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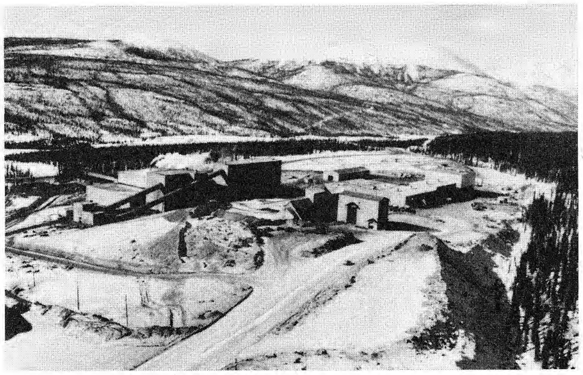
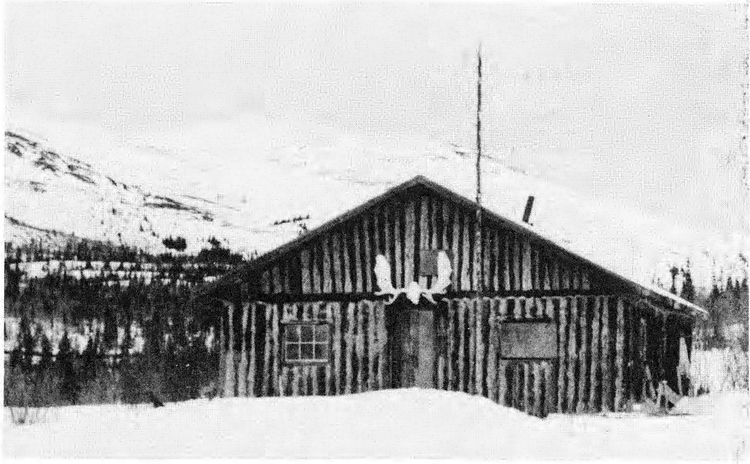
THE DEVELOPMENT OF THE VANGORDA PLATEAU ORE DEPOSITS

Technical Data Summary
&
Mill Modifications

CYPRUS ANVIL





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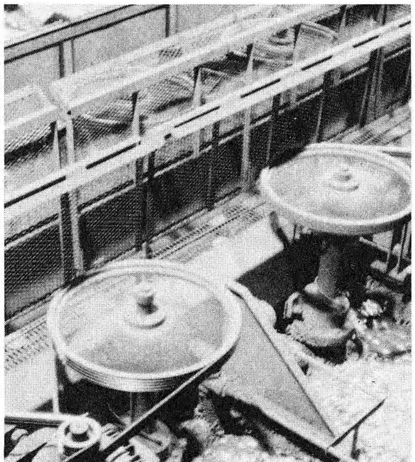

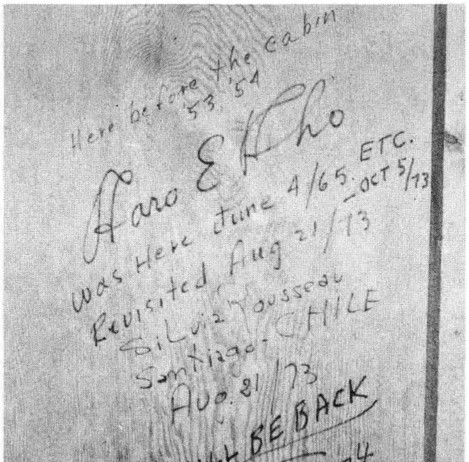

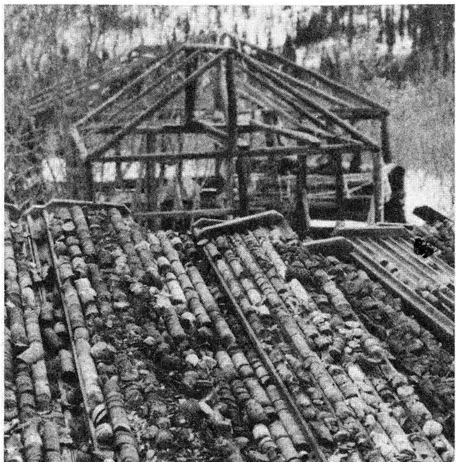

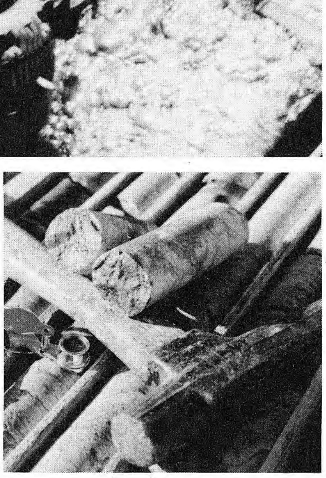

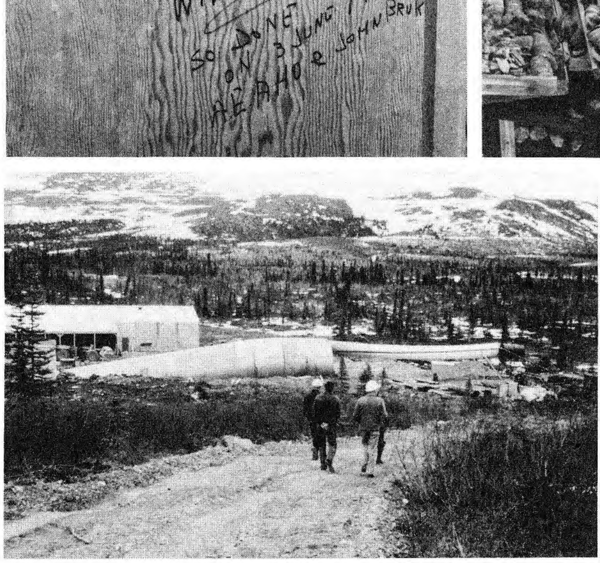

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INTRODUCTION

INTRODUCTION

1.0 INTRODUCTION

In 1978 when consideration was first given to the development of the Vangorda Plateau deposits an enormous amount of data was required in order that a decision to bring the deposits into production could be made. Some of the critical data was available as the result of work by other companies notably Noranda and Kerr Addison, but much of the information simply did not exist.

The generation of the mining and metallurgical data and also of the size and cost of the necessary support facilities was of critical importance to the overall development schedule. The Feasibility and Development Group which was formed in June of 1979, designed a schedule of events and almost immediately implemented programs to achieve the various objectives of the development schedule on a timely basis.

The schedule comprised of three main sections, milling strategy, transportation of ore and mining sequence, and design of support systems. The work contained in this presentation indicates the methods used to derive the milling parameters and the required modifications to the Cyprus Anvil Concentrator. The remaining sections will be published later in 1981-82 when data collection is completed for the appropriate areas of interest.

By year end of 1979, a series of detailed plans had been devised by the Feasibility and Development Group. Laboratory and Pilot Plant tests programs had been implemented to determine metallurgical response by ore type and also average metallurgical response by deposit. Surveys of operations were conducted on a world-wide basis to determine the best operational solutions to various problems and several plant visits were organized to permit first hand studies of new equipment concepts.

Preliminary mill design commenced in late 1979 and involved input from many sources including Faro mill operations personnel, members of the Feasibility and Development Group, Kilborn Engineers and some outside consultants. Mill design was completed by the middle of 1980 and an A.F.E. for the first phase development program submitted at that time. Following approval by the Directors of Cyprus Anvil, predesigned construction schedules were implemented and modifications to the existing plant commenced in June 1980.

At the time of writing, April 1981, the project is on schedule and the modified mill will start up as planned in September 1981 at the design rate. Ancillary projects including the Down Valley Tailings Dam and the additional electrical generating facilities are destined to be completed on schedule in October 1981.

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METALLURGY

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2.4 VANGORDA METALLURGY

2.5 COMPATABILITY TESTING

2.6 SPECIAL CHARACTERISTICS OF THE CONCENTRATES

2.0 SUMMARY

2.0 SUMMARY

Metallurgical test data from laboratory and pilot plant investigations was assembled and carefully analyzed. Following detailed analysis of all data, and based on the design performance of the modified concentrator and experience of northern operations, plant metallurgy for each ore type was predicted.

TABLE 1

PREDICTED METALLURGY BY ORE BODY

ORE SOURCE	CONC.	METALLURGY									
		ASSAYS						DISTRIBUTION			
		Pb	Zn	Au	Ag	Hg	As	Pb	Zn	Au	Ag
Faro-Zone III	Lead	67	*	0.75	600	40	0.03	87.5	*	33	65
	Zinc	*	53.5	*	*	300	0.01	*	88.5	*	*
Grum - Open Pit	Lead	60	*	3.5	750	90	0.10	80	*	33	65
	Zinc	*	55	*	*	650	0.05	*	83	*	*
Vangorda - Open Pit	Lead	50	*	6.5	575	60	0.25	80	*	35	55
	Zinc	*	52.8	*	*	300	0.10	*	77	*	*

Notes: a) Silver, Gold and Mercury assays in terms of g/tonne.

b) Data refers to predicted metallurgy at rated tonnages with design feed grades.

2.1 INTRODUCTION

2.1 INTRODUCTION

The determination of the metallurgical response of each of the three ore types was a long and complex process. Clearly the response of an ore is governed by two principal factors - mineralogical considerations and also treatment conditions.

Fortunately detailed mineralogical studies by the Exploration and Geological groups had revealed that each of the three deposits was made up of varying proportions of three major mineralogical components. These components were designated Type G indicating a baryte rich ore, Type E representing a pyritic specie and Type A which covered a range of mineralized quartzites with variable graphite contents.

Armed with this information, testwork proceeded with efforts being directed toward the determination of the metallurgical response of each of the three major mineralogical types. Then by using the geological model for each deposit, the proportion of each type present was determined, and hence the mean metallurgical response calculated.

As laboratory and pilot plant testwork proceeded it became apparent that certain treatment factors for optimum metallurgy were common to all ore types. A fine primary grind of at least 50 microns

P_{80} appeared quite critical and there were very strong indications that extremely fine regrinding was also necessary to the attainment of acceptable concentrate grades.

In the following section the derivation of metallurgical response by ore deposits using the technique outlined above is detailed. Some of the data requires further verification by means of additional testwork and other data, notably that from the Grum deposit is currently being verified by additional laboratory testwork on new samples of ore.

2.2 FARO ZONE III METALLURGY

2.2 FARO ZONE III METALLURGY

2.21 Preliminary Laboratory Testwork

Interest in the effects of finer primary grinding on Anvil metallurgy was stimulated by testwork performed at Faro during the fall of 1978 on a pyrrhotitic ore type. The indications from this and subsequent work coincided with data from other sources and suggested that optimum metallurgy for most Faro ore types would be obtained by primary grinding to at least the 50 micron P_{80} range. Table 2 shows the sources of data relating to work to determine optimum grind for Faro ore types.

TABLE 2

PRIMARY GRIND LEVEL FOR OPTIMUM METALLURGY

DATA SOURCE	TYPE OF TESTS	NO. OF TESTS	OPTIMUM GRIND P_{80}	%-325#
Cyprus Anvil Test Laboratory	Rougher & Cleaner Tests	100	40-45	85
Kamloops Research Laboratory	Rougher & Cleaner Tests	120	45-50	80
Sachtleben - FDR	Cleaner Testing	30	52	78
Mitsui Mining & Smelting	Rougher & Cleaner Tests	30	37-70	80-93

Notes: a) Numbers of tests are estimates only.

b) Japanese work identified pyrrhotitic ore as benefiting most from fine grinding.

c) Tests are all laboratory scale work on 1.0 or 2.0 kg samples.

Detailed laboratory testwork by ore type was then planned and carried out at the Kamloops Research Laboratory in order to ensure that these preliminary results were representative of the majority of ore types (Ref. Report KM008 - June 10th 1979). Parallel work at the Cyprus Anvil laboratory on a different suite of samples, but encompassing the same test methods yielded the similar results. All data from this latter phase of the program is summarized in the Table 3 below; For completeness, the pilot plant data, described in detail later, is also shown.

TABLE 3

METALLURGICAL IMPROVEMENTS WITH FINER GRINDING

DATA SOURCE	GRIND RANGE INVESTIGATED		METALLURGICAL IMPROVEMENTS ACROSS RANGE			
	(P_{80} microns)	(%-325#)	LEAD		ZINC	
			GRADE	RECOVERY	GRADE	RECOVERY
Rougher Tests KRAL-78	135 - 40	50 - 85	7.0	*	2.50	*
Rougher Tests CAMC 78-79	125 - 37	45 - 90	11.3	*	5.9	*
Cleaner Tests KRAL-79	125 - 45	45 - 80	2.5	7.0	2.1	9.5
Cleaner Tests CAMC-79	125 - 40	45 - 85	3.2	5.2	4.4	4.0
Plant Test Nov.13-17/78	125 - 60	45 - 70	3.0	5.8	3.0	4.0
Cleaner Tests KRAL-79	145 - 30	30 - 90	*	9.6	*	12.0
Lakefield Pilot Plant I-79	130 - 45	40 - 85	5.9	3.6	3.6	5.9
II-79	140 - 55	35 - 80	4.5	1.0	4.5	4.5

Notes: a) * signifies that results were adjusted to a constant grade or recovery for the purposes of analysis of data.

b) Pilot plant data refers to the two separate samples tested.

c) CAMC - Cyprus Anvil Mining Corporation - Faro
KRAL - Kamloops Research and Assay - Kamloops
Lakefield - Lakefield Research Laboratory - Ontario

2.22 Pilot Plant Test Program

In order to verify the results of the laboratory test program, a pilot plant investigation was planned and executed. The work was performed at the testing facility of Lakefield Research of Canada, Limited, Lakefield, Ontario, during September and October of 1979.

During the course of the investigation at Lakefield, two bulk samples were subjected to a detailed investigation. The technical aspects of the work are described in Lakefield Reports LR 2202 Volumes I-IV. Supervision and direction of the testing was the responsibility of W. Muir, Plant Metallurgist, Cyprus Anvil Mining Corporation and P.J. Brown, Consulting Metallurgist, Met Engineers Ltd. At all times during the program, the senior Lakefield metallurgist, Serge Bulatovic provided the indespensible link between client and pilot plant operating personnel.

The work was performed using the standard reagent pattern currently employed at Anvil and with a flotation circuit designed to approximate to that proposed in the modified circuit at Anvil. Grinding and regrinding effects were the principle parameters investigated in the program. The program had two major objectives which were:

(a) The determination of the optimum economic grind level.
(b) Estimation of plant metallurgy at this optimum point.

2.23 Determination of the Optimum Grind

Using the results of the pilot plant program at various grind levels and by estimating the capital expenditure needed to achieve the various grind levels, it became possible to estimate optimum economic grind level. This calculation was performed by using the metallurgical data generated by the pilot plant work, a capital expenditure estimate generated by Kilborn Engineering Ltd. for various grinding configurations and an operating unit cost increment. The results of this study are summarized in table 4 below, and confirm that the economic and the metallurgical optima are near coincident at a P_{80} of about 50 microns.

TABLE 4

OPTIMUM ECONOMIC GRIND LEVEL

CASE	PRIMARY H.P.	REGRIND H.P.	TOTAL H.P.	GRIND TARGET P_{80}	CAPITAL COST $\$x10^6$	OPERATING COST INCREASE $\$x10^6$	RELATIVE NET PRESENT VALUE $\$x10^6$
1	11,700	1,350	13,050	50	43.0	6.8	153.9
2	9,200	2,350	11,550	70	43.1	5.1	129.8
3	7,700	1,850	10,550	100	38.6	4.3	112.7
4	5,200	1,350	6,550	130	0	0	117.4

Notes:
a) Reference - Memo Brown to Taggart Dec. 16, 1979 "Optimum Grind Calculations".
b) Capital cost includes power plant: Clearly case 4 is with no expansion.
c) Operating cost includes manpower, steel and reagents.
d) Data assumes milling a 50:50 blend of Grum and Anvil Zone III ore.

In reviewing the data in table 4, it is important to note that for the four alternatives considered that the difference in capital and installed power is very small indeed. The difference in predicted plant metallurgy for each case is significant.

The calculation of Relative Net Present Value was achieved by making certain assumptions about costs and prices: The values computed are appropriate for comparative purposes but may not be used as absolute values.

2.24 Predicted Metallurgy at Optimum Grind

Pilot plant data at the economic optimum grind level of 50 microns was averaged for each of the two bulk samples tested and the results used as a basis to predict the plant metallurgy at the optimum grind.

Both ore samples treated were taken from the Faro Zone I pit and were selected by the Engineering department of Cyprus Anvil as being representative of the ore type known as 4E. Type 4E is a predominately pyritic specie which comprises the bulk of the ore in the Zone III ore body. Since both samples suffered moderate to severe en route oxidation, some of the minerals were rendered non flotable - this was especially evident in sample No. 1. Shown below in Table 5 are the sulphide and non-sulphide contents of the two samples illustrating the degree of sample oxidation.

TABLE 5

PILOT PLANT BULK SAMPLES - HEAD ASSAYS

SAMPLE	ASSAYS %			
	Pb	Pb Ox.	Zn	Zn Ox.
Bulk Sample No.1	2.35	0.40	3.85	0.25
Bulk Sample No.2	2.40	0.35	4.34	0.22

Note: "Oxide components" encompass all non-sulphides
including carbonates, hydroxides, etc.

Because of the oxidized condition of the samples, corrections were made to recorded data to compensate for the anamolous oxide contents. The corrections were based on experience, and established practice which required that a certain fraction of the reported oxide be removed from the head assay in the recovery calculations. These corrections had the effect of increasing the reported pilot plant lead recovery substantially; zinc metallurgy was only marginally influenced.

TABLE 6

CALCULATED SULPHIDE HEAD ASSAYS

SAMPLE	ASSAYS % LEAD ASSAY HEAD	ASSAY OXIDE	CALCULATED EFFECTIVE HEAD	ZINC ASSAY HEAD	ASSAY OXIDE	CALCULATED EFFECTIVE HEAD
Bulk Sample No. 1	2.35	0.40	2.05	3.85	0.25	3.72
Bulk Sample No.2	2.40	0.35	2.14	4.34	0.22	4.23

In keeping with conservative methods of predicting data used elsewhere in this study, it was assumed that, of the oxidized lead and zinc, only one quarter would be recoverable; (Table 6) Thus the calculated effective lead assays were derived. The results of the pilot plant data modified in this manner are shown below in table 7 and 8.

TABLE 7

LAKEFIELD PILOT PLANT TEST RESULTS

PILOT PLANT SAMPLE NO.1

50 MICRON GRIND

TESTS 9, 10 and 24

	WEIGHT %	ASSAYS Pb	Zn	DISTRIBUTION Pb	Zn
Feed	100.0	2.03	3.66	100.0	100.0
Lead Conc.	2.58	68.4	2.6	86.9	1.8
Zinc Conc.	6.35	0.38	52.1	1.2	90.4
Tails	91.71	0.27	0.31	11.9	7.8

Notes: a) See notes under table no. 8.

TABLE 8

PILOT PLANT SAMPLE NO.2

50 MICRON GRIND

TESTS 32, 33, 34 and 35

	WEIGHT %	ASSAYS Pb	Zn	DISTRIBUTION Pb	Zn
Feed	100.0	2.12	4.26	100.0	100.0
Lead Conc.	2.62	73.6	2.6	91.0	1.6
Zinc Conc.	6.99	0.48	55.2	1.6	90.6
Tails	90.39	0.17	0.37	7.4	7.8

Notes: a) Head assays were determined from the pilot plant grinding circuit overflow with a correction made for non-recoverable oxide species.

b) Weight percentages taken from assay calculations averaged for the pilot plant runs noted. Checked also by three product formula method.

c) Concentrate and minor element determined by assaying pilot plant composites.

In studying the overall results of the two test series two factors were considered significant:

i) The sample feed grades were quite low compared to predicted mill feed for Zone III: It would be quite reasonable to expect better results with the higher feed grade.

TABLE 9

COMPARISON OF HEAD ASSAYS

SAMPLE	HEAD ASSAYS	
	Pb	Zn
Bulk Sample No. 1	2.35	3.85
Bulk Sample No. 2	2.40	3.85
Zone III Average Estimated	2.90	4.50

Note: a) Zone III data is from latest mine model and does not include phase 6 ore.

ii) The results from sample No.2 were considered to be quite exceptional. There were several indications to support this view, especially the high zinc concentrate grades which suggested a somewhat atypical zinc mineral structure, the rapid rate of flotation of the minerals, and the results from the Faro plant when treating similar material. Also the sample exhibited most unusual precious metal distributions.

The pilot plant data was analyzed and a conservative metallurgical balance constructed for actual plant operation. For comparison are shown some locked cycle test results generated on samples of Zone III drill core. (Table 10)

TABLE 10

PLANT METALLURGICAL BALANCE

DATA SOURCE	CONCENTRATE	ASSAYS Pb	ASSAYS Zn	DISTRIBUTION Pb	DISTRIBUTION Zn
Pilot Plant Work Sample No.1	Lead	68.4		86.9	
	Zinc		52.1		90.4
Pilot Plant Work Sample No.2	Lead	73.6		91.0	
	Zinc		55.2		90.6
Locked Cycle Tests	Lead	77.2		88.8	
	Zinc		53.8		83.5
Predicted Plant Results Faro Zone III	Lead	67.0		87.5	
	Zinc		53.5		88.5

Note: a) All data refers to metallurgy at a P_{80} of 50 microns.

b) Locked cycle tests performed at Cyprus Anvil during 1980 on samples of Zone III drill core. Data reported in January 1981 in a report by A. McIntyre.

2.25 Gold, Silver and Minor Elements

Based on assays of diamond drill composites, the Zone III ore was estimated to contain appreciable quantities of silver and minor amounts of gold. Averaged data indicated that the mean silver content would be 35 g/tonne Ag while gold would be about 0.065 g/tonne Au. The estimate of the predicted gold and silver metallurgy was conducted as follows:

i) Silver Metallurgy

Two pilot plant runs using sample No. 1 were assayed for silver; the results are shown below.

TABLE 11

SILVER RECOVERY AND GRADE AT 50 microns GRIND

TEST NO.	PRODUCT	WEIGHT %	ASSAYS %, g/tonne			% DISTRIBUTION		
			Pb	Zn	Ag	Pb	Zn	Ag
PP-9	Pb Cl. Conc.	2.58	69.1	2.4	501.13	75.8	1.6	60.3
	Zn Cl. Conc.	6.56	0.43	52.5	38.32	1.2	88.8	11.7
	Zn Comb. Tail.	90.86	0.59	0.41	6.62	23.0	9.6	28.0
	Cyclone O'Flow	100.00	2.35	3.88	21.46	100.00	100.00	100.00
PP-10	Pb Cl. Conc.	2.61	68.2	2.82	486.38	76.7	1.9	58.2
	Zn Cl. Conc.	6.24	0.34	52.0	34.86	0.9	85.6	10.0
	Zn Comb. Tail.	91.15	0.57	0.52	7.62	22.4	12.5	31.8
	Kason U'Size	100.00	2.32	3.79	21.82	100.00	100.00	100.00

Assuming that the silver recovery increases linearly with lead recovery as with most Faro area ore types, then increasing the lead recovery to 87.5% should result in some increase in silver recovery from the reported 58-60% level. Based on experience and the long term silver recovery data at Anvil, a silver recovery figure of 65% was selected.

Unfortunately the bulk samples were quite low in silver content - about 21.5 g/tonne Ag: correction for this was achieved by calculating the quantity of silver recovered into the lead concentrate at fixed recovery (e.g. 600 g/tonne silver in the lead concentrate) from a predicted mill feed of 34 g/tonne Ag.

ii) Gold Metallurgy

Some support for both the gold and silver predicted recovery levels were obtained by reassaying a locked cycle test on samples from Faro Zone III obtained in the 1978 drilling program. The data in Table 12 shows that a gold recovery of 35-40% may be anticipated in the lead concentrate. Using the predicted mine model feed grade and a gold recovery of 33% the gold concentration in the lead concentrate was calculated (Table 14).

TABLE 12

GOLD AND SILVER METALLURGY - LABORATORY TEST DATA

	ASSAYS (g/tonne)		DISTRIBUTION	
	Au	Ag	Au	Ag
Feed	0.17	44	100.0	100.0
Lead Concentrate	1.20	530	38.8	66.3
Zinc Concentrate	0.70	39	37.0	8.0

Notes:
- a) Composite obtained from a very limited drill program.
- b) Head assays higher than mine model predicts.
- c) Data from cycle test No.8 Ref.LR2082 using Faro Zone III ore.

iii) Mercury and Arsenic

Lacking specific assay data on Zone III lead and zinc concentrate contaminants, minor element assays for the year 1979 were averaged and used to predict mercury and arsenic concentrations.The averaged data for 1979 concentrations of minor elements is shown below in Table 13.

TABLE 13

MINOR ELEMENTS - FARO ZONE III

CONCENTRATE	ASSAYS g/tonne	
	Hg	As
Lead Concentrate	40	300
Zinc Concentrate	300	100

2.26 Predicted Plant Metallurgy

TABLE 14

PREDICTED PLANT PERFORMANCE - ZONE III ORE

	WEIGHT	ASSAY				DISTRIBUTION			
		Pb	Zn	Ag	Au	Pb	Zn	Ag	Au
Feed	100.0	2.9	4.6	35	*	100.0	100.0	100.0	100.0
Lead Conc.	3.79	67.0	3.0	600	0.75	87.5	2.5	65.0	33.0
Zinc Conc.	7.61	0.5	53.5	40	*	1.3	88.5	8.7	*
Tails	88.60	0.37	0.47	10	*	11.2	9.0	26.3	*

Note: a) The average gold content of the lead concentrates from Zone III will be below payable limits. However, due to the irregular occurrance of gold in the ore it is possible that payable gold will be encountered in some shipments.

2.3 GRUM METALLURGY

2.3 GRUM METALLURGY

2.31 Preliminary Laboratory Testwork

Metallurgical investigations of the Grum ore body by Kerr Addison Ltd. and Noranda Mines were preceded by extensive mineralogical studies in which the entire array of ore types, believed to number twelve, were individually characterized. Having identified the various ore types, metallurgical testwork commenced at Lakefield on individual samples of each ore type, and also on composites of ore types. This compositing of ore types was a relatively late step when it was realized that many of the so called ore types were in fact metallurgically indistinguishable. The concept of compositing samples for testing was very acceptable since the costs of individually testing each ore type were prodigious.

The testwork, which was performed exclusively at Lakefield Research Ltd., Lakefield, Ontario, commenced in 1976 and culminated in late 1977 with long series of pilot plant tests. The work was directed principally by Mr. K. Konigsmann, Chief Metallurgist, Noranda Mines, and by others in the Noranda Milling Committee. A very brief synopsis of the excellent work performed at Lakefield is given below.

2.32 Critical Results from Laboratory Testwork

The earliest work performed at Lakefield demonstrated the need for a fine primary grind on Grum ores in the range 50 microns P_{80}. This observation was strongly supported by the results of the meticulous microscopic studies of Dr. Carson of Noranda Mines, which indicated an unusually fine intergrowth of mineral crystals in most ore types.

Many laboratory test series were performed to illustrate the effect of variations of primary grind on metallurgy. Typical of the tests are the data shown below in tables 15 and 16; they refer to testwork on a composite made up of seven of the twelve ore types.

TABLE 15

EFFECT OF FINENESS OF PRIMARY GRIND

TEST NO.	TIME Min.	%-200 Mesh	PRODUCTS	WEIGHT %	ASSAY% Pb	Zn	% DISTRIBUTION Pb	Zn
153	30	87.1	Pb Cleaner Concentrate Pb Rougher Concentrate	3.64 17.93	65.9 15.3	3.69 7.54	78.2 89.8	2.2 22.2
179	25	79.2	Pb Cleaner Concentrate Pb Rougher Concentrate	3.30 19.21	63.3 14.0	4.04 7.65	70.0 89.9	2.2 24.1
180	20	68.1	Pb Cleaner Concentrate Pb Rougher Concentrate	3.01 18.81	65.9 14.0	3.89 7.66	67.1 88.8	2.0 24.2

Notes: a) Data Source L.R. 1991 Vol. 7.

b) Comparative data for the effect of grind on the zinc circuit not reported for the tests.

Having established that adequate mineral liberation could be obtained at a relatively fine primary grind, attention was then focused on improving the lead concentrate grade. In almost all samples the investigators found that an extremely fine regrind of the lead rougher concentrate was mandatory in order to achieve reasonable concentrate grades at acceptable recoveries. Typical of the results obtained are those shown below in Table 16. Data in this table refers to work on the pilot plant composite which encompassed components from most ore types. The effect of regrinding on lead final concentrate grade, at almost constant recovery, is obvious.

TABLE 16

EFFECT OF LEAD CONCENTRATE REGRIND ON LEAD CLEANING

TEST NO.	PRIMARY GRIND (MIN.)	Pb REGRIND		PRODUCT	WEIGHT %	ASSAY %		% DISTRIBUTION	
		TIME (MIN)	% PASS 10 u			Pb	Zn	Pb	Zn
42	30	10	31.5	Pb Cleaner Concentrate	12.12	38.6	12.9	81.0	15.6
				Pb 1st Cleaner Conc.	18.63	27.3	13.7	88.0	25.5
				Pb Combined Tailing	81.37	0.85	9.19	12.0	74.5
43	30	20	37.2	Pb Cleaner Concentrate	10.54	45.1	12.9	80.4	13.7
				Pb 1st Cleaner Conc.	18.24	28.8	14.1	89.0	26.0
				Pb Combined Tailing	81.76	0.80	8.97	11.0	74.0
45	30	30	47.4	Pb Cleaner Concentrate	8.75	51.6	11.3	76.1	9.9
				Pb 1st Cleaner Conc.	16.08	32.4	14.0	87.7	22.6
				Pb Combined Tailing	83.92	0.87	9.14	12.3	77.4
46	30	40	58.0	Pb Cleaner Concentrate	7.97	54.6	10.7	73.4	8.1
				Pb 1st Cleaner Conc.	16.54	31.3	14.1	87.3	23.0
				Pb Combined Tailing	83.46	0.89	9.2	12.7	76.0
47	30	50	62.0	Pb Cleaner Concentrate	7.96	59.7	9.53	80.0	7.6
				Pb 1st Cleaner Conc.	14.56	35.8	13.6	87.7	19.9
				Pb Combined Tailing	85.44	0.86	9.32	12.3	80.1

Notes: a) Data source L.R. 1991 Vol. 10

b) Sample: Pilot plant composite.

The route to acceptable metallurgy with Grum ore types then was established; a fine primary grind (P_{80} 50-60 microns) followed by a very fine regrind of the lead rougher concentrate. (P_{80} 15-20 microns)

In the course of these detailed studies it was found that the lead regrind mill exerted a significant influence on the zinc metallurgy; probably due to the considerable quantity of zinc minerals reporting in the lead rougher concentrate. In general however, the zinc metallurgy did not pose a major problem. The Grum zinc mineral was observed to be quite low in interstitial iron and manganese and, apart from losses in the lead cleaner circuit products, zinc recovery was good. The laboratory results suggested that zinc concentrate grades in excess of 55% Zn could be fairly easily achieved provided that both the lead and zinc regrinding circuit were optimized.

TABLE 17

EFFECT OF ZINC REGRIND ON ZINC CLEANING

TEST No.	Zn CLEANER FEED %-20μ	%-10μ	PRODUCT	ASSAY Pb	Zn	DISTRIBUTION Pb	Zn
41,47	78	50	Zinc Cleaner Conc.	2.3	56.5	5.7	80.6
			Zinc Flotation Tail	1.2	1.6	16.8	12.6
42,43,44	39	22	Zinc Cleaner Conc.	2.5	51.0	6.3	73.5
			Zinc Flotation Tail	1.3	2.3	17.3	17.6

Note: a) Data source LR2027.

b) Data refers to laboratory batch cleaner tests on a composite of the major ore types.

2.33 Pilot Plant Test Program

As the laboratory testwork neared completion at Lakefield in mid 1977, underground operations at the Grum site were directed toward obtaining a bulk sample for pilot plant testwork. The development adit had by this time penetrated the main sulphide zone and the bulk sample was taken principally from this area. Unfortunately sample make-up was somewhat biased because of the limited availability of representative quantities of each specific ore type. The sample bias was reflected in the anomalously high metal contents of the pilot plant composite.

mlwl

TABLE 18

HEAD ASSAY - PILOT PLANT COMPOSITE

	ASSAYS %					ASSAYS g/tonne		
	Cu	Pb	Zn	Fe	As	Au	Ag	Hg
Sample	0.13	6.1	10.0	20.5	0.23	1.4	98	80

Despite the problems with the sample bias there was no alternative but to commence testwork and to attempt to compensate, or correct the test results for the high grade sample when predicting probable plant metallurgy.

To assist in the prediction of plant metallurgy and as a guide in the extrapolation of results several locked cycle tests were carried out on various composites of Grum ores by Lakefield Research.

TABLE 19

LOCKED CYCLE TEST DATA - GRUM COMPOSITES

TEST NO.	COMPOSITE NO.	PRODUCT	WEIGHT %	ASSAYS %		% DISTRIBUTION	
				Pb	Zn	Pb	Zn
207	1	Pb Cleaner Conc. Zn Cleaner Conc. Zn Flot.Tailing	4.77 9.18 86.05	66.7 0.58 0.95	4.56 57.40 1.43	78.4 1.3 20.3	3.2 78.5 18.3
		Head (Calc.)	100.0	4.06	6.71	100.0	100.0
217	1	Pb Cleaner Conc. Zn Cleaner Conc. Zn Flot. Tailing	6.12 8.91 84.07	62.2 0.40 0.40	5.67 59.70 1.49	91.0 0.9 8.1	5.0 76.7 18.3
		Head (Calc.)	100.0	4.18	6.93	100.0	100.0
230	3	Pb Cleaner Conc. Zn Cleaner Conc. Zn Flot. Tailing	5.82 13.75 80.43	62.5 0.39 1.02	6.16 57.80 1.25	80.6 1.2 18.2	3.9 85.4 10.7
		Head (Calc.)	100.0	4.51	9.31	100.0	100.0
237	2	Pb Cleaner Conc. Zn Cleaner Conc. Zn Flot. Tailing	9.12 14.31 76.57	59.7 0.56 0.41	7.47 55.7 0.78	93.3 1.4 5.3	7.4 86.2 6.4
		Head (Calc.)	100.0	5.84	9.25	100.0	100.0
241	2	Pb Cleaner Conc. Zn Cleaner Conc. Zn Flot. Tailing	9.06 14.49 75.87	59.1 0.46 0.44	7.24 56.6 0.57	91.4 1.2 5.7	6.9 86.0 4.5
		Head (Calc.)	100.0	5.86	9.53	100.0	100.0

Notes: a) The composites 1, 2, and 3 are described in compositional detail in L.R. 1991 Vol. 8.

b) Tests carried out at 50-60 micron grind.

After about thirty preliminary tests, the regrinding circuits were optimized, the reagent scheme balanced and the optimum primary grind established at between 50-55 microns P_{80}. In the course of the work, and subsequent to the preliminary tests, several interesting observations were made.

a) Initially it appeared that lead regrind was not quite as critical as the laboratory work had indicated. Presumably this was because much of the laboratory work was carried out with samples averaging 9-10% combined metal; the higher grade pilot plant samples would naturally require less regrinding.

b) The collector consumption in the pilot plant was considerably lower than had been observed in the laboratory. Again this could be related to the relatively coarse mineralization associated with the high grade sample.

c) The zinc circuit first cleaner pH was found to be critical. Even very small variations (± 0.5 units at pH 11.5) exercised enormous effects on the zinc concentrate grade.

d) High cyanide additions, in total about 250 g/tonne were found to be critical to attainment of good metallurgy.

The pilot plant program was directed by K. Konigsmann of Noranda Mines who was aided by D.M. Wyslouzil of Lakefield Research. A total of 53 tests were performed using a flotation reagent scheme very similar to that employed at Anvil (and incidentally almost identical to that used in the Anvil Pilot Plant program). The results are shown below in Table 20.

TABLE 20

GRUM PILOT PLANT TEST DATA

TEST NO.	METALLURGY			
	LEAD GRADE	LEAD RECOVERY	ZINC GRADE	ZINC RECOVERY
1	49.6	75.9	58.9	48.9
2	53.1	74.3	56.0	73.5
3	44.7	80.7	56.0	70.9
4	44.1	80.6	50.2	66.2
5	47.6	75.6	55.8	66.1
6	59.0	81.8	54.3	78.5
7	46.6	83.1	52.6	72.3
8	49.9	79.7	49.6	78.0
9	62.4	77.8	55.1	73.8
10	44.9	79.2	36.9	74.0
11	42.2	81.6	42.6	72.7
12	47.9	74.5	38.2	76.4
13	46.4	77.8	46.0	77.6
14	52.8	77.3	47.0	78.6
15	57.2	74.8	42.3	78.7
16	56.2	76.2	49.3	75.8
17	59.2	76.2	43.3	81.4
18	57.5	74.8	50.0	79.9
19	57.9	74.2	51.5	76.7
20	59.7	70.8	48.3	74.9
21	53.6	78.1	53.9	76.1
22	62.3	71.7	54.2	78.1
23	52.3	71.0	56.3	69.1
24	58.4	63.0	52.0	81.1
25	59.3	77.4	52.1	81.1
26	57.0	76.0	56.4	76.0
27	58.4	77.7	48.3	79.8
28	51.9	78.0	46.8	75.6
29	56.4	77.6	54.5	74.2
30	54.9	75.4	50.0	78.0
31	53.7	76.3	55.0	75.4
32	65.4	78.1	54.5	79.8
33	59.2	80.4	51.8	79.0
34	64.8	77.6	53.5	77.4
35	66.6	75.7	57.4	72.0
36	54.1	77.5	52.6	78.0
37	63.1	78.3	50.2	82.3
38	60.6	77.0	53.3	79.9
39	56.9	75.3	50.0	76.6
40	57.2	75.2	56.0	72.9
41	62.7	77.2	57.4	79.6
42	60.9	75.6	48.4	79.9
43	60.7	75.9	52.1	75.5
44	54.1	77.4	52.4	65.1
45	46.8	76.5	53.8	71.7
46	55.2	76.8	50.8	73.6
47	64.6	77.9	55.5	81.5
49	62.7	76.0	56.5	78.3
50	60.8	78.1	53.6	80.6
51	65.8	74.4	52.6	82.6
52	58.3	77.5	54.0	78.0
53	60.9	75.5	54.7	78.9

Notes: a) Data source progress report No.11 Vol. I-IV L.R. 2027.

b) All results from tests using a standard flowsheet and one composite ore sample.

2.34 Analysis of Pilot Plant Results

The data from the tests in the latter part of the program was examined in minute detail by the Noranda Milling Committee. Noting that the sample tested comprised principally of massive sulphide ore types, the committee elected to produce two metallurgical balances: One representing the massive sulphide ore metallurgy while the other indicated their estimate of the overall average Grum metallurgy. This latter estimate was based on the results of laboratory cleaner and the locked cycle tests on various samples.

The committee were of the opinion that the true average metallurgy of all Grum ores would be somewhat better than for the sulphide zone ores. Accordingly they modified their best pilot plant using the results of the locked cycle tests to reflect this belief and increased both lead and zinc recoveries with the same concentrate grades. The two balances are shown below in Table 21.

TABLE 21

PREDICTED METALLURGY FROM PILOT PLANT RESULTS

ORE TYPE	CONCENTRATE	ASSAYS				DISTRIBUTION			
		Pb	Zn	Au	Ag	Pb	Zn	Au	Ag
Massive Sulphide	Lead	62	10	4.8	925	77	*	33	72
	Zinc	2.5	56	*	*	*	81	*	*
Average	Lead	62	8	5.1	950	80	*	33	72
	Zinc	2.0	56	*	*	*	84	*	*

Notes: a) Data source - Noranda Milling Committee Report Dec. 1977.

b) Au and Ag in g/tonne.

2.35 Gold, Silver and Minor Elements

Some assays on three pilot plant runs yielded some interesting results on precious metal concentrations and distributions. Concentrates collected during various pilot plant runs were assayed for mercury and arsenic. The results are discussed below in section 2.36.

TABLE 22

GOLD & SILVER METALLURGY - GRUM PILOT PLANT

TEST NO.	PRODUCT	WEIGHT %	ASSAYS (g/tonne) Au	Ag	% DISTRIBUTION Au	Ag
PP25	Pb Cleaner Concentrate	8.02	3.40	857	48.8	72.5
	Zn Cleaner Concentrate	16.04	0.15	73	4.9	12.4
	Zn Combined Tailing	75.94	0.30	19	46.3	15.1
	FLOTATION FEED (calc)	100.00	0.68	95	100.0	100.0
PP35	Pb Cleaner Concentrate	6.77	4.46	950	50.6	74.8
	Zn Cleaner Concentrate	12.30	0.15	69	3.4	9.8
	Zn Combined Tailing	80.93	0.30	16	46.0	15.4
	FLOTATION FEED (calc.)	100.00	0.68	86	100.0	100.0
PP37	Pb Cleaner Concentrate	7.32	3.77	950	46.6	77.2
	Zn Cleaner Concentrate	16.72	0.30	74	9.8	13.8
	Zn Combined Tailing	75.96	0.30	11	43.6	9.0
	FLOTATION FEED (calc.)	100.00	0.68	90	100.0	100.0

Notes: a) Data source - Progress Report No.11 Vol.I L.R.2027. Gold & Silver assays reported only on these tests.

b) Each data set based on an 8 hour duration pilot plant run.

2.36 Predicted Plant Metallurgy

Because of the significant effect which mill feed grade exercises on metallurgy and since the pilot plant sample was known to be biased, there are reasons to believe that the Grum average metallurgy, predicted by the Noranda Milling Committee, may be optimistic. The principal reasons are as follows:

a) Lead grade was difficult to achieve in the pilot plant, and lead recovery was quite low in those tests in which concentrate grades in excess of 60% Pb were obtained.

b) Silver recovery at 72% probably reflects the anomalously high silver content in the pilot plant feed. Silver recovery into the lead concentrate at Cyprus Anvil averages somewhat less than 60% and is expected to reach 65% only at very high lead recoveries.

Based on actual experience of the operation of a metallurgical plant in the Faro area with the constraints of climate, limited manpower availability and variable ore types, a conservative approach would indicate that a somewhat more pessemistic balance be adopted. Accordingly the lead concentrate grade was reduced from 62% to 60% Pb, but the lead recovery was kept constant. Silver recovery was reduced to 65% to reflect lower silver grade expected in the diluted ore from the Grum deposit.

The gold in the mill feed was assumed to be proportional to the lead content; hence the diluted feed grade required that the pilot plant gold head assay be reduced by about 60%. A gold recovery figure of 33% was used in the calculations - identical to that assumed for Zone III ore.

TABLE 23

PREDICTED METALLURGY - GRUM ORE

CONCENTRATE	ASSAYS						DISTRIBUTION			
	Pb%	Zn%	Au	Ag	As	Hg	Pb%	Zn%	Au%	Ag%
Average Mill Feed-Lead Conc.	60	11	3.5	750	100	90	80	*	33	65
-Zinc Conc.	2.5	55	*	*	50	650		83		

Notes: a) Au, Ag, As, and Hg assays in terms of g/tonne.

b) Silver recovery recalculated on the basis of approximately 50 g/tonne Ag in the mill feed.

c) Average mill feed including dilution, will be about 9% combined lead and zinc metal.

2.4 VANGORDA METALLURGY

2.4 VANGORDA METALLURGY

2.41 Preliminary Laboratory Work

Since its discovery in the mid fifties the Vangorda ore body remained, until quite recently a metallurgical enigma. Early metallurgical work produced erratic results and even the most optimistic metallurgist could predict only the production of a bulk lead-zinc concentrate. Probably the reason for the very poor initial metallurgical results was that the early diamond drill core was of small diameter and hence core recovery was poor, incidentally negating the chance of an accurate ore body model, and ensuring the rapid oxidation of the minerals, thus preventing the generation of reproducible laboratory test data.

Redrilling in the late sixties, and recognition of at least three ore types by the various investigators led to the generation of a few groups of reasonably encouraging metallurgical results. After about five years of sporadic testwork in many laboratories, there emerged a picture of an ore which depicted an extremely fine mineral crystal intergrowth, very markedly activated iron and zinc minerals, and a prediliction for rapid and deleterious mineral oxidation.

The Table 24 below is a synopsis of the early published work on Vangorda ore - unfortunately very few of the investigators noted the ore type being tested. The testwork covers a wide range of metallurgical conditions varying from lime to soda ash modulated circuits, various collectors and depressants and several cleaning schemes. All work exhibited one common factor however - Vangorda ores required an extremely fine primary grind of the order 30-50 μ P_{80} to permit any sort of separation to be achieved.

TABLE 24

PRELIMINARY LABORATORY TEST DATA - VANGORDA ORE

REF.	FEED		LEAD CONC.		ZINC CONC.		TAILINGS		GRIND	SOURCE
	% Pb	% Zn	GRADE	REC.	GRADE	REC.	% Pb	% Zn	MESH #	
1	4.0	4.9	49.9	77.0	49.4	60.4	0.15	0.66	99%-200	Dowa Mining Company Report. March 25, 1969.
2	4.6	4.7	56.6	83.8	54.4	79.7	0.63	0.54	64%-325	Galligher Company Report. July 17,1969. Series 2 tests.
3	4.1	5.2	51.7	89.7	55.4	78.1	0.38	0.65	82%-325	Brunswick Mining & Smelting Report. 1969
4	3.2	5.2	46.1	77.6	53.0	49.5	-	-	-	Brunswick Mining & Smelting attachment "Refractory Ores"
5	4.0	4.9	49.8	77.0	49.4	60.4	-	-	-	Brunswick Mining & Smelting attachment "Refractory Ores"
6	3.9	5.1	56.5	72.5	50.8	51.5	-	-	-	Brunswick Mining & Smelting attachment "Refractory Ores"
7	4.4	5.0	22.7	85.2	24.2	60.0	0.52	0.44	82%-325	Noranda Report "Vangorda & Brunswick M & S Ore Samples" December 9-13, 1969
8	5.8	6.1	29.1	85.4	11.0	60.1	0.54	1.22	80%-325	Noranda Report "Preliminary Test work on Vangorda Cores" April 2, 1970.
9	3.4	3.7	22.3	75.6	12.5	64.2	0.58	0.70	80%-325	Noranda Report "Preliminary Test work on Vangorda Cores" April 2, 1970.
10	1.6	2.7	37.4	77.9	49.2	77.4	0.34	0.36	55%-325	Noranda Report "Preliminary Locked Tests" Feb. 26, 1975
11	3.6	4.1	53.6	77.9	51.8	77.3	0.64	0.63	63%-325	Noranda Report "Preliminary Locked Tests" Feb. 26, 1975
12	3.7	6.8	48.7	78.4	54.4	80.4	0.51	0.74	77%-325	Noranda Report "Preliminary Locked Tests" Feb. 26, 1975
13	6.2	10.5	25.0	93.2	48.6	64.6	0.34	0.72	72%-325	Noranda Report "Preliminary Locked Tests" Feb. 26, 1975
14	3.1	4.4	61.7	78.3	55.3	86.6	0.19	0.14	80%-325	Dowa Mining Company "Metallurgical Test of the Vangorda Ore" May 1975.

Notes:

a) All data shown in this table was abstracted from laboratory reports. Various schemes were employed but the soda ash - cyanide reagent scheme was most favoured.

b) The importance of a fine primary grind and the need for regrinding was repeatedly stressed by the investigators.

2.42 Detailed Type Testing

In the Summer of 1979 the Vangorda deposit was redrilled by Cyprus Anvil with relatively large diameter NQ drill bits and the core subjected to a meticulous geological examination and logging. The core was divided into three main geological species, type 4G, 4E and 4A, and then subdivided again according to assay range of lead and zinc.

Each group of samples was then subjected to detailed metallurgical testwork at the Kamloops Research Laboratory: The tests being designed to determine the effect of primary grind and regrinding on metallurgy when the particular ore type was treated with the Anvil reagent scheme. During the period June to November a total of 55 open circuit cleaner tests were completed on the three major ore types under study.

The results, are shown in summary form below in Table 25 as generated and also in a corrected form based on established laboratory procedures for the evaluation of metallurgical test data. The corrections involve the redistribution of the cleaner tailings to one or other of the concentrates according to experience.

Of special note are the extreme finess of grind employed in most tests ranging from 20-50 microns P_{80} in the primary grind. That these very fine grinds can be achieved is due to the extreme friability of the Vangorda ores.

Derivation of the metallurgy by ore type was fairly simple and was achieved by averaging the best corrected results obtained in the laboratory testwork. Almost invariably the best results occurred at very fine grind levels coupled with fine regrinding. Since the tests covered a wide range of test conditions for each major ore type, considerable attention was addressed to selection of the results for inclusion in each data set.

At no time during this phase of the testwork, nor in subsequent work performed in 1980, was a point detected beyond which fine grinding exercised a deleterious effect on metallurgy. These latter studies were conducted in the range 15-40 microns primary grind, with samples of various ore types (Ref. KM032 October 1980). As a point of general interest more and more complex sulphide operations are employing extremely fine primary grind levels. Meggen and Brunswick grind to 30-40 while at Huelva and Aznalcolla in Spain the primary grind is performed at less than 30 μ .

TABLE 25

VANGORDA METALLURGY BY ORE TYPE

TEST NO.	TEST METALLURGY				CORRECTED METALLURGY				GRIND TIME (min.)	GRIND % 325 MESH	ORE TYPE
	LEAD		ZINC		LEAD		ZINC				
	GRADE	RECOVERY	GRADE	RECOVERY	GRADE	RECOVERY	GRADE	RECOVERY			
1	45.5	43.9	54.1	69.6	45.5	58.0	54.1	74.7	15	94	1A4G
2	36.4	88.4	55.7	66.0	36.4	90.3	55.7	68.6	15	97	1A4G
3	55.7	85.6	56.7	76.4	55.7	87.1	56.7	79.0	15	96	1A4G
4	54.6	87.2	54.9	76.2	54.6	89.2	54.9	78.7	15	98	1A4G
5	57.8	83.2	56.6	75.2	57.8	85.7	56.6	77.7	15	97	1A4G
6	40.3	84.2	55.1	63.3	40.3	86.7	55.1	67.4	10	86	1A4G
7	48.2	85.0	57.5	62.3	48.2	87.5	57.5	67.9	10	88	1A4G
8	60.8	77.5	54.2	61.8	60.8	81.3	54.2	69.3	10	89	1A4G
13	57.3	83.1	55.9	71.7	57.3	83.4	55.9	75.2	15	94	1A4G
14	61.4	79.7	55.9	77.9	61.4	81.4	55.9	80.6	15	94	1A4G
9	36.4	72.5	47.7	63.3	36.4	76.5	47.7	69.3	10	91	1B4G
10	45.9	76.2	51.4	72.3	45.9	79.2	51.4	75.3	15	94	1B4G
11	49.6	67.9	52.4	75.9	49.6	73.6	52.4	81.6	15	97	1B4G
12	41.8	73.6	54.1	71.5	41.8	77.2	54.1	74.5	15	95	1B4G
15	41.3	71.3	51.6	76.0	41.3	74.4	51.6	78.3	15	95	1B4G
16	50.0	77.4	55.0	78.9	50.0	80.2	55.0	81.6	25	99	1B4G
17	49.3	80.8	55.1	74.8	49.3	82.1	55.1	76.8	15	95	1B4G
18	53.6	69.6	56.8	46.5	53.6	72.6	56.8	61.5	15	93	1B4G
19	56.9	67.9	58.1	59.8	56.9	72.5	58.1	67.4	20	98	1B4G
34	43.3	78.2	53.0	57.6	43.2	81.2	53.0	64.5	20	97	2A4E
30	33.7	79.6	37.2	68.7	33.7	81.3	37.2	71.7	10	80	2B4E
31	39.9	77.7	41.2	67.5	39.9	80.7	41.2	72.5	15	91	2B4E
32	44.4	77.5	51.0	65.8	44.4	79.5	51.0	70.9	20	97	2B4E
33	29.7	74.6	40.6	59.9	29.7	77.6	40.6	65.0	5	57	2B4E
35	50.7	78.8	49.8	65.7	50.7	82.8	49.7	72.0	20	99	2B4E
20	37.9	51.9	44.9	17.7	37.9	56.3	44.9	43.2	15	93	2C4E
21	42.9	53.6	50.4	43.1	42.9	61.1	50.4	55.6	20	97	2C4E
22	45.2	74.2	52.3	69.5	45.2	76.3	52.3	74.3	20	99	2C4E
23	33.4	72.5	42.7	74.6	33.4	75.5	42.7	77.6	15	91	2C4E
24	33.7	73.3	44.6	63.2	33.4	76.3	44.6	69.2	15	92	2C4E
25	38.4	58.0	42.4	73.1	38.4	64.0	42.4	77.0	15	98	2C4E
36	34.2	59.9	45.8	57.3	34.2	65.9	45.8	67.3	10	84	2C4E
41	36.5	55.5	41.4	55.3	36.5	62.5	41.4	67.0	5	60	2C4E
37	23.3	66.9	32.1	50.4	23.3	72.9	32.1	59.4	5	49	2D4E
38	37.2	74.7	46.7	53.3	37.2	77.7	46.7	61.0	20	97	2D4E
39	37.7	65.6	45.3	57.8	37.7	68.6	45.3	63.8	15	86	2D4E
40	42.5	60.6	42.6	61.2	42.4	64.0	42.6	67.2	10	71	2D4E
42	38.9	76.3	46.8	69.9	38.9	76.3	46.8	75.9	5	37	3A4A
43	58.9	70.0	52.4	58.8	58.9	76.0	52.4	69.8	10	56	3A4A
44	67.2	62.6	49.9	76.2	67.2	70.6	49.9	82.2	15	76	3A4A
45	58.6	78.8	56.5	69.8	58.6	81.8	56.5	75.8	25	88	3A4A
46	48.8	80.5	51.5	77.2	48.8	83.5	51.5	81.2	25	88	3B4A
47	33.4	82.8	50.3	69.9	33.4	84.8	50.3	77.9	15	73	3B4A
48	31.9	82.8	44.8	72.2	31.9	83.8	44.8	75.2	10	54	3B4A
49	40.2	71.8	45.5	79.4	40.2	74.8	45.5	83.4	10	62	3C4A
50	36.9	75.0	46.3	76.0	36.9	79.0	46.3	80.0	15	74	3C4A
51	31.0	72.8	52.2	73.1	31.0	77.8	52.2	78.1	20	85	3C4A
56	30.6	74.2	52.1	74.2	30.6	79.2	52.1	79.2	25	89	3C4A
52	35.2	80.4	53.9	66.1	35.2	82.4	53.9	72.1	25	90	3D4A
53	27.2	72.7	45.5	69.8	27.2	74.7	45.5	74.8	10	59	3D4A
54	21.8	63.0	42.3	73.8	21.8	68.0	42.3	75.5	15	70	3D4A
55	20.9	75.4	48.6	66.6	20.9	79.3	48.6	72.5	20	87	3D4A

Notes:
a) Test metallurgy is recovery at final cleaner concentrate grade.

b) Corrected metallurgy was arrived at by redistribution of the cleaner tailings.

c) Grind time refers to time in the laboratory test mill: Product P_{80} is proportional to grind time.

2.43 Gold, Silver and Minor Elements

In addition to the normal assays used for the computation of recoveries, the concentrates were assayed for silver, gold, mercury and arsenic. This data for selected tests is shown below in table 26 which shows average gold and silver concentrations and estimated recoveries. Averaged mercury and arsenic concentrations are reported in Tables 27 and 28 by type and for the Vangorda orebody.

TABLE 26

GOLD AND SILVER METALLURGY - VANGORDA ORE TYPES

ORE TYPE	TEST NO.	RECOVERY		GRADE	
		GOLD	SILVER	GOLD	SILVER
1A4G	3	54.6	70.6	4.8	692
1A4G	4	47.4	71.7	3.8	641
1A4G	5	42.2	63.3	4.1	685
AVERAGE	-	48.1	68.5	4.2	673
1B4G	17	28.9	62.7	7.5	584
1B4G	16	33.3	70.9	8.9	679
1B4G	11	23.5	56.3	7.2	588
1B4G	18	30.6	51.5	10.3	620
1B4G	19	23.1	50.3	8.9	696
AVERAGE	-	27.9	58.3	8.6	633
2A4E	34	59.7	81.5	7.5	638
2B4E	32	57.1	53.9	8.2	582
2B4E	35	60.2	64.4	8.9	714
2C4E	21	39.4	36.2	8.2	384
2C4E	22	39.0	42.5	6.5	360
2D4E	38	60.2	51.3	11.7	353
2D4E	39	52.8	42.7	12.3	353
AVERAGE	-	52.6	53.2	10.3	483
3A4A	44	7.0	46.4	2.4	822
3A4A	45	10.4	59.9	2.7	802
3B4A	46	9.6	61.0	2.2	579
3C4A	49	13.5	51.9	2.1	381
304A	52	17.0	67.9	2.5	415
AVERAGE	-	11.5	57.4	2.4	600

Notes: a) Recoveries calculated from composite head assay and unadjusted test distribution data.

b) It is estimated that silver and gold recoveries are accurate to within ±10%.

c) Au and Ag in g/tonne.

2.44 Predicted Plant Metallurgy

An approximate mine model based on preliminary cross sections generated by D. Hanson indicated the relative occurrance of the three major ore types. This data was then used to produce a weighted metallurgical balance for all elements of interest in the Vangorda ore body. See tables 27 and 28 below.

TABLE 27

VANGORDA METALLURGY BY ORE TYPE

ORE TYPE	RELATIVE OCCURRENCE IN DEPOSIT %	PREDICTED METALLURGY									
		ASSAYS						DISTRIBUTION			
		Pb	Zn	Au	Ag	Hg	As	Pb	Zn	Au	Ag
4G Baritic	31	53	*	6.0	650	80	50	84	*	40	65
		*	55	*	*	440	20	*	80	*	*
4E Pyritic	39	49	*	10.0	480	60	200	77	*	50	50
		*	51	*	*	250	50	*	72	*	*
4A Quartzitic	30	48	*	2.4	600	40	500	81	*	15	55
		*	53	*	*	250	200	*	79	*	*

Notes:
- a) Data from best adjusted test data from type testing program.
- b) Relative occurrence data calculated from the mine model available in 1980.
- c) Au, Ag, As, and Hg, in g/tonne.

TABLE 28

PREDICTED PLANT METALLURY - VANGORDA ORE

CONCENTRATE	ASSAYS						DISTRIBUTION			
	Pb	Zn	Au	Ag	Hg	As	Pb	Zn	Au	Ag
Lead	50	*	6.5	575	60	250	80	*	35	55
Zinc	*	52.8	*	*	300	100	*	77	*	*

Notes: a) Metallurgy calculated from weighted average of type testing data.

b) Au, Ag, As, and Hg, in g/tonne.

2.5 COMPATABILITY TESTING

2.5 COMPATABILITY TESTING

2.51 Laboratory Testwork

Having established the metallurgy for each ore type and then deducing the overall metallurgy for each ore body, the next point to consider was the feasibility of milling the various ores concurrently. As discussed earlier in this section, both pilot plant studies and all recent laboratory testwork have been designed to emulate achievable conditions in the modified Cyprus Anvil mill.

Since the Vangorda material is only a small fraction of the total remaining ore in the area, interaction effects were studied only between Grum and Cyprus Anvil ore. The testwork described here was designed in cooperation with W. Muir, Plant Metallurgist at Faro, and various members of the Feasibility and Development Group. The work was performed at the Lakefield Research Laboratory in 1979.

The testwork comprised several open circuit cleaner tests utilizing the standard flowsheet, with grinding at about 50 microns P_{80}, and a reagent pattern approximating to that employed at Cyprus Anvil. The samples consisted of drill core from the Faro deposit and the screened remnants of the pilot plant sample treated at Lakefield during the 1977 Grum testwork. Sample composition was as follows:

TABLE 29

HEAD ASSAYS OF COMPOSITES

SAMPLE	ASSAYS %				
	Cu	Pb	Zn	Fe	S
Grum	0.14	4.98	9.13	33.10	33.9
Cyprus Anvil	0.18	3.15	4.43	34.80	32.2

Notes: a) Grum sample originated from materials stored at Lakefield.

b) Cyprus Anvil sample was provided by W. Muir plant metallurgist and comprised principally Type 4E Pyritic species.

2.52 Base Metallurgy by Ore Type

First standard open circuit cleaner tests were performed on Grum and Cyprus Anvil samples to obtain data about the base conditions. The results of two test pairs are shown below in Table 30.

TABLE 30

BASE METALLURGY FOR GRUM & CYPRUS ANVIL SAMPLES

PRODUCT	WEIGHT %	ASSAYS % Pb	ASSAYS % Zn	% DISTRIBUTION Pb	% DISTRIBUTION Zn
CYPRUS ANVIL ORE					
Pb Cleaner Concentrate	3.01	74.9	1.81	81.7	1.3
Pb 1st Cleaner Conc.	4.61	54.8	4.31	91.3	4.6
Zn Cleaner Concentrate	6.04	0.36	52.80	0.8	74.3
Zn Rougher Concentrate	30.81	0.31	12.9	3.9	92.9
Zn Flotation Tailing	64.68	0.20	0.17	4.6	2.5
CYPRUS ANVIL ORE					
Pb Cleaner Concentrate	3.06	72.8	2.03	79.8	1.4
Pb 1st Cleaner Conc.	5.68	44.9	5.21	91.4	6.8
Zn Cleaner Concentrate	5.48	0.34	53.8	0.7	67.8
Zn Rougher Concentrate	13.00	0.48	29.5	2.2	88.3
Zn Flotation Tailing	81.32	0.22	0.26	6.4	4.9
GRUM ORE					
Pb Cleaner Concentrate	6.49	60.3	7.52	81.3	5.4
Pb 1st Cleaner Conc.	20.65	21.6	12.20	92.5	27.9
Zn Cleaner Concentrate	10.27	0.50	51.70	1.1	58.5
Zn Rougher Concentrate	27.25	0.65	22.80	3.7	68.4
Zn Flotation Tailing	52.10	0.36	0.65	3.8	3.7
GRUM ORE					
Pb Cleaner Concentrate	6.85	56.2	9.12	80.6	6.9
Pb 1st Cleaner Conc.	15.12	28.8	12.4	91.2	20.7
Zn Cleaner Concentrate	10.21	0.63	52.8	1.3	59.6
Zn Rougher Concentrate	21.60	0.70	31.1	3.2	74.1
Zn Flotation Tailing	63.28	0.43	0.73	5.6	5.2

2.53 Tests with Ore Mixtures

Two more tests were then performed with a 50:50 mixture of Grum and Anvil ores. The test procedures were the same and the results are reported below in Table 31.

TABLE 31

METALLURGY OF MIXTURE 50:50 GRUM & ANVIL

PRODUCT	WEIGHT %	ASSAYS %		% DISTRIBUTION	
		Pb	Zn	Pb	Zn
Pb Cleaner Concentrate	4.72	68.6	4.57	85.1	3.2
Pb 1st Cleaner Conc.	12.27	28.8	10.70	92.8	19.6
Zn Cleaner Concentrate	8.38	0.51	52.5	1.1	65.9
Zn Rougher Concentrate	23.80	0.53	21.60	3.3	76.9
Zn Flotation Tailing	63.93	0.23	0.36	3.9	3.5
Pb Cleaner Concentrate	4.41	69.8	5.30	80.5	3.5
Pb 1st Cleaner Conc.	9.30	37.3	9.65	91.2	13.5
Zn Cleaner Concentrate	8.01	0.38	53.50	0.8	64.9
Zn Rougher Concentrate	20.87	0.56	25.8	3.1	81.7
Zn Flotation Tailing	69.83	0.31	0.45	5.8	4.8

By rearranging and averaging the data in Tables 29 and 30 it is possible to show that the results for the mixture of ores falls, as expected, between that recorded for the two ore sources when separately tested.

An interesting point about these results is that the blended mixture produced results which were slightly better than the arithmetic average of both ore types individually. This is a peculiar but not unique occurrance with blends of ores and is probably due to preferential grinding of the relatively softer Grum ore.

TABLE 32

SUMMARY OF TEST DATA

CONDITION		ASSAYS		DISTRIBUTION	
		Pb	Zn	Pb	Zn
Anvil - Base Metallurgy	Lead Zinc	73.9 *	* 53.3	80.8 *	* 71.1
Grum - Base Metallurgy	Lead Zinc	58.3 *	* 52.3	81.0 *	* 59.1
50:50 Blend Metallurgy	Lead Zinc	69.2 *	* 52.8	82.8 *	* 65.1

Notes: a) All data from LR 2176 No. 5

b) Test data refers to batch tests which are not directly compable to the predicted plant data.

TABLE 33

COMPARISON OF

ANTICIPATED AND ACTUAL METALLURGY FOR BLENDED ORES

CONDITION		ASSAYS		DISTRIBUTION	
		Pb	Zn	Pb	Zn
Anticipated Metallurgy (Arithmetic Average)	Lead Zinc	66.1 *	* 52.8	80.9 *	* 65.1
Actual Metallurgy	Lead	69.2	*	82.8	*
	Zinc	*	53.0	*	65.4

Notes: a) Again data refers to batch cleaner test results, unadjusted for redistribution of cleaner tails.

2.6 SPECIAL CHARACTERISTICS OF THE CONCENTRATES

2.6 SPECIAL CHARACTERISTICS OF CONCENTRATES

There are several characteristics of mineral concentrates which are of considerable significance in shipping and sales which have not yet been discussed. The three most important factors, aside from major metal contents, are concentrate flow moitures, moisture content and trace element analyses.

2.61 Flow Moistures

Flow moisture is an approximate physical test method which permits an estimate to be made of the point at which plastic flow of concentrates might occur. Clearly the onset of plastic flow will depend on mean particle size, particle shape and the range of various sizes present. The effect of plastic flow of large concentrate masses on ship stability and rolling moment are apparent.

To determine the effect of finer grinding of the concentrates on flow moisture, samples originating from the fine grind tests at Lakefield Research were subsequent to a flow moisture test. The results of these standard tests are summarized below.

TABLE 34

FLOW MOISTURE

	NORMAL	FINE GRIND
Lead	8.1	9.7
Zinc	9.8	11.2

Notes: a) Normal - Nov. 1979 test result for a six month certificate.

b) Normal P_{80} for concentrates 30-40 microns.

c) Fine grind P_{80} 10-20 microns.

d) The increase in flow moisture point with finer grinding was unexpected. Usually flow moistures tend to decrease with finer particle size.

2.62 Moisture Contained in the Concentrates

The upgraded dewatering plant will accomodate all the planned concentrate production and produce concentrates containing 4.5% moisture under ideal conditions. However, these ideal conditions assumed in the theoretical calculations of dryer capacity seldom exist in practice. Therefore, as shown in the table below the moisture contents will decrease from present levels but will probably not immediately reach the target moisture content of 4.5%. The probable initial plant performance is shown below in Table 35.

TABLE 35

MOISTURE CONTENTS OF CONCENTRATES

CONDITION	LEAD CONC. % WATER	ZINC CONC. % WATER
1978-1979 Averages	5.2	6.6
Theoretical Plant Performance	4.5	4.5
Initial Predicted Performance	4.7	5.0

2.63 Spectrographic Analyses

Some trace elements exercise considerable influence on the smelter process and as such, incur significant penalties to the seller of concentrates. The detection of these minor or trace elements is usually achieved by performing a spectrographic analysis of the concentrates.

TABLE 36

SPECTROGRAPHIC DATA - FINAL CONCENTRATES

(Results from Can-Test Vancouver)

ELEMENT		VANGORDA						GRUM		ANVIL ORE	
		TYPE 4A		TYPE 4E		TYPE 4G		PILOT PLANT COMPOSITE		PILOT PLANT COMPOSITE	
		ZINC	LEAD	ZINC	LEAD	ZINC	LEAD	ZINC	LEAD	ZINC	LEAD
Aluminum	Al	1.	1.	1.	0.1	0.2	0.2	0.1	0.05	N.D.	N.D.
Antimony	Sb	N.D.	0.1	N.D.	0.05	N.D.	0.3	N.D.	0.15	N.D.	N.D.
Arsenic	As	TRACE	*	N.D.	0.03	N.D.	0.05	0.15	N.D.	0.01	0.0002
Barium	Ba	0.1	0.3	*	*	*	*				
Beryllium	Be	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.
Bismuth	Bi	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.
Boron	B	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.
Cadmium	Cd	TRACE	N.D.	TRACE	N.D.	TRACE	N.D.	0.05	0.1	0.05	N.D.
Calcium	Ca	1.	0.5	2.	1.	1.	1.	0.10	0.07	0.05	0.07
Chromium	Cr	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.
Cobalt	Co	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.
Copper	Cu	0.3	0.1	*	*	0.1	0.3	0.2	0.5		
Gallium	Ga	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.
Iron	Fe	MAJOR	MAJOR	MAJOR	MAJOR	MAJOR	MAJOR	MAJOR	MAJOR	MAJOR	MAJOR
Lead	Pb	*	MATRIX	*	MATRIX	*	MATRIX	MAJOR	MATRIX	MAJOR	MATRIX
Magnesium	Mg	1.	0.1	2.	0.3	1.	0.5	0.02	0.02	0.1	0.01
Manganese	Mn	0.1	0.07	0.3	0.2	0.2	0.3	0.03	0.01	0.02	0.01
Molybdenum	Mo	N.D.	N.D.	TRACE	TRACE	N.D.	TRACE	N.D.	N.D.	N.D.	N.D.
Niobium	Nb	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.
Nickel	Ni	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.
Potassium	K	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.
Silicon	Si	3.	5.+	0.5	1.	0.3	1.	1.5	1.5	0.1	0.1
Sodium	Na	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.
Strontium	Sr	TRACE	0.001	0.03	0.05	0.01	0.05	N.D.	N.D.	N.D.	N.D.
Tantalum	Ta	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.
Thorium	Th	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.
Tin	Sn	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.
Titanium	Ti	0.1	0.3	0.01	0.01	0.01	0.01	N.D.	0.01	N.D.	N.D.
Tungsten	W	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.
Uranium	U	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.
Vanadium	V	0.001	0.005	0.001	0.001	0.001	0.001	N.D.	N.D.	N.D.	N.D.
Zinc	Zn	MATRIX	*	MATRIX	*	MATRIX	MAJOR	MATRIX	*	MATRIX	*

Notes:

a) Percentages of the various elements expressed in these analyses may be considered accurate to within plus or minus 35 to 50% of the amount present.

b) Semi-quantitative spectrographic analytical results for gold and silver are normally not of a sufficient degree of precision to enable calculation of the true value of ores. Therefore, should exact values be required, it is recommended that these elements be assayed by the conventional Fire Assay Method. Quantitative and Fire Assays may be carried out on the retained pulp samples.

c) Silicon, aluminum, magnesium, calcium and iron are normal components of complex silicates.

d) MATRIX - Major constituent
MAJOR - Above normal spectrographic range
TRACE - Detected by minor amounts
N.D. - Not detected
* - Suggest assay (above 0.3%)

3
GRINDING

GRINDING

INDEX - GRINDING

3.0 SUMMARY

3.0 SUMMARY

The need for increased rod mill capacity and additional fine ore storage have been satisfied in the modified mill design by installing a completely new rod mill - ball mill grinding circuit to the West of the existing circuit. The total capacity of the modified circuits will be greater than that of the existing circuit even at the target grind levels of 50 microns P_{80}.

The design calculations for sizing the milling equipment were based on conservative estimates of both work index and operating time. Standard Bond procedures were used in the power calculations except for the determination of lead regrind mill capacity, where some adjustments were made to compensate for the relative softness of the lead mineral. This latter procedure was in accord with standard industry practice and yielded results which were in good agreement with operational practice in other concentrators.

Estimates of steel consumption were also derived using projected historical data; These estimates assured design operating conditions. Recommended media sizes were also calculated and again refer to operation at design conditions.

3.1 INTRODUCTION

3.1 INTRODUCTION

One of the principal concerns in designing the modified grinding circuit was to ensure minimum interference with normal plant operations during the construction program. Accordingly a completely new milling circuit was designed to be installed to the West of the existing circuit; the concept being that the present circuit would continue to function as usual through the construction period and, after the commissioning of the new circuit, the tonnage to the existing circuit would be reduced. Approximate calculations indicated that the existing circuit could produce the target grind of 50 microns P_{80} at a throughput rate of 200 tonnes/hour. The new circuit was then designed on the basis of treating 250 additonal tonnes/hour.

The design basis was quite conservative and reflected the difficulties of operating a complex plant in a remote location. For instance plant operating time to achieve planned tonnages at rated throughput would be only 86%. It is anticipated that actual plant operational capability will be in excess of 90%.

The design calculations were persued using standard Bond methods and by employing appropriate empiritical corrections which are standards in the industry. These calculations indicated that one new rod mill with a 1500 HP motor and two in-series ball mills each with 2500 HP motors would treat the desired tonnage and achieve the target grind level of 50 microns.

Similar calculations for the regrind mills indicated that the lead regrind mill will have to be very carefully controlled to achieve the desired grind; Both zinc regrind mill in parallel would be required to treat the zinc rougher concentrate.

3.2 FINE ORE STORAGE

3.2 FINE ORE STORAGE

One of the most critical areas, and often one which receives the least attention in mill design, is that of fine ore storage. There are two aspects of fine ore storage which exert considerable influence on the operational capability of the concentrator:

a) True storage capacity at rated tonnage.

b) Ease of withdrawal of ore from the storage bins under various climatic conditions.

3.21 Total Storage Capacity

Installed bin storage capacity varies greatly from one operation to another, and is influenced by considerations of crusher plant operating time and constraints imposed by climatic conditions. The data shown below in Table 1 indicate a fairly wide range of installed fine ore storage capacities. It is noteworthy that most operations in cold climates install in excess of two days capacity at rated tonnage.

TABLE 1

FINE ORE STORAGE CAPACITY - COMPLEX METALLURGY OPERATIONS

MINE	MILLING RATE (tonne/day)	FINE ORE STORAGE (tonnes)	THEORETICAL STORAGE TIME (days)
Black Moutain	3250	6000	1.85
Balmat	3450	3630	1.05
Kerr American	870	870	1.00
Edwards	340	340	1.00
Young	5010	3450	0.69
New Market	2720	2180	0.80
Jefferson City	1550	2270	1.46
Elmwood	2720	910	0.33
Brushy Creek	4540	910	0.20
Magmont	3810	3630	0.95
Ozark	4850	5440	1.12
Buick	5440	5440	1.00
Leadville	910	2720	3.00
Park City	580	1910	3.28
Bulldog Mountain	290	290	1.00
Tintac	680	540	0.80
Lucky Friday	620	910	1.62
Woodlawn	1000	2000	2.00
Star Unit	930	1180	1.27
Pend Oreille	1270	1630	1.29
Idarado	1500	1810	1.21
Bunker Hill	2120	2990	1.41
Pine Point	9980	8160	0.82*
Nanisivik	1810	4540	2.50*
Mattagami	3490	10890	3.12*
Mattabi	2720	6350	2.33*
H.B.M. & S. (Flin Flon)	6800	3180	0.47*
Sherritt (Ruttan)	9070	9070	1.00*
Noranda (Geco)	4540	7260	1.60*
Orchan	1860	4720	2.54*
Selco (South Bay)	450	730	1.60*
Texas Gulf	9070	24950	2.75*
Western	910	2270	2.50*

* Cold climatic conditions for most of year.

Note: All operations cited are either copper-lead-zinc or lead-zinc operations.

By grouping and averaging the data shown in table 1, it becomes apparent that the majority of cold weather operations are provided with an average of two days fine ore storage at rated capacity. In table 2, the present and proposed capacity at Anvil are shown together with averaged data from table 1.

TABLE 2

FINE ORE STORAGE CAPACITY

CONDITION	THEORETICAL STORAGE CAPACITY* (days)	ESTIMATED TIME CAPACITY (days)
Average - All Mills	1.5	1.2
Average Cold Zone Mills	2.0	1.5
Anvil - Present	1.0	0.75
Anvil - Future	1.6	1.2

Notes: a) Theoretical Capacity of a storage system is usually about 1.3 times the true capacity due to hangups, bin freezing, and poorly designed withdrawal systems.

b) Actual Ba capacity is also influenced by many factors, including climate, moisture content, size distribution and bin geometry

On studying the present Anvil concentrator, it becomes apparent that the only logical expansion of the fine ore storage area is to install two more bins of the same overall size, in line with and to the West of the existing bins, to give a total effective storage capacity in excess of 10,000 tonnes. To install more capacity to more closely emulate similar operations would be desirable, very expensive, and difficult to justify.

3.22 Ease of Withdrawal

The existing storage bins are equipped with a slot feeder withdrawal system, which after about ten years of continuous operation, has proved to be reliable, effective and easy to maintain under normal operating conditions. In the expansion of the ore storage area it is proposed to install a similar slot feeder system incorporating wider slots and additional wear protection along the slot lips, under the two new bins. Other withdrawal systems were considered, however since the predominant industry trend is toward slot feeder systems, because of reasons of reliability and live area considerations, and because of the satisfactory record of the feeder systems at Anvil, these other systems were rejected.

The bins will be supplied with fine ore by extending the present tripper car system. While tripper cars do have operational problems, the alternative, a shuttle conveyor is not viable over the required distribution distance. Once again the industry trend is very strongly toward tripper systems: The more sophisticated milling operations do however provide significant additional instrumentation to ensure good tracking and automatic car positioning.

3.3 ROD MILL CAPACITY

3.3 ROD MILL CAPACITY

A cursory examination of the design criteria suggested that no increase in rod mill capacity would be required in order to sustain a milling rate of 450 tonne/hour in the modified circuits. However, the excessive spillage of material around the rod mills during the last few years operation, and the very short life of the cyclone feed pump impellers and suction side liners, indicated that the present rod mill capacity was inadequate.

The volume flow through each of the existing rod mills is at times so great at the rated tonnage of 135 tonne/hour that, with a significant proportion of the ore types encountered at Anvil, spillage from the feed chutes and discharge trommels necessitated either an involuntary plant shutdown or at least a significant reduction in throughput. This condition has recently become increasingly associated with the quartzitic and pyritic ore types. These ore types predominate in the remaining Faro ore reserves and form a significant proportion of the known Grum and Vangorda ore reserves.

To investigate further the problems of the existing rod milling units, a detailed survey was undertaken to determine the practical operational capacity of small rod mills in other operations. The survey covered mining operations in various parts of the world involving many different ore types and modes of operation. The data is summarized in Table 3.

TABLE 3

SURVEY OF DATA - ROD MILL CAPACITY

SOURCE	COUNTRY	ROD MILL SIZE (ft)	MOTOR (HP)	TON /HOUR (ST/HR)	UNIT POWER KWH/S.T.	KWH/TONNE
Lachor	Canada	7 x 10	200	50	2.99	3.29
Algom-Nordic	Canada	8 x 12	400	80	3.74	4.12
Milliken	Canada	8 x 12	700	125	4.18	4.61
Algom-Quirke	Canada	8 x 12	400	70	4.27	4.70
Pheonix	United States	8 x 12	400	80	3.74	4.12
Woodlawn	Australia	8 X 13	500	60	4.55	5.02
N.B.H.C.	Australia	8 x 12	500	115	3.24	3.57
Friedensville	United States	8 x 12	350	55	4.70	5.18
Irado	United States	8 x 12	350	80	3.27	3.60
Tsumeb	South West Africa	9 x 10	350	60	3.27	3.60
Gold Fields	South Africa	9 x 12	450	72	4.67	5.15
Paragasha	Peru	9 x 12	350	60	4.35	4.80
Default	Canada	9 x 12	450	100	3.36	3.71
Heath Steel	Canada	9 x 12	500	95	3.93	4.34
Newfoundland Zinc	Canada	9 x 12	450	90	3.74	4.12
Gold Fields	South Africa	9 x 13	500	100	3.74	4.12
New Imperial	Canada	9 x 13	450	85	3.95	4.35
Craigmont	Canada	9.5 x 12	600	110	4.08	4.50
Mattagami	Canada	10 x 13	600	170	2.64	2.90
Stanleigh	Canada	10.5 x 14	800	130	4.60	5.07
Can. Met.	Canada	10.5 x 14	700	125	4.18	4.61
Stanrock	Canada	10.5 x 14	700	140	3.74	4.12
Brunswick	Canada	10.5 x 14	700	150	3.49	3.85
Texas Gulf	Canada	10.5 x 15	850	145	4.38	4.82

Note: a) Data sources were from published flowsheets and personal contacts.

Inspection of the data in table 3, indicates that most operations average considerably less than 80 S.T./hour, with a 9.0' diameter mill. Even allowing for the high density of the Faro area ore it is obvious that the present load on the rod mills is excessive.

The summary table 4 below demonstrates quite clearly that the existing rod mills (9.0' diameter) have a maximum operational capacity of about 80 tonnes/hour.

TABLE 4

SUMMARY OF ROD MILL DATA - DIAMETER VS CAPACITY

MILL DIAMETER (ft)	(mm)	OPERATIONAL CAPACITY S.T./HOUR	TONNES/HOUR	AVERAGE POWER DRAWN KW/S.T.	KW/TONNE
8.0'	2438	65	59	3.85	4.24
9.0'	2743	83	75	3.88	4.28
10.0'	3048	137	124.3	3.91	4.31

Note: a) Data was obtained by averaging all information for a specific mill size. Note power drawn is virtually constant per tonne milled.

The extent to which the present rod mills are overtaxed can be determined by refering to the data in table 5. It is apparent that the new configurations will permit the existing rod mills to function in a more reasonable fashion with an acceptable throughput rate.

TABLE 5

OPERATION OF 9.0' DIAMETER ROD MILLS

CONDITION	THROUGHPUT S.T./HOUR	THROUGHPUT TONNES/HOUR	UNIT POWER DRAWN KWH/S.T.	UNIT POWER DRAWN KWH/TONNE
Anvil Present	150	135	2.24	2.47
Survey Data	83	75	3.88	4.28
Anvil Proposed	74	67	4.54	5.01

Notes: a) Survey data from table 4.

b) Proposed data based on 200 tonnes/hour throughput on existing circuits.

3.4 DESIGN CALCULATIONS

3.4 DESIGN CALCULATIONS

Base Data:

These design calculations were performed in order to determine power requirements for all the grinding circuits operating in the new mode. Preliminary calculations had indicated that an additional 1500 HP of rod mill capacity and 5000 HP of ball mill capacity would suffice to produce the target grind levels at 250 tonnes/hour. Ancilliary calculations showed that with the new circuit operating at 250 tonnes/hour, the existing circuit could easily accommodate the remaining 200 tonnes/hour at the target flotation feed grind.

The power calculations were performed in accordance with Bond's formula, employing where applicable, the empirical correction factors for mill diameters, reduction ratios and fine grind effects. The following were regarded as base data for the final Bond calculations to determine product size and power requirements.

Calculation of power requirements for the regrind mills were performed in a similar fashion to that described above. Some adjustments were made to calculated power requirements to allow for the relatively softer minerals.

3.41 Work Index

Data from plant measurements laboratory test programs, comparative studies and the classic impact test method were gathered for each orebody and ore type. The data were arranged by orebody and the highest average value of W_1 observed taken as a design base: (e.g. W_1 = 13.224 KWH/tonne from Faro Zone III ore types)

There are of course, some inherent problems in using the high work index value for a design base. Clearly the mill, when handling softer Grum and Vangorda ores, will produce a much finer grind. (The mill operations personnel will then have the opportunity to increase tonnage and stabilize the primary grind P_{80} at the target levels). The table 6 below indicates the extent of the problem. Note that Vangorda ores have an optimum primary grind P_{80} of about 30-35 microns (Ref. Report KM 008 - Vangorda type 4G and 4E).

TABLE 6

PRIMARY GRIND P_{80} VALUES AND WORK INDEX

ORE SOURCE	WORK INDEX W_I	PRIMARY GRIND P_{80} AT RATED TONNAGE P_{80} microns	MAXIMUM TONNAGE AT TARGET GRIND tonnes/hour
Faro Zone III	13.22	50	450
Grum Ore	12.03	39	483
Vangorda Ore	9.24	30	507

Notes: a) For values of W_I see Survey Data.

b) Vangorda Maximum tonnage calculated at P_{80} = 33 μ. Grum P_{80} at 50 μ.

c) Work index data from Table 7.

d) Rated tonnage is defined as 450 tonnes/hour.

3.42 Tonnage Division

Calculations and power draw data indicated that apart from overloaded rod mills, the existing circuit was also suffering from an excessive throughput in the primary ball mills. Thus even if the present rod mill capacity were not to be increased an additional and critical production bottleneck would be encountered in the primary ball mills.

In order to reduce the effects of the mill modification construction work on production, and to solve the mill loading problems,the most logical route appeared to be to reduce tonnage to the existing circuit until the desired grind could be produced and to simultaneously install a new grinding circuit comprising both rod and ball mills to accomodate the remainder of the planned throughput.

Approximate Bond calculations indicated that the existing circuit could handle 200 tonnes/operating hour while the new circuit would have to be designed to treat the remaining 250 tonnes/hour. Accordingly detailed Bond calculations were performed applying where necessary the appropriate corrections to calculate product sizes and probable power draws for each unit.

TABLE 7

SOURCES OF WORK INDEX DATA

VALUE OF WORK INDEX			DATA SOURCE	ORE TYPE
	KWH/SDT	KWH/ Tonne		
	10.78		1978 Anvil Mill Records	Anvil Mill Feed
	12.32		1979 Anvil Mill Survey	Anvil Mill Feed
	11.59		Lakefield Pilot Plant L.R.2202/06	Anvil Pyritic Ore
	12.96		Lakefield Pilot Plant L.R.2202/06	Anvil Oxide Ore
Average	11.91	13.13		
	11.59		Lakefield Pilot Plant L.R.1191/11	Grum Bulk Sample (Tests 14-24)
	10.30		Lakefield Pilot Plant L.R.1991/11	Grum Bulk Sample (Tests 26-30)
	13.50		Lakefield Pilot Plant L.R.1991/11	Grum Bulk Sample (Tests 31-34 38-44)
	9.40		Lakefield Laboratory Work L.R.1991/10	All Grum Ore Types Averaged Data
	10.25		Lakefield Laboratory Work. L.R.1991/7	Grum Types C, D, G, F.
	10.46		Lakefield Laboratory Work L.R.2027	Grum Types A,B, C,D,F,K,J.
Average	10.92	12.03		
	8.20		Noranda Mines Report #5. Aug. 29, 1975	High Grade Vangorda
	7.40		Dowa Mining Co. Feb.1965: R1031	Vangorda Core Samples
	9.40		Noranda Mines Ltd. Report #1	Vangorda Pyrite Ores
	8.50		Noranda Mines Ltd. Report #1	Vangorda Baritic Ore
Average	8.38	9.24		

3.43 Correction Factors

The following correction factors, applied where necessary to the value of work index, were utilized in the detailed Bond calculations. These factors have been derived from empirical and theoretical studies and are the accepted method of Bond calculation correction.

Basically Bond's work was carried out at a time when 8.0' diameter mills were the largest available and grinding as a preparation for flotation was a poorly understood stage in mineral beneficiation. Thus as mill sizes increased and finer grinds were demanded by more complex ores, empiritical correction factors were developed and applied to Bond's basic formulae.

(i) Diameter Correction: (EF_3)

For mills greater than 8.0' diameter a correction is required to compensate for more efficient grinding. This factor is described by the equation:

$$EF_3 = \frac{8}{I.D.}^{0.2}$$

Where I.D. = diameter inside liners.

Where ID $\geq$ 12.5': $EF_3 \rightarrow 0.914$

(ii) Oversize Feed Correction: (EF_4)

This factor reflects the relative inefficiency of attempting to feed oversize material to a grinding mill. The factor is described as follows:

Feed size (microns) $F = 4000 \sqrt{\frac{13}{W_I}} = Fo$ (Efficiency feed in microns for ball mills)

$F = 6000 \sqrt{\frac{13}{W_I}} = Fo$ (Efficiency feed size in microns for rod mills)

(iii) Fineness Correction Factor: (EF_5)

Since the Bond calculation method becomes increasingly inaccurate below 50 microns, a correction factor for fine grinding exists and is defined as follows:

$EF_5 = \frac{(P + 10.3)}{1.145P}$ Where P = Product P_{80}

(iv) Reduction Ratio Correction: (EF_7)

Factor applies in ball mill grinding stages where the value of $R_R \geq 6$.

$$\text{Reduction Ratio } R_R = \frac{2\ (R_R - 1.35) + 0.26}{2\ (R_R\ \ 1.35)}$$

Where R_R = Feed P_{80}/Product P_{80}

Also check for rod milling design:

R_R (Rod Mills) $= 8 + \frac{5L}{D}$ Where L is mill length and D is the diameter.

(iv) Reduction Ratio Correction: (EF_7)

Factor applies in ball mill grinding stages where the value of $R_R \geq 6$.

$$\text{Reduction Ratio } R_R = \frac{2\,(R_R - 1.35) + 0.26}{2\,(R_R - 1.35)}$$

Where R_R = Feed P_{80}/Product P_{80}

Also check for rod milling design:

R_R (Rod Mills) $= 8 + \frac{5L}{D}$ Where L is mill length and D is the diameter.

(v) Power Transmission Efficiency:

Electrical power drawn at the mill motor decreases significantly through various losses encountered in the mill motor - clutch - pinion - gear system; The losses are collectively known as transmission losses.

Usual design with synchronous motor systems through a plate clutch calls for a transmission loss of about 3-4%. To ensure functional design, losses were assumed for these computations to be 5.0% of power drawn at the mill motor.

(vi) Power Draw Calculations:

Another problem with grinding circuit calculations is that mills of a particular size draw finite power quantities dependent upon critical speed fraction, charge and media type. Thus before meaningful grinding calculations can commence it is necessary to determine the maximum power drawn by the units in question. The following equations are relevant and refer to power drawn per unit weight of media.

Rod Mills $$\text{Power/Ton} = 1.07\sqrt[3]{D}(6.3 - 5.4\ VP)\ C.S.$$

Ball Mills $$\text{Power/Ton} = 3.10.D.^{0.3}(3.2 - 3\ VP)\ C.S.\left(1 - \frac{0.1}{2^{9-10C.S.}}\right)$$

Charge Weight $$\text{Tons} = \frac{D^2}{4}\pi VP\ L.\ \frac{Cd.}{2000}$$

Where D = Diameter inside liners (feet)

VP = Percent of mill volume loaded

C.S = Percent of critical speed

L = Length of grinding compartment

Cd = Media density 260 lb/ft

3.44 Detailed Bond Calculations

Before commencing the detailed calculations it is worth noting the objectives of the design calculations:

a) To provide a functional design which will assure that plan tonnages of ore can be treated to produce the grind levels dictated by economics and metallurgical constraints.

b) To ensure that each stage of the grinding circuit design is compatible with reasonable demand on equipment and conforms to standard design formula.

The table 8 below compares the present circuit with the proposed grind objectives.

TABLE 8

COMPARISON OF GRINDING CIRCUIT PRODUCTS

AREA	PARTICLE P_{80} - MICRONS EXISTING CIRCUIT	EXPANDED CIRCUIT
Flotation Feed	120 - 130	45 - 50
Lead Concentrate	35 - 50	14 - 16
Zinc Concentrate	35 - 40	18 - 20

Notes: a) Lead and zinc concentrate sizing analysis for present circuit taken from recent cy closizer reports: Ref. Met Engineering Ltd. October 19, 1979. Report work performed at University of Alberta.

b) Concentrate sizing analysis for expanded circuits based on pilot plant results: Ref. Lakefield Research 2202, No.6, Volume II.

3.45 Existing Grinding Circuits

The basic criteria for these calculations were discussed in the previous section. Design is for 200 tonnes/hour with a work index of 13.224 KWH/tonne and with a plant feed size F_{80} = 12,500 microns.

Based on the experience of others and at Anvil the rod mills are expected to draw about 400 HP each. Then work factor becomes:

$$W_o = \frac{3 \times 400 \times 0.747 \times 0.95}{200}$$

$$= \underline{4.26 \text{ KWH/Tonne}}$$

With $F_{80} = 12.500$: $\sqrt{F_{80}} = 111.8$: $\frac{10}{\sqrt{F_{80}}} = 0.0894$

Then Bonds Formula $W_o = W_I \left(\frac{10}{\sqrt{F_{80}}} - \frac{10}{\sqrt{F_{80}}}\right)$, but

since the mills are a different diameter from the standard

used by Bond, a diameter correction (EF_3) must be applied.

Then $4.26 = 13.224 \times 0.992 \left(\frac{10}{\sqrt{P_{80}}} - 0.894\right)$

Rod Mill discharge P_{80} = 582 microns.

The small ball mills will probably draw about rated load and thus work factor becomes

$$W_o = \underline{\frac{3 \times 450 \times 0.747 \times 0.95}{200}}$$

$$= \underline{4.79 \text{ KWH/tonne}}$$

With the new effective F_{80} = 582 the $\sqrt{F_{80}}$ = 24.1, and $\frac{10}{\sqrt{F_{80}}}$ = 0.414

Again the diameter correction Ef_3 is applicable.

$$W_I = 13.224 \times 0.992 = 13.11 \text{ KWH/tonne}$$

$$W_o = W_I \left(\frac{10}{\sqrt{P_{80}}} - \frac{10}{\sqrt{F_{80}}} \right)$$

$$4.79 = 13.11 \left(\frac{10}{\sqrt{P_{80}}} - 0.414 \right)$$

Then Ball Mill discharge P_{80} = 164.8 microns.

The final stage for the existing circuit is to determine the effect of the large ball mill. Here it becomes necessary to employ a diameter correction (EF_3) of 0.914 and a low reduction ratio correction (EF_7) based on a P_{80} estimate of 50 microns for the unit. (EF_7 = 1.067)

Ancillary calculations show that the large mill will draw only 2350 HP under load. Then the work factor

$$W_o = \frac{1 \times 2350 \times 0.747 \times 0{:}95}{200}$$

$$= \underline{8.34 \text{ KWH/tonne}}$$

Finally the Bond calculation can be applied using the two correction factors to modify W_I:

$$\text{Thus } W_I = 13.224 \times EF_3 \times EF_7$$

$$= 13.224 \times 0.914 \times 0.1067$$

$$= \underline{12.89 \text{ KWH/tonne}}$$

Then with $F_{80} = 164.8$ and $\sqrt{F_{80}} = 12.83$ then $\frac{10}{\sqrt{F_{80}}} = 0.779$

$$W_o = \left(\frac{10}{\sqrt{P_{80}}} - \frac{10}{\sqrt{F_{80}}} \right)$$

$$8.34 = 12.89 \left(\frac{10}{\sqrt{P_{80}}} - 0.779 \right)$$

$$\underline{P_{80} = 49.1 \text{ microns}}$$

3.46 New Grinding Circuit

Calculations were based on treating 250 tonnes/hour with the same plant feed size as the existing circuits. These calculations are somewhat more complex because of the numerous Bond corrections which were applied. Preliminary calculations based on an approximation indicated that the new rod mill could draw 1250 HP while the ball mills would each draw 2350 HP.

$$W_o = \frac{1 \times 1250 \times 0.747 \times 0.95}{250}$$

$$= \underline{3.55 \text{ KWH/tonne}}$$

A diameter correction (EF_3) is applicable although a feed size correction is not. Thus the modified work index

$$W_I = 13.224 \times 0.924 = \underline{12.22 \text{ KWH/tonne}}$$

Now $F_{80} = 12{,}500$ and $\sqrt{F_{80}} = 111.8$ and $\frac{10}{F_{80}} = 0.0892$

Then $W_o = W_1 \left(\frac{10}{\sqrt{P_{80}}} - \frac{10}{\sqrt{F_{80}}} \right)$

$$3.55 = 12.22 \left(\frac{10}{\sqrt{P_{80}}} - 0.0892 \right)$$

Rod Mill Discharge $\underline{P_{80} = 693 \text{ microns.}}$

The first ball mill requires a diameter correction and is marginal for a reduction ratio correction.

$$\text{Work Index} = 13.224 \times EF_3 \times EF_7$$

$$= 13.224 \times 0.914 \times 1.04$$

$$= \underline{12.57 \text{ KWH/tonne}}$$

Now $F_{80} = 693$ and $F_{80} = 26.3$ and $\frac{10}{F_{80}} = 0.380$

$$\text{Work Factor } W_o = \frac{1 \times 2350 \times 0.747 \times 0.95}{250}$$

$$= \underline{6.67 \text{ KWH/tonne}}$$

$$W_o = W_I \left(\frac{10}{P_{80}} - \frac{10}{F_{80}} \right)$$

$$6.67 = 12.57 \left(\frac{10}{P_{80}} - 0.380 \right)$$

<u>Ball Mill discharge P_{80} = 120 microns.</u>

The second ball mill exhibits precisely the same work factor as the first but the effect of EF_7 is much greater.

$$\text{e.g. Work Index } W_I = 13.224 \times EF_3 \times EF_7$$

$$= 13.224 \times 0.914 \times 1.08$$

$$= \underline{13.05 \text{ KWH/tonne}}$$

Again applying Bonds formula with F_{80} = 120 and F_{80} = 10.95 and $\frac{10}{F_{80}}$ = 0.913

$$W_o = W_I \left(\frac{10}{P_{80}} - \frac{10}{F_{80}} \right)$$

$$6.67 = 13.05 \left(\frac{10}{P_{80}} - 0.913 \right)$$

Ball Mill discharge P_{80} = 49.5 microns.

All the critical grinding circuit data deduced in the above calculations is summarized below in Table 9 for the modified mill circuits.

TABLE 9

SUMMARY OF DATA - ANVIL GRINDING CIRCUITS - MODIFIED MILL

MILL SIZE (FT)	MOTOR (HP)	POWER DRAWN (HP)	STAGE WORK INDEX KWH/TONNE	STAGE WORK FACTOR KWH/TONNE	PRODUCT SIZE P_{80} (MICRONS)
EXISTING CIRCUIT:					
9' x 12' rod	450	400	13.12	4.26	582
9' x 12' ball	450	450	13.12	4.79	165
13.5' x 22' ball	2500	2350	12.89	8.34	49
NEW CIRCUIT:					
12.5' x 16' rod	1500	1250	12.22	3.55	693
13.5' x 22' ball	2500	2350	12.57	6.67	120
13.5' x 22' ball	2500	2350	13.05	6.67	50

Notes: a) Power drawn calculated for the existing circuit with reduced tonnage is based upon several assumptions - e.g. less media/mill and different operating conditions.

b) Power draws in the new circuit are based on a range of theoretical conditions.

3.47 Regrinding Circuits

Based on test results from Lakefield Research pilot plant studies, the optimum P_{80} sizes for the concentrates were identified. The data are summarized below.

TABLE 10

ORE SOURCE	ESTIMATED REGRIND P_{80} SIZES (P_{80} microns) LEAD CONCENTRATE	 ZINC CONCENTRATE
Anvil	15	20
Grum	15	20
Vangorda	10 - 15.0	20

Notes:
- a) Anvil Data: Ref. Lakefield Report LR2202 No.6.
- b) Grum Data: Ref. Milling Committee Report, Noranda 1977.
- c) Vangorda Data: Average of best results from test work.

Using these results as a basis for calculation and assuming that efficient cyclones, cutting at the desired size are available, it is possible to estimate the power requirements for various levels of regrinding using Bond's formula.

It should be noted however that true sizing optima do not exist since in all probability all three ore types would benefit from regrinding to less than 10 microns. However there is a practical limit to regrinding imposed by sliming considerations and dewatering constraints.

a) Lead Regrinding Mill:

The lead mineral galena, is extremely friable due to its cubical crystal structure and weak intercrystalline bonding. This phenomenon has an appreciable effect on required grinding power calculations since the lead mineral tends to be preferentially reduced in grinding processes.

Utilizing results generated in the Lakefield Pilot Plant (Ref. L.R.2202 - No.6), a primary grind P_{80} of 50 microns in tests Nos. 9 and 10 resulted in an indicated rougher concentrate P_{80} of 30 microns or 60% passing 15 microns. Reference to the lead circuit mass flow drawing 10-F-02 Rev.C, indicates that 74 tonnes of lead concentrate will report to the regrind sump and presumably the passing 15 microns material will report to the cyclone overflow on the first pass. (e.g. Tonnes to Regrind Mill 74.0 x 0.40 = 29.6 tonnes.)

The regrind mill if charged correctly and operated at high density should draw about 475 HP from a 450 HP motor. Thus it is possible to calculate the work factor

$$W_o = \frac{475 \times 0.747 \times 0.95}{29.6}$$

$$= \underline{11.39 \text{ KWH/tonne}}$$

Now F_{80} = 30 and $\sqrt{F_{80}}$ = 5.48: Therefore $\frac{10}{\sqrt{F_{80}}}$ = 1.83

P_{80} = 15 and $\sqrt{P_{80}}$ = 3.87: Therefore $\frac{}{\sqrt{P_{80}}}$ = 2.58

Substituting in Bonds formula

$$W_o = W_1 \quad \left(\frac{10}{\sqrt{P_{80}}} - \frac{10}{\sqrt{F_{80}}} \right)$$

$$W_o = 13.224\ (1.58 - 1.83)$$

$$W_o = \underline{9.92\ \text{KWH/tonne}}$$

This is the uncorrected work factor and because of the mill operation two factors will have to be applied EF_3 and EF_7. Note that the reduction ratio is now only 2.0.

$$\begin{aligned} \text{Modified work index} &= W_1 \times EF_3 \times EF_7 \\ &= 13.224 \times 0.992 \times 1.20 \\ &= \underline{15.74\ \text{KWH/tonne}} \end{aligned}$$

Thus work factor corrected is now

$$\begin{aligned} W_o &= 15.74\ (2.58 - 1.83) \\ &= \underline{11.8\ \text{KWH/tonne}} \end{aligned}$$

CALCULATED APPLIED HP = 467 HP

Thus the theoretical work required to grind the rougher concentrate is close to the maximum which the regrind mill is capable of delivering (11.8 KWH/tonne vs 11.4 KWH/tonne). However in the calculations, the work index was assumed the same for rougher concentrates as for the original ore: This is very unlikely - the sulphide rich rougher concentrate is almost certainly very significantly softer than the overall ore. For this latter reason, the fine grinding correction EF_5 was not applied to the lead regrinding calculations.

b) Zinc Regrinding

The zinc sulphide mineral is tetragonal in structure and is quite resistant to grinding forces. Thus in the Pilot Plant work there was little evidence to suggest that there existed any selective overgrinding of the mineral. Zinc rougher concentrate production was estimated at about 64 tonnes/hour (drawing DF-10-03) and cyclosizer data showed that about 40% would pass 20 microns.

Tonnage to Regrind = 64.0 x 0.60 = 38.4 tonnes/hour

Now $F_{80} = 50$ and $\sqrt{F_{80}} = 7.07$ and $\frac{10}{\sqrt{F_{80}}} = 1.41$

$P_{80} = 20$ and $\sqrt{P_{80}} = 4.47$ and $\frac{10}{\sqrt{P_{80}}} = 2.24$

Substituting in Bond's Formula

$$W_o = W_1 \left(\frac{10}{\sqrt{P_{80}}} - \frac{10}{\sqrt{F_{80}}} \right)$$

$$= 13.224 \ (2.24 - 1.41)$$

$$= \underline{10.97 \text{ KWH/tonne}}$$

Corrections to work index will be for EF_3, EF_5 and EF_7. Again EF_7 is high because the reduction ratio is relatively low and EF_5 was applied because the zinc mineral was considered to be as hard as the remainder of the mineral assemblage.

$$\text{Corrected Work Index} = W_I \times EF_3 \times EF_5 \times EF_7$$

$$= 13.22 \times 0.992 \times 1.11 \times 1.32$$

$$= 19.22 \text{ KWH/tonne}$$

Thus the new work factor will be

$$W_o = 19.22 \ (2.24 - 1.41)$$

$$= 15.95 \text{ KWH/tonne}$$

Therefore applied power 15.95 x 38.4 = 613 KWH

<u>Calculated Applied HP = 820 HP</u>

Thus zinc rougher concentrate regrinding will require that both regrind mills be used. Very probably the best metallurgical advantage will accrue through using the mills in parallel with a cluster of small diameter cyclones on each mill.

c) Comparison of Data

The philosophy adopted during the calculations was therefore that the relative friability of the sulphides would more or less compensate for the decreased grinding efficiency at low reduction ratios. It is interesting to observe that, in their deliberations the Noranda Milling Committee elected to follow the same route toward assessment of regrind power needs and arrived at substantially the same results.

TABLE 11

REGRINDING POWER REQUIREMENTS

DATA SOURCE	ORE	POWER KWH/tonne LEAD CONC.	 ZINC CONC.
Noranda Milling Committee	Grum Composite	9.70	14.6
Lakefield Report LR2027 No.11	Grum Composite	11.3	7.6
Faro Ore Calculations	Anvil Composite	11.8	15.9

Notes: a) Power for regrinding on Grum refers to data from tests 22-37.

b) Unfortunately during the pilot plant work on Zone III the energy demand meters failed to register due to the very small concentrate quantities involved.

From table 11 it is apparent that the result of the Anvil regrind power calculations are in good accord with those deduced by the Noranda Milling Committee for Grum ores. As noted earlier Vangorda ores are relatively soft compared to either Faro or Grum ore. Thus if sufficient power is available for the harder ores, there will be an excess for softer ores such as Vangorda types 4G and 4A.

3.48 Maximum Power Draw Calculations

In the new circuit, the large mills represent a considerable part of the total power used in the mill. However because of various constraints, the mills will not draw rated power. Hence in the previous sections, the Bond calculations were performed using a calculated maximum power draw. The relevant data is shown below in summary form in table 12 indicating conditions for the maximum theoretical power draws.

TABLE 12

MAXIMUM POWER DRAW CALCULATIONS

UNIT ft.	CRITICAL SPEED %	UNIT POWER KW/TON MEDIA	MAXIMUM Draw Power KW	HP	COMMENTS
Rod Mill 12.5 x 16.0	68	6.93	973	1304	Culled 40% volume
	68	6.48	1023	1371	Culled 45% volume
	74	6.48	1058	1429	Culled 40% volume
	74	7.05	1113	1492	Culled 45% volume
Ball Mill 13.5 x 22.0	68	7.79	1565	2099	Forged Steel Balls
	68	7.79	1406	1882	Cast Iron Balls
	74	8.41	1690	2266	Forged Steel Balls
	73	8.41	1518	2032	Cast Iron Balls
	78	9.03	1815	2434	Forged Steel Balls
	78	9.03	1630	2182	Cast Iron Balls

Notes: a) Culling is often essential in large rod mills to ensure draw maximized. Placer Development practice this technique in most of their operations using rod mills of diameter greater than 12.0 feet.

b) Cast balls have a bulk density of only 90% of forged steel balls.

The data in the table above shows that good estimates for power drawn by the grinding mills would be 1250 hp for the rod mill and 2350 hp for the ball mills. As a matter of general interest, the present 13.5' x 22.0' mill at Cyprus Anvil usually draws 2350 hp at the motor.

3.5 ESTIMATED STEEL CONSUMPTION

3.5 ESTIMATED STEEL CONSUMPTION

Liner and media wear are usually predicted on the basis of past experience or from empiritical calculations utilizing the abrasion index concept. Fortunately the very considerable amount of historical data available from Cyprus Anvil records made prediction of liners and media consumption, relatively straight forward. The adjustments made to historical data are shown and discussed in the appropriate section below.

3.51 Sizing Media

Rods: In the expanded mill, two rod sizes will be used, small rods 3 1/2" x 12' for the existing rod mills and probably 3 1/2" x 16' for the new large rod mill. Presently the 3 1/2" diameter rods exhibit a wear rate of 380g/tonne; lower media charge levels in the mills and probably lower pulp densities will result in decreased power draw and probably less rod breakage too.

At rated capacity, the new rod mill will not be drawing full power and can most probably be operated with a relatively small media load and average pulp densities e.g. 80-85% solids at specific gravity 3.7-4.1. The maximum rod size of 3 1/2" was determined from the empiritical sizing equation shown below:

$$R = \left(\frac{F^{0.75}}{160}\right) \qquad \frac{\sqrt{W_i} \quad \times S_g}{(\%C_s) \quad \sqrt{D}}$$

R = Diameter of rod in inches

F = Feed size 80% passes in microns

W_I = Work Index

S_g = Specific Gravity

C_s = Critical Speed

In this case R = 3.25 - 3.50"

With these size rods in the large mill it would be reasonable to expect that rod consumption would be somewhat less than the present rod consumption for the small mills despite the greater tumbling distance.

Rod Consumption: Since the media consumption is principally related in a linear fashion to applied power, the normal method of calculation of consumption is to use the ratio of applied power method: The method outlined below in Table 13 where a 0.80 correction factor is applied to projected usage.

TABLE 13

GRINDING ROD CONSUMPTION

GRINDING ROD SIZE	POWER DRAWN (HP)		RATIO POWER DRAWN	CONSUMPTION (g/tonne)	
	Present	Projected		Present	Projected
3.5" (89mm)	3 x 450	3 x 400 1 x 1250	1.81	380	550

Another estimation method is the specific consumption technique. This is outlined below and gives similar results to the ratio of power drawn method.

TABLE 14

GRINDING ROD CONSUMPTION

CONDITION	POWER DRAWN		SPECIFIC CONSUMPTION	
	HP	KW	g/KWH	g/Tonne
Present	3 x 475	1065	150	375
Projected	3 x 400 1 x 1250	1830	140	570

3.51 Sizing Media (cont'd)

Balls: As the extensive and detailed work by many investigators (Bergmann et.al.) have shown, proper selection of grinding ball size is critical to the production of an optimized feed size prior to flotation. Too large a ball will result in wasted energy while balls which are too small will quite fail to smash the larger mill feed particles.

Calculation of the maximum ball size for any process is easily determined by using the formula shown below:

$$B = \sqrt{\frac{F}{K}} \qquad \sqrt[3]{\frac{S_g W_i}{(\%C_s)\sqrt{D}}}$$

Where K = 350 For wet overflow ball mills.

The table below compares present ball size selection with that predicted by the above equation.

TABLE 15

MAXIMUM BALL SIZE

STAGE	FEED SIZE F_{80} microns	MEDIA SIZE (inches) Calculated	Used Presently	Recommended
Primary Ball Mills	600 - 700	1" - 1 1/2"	2"	1 1/2"
Secondary Ball Mills	150 - 160	1"	1.5"	1 1/2"
Regrind Mills	30 - 40	Less than 1"	1.5"	1"

Now these calculations and the recommended media sizes are based on treating ore at rated throughput. If throughputs are to be increased above design or open circuit tertiary crushing is utilized, and the unit process F_{80} increases, then larger balls will be required. In the case of ultrafine grinding, such as regrinding there is a practical limit to the ball size which can be employed. Generally in the mining industry the smallest ball which can be obtained and utilized is about 1" diameter.

Ball Consumption: Media consumption has been estimated assuming that the present ball sizes are utilized rather than the recommended sizes. The data, derived by the ratio of power draw method is shown below in Table 16. (It is considered unlikely that ball size will have any significant effect on steel consumption.)

TABLE 16

GRINDING BALL CONSUMPTION

TYPE OF SUPPLY	POWER DRAWN		RATIO POWER DRAWN	CONSUMPTION	
	Present	Projected		Present kg/tonne	Future kg/tonne
Grinding Balls 2" (51mm)	3 x 500	3 x 450 1 x 2350	2.47	0.35	0.735
Grinding Balls 1 1/2" (38mm)	1 x 2350 2 x 400	3 x 400 2 x 2350	1.87	0.70	1.113
TOTALS				1.05	1.848

Notes:
a) Future consumption was calculated from the ratio of power drawn and a correction factor of 0.85.

b) Power drawn in present case assumes only two regrind mills in action.

c) Present data refers to 1978-79 consumption. which occurred while treating Type E material.

3.52 Liner Consumption

The most accurate method of predicting liner wear is to have available operating data over a long period. Fortunately the type of data is available for most of the unit operations under consideration.

Because liners are often replaced before they are completely worn down it is not really practical to use a specific consumption method, as is used with grinding media. Indeed with the advent of rubber liners this method is almost impossible to employ. Liner consumption then is usually expressed in terms of tonnes/set - e.g. the number of tonnes which can be treated before a liner replacement is required. Even this technique runs into problems however when the unit concerned does not treat the full tonnages.

In Table 17 are shown consumption in terms of the estimated replacement interval assuming treatment of 3.4 x 10^6 tonnes per year at rated capacity.

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TABLE 17

GRINDING LINERS CONSUMPTION

TYPE OF SUPPLY	REPLACEMENT INTERVAL (years)		COMMENTS
	Present	Predicted	
Rod Mill Liners (9.0')	0.6 - 0.7	1.0 - 1.2	Predicted on basis of tonnage treated.
Rod Mill Liners (12.5')	N.A.	0.6 - 0.8	Estimated from diameter ratio.
Ball Mill Liners (9.0')	3.0 - 3.5	4.0 - 5.0	
Regrind Mill Liners (9.0')	5.0 - 6.0	3.0 - 4.0	Reflect greater utilization of mills.
Ball Mill Liners (13.5')	3.5 - 4.5	3.0 - 3.5	Two of mills in series on new circuit.

Notes: a) All ball mills are expected to be equipped with rubber liners.

b) Rod mill liners assumed to be steel Noranda wave type. Recent tests at Faro with Ni hard have yeilded excellent results which surpass those reprinted in table 17.

c) Replacement intervals in terms of years at rated capacity of $3.4\text{x}10^6$ tonnes/year.

4
FLOTATION

FLOTATION

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4.2 CIRCUIT DESIGN FACTORS INFLUENCING THE FLOTATION PROCESS

4.3 DETERMINATION OF FLOTATION CAPACITY

4.4 ANCILLARY EQUIPMENT IN FLOTATION

4.0 SUMMARY

4.0 SUMMARY

In the Cyprus Anvil flotation section major renovations will occur with the removal of all the obsolete Wemco 76 cells and their replacement with the newly developed large volume Outokumpu 38 m^3 supercharged flotation cell. These new cells will be arranged so as to provide rougher scavenger and first cleaner capacity for both lead and zinc circuits. The existing supercharged Denver DR-200 cells will remain on site and serve as the second and third cleaner stages.

The installation of an on-stream analyzer system for up to fourteen pulp streams and capable of analyzing for four elements plus some sophisticated mass flow monitors will ensure functional control capability in these new circuits. Additional systems will provide the necessary logistical support for the flotation process control.

The provision of a new reagent mixing and storage building will ensure that reagents are dissolved and diluted in a safe and consistent manner. A new reagent distribution system for all reagents will greatly enhance metallurgical control capability in the flotation plant.

4.1 INTRODUCTION

4.1 INTRODUCTION

The flotation process is without doubt the most critical and at the same time the least understood phase in mineral separation operations. Basic research into flotation phenomena has continued at about the same level for at least two decades and innovations in treatment schemes have for the most part developed in laboratories associated with operating mines.

Similarly the development of new flotation equipment was until recently extremely slow due to a lack of demand for innovation by the operators. However during the past five years there has occurred a major technological renaissance. Plants are now being designed on a totally different basis, that of large unit operations equipped with sophisticated control apparatus and directed by advanced control strategies.

The realization that grinding was not simply a comminution stage but a crucial flotation feed preparation phase was the first step in development of the new design philosophies. Next the concept that air was one of the most critical reagents in the flotation system was developed. Finally simple control strategies, based on computerized sub-systems, developed and used easy to achieve mechanical changes to effect process control.

In the expansion of the flotation section of the concentrator considerable attention was directed toward utilization of these latest technological advances. Large volume cells manufactured by the Outokumpu Company of Finland, were chosen to replace the obsolete Wemco cells, and provisions were made for the installation of an on-stream analyzer system and a sophisticated array of monitoring apparatus. This new equipment will ensure that optimum efficiency is achieved in the minerals separation process and will provide the basis for the future development of a fully automatic centralized process control system.

In the following sections the principal factors influencing flotation are considered first and then the methods used to ensure the required flotation residence times and hence equipment needs outlined. Finally the ancilliary systems on which the flotation process is so dependent for smooth operation, are discussed briefly.

4.2 CIRCUIT DESIGN FACTORS INFLUENCING THE FLOTATION PROCESS

4.2 CIRCUIT DESIGN FACTORS INFLUENCING THE FLOTATION PROCESS

There are a myriad of factors which may influence the efficiency of a flotation process on a complex high sulphide ore. Unfortunately most of these influences are inter-related and hence are extremely difficult to accurately quantify. Design of a flotation circuit however requires that some attempt be made to recognize the major influences and to design the circuit accordingly.

In the following section, the factors regarded as significant in the flotation treatment of lead-zinc ores, are discussed. It is not intended that each factor be subjected to a detailed treatise, but simply to record that these factors were taken into account in the design calculations. The factors affecting flotation are discussed and where appropriate the discussion is supported with examples and data from other mines. It should be noted that the effects of each factor on a specific operation should not be directly applied to the Anvil operation.

4.21 Pulp Density

Pulp density is often a major influence on flotation circuit behaviour. High pulp densities often result in the occlusion of non activated particles into the froth bed and also increase the rate of flotation of ultra fines. The effects of high pulp density are best seen in complex metal circuits with quite high specific gravity ores, and show up as increasingly non selective flotation as pulp densities increase. The data below is typical of the type of problems encountered in circuits operated at high pulp density.

TABLE 1

EFFECT OF PULP DENSITY

ON ZINC CLEANING

PULP DENSITY %	INSOLUBLE CONTENT IN CONCENTRATE %
5	1.50
10	2.05
15	2.50
20	3.05
25	3.05
30	3.45

Notes: a) Data source - Kamioka Mine Recycle Water Study 1975-77.

To some extent the use of supercharged cells does tend to offset minor increases in pulp density, since this type of cell does disperse flotation air in relatively small bubbles and hence aids selective flotation. However in terms of general practice, even the best flotation cells cannot compensate for very high densities. As a general rule complex sulphide ores require densities of about 35–45% solids by weight, or 10–15% by volume in the rougher stages while cleaner circuits benefit from the lowest density which can be designed into the circuit. Typically lead-zinc cleaning takes place at 10–20% solids by weight or 2–5% by volume. The table below summarizes current practice.

TABLE 2

DATA SURVEY OF PULP DENSITY RANGES

ORE TYPE	MINERALS	PERCENT SOLIDS BY WEIGHT			
		ROUGHERS		CLEANERS	
		RANGE	MEAN	RANGE	MEAN
Massive Sulphide Ore	Copper and iron flotation	28-40	34	10-30	20
Lead-Zinc ore	Lead flotation	30-40	35	10-30	20
	Zinc flotation	20-30	25	10-25	18
Disseminated polymetalic ore	Copper flotation	18-33	15	10-23	16
	Pyrite and gold flotation	18-40	30	18-30	24
	Lead flotation	25-35	30	15-20	18

Notes: a) Data source - V.A. Glembotskii
Flotation 1965.

4.22 Rate of Flotation and Residence Time

The key factor in determining a flotation circuit performance is the rate of flotation of the various mineral species of interest. Rate of flotation is best described as approximating to a reaction of the type.

In ${}^{f}/_{t}$ = K.T. Where f and t are the assays of the feed and tails

Where T mean flotation or residence time

Where K is a rate constant.

Now it is apparent that the value of t will decrease as the value of T increases. e.g. metal recovery is proportional to flotation time or the residence time in the process. Clearly too there is a limiting value of t below which increases of flotation time (T) will not result in further reductions of the tailing assay.

Thus the rate of flotation governs mineral recovery for a particular residence time. Assuming that rate of flotation is fixed then residence time must be increased to improve recovery up to the limiting value of the tailings assay or to some other specified point.

Some typical flotation rate data is shown below in table 3 for lead-zinc mills operated by the Boliden group of mines.

TABLE 3

FLOTATION TIME T (min per cell) and

FLOTATION RATE k min^{-1}; ln f/t = kT;

f and t = assay of feed and tailing

ORE	FEED TONS PER HOUR	LEAD ROUGHER-SCAVENGER FLOTATION			ZINC ROUGHER-SCAVENGER FLOTATION		
		NUMBER OF CELLS	GALENA		NUMBER OF CELLS	SPALERITE	
			T min per cell	K min^{-1}		T min per cell	K min^{-1}
Laisvall	75	28	1,7	0,066	-	-	-
Vassbo	30	34	2,1	0,054	-	-	-
Saxberget	20	32	3,1	0,027	32	3,3	0,024
Svardsjo	30	28	4,3	0,011	28	4,7	0,021
Garpenberg N	36	28	3,6	0,027	-	-	-
Garpenberg	43	28	3,6	0,017	28	2,5	0,033
Langdal	80	34	1,5	0,027	26	1,8	0,061
Ravliden	20	16	4,6	0,025	14	4,7	0,037

* Data from 1970 Lead-Zinc Symposium

For galena at Laisvall and Vassbo the flotation rate is two to three times faster than it is for the more complex ores at the other plants.

The rate in the lead and zinc cleaners varies between 0.5-1.5 tonnes per m^2/hour in lead cleaners and 0.5-2.00 tonnes/m^2/hour in the zinc cleaners.

4.23 Air Dispersion and Aeration in Pulps

Air is one of the most critical components of a flotation system. Gases contained in the air are adsorbed onto mineral surfaces, cause oxidation, affect solubility of minerals, and as bubbles, form the transport system to the pulp surface for the selected minerals.

TABLE 4

EFFECT OF AERATION ON LEAD RECOVERY AT CONSTANT GRADE

LEAD GRADE % Pb	LEAD RECOVERY WITH AERATION %	LEAD RECOVERY WITHOUT AERATION %
30	72	56
25	82	74
20	85	81

Data Source - Vangorda Test V133-138
- Noranda Mines Report 1965

The dispersion of air into the pulp is an important factor in determination of flotation efficiency. For maximum efficiency a flotation machine must evenly disperse bubbles into the pulp to permit bubble - particle contact. Average bubble size should be about 0.50 mm diameter equivalent according to theoretical and practical studies. Also dissolution of air into a pulp is advantageous - a distinct advantage of deep cell design.

Intensity of flotation is generally believed to be controlled by the amount of air present in the flotation pulp. The quantity of air influences the rate of oxidation of minerals, thus altering their flotation behaviour. Often this phenomenon is utilized to enhance mineral separation and aeration stages are incorporated in many complex metal circuits. Table 4 shows some typical results of aerations of pulp.

4.24 Temperature Effects

Over the last few years, especially since the advent of rubber lined grinding mills, many operations have reported significant metallurgical improvements at elevated pulp temperatures. This is not really surprising since rate of reaction at phase interfaces invariably increases as pulp temperatures rise.

Characteristically even small increases in temperature will cause substantial increases in rate of flotation due probably to changes in rate of surface oxidation at higher temperatures. In the case of lead-zinc ores, several operations report that heating of zinc flotation pulps by the introduction of steam results in improved zinc metallurgy. Currently this practice is employed at Tintac (USA), Tverrfjellet (Norway), Meggen (Germany), Brunswick and Nanisivik (Canada) and the Zinc Corporation (Australia). The data in the table below indicates the extent of the advantage of pulp heating at Brunswick.

The presence of an apparent optimum at 30°C is interesting: The Meggen and Tverrfjellet operation have not reported an optimum temperature but operate their zinc circuits at 28 - 30°C as do the Australians at Zinc Corporation's Rosebury concentrator. Nanisivik with an operation in the high Arctic, heat the flotation pulp to 27°, but have not reported an optimum.

At Cyprus Anvil there are no plans for direct heating of the pulp however, the combined effects of recycling 30-50% of the total mill water supply, the use of a steam heating system in the concentrate stock tanks and the greatly increased grinding circuit power, could elevate the overall pulp temperature from the present 8 - 10°C to about 15°C.

TABLE 5

TEMPERATURE EFFECT ON NET SMELTER RETURN

PULP TEMPERATURE °C	CHANGE IN NET SMELTER RETURN $/tonne/milled
18	0.30
24	0.60
28	0.90
30	1.35
34	0.75

Notes: a) Data Source - Lead - Zinc Handbook 1970

b) Change in N.S.R. recalculated from original data from Brunswick Mines.

4.3 DETERMINATION OF FLOTATION CAPACITY

4.3 DETERMINATION OF FLOTATION CAPACITY

In the previous section the various factors affecting flotation were briefly discussed. This next section will draw heavily upon these discussions in this, the derivation of required flotation capacity.

4.31 Pulp Density for Various Circuits

Using the data generated in the previous section and the results of both pilot plant work and survey data, pulp densities were selected for the new circuits which would provide optimum conditions for flotation. The densities selected are shown below compared with surveyed data.

TABLE 6

PULP DENSITY DATA

(% weight)

	SURVEY DATA	PRESENT PLANT	NEW PLANT
Rougher/Scavenger	30-40	60-70	35-45
First Cleaners	15-25	40-50	20-25
Final Cleaners	10-20	40-45	15-20

Notes:
a) Survey data from Table 2.
b) Present plant data from plant survey data.
c) New plant data - calculated design.
d) All data assumes a specific gravity of 3.7 - 4.1.

4.32 Residence Times

The pilot plant studies on both Grum and Anvil ores indicated that because of the inherent rate of flotation of the minerals at the target grinds and conditions, certain mean residence times were required. By computing mass balances utilizing the pulp densities adopted, and the required residence times, the values of cell capacity needed for each stage were easily determined. The table below indicates the calculated residence times for the present circuit, the proposed circuit, and for comparison the pilot plant data too.

TABLE 7

A COMPARISON OF RESIDENCE TIMES
(minutes)

FLOTATION STAGE	ANVIL PRESENT	GRUM PILOT PLANT	ANVIL PILOT PLANT	ANVIL PROPOSED
Conditioning/Aeration	0	5.0	5.0	5.0
Lead Rougher/Scavenger	16.9	18.0	28.1	27.0
Lead 1st Cleaner	19.8	9.0	14.2	24.0
2nd Cleaner	6.8	3.5	12.8	20.1
3rd Cleaner	20.0	2.5	7.0	19.6
4th Cleaner	*	2.0	*	*
Zinc Conditioners	0	12.0	6.0	5.0
Zinc Rougher/Scavenger	25.9	23.5	14.5	26.7
Zinc 1st Cleaner	24.5	20.0	27.0	37.0
2nd Cleaner	19.5	6.5	15.0	17.3
3rd Cleaner	15.0	6.0	10.0	11.9
4th Cleaner	*	5.5	6.4	*

Notes: a) Sources of data were as follows:

Anvil Present - In plant measurements
Grum Pilot Plant - Ref. Report L.R.2027 No.11 - Vol. 1
Anvil Pilot Plant - Ref. Report L.R.2202 No.6 - Vol. 1
Anvil Proposed - Calculation Kilborn Drawing 10F-02/03.

Of special note when considering residence times are the aerators and conditioner stages for lead and zinc respectively. Almost invariably pre lead flotation aeration, especially with high iron ores, enhances lead-iron separation (See table 4 for aeration data). Normally aeration for at least 15 minutes is recommended as a practical minimum. However, two OK 38 cells, located at the head of the lead roughers with a five minute retention or residence time and their superior air dispersive capability should be sufficient to aerate the pulp and stimulate flotation. Similarly two more OK 38 cells, acting as zinc conditioners, will provide excellent pre-zinc flotation conditioners.

4.33 Installed Flotation Capacity

The capacity of flotation cells installed in each stage of the circuit are outlined below. For the purposes of comparison the present cell capacity is shown also.

TABLE 8

PRESENT AND PROPOSED CELL CAPACITY

UNIT PROCESS	PRESENT CIRCUIT (ft^3)	PROPOSED CIRCUIT (ft^3)
Lead Rougher/Scavenger	6000	13500
Lead Cleaners	6300	14100
Zinc Rougher/Scavenger	9000	13500
Zinc Cleaners	7100	14100
TOTAL	25400	55200

The doubling of installed cell capacity appears at first sight to be somewhat excessive, despite meeting the obvious constraints of residence time and pulp density. However inspection of data from other properties indicates that the installed cell capacity per tonne milled is actually within normal range for complex sulphide ore separations.

TABLE 9

INSTALLED CELL CAPACITY PER TONNE TREATED

MINE	INSTALLED CAPACITY/TONNE ft^3/tonne	m^3/tonne
Anvil Present	2.80	0.079
Anvil Proposed	6.09	0.172
Brunswick	6.45	0.183
Orchan	6.06	0.172
Lac Dufault	6.39	0.181
Mattagami	4.08	0.116
Mattabi	5.07	0.144
Tintac	6.66	0.189

Data Source - Noranda Milling Committee
- Lead Zinc Symposium 1970.

4.34 Critical Flotation Circuit Parameters

Finally and to summarize all the relevant flotation circuit design data, are shown estimated solid contents, stage recoveries, metal contents and specific gravity of solids. (Table 10 & 11)

TABLE 10

LEAD FLOTATION

STAGE	FEED M.T.P.H.	% SOLIDS	FLOTATION TIME(min)	STAGE RECOVERY	CONC. GRADE	SOLIDS S.G.
Roughers	564	45	11	80	19.5	4.0
Scavengers	485	40	16	45	3.0	4.0
First Cleaners	176.0	25	24	85	30.0	5.0
Second Cleaners	89	25	20.1	75	53.0	5.5
Third Cleaners	33	20	19.6	70	67.0	6.0
OVERALL				87.5	67.0	

TABLE 11

ZINC FLOTATION

STAGE	FEED M.T.P.H.	% SOLIDS	FLOTATION TIME(min)	STAGE RECOVERY	CONC. GRADE	SOLIDS S.G.
Roughers	472	39	10.6	80	34.1	4.0
Scavengers	408	35	16.1	60	7.0	4.0
First Cleaners	115	25	37.0	85	39.0	4.2
Second Cleaners	89	25	17.3	80	50.5	4.2
Third Cleaners	54	20	11.9	75	53.5	4.2
OVERALL				88.5	53.5	

Notes: a) Data contained in tables 10 and 11 relate to Kilborn Mass Flow Dwgs.

4.4 ANCILLARY EQUIPMENT IN FLOTATION

4.4 ANCILLARY EQUIPMENT IN FLOTATION

In addition to the more obvious equipment requirements in the modified flotation area, there are a number of ancillary systems which will be necessitated to ensure functional operation of the flotation section.

4.41 Reagent Handling System

Optimum control of a flotation circuit depends heavily upon the capability of delivering precisely controlled reagent dosages to the appropriate place in the circuit. The reagent delivery system must be of high reliability and be supported by sufficient storage and mixing vessels to ensure continuous operation under the exigencies of normal concentrator operation.

The existing reagent storage, mixing and delivery systems are inadequate, and in some respects unsafe; Certainly the present cyanide mixing and storage systems do not conform to industry norms for handling this dangerous substance. The features of the proposed new system are discussed briefly below in a general description of the new reagent area. This area will be located just to the north of the present concentrator separated from it and housed in a new 60' x 120' light frame structure.

a) Storage Capacity:

The new area will provide for increased storage space for all solid and liquid reagents in concentrated form. In addition, the entire cyanide system from the dissolution phase to the storage tanks will be completely enclosed and remotely operated by a discrete control module. Provision will also be made for flammable liquid storage, under conditions and in a zone acceptable to the insurance brokers. In all areas, ventilation will conform to regulation requirements.

Storage capacity for dry soda ash will be increased from the present 100 tonnes in two small bins located inside the concentrator building, to about 300 tonnes in a free standing, pneumatically loaded tank situated close to the new reagent building.

b) Mixing Bay:

All reagents will be mixed or diluted in one area. Soda ash will be slurried and partially dissolved and then delivered to the appropriate grinding or flotation circuit via a pressurized loop system. This system will permit good soda ash distribution in the grinding mills, ensuring that this critical phase of flotation feed preparation is performed at near constant pH. Other reagents will be mixed, diluted and fed via pressurized loop and solenoids to the desired addition points. In all cases sufficient storage for ready to use reagents will be provided to ensure smooth operation on weekends and over statutory holidays.

c) Delivery System:

The design of a reagent delivery system for a complex metallurgical circuit must be such that some flexibility be built into the system; In addition of course the system must be reliable and precise enough to exactly dispense the desired reagent quantity. The table below indicates the general distribution of reagents to the various unit operations.

TABLE 12

REAGENT ADDITION POINTS

STAGE	REAGENT								
	Na_2CO_3	NaCN	Z-11	MIBC	$CuSO_4$	$Ca(OH)_2$	1012	PERCOL	AERODRI
Rod Mills	X	X							
Ball Mills									
Aerators			X						
Lead Rougher/Scavenger			X	X					
Lead Regrind	X	X	X						
Lead Cleaners	X								
Conditioners			X		X	X			
Zinc Rougher/Scavenger			X				X		
Zinc Regrind			X		X	X			
Zinc Cleaners						X			
Stock Tanks									X
Thickeners								X	

Notes:
a) Soda ash would be added to all grinding mills.
b) Cyanide would be added in proportion to applied power.
c) Lime addition to the zinc conditioners would precede copper sulphate addition.

TABLE 13

SUMMARY OF DATA - REAGENT CONSUMPTION

(all data in grams/tonne mill feed)

REAGENT	1979 MILL ACTUAL	GRUM PILOT PLANT	ANVIL PILOT PLANT	PROJECTED EXPANDED MILL
Lime	1100	2500	4500	900
Soda Ash	1600	1800	9000	1400
Copper Sulphate	325	1050	750	400
Xanthate	230	190	185	180
Cyanide	100	150	290	135
M.I.B.C.	20	40	55	15
Dow 1012	4	35	40	2
Sodium Sulphide	0	0	0	0
Flocculant	3	N/A	N/A	5
Alcopol	30	N/A	N/A	10

Notes:
a) N/A indicates not applicable

b) 1979 mill taken from the December 1979 Report

c) Pilot Plant data is derived from a rough average reagent used in all tests in series.

d) Sodium sulphite is still undergoing evaluation at the time of writing.

TABLE 14

REAGENT STORAGE CAPACITY

REAGENT	CONSUMPTION PROJECTED g/tonne	SYSTEM CAPACITY (DAYS)		REAGENT CONCENTRATION AT ADDITION POINT %
		TOTAL	MIXED: READY TO VSE	
Na_2CO_3	1400	15-20	N.A.	10
NaCN	135	4-5	3-4	10
Xanthate	180	5-6	3-4	10
C_4SO_4	400	10-15	7-10	14
$Ca(OH)_2$	900	10-15	2-3	25
M.I.B.C.	15	3-4	3-4	100
DOW1012	2	15-20	15-20	100
Aerodri	10	7-10	7-10	100
Percol	5	5-6	3-4	1

Notes:
- a) Capacity based on 9,300 tonnes/day at projected comsumption.
- b) Concentration data is a best estimate.
- c) Total system capacity does not include reserve capacity of unmixed reagents on site.

4.42 Compressed Air System

The removal of the induced air Wemco 76 cells and their replacement with the large supercharged OK 38 cells, created a need for a large volume of low pressure air. The Outokumpu cells require air at a somewhat higher pressure than that supplied by the existing Spencer blowers supplying the Denver DR-200 cells. For this reason, a new set of Spencer blowers was included in the expansion package.

TABLE 15

AIR REQUIREMENTS BY CELL

CELL FUNTION	REQUIREMENTS		INSTALLED	
	PRESSURE (psi)	VOLUME/CELL (ft^3/min)	PRESSURE (psi)	VOLUME (ft^3/min)
Outokumpu OK 38				
Lead Rougher/Scavenger	4.4	600-850	4.4	3x14100
Zinc Rougher/Scavenger				
Lead First Cleaner	4.4	800-1000		
Zinc First Cleaner				
Denver DR-200				
Lead Second-Third Cleaners	3.0	140-160	3.0	3x9800
Zinc Second-Third Cleaners				

The three new blowers, with two operating at any given time, will be housed in the area currently used for reagent storage and mixing. The existing Spencer blowers will be relocated to serve the DR-200 cleaner cells.

4.43 On-Stream Fluorescence Analysis

An Outokumpu Courier 3000 on-stream system will be installed to monitor the flotation circuit behaviour. The system which is supplied complete with the necessary ancilliary equipment, is capable of handling 14 pulp streams simultaneously, although design concepts for the initial installation will require that six streams will be employed. The concept for the deployment of the additional analyzer capacity is shown below in Table 16 together with the approximate time from envisaged for each stage of the work.

TABLE 16

REPORTED METALLURGICAL IMPROVEMENTS - ON-STREAM CONTROL

PLANT	METAL	IMPROVEMENTS %		COMMENTS
		GRADE	RECOVERY	
CANADA				
Flin-Flon, Manitoba	Cu Zn	1.70 1.35	0 0	Six year sample period.
Strathcona, Ontario	Cu	1.5-2.0	0	On-stream control reduced variances of rougher concentrate grade significantly.
Inco, Ontario	Ni	0.46	2.0	Grade improvement was achieved at 8-9% concentrate grade (pentlandide). Recovery improvement is an estimate only.
Froode-Stobie, Ontario	Ni	0	2-3	Data quoted in response to question as an informed estimate.
Ecstall, Ontario	Ni	0.63 0.80		Data recorded over two years of plant operation. Reagent reduction of 10% was also recorded.
Lac Dufault, Quebec	Cu	0	1-2	Data quoted on basis of several years plant operation Reagent savings in excess of 25%.
AUSTRALIA				
New Broken Hill	Zn	0	1-2	Estimated after two years operation.
FINLAND				
Vihanti Pyhasalmi	Cu Zn	0 0	2.5 2.0	Data quoted in general review of automated mills in Finland after several years experience.
JAPAN				
Kamioka	Pb Zn	0.2 1.78	3.7 0.66	Automatic control systems are applied to the bulk flotation circuit. Frother consumption showed a significant decrease, as did cyanide usage.

TABLE 17

ON-STREAM ANALYZER - APPLICATION BY PHASE

PHASE	PURPOSE	STREAMS	ESTIMATE TIME TO COMPLETE
I	To familiarize operators with unit and to perform the necessary calibration tasks.	Lead Scavenger Tail Zinc Retreat Tail Lead Circuit Feed Lead Concentrate Zinc Concentrate Tails. - Final	Six Months
II	To improve metallurgical understanding of circuit roughers and cleaners.	Zinc Scavenger Tails Lead 1st Cleaner Tails	Six Months
III	Improve cleaner circuit control and assist in setting parameters.	Lead 2nd Cleaner Tails Zinc 2nd Cleaner Tails Lead Rougher Concentrate Zinc Rougher Concentrate	Six Months
IV	To provide data for setting up mechanical controls and and air for Outokumpu cells.	Lead Rougher Tails Zinc Rougher Tails	Six Months

Apart from the analyzer hardware, the system is complimented by a P.D.P.11 digital data processor with a 16 bit word length and a 16K core memory. The function of this unit is to convert the output signals from the analyzer amplifier scaler unit to a form suitable for use by a high speed typewriter. Courier software to achieve these objectives, including a theoretical fluorescence model are included as part of the P.D.P.11 package. Some companies have, to their great cost, attempted in the past to generate their own improved "model"; This idea was considered but was then rejected because of the lack of available specialist personnel at Anvil.

In conjunction with the analyzer will be the instrumentation associated with each sampling station. This instrumentation which will be principally of an indicating function will be as follows:

a) Sample pump on/off indication at control centre.
b) Low flow indicator/alarm in X-ray cell.
c) Low level indicator/alarm in the cell head tanks.
d) Mylar film rupture alarm.

A substantial inventory of spare parts including spectrometer, x-ray generator source and spare processor modules are included in the package, as well as a 5 tonne capacity air conditioner and humidity controller. These spare modules will permit the operating staff to send out the sophisticated electronics of the system for check and scheduled maintenance.

4.5 REAGENT CONSUMPTION

4.5 REAGENT CONSUMPTION

4.51 Comparative Reagent Consumption

Reagents utilized in flotation processes are almost exclusively surface active chemicals - e.g. their reaction depends upon the amount mineral surface available. Thus with increased surface area it would be logical to expect reagent consumption to increase linearly. In fact reagent usage is governed by other factors, including temperature of the pulp; pulp constituents, reagent interaction, reagent decay rates and metering system efficiencies.

In this section is outlined the rationale for the reagent consumptions predicted for the modified plant operating at rated tonnage at the target grind. The predicted consumptions were based on actual operation and the two sets of pilot plant data. Shown below are the summarized results.

TABLE 18

SUMMARY OF DATA - REAGENT CONSUMPTION

(all data in grams/tonne mill feed)

REAGENT	1979 MILL ACTUAL	GRUM PILOT PLANT	ANVIL PILOT PLANT	PROJECTED EXPANDED MILL
Lime	1100	2500	4500	900
Soda Ash	1600	1800	9000	1400
Copper Sulphate	325	1050	750	400
Xanthate	230	190	185	180
Cyanide	100	150	290	135
M.I.B.C.	20	40	55	15
Dow 1012	4	35	40	2
Sodium Sulphite	0	0	0	0
Flocculant	3	N/A	N/A	5
Alcopol	30	N/A	N/A	10

Notes: a) N/A indicates not applicable.

b) 1979 mill taken from December 1979 Report.

c) Pilot Plant data is derived from a rough average reagent used in all tests in series.

It is possible that the control of the reagent dispensing system is perhaps the most critical of all when assessing reagent consumption. In the modified plant the reagent systems have been designed to provide the most precise control of reagents and thus affect the maximum savings.

Soda Ash:

Present: 1,600 g/tonne Projected: 1,400 g/tonne

Soda Ash consumption is expected to be reduced despite the finer grinding. The following were considered:

a) Probably the present lead flotation pH of 9.8-10.1 is too high: Generally Noranda operate their concentrators at pH 9.2-9.4. Operation at this level would result in a very significant decrease in soda ash consumption.

b) Use of soda ash in the lead cleaners, especially the regrind mill will increase consumption. The extent of the increase is not known but pilot plant work indicates that about 25%-30% of the total soda ash could be used in the cleaner circuit.

c) The planned conversion of the current dry soda ash system, to a more practical, solution or slurry system, will reduce overall consumption by permitting more efficient reagent utilization, and the elimination of reagent loss due to spillage.

On average then it is expected that these factors influencing soda ash consumption will tend to balance each other and that soda ash consumption will be reduced slightly to the 1400 g/tonne level.

Lime:

Present: 1,100 g/tonne Projected: 900 g/tonne

Based on data from Grum and Anvil pilot plant work it is possible that total lime consumption will decrease despite the increased area available for surface oxidation of iron minerals. The following factors were considered:

a) In both Grum and Anvil pilot plants the lead cleaner circuits were modulated with soda ash rather than lime.

b) Pilot plant work (Ref. L.R.2202) indicated that lime was required only in the zinc roughers and the zinc first cleaners. The pH was then allowed to decrease from the pH 11.0-11.5 set point in the subsequent stages of cleaning.

c) Recycling thickener overflows to the zinc cleaner launder sprays will result in significant recycling of water at pH 10-10.5.

d) The use of conditioners to provide retention time while Ca^{2+} ion depression of pyrite proceeds, will probably result in a reduction of lime consumption.

Copper Sulphate:

Present: 325 g/tonne Projected: 400 g/tonne

A significant increase in copper consumption is expected due to several major factors.

a) It is generally considered that as surface area increases, so will copper sulphate demand. Since new surface area created is related to increases in applied grinding power, the modified circuits will cause an increase of new surface area and hence copper sulphate requirements, roughly in proportion to the new grinding power applied. Allowing for fine grind effect area increases by a factor of 0.85 for each doubling of power applied.

b) Relatively minor changes in copper sulphate demand may accrue due to improved reagent control systems, use of the on-stream analysis to relate copper addition to the zinc head assay, and the utilization of the zinc conditioners. In general as control strategies improve copper sulphate consumption is reduced.

c) It is worth noting that the pilot plant mode of operation, that of controlling froth by using copper sulphate, is atypical and results in excessive activator consumption.

Xanthate:

Present: 230 g/tonne Projected: 180 g/tonne

The slight reduction shown in xanthate consumption may be pessimistic. As indicated below, there are good reasons to expect greater reductions:

a) Xanthate consumption at Anvil is very high compared to almost any other lead-zinc operation in the world. The reasons for this anomaly are complex but chief amongst them is the control strategy in the lead cleaners where attached xanthates are desorbed with lime. In a soda ash modulated circuit this will not occur and overall xanthate consumption should fall significantly.

In the zinc circuit, the use of xanthates to "pull" cleaner circuits where the current cells are mechanically inept results in wastage. This situation will be corrected in the circuit with the increased cell capacity and better cell mechanics.

b) The use of lead conditioners-aerators will improve the distribution of xanthate in the pulp and hence reduce consumption.

c) On-stream monitoring of the lead circuit feed and the increased residence times in the lead roughers will permit the adoption of a "starvation collection" scheme which will greatly enhance metallurgy while reducing collector consumption.

d) After the implementation of the on-stream analyzer system, it should be possible to ratio $CuSO_4$/xanthate in fixed proportions - again a significant reduction in collector usage may be expected. Some Finnish and Swedish mines currently employ this practice with marked success.

e) Recycling water through an ionic flotation system will result in a small increase in xanthate consumption. At this time however it is not known if the treated water will contain appreciable quantities of dissolved xanthates which would, when returned to the zinc circuit, effect a reduction in collector use in that area.

Cyanide:

Present: 100 g/tonne Projected: 135 g/tonne

The effects of cyanide on lead flotation are very complex, however as a generalization, it can be said that cyanide depresses pyrite and sphalerite due to ionic adsorption, and effects chemical changes on the surface of lead sulphides which enhance xanthate adhesion.

a) Cyanide consumption is very dependent on surface area, therefore it can be expected that the new circuits with the finer grind will require more cyanide than present operations.

b) Improvements in the cyanide distribution system will make the best use of the cyanide added to the circuits, and it may be possible to effectively ratio cyanide to soda ash demand at a fixed pH with the more sophisticated soda ash and cyanide feeding arrangements.

c) Finer concentrate regrinding will involve more cyanide consumption to ensure adequate depression of the newly liberated iron and zinc minerals from the lead concentrates. Currently consumption in the regrind mill is about 10 g/tonne mill feed; this figure could rise to 25 g/tonne with the ultra-fine lead concentrate regrind.

Frothers:

M.I.B.C. Present: 20 g/tonne Projected: 15 g/tonne
DOW1012 Present: 4 g/tonne Projected: 2 g/tonne

With improved reagent control systems and better mechanical capabilities with the cleaner cells, minor reductions in frother consumption may be possible. Currently about 15% of the total frother consumed is to correct wayward froth conditions which should be adjusted by mechanical means.

It is entirely possible that recycling thickener overflows to the zinc circuit will almost eliminate the need for the frother 1012; the frothing characteristics of the filter aids are well know. Conversely recycling may result in a build-up of Ca^{2+} which will cause the circuit water to harden slightly, possibly requiring some extra frother to maintain a stable foam.

Sodium Sulphite:

Present: None Projected: None

This subtle reagent is probably the most misunderstood in the entire reagent array. Sodium sulphite is a powerful depressant for both iron and zinc minerals ranking somewhat below gaseous SO_2 treatment in effect. Numerous laboratory tests have shown that zinc metallurgy is almost always enhanced by small additions of sulphite in the 1,000 g/tonne range. There is little doubt too that as the fineness of grind is increased, the need for sulphite could decrease. Laboratory tests could determine the effectiveness of sodium sulphite at fine grinds.

Thus, no provision has at this time been made for sodium sulphite usage. In the event of oxide ore treatment where sulphite is very effective, some temporary feeding arrangements will have to be made. Should the laboratory program of testing indicate an advantage for sulphite modulated circuits, a new storage and feeding system will have to be separately justified.

Flocculant:

Present: 3 g/tonne Projected: 5 g/tonne

Improved mixing and a better control of flocculant addition could effectively reduce consumption. However, the ultra-fine concentrates envisaged in the expanded mill design, will require that more flocculant be used to ensure thickener overflow clarity.

Because of the need to recycle water from the thickener overflows to the zinc cleaner circuit it would appear that addition of lime as a flocculant is precluded hence more flocculant of the complex polymeric type will be required. Data from Lakefield testing indicates too that possibly two different flocculants might be more efficacious than the present ubiquitous Percol 351.

Filter Aid:

Present: 30 g/tonne Projected: 10 g/tonne

Filter aids are essentially surface tension reducers and depend for their effect on homogeneous dispension throughout the filter feed pulps. Currently a poorly designed distribution system wastes 20-30% of the total filter aid consumed, and the effect of the reagent actually added to the pulps is diminished, due to incomplete mixing.

The installation of stock-tanks and flow regulators on the thickener underflows and an improved reagent distributor system will eliminate waste and greatly enhance the effect of the reagent. Very possibly the estimated consumption of 20 g/tonne is pessimistic.

TABLE 19

REAGENT ADDITION BY AREA

ADDITION POINT	MAXIMUM REAGENT FLOWS (CC/MIN)								
	Na_2CO_3	NaCN	Xanthate	$CuSO_4$	M.I.B.C.	1012	$Ca(OH)_2$	Aerodri	Flocculant
Rod Mills A	Yes	4,500							
B		4,000							
Ball Mills									
Cyclopal Overflows			2x3,000						
Lead Aeration					140				
Lead Roughers									
Lead Scaveneers			5,000		70				
Lead Regrind	Yes	1,000							
Lead Cleaner 1	Yes		5,000		50				
Lead Cleaner 2									
Lead Cleaner 3	Yes								
Zinc Conditioner			10,000	2,500		40	Yes		
Zinc Roughers									
Zinc Scaveneers			5,000						
Zinc Regrind							Yes		
Zinc Cleaner 1			5,000	5,000		10	Yes		
Zinc Cleaner 2									
Zinc Cleaner 3									
Lead/Zinc Stock Tanks								2x250	
Lead/Zinc Thickeners									2x5,000

Notes:
a) These maximum flow estimates were made by metallurgical staff on site (1979-1980).
b) Yes indicates that flow will be directed to that point and controlled by a pH probe.
c) Solution strengths are 10% except for lime which is 20% and the remainder are full strength. Flocculant is delivered as a 1% solution.

5
DEWATERING

DEWATERING

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5.0 SUMMARY

5.0 SUMMARY

Calculations based on laboratory test data and plant experience detailed in this section, show that the proposed modifications to the Cyprus Anvil mill will necessitate some increases in dewatering plant capacity. A re-organization of the existing thickener tanks will provide adequate settling area for the very fine concentrates under normal conditions, however the use of adequate dosages of suitable flocculants is recommended. The thickener operations will be facilitated by the installation of underflow density controlled systems and duplex diaphragm pumps supplying concentrate stock tanks.

Considerable increases in filter capacity are planned to accommodate the finer grind of the concentrates. Complete renovations of the dryer feed belt configuration and a reassignment of dryer duty will permit the concentrates to be adequately dried in the final dewatering phase. In addition, the probability of cross contamination of concentrates in the dewatering system will be greatly reduced by dividing the dryer basement into dedicated lead and zinc areas and recycling liquid spillage to the appropriate thickener.

5.1 INTRODUCTION

5.1 INTRODUCTION

As the extensive laboratory and pilot plant testwork has demonstrated enhancements in metallurgical results require that the concentrates be ground very fine indeed to ensure maximum liberation of intergrown mineral crystals. Finer grinding of concentrates does however create significant additional problems for the dewatering section. A reduction of average particle size from 30 microns to 15 microns for example reduces median terminal settling velocities by a factor of eight - e.g. the effective thickener area should be eight times as great. Similar approximations can be made for filtering efficiency fall off, with decreasing particle size - here though, the perceived effect is usually a thinner filter cake and an increased cake moisture. The dryers are least affected by particle size, although moisture entrainment in the dryer discharge pellets and excessive dust recirculation through the scrubbers are factors to be considered.

With the increased grinding and regrinding effects then substantial additional loads will be placed on the dewatering section. Further because dewatering is the ultimate function in a metallurgical plant it is mandatory that no bottlenecks develop in this area.

Accordingly design parameters were selected to accomodate the majority of worst anticipated conditions. Reorganization of the existing thickener tanks, a large increase in filtering capacity, and a reassignment of dryer functions will permit smooth operation at design tonnages, feed grade and final concentrate grind levels.

5.2 DEWATERING DESIGN - BASIC PARAMETERS

5.2 DEWATERING DESIGN CALCULATIONS - BASIC PARAMETERS

The function of dewatering of the concentrates is accomplished by means of three unit processes, thickening, filtering and drying. In performing design calculations for a dewatering section it is important to recognize that significant excess capacity must be built into each unit process: The function of this excess capacity is to absorb short term load fluctuations on each of the unit processes and also to preclude the formation of a production bottleneck in the series of dewatering operations. In the table below are shown the design tonnage capacity data compared with the plan production tonnages.

TABLE 1

DEWATERING CIRCUIT DESIGN CAPACITY

CONCENTRATE	PLAN PRODUCTION		DESIGN PRODUCTION		DESIGN FACTOR %
	tonne/hour	tonne/day	tonne/hour	tonne/day	
Lead	21.2	510	25.2	605	1.19
Zinc	36.7	881	53.2	1277	1.45

Notes: a) Plan production based on 1984 output and corrected for 0.85 operating time factor. 1984 was selected as a base point because in that year concentrate output will reach a maximum, based upon case 4BV development schedule.

Because zinc content of the mill feed is almost invariably greater than the lead content, even quite a small percentage variations in zinc mill feed grade exercise a significant effect on concentrate output and hence dewatering circuit load: Accordingly a larger design factor was built into the zinc base calculations.

Design criteria then were generated on the assumption that the tonnages shown in table 1 could be dewatered at the target concentrate grinds to eventually produce concentrates containing an average of 4.5% water and clear thickener overflows. Basic sources of unit capacity data were of course the theoretical design calculations but these were extensively supported by laboratory testwork and also by data derived from surveys of other operations.

5.3 THICKENING

5.3 THICKENING

5.41 Laboratory Test Data

Extensive laboratory tests were carried out on various occasions with concentrates from both Anvil and Grum deposit ores. The tests, which were of the standard terminal velocity type and also compression tests were almost exclusively performed at Lakefield Research Laboratory. As is normal with settling tests, variation in pulp pH, flocculant type and gravity were studied.

The data shown below has been abstracted from the more voluminous test reports referenced below. Note the effect of flocculant addition on unit load in tests T4, T5 and T9.

TABLE 2

LEAD & ZINC CONCENTRATE THICKENING TESTS

ORE SOURCE	TEST NO.	pH	FLOCCULANT g/tonne	FINAL DENSITY %	SETTLING RATE m/h FEED ZONE	SETTLING RATE m/h COMPRESSION ZONE	UNIT LOAD tonnes/ft^2/day	UNIT LOAD tonnes/m^2/day
Faro -Lead	T1	10.0	-	75	1.5	0.7	1.43	15.4
	T2	11.5	-	74	1.7	0.6	1.43	15.4
	T3	11.0	-	74	1.8	0.6	1.30	14.0
	T4	10.0	20	73	0.6	0.5	2.26	24.4
	T5	10.0	10	73	5.6	0.6	2.26	24.4
Grum -Lead	L1	8.0	-	77	1.0	0.07	0.62	6.7
	L2	9.0	-	76	0.8	0.2	0.94	10.2
	L3	10.0	-	76	0.9	0.2	0.94	10.2
Faro -Zinc	T6	9.0	-	73	0.4	0.13	0.42	4.5
	T7	11.3	-	73	0.3	0.10	0.39	4.2
	T8	11.9	-	73	0.3	0.08	0.36	3.9
	T9	11.9	20	68	0.9	0.14	0.52	5.6
Grum -Zinc	Z1	8.1	-	60	0.20	0.14	0.32	3.5
	Z2	10.0	-	57	0.22	0.11	0.27	2.9
	Z3	11.0	-	59	0.26	0.11	0.27	2.9
	Z4	12.0	-	58	0.27	0.14	0.32	3.5

Notes:
a) Data from Lakefield pilot plant test concentrates for both Grum & Cyprus Anvil. Ref. L.R.2027 & L.R.2202.

b) Concentrates tested in anticipated concentrate grind range. Pb P_{80} = 15 microns: Zn P_{80} = 20.

c) Thickener area base data derived from compression zone settling rates only.

Now laboratory test data usually indicates that the required thickener settling area is considerably less than normal operating practice demands. This phenomenon arises because of the presence of several efficiency reducing factors commonly found in normal operations:

i) Thickeners operate with considerable circulating loads generated by recycling of scrubber and filtrate fluids.

ii) Upcurrents in the thickeners due to excessive volumes of feed and the use of water and air to relieve cone congestion, contribute to impaired settling of fine particles.

iii) Laminar flow conditions, while often encountered under laboratory settling test conditions, seldom prevail in a plant, thus reducing effective settling rates.

While laboratory tests provide some order of magnitude data for thickener sizing it is apparent, from the divergence of calculated thickener areas indicated by the testwork, that some other means must be found to support the laboratory calculations. The most obvious source of supporting data is operating practice in other mines.

5.32 Survey Data

The data gathered in the survey of other operations is shown below in table 3. The unit settling area per tonne of solids was calculated from the reported concentrate output data based on a plant operating continuously.

TABLE 3

UNIT SETTLING AREAS IN PLANT PRACTICE

PLANT	CONCENTRATE	CONCENTRATE SIZING P_{80} microns	UNIT LOAD tonnes/ft^2/day	tonnes/m^2/day
Brunswick	Copper	20	0.06	0.65
	Lead	15	0.30	3.24
	Zinc	20	0.30	3.24
Sherrit-Fox	Copper	50	0.20	2.18
	Zinc	50	0.31	3.35
Sherrit-Ruttan	Copper	70	0.19	2.05
	Zinc	70	0.06	0.65
Texas Gulf	Copper	20	0.14	1.51
	Lead	20	0.02	0.22
	Zinc	20	0.05	0.54
Sullivan	Lead	30	0.40	4.32
	Zinc	30	0.40	4.32
Dufault	Copper	40	0.06	0.65
	Zinc	40	0.02	0.25
Santander	Lead	60	0.04	0.45
	Zinc	60	0.12	1.30
Masua	Lead	35	0.03	0.30
	Zinc	30	0.20	2.18
Tintac	Lead	30	0.03	0.30
	Zinc	30	0.06	0.65
Bunker Hill	Lead	40	0.12	1.32
	Zinc	30	0.12	1.35
Ozark Lead	Lead	40	0.21	2.22
	Zinc	50	0.04	0.40
Flin Flon	Copper	60	0.16	1.70
	Zinc	40	0.14	1.47
Mattabi	Copper	50	0.03	0.32
	Zinc	55	0.13	1.42
Nanisivik	Lead	70	0.13	1.42
	Zinc	50	0.23	2.49

Notes: a) Concentrate sizing P_{80} are estimates only.

As an adjunct to the reported survey data, the current design methods of several major engineering companies was noted; The results are quite interesting and reflect the very conservative trends in the thickener design practice. There is no doubt that this conservatism springs from the well merited lack of confidence in laboratory test data.

TABLE 4

CURRENT THICKENER DESIGN PRACTICE

COMPANY	DESIGN CRITERIA tonnes/m^2/day	Data Source
Inco (new mills)	1.8	Inco Report to C.M.P./78
Highmont	1.5	Teck Corp.
Mattabi	1.3	Noranda
Island Copper	1.6	Placer Development
Placer(all mills)	1.1 - 2.2	Placer Development
Kilborn	1.8	Kilborn Vancouver

Notes: a) These data represent minimum design criteria which would be adjusted for grind, water temperature and average specific gravity.

5.33 Sizing Calculations

In considering all sources of information on thickener design, it becomes apparent that there is a wide divergence of opinion regarding basic data handling. First, the laboratory test results may be analysed by one or more of several methods and the results interpreted in various ways. Since, with the same data set, each of the methods produces a completely different answer, it is probably reasonable to use an average of the laboratory results and also, the somewhat more conservative data from the survey of other operations.

TABLE 5

SUMMARY OF THICKENER DESIGN DATA

CONCENTRATE	DESIGN CRITERIA (tonnes/m^2/day)		
	Laboratory Results	Survey of Mills	Design Practice-General
Lead	6.7 - 15.4	0.3 - 4.3	1.5
Zinc	2.9 - 4.5	0.3 - 3.2	1.5

Notes: a) Design criteria for the laboratory tests are results for pulps without flocculant additions and at flotation pH levels.

b) Settling rates were determined at room temperature: 18-20°C.

Current plans call for the existing 90 and 65 ft. diameter thickeners to be used concurrently for treatment of the zinc concentrates while the 75 and 40 ft. diameter units will be used for lead concentrates. Assuming all of the thickener area is useful, the loading per unit area can be calculated.

TABLE 6

UNIT LOADING ON NEW THICKENER CONFIGURATION

CONCENTRATE	Total Area for Settling ft^2	m^2	DAILY TONNAGE DMT	UNIT LOAD tonne/ft^2/day	tonne/m^2/day
Lead	5672	525	605	0.107	1.15
Zinc	9675	896	1277	0.213	2.3

Note: a) Settling area for lead and zinc compares ore thickener and ore overflow clarifier unit.

Comparing the data in table 5 and table 6, it is apparent that the unit loading on the new thickener configuration will be within the range of survey and design pratice data.

5.4 FILTERING

5.4 FILTERING

The process of filtration in which the thickened concentrate slurries are dewatered to a semi-dry cake appears deceptively simple. The filtration stage is however very complex and incorporates two distinct phases: An appreciation of the factors affecting each of these phases is of some value in filter design work. The phases are as follows:

a) Cake Formation:

In this phase the filter is operated in a mode to produce a cake adhering to the filter medium. The key factors here are pressure drop across the cake, solids content of the pulp, and the viscosity of the fluid. Clearly rate of cake formation will be strongly influenced by the vacuum pump efficiency, cake formation time, mean particle size and to a lesser extent, solids content and temperature of the pulp.

Formation is governed by the following equation:

$$\text{Formation Rate} = \sqrt{\frac{K.W.\Delta P}{\mu\ \alpha\ \theta_f}}$$

Where: K and α are constants for a particular data ref.

W - Pulp solids content

ΔP = Pressure drop across the cake

μ = Fluid viscosity ($\mu \propto \frac{1}{T\ ^{o}C}$)

θf = Formation time.

b) Cake Dewatering Rate:

Cake dewatering rate is regulated by cake porosity which in turn is governed by particle size distribution. No other factor influences dewatering rate to the same degree as particle size given that vacuum and pulp temperature remain relatively constant.

The equation below indicates the relative dependence of dewatering rate on various factors:

$$\text{Dewatering} \propto (d.\ Mr.\ FA)$$

Where: d = Particle size distribution parameters.

Mr = Minimum residual cake moisture for d and other parameters for testwork.

$$FA = \frac{\Delta P}{M} . Q . \theta_d$$

Where : M = Cake output per cycle.

Q = Gas flow through cake.

θd = Dewatering time.

ΔP = Pressure drop across the cake.

Note how the dewatering rate and to an even greater extent the cake formation rate are governed by factors relating to mean particle size or temperature. Clearly then since particle size distribution is governed by metallurgical considerations filter operational efficiency will be determined by pulp temperature and the dewatering time.

5.41 Laboratory Test Data

Laboratory disc filter tests have been performed on samples of lead and zinc concentrates from Anvil and Grum ore bodies over the range of expected conditions. During the tests concentrate grind levels were in the P_{80} = 15-20 micron range. The test results are shown below in tables 7 and 8.

TABLE 7

RESULTS OF LABORATORY FILTERING TESTS - ANVIL ORE

TEST NO.	^oC	SLURRY pH	SLURRY % SOLIDS	FILTER CLOTH	FILTER AID* g/tonne	FILTER TIME (sec) FORM	FILTER TIME (sec) DRY	FILTER TIME (sec) CYCLE	CAKE RATE $t/m^2/h$	CAKE MOISTURE %	FILTRATE RATE $l/m^2/h$	FILTRATE CLARITY g/l
LEAD												
F-1	R.T.	10.8	50	POPR901F	-	30	90	180	.467	12.1	473	2.5
F-2	R.T.	10.8	50	POPR901F	-	30	180	315	.259	12.8	258	2.5
F-4	R.T.	10.8	50	POPR901F	-	60	360	630	.181	11.6	187	2.5
F-5	R.T.	10.8	71	POPR901F	-	15	90	158	.530	13.2	159	1.2
F-10	R.T.	10.8	62	POPR825F	10	30	180	315	.295	11.4	160	0.32
F-11	R.T.	10.9	62	POPR825F	10	30	360	585	.161	12.9	90	0.32
F-12	50	10.9	62	POPR825F	10	30	180	315	.528	11.1	307	0.27
F-13	50	10.9	62	POPR825F	10	30	360	585	.280	11.1	152	0.27
F-14	70	10.9	62	POPR825F	10	30	180	315	.713	7.5	356	0.27
F-15	70	10.9	62	POPR825F	10	30	360	585	.376	9.8	185	0.27
ZINC												
F-16	R.T.	11.0	50	POPR901F	-	30	90	180	.331	14.9	333	3.2
F-17	R.T.	11.0	50	POPR901F	-	30	180	315	.247	14.8	197	3.2
F-18	R.T.	11.0	50	POPR901F	-	30	90	180	.421	14.8	312	3.2
F-19	R.T.	11.0	66	POPR925F	-	15	90	158	.522	15.1	172	1.7
F-20	R.T.	11.0	66	POPR925F	-	15	180	293	.304	13.9	110	1.7
F-21	R.T.	11.0	66	POPR925F	-	15	360	563	.146	14.1	62	1.7
F-22	R.T.	11.0	62	POPR925F	-	15	180	293	.269	14.6	128	0.97
F-23	R.T.	11.0	62	POPR925F	-	15	360	563	.126	14.6	59	0.97
F-24	R.T.	11.0	62	POPR925F	10	15	180	293	.243	12.4	111	0.97
F-25	R.T.	11.0	62	POPR925F	10	15	360	563	.109	11.2	50	0.55
F-26	50	11.0	62	POPR925F	10	15	180	293	.295	11.5	162	0.55
F-27	50	11.0	62	POPR925F	10	15	360	563	.167	11.2	87	0.55
F-28	70	11.0	62	POPR925F	10	15	180	293	.387	11.7	176	0.55
F-29	70	11.0	62	POPR925F	10	15	360	563	.198	12.1	88	0.55

Note: Temperature data as reported from test sheets R.T. approximates 20^oC.

TABLE 8

RESULTS OF LABORATORY FILTERING TESTS - GRUM ORE

TEST NO.	SLURRY % SOLIDS	SLURRY pH	FILTER CLOTH	FILTER AID* g/tonne	FILTER TIME (sec) FORM	FILTER TIME (sec) DRY	FILTER TIME (sec) CYCLE TOTAL	FILTER CAKE RATE $t/m^2/h$	FILTER CAKE MOISTURE %	FILTRATE $l/m^2/h$	CLARITY g/l
LEAD											
L-4-1	55	8.3	POPR853F	No	60	300	540	0.30	14.4	158	N.D.
L-4-2	55	8.3	POPR853F	No	30	300	495	0.19	12.5	102	N.D.
L-4-3	55	8.3	POPR853F	No	60	300	540	0.36	13.6	172	N.D.
L-5-3	55	8.2	POPR851	No	60	600	990	0.18	11.6	90	N.D.
L-5-4	65	8.2	POPR851	No	30	300	495	0.42	13.5	118	N.D.
L-5-5	65	8.2	POPR851	No	30	500	945	0.22	12.4	53	N.D.
ZINC											
Z-5-1	55	11.1	POPR851	-	60	180	360	0.31	18.6	97	N.D.
Z-5-2	55	11.1	POPR851	-	60	300	540	0.18	18.6	65	N.D.
Z-5-3	55	11.1	POPR851	-	60	480	810	0.12	17.6	43	N.D.
Z-5-4	55	11.1	POPR851	-	30	300	495	0.15	18.4	47	N.D.
Z-5-5	55	11.1	POPR851	-	30	480	365	0.18	18.7	66	N.D.

N.D.: Not Determined.

Data reported at R.T. of about 18-20°C.

Of special note in the above tables are the reported cake moisture levels, temperature effects and effect of filter aid on filter performance. Table 9 below summarizes the pertinent data.

TABLE 9

EFFECTS OF PULP TEMPERATURE ON FILTERING RATE

CONCENTRATE	TEMPERATURE °C	CYCLE TIME SEC	FILTERING RATE tonnes/m^2/hour	CAKE MOISTURE %
Faro - Lead	18	315	0.295	11.4
	50	315	0.528	11.1
	70	315	0.713	7.5
Faro - Zinc	18	293	0.243	12.4
	50	293	0.295	11.5
	70	293	0.387	11.7

Notes: a) Data is as reported - room temperature is assumed to be 18°C. See L.R.2202 Vol.IV.

b) Tests carried out with target grind levels, fixed pH values in the anticipated range and 10g/tonne filter aid.

Assuming that the normal range of filter operating conditions will be straddled by the data sets shown in table 7 and 8, then it is reasonable to average all the data from the laboratory tests.

TABLE 10

SUMMARY OF LABORATORY DISC FILTER TEST DATA

CONCENTRATE	PULP CHARACTERISTICS SOLIDS %	pH	TEMP°C	FILTER AID g/tonne	CAKE MOISTURE % H_2O	MEAN FILTERING RATE tonne/ft²/hr	tonne/m²/hr
Lead - Anvil	60-65	10.8	18-70	10	11.3	0.036	0.389
- Grum	55-60	8.2		0	13.0	0.013	0.139
Zinc - Anvil	55-60	11.0	18-70	10	13.0	0.025	0.269
- Grum	55-60	11.1		0	18.4	0.009	0.094

Notes: a) Concentrates reground to P_{80} = 15-20 microns.

b) Vacuum not recorded during tests but believed to be 22 - 24 inches mercury.

c) Data from pilot plant test progress Lakefield Research Reports, L.R.2027 & L.R.2202.

d) Filter aid used. - Drimax 1234.

5.42 Survey Data

A survey of the relevent literature indicated the range of operational capacity per unit area for base metal operations. As expected filtering rate on average tended to be inversely proportional to grind P_{80} of the concentrates. Few of the operations reported a surfeit of filter capacity although several mills practice one or two shift filtration by providing a suitable storage tank capacity for thickened concentrates. A summary of the survey data is shown below in table 11.

TABLE 11

SURVEY OF OPERATING PRACTICE

PLANT		FILTER TYPE	FILTERING RATE tonnes/ft^2/day	FILTERING RATE tonnes/m^2/day	MEAN CONCENTRATE SIZE P_{80} microns
Heath Steel	- Copper	D.O.L.	0.235	2.30	40 - 45 microns
	- Lead	D.O.L.	0.332	3.25	
	- Zinc	D.O.L.	0.405	3.97	
Brunswick	- Copper	D.O.L.	0.162	1.59	30 - 35 microns
	- Lead	D.O.L.	0.273	2.67	
	- Zinc	D.O.L.	0.386	3.78	
	- Bulk	D.O.L.	0.106	1.04	
Sherritt-Fox	- Copper	D.O.L.	0.221	2.17	25 - 30 microns
	- Zinc	D.O.L.	0.265	2.60	
Sherritt-Ruttan	- Copper	Eimco	0.664	6.51	70 - 75 microns
	- Zinc	Eimco	0.202	1.98	
Texas Gulf	- Copper	Eimco	0.273	2.67	15 - 20 microns
	- Lead	Eimco	0.041	0.40	
	- Zinc	Eimco	0.182	1.78	
Sullivan	- Lead	Drum	1.333*	13.03	25 - 30 microns
	- Zinc	Eimco	0.347	3.40	30 - 35 microns
Dufault	- Copper	Eimco	0.114	1.12	N/A
	- Zinc	Eimco	0.032	0.31	N/A
South Bay	- Copper	Eimco	0.144	1.56	40 - 45 microns
	- Zinc	Eimco	0.184	1.98	
Willroy	- Copper	American	0.101	1.09	35 - 40 microns
	- Zinc	American	0.179	1.93	
Brenda	- Copper	Eimco	0.208	2.25	30 - 35 microns
Orchan	- Copper	D.O.L.	0.073	0.79	N/A
	- Zinc	D.O.L.	0.542	5.85	N/A

Notes:
a) Unless otherwise noted all filters are disc type.
b) Filtering rate calculations based on continuous operation and on reported concentrate output.
c) N/A. Data not reported in reference text.

Besides providing useful information about filtering rates, there came to light several pertinent operational points which are of interest.

i) The use of filter aids is increasing. Almost all operations report that dosage rate and dilution are critical to effectiveness, although there does not seem to be a preference for one particular manufacturer.

ii) There does not appear to be a preferred filter manufacturer, for mineral concentrate dewatering, although the majority of plants have installed disc type units from Dorr Oliver or Eimco.

iii) Filter bag type varies widely, although synthetic materials (polypropylene) are sometimes reported to be preferred due to cost advantage.

iv) Vacuum was reported to be in excess of 20 inches of mercury in most installations. Fine grind applications frequently reported 22-23 inches of mercury at the filter head.

5.43 Sizing Calculations

Using the average range of laboratory data and comparing this data with the relevant information from the survey, indicated that the laboratory data yielded optimistic results.

TABLE 12

FILTER CAPACITY DATA

Concentrate	FILTERING RATE tonnes/m^2/day Laboratory	Survey
Lead	6.3	3.0
Zinc	4.4	2.7

Notes: a) Laboratory data from table 10.
b) Survey data from relevant data in table 11.

Assuming that the survey data is more likely to be correct than the laboratory results and recognizing the extremely fine grind P_{80} values for the Faro area concentrates, it was deemed prudent to install sufficient filter capacity to meet the unit filter area predicted by the survey data.

To achieve this objective five new filters will be installed replacing the existing units for a total of 582 m^2 of theoretical filter cloth area. The calculated concentrate load per unit area of filter would then be as follows:

TABLE 13

DESIGN FILTERING RATE

CONCENTRATE	FILTER AREA ft^2	FILTER AREA m^2	TONNAGE/DAY tonnes	DESIGN FILTERING RATE tonnes/ft^2/day	DESIGN FILTERING RATE tonnes/m^2/day
Lead	3456	320	605	0.175	1.89
Zinc	5530	512	1277	0.231	2.49

Notes: a) Filter area is the disc surface area with no allowance for head shaft or depth of immersion in filter boot.

b.) Calculations based on using 10.5' diameter disc, 10 per filter unit, except the unit feeding the oil fired dryer which will have 12 discs.

5.5 DRYING

5.5 DRYING

As part of the dewatering section, upgrading, the dryer section will be extensively modified. The removal of the bulk production system will permit one of the loadout belts to be removed and incidentally should significantly reduce dust production from the swing conveyor. The dryers will have to be repositioned in order to reach the required belts, but the trouble involved in repositioning will be repaid by developing separate lead and zinc filter dryer areas. The decrease in internal concentrate contamination should be most evident as a result of this change.

5.51 Thermal Capacity Calculations

The dryer design capacity calculations were performed using a standard formula and employing corrections for combustion chamber losses, radiation losses and other miscellaneous thermal effects. The data indicates that the existing dryers will, if correctly operated, produce an acceptable moisture level in the concentrates delivered to the loadout design conditions.

It is worth emphasizing that the dryer capacity calculations have been based on coal of low thermal value similar to that experienced during the 1978-1979 operation. If by improved selective mining and by cleaning of the coal prior to combustion, the thermal value could be improved some increase in dryer treatment capacity could be expected.

TABLE 14

DRYER CAPACITY CALCULATIONS

PRODUCT	QUANTITY tonnes/day	DRYER USED	AVAILABLE CAPACITY BTHU x 10^6	DESIGN REQUIREMENTS BTHU x 10^6
Lead Conc.	605	2x5'dia.x40'	20.0	16.3
Zinc Conc.	1277	1x6'dia.x44' 2x5'dia.x40'	35.0	34.7

Notes: a) Design requirements are for moisture reduction to 4.5% at the dryer discharge.

b) Thermal value of coal used in sizing/ capacity calculations 8400 BTHU/lb.

5.52 LEAD DRYERS

In these dryer calculations basic criteria are tested first, heat requirements calculated, losses determined and finally the fuel quantity calculated.

Lead Rotary Dryer Capacity Calculations

Dryer #3 and 4, 5' Dia. x 40' long, Coal-Fired

TABLE 15

CALCULATION BASE DATA

(per dryer)

Material to be dried		Lead Concentrate	
Maximum dryer feed		13.9	STPH
Moisture in feed		12	%
Moisture in product (max. allowable)		4.5	%
Moisture removed		1.24	STPH
Air temperature entering dryer from combustion chamber	T_1	1200	^oF
Gas temperature leaving dryer	T_2	300	^oF
Material entering dryer	T_3	60	^oF
Material leaving dryer	T_4	204	^oF
Ambient air temperature	T_5	70	^oF (ave)
Specific heat of concentrate (sh)		0.051	BTU/lb ^oF
Specific heat of air		0.24	BTU/lb ^oF
Specific heat of water		1.00	BTU/lb ^oF
Enthalpy of water at T_3		28.06	BTU/lb
Enthalpy of water at 204^oF (T_4)		180.7	BTU/lb
Latent heat of fusion of ice		144.0	BTU/lb
Enthalpy required to evaporate water at 212^oF		970.3	BTU/lb
Enthalpy of super heated steam at T_2		1192.8	BTU/lb
Dryer radiation loss		400	BTU/hr/ft^2
Combustion chamber radiation loss		2000	BTU/hr/ft^2
Hours per day of operation		24.0	Hrs/day
Gross heating value of coal used		8400	BTU/lb
Enthalpy of water at $T_5$$^oF$		28.1	BTU/lb
Elevation above sea level		3860	Ft

Note: a) All calculations shown below are in imperial units only. Where units are not shown, imperial units should be assumed.

1. Moisture Driven Off

$$\frac{379 \text{ TPD}}{24 \text{ hrs/day}} = \text{Wet Feed (TPH)} = 15.80 \text{ TPH} = A$$

$$\text{Dry Feed} = 15.80 \text{ TPH} \times \frac{(100 - 12)}{100} = 13.9 \text{ TPH} = B$$

$$\text{Product (4.5\% Moisture)} = 13.9 \text{ TPH} \times \frac{(100 \times 4.5)}{100} = 14.56 \text{ TPH} = C$$

Moisture Driven Off (A-C) 15.80 - 14.56 = 1.24 TPH = D

Moisture in Product (C-D) 14.56 - 13.9 = 0.66 TPH = E

2. Heat Requirements

2.1 Heat loss in product discharging at 204^{O}F = T_4

$$13.9 \text{ TPH} \times \frac{2000 \text{ lb}}{T} \times 0.51 \frac{\text{BTU}}{\text{lb}/^{O}\text{F}} (204-60)^{O}\text{F} \qquad = 204{,}163 \text{ BTU/hr}$$

2.2 Heat loss in water discharging at 204^{O}F

$$0.66 \text{ TPH} \times \frac{2000 \text{ lb}}{T} \times (204-60)^{O}\text{F} \qquad = 190{,}080 \text{ BTU/hr}$$

2.3 Heat loss in elevating excess water to boiling point - T_4

$$1.24 \text{ TPH} \times \frac{2000 \text{ lb}}{T} (210-60) \text{ BTU/hr} \qquad = 372{,}000 \text{ BTU/hr}$$

2.4 Heat loss in evaporating water

$$1.24 \text{ TPH} \times \frac{2000 \text{ lb}}{T} \times 970.3 \text{ BTU/lb} \qquad = 2{,}406{,}344 \text{ BTU/hr}$$

2.5 Heat loss in super heating water to 300^{O}F = T_2

$$1.24 \text{ TPH} \times \frac{2000 \text{ lb}}{T} (1102.8 - 1150.4) \text{ BTU/lb} \qquad = 105{,}152 \text{ BTU/hr}$$

(H300 + H212 from steam tables)

2.6 Heat loss by radiation from dryer plus combustion changer

Dryer losses

$$400 (\pi \times 5' \times 40') + \frac{2 \pi 5^2}{2} \qquad = 267{,}035 \text{ BTU/hr}$$

Combustion Chamber losses

2000(7.2 x 15.2 x 2)+(15.2 x 17.8 x 2)+(7.2x17.8x2)=2,032,640 BTU/hr

2.7 TOTAL HEAT REQUIRED FROM GASES = 5,577,414 BTU/hr = F

3. Air Requirements to Transmit Heat

3.1 Amount of air required

$$\frac{5,577,414 \text{ BTU/hr}}{1200\text{-}300^{\circ}\text{F} \times 0.24 \dfrac{\text{BTU}}{\text{lb}/^{\circ}\text{F}}} = 25,821 \text{ lb/hr} = G$$

3.2 Heat loss in exhaust gases

$$\frac{25,821 \text{ lb}}{\text{hr}} \times (300-70)^{\circ}\text{F} \times 0.24 \frac{\text{BTU}}{\text{lb}^{\circ}\text{F}} = 1,425,319 \text{ BTU/hr} = H$$

3.3 Heat loss in moisture in exhaust gases

Assume 70% relative humidity

$$\begin{array}{l}\text{Moisture content} \\ \text{(from)} \\ \text{Psychometic chart}\end{array} = \frac{0.011 \text{ lb Moisture}}{\text{lb dry air}} \times 25,821 \text{ lb/hr} = 284 \text{ lb/hr} = I$$

$$\frac{284 \text{ lb}}{\text{hr}} \times 1192.8 - 28.1 \frac{\text{BTU}}{\text{lb}} = 220,775 \text{ BTU/hr} = I$$

TOTAL = 7,333,508 BTU/hr = K

4. Net Heating Value of Fuel

4.1 Coal - gross heating value = 8400 BTU/lb

Assay	As Received	Dry Basis
% moisture	3.56	--
% ash	34.39	35.66
% volatile	24.28	25.18
% fixed carbon	37.77	39.16
	100.00	100.00
% S	0.47	0.49

For design purposes - Coal quality is as follows:

Use % moisture	4%
% ash	35%
BTU/lb	8,400

Heat loss from moisture in exhaust gases:

.04 lb H_2O x (300-70) = 9.2 BTU/lb fuel

.04 lb H_2O x 970.3 BTU/lb = 38.81 BTU/lb fuel

.04 lb H_2O x (1192.8 - 28.1) = 46.59 BTU/lb fuel

Heat loss for equivalent amount of air

.04 x .24 x (300-70) = 2.21 BTU/lb fuel

Fuel requirement to generate

heat required from gases = $\frac{F}{M} = \frac{5,577,414}{8,400}$ = 664.0 lb fuel

Net heat loss for water from combustion = 96.81 BTU/lb fuel = L
Gross heating value of coal = 8,400 BTU/lb fuel = M
Net heating value (M-L) = 8,303.2 BTU/lb fuel = N

4.2 Combustion Losses

1 lb fuel = 8,303.2 BTU/lb=N

$c + O_2 \longrightarrow CO_2$

Air requirements:

$$\frac{\text{Moles air}}{\text{Moles fuel}} = \frac{1 \text{ mole } O_2}{\text{mole fuel}} \times \frac{1 \text{ moles air}}{\text{moles } O_2} \qquad 4.76\ \frac{\text{moles air}}{\text{moles fuel}}$$

Assume 20% excess air = 4.76 x 1.2 = 5.71 $\frac{\text{moles air}}{\text{moles fuel}}$

$$\frac{\text{lb air}}{\text{lb fuel}} = \frac{5.7 \times 28.95}{12} = 13.77\ \frac{\text{lb air}}{\text{lb fuel}}$$

Heat loss in fuel

1.0 (300 - 60) .24 = 57.6 BTU/lb fuel

Heat loss in air

13.77 (300 - 60) x .24 = 793.2 BTU/lb fuel

TOTAL = 850.8 BTU/lb fuel=0

Net heating value of fuel = N-0 = 7452.4 BTU/lb

Fuel Availability = $\frac{7452.4}{8,400}$ x 100% = 88.7 %@T_2°F =P

4.3 Combustion Losses

Gross heat requirements

100 x $\frac{7,333,508 \text{ BTU/hr}}{88.7}$ = 8,267,766 BTU/hr = Q

4.4 Heat loss in burning fuel

(Q-K) BTU/hr = 934,258 BTU/hr

5. Heat Distribution

		BTU/hr
2.1	Heat loss in product discharging at 204^{o}F	204,163
2.2	Heat loss in elevating water to 204^{o}F	190,080
2.3	Heat loss in elevating excess water to boiling point	372,000
2.4	Heat loss in evaporating water	2,406,344
2.5	Heat loss in superheating water to T_2^{o}F	105,152
2.6	Heat loss by radiation from dryer	2,299,675
3.2	Heat loss in exhaust gases	1,425,319
3.3	Heat loss in water due to relative humidity of air	330,775
4.4	Heat loss in burning fuel	934,258
	Total Heat Requirements	8,267,766 BTU/hr

6. Fuel Requirements

$$\frac{\text{BTU}}{\text{BTU/lb fuel}} = \frac{8,267,766}{8,400} \qquad \underline{984 \text{ lb fuel/hr} = R}$$

5.53 ZINC DRYERS

Zinc Rotary Dryer Capacity Calculations

Dryer #5, 6' Dia. x 44' long, Oil-Fired

TABLE 16

CALCULATION BASE DATA

Material to be dried		Zinc Concentrate	
Maximum dryer feed		22.05	STPH
Moisture in feed		13	%
Moisture in product (max. allowable)		4.5	%
Moisture removed		2.25	STPH
Air temperature entering dryer from combustion chamber	T_1	1200	^{o}F
Gas temperature leaving dryer	T_2	300	^{o}F
Material entering dryer	T_3	60	^{o}F
Material leaving dryer	T_4	204	^{o}F
Ambient air temperature	T_5	70	^{o}F (ave)
Specific heat of concentrate (sh)		0.128	BTU/lb ^{o}F
Specific heat of air		0.24	BTU/lb ^{o}F
Specific heat of water		1.00	BTU/lb ^{o}F
Enthalpy of water at T_3		28.06	8TU/lb
Enthalpy of water at 204^{o}F (T_4)		180.7	BTU/lb
Latent heat of fusion of ice		144.0	BTU/lb
Enthalpy required to evaporate water at 212^{o}F		970.3	BTU/lb
Enthalpy of super heated steam at T_2		1192.8	BTU/lb
Dryer radiation loss		400	BTU/hr/ft^2
Combustion chamber radiation loss		2000	BTU/hr/ft^2
Hours per day of operation		24.0	Hrs/day
Gross heating value of No.2 fule oil		17000	BTU/lb (140,000BTU/gal)
Gross heating value of coal used		8400	8TU/lb
Enthalpy of water at T_5^{o}F		28.1	BTU/lb
Elevation above sea level		3860	Ft

1. Moisture Driven Off

$$\frac{\text{TPD}}{\text{Hrs. /Day}} = \text{TPH Feed} = 25.34 \text{ TPH} \quad \text{A}$$

$$\text{Dry Feed} = 25.34 \text{ TPH} \times \frac{(100 - 13)}{100} = 22.05 \text{ TPH} \quad \text{B}$$

Product (4.5 % Moisture)

$$= 22.05 \text{ TPH} \times \frac{(100 - 4.5)}{100} = 23.09 \text{ TPH} = \text{C}$$

Moisture Driven Off (A-C) = 2.25 TPH = D

Moisture in Product (C-B) = 1.04 TPH = E

2. Heat Requirements

2.1 Heat loss in product dischargin at 212^{O}F = T_4

$$22.05 \text{ TPH} \times \frac{2000 \text{ lb}}{\text{T}} \times 0.128 \frac{\text{BTU}}{\text{lb}/^{O}\text{F}} (204-60)^{O}\text{F} = 812,851 \text{ BTU/hr}$$

2.2 Heat loss in elevating water in product

$$1.04 \text{ TPH} \times \frac{2000 \text{ lb}}{\text{T}} \times (204 - 60) = 299,520 \text{ BTU/hr}$$

2.3 Heat loss in elevating excess water to boiling point - T_4

$$2.25 \text{ TPH} \times \frac{2000 \text{ lb}}{\text{T}} (204 - 60) \text{ BTU/lb} = 648,000 \text{ BTU/hr}$$

2.4 Heat loss in evaporating water

$$2.25 \text{ TPH} \times \frac{2000 \text{ lb}}{\text{T}} \times 970.3 \text{ BTU/lb} = 4,366,350 \text{ BTU/hr}$$

2.5 Heat loss in super heating water to 300^{O}F = T_2

$$2.25 \text{ TPH} \times \frac{2000 \text{ lb}}{\text{T}} (1192.8 - 1150.4) \text{ BTU/lb} = 190,800 \text{ BTU/hr}$$

H300 + H204 from steam tables

2.6 Heat loss by radiation from dryer + combustion chamber

Dryer losses

$$400 \text{ BTU/ft}^2 (\pi \times 6 \times 44) = \frac{2 \times \pi\ 36)}{4} = 354,372 \text{ BTU/hr}$$

Combustion Chamber losses

$$2000 \text{ BTU/ft}^2\ \pi \times 5.5 \times 13.8 + \pi \times 4.25 \times 10 = 743,929 \text{ BTU/hr}$$

2.7 <u>Total Heat required from gasses = 7,415,822 BTU /hr</u> = F.

3 Air Requirements to transmit Heat

3.1 Amount of air required

$$\frac{7{,}415{,}822\ \text{BTU/hr}}{1200-300^{O}\text{F} \times 0.24\ \frac{\text{BTU}}{\text{lb}/^{O}\text{F}}} = 34{,}333\ \text{lb/hr} \qquad G$$

3.2 Heat loss in exhaust gases

$$\frac{34{,}333\ \text{lb}}{\text{Hr}} \times (300 - 70)^{O}\text{F} \times 0.24\ \frac{\text{BTU}}{\text{lb}\ ^{O}\text{F}} = 1{,}895{,}182\ \text{BTU/hr} = H$$

3.3 Heat loss in moisture in exhaust gases

Assume 70% relative humidty

Mositure content (from) Psychrometic chart $= \frac{0.011\ \text{lb Moisture}}{\text{lb dry air}} \times 34{,}333\ \text{lb/hr} = 378\ \text{lb/hr} \qquad I$

$$\frac{378\ \text{lb}}{\text{hr}} \times 1192.8 - 28.1\ \frac{\text{BTU}}{\text{lb}} = 440{,}257\ \text{BTU/hr} = J$$

TOTAL = 9,751,261 BTU/hr = K

4. Net Heating Value of Fuel:

4.1 Net Heating Value of No. 2 Fuel Oil

From Published stack emission data
15% excess air (by volume)

CO_2 - 12.9 % - 1390 SCF - 160 lb/million BTU
CO - 0.002% - .22 SCF - 0.016 lb/million BTU
O_2 - 2.9 % - 310 SCF - 25 lb/million BTU
N_2 - 84.2 % - 9080 SCF - .670 lb/million BTU
H_2O - - - 1420 SCF - 65 lb/million BTU

4.2 Heat loss from moisture in exhuast gases

$$\frac{7,415,822}{1,000,000} \times 65 \text{ lb water} \times (204 - 70)^{\circ}\text{F} = 64,592 \text{ BTU/lb Fuel}$$

$$\frac{7,415,822}{1,000,000} \times 65 \text{ lb water} \times 970.3 \text{ BTU} = 467,712 \text{ BTU/lb Fuel}$$

$$\frac{7,415,822}{1,000,000} \times 65 \text{ lb water} \times (1102.8 - 28.1)\text{BTU/lb} = 561,419 \text{ BTU/lb Fuel}$$

Heat loss for an equivalent amount of air

$$\frac{7,415,822}{1,000,000} \times 65 \text{ lb} \times 0.24 \text{ BTU } (300 - 70)^{\circ}\text{F} = 26,608 \text{ BTU/lb Fuel}$$

Fuel Requirement to Generate Heat Required from gasses $= \dfrac{7,415,822}{17,000}$ = 436 lb Fuel

Net heat loss for water from combustion $= \dfrac{1,120,331 \text{ BTU/lb Fuel}}{436}$ = 2,570 = L

Gross Heating value of No. 2 fuel = 17,000 BTU/lb Fuel = M

Net Heating value (M-L) = 14,430 BTU/lb Fuel = N

4. Net Heating Value of Fuel:(con't)

4.2 Combustion Losses (con't)

$$C + O_2 \longrightarrow CO_2$$

Air Requirements:

$$\frac{\text{Moles air}}{\text{Moles fuel}} = 1\ \frac{\text{mole } O_2}{\text{mole fuel}} \times \frac{1}{0.21}\ \frac{\text{moles air}}{\text{moles } O_2} = 4.76\ \frac{\text{moles of air}}{\text{moles of fuel}}$$

$$\text{Assume 20\% excess air} = 4.76 \times 1.2 = 5.71\ \frac{\text{moles air}}{\text{moles fuel}}$$

$$\frac{\text{lb air}}{\text{lb fuel}} = \frac{5.7 \times 28.95}{12} = 13.77\ \frac{\text{lb air}}{\text{lb fuel}}$$

Heat loss in fuel

1.0 (300 - 60) .24 = 57.6 BTU/lb fuel

Heat loss in air

13.77 (300 - 60) x .24 = 793.2 BTU/lb fuel

TOTAL = 850.8 BTU/lb fuel = O

Net heating value of fuel = N-O = 7452.4 BTU/lb

$$\text{Fuel Availability} = \frac{7452.4}{8,400} \times 100\% = 88.7\ \%\ @\ T_2\,^{o}F = P$$

4.3 Combustion Losses (con't)

Gross heat requirements

$$100 \times \frac{9,745,129}{88.7}\ \text{BTU/hr} = 10,986,617\ \text{BTU/hr} = Q$$

4.4 Heat Loss in Burning Fuel

(Q-K) BTU/hr = 1,241,488 BTU/hr

5. Heat Distribution:

		BTU/hr
2.1	Heat loss in product discharging at $204^{\circ}F$	674,611
2.2	Heat loss in elevating water to $204^{\circ}F$	247,680
2.3	Heat loss in elevating excess water to boiling point	561,000
2.4	Heat loss in evaporating water	3,628,922
2.5	Heat loss in superheating water to T_2 $^{\circ}F$	158,576
2.6	Heat loss by radiation from dryer	2,299,675
3.2	Heat loss in exhaust gases	1,894,149
3.3	Heat loss in water due to relative humidity of air	439,092
4.4	Heat loss in burning fuel	1,241,488
	Total Heat Requiremnent =	11,145,493 BTU/hr

6. Fuel Requirements:

$$\frac{\text{BTU/hr}}{\text{BTU/lb fuel}} = \frac{11{,}145{,}493}{8{,}400} = \underline{\underline{1{,}327 \text{ lb fuel/hr} = R}}$$

Zinc Rotary Dryer Capacity Calculations

Dryers #1 and 2, 5' Dia. x 40' long, Coal-Fired

TABLE 17

CALCULATION BASE DATA

(per dryer)

Material to be dried		Zinc Concentrate	
Maximum dryer feed		18.3	STPH
Moisture in feed		13	%
Moisture in product (max. allowable)		4.5	%
Moisture removed		1.87	STPH
Air temperature entering dryer from combustion chamber	T_1	1200	^oF
Gas temperature leaving dryer	T_2	300	^oF
Material entering dryer	T_3	60	^oF
Material leaving dryer	T_4	204	^oF
Ambient air temperature	T_5	70	^oF (ave)
Specific heat of concentrate (sh)		0.128	BTU/lb ^oF
Specific heat of air		0.24	BTU/lb ^oF
Specific heat of water		1.00	BTU/lb ^oF
Enthalpy of water at T_3		28.06	BTU/lb
Enthalpy of water at 204^oF (T_4)		180.7	BTU/lb
Latent heat of fusion of ice		144.0	BTU/lb
Enthalpy required to evaporate water at 212^oF		970.3	BTU/lb
Enthalpy of super heated steam at T_2		1192.8	BTU/lb
Dryer radiation loss		400	BTU/hr/ft^2
Combustion chamber radiation loss		2000	BTU/hr/ft^2
Hours per day of operation		24.0	Hrs/day
Gross heating value of coal used		8400	BTU/lb
Enthalpy of water at T_5 ^oF		28.1	BTU/lb
Elevation above sea level		3860	Ft

1. Moisture Driven Off

439.2 TPD Product at 4.5% Moisture Required

Feed in at 13% Moisture $\frac{439.2}{0.87}$ = 504.8 TPD (wet)

$\frac{504.8 \text{ TPD}}{24 \text{ hrs PD}}$ = 21.03 TPH Feed = 21.03 TPH A

Dry Feed = 21.03 TPH x $\frac{(100 - 13)}{100}$ = 18.3 TPH B

Produce (4.5% Moisture = 18.3 TPH ÷ $\frac{(100 - 4.5)}{100}$ = 19.16 TPH = C

Moisture Driven off (A-C) 21.03 - 19.16 = 1.87 TPH = D

Moisture in Product (C-B) 19.16 - 18.3 = 0.86 TPH = E

2. Heat Requirements

2.1 Heat loss in product discharging at 204°F = T_4

$$18.3 \text{ TPH} \times \frac{2000 \text{ lb}}{\text{T}} \times 0.128 \frac{\text{BTU}}{\text{lb}/^{\text{o}}\text{F}} (204 - 60)^{\text{o}}\text{F} = 674{,}611 \text{ BTU/hr}$$

2.2 Heat loss in water discharging at 204°F

$$0.86 \text{ TPH} \times \frac{2000 \text{ lb}}{\text{T}} \times (204 - 60)^{\text{o}}\text{F} = 247{,}680 \text{ BTU/hr}$$

2.3 Heat loss in elevating excess water to boiling point - T_4

$$1.87 \text{ TPH} \times \frac{2000 \text{ lb}}{\text{T}} (210 - 60) \text{ BTU/lb} = 561{,}000 \text{ BTU/hr}$$

2.4 Heat loss in evaporating water

$$1.87 \text{ TPH} \times \frac{2000 \text{ lb}}{\text{T}} \times 970.3 \text{ BTU/lb} = 3{,}628{,}922 \text{ BTU/hr}$$

2.5 Heat loss in super heating water to 300°F = T_2

$$1.87 \text{ TPH} \times \frac{2000 \text{ lb}}{\text{T}} (1192.8 - 1150.4)\text{BTU/lb} = 158{,}576 \text{ BTU/hr}$$

(H300 + H212 from steam tables)

2.6 Heat loss by radiation from dryer + comustion chamber

Dryer losses

$$400 \; (\pi \times 5' \times 40') + \frac{2\pi 5^2}{2} = 267{,}035 \text{ BTU/hr}$$

Combustion Chamber losses

$$2000 \; (7.2 \times 15.2)+(15.2 \times 17.8 \times 2)+(7.2 \times 17.8 \times 2) = 2{,}032{,}640 \text{ BTU/hr}$$

2.7 Total Heat required from gases = 7,411,888 BTU/hr = F

3. Air Requirements to transmit Heat:

3.1 Amount of air required

$$\frac{7{,}411{,}888 \text{ BTU/hr}}{1200\text{-}300^{\circ}\text{F} \times 0.24 \ \frac{\text{BTU}}{\text{lb}/^{\circ}\text{F}}} = 34{,}314 \text{ lb/hr} = \text{G}$$

3.2 Heat loss in exhaust gases

$$\frac{34{,}315 \text{ lb}}{\text{Hr}} \times (300\text{-}70)^{\circ}\text{F} \times 0.24 \ \frac{\text{BTU}}{\text{lb}^{\circ}\text{F}} = 1{,}894{,}149 \text{ BTU/hr} = \text{H}$$

3.3 Heat loss in moisture in exhaust gases

Assume 70% relative humidity

$$\begin{array}{c}\text{Moisture content} \\ \text{(from)} \\ \text{Psychrometic chart}\end{array} = \frac{0.011 \text{ lb Moisture}}{\text{lb dry air}} \times 34{,}314 \text{ lb/hr} = 377 \text{ lb/hr} = \text{I}$$

$$\frac{377 \text{ lb}}{\text{hr}} \times 1192.8 - 28.1 \ \frac{\text{BTU}}{\text{lb}} = 439{,}092 \text{ BTU/hr} = \text{J}$$

TOTAL = 9,745,129 BTU/hr = K

4. Net Heating Value of Fuel

4.1 Coal - gross heating value = 8,400 BTU/lb

	As Received	Dry Basis
% moisture	3.56	---
% ash	34.39	35.66
% volatile	24.28	25.18
% fixed carbon	37.77	39.16
	100.00	100.00
% S	0.47	0.49
BTU/kg	17,826	18,484

For design purposes -

use % moisture	4%
% ash	35%
BTU/lb	8,400

Heat loss from moisture in exhaust gases:

.04 lb H_2O x (300 - 70) = 9.2 BTU/lb fuel

.04 lb H_2O x 970.3 BTU/lb = 38.81 BTU/lb fuel

.04 lb H_2O x (1192.8 - 28.1) = 46.59 BTU/lb fuel

Heat loss for equivalent amount of air

.04 x .24 x (300 - 70) = 2.21 BTU/lb fuel

Fuel Requirement to Generate Heat Required from gases = $\frac{F}{M} = \frac{7,411,888}{8,400}$ = 882.4 lb fuel

Net heat loss for water from combustion = 96.81 BTU/lb fuel = L

Gross Heating value of coal = 8,400 BTU/lb fuel = M

Net Heating value (M-L) = 8,303.2 BTU/lb fuel = N

4.2 Combustion Losses

1 lb fuel = 8,303.2 BTU/lb = N

4. Net Heating Value of Fuel (con't)

4.3 Combustion Losses

1 lb Fuel = 14,430 BTU/lb = N

Air Requirements

Wt. O_2 required per lb of fuel burned:

C; $0.873 \times \frac{32}{12}$ = 2.328 lb

H_2; $0.1165 \times \frac{16}{2}$ = .932 lb

O_2; 0.0027 = -.0027

N_2; inert = -------

Ash; inert = -------

S; $0.0022 \times \frac{32}{32}$ = $\underline{.0022}$

3.260 lb

Therefore Wt. air required

$3.26 \times \frac{(0.768)}{(0.232)}$ = 10.79 lb N_2

Wt. Air = 10.79 + 3.26 = 14.05 lb air/lb fuel

Assume 20% excess air

14.05 x 1.2 = 16.86 lb air/lb fuel

Heat loss in fuel

1.0 (300 - 60) .5 = 120 BTU/lb Fuel

Heat loss in air

16.86 (300 - 60) x .24 = $\underline{971}$ BTU/lb Fuel

TOTAL = $\underline{\underline{1,091}}$ BTU/lb Fuel 0

4.3 Combustion Losses (con't)

Net heating value of fuel = N-O = 13,339 BTU/lb

Fuel Availability = $\frac{13,339}{17,000}$ x 100% = 78.5% @ T_2 °F (P)

Gross heat requirements

$\frac{100 \times 9,751,261 \text{ BTU/hr}}{78.5\%}$ = 12,421,989 BTU/hr (Q)

4.4 Heat Loss in Burning Fuel

(Q-K) BTU/hr = 2,670,728 BTU/hr

5. Heat Distribution

		BTU/hr
2.1	Heat loss in product discharging at 212°F	812,851
2.2	Heat loss in elevating water in product	299,520
2.3	Heat loss in elevating excess water to boiling point	648,000
2.4	Heat loss in evaporating water	4,366,350
2.5	Heat loss in superheating water to T_2°F	190,800
2.6	Heat loss by radiation from dryer and combustion chamber	1,098,301
3.2	Heat loss in exhaust gases	1,895,182
3.3	Heat loss in water due to relative humidity of air	440,257
4.4	Heat loss in burning fuel	2,670,728
	Total Heat Requirement =	12,421,989 BTU//hr

6. Fuel Requirements:

$$\frac{12{,}421{,}989 \text{ BTU/hr}}{17{,}000 \text{ BTU/lb fuel}} = \underline{\underline{731 \text{ lb fuel/hr} = R}}$$

6
POWER

POWER

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PAGE NO.

6.0 SUMMARY

6.0 SUMMARY

Calculations based on most recent estimates of Cyprus Anvil power requirements show that the modified mine-mill complex will require a peak demand of about 25MW and a projected annual total consumption of 150-160 GWH. Compared with present peak demand of 16MW and energy consumption of 100 GWH per annum.

Studies of the N.C.P.C. system capacity showed that the existing system could supply the projected demand although some additional generating capacity would be needed to maintain the system reserve considered necessary for a public utility. The most economic method of supplying the required reserve was to install two 3MW gas turbine generators at the Faro plant of N.C.P.C. and to provide 2.5MW of diesel capacity at the minesite. This latter unit would also provide reliable emergency power at the minesite during power outages.

Additional work considered alternative sources of power, including thermal coal fired generators and new hydro electrical schemes, to supply future development of the Vangorda Plateau.

6.1 INTRODUCTION

6.1 INTRODUCTION

By using the equipment register of the modified mill and estimating the probable additional mine equipment size it became possible to forecast the probable connected electrical load. Further by employing a load factor of about 0.60 the peak demand and the annual energy consumption were calculated for the operation. The calculations showed that a peak demand of 25MW and an annual load of about 155 GWH could be anticipated.

A study by Monenco Ltd. of Calgary of the N.C.P.C. generating system capacity indicated that, by using a combination of thermally generated and existing hydro-electric capacity, the projected power demands on the system could be satisfied. However, the same studies showed that the essential system reserve capacity, would be depleted. Thus a system generating capacity would have to be increased by a total of about 9MW of "reserve only" capacity.

Following a lengthy analysis of data it was decided to install two 3MW gas turbine units at the N.C.P.C. Faro plant and one 2.5MW diesel unit at the minesite to provide the required reserve until the WH #4 generator is commissioned in mid 1983. The diesel unit will act both as a contributor to reserve capacity and as an emergency power generator. One of the two gas turbine units is the property of N.C.P.C. and will be moved to Faro during the spring of 1981.

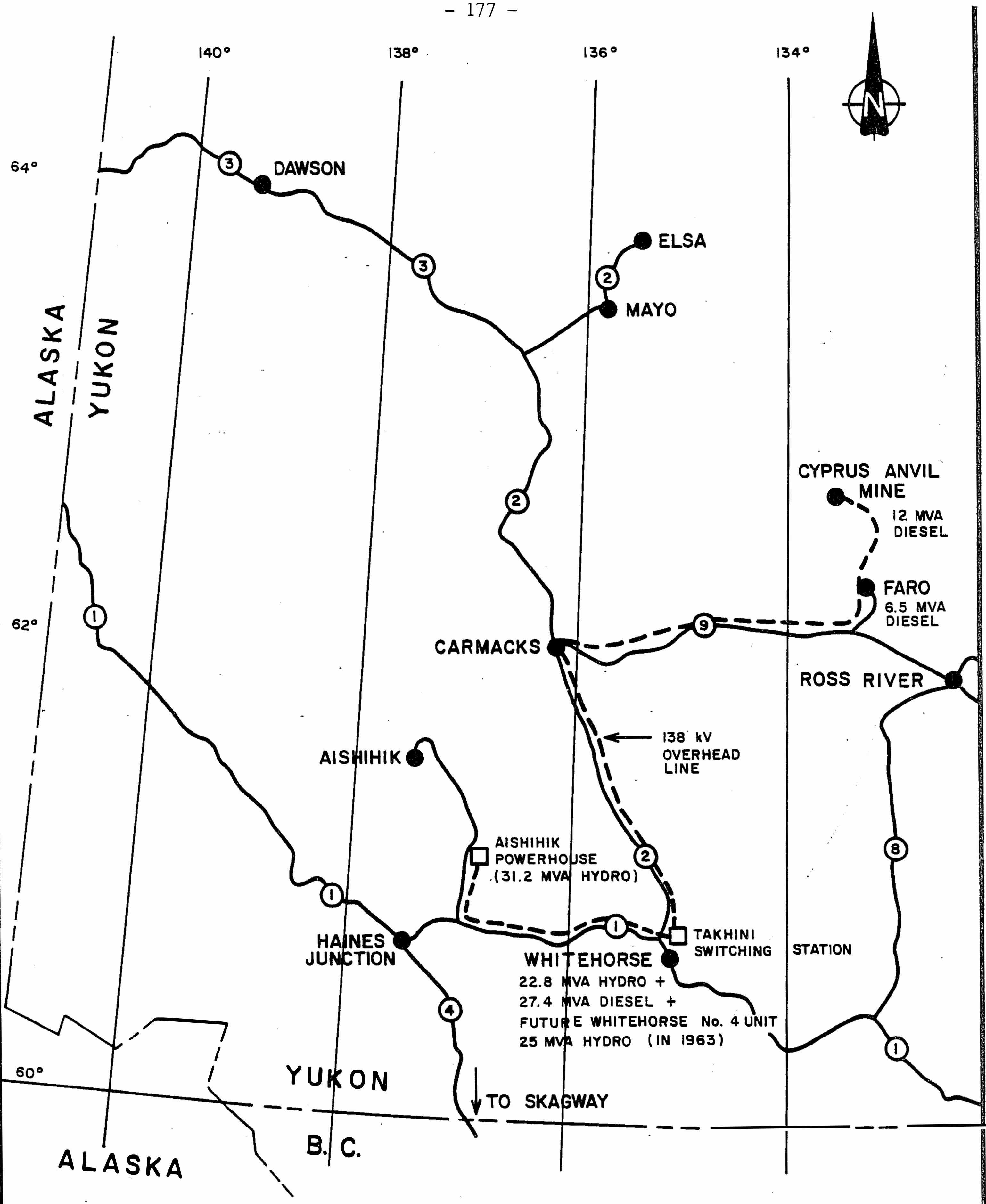


FIG. 6-1

NCPC YUKON POWER SYSTEM

Drawn by: C. L. C.

6.2 THE N.C.P.C. GENERATING SYSTEM

6.2 THE N.C.P.C. GENERATING SYSTEM

6.21 Generating Capacity

The N.C.P.C. electrical distribution grid provides power to Faro and the minesite, and also services Whitehorse, Carmacks, Ross River, and several of smaller communities. (Figure 6-1). The total installed capacity of the existing system is 77.7 MW. Following the installation of the fourth hydroelectric turbine at the Whitehorse Rapids plant in 1983, the system capabilities will be increased to 97.7 MW. The table 1 below summarizes the N.C.P.C. generating capabilities and includes the small diesel unit located at Faro, installed primarily to provide emergency domestic power to the communities of Faro and Ross River in the event of a power outage.

TABLE 1

N.C.P.C. GENERATING CAPACITY

SOURCE		PRESENT CAPACITY (MW)	POST MID 1983 CAPACITY (MW)
Hydro:	Aishihik	31.0	31.0
	Whitehorse Rapids	20.0	40.0
	Sub-Total	51.0	71.0
Diesel:	Faro Plant	5.1	5.1
	Whitehorse Rapids	21.6	21.6
	Sub-Total	26.7	26.7
TOTAL		77.1	97.7

Notes:
a) Before mid 1980, the Faro plant output was limited to about 4.2 MW due to cooling problems. As expected this limitation was corrected and the unit achieves its full nameplate rating of 5.1 MW in 1981.

b) The optimum operating level of Aishihik is 12.5 MW per generator. During the summer the energy output of this plant is limited in order to store water for the winter.

c) The new hydro unit at Whitehorse could suffer limitations during the winter months.

6.22 Reserve Capacity

Appropriate utility system reserve capacities are normally computed by adding five percent of the anticipated peak demand to the capacity of the largest single generator that could be taken "off-line"; in this case of the two hydroelectric turbines (15.5 MW) located at Aishihik. The forecast system peak demand 1980-81 is 55.5 MW.

Thus:

$$\begin{aligned}\text{Reserve capacity (MW)} &= \text{capacity largest unit} + (0.5\text{xpeak demand})\\ &= 15.5 + (0.05 \times 55.5)\\ &= \underline{18.3\ \text{MW}}\end{aligned}$$

As shown in table 2 below the system has adequate capacity to satisfy the current requirements. The additional 9.0 MW load imposed by the development programs however, will result in a 10.0 MW shortfall in the system reserve capacity, which will be largely offset by the installation of approximately 9 MW of thermal generating capacity. The small (1.0 MW) and temporary reduction in reserve capacity is not sufficient to justify the installation of additional generating equipment, particularly when considering the excess reserve system capacity that will exist once the new hydroelectric turbine is installed and operating by mid 1983.

TABLE 2

N.C.P.C. SYSTEM CAPACITY vs. REQUIREMENTS

CASE	PERIOD	PEAK DEMAND MW	CALCULATED RESERVE MW	REQUIRED GENERATING MW	ACTUAL CAPABILITY MW	SURPLUS (SHORTAGE) MW
1	Forecast 1980/81	55.5	18.3	73.8	77.7	3.9
2	Forecast plus Vangorda Development Load	68.8	18.9	87.7	77.7	(10.0)
3	As for case 2 plus 9 MW additional installed	68.8	18.9	87.7	86.7	(1.0)
4	As for case 2 plus additional 120 MW Hydro Unit at Whitehorse	68.8	18.9	87.7	106.7	19.0

Note: a) Data includes provision for normal load growth.

b) Case 2 is hypothetical base case.

6.3 SUPPLEMENTARY POWER GENERATING

6.3 SUPPLEMENTARY POWER GENERATING

6.31 Alternative Systems Studied

Network analyses of the system were integrated with predicted total grid loadings to establish the need for an additional 9 MW generating facility, that would be required to maintain the system reserve. A number of power generating alternatives were considered including, hydroelectric generating systems, gas turbines, and diesel engine generators.

Hydroelectric generating potential in the immediate area was found to be limited. Nevertheless, studies are scheduled to develop an inventory of possible sites and identify their respective generating potentials.

Gas turbines were found to be less expensive than stationary diesel generating units in terms of capital cost, although their inferior fuel efficiency characteristics, render these machines more suitable for stand-by and peaking service. Diesel engine power generating units are normally favoured to supply long term energy needs. These units require a higher initial capital investment, however this disadvantage is offset by the improved reliability and a somewhat lower unit cost of energy produced.

6.32 System Selection

The selected system to maintain the essential reserve capacity at the 18-20 MW level was something of a compromise. The system comprised of two gas turbine generators of about 3.0 MW rated capacity each. The remainder of the reserve capacity would be supplied by a diesel generator with a nominal 2.5 MW capacity.

TABLE 3

NEW THERMAL GENERATORS IN SYSTEM

TYPE	NUMBER	LOCATION	ESTIMATED OUTPUT AT MAXIMUM LOAD MW
Cullen Gas Turbine	2	N.C.P.C. Plant-Faro	2 x 3.0
Midwest (EMD) Diesel	1	Minesite	1 x 2.5

The location of the units was dictated by a number of considerations. Installation at Whitehorse would have been acceptable but the line losses in transmission would have increased to 15-16% of the power generated making the additional power very expensive. Installation at the minesite of all the capacity was studied in detail but was rejected because of both operating and capital costs.

Finally it was decided to install two gas turbine generators at the N.C.P.C. plant at Faro and one diesel unit at the minesite. The capital cost was relatively low and the concept provided for a small but reliable unit to be installed at the minesite to act as an emergency generator. Also this arrangement placed most of the capacity directly in the hands of the N.C.P.C. operators at the Faro plant.

6.33 Ancillary Equipment

The main transformer at the minesite (138-4.16 KV) is fully loaded at 16 MVA and at times to close to overload - a condition which markedly increases the probability of failure. With the increased load on the system additional transformer capacity is clearly required.

As an integral part of the mill modifications an identical 16 MVA transformer will be produced and installed in parallel with the existing unit. This will permit the additional load to be transformed to 4160 V and significantly reduce the risk of transformer failure.

In addition to the above a mine isolation transformer will be included in the circuit to improve the power characteristics of the mine electrical supply.

6.4 POWER SUPPLY TO THE VANGORDA PLATEAU

6.4 POWER SUPPLY TO THE VANGORDA PLATEAU

Power to the Vangorda Plateau minesites will be supplied by a branch line from the existing 138 KV transmission line between the minesite and Faro. A transformer and substation will be located adjacent to the new installations. (Figure 6-2)

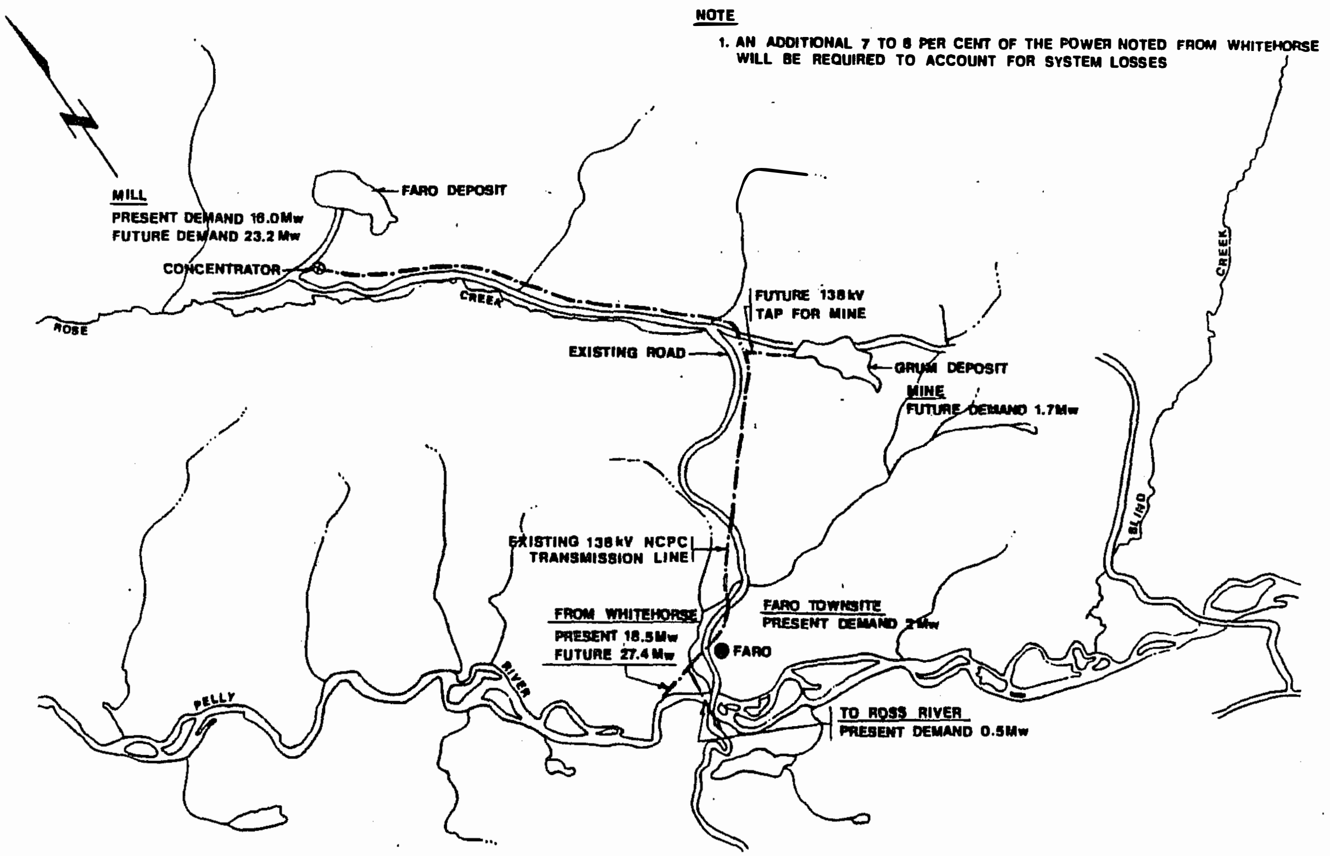


FIG. 6-2
N.T.S.

POWER DISTRIBUTION REQUIREMENTS FARO AREA

At this time (April 81) it is not possible to accurately access the power requirements of the new mine complexes, at the Vangorda and Grum minesites.

6.5 LONG TERM POWER SUPPLY

6.5 LONG TERM POWER SUPPLY

6.51 Hydro Electrical Sites

The N.C.P.C. is presently considering a number of potential hydroelectric sites within the Yukon with a view to commencing staged development programs to provide sufficient energy for long term territorial growth. Areas currently under study are listed below with their respective potential capacities.

TABLE 4

POTENTIAL HYDROELECTRIC DEVELOPMENTS

POTENTIAL SITE	POTENTIAL CAPACITY MW
Eagle Rock	75
Francis River	70
Teslin River	45
Little Salmon River	15

In view of the transmission losses referred to earlier, and the increasing cost of thermal generated energy, there will be considerable advantage to Cyprus Anvil if any one of these sites could be developed.

6.52 Thermal Generation

In addition to these studies conducted by the N.C.P.C., Cyprus Anvil will continue to evaluate the feasibility of thermal power generating systems including local and remote coal fired generators and other fossil fuel systems. To date studies have suggested that a base load coal fired plant could be a possible source of future electrical power. More work is required to determine mineable coal reserves in the locations studied, environmental impact and cost per unit of power generated.

7
ENVIRONMENT

ENVIRONMENT

INDEX - ENVIRONMENT

7.0 SUMMARY

7.0 SUMMARY

The proposed modifications to the existing concentrator operation have been designed to improve environmental conditions. Four specific areas were identified in the environmental assessment as requiring attention:

a) Respirable Dust Levels

These levels have been in excess of safe levels in some areas of the operation, necessitating protective apparatus for employees in these areas and regular physiological tests at specified intervals.

By covering the new flotation cells with steel plate and by significantly reducing the number of conveyor transfer points in the filtering and drying area, the overall dust level in the concentrator will decline significantly.

b) Noise Levels

Typically concentrators have elevated noise levels. Hearing protection is required in most areas where prolonged employee exposure is necessitated during a normal working shift.

The increased use of rubber mill liners of sound absorbing materials in some motor housings and gearbox shields and a significant overall reduction in the numbers of small high speed motors will reduce noise in most areas to relatively safe levels.

c) Water Consumption

Total water consumption in the operation is very high, although to date, the consumption has been within the water license requirements. To perform at best efficiency the metallurgical techniques needed for ore treatment in the new circuits, additional quantities of water, beyond present license limits are required.

To avoid further demand on the local water supply and storage system, the concentrator modifications include a simple water reclamation scheme within the concentrator. Included in the overall concentrator modifications are plans for reclamation of water from the tailings ponds, subsequent if necessary, to a significant aging period. The execution of these plans is contingent upon the completion of a considerable amount of laboratory testwork designed to evaluate the effects of recycling water on the flotation system.

7.1 INTRODUCTION

7.1 INTRODUCTION

The field of environmental studies is a relatively new factor in mine or mill development studies, but is one of growing significance. In this very brief section devoted to the environmental aspects of the development of the Vangorda Plateau ore bodies, are outlined the basic environmental concepts used in the design studies. It may be argued that some of the detail should more correctly be labelled as employee health, safety measures or water conservation concepts, however all have been grouped together here as a matter of expediency.

The primary concern of the environmental design concepts was toward ensuring that working conditions within the concentrator be improved by reducing dust levels and noise in the expanded work zones. Next in importance were that the constraints of water conservation be met and finally that the more aesthetic considerations of site appearance and ecological damage be considered.

Improvements in any of these areas will enhance employee safety and strengthen morale by furnishing an improved environment for those working and living in the Faro area.

7.2 RESPIRABLE DUST LEVELS

7.2 RESPIRABLE DUST LEVELS

Because of the nature of the metallurgical operations at Anvil, considerable quantities of dust are generated and released into the surroundings. Of some concern are the levels of dust, especially the ultra-fine particulates, to which some employees are routinely exposed. Inhalation of even chemically neutral particulates may be hazardous and on average, the dusts to which Anvil workers are exposed are, in general, chemically active.

Careful studies conducted over the last five years at the minesite have been used as a design base for planning and achieving improved air quality in the concentrator area. In the present mill expansion design, significant efforts were directed toward continuing the established trend of enhanced air quality in all working areas.

As shown in Table 1 below, significant reductions in airborne dust levels have been achieved in most areas of the plant in recent years. However in some areas, the concentrations of lead in the air (about one third of which is in the reactive form of lead oxides), approaches and occasionally exceeds desirable levels.

TABLE 1

AIR QUALITY IN THE CONCENTRATOR

	A R E A							
PERIOD	GRINDING		FLOTATION		FILTERS		LOADOUT	
	DUST$_3$ mg/m^3	LEAD$_3$ ug/m^3	DUST$_3$ mg/m^3	LEAD$_3$ ug/m^3	DUST$_3$ mg/m^3	LEAD$_3$ ug/m^3	DUST$_3$ mg/m^3	LEAD$_3$ ug/m^3
1973-74	2	250	4	400	14	2200	60	8500
1979	1	80	1	100	1.5	400	2	250
Target	10	150	10	150	10	150	10	150

Notes: a) Dust refers to any particulate.

b) Lead refers to the total lead. As a point of interest dust collected during these surveys suggests that at least 30% of the total lead present in the air is in the form of non-sulphide compounds of relatively high.

c) Regulation is the adopted standard level for Cyprus Anvil Mine. The American Conference of Industrial Hygenists has suggested that this standard is appropriate.

Further significant reductions in respirable dust levels throughout the concentrator are anticipated due to a number of fundamental design criteria in the expanded mill which are detailed below:

i) The principle source of dust emissions in the flotation area are the flotation cells. As the rising air bubbles coalesce and burst at the foam-air interface, ultra-fine particulates are ejected at considerable velocities into the atmosphere. Naturally, these particulates are rich in minerals and surface active agents.

To reduce dust from this source, the new rougher/ scavenger cells will be covered with steel plate. At this time, there are no plans to cover the cleaner cells, but this is an obvious next step for the operations group to adopt once the need for observing the flotation froths is reduced by the installation and successful operation of the on-stream analysis unit.

ii) The most prolific dust source in the filter dryer area is the transfer conveyor system feeding the dryers and the loadout belts. Present plans indicate that the new system, with a much simpler conveyor- transfer arrangement, and the elimination of the swing conveyor, will result in a significant decrease in suspended dust loads.

iii) Better filter-dryer control systems should, while reducing overall product moisture, almost eliminate periods of over-drying and hence dusting conditions from the dryers. This will result in much improved work conditions in the load-building and loadout conveyor galleries.

7.3 NOISE LEVELS

7.3 NOISE LEVELS

Average noise levels in the concentrator, have always been near or above the threshold level at which hearing protection is required. Since the recognition of the problem in 1976, and the implementation of audio-testing of new employees, there have been several cases of significant receptor damage amongst exposed employees. The table below indicates the extent of the problem.

TABLE 2

CONCENTRATOR NOISE LEVELS BY AREA.(DbA)

"SAFE" LEVEL	CRUSHING	GRINDING	FLOTATION	FILTERING
85	95 - 105	93 - 95	85 - 90	85 - 87

Notes: a) Data from 1979 averaged survey by area in the concentrator.

b) "Safe" level is recommended maximum exposure level (B.C.-W.C.B. 1979-80), for continuous 8 hour employee exposure.

Since milling operations are often extremely noisy, the most obvious solution to the noise problem is to provide protective gear to exposed employees and to limit employee exposure to high sound pressure areas. This approach, coupled with a well planned employee testing program, has been quite successful to date.

In the modified mill however, there are some design features which should greatly reduce average noise levels. These features are detailed below:

(i) In the new milling section, the large ball mills will be equipped with rubber liners thus effecting a reduction in noise and incidentally resulting in a marked increase in pulp temperature.

It is possible then that in the grinding circuit alone, more than 90% of the installed power could be transmitted to rubber lined mills. Significant reduction in average sound pressure levels should accrue.

(ii) Noise emission from pumps and mill motors is also a source of excessive sound pressures. Plans now call for the replacement of many of the small high speed motors with much fewer larger, and in some cases, lower speed motors.

(iii) In the flotation area much of the sound pressures originate from drive belts on the pumps and flotation machines; this occurs because the machinery is loaded to maximum tolerances. The design of the modified circuit, at rated tonnage will dictate that the larger drives will not be stressed to maximum loadings and hence noise levels will be reduced.

7.4 WATER CONSUMPTION

7.4 WATER CONSUMPTION

Use of water in the operations at Faro, has received very little attention during the last decade. However, with the present trends toward conservation, the attitude of the water licensing authorities has become somewhat firmer and it would be reasonable to expect that no increase in license quantities could be anticipated above the annual 2.0 x 10^9 imperial gallons, now allocated to the Faro operation under the current license.

Using this criteria as a design base, a scheme for water recirculation was devised to satisfy the increased water demand from the modified process. The scheme consists of two distinct phases which are outlined in the following sections.

7.41 Internal Recirculation

Reduction of flotation densities throughout the circuit results in significant additional water demand. In order to satisfy this demand, water will be recirculated from two main sources within the concentrator; the thickener overflows and the cooling and scrubbing system effluents. In addition the recycling of the lead retreat tails to the head of the lead roughers will conserve considerable quantities of water. The table below shows the present consumption of water and the projected usage.

TABLE 3

ESTIMATED WATER DEMAND BY AREA - MILL ONLY

AREA	PRESENT MILL		MODIFIED MILL	
	m^3/hour	IGPM	m^3/hour	IGPM
Grinding	290	1060	330	1210
Flotation - Lead	220	810	370	1360
- Zinc	165	600	365	1340
Dewatering & Miscellaneous	400	1470	415	1520
TOTAL	1075	3940	1480	5430

Notes: a) Demand data at rated tonnage. April 1979 survey.

b) See drawing no. 10-F-07 Rev. 0.

c) License quantities are equivalent to 4150 IGPM (1130 m^3/hr.) at 92% operating.

Recirculated water in the mill may be divided into two types, clean water which may be reused without treatment and water which requires some pre-treatment before reuse. As shown in the table below almost 25% of the total water demand can be recirculated without treatment; another 20% will require treatment before reuse.

TABLE 4

ESTIMATED RECIRCULATION WATER QUANTITIES & SOURCES

WATER QUALITY	AREA	QUANTITY		COMMENTS
		m^3/hr	IGPM	
Dirty Water	Thickener Overflows	250	920	Probably this fluid will require treatment before reuse. Planned treatment scheme is to use ionic flotation.
Clean Water	Dust Collectors	60	220	This water is quite clean and can be used almost anywhere in the process.
	Air Conditioners	60	220	
	Vacuum Seals	90	330	
	Cooling Water	90	330	
	Miscellaneous Other	30	110	
	Total Recirculation	580	2130	

Ref. Drawing 10-F-07 Rev. 0

7.42 Ionic and Particulate Flotation

The internal recirculation of some fluids, notably the thickener/ scrubber overflows require some pre-cleaning stage prior to reuse in the process. Currently under study is a Japanese process which, on similar fluids, yields remarkable results in terms of removal of deleterious metal ions from the process fluids.

The process which is essentially one of flotation depends for its effectiveness upon the production of insoluble metal xanthates within the flotation process and their subsequent removal as a concentrate - hence effectively deionizing the fluid. Some reported data on this type of process are shown in the table below.

TABLE 5

ION FLOTATION DATA - ZINC THICKENER OVERFLOW

QUANTITY TREATED	m^3/day	ASSAYS (g/tonne)				DISTRIBUTION			
		Cu	Zn	Cd	CN	Cu	Zn	Cd	CN
Feed	2200	22	B1	0.4	29	100.0	100.0	100.0	100.0
Concentrate	1	127000	470000	2000	7000	98.0	98.2	98.2	26.0
Tailing	2199	0.5	1.5	0.01	21.5	2.0	1.8	1.8	74.0

Notes: a) Data from Kamioka Mines.

b) Concentrate tonnage about 1.5 tonnes/day.

Thus it is possible to estimate that total fresh water demand will continue at the current level of 1050–1100 m^3/hour, of which about 900 m^3/hour will be utilized in the process. Approximately 550-600 m^3 will be recirculated to provide the necessary dilution in the flotation circuits.

7.43 External Recirculation

If water usage is further limited in the future by changing legislation, recirculation outside of the concentrator will have to be practiced. At this time only two stages of additional recycling have been given consideration.

(i) Thickening & recycling

This concept envisages thickening of the zinc retreat-first cleaner tailings, flotation cleaning and the recycling for a yield of about 5000 m^3/day of process water. This system has the disadvantage of not allowing the fluids to "age" - a critical aspect if reuse in flotation circuits is to be seriously considered. Conversely this type of arrangement would provide a warm water recycle - demonstrably advantageous in lead/zinc flotation.

(ii) Recycling from the tailings pond

This concept calls for a pump station to be located in the tailings treatment system and should permit most of the process water to be recycled after "aging". Allowing for seepage and entrainment, it would be reasonable to anticipate about 85% of the total water utilized in the process could be recycled e.g. about 22-24,000 m^3/day.

These two schemes are by no means mutually exclusive and could proceed in parallel. Laboratory testwork conducted over a reasonable period using both "aged" and filtered water would provide a solid design base for future water conservation.

In summary then future use of water at Cyprus Anvil may be significantly reduced by a series of conservation measures. The speed at which these measures proceed will depend upon the results of tests aimed at determining effect of recycling water on plant metallurgy and also the efficiency of the ionic flotation scheme described earlier. The table below demonstrates estimated water consumption by conservation phase.

TABLE 6

WATER CONSUMPTION AND RECIRCULATION

PHASE	TOTAL USE		FRESH WATER		RECYCLED WATER	
	m^3/hour	IGPM	m^3/hour	IGPM	m^3/hour	IGPM
Present Operation	1050 - 1150	3850 - 4220	1050 - 1150	3850 - 4220	0	0
I-Clean Water Recycle II-Dirty Water Recycle	1400 - 1500	5130 - 5500	900 - 950	3300 - 3500	580	2130
III-Cleaner Tails Thickening	1400 - 1500	5130 - 5500	700 - 750	2570 - 2750	200	730
IV-Recycling from Pond	1400 - 1500	5130 - 5500	200 - 400	730 - 1460	400	1460

Notes: a) For technical reasons phases I & II should proceed simultaneously.

b) All data based on current best estimates and operation at rated tonnage.

7.44 Water Recirculation in Other Mines

A recent survey of Canadian operations indicates that almost all mines started up during the last five years practice recirculation to some extent. Table 7 below shows the majority of all operating mines practice recirculation.

TABLE 7

RECYCLE WATER PRACTICE

SURVEY OF ALL CANADIAN MILLS

MILL PRACTICE	TOTAL RECYCLED %
Recycle	70
No Recycle	30

Note: a) Data from C.I.M. 1978 Milling Review 1978, covering approximately sixty operations.

It is of interest to compare recirculation practiced by mine type. Simple single mine operations are more likely to recirculate than complex base metal massive sulphide mines, although it would appear that most mines are able to reach 50% recirculation without major problems. Table 8 summarizes the data.

TABLE 8

RECYCLE WATER PRACTICE BY MILL

PLANT NAME	PRODUCT	RECYCLE (% of Total)
Stekenjokk	Cu/Zn	30
Asarco	Pb/Zn	60
Mattagami	Pb/Zn	75
Geco	Pb/Zn	100
Willroy	Pb/Zn	80
Cominco - Sullivan	Pb/Zn	50
Kamioka	Pb/Zn	90-100
Texas Gulf	Cu/Pb/Zn	60
Kopparberg	Cu/Pb/Zn	25
Brunswick	Cu/Pb/Zn	50
Kristineberg	Cu/Pb/Zn	55
Mattabi	Cu/Pb/Zn	70
Opemiska	Cu	70
Gibraltar	Cu	80
Granduc	Cu	85
Madelaine	Cu	65
Whitehorse Copper	Cu	80
Thierry	Cu	75
Afton	Cu	50
Endako	Mo	85
Consolidated Durham	Mo	90
Brenda	Cu/Mo	88-90
Similkameen	Cu/Mo	75
Utah	Cu/Mo	40
Inco - Coppercliff	Ni	95
Inco - Clarabell	Ni	90
Umex	Ni	65

Ref. Data from published literature, Circa 76-79 may not now be accurate due to greater demands and environmental constraints in recent years.

7.5 DOWN VALLEY TAILINGS POND

7.5 DOWN VALLEY TAILINGS POND

7.51 Down Valley Tailings Pond Expansion

Changes in the mill design, the decision to process Vangorda Plateau ore in the Faro concentrator, and the continuing deterioration in effluent quality from the existing tailing pond all combine to make a new tailings disposal scheme a matter of some urgency. The proposed "Down Valley" design has the following basic features:

a) The capacity is large enough to accommodate all known economic reserves in the Anvil Range.

b) Seepage into Rose Creek is controlled by keeping Rose Creek above the level of the tailings, and by building an essentially water tight dam.

c) Stability of the tailings dyke is enhanced by not using tailings in its construction.

d) Water quality is improved by using a two stage design in which tailings are impounded in a first pond and relatively clear effluent is contained in a second pond. The effluent will spend most of its retention time in the second pond, separated from the tailings. The decant will not join the main Rose Creek flow for 700m downstream from the dam, providing space for additional water quality treatments if needed. These features should all help to alleviate the many problems which have plagued the present tailings pond.

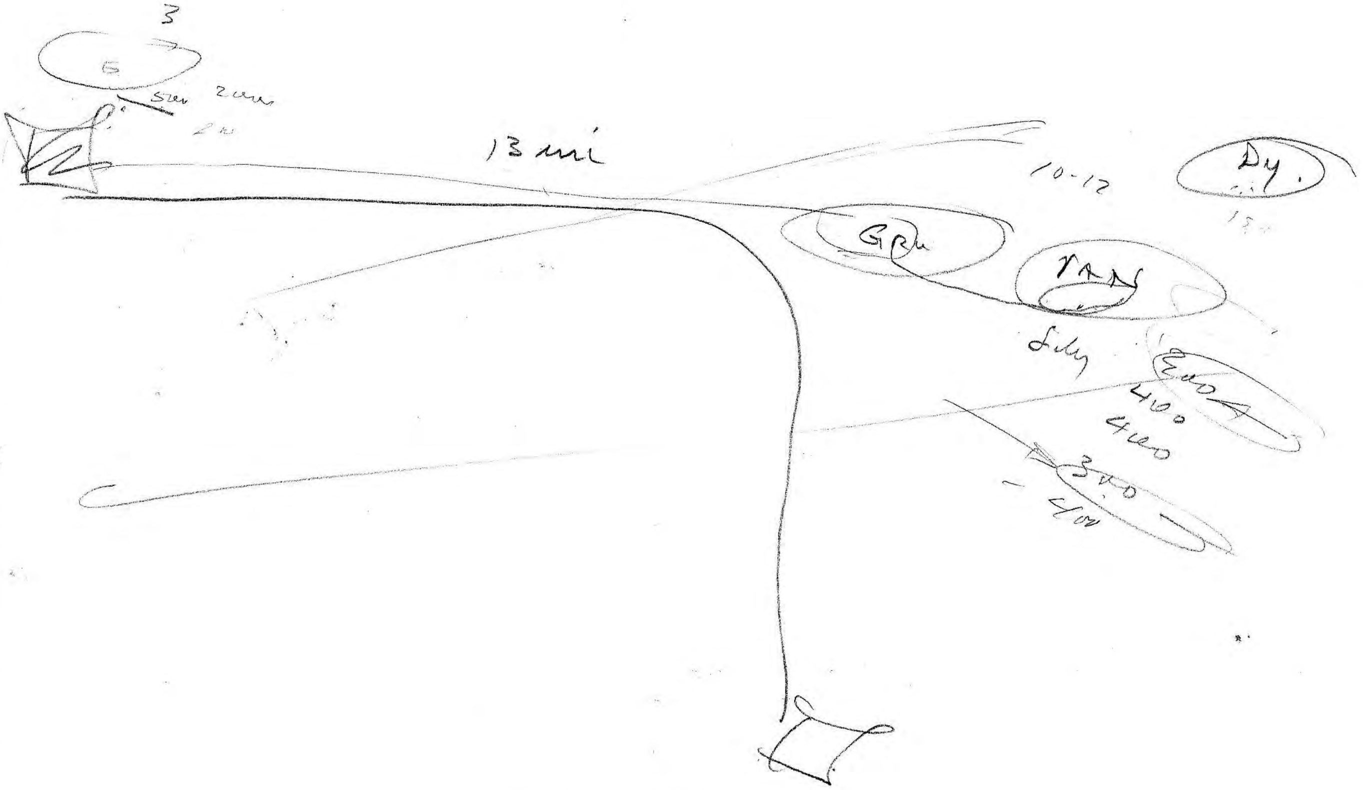
The Rose Creek diversion canal will have a total length of about 4,000m, and will be designed to accomodate a one hundred year flood with no damage. A five hundred year flood would be required to overtop the diversion banks.

The tailings pond dam will be constructed with a spillway large enough to handle the full flow of Rose Creek, so that in the unlikely event of a failure of the diversion, the tailings pond dam would not be breached.

The problem of long term stabilization of the tailings is exacerbated by the large increase in tailings surface area that will result when the new tailings scheme is put into operation. At the same time, the opportunity to study this problem will be available for the first time, since some tailings areas will soon become inactive and available for reclamation research.

8
G.A. DRAWINGS

GENERAL ARRANGEMENT DRAWINGS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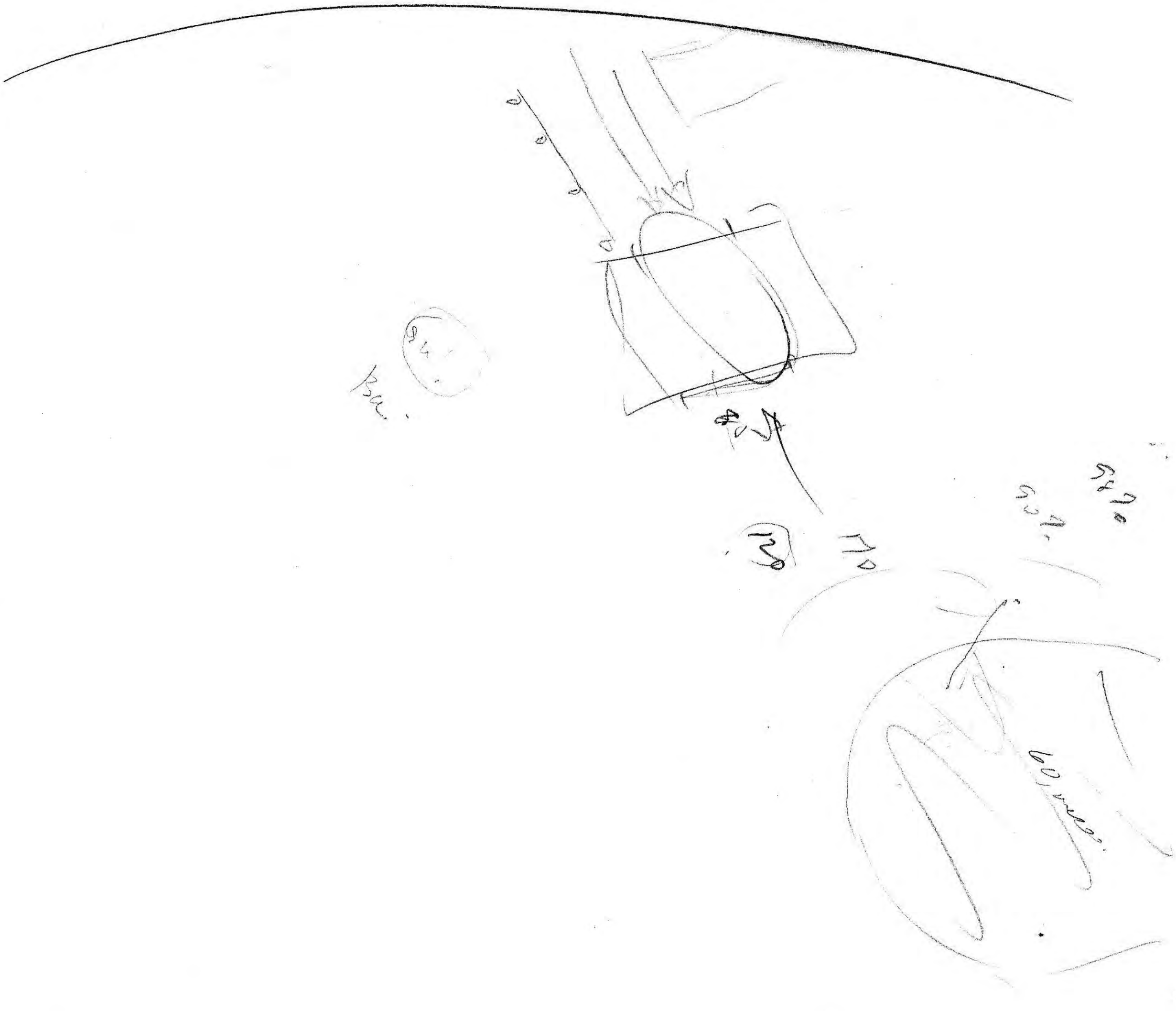
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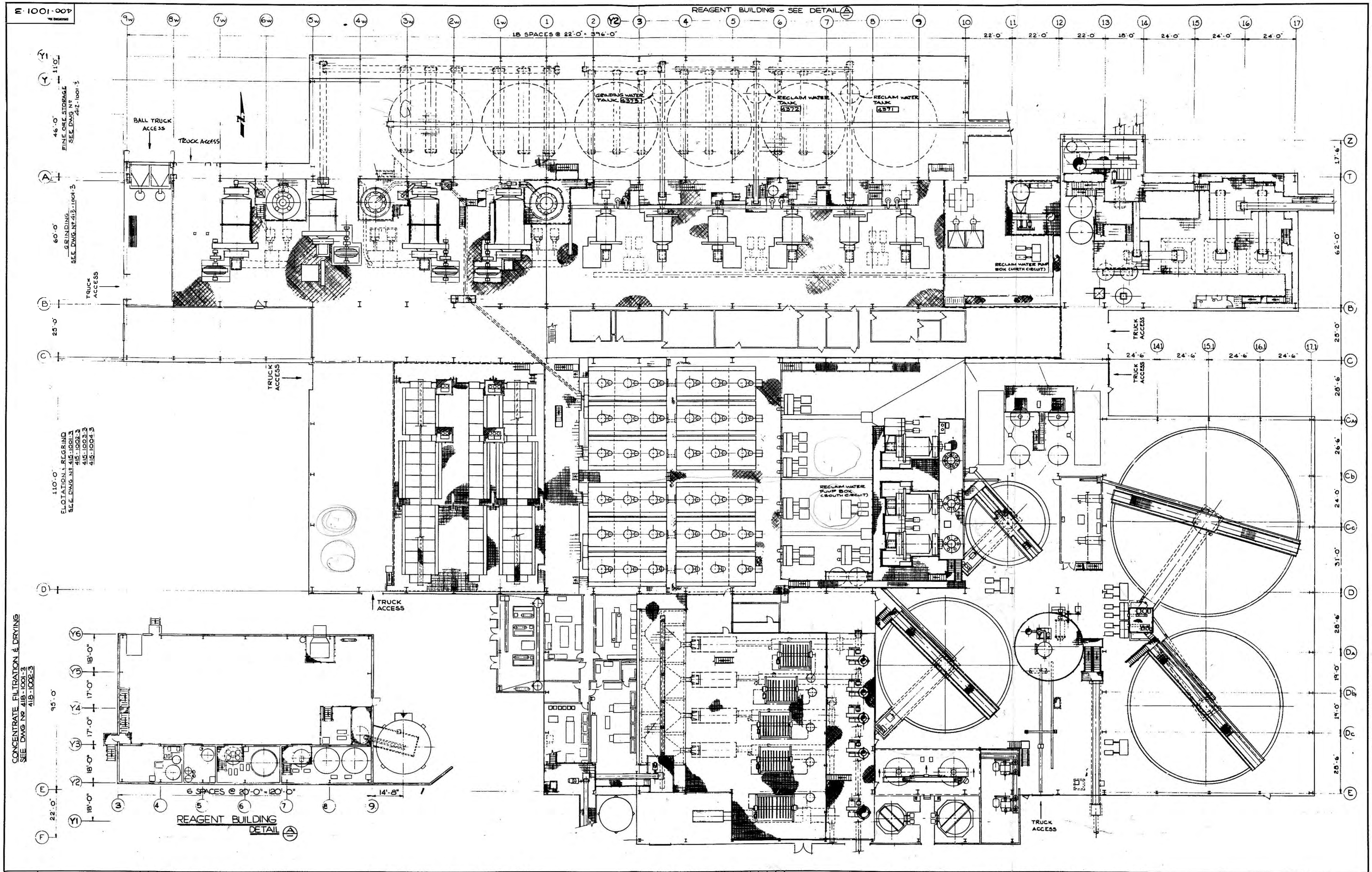

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B	ELECTR. C11 REFERENCE	AUG 15/80	GFA
A	ISSUED FOR CLIENT'S APPROVAL	MAY 30/80	E.L.

REVISIONS

SECTION: MECHANICAL
SCALE: 1/16" = 1'-0"
DESIGNED BY:
DRAWN BY: B. ROSS 20/12/79
CHECKED BY:
APPROVED BY:

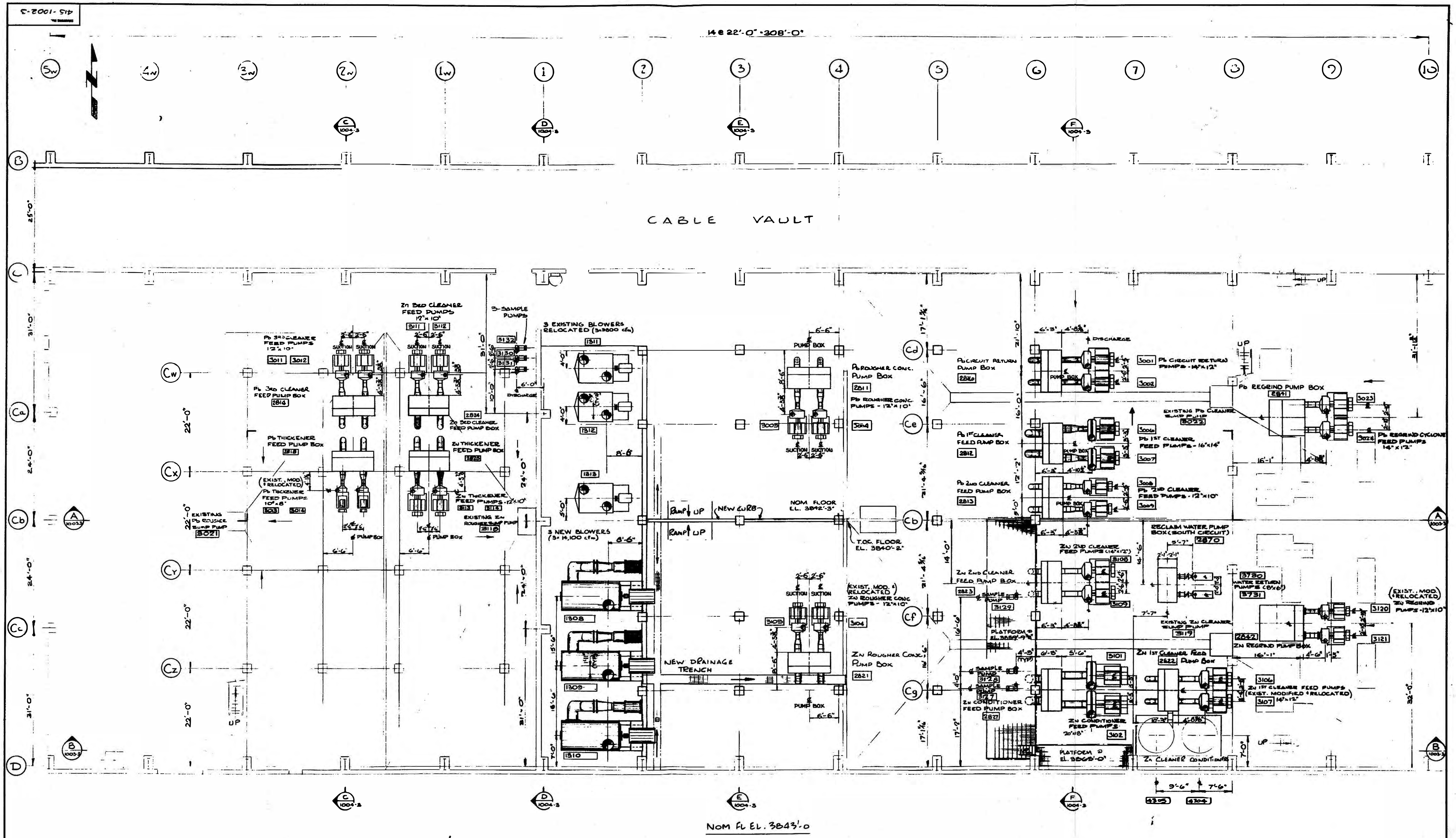
CLIENT: CYPRUS ANVIL MINING CORP
VANGORDA PLATEAU DEVELOPMENT
LOCATION: FARO YUKON TERRITORY

KILBORN

TITLE
MILL MODIFICATIONS
CONCENTRATOR
GENERAL ARRANGEMENT
PLAN

ACCT. No.
PROJECT No. 7436
DIVISION No.
DRAWING NUMBER 400-1001-3
REV. C

DWG. NO.	REFERENCE DRAWINGS



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415-1001 3	G A - OPERATING FLOOR

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1	ISSUED FOR CONSTRUCTION C-11	SEPT 26/80	RL
B	ISSUED FOR TENDER C-11	AUG 18/80	RAJ
A	ISSUED FOR CLIENTS APPROVAL	MAY 30/80	RR

REVISIONS

SECTION: MECHANICAL
SCALE: 1/8" = 1'-0
DESIGNED BY:
DRAWN BY: R.A
CHECKED BY: AW 29.9.80
APPROVED BY: RPN 29/9/80

CLIENT: CYPRUS ANVIL MINING CORP VANGORDA PLATEAU DEVELOPMENT
LOCATION: FARO YUKON TERRITORY

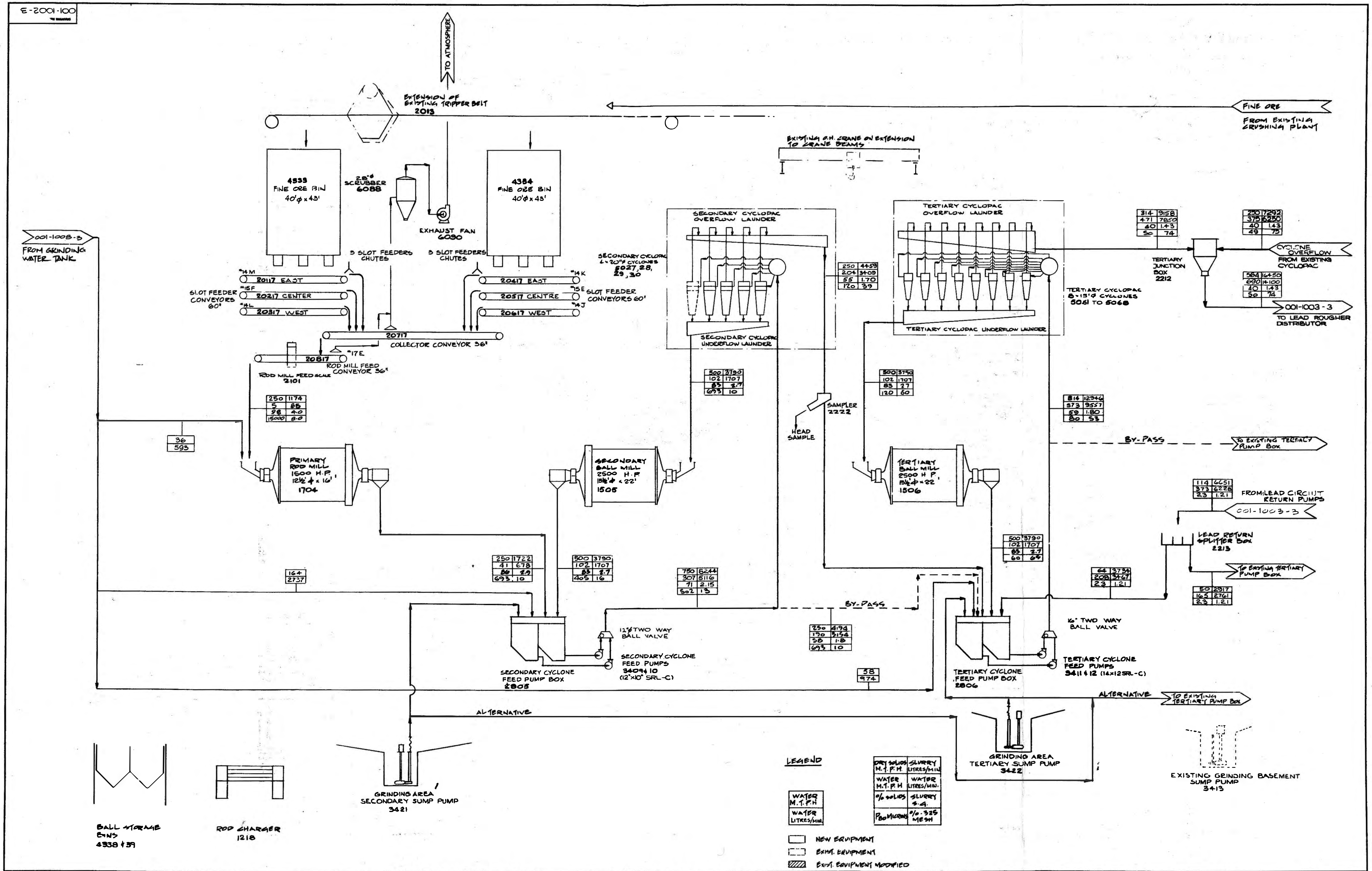
KILBORN

TITLE
MILL MODIFICATIONS
FLOTATION & REGRIND AREA
GENERAL ARRANGEMENT
GROUND FLOOR PLAN

C-11
ACCT. No.
PROJECT No. 7436
DIVISION No.
DRAWING NUMBER 415-1002-3
REV. 2

NCI-2508-K.E.L.

E-2001-100



LEGEND

WATER M.T.P.H	
WATER LITRES/MIN.	

DRY SOLIDS M.T.P.H.	SLURRY LITRES/MIN
WATER M.T.P.H	WATER LITRES/MIN.
% SOLIDS	SLURRY S.G.
P80 MICRONS	% -325 MESH

- ☐ NEW EQUIPMENT
- ⬚ EXIST. EQUIPMENT
- ▨ EXIST. EQUIPMENT MODIFIED

NO.	DESCRIPTION	DATE	BY
C	ISSUED FOR REFERENCE	JAN/12/81	RL
B	REVISED & ISSUED FOR CLIENT APPR.	MAY 30/80	RL
A	ISSUED FOR CLIENT APPROVAL	15 FEB. 80	KK

REVISIONS

DWG. NO. | REFERENCE DRAWINGS

SECTION: METALLURGICAL
SCALE: —
DESIGNED BY: R.M.N. DATE: DEC 79
DRAWN BY:
CHECKED BY: RMN MAR. 11 '80
APPROVED BY: RPW 8/7/80

CLIENT: CYPRUS ANVIL MINING CORP.
VANGORDA PLATEAU DEVELOPMENT
LOCATION: FARO . YUKON TERRITORY

KILBORN

TITLE
MILL [illegible]
FINE ORE BINS AND
GRINDING EXTENSION
FLOWSHEET

ACCT. No.
PROJECT No. 743[illegible]
DIVISION No.
DRAWING NUMBER 00 [illegible]
REV. C

LEAD FLOTATION

STAGE	FEED MTPH.	% SOLIDS	FLOTATION TIME (MIN)	STAGE RECOVERY	CONC GRADE	SOLIDS S.G.
ROUGHERS	564	40	9.3	80	19.5	4.0
SCAVENGERS	485	43	16.4	45	3.0	4.0
FIRST CLEANERS	176	25	24	85	30.0	5.0
SECOND CLEANERS	89	25	20.1	75	53.0	5.5
THIRD CLEANERS	33	20	19.6	70	67.0	6.0
OVERALL				87.5	67.0	

NOTE FLOTATION TIMES CALCULATED ON TOTAL FEED TO BANK, INCLUDING RECIRCULATING LOADS.

LEGEND

(A) AUTOMATIC SAMPLER

(L) LAUNDER SPRAY

DRY SOLIDS MTPH	SLURRY LITRES/MIN	
WATER MTPH	WATER LITRES/MIN.	
% SOLIDS	SLURRY S.G	

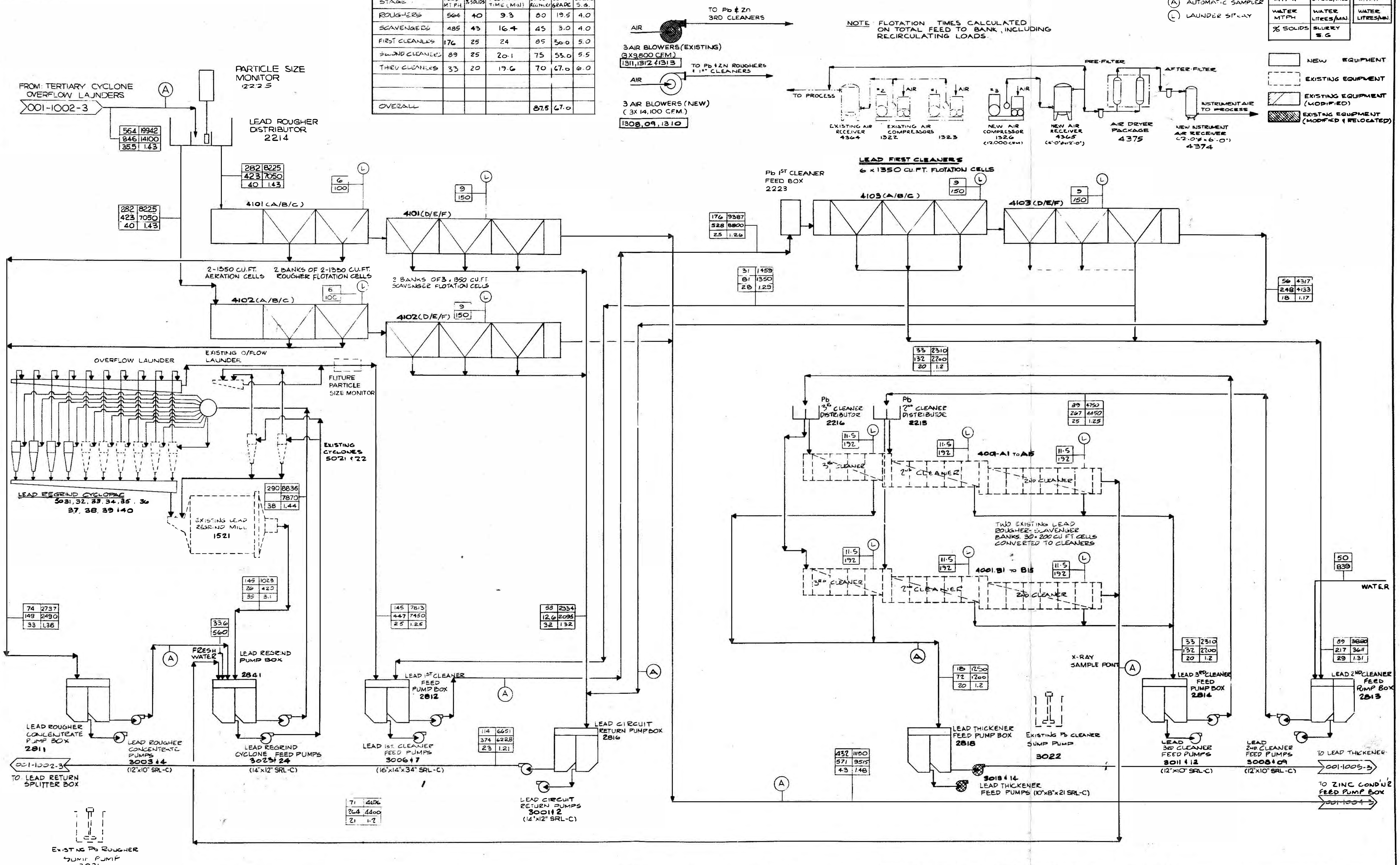
WATER MTPH
WATER LITRES/MIN.

☐ NEW EQUIPMENT

⬚ EXISTING EQUIPMENT

▨ EXISTING EQUIPMENT (MODIFIED)

▩ EXISTING EQUIPMENT (MODIFIED & RELOCATED)



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B	REVISED & ISSUED FOR CLIENT APP	MAY 30/80	RL
A	ISSUED FOR CLIENT APPROVAL	15 FEB. 80	KK

REVISIONS

		DATE
SECTION: METALLURGY		
SCALE: NONE		
DESIGNED BY: R N		DEC '79
DRAWN BY: W/R		DEC 5/79
CHECKED BY: RMN		MAR 14 '80
APPROVED BY: RPN		8/7/80

CLIENT: CYPRUS ANVIL MINING CORP
VANGORDA PLATEAU DEVELOPMENT
LOCATION: FARO, YUKON

KILBORN

TITLE

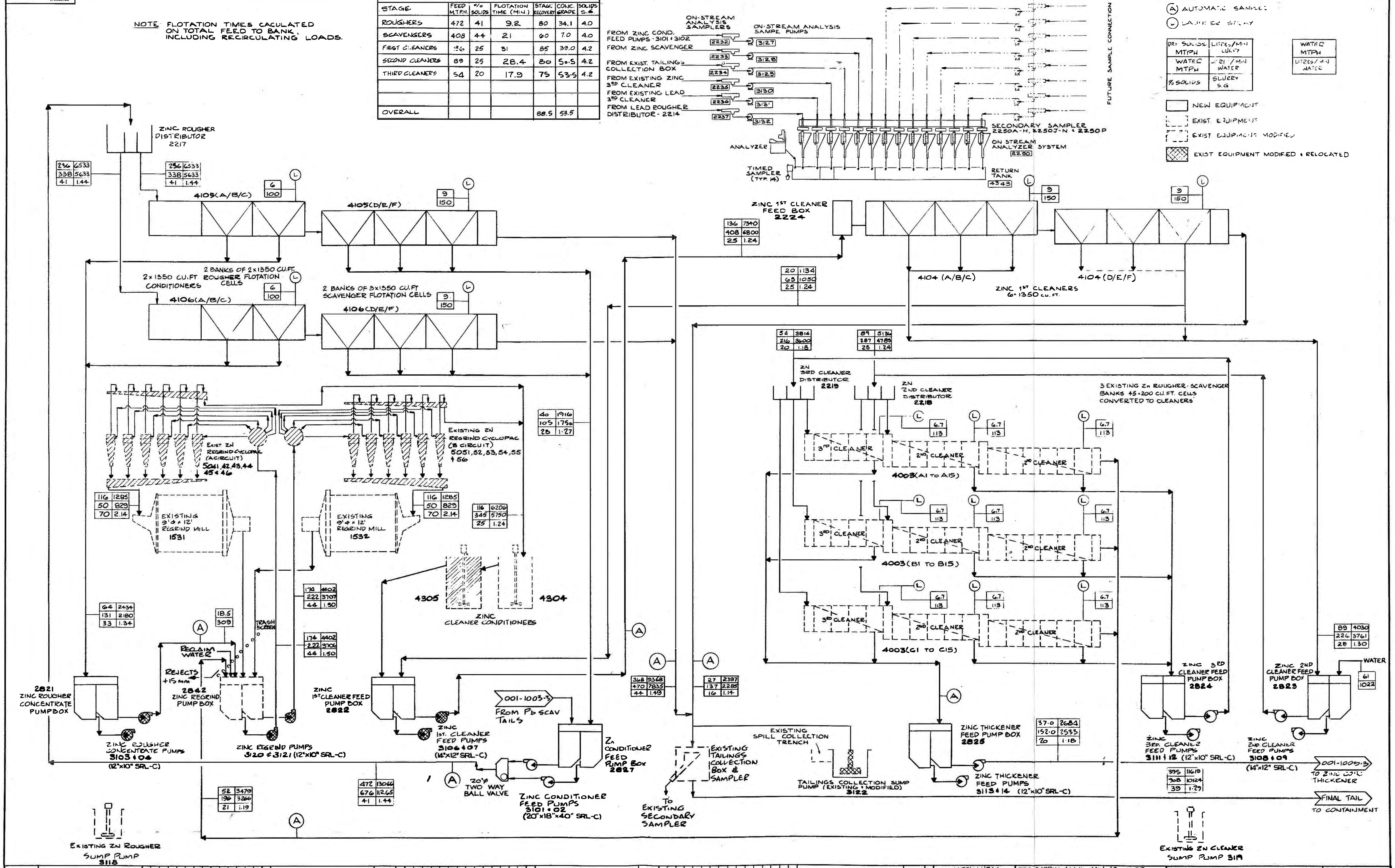
MILL MODIFICATIONS
LEAD ROUGHER, CLEANER,
FLOTATION & REGRIND
FLOWSHEET

ACCT. No.
PROJECT No. 7436
DIVISION No.
DRAWING NUMBER 001-1003-3
REV. C

NOTE FLOTATION TIMES CACULATED ON TOTAL FEED TO BANK, INCLUDING RECIRCULATING LOADS.

ZINC FLOTATION

STAGE	FEED M.T.P.H.	% SOLIDS	FLOTATION TIME (MIN.)	STAGE RECOVERY	CONC. GRADE	SOLIDS S.G.
ROUGHERS	472	41	9.2	80	34.1	4.0
SCAVENGERS	408	44	21	60	7.0	4.0
FIRST CLEANERS	136	25	31	85	39.0	4.2
SECOND CLEANERS	89	25	28.4	80	50.5	4.2
THIRD CLEANERS	54	20	17.9	75	53.5	4.2
OVERALL				88.5	53.5	



LEGEND

Ⓐ AUTOMATIC SAMPLER

Ⓛ LAUNDER SPRAY

DRY SOLIDS MTPH	LITRES/MIN SLURRY
WATER MTPH	LITRES/MIN WATER
% SOLIDS	SLURRY S.G

WATER MTPH
LITRES/MIN WATER

☐ NEW EQUIPMENT

⬚ EXIST. EQUIPMENT

⬚ EXIST EQUIPMENT MODIFIED

▨ EXIST EQUIPMENT MODIFIED & RELOCATED

DWG. NO.	REFERENCE DRAWINGS

NO.	DESCRIPTION	DATE	BY
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B	REVISED & ISSUED FOR CLIENT APPL	MAY 30/80	RL
A	ISSUED FOR CLIENT APPROVAL	15FEB.80	KK

REVISIONS

SECTION: METALLURGY
SCALE: NONE
DESIGNED BY: R.M.N — DEC '79
DRAWN BY: W. KUPFER — DEC 11/79
CHECKED BY: RMN — MARCH '80
APPROVED BY: RPW — 8/7/80

CLIENT: CYPRUS ANVIL MINING CORP
VANGORDA PLATEAU DEVELOPMENT
LOCATION: FARO. YUKON TERRITORY

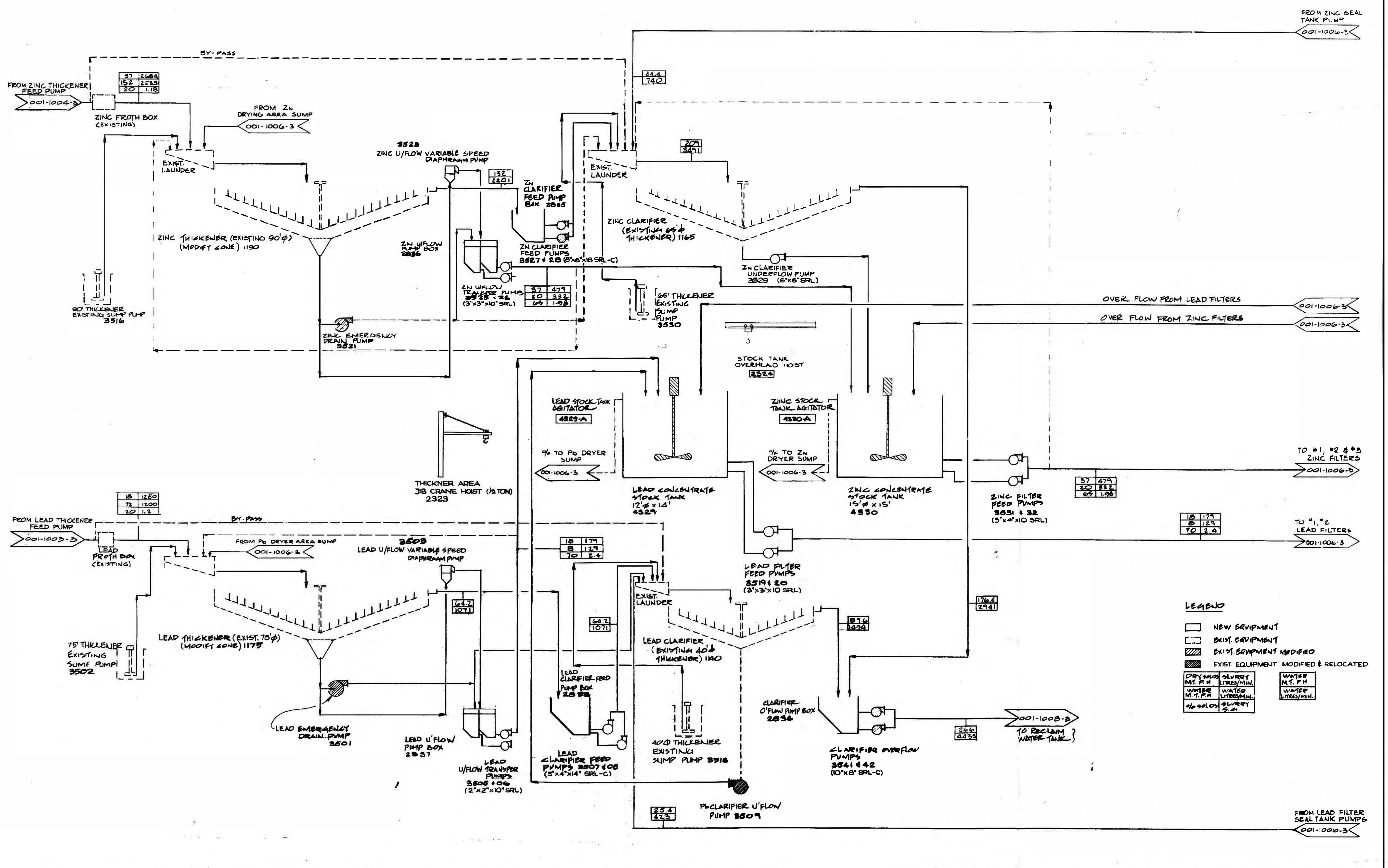
KILBORN

TITLE
MILL MODIFICATIONS
ZINC ROUGHER, CLEANER,
FLOTATION & REGRIND
FLOWSHEET

ACCT. No.
PROJECT No. 7436
DIVISION No.
DRAWING NUMBER 001-1004-3
REV. C

NCI-253 E-K.E.L.

E-S001-100



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B	REVISED & ISSUED FOR CLIENT APP'L	MAY 30/80	RL
A	ISSUED FOR CLIENT APPROVALS	15 FEB. 80	KK

REVISIONS

DWG. NO. | REFERENCE DRAWINGS

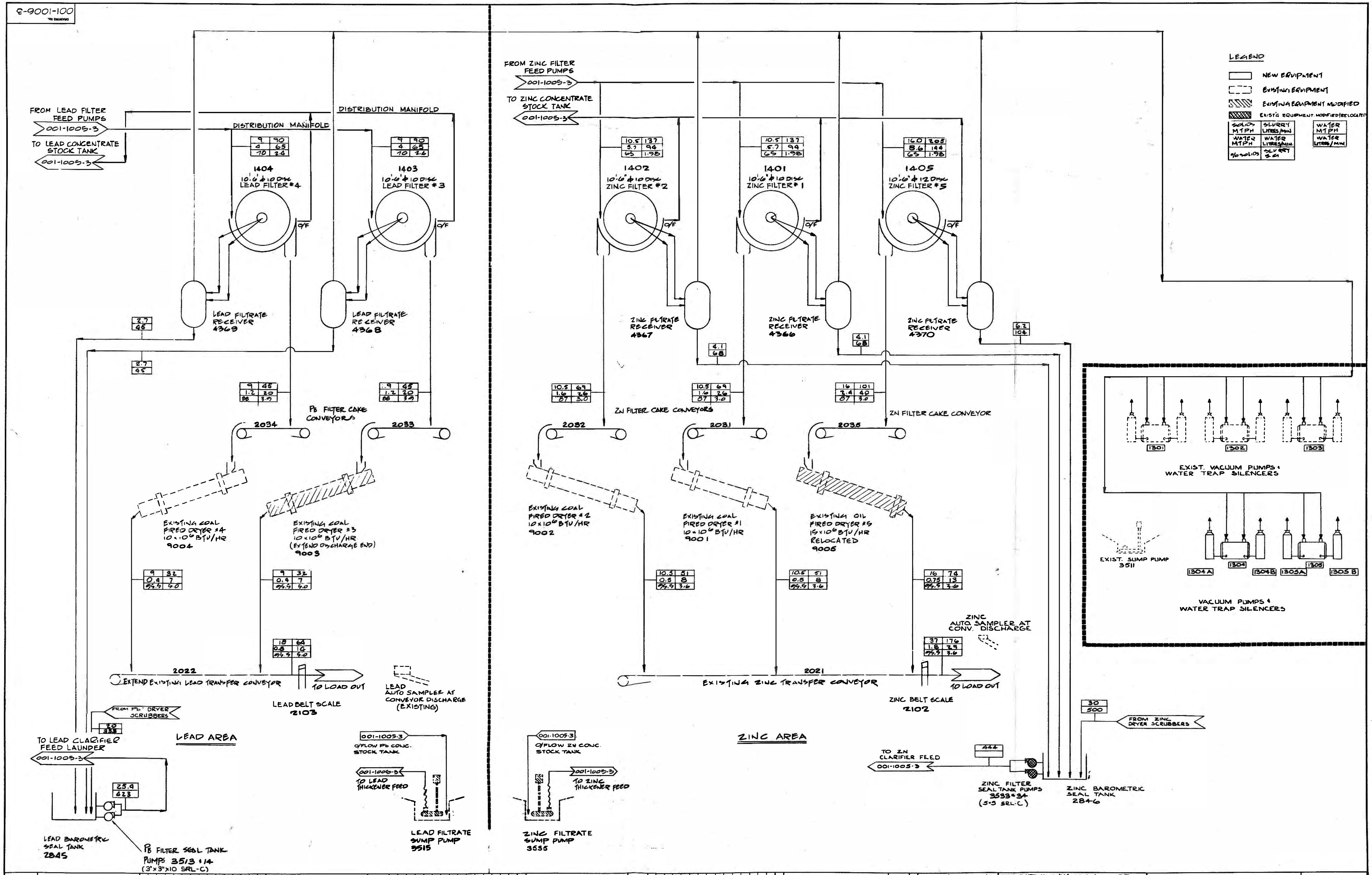
SECTION: METALLURGICAL
SCALE: —
DESIGNED BY: R.M.N DATE DEC 79
DRAWN BY: P.L.
CHECKED BY: RMN MARCH '80
APPROVED BY: RRW 8/7/80

CLIENT: CYPRUS ANVIL MINING CORP.
VANGORDA PLATEAU DEVELOPMENT
LOCATION: FARO, YUKON TERRITORY

KILBORN

TITLE
MILL MODIFICATIONS
CONCENTRATE THICKENING
FLOWSHEET

ACCT. No.
PROJECT No. 7436
DIVISION No.
DRAWING NUMBER 001-1005-3
REV. C



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B	REVISED & ISSUED FOR CLIENT APP'L	MAY 30/80	RL
A	ISSUED FOR CLIENT APPROVAL	15 FEB 80	KK

REVISIONS

SECTION: METALLURGICAL
SCALE: —
DESIGNED BY: Ran — DEC. 75
DRAWN BY: P.J.
CHECKED BY: Kmn — MARCH '80
APPROVED BY: RPW — 8/7/80

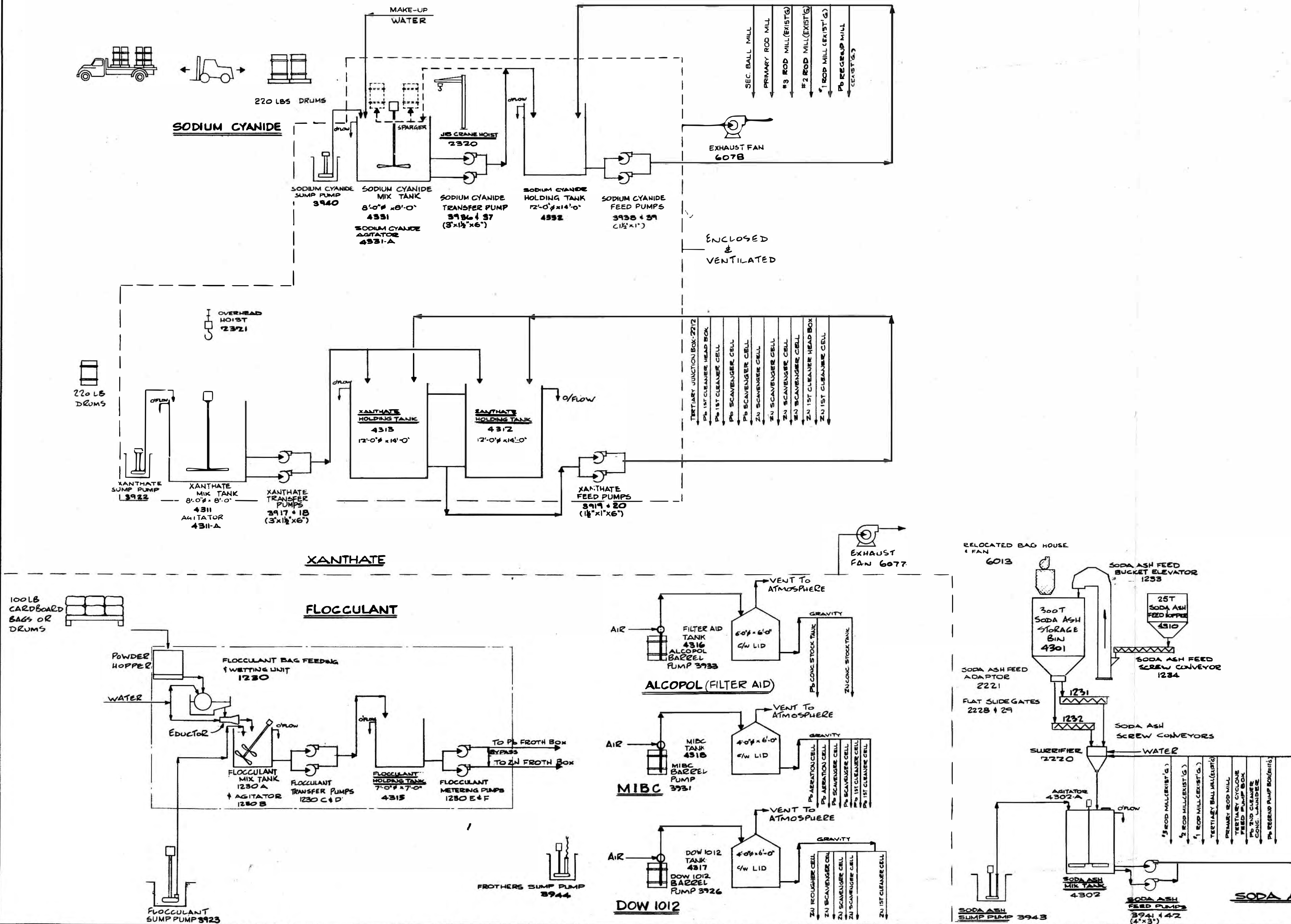
CLIENT: CYPRUS ANVIL MINING CORP.
VANGORDA PLATEAU DEVELOPMENT
LOCATION: FARO, YUKON TERRITORY

KILBORN

TITLE
MILL MODIFICATIONS
CONCENTRATE FILTRATION &
DRYING FLOWSHEET

ACCT. No.
PROJECT No. 7436
DIVISION No.
DRAWING NUMBER 001-1006-3
REV. C

001-1007-3



NOTES

(I) EXISTING COPPER SULPHATE SYSTEM REMAINS UNCHANGED

(II) TANK SIZES CATER FOR POSSIBLE FUTURE EXPANSION TO 15000 tpd

(IV) EXISTING LIME SYSTEM REMAINS UNCHANGED

LEGEND

☐ NEW EQUIPMENT

▨ EXISTING EQUIPMENT MODIFIED & RELOCATED

DWG. NO.	REFERENCE DRAWINGS
001-1027-3	P & I DIAGRAM
001-1028-1	"

NO.	DESCRIPTION	DATE	BY
2	ISSUED FOR REFERENCE	JAN 12/81	RL
1	ISSUED FOR CONSTRUCTION - C9	SEPT. 19/80	RL
B	ISSUED FOR CLIENT APPROVAL	MAY 30/80	RL
A	ISSUED FOR CLIENT APPROVAL	MAR 4/80	KK

REVISIONS

SECTION:	METALLURGICAL	DATE
SCALE:	NONE	
DESIGNED BY:	R MN	DEC '79
DRAWN BY:	C.C.	AUG 22 '79
CHECKED BY:	RMN	MARCH '80
APPROVED BY:	RFW	8/7/80

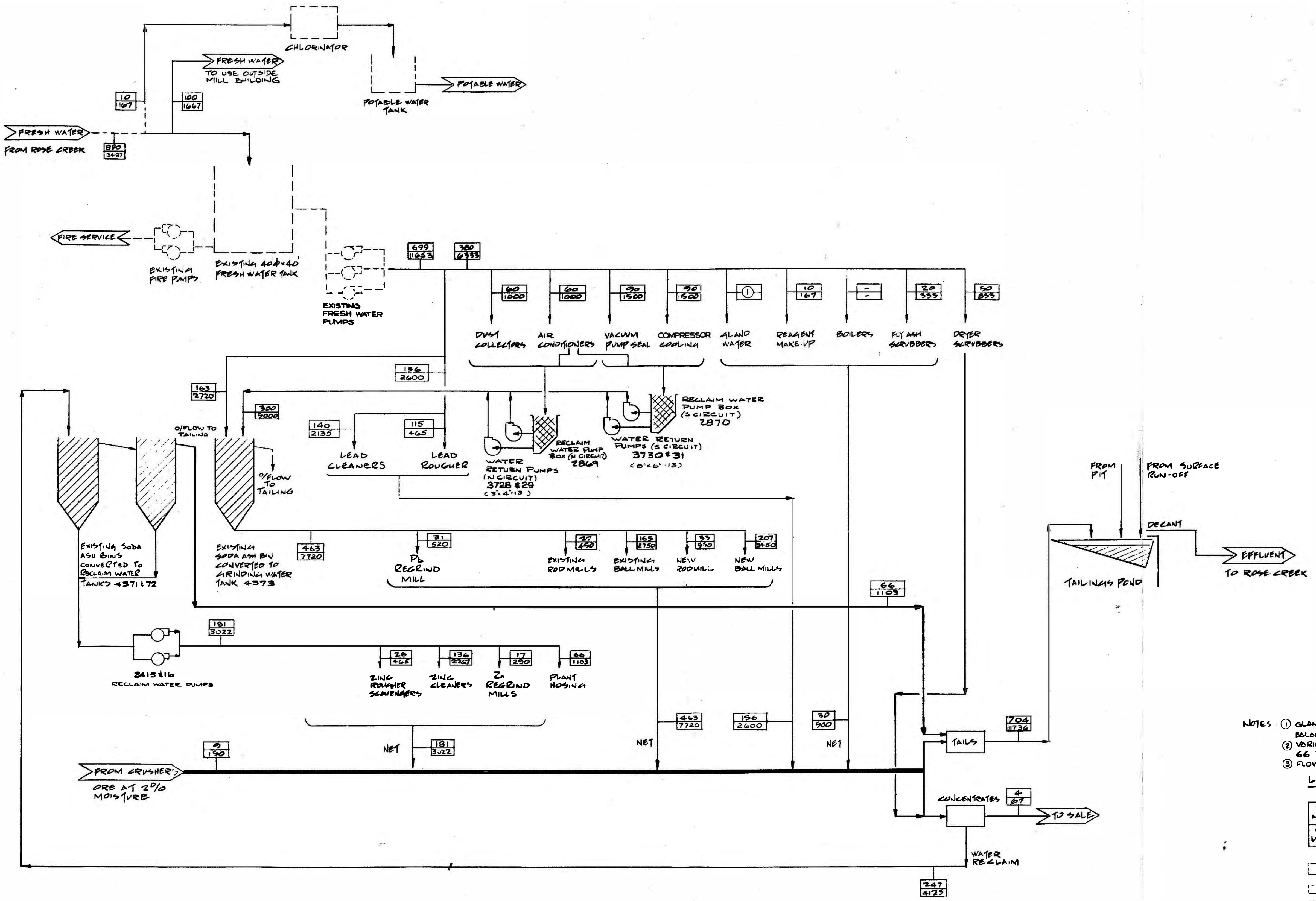
CLIENT: CYPRUS ANVIL MINING CORP
VANGORDA PLATEAU DEVELOPMENT
LOCATION: FARO YUKON TERRITORY

KILBORN

TITLE
MILL MODIFICATIONS
REAGENT HANDLING & MIXING
FLOWSHEET

ACCT. No.
PROJECT No. 7436
DIVISION No.
DRAWING NUMBER 001-1007-3
REV. 2

DRAWING No. 001-1008-3



NOTES
① GLAND SERVICE INCLUDED IN MATERIAL BALANCE WITH LAUNDER SPRAY MAKE-UP
② VARIABLE DEMAND UP TO MAXIMUM OF 66 TONNES/HR TOTAL. BALANCE TO TAILINGS
③ FLOW BASED ON TREATING 3,402,000 TONNES/YR

LEGEND

WATER MTPH
WATER LITRES/MIN

- ☐ NEW EQUIPMENT
- ⬚ EXISTING EQUIPMENT
- ▨ EXISTING EQUIPMENT MODIFIED
- ▩ EXISTING EQUIPMENT RELOCATED

NO.	DESCRIPTION	DATE	BY
D	ISSUED FOR REFERENCE	JAN 12/81	RL
C	GENERAL REVISION		
B	REVISED & ISSUED FOR CLIENT APP'L	MAY 30/80	EL
A	ISSUED FOR CLIENT APPROVAL	15 FEB 80	KK

REVISIONS

DWG. No. | REFERENCE DRAWINGS

SECTION: METALLURGICAL
SCALE: —
DESIGNED BY: R.M.N — DEC '79
DRAWN BY:
CHECKED BY: RMN — MARCH '80
APPROVED BY: R.R.W — 8/7/80

CLIENT: CYPRUS ANVIL MINING CORP
VANGORDA PLATEAU DEVELOPMENT
LOCATION: FARO, YUKON TERRITORY

KILBORN

TITLE
MILL MODIFICATIONS
WATER RECIRCULATION
PHASE 1 - FLOWSHEET

ACCT. No.
PROJECT No. 7436
DIVISION No.
DRAWING NUMBER 001-1008-3
REV. D