold Refinery		4,677,214	32.41
	Sub Total	10,500,000	83.00
	Indirect Costs	5,250,000	
	Total	15,750,000	83.00
	Contingency	3,150,000	
	TOTAL	18,900,000	

Total Op. Cost 6,300,000 $/annum

APPENDIX B

Refer to "Expatriate Resources Ltd. 138 kV Transmission Line Prefeasibility Study", prepared by Ian Hayward International Ltd., November 1999. (Not bound with this report).

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Tel/fax +44-1344-874704

Monday, 20 September 1999

Brad Marchant
Biomet Mining Corporation
Unit 360
3771 Jacombs Road
Richmond BC
V6V 2L9
Canada

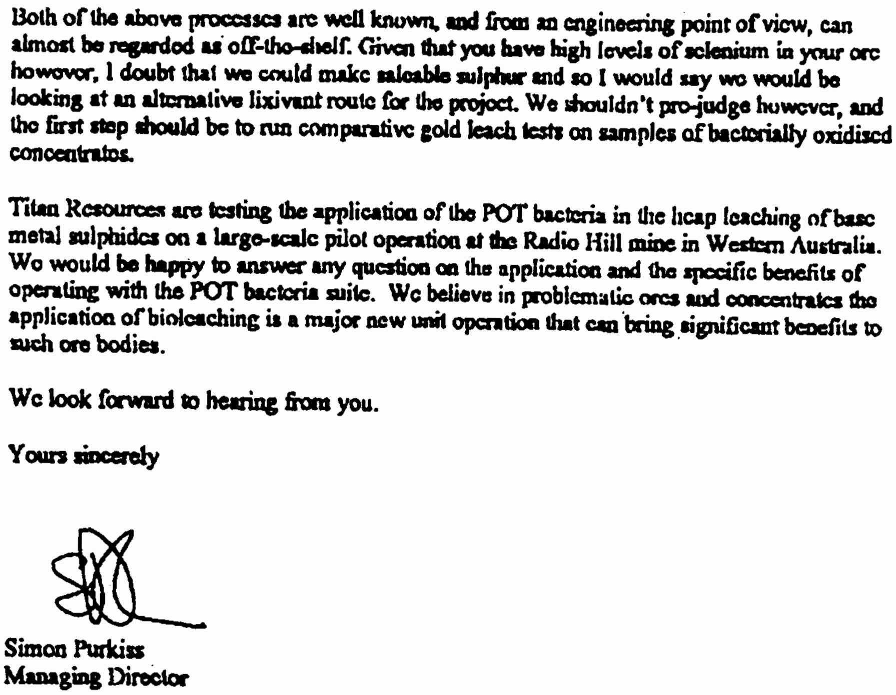
Dear Brad,

Re Wolverine: Copper/Gold Bio-Leach

In reply to your recent fax to Ross Gilders at RPC with some questions on the bioleach route that we would propose for the Wolverine deposit. I hope the following serves to clarify matters.

We are able to run our bacterial leach process over a wide range of pH and temperature. The upper pH limit is set by iron precipitation, which tends to come into play at above pH2.0, though even above this the leach can be made to work under certain conditions. At the other end of the scale we have operated very well at pH 0.8 and have yet to find a lower limit of operation. The temperature range for optimum leach kinetics is between 40°C and 65°C. The pulp density of a tank leach is nominally 20% solids (w/w), in gold ore processing. With base metals on the other hand, the density tends to be lower, closer to 15%, though this is dictated by the final solution tenors obtainable from the leach and the tolerance of the bacteria themselves to high base metal concentrations in solution.

Our copper specific bacterial culture does not oxidise the sulphur contained in the mineral to sulphate but rather to elemental sulphur. This has the advantage of requiring less oxygen and producing less heat, but has the disadvantage of making subsequent cyanide consumption unacceptably high. This can be controlled by introducing a sulphur oxidising bacteria to the system and generating a mixed culture. Even then, the cyanide consumption would be high because of the residual copper and elemental sulphur, and 5 to 10kg/t NaCN might be expected. This situation can be avoided by one of two ways. Firstly, the use of an alternative lixivant that is not effected by either copper or sulphur, i.e. ammonium thiosulphate. There is a great deal of information about this leach in the literature, especially in the Randol volumes, and more recently, Newmont have developed the process for their gold bio-heap leach process. The second alternative is to employ the perchlorethylene process for recovering the sulphur as a saleable product. This process was first used by Falconbridge and has now been taken up by CESL specifically to enable gold recovery from their pressure leach residues.



Both of the above processes are well known, and from an engineering point of view, can almost be regarded as off-the-shelf. Given that you have high levels of selenium in your ore however, I doubt that we could make saleable sulphur and so I would say we would be looking at an alternative lixivant route for the project. We shouldn't pre-judge however, and the first step should be to run comparative gold leach tests on samples of bacterially oxidised concentrates.

Titan Resources are testing the application of the POT bacteria in the heap leaching of base metal sulphides on a large-scale pilot operation at the Radio Hill mine in Western Australia. We would be happy to answer any question on the application and the specific benefits of operating with the POT bacteria suite. We believe in problematic ores and concentrates the application of bioleaching is a major new unit operation that can bring significant benefits to such ore bodies.

We look forward to hearing from you.

Yours sincerely

Simon Purkiss
Managing Director