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Attention: Mr. J.A. Abramo

Dear Sir:

Re: Anderson Uranium Mill

We submit herewith our report "Metallurgical Summary and Mill Design Criteria for the Anderson Uranium Deposit", May 1978.

The recommended flowsheet, after an additional campaign of test work, has proven to be quite similar to that selected and presented in our report of September 1977 "Metallurgical Summary and Preliminary Design Data for the Anderson Uranium Deposit". The most significant change is substitution of a semi-autogenous grinding mill and rod mill for the crushing and grinding circuit chosen for the preliminary report. The overall flowsheet and design data have been considerably refined and validated from the additional test work.

This report has been issued for use by Minerals Exploration Company, Union Oil Company, and the engineering contractor selected for the detailed design. We request that it not be distributed to other companies or individuals.

We appreciate this opportunity to be of service.

Yours very truly,


M.E. Grimes, P. Eng.

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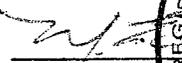
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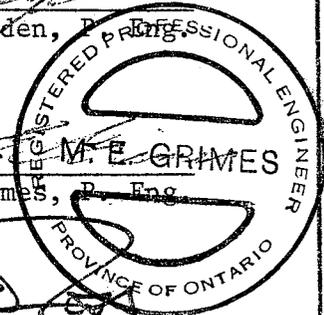
MINERALS EXPLORATION COMPANY

Metallurgical Summary and Mill Design Criteria
for the Anderson Uranium Deposit


A.S. Hayden, P. Eng.


M.E. Grimes, P. Eng.


A.H. Ross, P. Eng.



Toronto, Ontario, Canada
May, 1978

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Appendix I Plan - Analytical & Testing Laboratory

1. INTRODUCTION

1.1 Preface

A. H. Ross & Associates have been engaged since March 1977 as metallurgical consultants to Minerals Exploration Company relative to recovery of uranium from the Company's Anderson deposit in Arizona. The following activities have been undertaken by A. H. Ross & Associates and are summarized in this report:

Direction, interpretation, and evaluation of the testwork performed by Hazen Research, Inc. and by equipment manufacturers.

Review of test results from the viewpoints of uranium recovery, capital cost, circuit alternatives, and optimization of operating cost.

Selection of a suitable process flowsheet.

Development of design data and criteria to serve as the basis for detailed design of a process plant.

1.2 Evaluation Bases and Premises

The estimates of metallurgical recovery and reagents consumption in this report have been based on an average mill feed grade of 0.069% U_3O_8 .

For economic studies to select process conditions and evaluate flowsheet alternatives, a selling price of U.S. \$43.20 per pound of U_3O_8 in yellowcake was specified by Minerals Exploration Company.

The evaluations, design data, and equipment sizing have been based on a mill feed rate of 2200 tons per day of ore for 365 days per year or 803,000 tons per year for the mill circuit up to solvent extraction. From solvent extraction through yellowcake processing, the equipment sizing, etc. have been based on a production of 3960 pounds per day of U_3O_8 .

1.3 Reference Documents

1. Hazen Research, Inc. report, "Alkaline leaching of Anderson Mine Samples", August 15, 1977.
2. Hazen Research, Inc. report, "Acid leaching of Anderson Mine Samples", August 24, 1977.
3. Hazen Research, Inc. report, "Uranium Recovery from Anderson Mine ore", March 15, 1978.
4. A. H. Ross & Associates report, "Metallurgical summary and preliminary design data for the Anderson uranium deposit", September 1977.
5. Morrison-Knudsen report, "Preliminary feasibility study, Anderson project, uranium mine and mill", December 1977.
6. Letter report from Enviro-Clear Division of Amstar Corporation, March 1, 1978.
7. Envirotech Corp. report, "Report of investigations. CCD thickening and counter-current vacuum filtration and washing of acid leached uranium ore residue from Anderson, Arizona", March 1978.

2. SUMMARY

On the premise that the samples supplied for metallurgical testing are representative, it has been demonstrated that the Anderson uranium deposit is amenable to treatment in an acid leach flowsheet. The recommended flowsheet is believed to be the most economical and reliable of the alternatives investigated. However, it is possible that a preferred flowsheet could be developed from a broader based and more lengthy experimental program.

High extraction and recovery of uranium has been demonstrated under relatively severe leaching conditions and with relatively high reagent consumptions. Laboratory tests produced yellow cake product of satisfactory purity using the flowsheet conditions recommended herein.

An average overall recovery of uranium of 88.5 percent from mill feed assaying 0.069% U_3O_8 has been estimated for the selected flowsheet when using the specified flowsheet conditions. Uranium losses, as a percent of uranium content in mill feed, are estimated as follows:

Leaching	9.0%
Countercurrent decantation	1.75%
Solvent extraction	0.25%
Unaccounted	<u>0.5%</u>
Total	11.5%

Annual production when processing 2000 tons of ore per day for 365 days per year would be 891,550 pounds U_3O_8 .

3. METALLURGICAL INVESTIGATIONS AND PROCESS SELECTION

3.1 Mineralogy

The Anderson uranium deposit is reported to occur primarily in two ore horizons separated from each other by a distance of about fifty feet. The host rock consists mostly of shale and siltstone, with both relatively hard and soft layers. The zones are highly variable with respect to uranium and carbonate content, with the higher carbonate content in the lower zone. In April 1977, Hazen Research, Inc. (HRI) analyzed a large number of drill core sections arising from drilling programs in 1975 and 1976. The CO₂ content was found to vary from 0.03 to 32%, the U₃O₈ content from 0 to 0.5%, and with no apparent correlation between them. Samples from a new drilling program in the autumn of 1977 showed a very similar pattern.

3.2 Ore Samples

The A.H. Ross & Associates preliminary report (Ref. 4) described the bulk samples and drill core samples that were used in testwork at Hazen Research, Inc. in early 1977.

A new drilling program was carried out by Minerals Exploration in the autumn of 1977 to provide a sufficient quantity of representative drill core for an additional metallurgical test program, and to further define ore reserves and mining plans. These core samples were analyzed at Mountain States Research and Development and at Hazen Research, Inc. for core densities and moisture content. Uranium and CO₂ content were determined by Skyline Labs, Inc. Bulk density of core material was found to be lower than anticipated, which led to a downward revision of ore reserve estimates.

A large number of individually coarse-crushed and bagged core sections from thirty-five drill holes were shipped to Hazen Research, Inc. in December 1977 from Minerals Exploration, with instructions for combining sets of samples into six different composites, each of which corresponds to a mining area. The composites, labelled A to F, are also intended to represent a time sequence of mining.

Each core section was riffled into halves, one of which was held in reserve, and the other made available for compositing. The six composites were dried at 50°C, crushed to minus 6 mesh, blended, and split into suitable portions for tests. The weight, specific gravity, and assay for each of the composites are given in Table 1. The calculated assay based on reported uranium content of the core sections is also recorded.

Table 1. Analysis of Composite Samples

<u>Composite</u>	<u>A</u>	<u>B</u>	<u>C</u>	<u>D</u>	<u>E</u>	<u>F</u>
Weight - lb	75	81	136	68	80	46
Specific gravity	2.36	2.41	2.28	2.38	2.41	2.47
Assays						
U ₃ O ₈ radiometric:						
U ₃ O ₈ Beta equivalent	0.054	0.082	0.079	0.055	0.035	0.041
U ₃ O ₈ Gamma equivalent	0.046	0.065	0.061	0.047	0.028	0.037
U ₃ O ₈ Beta/Gamma calculated	0.065	0.11	0.10	0.066	0.040	0.049
U ₃ O ₈ fluorimetric	0.064	0.092	0.084	0.066	0.042	0.049
U ₃ O ₈ volumetric	0.061	0.100	0.091	0.067	0.045	0.049
V ₂ O ₅	0.089	0.062	0.15	0.10	0.066	0.14
Mo	0.005	0.009	0.015	0.005	0.008	0.002
CO ₂	8.0	9.1	3.4	2.7	6.2	1.6
PO ₄	0.16	0.13	0.17	0.14	0.15	0.17
Cl	0.01	0.01	0.01	0.01	0.01	0.01
S	0.49	0.70	1.08	1.03	0.96	0.02
Calculated U ₃ O ₈ assay	0.060	0.096	0.088	0.067	0.042	0.054

In addition to the individual composites, a mixed B-C composite and two master composites containing portions of all six individuals were prepared. They were proportioned as follows:

<u>Individual Composite</u>	<u>Weight %</u>		
	<u>Master Composite 1</u>	<u>Master Composite 2</u>	<u>Composite B - C</u>
A	16.67	16	
B	16.67	14	39
C	16.67	22	61
D	16.67	19	
E	16.67	21	
F	<u>16.67</u>	<u>8</u>	
Total	100	100	100

The proportions used for Master Composite 2 were supplied by Minerals Exploration as representative of the expected tonnage proportions among the six individual mining areas. Composite B - C was made up in the proportions of B and C from Master Composite 2. It was selected for definitive leaching testwork because it had indicated the poorest settling behaviour in leach amenability tests. At the same time, it was considered to be indicative of ore requiring higher than average power input.

3.3 Grinding

3.3.1 Hazen Research Tests

3.3.1.1 Comparative grindability of composites

During the alkaline leaching testwork in early 1977, a Bond work index of 12.8 was determined in grinding from minus 6 mesh to minus 65 mesh. This grindability measure was employed in formulating the preliminary design for grinding in the A. H. Ross report of September 1977 (Ref. 4).

A wide variation in composition both within and between drill core sections in chemical composition and appearance was evident from earlier samples and from the autumn 1977 drilling program. Consequently, it was decided to carry out comparative grindability tests at Hazen Research on the six individual ore composites.

The first comparisons were made by wet grinding at 50% solids 1 kg of minus 6 mesh ore to approximately 2% plus 28 mesh in a laboratory ball mill, in preparation for comparative leach amenability tests among the six composites. The time required to reach the desired grind was noted. Screen analyses of the minus 6 mesh and ground ore was measured for each test. Significant differences were evident with time requirements ranging from 7 minutes to 12 1/2 minutes. However, differences among the minus 6 mesh feeds tended to mask the grindability data.

A more reliable comparison was made by first screening the minus 6 mesh ore to the same approximate size distribution for each of the six composites, then grinding 1 kg of each ore for 10 minutes in a laboratory ball mill. The results of these tests are given in Table 2.

Table 2. Ore Hardness - Comparative Grinds

<u>Grinding Conditions:</u>			
Ore feed	Minus 6 mesh, all composites of similar size distribution, 1000 g		
Ball mill	7-5/8" ID x 7-1/8", 85/90 rpm		
Ball charge	Size	No.	Wt., kg
	Inch Diam		
	1-1/2 to 1-3/4	7	1.54
	1 to 1-1/4	33	2.25
	3/4	17	0.52
	1/2	25	0.27
% solids	50		
Time	10 minutes		

<u>Screen Analysis</u>		Weight Distribution, %					
Size Mesh	Minus 6 mesh Feed	Composite A	Composite B	Composite C	Composite D	Composite E	Composite F
+8	6.7 - 7.5						
8 x 10	17.6 - 19.8						
10 x 14	15.0 - 16.7						
14 x 20	11.3 - 12.0						
20 x 28	8.7 - 9.0	5.0	4.8	4.5	2.5	6.6	0.5
28 x 35	6.6 - 6.7	2.1	2.0	1.7	0.8	1.5	0.6
35 x 48	5.6 - 6.0	7.5	7.4	6.8	4.2	6.7	1.6
48 x 65	4.1 - 4.7	9.5	9.3	9.2	6.7	9.1	3.4
65 x 100	3.9 - 4.6	10.6	10.4	11.0	9.3	10.0	6.4
100 x 150	3.0 - 4.6	9.4	9.4	10.0	8.7	9.4	8.7
150 x 200	2.7 - 5.5	6.6	6.0	6.3	6.1	5.9	8.5
- 200	8.4 - 10.0	49.3	50.7	50.5	61.7	50.8	70.3

It is evident that Composite F is much easier to grind than the average, with Composite D following. The differences between the other composites were small. D and F together are estimated to make up about 27% of the Anderson deposit.

3.3.1.2 Comparative ball milling on Master Composite 1

Master Composite 1 was used for initial leaching tests at two different grinds. The minus 6 mesh feed was ball milled for 7 minutes and for 11 minutes. These times were within the range previously noted for the individual composites. The 7 minute grind yielded 1.9% plus 14 mesh and 34.8% - 325 mesh; the 11 minute grind yielded 1.9% plus 28 mesh and 39.9% - 325 mesh.

3.3.1.3 Comparative ball and rod milling

Settling tests on leached pulp indicated a problem with fines, hence it was decided to test a laboratory rod mill against the ball mill. Rod milling was carried out on Composite C for 2, 4, 5, 7 and 10 minutes. The resultant size distributions were compared to ball milling data on the same composite. The data are presented in Table 3. It is concluded that for the same top size distribution, the rod mill gave a narrower overall size distribution with fewer fines.

All subsequent grinding to prepare leach feed samples was carried out in the laboratory rod mill, on the basis of these tests.

Table 3. Rod versus Ball Milling in HRI tests
on Composite C ore

Top size Mesh size	about 5% +28 mesh cumulative weight % passing		about 2% +28 mesh cumulative weight % passing	
	rod mill 4 min.	ball mill 10 min.	rod mill 5 min.	ball mill 12 min.
20	99.0	100	100	100
28	94.9	95.5	98.4	97.6
35	85.0	93.8	94.5	96.6
48	73.1	87.0	82.2	91.3
65	65.1	77.8	71.0	81.9
100	56.8	66.8	60.0	69.9
150	49.1	56.8	48.2	58.3
200	42.2	50.5	29.6	50.3
270	36.3	-	36.9	43.7
325	34.3	-	36.5	38.6

3.3.1.4 Bond work index

A standard Bond work index determination was carried out on Master Composite B - C, in grinding from minus 6 mesh to minus 28 mesh. The Bond work index was found to be 11.8. This agrees quite well with the index of 12.8 found on the earlier sample from the alkaline leaching test program.

3.3.2 Koppers (Hardinge) tests

3.3.2.1 Samples

Two samples of split core, considered to be "hard" and "soft", were selected by Minerals Exploration personnel from drill core obtained in the autumn 1977 program and were sent to the Koppers Company, Inc., Hardinge Operation, York, Pa. in November 1977 for their appraisal. An opinion was requested on amenability to semi-autogenous grinding.

In January 1978, a five pound sample of Master Composite 2 was sent from Hazen Research to Koppers, as representative of the average ore. At the same time, all available comparative grinding data obtained by Hazen Research was also provided to Koppers.

In March 1978, after a meeting to discuss test results to date, two additional samples, each about 4 kg, were sent from Hazen to Koppers. These were from Composites A and E which had required the greatest grinding input in HRI comparative tests.

3.3.2.2 Test results on the November and January samples

Koppers Company reported the following information:

	<u>Broken ore</u> <u>"soft"</u>	<u>Broken ore</u> <u>"hard"</u>	<u>Master</u> <u>Composite 2</u>
estimated natural grain size:	48 mesh	20 mesh	10 mesh
estimated hp.h/ton to give 2 - 3% plus 28 mesh:	3 - 4	8 - 10	10 - 12

On Master Composite 2, an initial grinding test at 70% solids was unsuccessful because of excessive pulp viscosity. Some 15% of the weight reported as plus 28 mesh at a grinding time estimated from the earlier tests. This test was repeated successfully at 65% solids.

Evidently the visual selection of 'soft' and 'hard' core material, whereas it indicated comparative grindability, did not bracket the average conditions. According to the tests on Master Composite 2, the average is tougher than the harder of the two selected samples.

From these results, the Koppers Company concluded that single stage milling was not feasible, and they proposed consideration of a semi-autogenous primary mill in open circuit followed by a rod mill in closed circuit with a scalper screen. They suggested that the softer ores would operate essentially in open circuit, and that the harder ores might require a 10 - 20% circulating load of screen oversize.

Koppers recommended a full scale pilot test of semi-autogenous grinding to develop reliable design data. However, if such a test were not undertaken, they proposed a conservative 12 - 13 hp.h/ton for the two grinding units, with about 40 - 45% of the power to the SAG mill.

3.3.2.3 Test results on the March samples.

These last samples were provided to Koppers, first to permit them to make a more precise determination of apparent power requirements for a SAG mill - rod mill combination, and second, to obtain comparative power required to grind to both nominal 2% +28 mesh and 2% +35 mesh. This latter comparison was requested to assist in overall process economic evaluations then in progress.

It was reported that both composites had very similar power requirements in grinding to the same top size, but composite E produced over 18% more -200 mesh size than did Composite A. Because of pulp viscosity, grinding was done at 55% solids. The estimated power requirements were reported as 6 - 7 hp.h/ton for 2% +28 mesh and 7 - 8 hp.h/ton for 2% +35 mesh.

3.3.3 Selection of Screen Size in Grinding

From the Koppers estimates of power required, the difference in grinding power costs would be about \$0.025/ton between about 2% +28 mesh and 2% +35 mesh with power at \$0.03/kW.h.

Thickening tests done by Envirotech (Eimco) used leached pulp originating from one standard grind, nominally -35 mesh. Hence there is no relative data from their work with respect to different grinds.

Nearly all leach tests carried out by HRI were also carried through a standard cylinder thickening test and a Kynch calculation for thickener area. Tests were made at nominal grinds of -14, -20, -28, -35 and -48 mesh, although there are relatively few comparisons under otherwise identical leaching conditions. The reproducibility of the thickening tests is very poor, as is the usual case. A reasonable estimate indicates an increase of area required in going from -28 to -35 mesh of $0.8 \text{ ft}^2/\text{ton}/\text{day}$, with confidence limits of $\pm 0.8 \text{ ft}^2$ for a range of 0 to 1.6 ft^2 . We estimate the cost of the finer grind, with respect to thickener area as about $\$0.26 \pm 0.26 / \text{ton}$ assuming $\$80/\text{ft}^2$ for thickener area and a 3.5 year amortization.

Offsetting the extra costs of finer grinding, a decrease in uranium loss in residue is looked for. A semi-theoretical calculation was carried out using actual assays of screen fractions, applied to other size distributions, to derive a weighted average residue assay for different feed grinds.

These calculations led to an estimated difference in residue assays between -20, -28, -35 and -48 mesh of about 0.0006% U_3O_8 on the average for each interval. It was further concluded that these differences had limits of $\pm 0.0004\%$.

From other leach tests, with a single residue assay for each test, the difference between grinds was also estimated at about 0.0006% U_3O_8 , with slightly wider limits of $\pm 0.0005\%$.

The risk of unacceptable uranium loss for coarser ore, and poor thickening for finer ore, reduced the practical choice to nominal -28 mesh and -35 mesh. The value of increased uranium extraction less grinding and thickening costs is estimated at some \$0.22/ton in favour of -35 mesh, but with a range of uncertainty from \$0.47/ton in favour of -28 mesh to \$0.90/ton in favour of the finer grind.

Test work was nearly all carried out on the hardest ore. A substantial portion of the ore is softer and will produce more fines for a given power input. In view of this, and the wide net cost estimates, it is considered prudent to decide on a nominal -28 mesh for design.

3.3.4 Basis for Design

The grinding circuit recommended for this project consists of a semi-autogenous primary mill in open circuit followed by a rod mill in closed circuit with a scalper screen.

The principal reason for this selection is to eliminate a crushing circuit in favour of semi-autogenous grinding. A further consideration is that all ore delivered to the millsite will pass an 18 inch square mesh grizzly.

The ability of a semi-autogenous grinding mill alone to produce a product of the described size distribution has not been demonstrated by pilot plant testwork. From the results of the preliminary laboratory-scale investigations by Koppers (Section 3.3.2), grinding of the harder ores in a SAG mill will result in an unacceptable amount of material coarser than the desired size. With softer ores, the results indicated that the product size would be acceptable. As the grinding circuit must be designed to treat the harder ores, a rod mill has been provided after the SAG mill.

When treating softer ores, the possibility exists of overgrinding to produce a product with a high percentage of material at the finer sizes. To avoid this, both mills should be equipped with a variable speed motor drive. With this provision, the grinding circuit would then have the capability to cope with ore variations.

The rod mill would receive the entire pulp discharge from the SAG mill. Only the coarse ore particles would be ground; the fines would pass virtually through the mill without further size reduction. This is a major reason for selecting a rod mill rather than a ball mill for this application. The design of the grinding circuit is premised on closed-circuiting the rod mill with a sievebend. The function of this equipment is to screen tramp oversize and possibly coarse ore from the leach circuit feed. The oversize from the sievebend would be returned to the rod mill, or would be discarded if found to be without value.

As balls are charged to the SAG mill, provision must be made to avoid small balls or tramp steel from entering the rod mill. One means of removal which can be considered is a magnet positioned at the SAG mill discharge.

The recommended grinding circuit should be capable of processing the wide range of ore types indicated by the core composites that have been tested. On the other hand, it is possible that pilot scale grinding tests would permit design of a less expensive installation. One scheme that might prove to be satisfactory would be to close-circuit the SAG mill with a cyclone, sievebend, or vibrating screen, as is practised at six United States mills that process uranium sandstone ores. Provision might then be made for secondary grinding if it were needed at a future date.

Another alternative that further testwork might justify would be to route the SAG mill discharge to a sievebend, with only the oversize going to the rod mill. The rod mill product would join the SAG mill discharge and be recycled to the sievebend. This scheme would be expected to permit a reduction in rod mill size and power consumption but would probably require pumping and screening of coarser ore particles. The use of cyclones or vibrating screens might be considered in this application.

3.4 Leaching

3.4.1 Testwork by Hazen Research

The early 1977 program at Hazen Research had included both alkaline and acid leaching, plus studies of organics removal which were necessary for an alkaline leaching circuit. Evaluation of the research findings by A. H. Ross & Associates as reported in the Design Data report of September 1977 (Ref. 4) had concluded that acid leaching was the more economical and more trouble-free route and had proposed the following average leaching conditions:

pulp density	55%
final pH	1.5
final emf	-500 mV
temperature	80°C
acid addition	500 lb/ton
sodium chlorate	12 lb/ton
retention time	10 h

This program was terminated to provide for an initial feasibility study, while recognizing that more testwork would be needed to develop definitive design data.

In the autumn of 1977, a new program of drilling was undertaken by Minerals Exploration from which additional ore was made available for further leaching and other tests. Tables 4 and 5 summarize the additional leaching tests carried out by Hazen Research, Inc. Settling tests and a Kynch calculation for unit area were made by HRI on most leach pulps and these data are also shown.

Table 4. Hazen Research Single Stage Leach Tests

Test No.	Ore Composite	Grind Time	% Solids	Temp °C	Time h.	Sodium Chlorate lb/ton	Total Acid lb/ton	Final emf -mV	Final Free Acid g/l	Calculated Acid Consumed lb/ton	Assays Heads
1	D	BM 10 min	40	80	12	44	223	390	11	190	0.066
2	A	BM 12.5 min	39	80	12	27	467	440	12	429	0.064
3	F	BM 7 min	43	80	12	12	144	710	12	112	0.049
4	B	BM 12 min	35	80	12	30	548	470	13	500	0.092
5	C	BM 12 min	43	80	12	32	278	330	14	241	0.084
6	E	BM 12 min	41	80	12	31	397	350	12	362	0.042
7	M 1	BM 7 min	39	80	12	23	356	390	15	309	0.064
8	M 1	BM 11 min	39	80	12	28	281	360	4	268	0.064
9	M 1	BM 11 min	37	80	12	23	371	390	16	317	0.063
10	M 1	BM	39	75	10	2.1	369	330	14	325	0.064
11	M 1	BM	38	75	10	6.3	361	340	13	319	0.065
12	M 1	BM	38	75	10	13	357	345	12	318	0.066
13	M 2	RM 5 min	39	75	6	6.3	355	330	9	327	0.064
14	M 2	RM 5 min	39	75	6	6.3	416	350	19	356	0.062
15	M 2	RM 5 min	38	75	6	28	418	385	19	359	0.062
16	M 2	RM 3.5 min	38	75	6	6.3	355	335	8	329	0.065
17	M 2	RM 3.5 min	37	75	6	6.3	417	350	18	356	0.063
18	M 2	RM 3 5 min	37	75	6	28	417	390	20	349	0.063
19	M 2	RM 6 min	39	75	6	6.3	355	330	11	321	0.063
20	M 2	RM 6 min	38	75	6	6.3	417	355	21	348	0.062
21	M 2	RM 6 min	39	75	6	27	417	385	20	354	0.068
22	B - C	RM 5 min	42	75	6	6.3	372	305	8	349	0.088
23	B - C	RM 5 min	41	75	6	6.3	434	335	19	379	0.086
24	B - C	RM 5 min	42	75	6	27	421	365	19	369	0.086
25	B - C	RM 3.5 min	41	75	6	6.3	366	315	8	341	0.086
26	B - C	RM 3.5 min	41	75	6	6.3	429	330	20	371	0.090
27	B - C	RM 3.5 min	41	75	6	27	416	365	20	355	0.090
28	M 2 *	RM 5 min	41	60	6	2.0	374	375	16	331	0.069
29	M 2 *	RM 5 min	39	60	6	6.2	446	400	30	362	0.067
30	M 2	RM 5 min	39	60	6	2.0	448	380	32	354	0.064
31	M 2	RM 5 min	40	60	6	6.2	373	380	17	325	0.064
32	M 2 *	RM 5 min	41	75	6	2.0	441	365	28	364	0.066
33	M 2 *	RM 5 min	42	75	6	6.2	375	365	14	339	0.066
34	M 2	RM 5 min	41	75	6	2.0	374	335	13	338	0.064
35	M 2	RM 5 min	41	75	6	6.2	445	370	27	375	0.066
36	B - C	RM 5 min	37	75	6	6.4	617	-	70	412	0.091
37	B - C	RM 5 min	37	75	6	6.4	793	-	120	424	0.089
38	B - C	RM 5 min	38	50	6	6.4	569	-	82	372	0.088
39	B - C	RM 5 min	37	50	6	6.4	738	-	120	379	0.088
40-A	B - C	RM 5 min	50	75	6	6.0	425	330	23	380	0.090
40-B	B - C	RM 5 min	50	75	6	6.0	425	390	22	380	0.090
41-A	B - C	RM 5 min	50	75	6	6.0	615	370	80	440	0.090
41-B	B - C	RM 5 min	50	75	6	6.0	615	370	76	430	0.090

* "Weathered" ore (cf text)

** BM ball milled, RM rod milled

*** MG 200 flocculant, Tests 1 - 9 used 0.15 lb/ton, thereafter, used 0.20 lb/ton

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Table 5 - Hazen Research Two Stage Leach Tests, Duplicated

Test No.	Ore Composite	Grind	Temp. °C	Time h.	Sodium Chlorate lb/T	Acid Addition per 400g ore lb/T	Final emf -mV	LEACHING			THICKENING								
								Final Free Acid g/l	Calculated Acid Consumed lb/T	Residue % U ₃ O ₈	Extraction % U ₃ O ₈	Flocculant Type lb/T	Feed % Solids	Overflow U ₃ O ₈ g/l	Underflow Area % Solids	Calculated Acid Consumed lb/T			
3 - A	B - C Head = 0.090 % U ₃ O ₈	RM 5m	50	2	-	1.44 l @ 63.3 g/l	456	20.1	314.0	-	(90)	MG-200	0.21	19.3	0.205	16.7	34	5.7	21.0
3 - B	B - C Head = 0.090 % U ₃ O ₈	RM 5m	75	4	6.0	-	440	88.4	62.3	0.0050	94.4	MG-200	0.22	28	0.213	-	-	-	-
3 - B	B - C Head = 0.090 % U ₃ O ₈	RM 5m	50	2	-	1.44 l @ 63.3 g/l	456	21.4	306.5	-	(90)	MG-200	0.30	19.2	0.231	17.8	37	5.3	20.5
3 - B	B - C Head = 0.090 % U ₃ O ₈	RM 5m	75	4	6.0	-	415	91.3	62.0	0.0052	94.2	MG-200	0.22	13.3	0.219	-	32.3	6.3	-

The first tests (1 - 6) were designed to use the earlier proposed average leaching conditions on the six different drill core composites. The first finding was that 55% solids was impractical because of very high pulp viscosity. Hence these tests, and all but a few special tests, used 38 - 43% solids in leaching. In spite of very high chlorate additions, the proposed -500 mV emf was not obtained except for Composite F. As can be observed from Table 4, a revised standard for oxidation potential of -350mV minimum would appear to be adequate.

The principal variable of the next twenty tests was grind, in its effects on uranium extraction, acid consumption and thickening. Levels of residual free acid to 20 g/l were also tested.

Having settled on one grind (5 minutes in a laboratory rod mill), eight tests (28 - 35) examined possible effects of "weathering" of the ore on leaching response. The idea was to simulate what might occur in stockpiling of mined ore for some months prior to milling. The laboratory weathering procedure consisted in holding batches of -6 mesh ore in trays in an oven at 50 - 55°C for four weeks, while spraying three times per week with water to give peak concentrations of 20% moisture.

The final eight single stage tests were intended to examine the possible advantages of a high residual free acid. Different temperatures were included as well as duplication of tests.

The evident increase in uranium extraction with the higher free acid should be exploitable in mill practise with a two stage leaching circuit. After a favourable preliminary economic analysis of a projected circuit, a duplicated two stage leaching experiment was designed. The experimental results are given in Table 5.

The evaluation of leaching factors to select a recommended flowsheet is summarized in sections 3.4.3 to 3.4.10. An important feature of the evaluations is the reproducibility of reported data. Almost all tests were balanced experimental designs which readily permitted statistical analyses. The most important determination was the reproducibility of U_3O_8 content of leach residue from test to test including all random errors. This was found to be a very satisfactorily low $\pm 0.0009\%$ U_3O_8 , on an individual test with 95% confidence. The same sort of analysis showed ± 10 lb/ton for acid consumption, but a very broad ± 1.6 ft² for thickening tests.

3.4.2 Leach Agitation Tests

HRI had found it necessary to leach at about 40% solids because high viscosity made it too difficult to go higher. However, test data were pointing to the desirability of higher free acid in leaching. A higher percent solids would reduce acid cost. To follow up, Mr. David Daigler, from Denver, conducted leach agitation tests in the HRI laboratories at 50% solids, on behalf of Mixing Equipment Company (Mixco) of Rochester, N.Y.

Mixco recommended as follows:

- (1) both "acid kill" and balance of leaching would be at 50% solids.
- (2) two tanks in series be used to handle the foam and high viscosity of the acid kill step. For tanks of 30,000 gallons, each would use a 125 hp agitator.
- (3) for the remaining leach tanks, if each were 60,000 gallons, a 150 hp agitator would be used in each.
- (4) power required should be ± 25 hp on the quoted figures.
- (5) a lower than one to one height to diameter ratio would be desirable in the acid kill tanks, but otherwise one to one is standard.

If it were desired to drop back to 40% solids in leaching, Mixco estimated the requirements as follows:

- (1) for 37,000 gallons acid kill tanks, use 100 hp agitators.
- (2) for 75,000 gallon normal leach tanks, use 125 hp agitators.

3.4.3 Effect of Grind

The Hazen Research March 1978 report (Ref.3) included considerable screen analysis data both on ore and on leach residues, as well as limited assays on ore screen fractions. In addition, the single stage leach tests reported in Table 4 show the differences in extraction for different nominal grind of leach feed.

It is concluded that a decrease in U_3O_8 content of residue in the range of 0.0004 to 0.0008% U_3O_8 is obtained when leaching under standard conditions for each increment of finer grinding tested, i.e. nominal -20, -28, -35 and -48 mesh. The figure used, say 0.0006% U_3O_8 , would have confidence limits of $\pm 0.0005\%$ U_3O_8 . Hence at \$43.20/lb U_3O_8 , the grinding increments could lead to extra recovery of value \$0.09 to \$0.95 per ton of ore.

No difference in acid consumption, or other leaching response could be attributed to grind.

3.4.4 Effect of % Solids and Retention Time

3.4.4.1 Effects on leaching response

Four of the HRI single stage leaching tests were carried out at 50% solids; the balance were at 40% solids. The available comparisons are very few but they do suggest a decrease in uranium extraction with higher percent solids, possibly because of less efficient mixing. On the other hand, higher percent solids necessarily reduces acid requirements for a given free acid concentration.

Retention time was recommended initially at 10 hours. The current program established that there was no gain in extraction by going beyond 6 hours whereas acid consumption was progressively increased with time. The data, in fact, shows no significant change in residue U_3O_8 between 4 hours and 6 hours but a small additional saving in acid consumption.

3.4.4.2 Capital cost of leaching equipment

For a given ore throughput, leaching tank volume is directly proportional to residence time and to pulp density. The latter is an inverse function of % solids. The total required leaching tank volume will be divided into a number of agitated tanks. Hence, cost of agitators also is a function of residence time and pulp density through their effect on leaching volume. In addition, in the Anderson case, the % solids markedly affects viscosity and the agitation power needed.

For a number of cases that seemed to be likely flow-sheet options, detailed cost estimates were prepared. The lowest cost alternatives of three flowsheet options are shown in Table 6. It is interesting that the high power requirements for agitators at 50% solids causes the total cost of tanks and agitators to be higher in Case A than in Case B. Case C requires much more tankage because of more dilute pulp, hence has the highest cost.

Table 6. Leaching Options, Equipment Costs

<u>Case</u>	<u>A</u>	<u>B</u>	<u>C</u>
Number of stages	one	one	two
First (acid kill) tanks			
% solids	50	40	17
time - h	2	1	1
number of tanks	2	1	1
size of tanks	19' x 22'	21' x 23'	24' x 27'
agitator hp per tank	<u>125</u>	<u>100</u>	<u>60</u>
Balance of one stage tanks and second stage tanks			
% solids	50	40	35
time - h	4	5	5
number of tanks	3	4	4
size of tanks	19' x 22'	21' x 23'	22' x 24'
agitator hp per tank	<u>100</u>	<u>75</u>	<u>75</u>
Total installed cost - \$	<u><u>1,020,000</u></u>	<u><u>962,000</u></u>	<u><u>1,249,000</u></u>

3.4.5 Effect of Leaching Temperature

3.4.5.1 Effect on uranium extraction

Leaching temperature was found to affect uranium extraction in tests at low to moderate free acid level, but not at high free acid level. The residue in the latter instance is already very low. At 15 g/l free acid, an increase from 60°C to 75°C in a 6 hour leach reduced the U_3O_8 by some 0.0008% U_3O_8 . Nearly all leaching was done at 75°C so there are few comparisons. This was chosen as the upper test figure rather than 80°C as used before, in order to ease the problem of materials of construction. The effect on leaching response was expected to be minimal.

3.4.5.2 Effect on acid consumption

Several sets of tests show that acid consumption is affected by temperature. In one instance an increase from 50°C to 75°C increased acid consumption by 30 lb/ton.

The interaction of these opposing responses is evaluated in section 3.4.8.

3.4.5.3 Steam requirements in leaching

The required leaching temperature determines the amount of heat to be supplied by process steam in the leaching circuit. Three factors go to make up net requirements, the total sensible heat from ambient to operating temperature, the convective and other heat losses at operating temperature, and the heat gain from the acid dissolution of the carbonates in the ore. The

heat of reaction has been estimated at $27,670 \times A$ Btu/h where A is lb acid consumed per ton of ore, and ore throughput is 2,000 ton/day.

3.4.6 Effect of Free Acid Level

From leaching of B - C Composite, the following averaged data can be extracted from the Hazen tests, all at nominal -35 mesh, 75°C , 6.0 lb/ton chlorate, for six hours:

% Solids	Free acid - lb/ton			Total Residue acid	% U_3O_8	\$/ton		
	acid g/l	Free acid	Consumed			Total Residue acid	U_3O_8	Total
42	10	28	344	372	0.010	5.95	8.64	14.59
41.5	21.5	61	367	428	0.0075	6.85	6.05	12.90
37	65	221	395	617	0.0055	9.87	4.75	14.62
37	113	385	408	793	0.0045	12.69	3.89	16.58

The costs are those used in the MK study (Ref. 5) of \$43.20/lb U_3O_8 and acid cost of \$30.00/ton of 93% acid.

Increased free acid clearly yields a higher uranium extraction in the range tested. The total cost of acid plus uranium value in residue shows a minimum in the vicinity of 21 g/l free acid. However, a two stage leach should reduce total acid requirements significantly for the high free acid cases, which could change the conclusions. This is examined in section 3.4.8.

3.4.7 Effect of Oxidant

The September 1977 A. H. Ross report (Ref. 4) had proposed 12 lb/ton chlorate as near optimum with a desired emf of -500 mV at the end of leaching. As noted in section 3.4.1, a lesser oxidation potential of about -350 mV appears to be adequate from the current tests. The original criterion was, therefore, abandoned in favour of testing fixed additions of sodium chlorate.

No significant difference in leaching response was found between 6 and 27 lb/ton chlorate over a range of levels of other factors. A few tests with only 2 lb/ton chlorate also gave equivalent leaching.

Although in practise a lower quantity of chlorate may be adequate for some of the ore deposit, as represented by the individual composites, it is deemed to be prudent for design to specify 6 lb/ton for the average condition.

3.4.8 Optimum Single Stage Versus Two Stage Leaching

3.4.8.1 Cases examined

A preliminary economic study of a two stage leaching circuit with high free acid at the end of leaching was made during the course of the Hazen test work. The study was sufficiently encouraging to warrant simulated two stage tests. The test circuit included an interstage thickener because the study had shown better overall economics than the alternative of filters. The Hazen data, shown in Table 5, were incorporated into a complete mill material balance. The essential parameters from the flowsheet balance are as follows:

Master Composite B - C, heads 0.090% U_3O_8
 first stage leach, 2 hr at 50°C
 second stage leach, 4 hr at 75°C
 sodium chlorate to second stage 6 lb/ton

% solids in grinding	55
% solids in first stage leach	16.5
% solids in interstage thickener U'F	35
% solids in second stage leach	32.4
% solids CCD thickeners U'F	38

CCD overflow recycle to first stage leach, ton/ton ore	4.25
pregnant solution (interstage thickener O'F), ton/ton ore	3.21
SX raffinate bleed (15% of pregnant), ton/ton ore	0.48
make-up water to CCD wash, ton/ton ore	1.06
free acid in recycle to first stage leach, g/l	62.8
free acid in pregnant liquor, g/l	20
free acid from second stage leach, g/l	115
free acid in CCD wash, g/l	14.2
final residue % U_3O_8	0.0051
acid consumption, first stage leach lb/ton	310
acid consumption, interstage thickener lb/ton	21
acid consumption, second stage leach lb/ton	<u>62</u>
	393
acid loss in solvent extraction, lb/ton	2
acid loss in raffinate bleed, lb/ton	19
acid loss in CCD tails, lb/ton	<u>57</u>
Total acid make-up required, lb/ton	471

Four cases were examined, three single stage options,
and one two stage option, as follows:

<u>Cases</u>	Leach % solids	Temp °C	<u>Free acid</u>		Acid consumed lb/ton	Total* acid lb/ton	Residue % U_3O_8
			g/l	lb/ton			
Single stage A	45	60	25	61	358	401	0.0099
Single stage B	45	75	25	61	379	422	0.0071
Single stage C	45	50	90	220	374	528	0.0053
Two stage D	1st	16.5	50	} (78)	393	471	} 0.0051
	2nd	34.2	50				

*after recovery of 30% of free acid in CCD for single stage cases

and therefore, not included.

10. maintenance 3.5% of capital/year
11. flocculant for interstage thickener 0.25 lb/ton at \$1.50/lb.
12. electric power balances out between total circuits.
13. solvent loss, single stage = \$243,900/year
two stage = \$183,300/year

3.4.8.3 Cost comparison

<u>Case</u>		<u>Single stage</u>			<u>Two stage</u>
		<u>A</u>	<u>B</u>	<u>C</u>	<u>D</u>
acid	\$/year	4,683,700	4,929,000	6,167,000	5,501,300
steam	\$/year	278,700	488,600	112,600	1,094,100
flocculant	\$/year				273,800
solvent	\$/year	243,900	243,900	243,900	183,300
maintenance	\$/year	149,700	149,700	149,700	194,000
U ₃ O ₈ loss	\$/year	<u>6,244,100</u>	<u>4,478,100</u>	<u>3,342,800</u>	<u>3,216,700</u>
total, operating plus U ₃ O ₈		11,600,100	10,289,300	10,016,000	10,463,200
capital	\$	4,278,100	4,278,100	4,278,100	5,542,200
capital @ 3.5 year	\$/year	<u>1,222,300</u>	<u>1,222,300</u>	<u>1,222,300</u>	<u>1,583,500</u>
total, all items	\$/year	12,822,400	11,511,600	11,238,300	12,046,700

The totalled cost are intended for comparison only and are not meant to be estimates of actual total milling costs. In particular, the value of uranium loss is included.

The two stage leach shows no cost advantage over the single stage leaching cases, in spite of the inclusion of a low residue assay.

Of the single stage leach cases, case A is the most expensive, whereas cases B and C are very close with C the least expensive. However, it was noted previously that the cost of neutralizing excess acid was not included in this case. If a reduction

to 25 g/l in leach discharge were required, which would compare directly to the other cases, about 200 lb/ton of lime would be required. At say \$60/ton, lime would cost about \$6.00/ton of ore, or \$4,380,000/year. A cost of this order, to be added to case C would make it the most expensive case. Consequently, case B has been selected as the preferred case.

3.4.9 Effect of Ore Variability

The six composites which were made up on instructions from Minerals Exploration range in head grade from 0.042% to 0.092% U_3O_8 . When leached under a standard set of conditions uranium extraction ranged from 85 to 95%. Acid consumption ranged from 112 to 500 lb/ton and considerable differences were observed in Hazen settling tests on leached pulp. To make it possible to carry out a reasonable leaching test program, blended composites were employed. The optimum leaching conditions derived herein are based on the blended composites, or average ore.

For operating cost estimates, conditions based on the average ore are employed.

Equipment selection must of necessity be based on "worst" conditions if all ore must be processed at stated throughput rates.

3.4.10 Selected Design Basis for Leaching

In section 3.4.8, single stage leaching at about 25 g/l free acid at the end of leaching was identified as the most economical of four cases chosen for detailed study, which included a two stage leach. In other sections of the leaching discussion, the effects of various factors was quantified so far as possible. The result of these analyses leads to the following recommended design basis, for the average Anderson ore, as represented by the weighted composite (Master Composite 2):

single stage leaching
grind, about 2% + 28 mesh
pulp density in leaching, 45% solids

sodium chlorate 6 lb/ton,

or final emf -350 MV

acid addition to give about 25 g/l free acid at
end of leaching

estimated total acid 415 lb/ton

leaching temperature 75°C

leaching time, 6 hours total including 1 - 2 hours
in acid kill stage.

For Master Composite 2, at approximately 0.066% U_3O_8 , the indicated leaching conditions should give a residue of about 0.0059% U_3O_8 , or about 91% extraction. It was noted in section 3.4.9 that the six composites, when leached under standard conditions yielded uranium extraction in the range of 85% to 95%, hence the behaviour of Master Composite 2 is consistent with these data. It is probable that adjustment of leaching conditions can ensure the 91% extraction on all ores, if they are as represented by the samples received. If the six composites truly represent ore variations which will be milled at different time periods, then acid requirements will vary from 160 to 550 lb/ton.

An important consideration in choosing a compatible set of leaching conditions for the Anderson case, was to avoid the commitment of too much capital expense to gain an estimated reduction in operating cost. This consideration is particularly valid where the "operating cost" includes the theoretical value of uranium in the leached residue. On the other hand, in leaching, the choice of 6 hours retention time was selected to give a safe volume of tankage so that other variables could be reduced in attempting to optimize the circuit. This sort of flexibility seems essential with the variability of the Anderson ore.

3.5 Liquid - Solid Separation

3.5.1 Hazen Research Test Data

Tests by equipment suppliers have been used for design purposes. Nonetheless, Hazen Research conducted standard cylinder settling tests and a Kynch calculation on almost all leached pulps, and their data is believed to be useful on a comparative basis between leaching tests.

The six different ore composites showed a wide variation in thickening (Table 4) but the low level of flocculant used tends to obscure the differences. Nonetheless, from these unit area observations, composites B and C were selected as representing poor thickening behaviour, and a combined B - C composite was made up for general testing.

The fineness of grind entered to leaching was shown to have an effect on thickening, but the poor reproducibility of calculated thickening area makes it most difficult to assign hard numbers. Nonetheless, it was concluded that there was an increase in required unit area of $0.8 \text{ ft}^2/\text{ton}/\text{day}$ in going from nominal -28 to -35 mesh. For a five stage CCD circuit, this difference in area would cost about \$700,000 for Anderson. On a 3.5 year amortization, it would be about \$0.26/ton of ore. The lack of precision is such that the cost should really be quoted as 0 to \$0.52/ton of ore for the change in nominal screen size. A few tests suggested that higher leaching temperature would require more thickening area but the data was not conclusive.

3.5.2 Enviro-Clear Thickening Tests

Thickening tests were carried out by Deltech, Inc. Denver, for the Enviro-Clear Division, Amstar Corporation (Enviro-Clear). Some 42.6 kg of ore were leached by HRI for these tests. Composite B - C at nominal -35 mesh was leached at 75°C for 6 hours at 41% solids, with 6.7 lb/ton chlorate added, and a residual free acid of 17 g/l.

The Enviro-Clear letter report of March 1, 1978 (Ref. 6) proposes design conditions as follows:

Unit area	0.7 - 0.8 ft ² /ton/day
Flocculant	0.25 lb/ton (stage 1)
Overflow	50 ppm solids
Underflow	30% solids

However, the letter indicates some dissatisfaction with the data generated, and suggests that optimum conditions were not reached.

An underflow concentration of 30% solids is not acceptable for design purposes. Because of the brevity of the Enviro-Clear report, and the absence of back-up analytical and correlating data, it is not possible to analyze the parameters proposed by Enviro-Clear or to determine design requirements for obtaining higher underflow densities.

Enviro-Clear thickeners are still considered a potentially viable option, but further testwork would be required to generate adequate design information.

3.5.3 Eimco Thickening Tests

3.5.3.1 Introduction

Both thickening and filtration studies were

conducted on leached ore pulp by the Eimco Process Machinery Division, Envirotech Corporation, Salt Lake City (Eimco). Two batches of ore were leached for these tests which extended into two weeks. B - C composite was used at nominal -35 mesh. Other conditions were 75°C, 6 hours, 42% solids, 6.6 lb/ton chlorate added and residual free acid in one batch of 18 g/l and in the other of 23 g/l.

3.5.3.2 Method of test analysis

Eimco have analysed the settling test data using their newly-developed Direct Underflow (DUF) method, rather than the more conventional Kynch method. Although the DUF method has some attractive features, there has been insufficient experience with it to warrant its use for a design basis at the present time. Consequently, Kynch analyses have been performed, which provide the basis for A. H. Ross and Associates design recommendations.

3.5.3.3 Flocculant choice

Scoping tests were conducted on a number of flocculants. Although flocculant type was not optimized, Dow MG 200 performed well and was chosen for use in the subsequent settling tests.

3.5.3.4 Effect of feed density

Qualitative observations indicated that better flocculant mixing was obtained at lower feed densities but a dependence of settling rate on feed density cannot be identified from the data.

3.5.3.5 Terminal densities

The average terminal density of all standard cylinder tests was 43.5% solids. Individual values ranged from 38.4% to 46.4% solids.

The "deep tube" test yielded 47.1% solids. Eimco recommend that 45% solids be considered as an upper limit due to a potential for solids buildup on the raking mechanism, and probable difficulty in pumping.

Densities of 38% and 40% solids were chosen for evaluation purposes.

3.5.3.6 Summary of test results

Table 7 summarizes the test data and indicated unit areas at 38 and 40% solids as determined by Eimco and A. H. Ross & Associates. It may be noted that the Eimco DUF method and the Kynch method give rise to quite different calculated unit areas in several of the tests.

Table 7 Thickener Unit Area Estimates

Stage	Test	Flocculant lb/ton	Feed % solids	Terminal density, % solids	Unit area, ft ² /ton/day			
					38% solids		40% solids	
					Eimco ⁽¹⁾	A.H.R. ⁽²⁾	Eimco	A.H.R.
1	1	0.187	18.6	44.6	5.7	4.6	9.5	5.3
	2	0.136	19.2	45.2	7.7	6.5	10.5	7.4
	3	0.301	17.4	40.0	13.0	6.5	82.0	7.2
	4	0.253	17.4	44.5	4.3	4.2	6.5	4.9
	5	0.236	21.1	44.1	4.0	-	7.5	-
	6	0.252	10.2	47.1	4.0	4.9	7.5	5.6
	7	0.242	26.4	46.4	4.0	-	7.5	-
	8	0.252	14.9	44.4	4.0	4.5	7.5	5.3
2	9	0.140	18.8	43.9	7.0	6.1	12.2	7.0
	10	0.182	18.7	42.1	8.7	7.1	17.0	8.5
	11	0.250	17.4	38.4	95.0	6.2	>100	7.1
	12	0.177	17.5	38.8	33.0	5.6	>100	6.9
	13	0.163	21.4	43.6	4.6	4.5	8.5	5.0
	14	0.180	10.2	41.7	4.6	5.6	8.5	7.0
	15	0.170	26.5	45.1	4.6	-	8.5	-
	16	0.180	14.9	43.5	4.6	5.0	8.5	5.6
3	17	0.047	18.8	44.9	8.5	7.1	11.9	8.0
	18	0.091	19.3	45.4	6.9	5.9	12.2	6.9
	19	0.025	17.3	41.6	14.5	8.6	25.0	9.2
	20	0.076	17.5	40.3	22.0	11.4	87.0	12.8
	21	0.069	21.3	45.0	5.0	5.3	7.5	6.1
	22	0.072	10.2	43.6	5.0	8.8	7.5	10.7
	23	0.072	26.5	45.4	5.0	-	7.5	-
	24	0.078	15.0	44.7	5.0	6.2	7.5	7.0
1	25	0.191	16.5	47.7	4.0	-	4.0	-

(1) Direct underflow method

(2) Kynch method. Includes a 25% scale-up factor.

3.5.3.7 Selection of optimum design conditions

Selection of optimum design conditions is influenced by a number of variables which affect capital and/or operating costs. The conditions and variables tested are listed as follows:

leach discharge density	45% solids
flocculant consumption	from test data
underflow densities	38 and 40% solids
thickener unit areas	6.5, 7, 7.5, and 8 ft ² /ton/day
pregnant/ore ratios	2.5 to 5.5
maximum allowable rise rate	0.1 gal/min/ft ²

The maximum allowable rise rate is estimated from the test data, and sets an upper limit for the pregnant/ore ratio.

In relating operating costs to capital costs, a simple payback period of 3.5 years was utilized.

Selected optimum design conditions are as follows:

unit area, ft ² /ton/day	7.0
thickener diameter, ft	140
pregnant/ore ratio	4.25
underflow density	38% solids
number of stages	5
flocculant consumption, lb/ton,	
stage 1	0.15
2	0.09
3	0.06
4,5	0.036
soluble loss, %	1.75

The Eimco report (Ref. 7) concluded that a unit area of 7.5 ft²/ton/day with 40% solids underflow density would be suitable for design. The A.H.Ross economic comparison, based on 38% solids underflow density led to the 7.0 ft²/ton/day as tabulated.

3.5.4 Eimco Filtration Tests

3.5.4.1 General

The testwork was conducted to provide design information suitable for both drum filters and belt extractors. Although the general test method used is intended for the design of a top feed (belt) filter, it was found through comparative tests to be applicable to drum filters also.

Wash efficiency data on a number of tests is based on lithium assays, where solutions had been previously spiked with a lithium compound. Although uranium analyses would have been preferred, the lithium based data have been used. The correlation between lithium and uranium based results appears to be close enough that any error introduced via use of the former should not have a significant effect on the choice of design conditions.

3.5.4.2 Flocculant

Scoping tests were conducted on various flocculants. Although flocculant type was not optimized, Jaguar MDD was selected for use in the testwork.

Eimco found that good mixing of flocculant with the pulp was difficult to achieve and that prior dilution of the feed pulp was necessary.

Provision of mechanically agitated flocculant mixing tanks may be required on plant scale.

The dilution of feed pulp mentioned above yielded better form filtration rates than undiluted pulp, which Eimco attributes to the better flocculant mixing. This feature is included in the filter circuit flowsheets evaluated by Eimco and by A. H. Ross & Associates.

3.5.4.3 Cake weight

Filter cake density is very low at 44 - 45 lb/ft³. The reason for the low density has not been determined, but may well be from pockets of the gas which is generated copiously in leaching.

3.5.4.4 Cake cracking

Cracking of the filter cake was found to occur after very short dry times of 0.05 to 0.10 minutes. As Eimco have noted, the prevention of cake cracking on drum filters would require the use of mist sprays to keep the cake surface damp. As the liquid supply would be primarily recycled acidic solutions, the filters would require hooding. This would be a relatively expensive and operationally awkward addition. In theory cake cracking should be a lesser problem with belt extractors, although it has not been adequately established in practice.

3.5.4.5 Pulp aging

Filtration rate was found to decrease as the time between leaching and filtering increased up to one hour. Beyond one hour, no significant change was detected. Testwork was conducted on pulp aged for over one hour.

3.5.4.6 Cake moistures

Cake moistures obtained in the tests were high, in the general range of 46 - 50% liquor.

3.5.4.7 Temperature and vacuum

The data developed is relative to a pulp and wash solution temperature of 40°C, and a vacuum range of 18 - 22 inches Hg.

3.5.4.8 Drum filter sizing

Full scale filtration rates (FSFR) were calculated from the data and compare with Eimco's figures as follows:

	<u>Eimco</u>	<u>A.H. Ross & Associates</u>
Stage 1: wash ratio	1.39	1.0
FSFR, lb/hr ft ²	12.5	23.8
flocculant, lb/ton	0.65	0.65
Stage 2: wash ratio	1.0	1.0
FSFR, lb/hr ft ²	15.7	17.0
flocculant, lb/ton	0.37	0.37
Stage 3: wash ratio	1.0	1.0
FSFR, lb/hr ft ²	16.3	17.0
flocculant, lb/ton	0.28	0.28

Wash rates determined filter sizing for all the above cases. After adjustment of Eimco's stage 1 rate to correspond to a 1.0 wash ratio, a comparison shows that Eimco have interpreted the data somewhat more conservatively than A. H. Ross & Associates for all three stages.

3.5.4.9 Belt extractor sizing

The testwork indicates that filtration rates are higher at a 1/4 inch cake thickness than at a 3/8 inch cake thickness, and 1/4 inch was therefore chosen for design purposes.

Calculated filter sizings are given below, with an Eimco example calculation for comparison. Flocculant consumption is 0.48 lb/ton in all 3 cases.

	<u>Eimco</u> <u>3 stage wash</u>	<u>A.H. Ross & Associates</u> <u>3 stage wash</u> <u>4 stage wash</u>	
form, min	0.15	0.16	0.16
dry, min	0.02	0.05	0.05
wash, min	0.295 ⁽¹⁾	0.20	0.20
dry, min	0.02	0.05	0.05
wash, min	0.255	0.20	0.20
dry, min	0.02	0.05	0.05
wash, min	0.255	0.20	0.20
dry, min			0.05
wash, min			0.20
final dry, min	<u>0.046</u>	<u>0.29</u>	<u>0.29</u>
total min	1.061	1.20	1.45
FSFR, lb/hr ft ²	41.9	37	31

(1) Eimco wash times cannot be directly compared with A. H. Ross & Associates' figures, since wash volumes are not equivalent.

As an inspection of the above table shows, the essential difference between Eimco and A. H. Ross & Associates interpretations is in the allowance for intermediate and final dry times. Eimco have chosen to minimize dry times due to the tendency of the cake to crack. A. H. Ross & Associates agree in principle, but have used the longer dry times to provide more flexibility,

and a more conservative filter sizing.

3.5.4.10 Wash Efficiency

Wash efficiency is determined as a function of wash ratio, and data is plotted on linear or semi-log paper as R (fraction remaining in cake after washing) vs n (wash ratio).

For first stage washing, Eimco have chosen design curves which can be defined as follows:

Extractors: $R = 0.186$ at $n = 1.0$

Drums: $R = 0.44$ at $n = 1.0$

In the case of extractors, R is increased by 0.05 for each subsequent wash stage. Wash efficiencies on second and third stage drums are unchanged.

Eimco have chosen a low efficiency for drum filter washing because of the tendency for cake cracking. A. H. Ross & Associates have used the same factor for wash calculations although it is thought to be possibly overly conservative. On the other hand, the Eimco efficiency factor used in this instance for belt extractors is believed to be overly optimistic for design purposes, based on operating performance information available to A. H. Ross & Associates.

A. H. Ross & Associates have consequently chosen an R factor for the extractors of 0.30 based on a wash ratio of 1.0.

For all stages after the first, the Eimco factors have been used.

3.5.4.11 Drum filter flowsheet

Three potential filter circuits were analysed. All employed dilution of feed solids to 30 - 32% for each stage, and wash ratios of 1.0 on each stage:

<u>Option</u>	<u>Number of stages</u>	<u>Pregnant/ore ratio</u>	<u>*Capital cost \$</u>	<u>Annual operating cost \$/year</u>
1	2	2.66	15,390,000	3,461,000
2	3	2.12	22,909,000	4,028,000
3	3	3.56	23,499,000	3,493,000

*Capital and operating costs have been derived as in Section 3.4.8.2. Value of uranium loss is included.

On the above basis, 3 stage filtration was not considered further.

3.5.4.12 Belt extractor flowsheet

Three potential filter circuits were analysed. All employed wash ratios of 1.0 (relative to final cake moisture) and feed densities of about 32%.

<u>Option</u>	<u>Number of stages</u>	<u>Pregnant/ore ratio</u>	<u>*Capital cost \$</u>	<u>Annual operating cost \$/year</u>
1	3	2.07	12,241,000	2,116,000
2	4	2.07	14,016,000	1,879,000
3	4	2.93	14,385,000	1,783,000

*As for drum filters, costs are derived as in Section 3.4.8.2.

Payback of option 2 over option 1 is 7.5 years.

Payback of option 3 over option 1 is 6.4 years.

Option 1 was selected for further comparisons.

3.5.5 Selection of Flowsheet

The three selection sets of conditions for thickeners, drum filters and belt extractors are compared below:

<u>Description</u>	<u>Capital cost \$</u>	<u>Operating cost \$/year</u>
CCD 5 stages	10,112,000	1,713,000
Drum filters 2 stages	15,390,000	3,450,000
Belt extractors 3 wash stages	12,241,000	2,116,000

On the above basis, a conventional CCD circuit of 5 stages is recommended. The details of the circuit are given in section 3.5.3.7.

3.6 Solvent Extraction

3.6.1 Cyclic Testing

Some 475 litres of leach solution were passed through a bench scale continuous solvent extraction unit at Hazen Research, Inc. The organic solvent was processed through loading and stripping for a total of about 12 cycles. Separation of organic and aqueous phases was good and uranium recovery was high.

For the last six days of the cyclic testing, assays of U_3O_8 in g/l were reported as follows:

<u>Date</u>	<u>Feed liquor</u>	<u>Raffinate</u>	<u>Loaded organic</u>	<u>Barren organic</u>	<u>Pregnant strip liquor</u>
Jan 31	0.149	0.0005	1.68	0.38	14.4
Feb 1	0.149	0.0009	1.92	0.46	13.0
Feb 2	0.149	0.0008	1.75	0.13	17.9
Feb 3	0.149/0.166	0.0005	1.87	0.17	20.8
Feb 6	0.166/0.168	0.0004	2.00	0.17	24.8
Feb 8	0.168/0.173	0.0005	2.04	0.10	24.6

A small amount of crud, thought to be derived from inadequately clarified feed solution, collected in the first extraction stage. A somewhat greater amount of sludge, thought to be caused by the presence of zirconium, was formed in the stripping section.

Mixing time, settling areas, and loading and stripping isotherms were found to be within normal values. Design values were selected consistent with practice as modified according to the test results.

3.6.2 Extraction Rate

Rate of transfer of U_3O_8 was measured when aqueous and organic solutions were mixed together. Equilibrium was attained in less than 0.25 minutes.

3.6.3 Extraction Settling Rate

Rate of organic/aqueous disengagement was measured by noting the width of the mixed phase band at various solution flowrates. The data so obtained were compared to that from an operating plant.

3.6.4 Stripping Method

Stripping by ammonia-ammonium sulfate was selected for the test work. This method is economically competitive with the other widely used process - salt stripping - and avoids the problem of meeting product sodium specification and the intensive product washing required for chloride removal. Interfering elements, such as molybdenum, which are not compatible with the ammonia-ammonium sulfate system were not encountered.

3.6.5 Stripping Rate

U_3O_8 equilibrium between the organic solvent and stripping solution was rapidly attained.

3.6.6 Strip Settlers

Settling rates as determined by emulsion band-flow rate measurement were satisfactory and a normal settling area was provided.

3.6.7 Regeneration

Sodium carbonate is most frequently used when regeneration of the organic solvent is required. However, in the HRI tests, a very slow settling emulsion was formed with sodium carbonate whereas a sodium bicarbonate solution performed satisfactorily.

Settling requirements were similar to the stripping process and an equally sized mixer-settler was chosen.

3.6.8 Sludge

The sludge formed in extraction was only of a minor amount and was considered to have been caused by incomplete solution clarification.

The sludge formed in stripping was somewhat more abundant and was thought to have been caused by the presence of zirconium. Means will be required (e.g. a self priming pump) to move such sludge to the regeneration stage, or alternatively, to the sludge treatment vessel. From other work, a zirconium sludge has been found to break when treated with either acid or soda ash.

The proper route depends on the amount of such sludge actually encountered and the disposition of the zirconium when the accompanying uranium is precipitated. It is considered that the subject can be left open and answers obtained if and when sludge becomes a problem.

3.7 Uranium Precipitation

A composite of the strip solution produced during the last sixty hours of the cyclic solvent extraction test was precipitated with ammonia to produce yellowcake. Reagent consumption, physical characteristics, and product purity were measured.

Precipitation was carried out at 40°C to pH 7.6 using aqueous ammonia, then settled and decanted. Thickened slurry was filtered then washed successively with dilute ammonium sulfate, and ammonia solution. The final cake was dried at 90 - 100°C.

Precipitation was virtually complete. Ammonia consumption was 0.18 lb/lb U_3O_8 . The product settled and filtered well. The dried yellowcake contained 87.2% U_3O_8 , 0.82% sulfate, 0.049% Mo and <0.01% V_2O_5 . These and all impurity analyses were within the concentrate feed specifications of Allied Chemical Company and Kerr-McGee Nuclear Corporation.

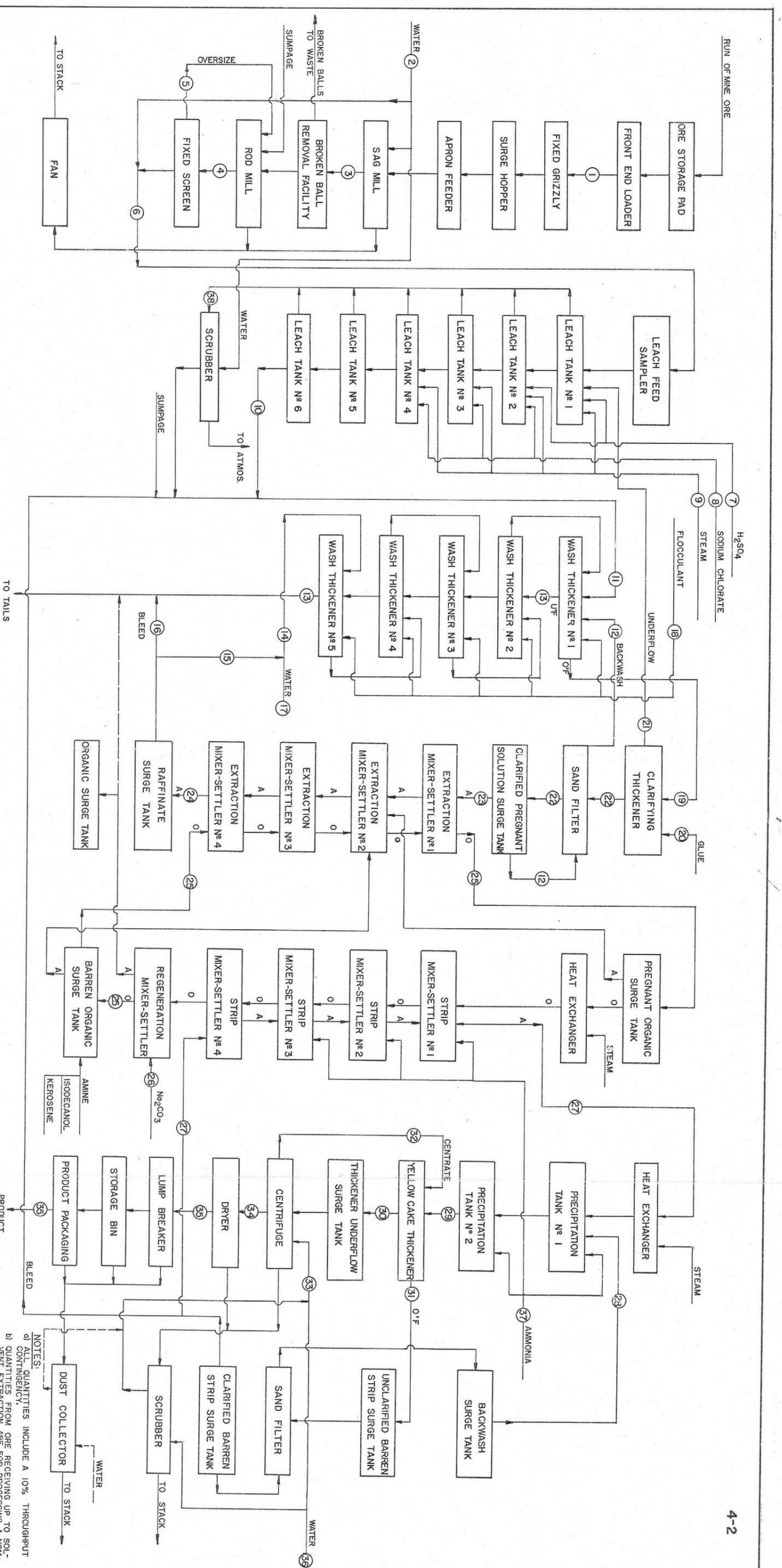
4. FLWSHEETS

4.1 Process Flowsheet

Drawing 4.1 presents the proposed Anderson mill process flowsheet in block form with the associated solids, liquid and total pulp flow for the principal streams. In order to avoid difficulty in following the flowsheet, the alternative routing of certain streams are not shown. These are, however, noted in the appropriate sections under Design Criteria. In particular, this applies to the clarifying thickener underflow, the sand filter backwash, and the aqueous underflow from the barren organic surge tank.

4.2 Yellowcake Washing Circuit-Soluble Sulfate Balance

The block flowsheet and material balance presented in 4.2 are referenced in Section 5.9 of the Design Criteria.



STREAM NUMBER	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23
SOLIDS ton/h.	91.67	-	91.67	110.00	18.33	91.67	-	91.67	-	91.67	91.67	-	91.67	-	-	-	-	-	-	-	-	-	-
LIQUID ton/h.	10.18	74.37	75.00	90.00	15.00	84.55	23.47	0.69	7.97	112.04	112.68	15.58	149.57	424.10	331.16	58.44	92.94	5.65	408.44	0.64	3.90	405.18	389.60
TOTAL FLOW gal/min.	194.5	297.5	453.4	544.0	90.7	491.5	51.6	2.1	31.9	601.3	603.9	62.3	751.2	1696.4	1324.6	233.8	371.8	22.6	1663.8	2.6	15.6	1820.7	1538.4
STREAM NUMBER	24	25	26	27	28	29	30	31	32	33	34	35	36	37	38	39	40	41	42	43	44	45	46
SOLIDS ton/h.	-	-	-	-	-	0.10	0.10	Trace	Trace	Trace	0.10	0.10	-	0.03	-	-	-	-	-	-	-	-	-
LIQUID ton/h.	389.60	41.20	0.40	330	0.23	3.53	0.20	4.50	1.17	1.04	0.05	-	1.04	-	11.53*	-	-	-	-	-	-	-	-
TOTAL FLOW gal/min.	1558.4	164.8	1.6	13.2	0.9	14.1	0.8	18.0	4.7	4.2	0.2	-	4.2	-	-	-	-	-	-	-	-	-	-

*) 854 CO₂ plus 3.0 H₂O, Net Loss is CO₂ only.

d) LEACHED ORE WEIGHT TAKEN AS EQUAL TO UNLEACHED ORE WEIGHT.

e) GAL. REFERS TO U.S. GALLONS.

f) QUANTITIES ARE BASED ON 24 HOUR DAY, SPECIFIC GRAVITY OF ORE OF 2.38 EXCEPT SULPHURIC ACID AND SODIUM CHLORATE OF 1.02.

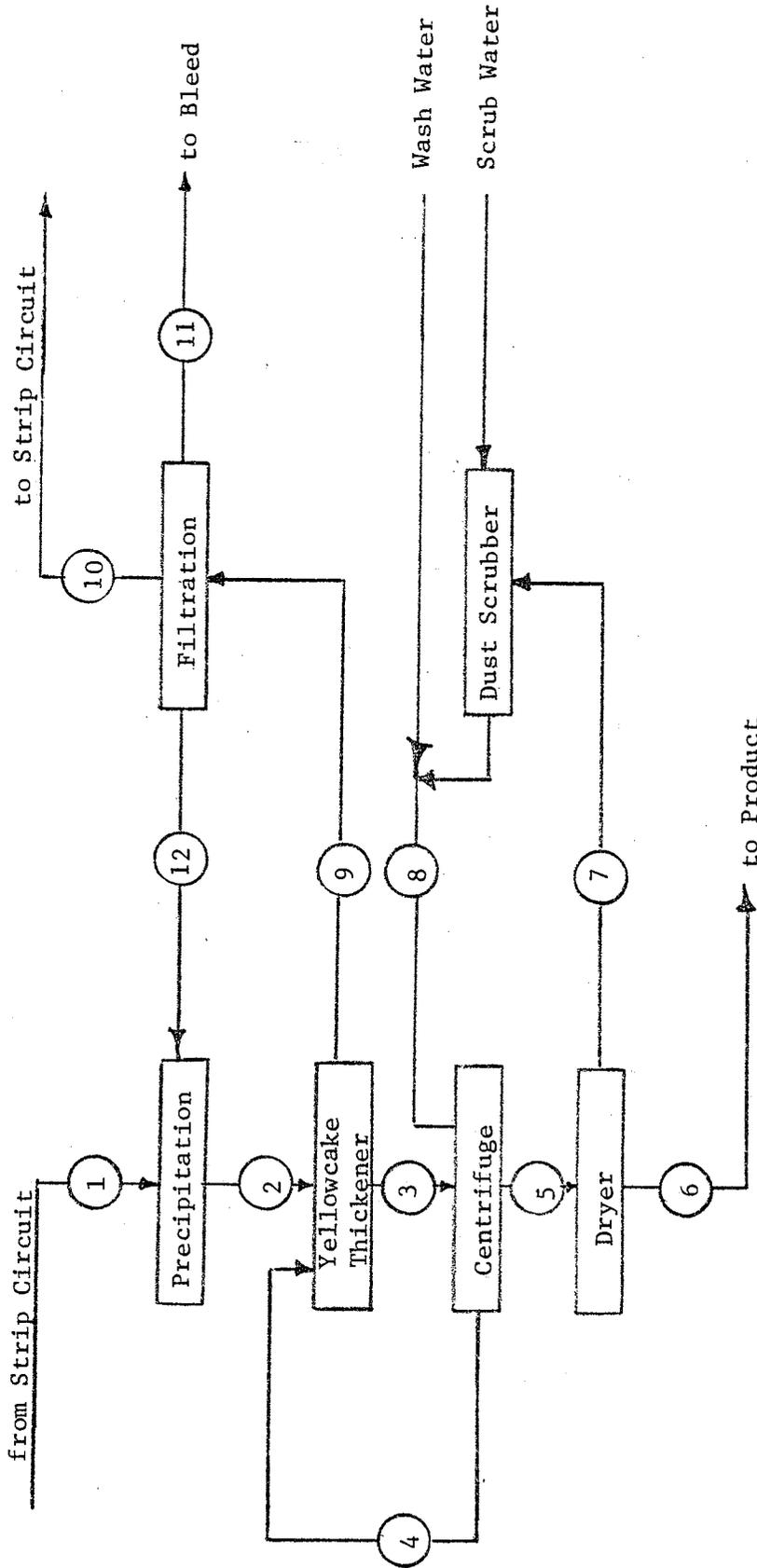
A. H. ROSS & ASSOCIATES
 CHEMICAL & METALLURGICAL ENGINEERS
 TORONTO - CANADA

MINERALS EXPLORATION COMPANY
 ANDERSON URANIUM MILL
 PROCESS FLOWSHEET

DATE: MAY 1978
 DRAWN BY: M.E.G.
 DRWG. No. 4.1

REVISED: MAY 23, 1978

4.2 Yellowcake Washing Circuit - Soluble Sulfate Balance



4 - 3

Stream:	1	2*	3	4	5	6	7	8	9	10	11	12
Solids:												
Yellowcake lb/day	4702	4754	4754	95	4754	4659	95	95	43			43
U ₃ O ₈ lb/day	3997	4041	4041	81	4041	3960	81	81	37			37
Liquid:												
Flow ton/day	79.20	84.64	4.75	28.15	1.58			24.98	108.03	79.20	23.39	5.44
Sulfate lb/day	22,626	23,626	1037	982	55	54			23,570	17,280	5,104	1186
Sulfate g/kg.	142.8	139.6	109.1	17.4	17.4				109.1	109.1	109.1	109.1

* 186 lb/day of sulfate is precipitated as insoluble sulfate.

5. Design Criteria5.1 Millsite Conditions

Site	Owner's property in Yavapai County, Arizona
Elevation	Approximately 1900 feet above sea level
Temperatures	Air minimum 28°F
	maximum 112°F
	Mill water 65° - 70°F
Ore	similar to air

5.2 Ore Characteristics

Specific gravity	range: 2.28 to 2.47
	design average: 2.38
Bulk density of as-mined ore compacted	110 lb/ft ³ on dry basis
	loose 100 lb/ft ³ on dry basis
Moisture	mine run 15% maximum
	design mill feed 10%

Ore composition, from weighted drill core assays

	<u>% on dry basis</u>
U ₃ O ₈ - fluorimetric	0.067
- volumetric	0.070
V ₂ O ₅	0.100
Mo	0.008
CO ₂	5.25
PO ₄	0.15
Cl	0.01
S	0.81

5.3 General Production Criteria

Operating days per year		365
Ore feed rate (dry basis)		
average		2000 ton/day
design		2200 ton/day
average annual rate		730,000 ton/year
Ore grade	average	0.069% U_3O_8
Overall recovery		88.5%
Uranium losses insoluble, in residue		9.0%
soluble, in CCD		1.75%
in solvent extraction		0.25%
unaccounted		0.5%
Uranium production,		
average		2443 lb/day U_3O_8
design of solvent extraction and yellowcake facilities (based on 2200 ton/day ore, 0.10% U_3O_8 , 90% recovery)		3960 lb/day U_3O_8
average U_3O_8 per year		891,690 lb

5.4 Ore Receiving and Grinding5.4.1 Design Basis

Ore pad	levelled area for ore storage in 30 ft high piles.
Operating schedule	24 hours per day, 7 days per week
Feeding method	front-end loader from stockpile
Ore feed rate	91.7 dry tons per hour
Feed size	minus 18 inch run-of-mine ore
Product size	98% passing 28 mesh (Tyler) minimum practical minus 325 mesh.
Pulp density, grinding circuit	55% solids
Type of grinding circuit	semi-autogenous primary mill (SAG mill) in open circuit followed by a rod mill in closed circuit with a scalper screen.

5.4.2 Equipment

Receiving hopper grizzly	18 inch by 18 inch grid openings
Ore feeder	variable speed apron feeder
Semi-autogenous mill	
diameter	16 feet
length	5 feet
motor	450 hp with vari-speed drive
ball load	up to 8% of mill volume
Rod mill	
diameter	9.5 feet
length	16 feet
motor	600 hp with vari-speed drive
Scalper screen	
type	sievebend, such as DSM screen
size	1 @ 6 ft wide, or 2 @ 4 ft wide
Circulating Load	up to 20% of throughput, on harder ore types.

5.4.3 General Instructions

- (1) Provision is to be made to contain and treat rainwater carry-off from the ore pad, and to minimize dusting by winds.
- (2) Facilities should be included for removing or breaking oversize ore that will not pass the grizzly.
- (3) A magnet and metal detector are to be installed on the conveyor belt feeding the SAG mill.
- (4) An integrating weightometer is required on the conveyor belt feeding the SAG mill.
- (5) Variable speed drives should be installed on the SAG and rod mills.
- (6) A fan is to be provided to vent the grinding mills at the discharge ends for removal of radon gas.
- (7) Suitable dust suppression and collection equipment is to be provided.
- (8) A ball storage area and equipment for adding balls to the SAG mill are to be provided.
- (9) A rod storage area and feeder are to be provided; an electrically-powered rod charger is preferred.
- (10) Provision is to be made for preventing the undersize balls from the SAG mill entering the rod mill.
- (11) Space should be provided for possible future installation of agitated pulp surge capacity ahead of leaching, if it should become desirable to accommodate variations in ore grindability by varying mill feed rate, rather than or, in addition to, speed variation of the grinding mills.

5.5 Leaching5.5.1 Design Basis

Pulp density

for tank sizing	45% solids (w/w)
for agitator design	50% solids (w/w)

Temperature

75°C

Free acid in leach discharge

20 - 25 g/l

emf at end of leaching

-350 mV

Sodium chlorate consumption

- average	6 lb/ton
- range	4 - 12 lb/ton
- solution strength	40%

Acid consumption

- average	415 lb/ton
- range	150 to 600 lb/ton

Retention time

- acid kill stage	1.5 hours
- remaining vessels	4.5 hours

5.5.2 Equipment

Leach tanks

total 6 in series

first two tanks

diameter 18 ft

height 20 ft

remaining four tanks

diameter 20 ft

height 22 ft

5.5.3 General Instructions

- (1) The first two leach tanks must contend with high pulp viscosity and foam. The tanks are to be 18 ft diameter by 20 ft high with a nominal pulp depth of 13 ft. These tanks will provide about 1.5 hours retention time at design flow rate.
- (2) The remaining four leach tanks are to be 20 ft diameter by 22 ft high. Pulp depth is to be 18, 19, 20, and 20 ft in these four tanks in succession with the greatest depth at the leach discharge.
- (3) The tank outlets are to be baffled to overflow pulp from the lower portion of a tank to the following tank. It is recommended that sloped covered launders be used for the overflow pulp between tanks because of high viscosity and foaming properties of the pulp. This will be necessary at the head of the circuit and optional thereafter.
- (4) Provision is to be made to by-pass any leach tank.
- (5) Mechanical agitators will be used in all leach tanks. The low height to diameter ratio in the first tanks plus relatively high power requirements will need special consideration in design and selection of agitation equipment.
- (6) Rubber wear pads are to be installed on the inside tank bottom of each leaching vessel beneath the agitator impellor. A suitable choice is 1/2 inch thick pure amber rubber with a durometer rating of 35.
- (7) Provision is to be made for heating of the leach pulp to 75°C in the second of the first two leach tanks in series by sparging of live steam into the two tanks. In the event that it is desired to operate the first two tanks at a lower

temperature, provision is to be made to supply all of the necessary steam to reach 75°C in the fourth vessel in series by sparging of live steam into the third and fourth vessels.

- (8) The acid required in leaching is to be added in the first two leaching vessels. The requirement may be as high as 600 lbs of acid per ton of ore. At least four addition points are necessary in each of the first two tanks to minimize local over-concentration of acid. The acid lines are to terminate above the pulp level and be sufficiently distant from the tank walls to prevent damage to the tank.
- (9) Provision is to be made to add sodium chlorate solution to the second, third and fourth leach tanks. The chlorate will be added as a 40% solution. Feed lines are to terminate above the pulp level.
- (10) All distribution piping is to be located above the leach tanks.
- (11) All leach tanks are to be covered and vented through a common wet scrubber. The covers must be able to support the weight of two heavy persons, and should be adequately protected against corrosion.
- (12) The leach tanks may be insulated and located outdoors.
- (13) The leach tank drain lines should be suitable for connection to a pump.

5.6 Countercurrent Decantation5.6.1 Design Basis

Number of thickener stages	5
Unit thickener area	7.0 ft ² /ton/day
Flocculant	
type	Dow MG - 200
dosage, first thickener	0.15 lb/ton ore
total to CCD	0.37 lb/ton ore
concentration, stock solution	0.3%
concentration, to thickeners	0.03%
Underflow pulp density	
for flowsheet	38% solids
for pump design	35% solids
Pregnant liquor	1560 gal/min
Wash solution, raffinate recycle water	65% - 100% of total raffinate balance

5.6.2 Equipment

Thickeners

number	5
diameter	140 feet
mechanism	centre pier

Underflow pumps

operating	diaphragm
spares	centrifugal

5.6.3 General Instructions

- (1) Normal pumping of thickener underflow is to be carried out with diaphragm pumps, with each pump having capacity to handle full flow at 35% solids.
- (2) Centrifugal pumps are to be provided as spares, one for each diaphragm pump.
- (3) Use two underflow lines per thickener, one for each pump.
- (4) The diaphragm pumps are to be located to give minimum suction lift.
- (5) CCD interstage pumping is to be accomplished by directing the diaphragm pump discharge and the thickener overflow solution into a pump box, with subsequent advance of the mixture to the next stage, with a centrifugal pump. The pump box should be as tall as possible to permit a variable head on the pump to compensate for flow fluctuation.
- (6) Piping is to be arranged so that any thickener may be bypassed.
- (7) Piping and pumps transporting thickener feed and overflow are to be designed to accommodate a maximum pregnant solution to ore ratio of 5.5:1.
- (8) Motorized rake lifting mechanisms and remote torque indicators are to be provided for each thickener.
- (9) Stub arms are not to be used in thickener mechanisms.
- (10) Thickeners are to be provided with underflow tunnels with good drainage characteristics; they can be about

7 ft high by 4 ft wide.

- (11) The thickeners general arrangement should minimize the length of suction lines to the diaphragm pumps.
- (12) The thickener tank walls should extend one foot above the solution level to decrease wave action due to winds and prevent solution loss over the wall.
- (13) The piping arrangement should permit recycle, advance and bypass of underflow solids plus recycle, advance and bypass of overflow solution.
- (14) The minimum recommended slope for thickener feed launders is $3/8$ inch per foot.
- (15) As most flocculant solutions degrade with time, it is advantageous to design for daily mixing of fresh flocculant. This also applies to glue make-up.
- (16) In order to achieve adequate flocculation, thickener feed, flocculant solution, and diluent solution should be mixed in a mechanically agitated tank for each thickener stage.
- (17) Provision should be made for staged flocculant addition (not less than 5 points) along the thickener feed launder, with a final addition into the thickener well.
- (18) The flocculant stock solution will be diluted before addition to the pulp. Dilution will be with wash solutions for thickeners 3, 4 and 5, and with No. 2 overflow solution for thickeners 1 and 2.

5.7 Clarification5.7.1 Design Basis

Pregnant liquor flow	1560 gal/min
Clarifying thickener	
feed solution solids	150 - 250 ppm
overflow solution solids	35 - 80 ppm
underflow pulp	5% solids, maximum
Flocculant	
type	glue
dosage	0.10 lb/ton
concentration	7 g/l
Filtered clarified solution	
solids content	10 ppm

5.7.2 Equipment

Clarifying thickener-surge tank	
diameter	70 feet
surge working volume	375,000 gallons
Clarifying filters	
type	downflow pressure sand filters
number of units	2 operating, 1 spare
size	11 feet diameter
Clarified pregnant solution tank working volume	375,000 gallons

5.7.3 General Instructions

- (1) Underflow from the clarifying thickener surge tank is to be directed to the first leach tank with provision to bypass to the first CCD thickener.
- (2) The surge capacity in the clarifying thickener is to be provided by additional tank volume above the solution withdrawal point. Solution withdrawal is to be made through an exterior ring main around the mid-section of the tank, with connections at three points.
- (3) Glue addition is to be made to the centre well of the clarifier-surge vessel.
- (4) A tunnel is not needed for the underflow from the clarifier-surge vessel; the underflow line may be located inside and on the tank bottom.
- (5) Clarifying filter backwash solution is to be returned to the first CCD thickener with provision to go to the second thickener. Clarified pregnant solution is to be used for backwashing.
- (6) Downflow clarifying filters have successfully used both garnet-anthracite and all anthracite filter media. In some installations a buildup of gypsum on the garnet has resulted in a progressive decrease in flowrate and eventually in the need to replace the media. In at least two recent installations, all anthracite beds are giving satisfactory results. Provision should be made for pH adjustment of pregnant solution in the clarifier by manual addition of acid.
- (7) If the millsite is subject to frequent sandstorms, a cover should be provided for the clarified pregnant liquor surge tank if located outdoors.

5.8 Solvent Extraction5.8.1 Design Basis

Pregnant liquor

design flow	1560 gal/min
average grade	0.131 g/l U_3O_8
maximum grade	0.211 g/l U_3O_8
U_3O_8 content, maximum	3960 lb/day
specific gravity	1.05

Extraction

number of stages	4
solvent composition	
Alamine 336 or equivalent	2.5% by volume
isodecanol or tridecanol	2.5% by volume
kerosene	95% by volume
solvent specific gravity	0.80
solvent specific heat	0.46 Btu/lb/ $^{\circ}$ F
organic loading	2.0 g/l U_3O_8
organic advance flow	165 gal/min
organic recycle flow	2080 gal/min

Stripping

number of stages	4
strip solution composition	150 g/l ammonium sulfate plus ammonia gas for pH control in the range of 4.0 to 4.3
advance flow	13.2 gal/min
recycle flow	55 gal/min
assay	25 g/l U_3O_8
specific gravity	1.11
pregnant organic flow	165 gal/min
pregnant organic temperature	40 $^{\circ}$ C
Ammonia required	0.40 lb/lb U_3O_8
Regeneration	
number of stages	1

barren organic flow	165 gal/min
regeneration solution composition	5% NaHCO ₃
advance flow	5 gal/min
recycle flow	70 gal/min

5.8.2 Equipment

Mixer-settlers

	Per stage	
	Mixer, working volume, gal	Settler ₂ area ft ²
extraction	3,100	2,100
stripping	930	160
regeneration	930	160

Tanks

	Number	Working volume, gal
barren organic	1	23,500
pregnant organic	1	23,500
organic sludge	1	40,000
raffinate surge	1	60 ft minimum diam x 10.5 ft high

5.8.3 General Instructions

- (1) The views of the Fire Insurance Underwriter should be obtained on the location of the solvent extraction circuit relative to the general plant layout.
- (2) Maximum overflow crest height on settler overflow weirs is to be 1.0 inches.
- (3) Provision is to be made for solvent recovery from the raffinate surge tank on an intermittent basis. The organic will be removed by skimming through an overflow pipe and will be returned to the No. 4 extraction mixer.

- (4) Combination mixer-settlers of rectangular design can be used. Organic depth in the settlers is to be 15 inches. Aqueous and organic overflow weirs, extending the full width of the settlers, are to be used.
- (5) A continuous bleed is to be taken from the bottom of the barren and pregnant organic surge tanks and returned to No. 3 strip mixer and to No. 2 extraction mixer respectively. When solvent regeneration is in operation, then the bleed from the barren organic surge tank should go to No. 2 extraction mixer.
- (6) A 'picket fence' baffle is to be located across the feed end of the settlers. A head loss of $1/4 - 1/2$ inch across the baffle should be used to size the pickets.
- (7) Make-up kerosene, alcohol and amine are to be added to the barren organic tank by separate pumps. Barrel type pumps are suitable for the alcohol and amine.
- (8) Mix tank working volume excludes freeboard and any space under the false bottom below the impeller.
- (9) The speed of the mixer impellers should be variable and the bottom clearance should be adjustable by raising and lowering the shaft.
- (10) The pregnant and barren organic surge tanks are to be covered and vented to atmosphere.
- (11) Provision is to be made for the introduction of approximately 15 cfm of air with the ammonia feeding the strip mixers. A sparge ring is recommended for dispersing the gases in the mixers.
- (12) A Teflon tube bundle or carbon block heat exchanger

is suitable for heating the organic. The actual operating temperature is expected to be lower than design capability.

- (13) The purpose of the organic sludge holding tank is to assist in the removal of interfacial crud from the settlers should this become necessary and for possible recovery of uranium from the spent regeneration solution. The tank should be equipped with an agitator mechanism, pipe lines for sulfuric acid, ammonia, process water and steam. An arrangement is needed for decanting this tank for supernatant disposal to tailings and for discharge of the yellowcake sludge to leaching and/or precipitation. Space should be provided for the possible future installation of a separate acidification tank, precipitation tank and thickener for processing spent regeneration solution.
- (14) Provision should be made for manual addition of sulfuric acid to No. 3 extraction settler aqueous overflow for pH adjustment of barren organic from the regeneration mixer-settler.
- (15) During detailed engineering, consideration should be given to the alternatives of separate mixer tanks or combined mixer-settler designs.
- (16) Piping and layout should be arranged to permit by-passing of the solvent regeneration mixer-settler unit.

5.9 Yellowcake Precipitation and Washing5.9.1 Design Basis

U_3O_8 production, maximum 3960 lb/day

Precipitation

strip solution flow 13.2 gal/min

strip solution, U_3O_8
concentration 25 g/l

number of stages two

pH - first stage 4.5 - 7.0

- second stage 6.5 - 7.5

temperature 60 - 70°C

ammonia consumption, approx. 0.20 lb/lb U_3O_8

Yellowcake, dry basis

specific gravity 5.7

specific heat 0.25 Btu/lb/°F

Yellowcake thickener

underflow density, for
flowsheet 33% solids

underflow density, for
pump design 20% solids

overflow solids 200 ppm

Thickener overflow filtration

feed flow 17.1 gal/min

filtrate solids, maximum 10 ppm

backwash clarified filtrate

Centrifuging

solids feed 194 lb/h

total solution feed 5.0 gal/min

discharge density 60% solids

solids in centrate 2% of feed

5.9.2 Equipment

<u>Tanks</u>	<u>Number</u>	<u>Working capacity gal. per unit</u>
precipitation	2	600
unclarified barren strip	1	2500
clarified barren strip	1	2500
thickener underflow	1	2500
filter backwash	1	250
centrifuge wash water	1	250

Thickener

diameter	18 ft.
bottom, sloped, height to diam	1/2
side walls	5 ft

Clarifying filter

type	downflow pressure sand filters
diameter	2 ft

Centrifuge

type	solid bowl
size	Bird 18 in x 28 in, or equivalent

5.9.3 General Instructions

- (1) The precipitation tanks are to be mechanically agitated. The tanks are to have baffles and the agitator is to be designed to provide only mild agitation of the yellowcake slurry.
- (2) The precipitation tanks and thickener tanks should be covered and vented with natural draft.
- (3) The solution balance in the yellow cake thickening and centrifuging circuit, page 4 - 3, is designed to produce dried yellowcake containing no more than 1.5%

soluble sulfate.

- (4) Circuit bleed from the clarified barren strip surge tank is planned to go to leach discharge.
- (5) Backwash from the sand filter will be returned to the precipitation tanks via an elevated surge tank.
- (6) The thickener feed well should be about 3 ft deep. No tunnel is required.
- (7) Good quality water should be used for the wash circuit and the hose connections. The area floor sump should discharge to the thickener.
- (8) Flocculant is not ordinarily required in the precipitation thickener; if found necessary, facilities can be provided later.
- (9) Feed to the centrifuge is to enter through an open pot; dilution water is added at this point. The design must provide adequate access for sampling and volume measurements.
- (10) The centrifuge discharge chute should be vented to the scrubber on the yellowcake dryer.
- (11) A steam heated heat exchanger is recommended on the pregnant strip flow to precipitation.
- (12) The uranium precipitation and washing area should be roofed and sided.

5.10 Yellowcake Drying and Packaging5.10.1 Design Basis

Yellowcake production

dry weight	4659 lb/day
contained U_3O_8	3960 lb/day

Dryer operation

moisture, dryer feed	40%
moisture, for dryer design	60%
moisture, dried product	2% maximum
solids maximum temperature	400°C
gas temperature, design	570°C at the hottest hearth
gas temperature, capability	650°C
dust to scrubber	2% of dryer feed

Product size, maximum 0.25 in

Bulk density, dried yellowcake

loose	80 lb/ft ³
compacted	120 lb/ft ³

5.10.2 Equipment

Dryer feeder	screw type
Dryer, type	multiple hearth
hearth area	100 ft ²
Dryer dust collector	high energy venturi type wet scrubber
Packaging dust collector	dry baghouse or wet scrubber with minimum make-up water requirements
Product storage bin	125 cubic feet
Product drums, size	55 gal
capacity	about 950 lb yellowcake

5.10.3 General Instructions

- (1) A wet scrubber is used to treat the dryer gases and to ventilate the centrifuge. The enclosure around the product size reduction unit, bin, and packaging station requires a separate collector. A dry baghouse is recommended, unless a wet scrubber can be specified which will permit total water make-up requirements within the allowable for the yellowcake washing circuit.
- (2) Dust collector stacks must be installed so as to permit measurement and sampling of flows.
- (3) The dryer is to be designed to permit access to all hearths without dismantling the unit.
- (4) Because of the variability of the U_3O_8 content of the plant feed, the turn-down capability of the dryer should be considered in its selection.
- (5) The dried product does not require cooling prior to discharge into the storage bin.
- (6) The lump breaker used on the dryer discharge is to be enclosed and vented to the packaging dust collector. A trommel mill unit is suggested for this service.
- (7) Provision should be made for vacuum cleaning in the dryer and packaging area.
- (8) The sump in the dryer - packaging area should discharge to the yellowcake thickener. The water used should be of good quality.
- (9) Full steel drums of yellowcake can be stacked to

a maximum of three high.

- (10) The dryer and drum packing station are to be in an enclosed area which is not to contain exposed pipe or ductwork, upon which dust could settle.
- (11) The walls and ceiling are to be sheathed to cover steelwork, etc., and to provide a smooth surface that can be readily painted. Floor trenches should not be used and the floor sump should be of minimum size.

5.11 Estimated Consumption of Reagents,
Supplies, and Process Steam

Based on 2,000 ton/day of ore:

	<u>per day</u>
Sulfuric acid (100% basis)	420 tons
Sodium chlorate	12,000 lb
Flocculant, Dow MG - 200	600 lb
Sodium carbonate	600 lb
Glue	200 lb
Amine	200 lb
Isodecanol or tridecanol	410 lb
Kerosene	470 gal
Ammonia, re: ore grade	1,340 lb
re: 3,960 lb/day U_3O_8	2,380 lb

Supplies

SAG mill balls	0.45 lb/ton
Rod mill rods	1.0 lb/ton
SAG mill liners	0.14 lb/ton
Rod mill liners	0.12 lb/ton
Product drums (900 lb/drum)	1,165 per year

Process steam

	<u>lb per hour</u>
Leaching	19,600
Heating of solvent	700
Uranium precipitation	900
Miscellaneous	<u>600</u>
	21,800

5.12 Mill Sampling Facilities

5.12.1 Moisture in Mill Feed

A manual sample is to be taken from the mill feed belt.

5.12.2 Mill Feed

An automatic sampler of the straight-line type is to be located on the leach feed pulp; this unit is to operate continuously. The initial sample is to be reduced by two stages of Vezin samplers. The third cut is to be stored in a small tank equipped with a mixer and is to be further reduced at the end of each shift by a third Vezin cutter.

The first cut is to be 3.40% of the feed stream, while the second and third stage samplers take 2.50 and 1.25% cuts to provide approximately 10 quarts of sample.

5.12.3 Leach Discharge

Access is to be provided to readily obtain a manual sample at the discharge from the leach circuit.

5.12.4 Mill Tailings

An automatic sampler of the straight-line type is to be located on the CCD thickener circuit pulp discharge; this unit is to operate continuously. The initial sample is to be reduced by a second stage automatic sampler which is to operate intermittently. Final sample volume should be approximately 10 quarts for each 8 hour shift.

Any raffinate sent to tailings disposal should enter the system downstream of the mill tailings sampler.

5.12.5 Yellowcake

A sample will be cut by hand from the top of each drum of yellowcake.

5.12.6 Solution Samples

Solution samples are to be taken to provide an 8 hour composite sample of the following streams:

Pregnant liquor (solvent extraction feed)

Raffinate

Pregnant organic

Barren organic

Pregnant strip liquor

Barren strip liquor

Washing circuit bleed stream

Regeneration aqueous discharge

5.12.7 General

All samplers, with the exception of those for mill feed and yellowcake will require corrosion resistant construction.

Sampling of process streams by the operator to control solvent extraction will be in addition to those listed above.

5.13 Main Installed Spare Pumps

This list of spare pumps is subject to change pending details of final flowsheet and review of general arrangement drawings.

Leach feed

Leach discharge

CCD thickener underflows

CCD thickener slurry advance

Tailings discharge

Clarifying thickener feed

Sand filter feed

Pregnant liquor feed to SX

Feed to raffinate tank

Discharge from raffinate tank

Feed to pregnant organic tank *

Feed to barren organic tank *

Feed to yellow cake precipitation

Feed to barren strip liquor sand filter

Discharge from clarified barren liquor surge tank *

Sulfuric acid feed to process

Sodium chlorate feed to process

Flocculant feed systems to process

Glue feed system to process

* No spare on discharge of tank on premise that the tank can be bypassed.

5.14 Standby Power Requirements

Standby power should be provided to operate the following equipment during a failure of the normal electrical power supply.

Leach agitator drive; all to be connected, but only one at a time to be driven.

Leach discharge pump

All CCD thickener drives and rake lifting devices

Thickener underflow diaphragm pumps

Slurry advance pump on each thickener

Tailings pumps

Solvent extraction area sump pumps

Yellowcake thickener drive and underflow pump

Yellowcake dryer drive and cooling fan

Plant lighting

Instrument air

Gland water pump

Steam boiler

5.15 Instrumentation

5.15.1 General

This outline of requirements for process instrumentation represents recommendations with regard to applications where instruments can be used effectively to aid the operator in achieving efficient mill operation. It is subject to change pending finalization of the general arrangement and flowsheet circuits, and process equipment selection.

Graphic flowsheet-type panels with light indicators may be worthwhile.

High level alarms are to be installed in critical floor sumps.

5.15.2 Grinding

Remote control of the ore feeder vari-speed drives from the grinding control panel.

An integrating weightometer on the semi-autogenous mill feed belt, which will control the ore feed and water addition to this mill. A Merrick Model E-310 is suggested.

Indicating-recording power meters for the semi-autogenous and rod mill motors.

Flowmeters on the water supply to the semi-autogenous mill feed and to the discharge of the scalping screen.

A density recorder-controller on the leach feed to control water addition to the scalper screen discharge.

A mass flow-indicating and totalizing system is recommended for the leach feed, for metallurgical accounting purposes.

5.15.3 Leaching

Conductivity indicating-controllers on the discharges from the first and second leach tanks to provide automatic adjustment of acid addition regardless of changes in ore type or throughput.

A conductivity indicator on the third and sixth leach tanks.

Flow rotameters on all acid feed lines to individual tanks, plus an indicating totalizer on the acid header feeding the system.

Manual control of sodium chlorate addition through a flow meter and control valve to the second, third, and fourth leach tanks.

Multi-point emf recorder with electrodes in the second, fourth, and sixth leach tanks.

Totalizing flow indicator on total sodium chlorate addition to leach.

Temperature controllers for steam addition to the first, second, third, and fourth leach tanks.

A multi-point temperature recorder to serve all leach tanks.

Flow totalizer on total steam supply to leaching.

5.15.4 CCD Thickening and Clarification

Flow recorder-controller with totalizer for raffinate recycle to CCD wash. Maintain set volume of recycle raffinate to provide desired CCD wash ratio.

Flow recorder-controller with totalizer for fresh water wash to CCD.

Dependent upon the type of distribution system selected, it may be desirable to have measurement of flocculant flow to each thickener to facilitate manual adjustment of flocculant additions. Metering pumps can also be considered.

High torque alarms, torque recorders and motorized lifting devices for the thickener rake mechanisms. Push button controls for the lifting devices should be located both on the thickeners and in the control room for the CCD area.

Gamma gauge density and magnetic flowmeter indicating-recorders on each thickener underflow are recommended, with provisions for future conversion to control of underflow pumping speed.

Bubbler type level indicator and high level alarm in the clarifying thickener-surge tank.

Flow indicator-controller on clarifying thickener underflow to leach or CCD alternative.

The instrumentation for the clarifying sand filters (flow-meters, switches and relays for automatic operation) is to be provided by the supplier of the filters.

A means of measuring solution clarity, such as an on-stream turbidimeter, should be provided for the clarifying thickener overflow and for clarified pregnant solution.

5.15.5 Solvent Extraction and Precipitation

Flow recorder-controllers on the following process streams to the solvent extraction circuit:

Pregnant solution flow
Barren organic to extraction
Pregnant organic to stripping

Barren ammonium sulfate solution to stripping
Soda ash solution to solvent regeneration

Flow totalizer on the raffinate discharge to tailings, for metallurgical accounting.

Temperature indicating controllers to control the steam supply to the pregnant organic and to the pregnant strip heat exchangers.

pH recorder-controllers for stripping, to maintain set pH values in the first three stripping stages by individual control of ammonia addition to these three strip mixers. The pH measurements are to be made on a small bleed stream withdrawn from each mixer stage through a wall outlet.

Ammonia gas is to be introduced into the mixers through ring spargers of Type 316 stainless steel construction. Air is to be mixed with the ammonia gas to reduce local ammonia concentration and the attendant possibility of local uranium precipitation.

A pH recorder-controller is required for uranium precipitation, to maintain a set pH in the second uranium precipitation tank. Manual addition of ammonia will be required for the first precipitation tank.

Bubbler-type level indicators on each of the following tanks:

Barren organic surge
Pregnant organic surge
Clarified pregnant liquor storage
Raffinate storage
Unclassified barren strip liquor surge
Clarified barren strip liquor surge
Organic sludge tank

Conductivity probes in all the solvent extraction system mixers to indicate continuous phase condition.

Controls for the ammonia storage and vaporizer system, to be supplied by the equipment supplier or as recommended by them.

5.15.6 Yellowcake Washing and Drying

Flowmeters on the following water and process solution lines:

Make-up water to dryer scrubber

Solution input to dryer scrubber

Scrubber solution to the centrifuge

Fresh water to the centrifuge

Make-up water to the packaging area scrubber, if a dry dust collection system is not used

A flow totalizer on the total fresh water supply to the yellowcake precipitation, washing, drying, and packaging areas, including hose stations which will give rise to sumpage return to these areas.

A flow indicating-totalizer on the barren strip liquor bleed stream.

A measuring pot with shut-off valve on the bottom discharge line of the feed to the centrifuge, in order to determine flow rate by measuring time to fill the pot to a fixed level. Fresh water to the centrifuge will also be added through this pot.

A variable speed pump to transfer yellowcake slurry from the yellowcake thickener underflow surge tank to the centrifuge, with the pump control located at the centrifuge station.

Dryer instrumentation, including draft indicators and controls, temperature indicators and controls, fuel system controls, alarms, etc. as supplied or recommended by the equipment supplier.

Instrumentation for the sand filter is to be supplied by the filter supplier.

5.16 Materials of Construction

5.16.1 General

The materials of construction specified herein are those which have been used successfully in operating uranium mills and are considered to be suitable for the process conditions selected for use in the Anderson mill.

The choice of construction materials is governed primarily by such process chemical and temperature requirements as:

Leaching temperature of 75°C.

Free acid concentration of 20 - 25 g/l in the leach pulp, with potentially higher concentrations in the first leaching tanks.

Addition of 6 lb/ton of sodium chlorate to the leaching circuit.

Tertiary amine, phase modifier, and kerosene in solvent extraction.

Free ammonia in stripping and precipitation.

Because early failure of bonded linings have sometimes been experienced, all contracts for lining of equipment should specify that the lining contractor accepts responsibility for the condition of the surface to be covered. It is recommended that the application of the lining be carried out by the supplier of the lining material.

5.16.2 Grinding

Pump tanks and launders in the grinding and neutral thickening areas should be of steel construction, lined with

rubber. Linatex is widely used for wet abrasion applications because of the relative ease in building up worn areas.

5.16.3 Leaching

Rubber-lined mild steel tanks, tank covers, and agitators are suitable. Because of the high operating temperature and acidic oxidizing conditions, detailed rubber specifications must be confirmed with suppliers. A number of laminates of different rubbers have been used successfully in similar applications.

Leach tank vent piping may be of fibre-glass reinforced polyester or polyethylene construction.

Steam injection pipes may be of Carpenter Alloy 20 or equivalent, or of carbon steel pipe covered and lined with hard rubber.

The outer surface of the leaching tanks should be painted with one primer coat and two finishing coats of acid resistant paint.

5.16.4 Thickeners

Thickener tanks in acid circuits have been successfully constructed from the following materials:

Wood stave

Rubber lined steel

FRP lined steel

FRP lined concrete bottom, wood stave wall

Type 316 stainless steel is preferred for the vertical shafts, rake arms and blades. FRP feed wells have been used successfully. Thickener cones should also be of Type 316 stainless steel.

If wood stave construction is used for the thickener walls, the supporting bands and turnbuckles should be covered with polyethylene tubing for acid protection.

A rubber lined straight-through Saunders-type valve is suggested for the thickener underflow lines.

5.16.5 Clarification and Solution Storage

The pressure sand filters should be rubber-lined mild steel. Type 316 stainless steel is to be used for the internal distribution piping.

Type 316 stainless steel is recommended for the clarifier rakes.

Solution holding tanks may be wood stave or mild steel lined with rubber or FRP. The raffinate tank should not be rubber lined.

5.16.6 Solvent Extraction

FRP lined mild steel is suitable for storage tanks and for mixer-settlers. FRP bonded to concrete, self-supporting FRP and Type 316 stainless steel can be considered for the mixer-settlers.

Type 316 stainless steel is preferred for the wetted parts of the extraction and stripping pumper-mixers.

Teflon and impervious graphite have been used for the steam heater on the pregnant organic stream. Stainless steel has failed on this service.

5.16.7 Yellowcake Precipitation, Washing and Drying

The uranium precipitation tanks may be wood stave or FRP coated mild steel; the mechanisms should be FRP, neoprene or nitrile rubber covered mild steel, or Type 316 stainless steel.

FRP lined mild steel is suggested for the strip liquor holding tanks.

The yellow cake thickener tanks can be FRP lined mild steel, epoxy coated mild steel, or wood stave. The thickener shafts, rake arms and rake blades should be of Type 316 stainless steel. FRP feed wells have proven satisfactory in this service.

The pregnant strip liquor heat exchanger should be of Teflon (or equivalent), or Type 316 stainless steel construction.

The wetted parts of the yellow cake centrifuge should be of Type 316 stainless steel construction. Stainless steel is suggested for the screw of the dryer screw feeder.

The wet scrubber should be Type 316 stainless steel.

5.16.8 Chlorate Storage

The sodium chlorate tanks may be of mild steel construction. As moderate to severe corrosion may be encountered, it is recommended that the inside of the tanks be painted with a phenolic resin coating. Self-supporting FRP may also be considered.

5.16.9 Pumps

Rubber lined pumps with Type 316 stainless steel trim are recommended for acid slurry service throughout the plant. Acid solution pumps may be of stainless steel or, where contact with organic solvent is not possible, may be rubber lined with stainless steel trim.

Type 316 stainless steel pumps or pumps of mild steel with PVC or phenolic covered impellers and casings have been used successfully in the solvent extraction circuit for

both aqueous and organic service. In order to reduce gland seal flows, vertical type pumps have been used where layout permits; otherwise, mechanically sealed horizontal pumps are employed.

In the precipitation and yellow cake washing circuits, the yellow cake slurry pumps may be Sandpiper type (air-operated diaphragm), O.D.S. or Moyno. Natural rubber stators are preferred for Moyno pumps; they should be over-sized (slow speed) and care must be taken to avoid interruption of feed to the pumps.

The sodium chlorate solution pumps should be of cast iron construction, self-priming and glandless. There must be no hydrocarbon based oil or grease lubrication.

Sulphuric acid pumps should be of mild steel construction and equipped with mechanical seals.

5.16.10 Concrete Surfaces Subject to Acid Spillage

Concrete protection practice in acid-leaching uranium mills has varied from nil to almost complete coverage. If no protection is provided, it is probable that some resurfacing will be required after 3 - 7 years' operation.

At the Exxon mill, in the leach area, the bottom level of the CCD pumphouses, and around the SX mixer-settlers, the column footings, pump bases and twelve inches of the outside wall were coated with an epoxy containing glass flakes. Two applications were used to provide a total thickness of 20 mils.

At the Elliot Lake operations in Canada, several types of coatings were tried. Denison Mines used asphalt in the leaching area and, after twenty years, the covering is still in service, although the concrete has broken up to some extent in the pH adjustment area. Epoxy paste,

applied by trowelling, was used extensively at Denison and at other Elliot Lake mills. With this covering, it is essential that the concrete be completely dry before application and, in the sumps and lower floor areas, which were not adequately dried, the coating failed. In the subsequent expanded leaching installation at Denison, the floors, piers and walls up to a height of 18 inches were protected with a silica loaded epoxy compound.

The areas around ion exchange and precipitation at the Shirley Basin mill of Utah International are protected with a polyester quartz sand mixture. The only cracking has been along expansion joints. The floor and walls of the CCD thickener tunnels have been covered with FRP up to a height of eighteen inches.

5.16.11 Piping

Carbon steel is the standard material for pulp slurries, unlined for neutral pulps, soft natural rubber lined for acid pulps. Victaulic couplings are usual but flanged connections should be employed in thickener underflow lines.

Material conducting hose has been used in limited areas such as:

- where abrasion is particularly severe
- to obtain long radius bends
- to decrease pipe vibration
- to facilitate maintenance on pipe connections

Where abrasion is not expected to be severe, high density polyethylene and ABS may be also considered for slurry lines. For security reasons, such materials should not be used on thickener underflow lines. FRP has low abrasion resistance and should not be used for slurry piping.

Acid solution lines may be of such material as PVC, FRP, ABS, rubber lined steel, etc.; the main criterion is installed cost. Failure of plastic lines in the manifolds around the sand bed filters, because of water hammer, has occurred. Rubber lined steel is suggested for this location.

PVC and FRP piping have often been used for the organic solvent. In the locations where a line breakage would drain the contents of a vessel, stainless steel is preferred. Buna-N rubber has satisfactory resistance to the organic and may serve for gaskets, sleeves, etc.

In the yellow cake precipitation-thickening area, PVC, FRP, rubber lined steel, etc. are satisfactory. Thickener underflow should be stainless steel or rubber lined steel.

For the tailings discharge, rubber lined steel, high density polyethylene, polypropylene and ABS have been used successfully. It is recommended that rubber lined steel be used for the initial portion of the tailings line next to the tailings pump.

Schedule 80 carbon steel pipe with welded fittings and flanged connections is commonly used for concentrated sulfuric acid. Small lines are blocked by iron sulfate, and sizes below 1 inch should be avoided.

5.17 Analytical and Testing Services5.17.1 Mine Control

Mine operation can be largely controlled using radiometric methods to delineate ore and waste both at the working faces and during transportation. Marginal material can be sampled for laboratory analysis to determine the final destination of the material.

The radiometric instruments require chemical analyses for their calibration.

Drill core from mine development would likely be analyzed both chemically and radiometrically.

5.17.2 Process Control

The basic mill samples and their frequency of collection are listed below on a basis of milling 2,000 tons per day of ore:

<u>Sample</u>	<u>Frequency</u>	<u>Elements to be Analyzed</u>
Feed to grinding	shift composite	H ₂ O
Leach feed	shift composite (possible); daily composite	U ₃ O ₈ , Mo, V ₂ O ₅ , CO ₂
Leach discharge liquid	shift or daily grab	emf, pH, Fe ⁺⁺ , Fe ⁺⁺⁺ , H ₂ SO ₄
solid	shift or daily grab	U ₃ O ₈
Final CCD thickener liquid	daily composite	U ₃ O ₈
solid	daily composite	U ₃ O ₈
Pregnant solution	shift composite	U ₃ O ₈ , pH, turbidity
Raffinate	shift composite	U ₃ O ₈
Pregnant organic	shift composite	U ₃ O ₈

Barren organic	shift composite	U_3O_8
Pregnant strip liquor	shift composite	U_3O_8, SO_4
Barren strip liquor	shift composite	U_3O_8, SO_4
Product	daily composite	U_3O_8, SO_4, H_2O
Organic	weekly grab	amine, alcohol
Regeneration aqueous	daily composite	Mo, U_3O_8 , pH
Washing circuit bleed	daily composite	U_3O_8
Tailings liquor	weekly composite from shift grab	$U_3O_8, V_2O_5, Mo, Fe, NH_3, Ra, Se$

5.17.3 Product Control

While the lot size has not been specified, it is estimated that approximately 1 lot per two weeks will be produced. Therefore, two to three samples of yellow cake would be submitted monthly for a check on specification requirements.

Depending on the contract stipulations, the following elements would be determined:

$U_3O_8, PO_4, \text{halides}, SO_4, Fe, H_2O, Mo$

Capability for analyzing for the following elements should exist, but their determination may not be routinely required:

$V_2O_5, F, As, CO_2, Th, Zr, Ca, Na, B, K$

5.17.4 Miscellaneous Services

Laboratory analytical services should be considered also for the following demands:

Additional exploratory work involving geological

samples; diamond drill or churn drill samples for future mining requirements.

Miscellaneous mill samples such as dust collection system products for U_3O_8 content, daily leach feed composite samples for screen analyses to control mesh of grind, provide stock solutions for mill operators' use, etc.

Miscellaneous samples generated by plant metallurgists such as screen assay test fractions for U_3O_8 , size distribution, leach test products, solvent extraction, test products, etc.

Miscellaneous infrastructure samples such as potable and boiler water treatment plant analyses, sewage treatment plant analyses, plant operator urine analysis for albumin, sugar, and U_3O_8 content. If the local hospital has the facilities, then the last item would probably be performed under medical auspices.

Miscellaneous plant effluent control including radium determinations.

5.17.5 Metallurgical Testing

The laboratory facility should include space and equipment for routine testing of all unit operations in the mill flowsheet for optimization of process variables.

Testing of samples from other properties, amenability procedures for custom ores, etc. may also be required from time to time.

5.17.6 Facilities

Sample preparation facilities will be necessary since all mine samples, plant feed, tailings and product samples, plus any solid samples from experimental test work would have to be suitably dried, ground, and individually blended.

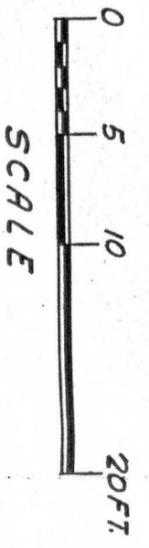
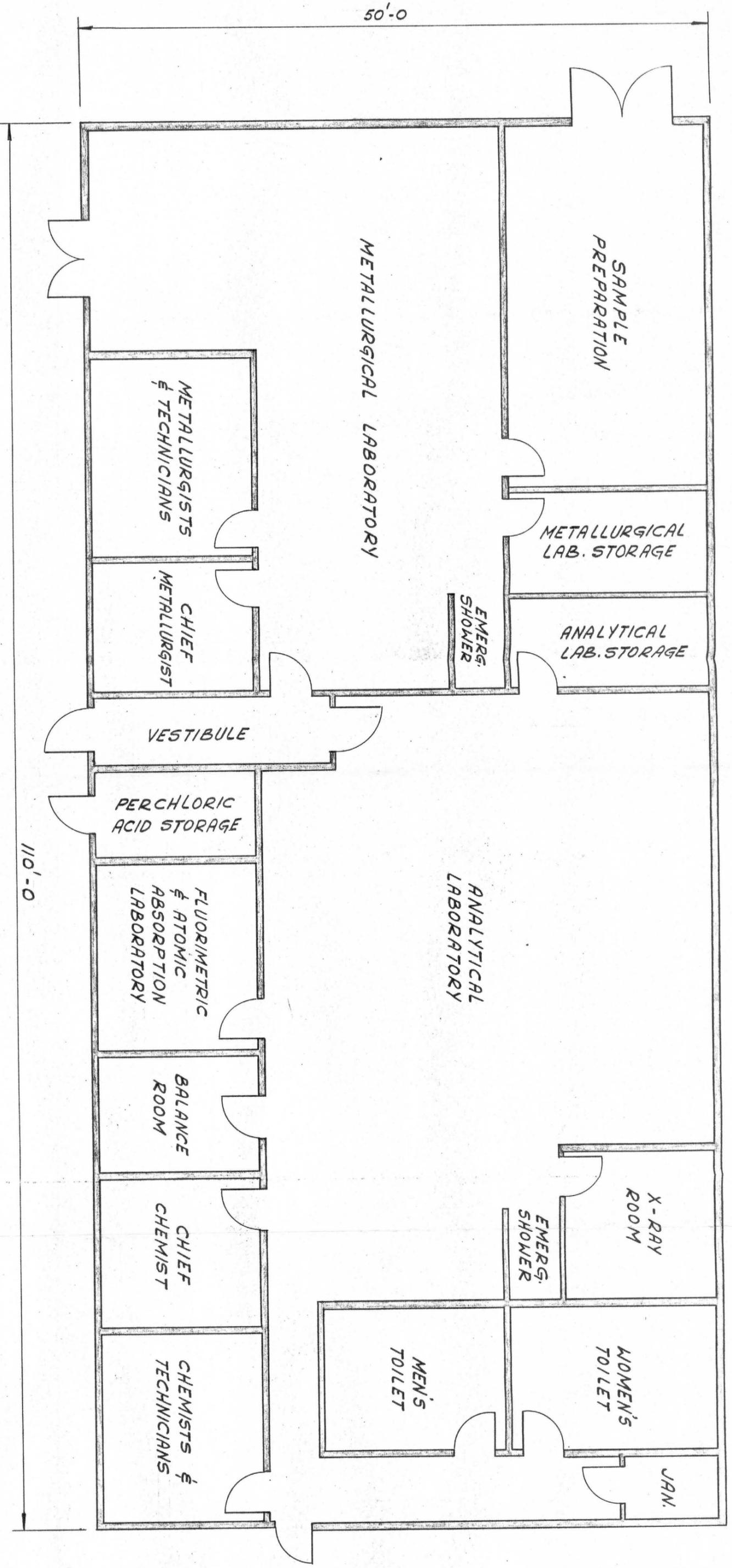
An operation of this magnitude requires a well equipped facility. The accompanying drawing shows the suggested layout of a suitable analytical and metallurgical laboratory plus offices. Efficient dust control is required throughout.

The arrangement assumes that an X-ray unit would be available for assaying. Although not necessary for routine mill analyses, an atomic absorption unit is very desirable for miscellaneous determinations, environmental water samples, etc.

6. Abbreviations

Definitions for the abbreviations used in this report are as follows:

Btu	British thermal units
°C	degrees Celsius or Centigrade
°F	degrees Fahrenheit
kW	kilowatt
kW.h	kilowatt hour
hp.h	horsepower hour
min	minute
h	hour
d	day
rpm	revolutions per minute
gal/min, gpm	gallons (U.S.) per minute
emf	electromotive force
mV	millivolts
lb	pound, avoirdupois
ton	short ton of 2,000 pounds
g	gram
kg	kilogram
l	liter
gal	gallon, United States measure
ft	foot
ft ²	square feet
ft ³	cubic feet
ID	inside diameter
diam	diameter
ppm	parts per million by weight
%	percent
\$	United States dollars
SAG	semi-autogenous grinding
FSFR	full scale filtration rate
CCD	counter-current decantation
ODS	Oliver diaphragm slurry



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