

THE UNIVERSITY OF MICHIGAN
INDUSTRY PROGRAM OF THE COLLEGE OF ENGINEERING

A TECHNICAL AND ECONOMIC EVALUATION OF A
SOLVENT EXTRACTION PROCESS FOR THE
EXTRACTION OF URANIUM FROM DOMESTIC ORES

M. E. Weech

December, 1957

IP-253

ACKNOWLEDGEMENT

The Engineering Research Institute of the University of Michigan sponsored this investigation and has approved distribution of the report by the Industry Program of the College of Engineering.

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1.0 INTRODUCTION

Due to a keen interest on the part of Engineering Research Institute personnel in uranium ore processing, funds were made available through the Director's office for an evaluation of a process comprising the latest technology available. These funds were to be used in a study of the feasibility and economics of such a process, pointing up both the economic and technical advantages to be gained, and so stimulate sufficient industrial interest to instigate research projects in this area.

This report summarizes the results of this study.

2.0 SUMMARY AND CONCLUSIONS

A process has been presented for the extraction of uranium from domestic ores utilizing a solvent-extraction separation method and containing provisions for recovery of acids that would normally be wasted as reaction products of the undesired components.

This process has further advantages in lower grinding costs and more compactness, and shows promise of producing a product of greater purity than conventional processes.

The total investment for a mill utilizing this process has been estimated at \$4,385,000. Processing costs per ton of ore are estimated at \$5.25, which should enable a plant to pay for itself in less than two years at current prices for the UO_3 product.

Pilot plant operation is believed necessary on this process prior to final plant design. The uncertainties believed to exist in the process and in the equipment design are listed.

3.0 DISCUSSION OF PROCESS

3.1 PRESENT PRACTICES IN THE FIELD

For many years the uranium-containing ores were processed mainly for their vanadium content with the uranium having secondary value. Consequently the early flowsheets worked out by the U. S. Bureau of Mines were designed for vanadium recovery with uranium recovery steps added to their process as its recovery became important.

One of the later flowsheets developed by the Bureau of Mines is shown in Fig. 1, which was taken from Ref. 1. This process is adapted to the processing of a high-lime ore. As can be seen, the chemical requirements for such a process are quite high. Appreciable quantities of soda, nitric acid, caustic, and sulfuric acid are required. In addition there are a number of filtration steps in the process which can be very troublesome on many of the ores. A discussion of the operation and difficulties of such a flow-sheet is given by Philippone in Ref. 2.

Newer processes have been developed which eliminate a few of the problems inherent in the Bureau of Mines flowsheet. These processes are based upon ion exchange resins that are selective in their absorption of a uranium sulfate complex. These are anion resins since the complex of sulfate and uranium ion has a negative charge. The processes are called resin-in-pulp or resin-in-column depending upon whether the exchange resins are fixed in a column or are contained in wire-screen baskets and lowered into the slurry of acid and finely ground ore. If a resin-in-column process is used, the leach liquor is separated from the solids by filtration, while in the resin-in-pulp flowsheet, a slurry of solids in the leach liquor is circulated through the resin. In this method the ore is ground finely enough so that there is no plugging of the resin bed with coarse ore particles. A brief schematic flowsheet of a resin extraction process is shown in Fig. 2. This flowsheet and other information on ion exchange processes were taken from Refs. 3, 4, and 5.

The difficulty with these processes is that the resin bed slowly accumulates "poisons" which cannot be eluted from the bed so that eventual replacement of the resin is necessary. Further, when the ores being treated contain considerable lime, the acid costs from these processes are high.

As an example, assume an ore of the following composition is being treated.

U ₃ O ₈	0.3 wt %
Ca O	12.0
Nondissolvables	Remainder

FIGURE -1
 BLOCK DIAGRAM OF SALT-
 ROAST, ACID-LEACH PROCESS

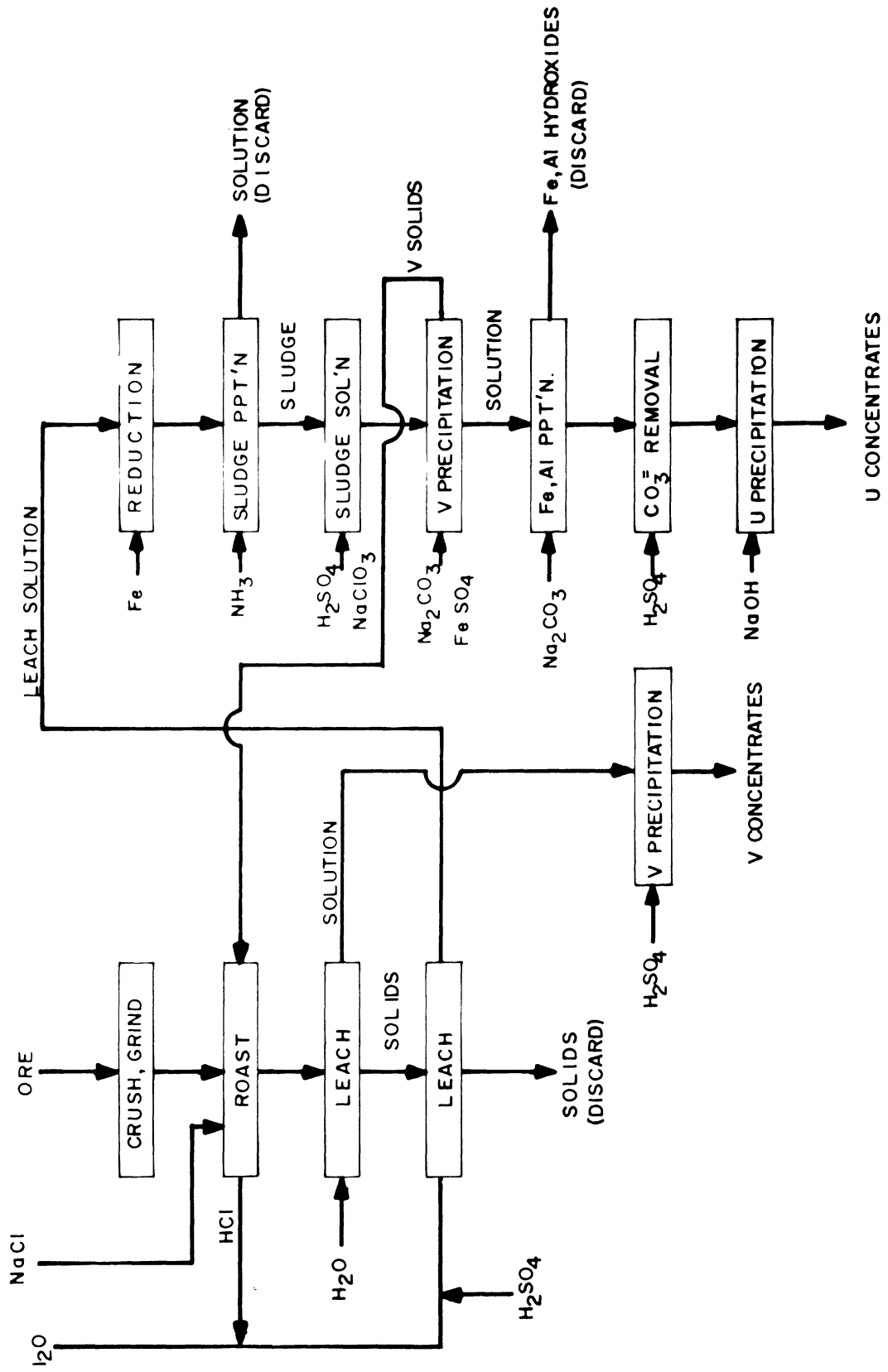
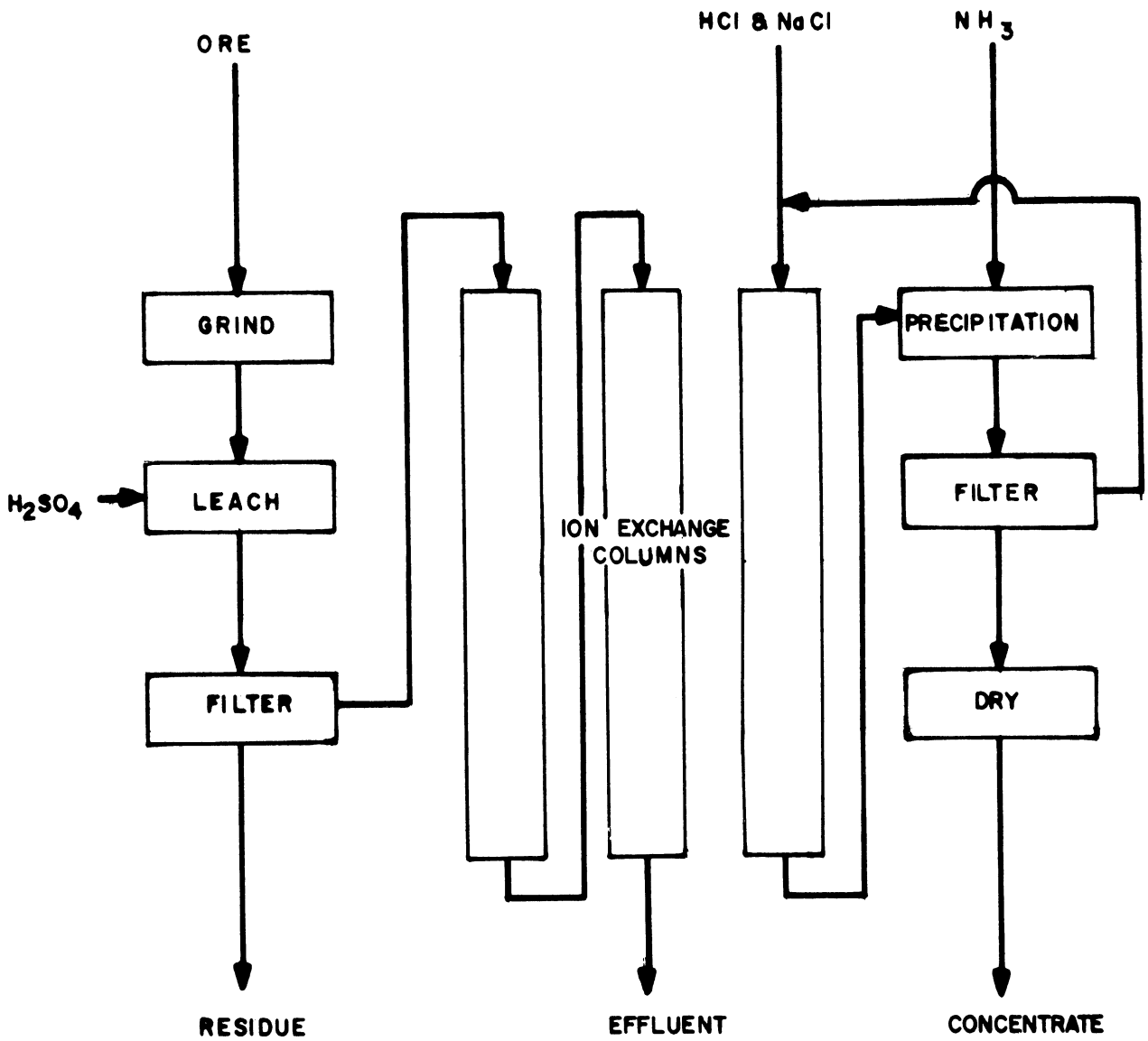


FIGURE-2
BLOCK DIAGRAM OF ION EXCHANGE
PROCESS FOR SULFATE SOLUTIONS



This ore has a reasonably high lime content but certainly not the highest that would be encountered. This ore would require approximately 421 lb of 100% H_2SO_4 per ton for leaching, not counting oxidants that would have to be added for rapid dissolution. Commercial 66° acid (93.2 wt %) is assumed to be purchased at a bulk price of 1.1 cents per lb. Freight in tank car quantities would add approximately 0.2 cents per lb to the base cost for a 300-mile haul, so the final cost of acid for leaching a ton of the ore would be \$5.87. Total yearly acid costs for a 1000-ton-ore-per-day plant operating 300 days per year would be \$1,761,000. This, of course, is the major cost but additional costs would be incurred for oxidants, eluting acids, and caustic or ammonia.

The basic difficulties in the present processes can then be summarized as follows:

There are large chemical requirements either in leaching acids or in subsequent steps or both.

Filtration steps within the process are cumbersome and slow.

Critical steps within the process such as pH adjustment can be troublesome where close tolerances are required. There are several pH adjustment steps in the Bureau of Mines flowsheet and also some in the ion exchange process.

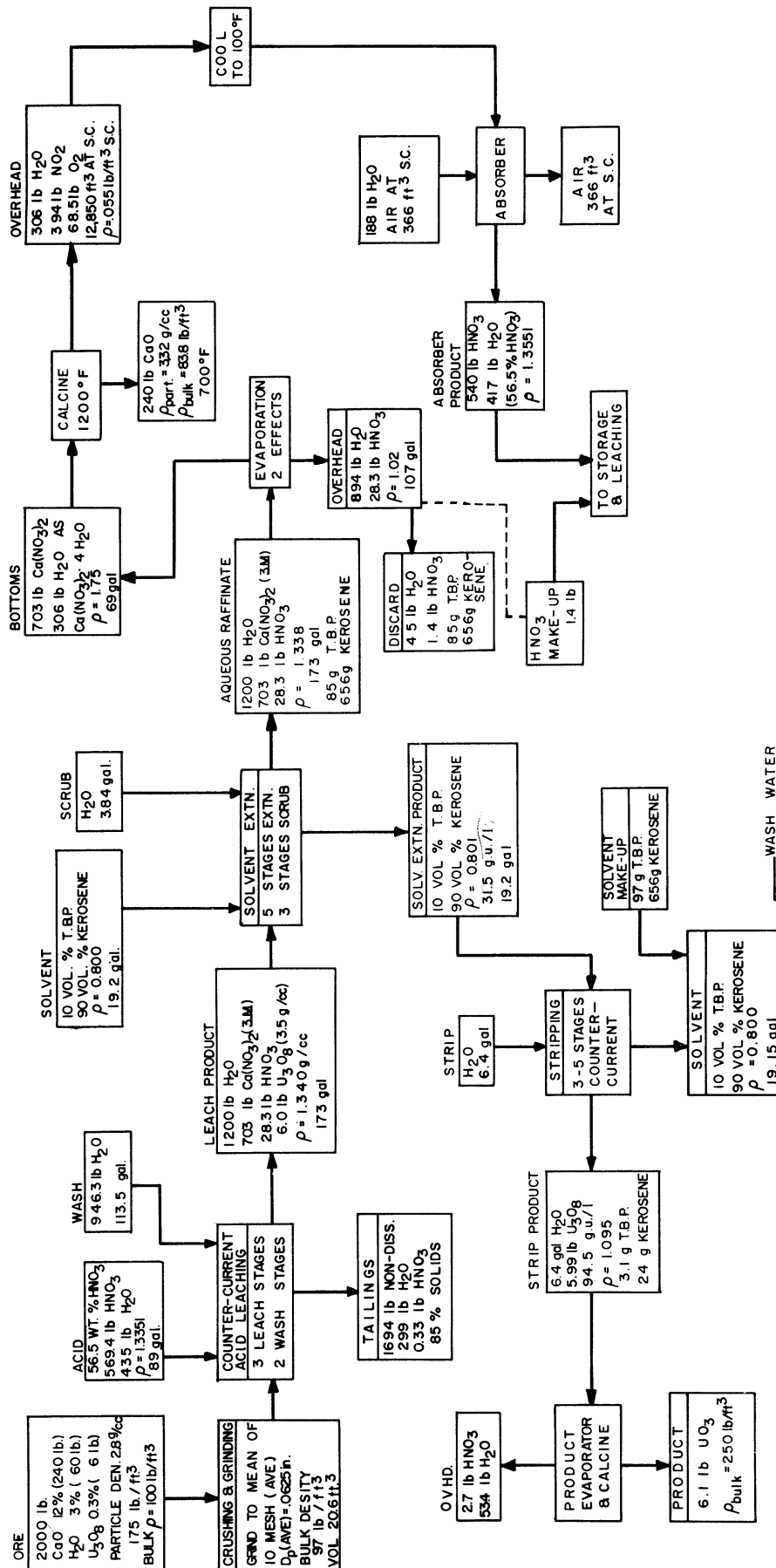
Product purity could be considerably improved in the existing processes.

3.2 PROPOSED PROCESS BASIS

In arriving at the process proposed here, attempts were made to eliminate or avoid the difficulties outlined above. Each of the steps outlined has been backed up with reference data found in the literature, which are known to be operable from past experience, or have been found to be true from laboratory experiments. While the process steps are firm, the optimum selection of equipment suitable for final plant operation remains to be done and should be proven with pilot plant information.

The basic process, shown in Dwg. 461-1128-U-1, consists of grinding the ore to an average particle size of 10 mesh and leaching the ore fines with nitric acid. The nondissolvable part of the ore is washed with water to remove entrained acid and then discarded. The rich liquor from leaching is combined with the washes and routed to countercurrent solvent-extraction equipment. In this equipment the aqueous solution is contacted with an organic extractant that is immiscible in the water phase and highly selective in extracting uranium from the aqueous phase as a uranyl nitrate. An organic

OVERALL FLOWSHEET



DESIGNED BY	APPROVED BY	SCALE	NONE
DRAWN BY	R. I. K.	CHECKED BY	DATE
ENGINEERING RESEARCH INSTITUTE		5/28/57	
UNIVERSITY OF MICHIGAN		ANN ARBOR MICHIGAN	
PROJECT		461-1128	
CLASSIFICATION		UNCLASSIFIED	
ISSUE	DATE	DWG. NO. B-461-1128-U-1	
		CHEMICAL FLOWSHEET	
		URANIUM ORE PROCESSING	
		PLANT	

extractant that fulfills these requirements would be n-tributyl phosphate mixed with kerosene or another suitable diluent.

The uranium-containing organic solvent is scrubbed with a small stream of distilled water in two or more scrubbing stages to remove impurities that may have been extracted with the uranium and is then routed to a second contactor where the uranium is stripped from the organic phase with distilled water. The uranium stripped out in the aqueous phase is quite pure and is concentrated in an evaporator to a thick syrup of $\text{UO}_2(\text{NO}_3)_2 \cdot 3\text{H}_2\text{O}$. This tri-hydrate is calcined at approximately 400°F which decomposes the uranyl nitrate to UO_3 , water, NO_2 , and O_2 . The UO_3 is packaged and shipped and the gases are reabsorbed in water to form nitric acid which can be reused in the process.

The organic solvent after being stripped of its uranium is scrubbed with a sodium carbonate solution and then washed with water. This treatment removes any decomposition products of tributyl phosphate that have formed, as well as any traces of metallic ions that were not removed in the stripping operation. After this treatment the organic solvent can be recycled back into the extraction apparatus.

The aqueous stream leaving the solvent-extraction apparatus contains all the components in the ore that dissolve in nitric acid. Generally these components would be calcium, some aluminum, iron, vanadium, and traces of other elements depending upon the nature of the ore. Nitric acid was consumed in the leaching step forming nitrate salts of these elements. A recovery of this acid is mandatory for an economically operating process. Recovery is accomplished by concentrating this solution in evaporators using steam and calcining the salts at $700\text{-}1200^\circ\text{F}$ to decompose the nitrates. UO_2 , O_2 , and steam are evolved, leaving the metal ion as the oxide. The gases are cooled and absorbed in water in a multi-stage absorption tower. The product of the absorption is nitric acid which is recycled back to the leaching step. Metal oxides are discarded, although some use could be found for them in cement or plaster manufacture.

A detailed discussion of the process steps follows.

3.3 STEPS IN THE PROPOSED PROCESS

3.1 Crushing and Grinding.—Crushing and grinding requirements are set both by the high uranium recoveries required in the leaching step and the least possible amount of fines carried over into the solvent-extraction step. Leaching requirements are best met by fine grinding, while the solvent-extraction step works best if very few fines are carried over from leaching to solvent extraction.

It appears that for the secondary ores such as carnotite⁶ that one stage of leaching for a -4 mesh in HCl would give uranium recoveries of 96% after leaching for 4 hr at 80°C. For -20 mesh ore at 80°C, 95% recoveries of uranium were obtained in 1/2 hr with no significant improvement in recoveries being noted for additional leaching times of up to 4 hr.

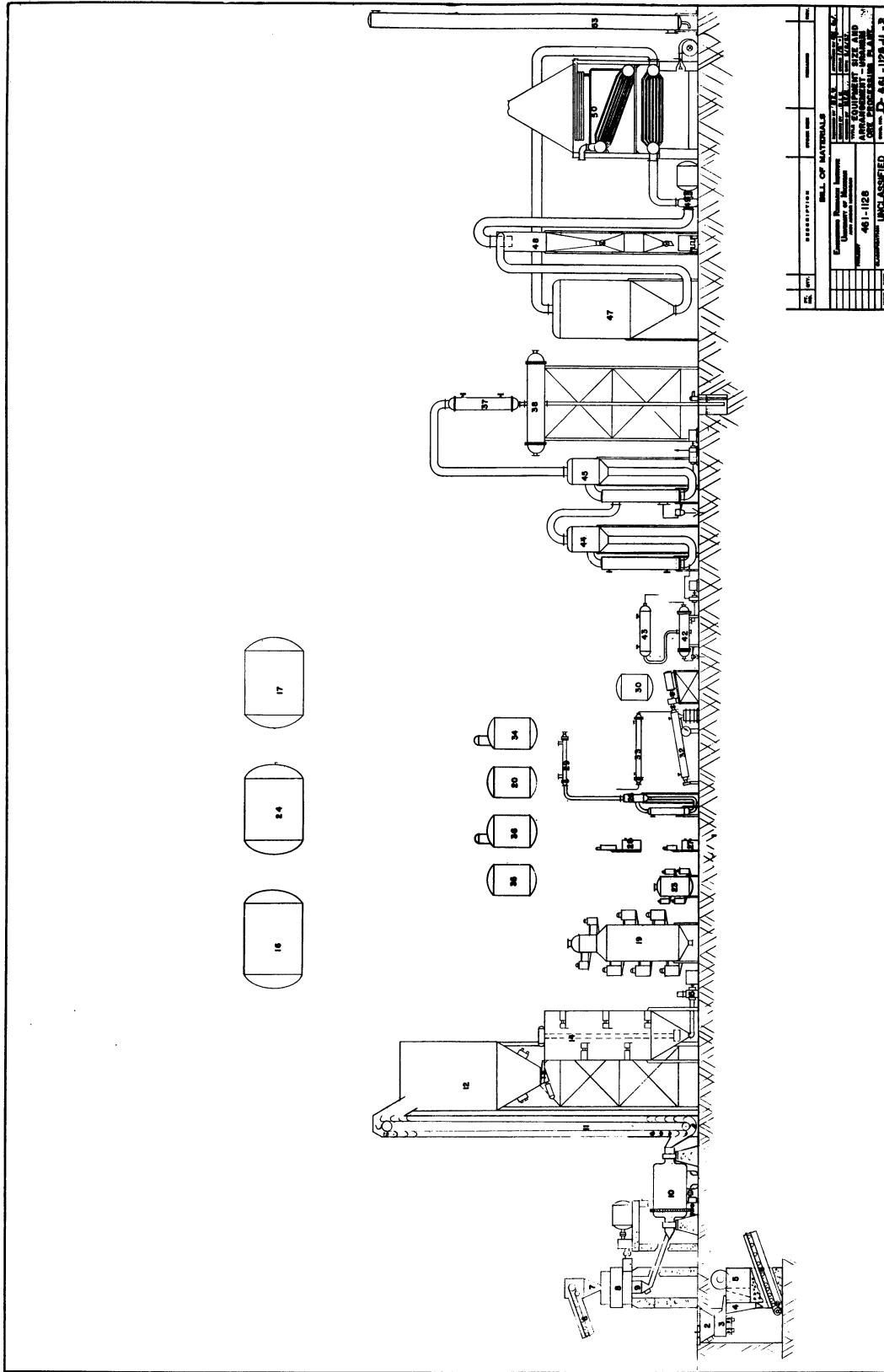
When the ores being treated are pitchblend, the leaching time is apparently quite different,⁷ times of up to 16 hr being required depending upon acid, oxidant strength, and temperature. However, some essential data regarding mesh size and method of leaching are missing from this reference. Since most ores to be processed will be western ores, an average 10-mesh particles size was selected, recognizing that the information upon which this selection is based is meager and that additional data on leaching rates for all the varieties of ore being processed will be required.

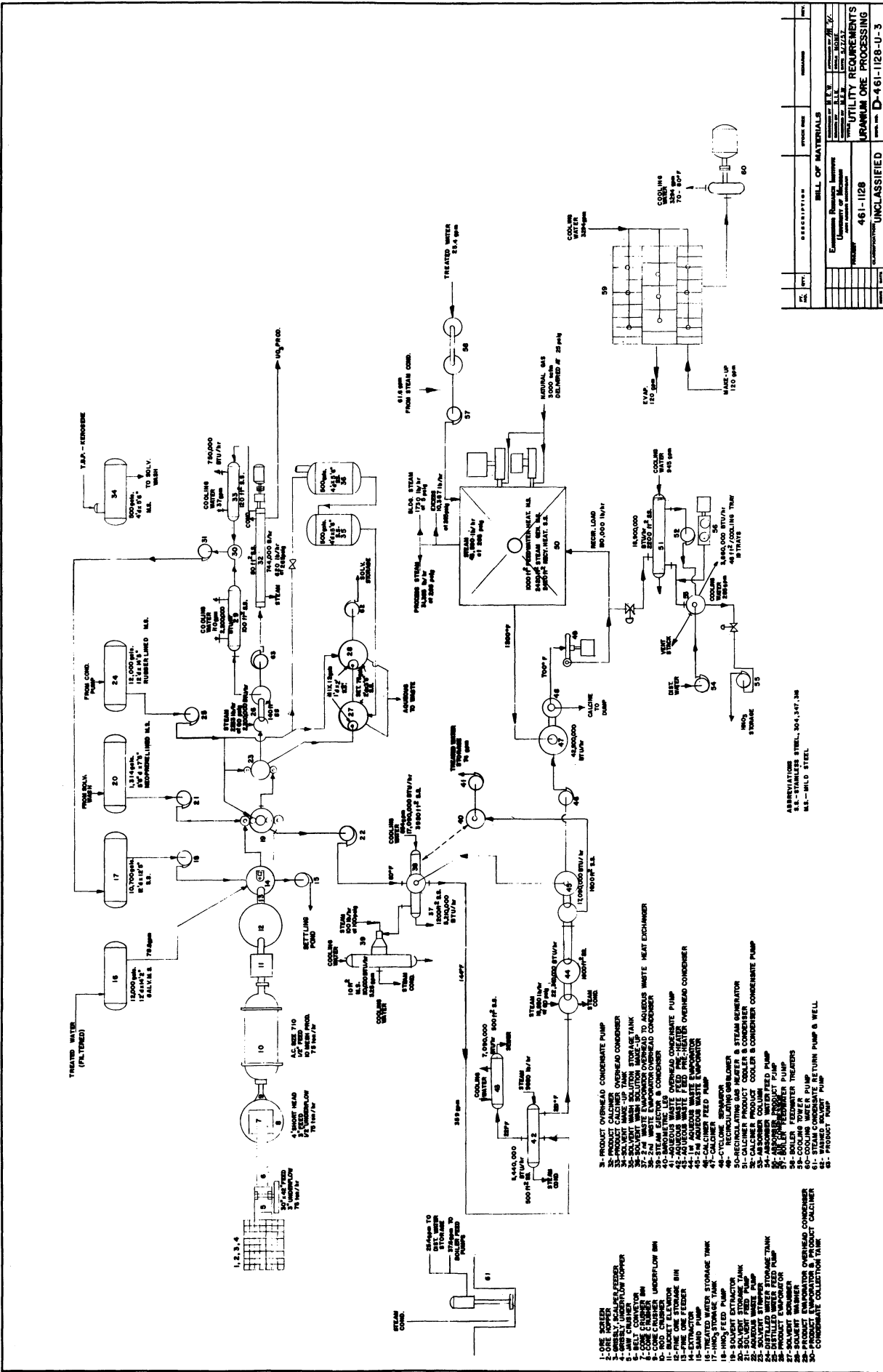
The equipment selected in this study is shown in Dwg. 461-1128-U-2 and U-3. Equipment numbers given here refer to these flowsheets. The equipment is composed of a coarse ore screen (1), equipped with electrically driven vibrator, through which the ore is dumped from carryall loaders. A combination scalper, feeder, grizzly unit (3) feeds the ore from the ore bin (2) into the grinding system. This scalper feeds the coarse ore into the jaw crusher (5) while pieces smaller than 3 in. fall through the hopper (4) and onto the conveyor belt (6). The underflow from the jaw crusher falls directly onto the conveyor belt. It is assumed that this section of the ore-crushing equipment composed of items (1) through (5) is located in close proximity to the ore storage areas. The belt conveyor with a length of approximately 125 ft conveys the crushed ore to the subsequent grinding equipment.

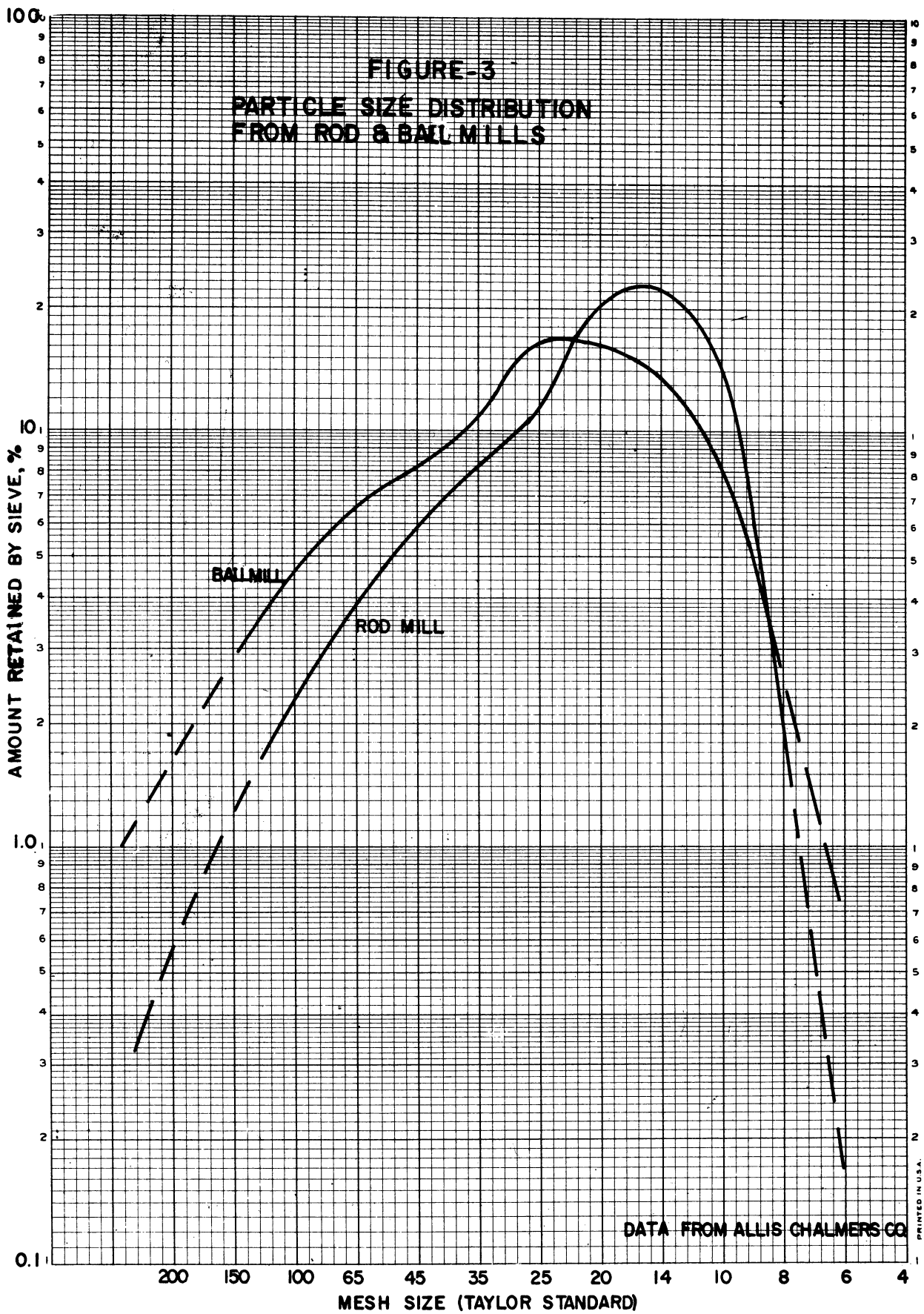
The discharge from the belt conveyor falls into bin (7) which guides the ore into the core crusher (8). This crusher reduces the ore size from a maximum of 3 in. to approximately 1/2 in. with the fines falling into the electrically vibrated bin (9). This bin feeds the ore into the rod mill (10), which reduces the particle size from 1/2 in. to a nominal -10 mesh. A bucket elevator (11) conveys the ground ore into the fine ore storage bin (12), where a hold-up of 4 hr is provided to allow some operating flexibility between the crushing grinding equipment and the subsequent extraction steps. All the crushing grinding equipment has been sized for greater than a 1000-ton-per-day capacity, with these sizes being the basis for subsequent pricing.

The selection of a jaw crusher rather than a gyratory was made on the basis of a lower first cost. This selection is subject to review if operating costs indicate that a gyratory is more desirable.

A rod mill was selected for final grinding rather than a ball mill since the spread in mesh size from a ball mill is narrower. This is illustrated in Fig. 3. The data in this figure were furnished by the Allis-Chalmers







Manufacturing Company. It is evident that the ball mill grinding to a nominal -8 mesh particle size would result in 1.6% passing through a 200-mesh screen. There would also be about 2.3% retained on a number 8 screen. The rod mill would give about 0.6% -200 mesh and a smaller proportion of oversize particles. It is important that the quantity of solids of -200 mesh be held to as low a value as possible. The reasons for this will be discussed in the section on leaching.

Since it is unimportant in subsequent steps just what the particle-size distribution is, as long as the -200 mesh fraction is minimized and the +8 mesh fraction is not excessive, there is no need for a recirculation system and consequently no closed circuit grinding is required. Obviously this simplifies the grinding section of the plant and reduces the over-all plant cost.

3.3.2 Leaching.—The ground ore is transported by a bucket elevator (11) into a fine ore storage bin (12). This bin is sized to provide a 4-hr ore hold-up at full plant capacity so that some degree of flexibility will exist in matching the crushing and grinding rates to the plant throughout.

The fine ore is fed from this bin by a vibratory feeder into a continuous leaching extractor (14) where the ore flows downward and the leaching liquor flows upward. The basic design of the ore extractor is illustrated in Fig. 4. The ore extractor consists of a tall cylindrical vessel with a conical bottom. Down the center of this vessel is a shaft to which is attached the scraper blades. As the shaft rotates, the solids are scraped toward the outside of the tray and fall over the edge and into the conical chamber below. Solids pile up at the downcomers from each tray to form a seal to prevent liquid by-passing. The solids are routed downward from tray to tray in this manner until they are discharged out of the bottom of the vessel as a slurry. Leaching acid enters below the fourth conical section from the top and combines with part of the wash water which flows upward after being introduced at the bottom conical section. The combined solution flows upward with alternate agitation and settling at each conical section until the solution runs out of the overflow at the top of the vessel. Sufficient wash water is introduced at the bottom to furnish adequate slurry liquid for discharge of the spent ore sands. Thus the bottom two conical sections are washing stages and the upper three sections serve as leaching stages. The size selection of this unit will be discussed in the following paragraphs.

The problems in ore leaching in this type of apparatus are in matching the dissolution rate with the unit throughout and at the same time keeping the velocity of the liquid through the unit low enough so that fines carried over from the ore are not excessive.

Saine and Brown⁶ give data on the leaching rate of a number of western United States ores. Their data on ore from the Uravan district in Colorado are summarized in Table I. With the exception of experiment B-16

TABLE I
Effect of Mesh Size on Direct Acid Leaching of
Raw Carnotite Ore
(80°C Leaches)

Exp. No.	Ore Mesh	1st Leach		2nd Leach		Tails			
		U %	V %	U %	V %	U		V	
		of Total	of Total	of Total	of Total	% of Wt	% of Total	% of Wt	% of Total
B-15	-4	96	42	0.5	7	0.004	4.0	0.485	51
B-16	-8	93	36	1.0	6	0.015	6.0	0.525	58
L-365	-20	97	42	0.5	11	0.004	2.5	0.445	47
B-17	-60	97	32	0.5	13	0.004	2.5	0.520	55
B-18	-80	97	36	1.5	12	0.002	1.5	0.500	52
B-19	-100	98	37	0.4	10	0.002	1.5	0.495	53

Average analysis of head samples:

U₃O₈: 0.15%

V₂O₅: 1.7%

Ore Sample: 100 g of raw Uravan ore

Acid Leach: 200 ml of 5% HCl agitated 4 hr at 80°C

in these data, it apparently makes some difference whether the ore is -4 or -100 mesh. The second leach contributes to over-all uranium recovery but not to any great extent.

If sample B-16, by comparison with the rest of the runs, is indicative of the experimental errors in the data, then the conclusion might be reached that mesh size is unimportant in leaching this particular ore. However, if B-16 were discarded as being nonrepresentative, then grinding to -100 mesh would represent a 2.5% gain in the uranium recovered from the tails.

It should be noted that these data are based on batch leaching of the ore and not on a continuous countercurrent system. While batch data are indicative of gross recoveries that could be achieved, there should be

some fundamental advantages in a countercurrent system which would result in greater uranium recoveries.

There remains the question of how much leaching time is required per stage. Saine and Brown⁶ also investigated this and their results are given in Table II. In this table, runs for equal acid concentrations are compared for leaching times varying from 1/2 to 4 hr. It can be seen that the quantity of uranium leached from the ore does not vary significantly for times within this period. This is evident particularly in runs S-375, L-364, and L-365. The data also show that the acid used for leaching makes little

TABLE II

Time Dependence of Direct Acid Leaching of Raw Carnotite Ore
(80°C Leach)

Exp. No.	% Acid	Leach Time (hr)	1st Leach		2nd Leach		Tails			
			U %	V %	U %	V %	U		V	
			of Total	of Total	of Total	of Total	% of Wt	% of Total	% of Wt	% of Total
L-350	1% H ₂ SO ₄	1	90	7	3	8	.009	7	0.790	85
L-351	1% H ₂ SO ₄	4	90	12	6	12	.006	4	0.560	76
L-352	5% H ₂ SO ₄	1	95	30	0.7	5	.006	4	0.550	65
L-353	5% H ₂ SO ₄	4	96	39	0.4	10	.005	4	0.510	51
L-355	20% H ₂ SO ₄	4	96	51	0.3	9	.005	4	0.400	40
L-363	1% HCl	4	90	--	5	--	.006	5	-----	--
S-372	2% HCl	2	93	27	2	8	.006	5	0.560	65
S-375	5% HCl	1/2	95	22	0.7	7	.005	4	0.565	71
L-364	5% HCl	1	95	22	1.5	9	.005	4	0.665	69
L-365	5% HCl	4	97	42	0.4	11	.005	3	0.445	47
L-367	20% HCl	4	96	--	0.3	--	.005	5	-----	--

Ore Sample: 100 gm of Uravan ore ground to -20 mesh.

Leach Description: 300 ml of leach solution per 100-gm ore at 80°C.

difference in the recoveries attained. For this reason no differences are expected in reaction times when HNO_3 is used rather than HCl , or H_2SO_4 .

On the basis of these data three countercurrent leaching stages were selected having a solids hold-up of 10-15 min per stage. It may be necessary to increase this time when more data are available on a wider variety of ores.

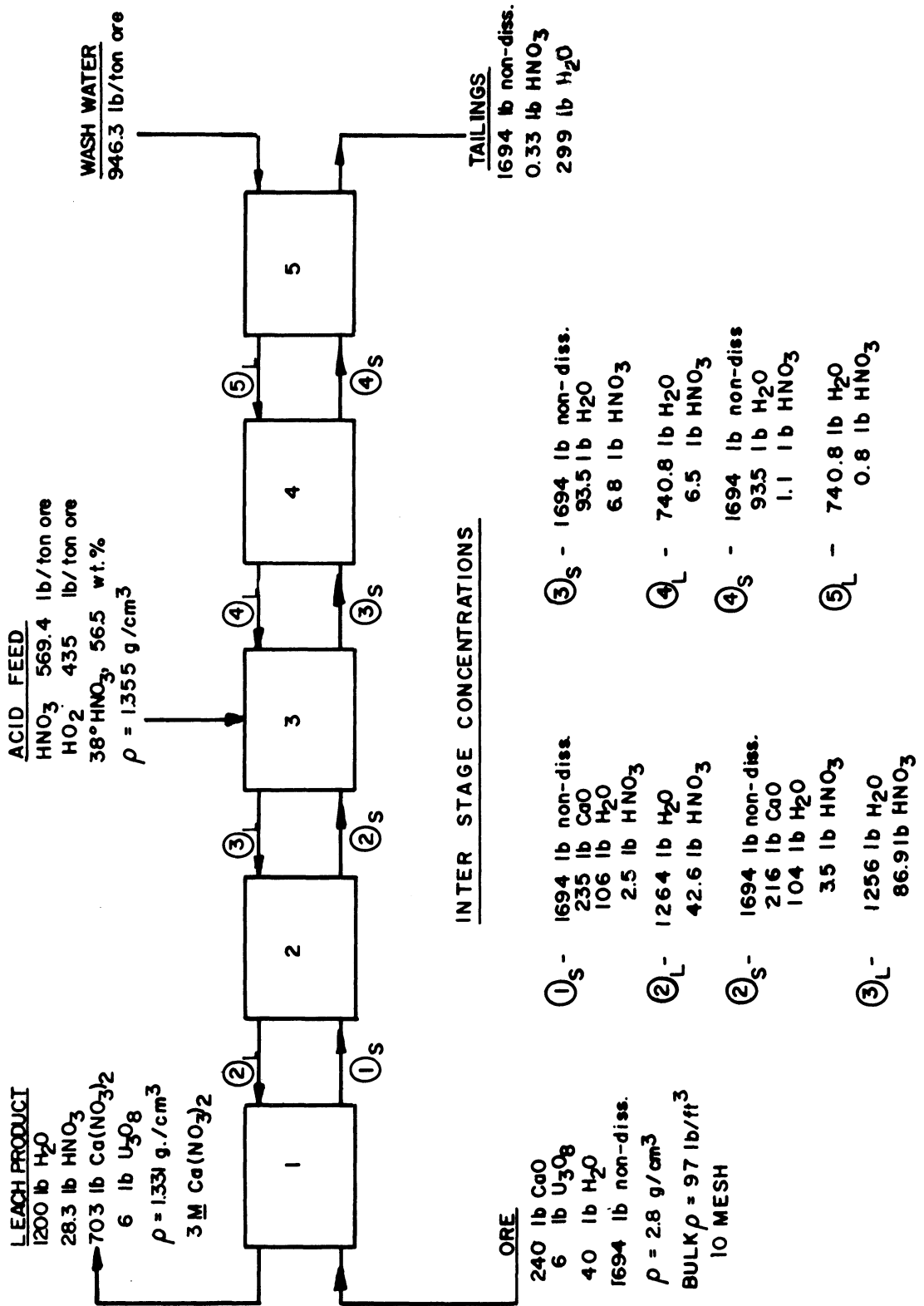
It is interesting to note that Saine and Brown found carnotite ore very easy to grind down to -80 mesh, whereas extensive ball-mill grinding was necessary to reduce the size to -100 mesh. This supports the author's opinion that the process proposed here would result in significantly lower grinding costs than conventional processes.

The leaching conditions were selected on the basis of the available data. It was necessary, of course, to alter the acid concentration in the leach liquor to higher concentrations to permit dissolution of all the lime in the particular type ore selected. This change should result in greater rather than lower recoveries. The conditions selected were used in a stage-to-stage leaching calculation to determine probable composition of the liquid in each stage and to determine the number of wash stages required to remove essentially all the acid from the sands. The results of this calculation are portrayed in Fig. 5. In this calculation only the lime was considered as a dissolvable component since only its performance was completely predictable at this time.

Most of the lime is dissolved in stage 3, where fresh acid contacts the solids coming from stage 2. It is assumed that this will be true of the uranium in the ore also. It should be noted that there is still unreacted HNO_3 present in the leach product, so leaching will also occur in stages 1 and 2. This is shown also in the decrease in lime concentration in the solids stream from stages 1 and 2. This calculation has also been valuable in predicting the number of washing stages required. It is evident that two such stages, 4 and 5, will reduce HNO_3 losses to a negligible quantity.

The internal area of the countercurrent leaching equipment is adjusted so that the upward flow of the leach liquor does not carry appreciable quantity of fines with it. This liquid velocity was set at a nominal value of 5×10^{-3} ft/sec. Even with this low velocity, solid particles below -200 mesh would be carried out of the leaching equipment and into the solvent-extraction apparatus. This is the chief reason for selecting grinding equipment that will give a minimum of fines in the -200 mesh region. Referring to Fig. 3, it is evident that up to 0.6 wt % of the ore could be expected to be carried over when a rod mill is used as the final grinder. This may be a pessimistic figure since some filtering action would occur as the liquor

FIGURE - 5
COUNTER CURRENT LEACHING OF ORE
CALCULATION RESULTS



passes through the ore fines. Certainly, however, any plant design should be based upon some anticipated solids carryover.

Several other types of equipment could be considered for the leaching application discussed above. It is the customary practice in the present plants to use large agitated vessels for leaching. These vessels are charged with the ground ore and the appropriate solution pumped in, and the ore and solution agitated for as long as 8 hr. The ore is allowed to settle and the supernate routed to ion-exchange extraction apparatus. The ore sands are then washed, resettled and decanted, and the sands pumped out. Present plants do not generally use countercurrent leaching of more than two stages with just one stage being the rule rather than the exception. The disadvantages of this system are the very large tanks required and the comparatively low throughput compared to the continuous type of equipment for equivalent capital investment.

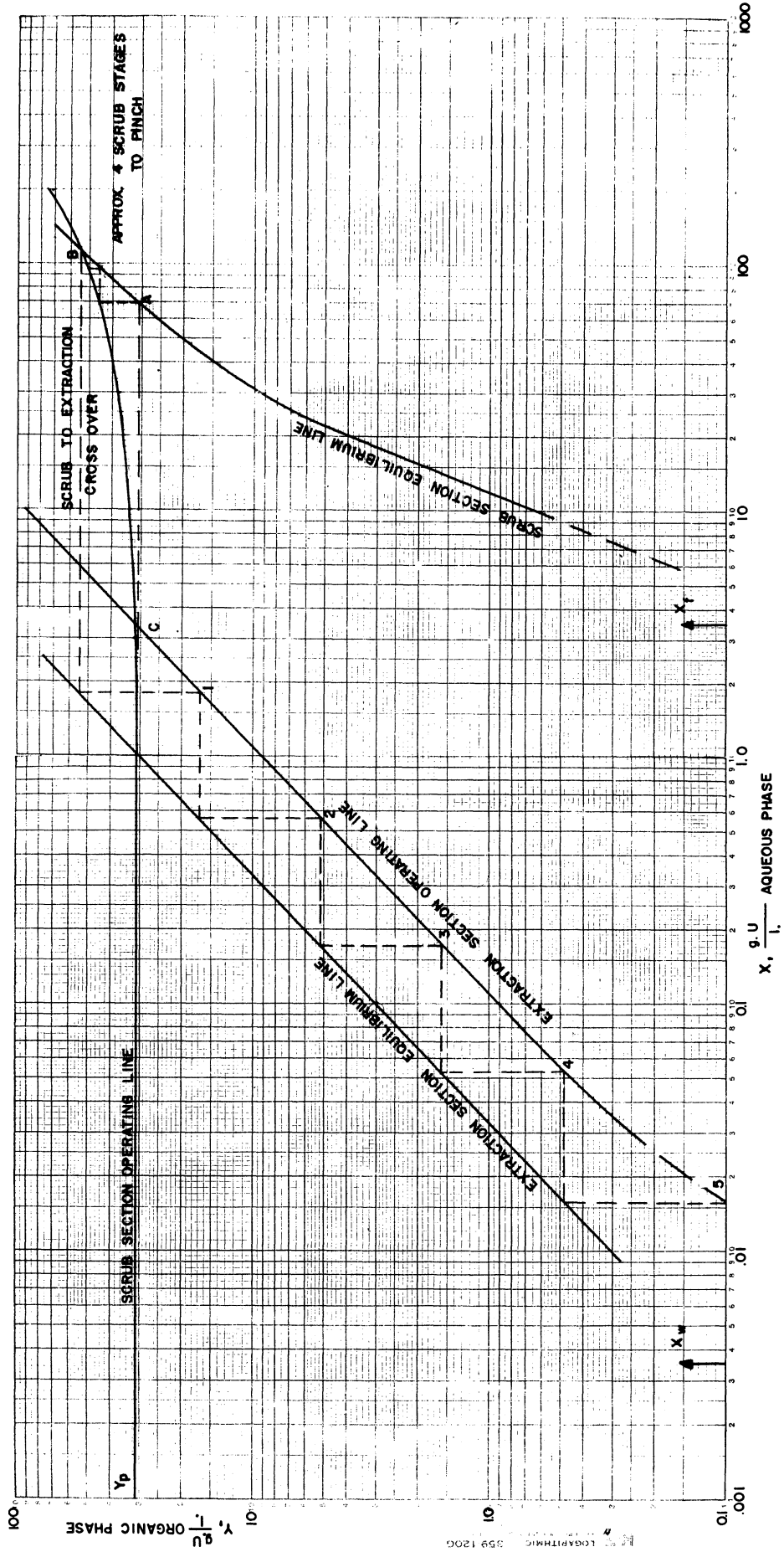
Another type of equipment that could be used is a percolation tower where ore is fed into the top of a tall column and the leach liquor percolated up through the fines. The conventional Dorr-Oliver leaching equipment could also be used. This type of equipment is not normally enclosed, so covers would have to be placed over it to prevent the escape of NO₂ fumes.

Comparing the advantages and disadvantages of all the possible types appears to indicate that the type portrayed here has a higher throughput and is comparable in first cost to other types. Maintenance on the unit selected should not be significantly more than on the other types with the possible exception of the simple tank-agitator variety.

3.3.3 Solvent Extraction and Stripping.—The rich leach liquor overflows from the extractor (14) to the solvent extractor (19). Thus the solvent-extraction aqueous feed rate is actually controlled by the controllers on the acid and wash streams to the extractor (i.e., leaching apparatus).

The solvent-extraction equipment contains five extraction stages and three scrub stages. The function of the extraction section is to extract uranium from the aqueous phase into the organic phase. Extraction efficiency of uranium is measured by a parameter called a distribution coefficient. This coefficient is the ratio of uranium concentration in the organic phase to the uranium concentration in the aqueous phase, using consistent units. The value of this coefficient varies with the nature of the organic extractant, the nitrate concentration in the aqueous phase, and the uranium concentration in both phases. The effects of other components in the system are relatively minor compared to those mentioned above. Distribution ratios are given for a wide variety of conditions by Moore⁸ and Granquist and Merrill.⁹ The data presented in Refs. 8 and 9 permit the calculation of extraction and stripping stage requirements in this process. This calculation is shown in Fig. 6 and will be discussed in later paragraphs.

FIGURE-6
 STAGE TO STAGE CALCULATIONS FOR THE URANIUM
 IN THE EXTRACTION-SCRUB SECTIONS OF THE
 SOLVENT EXTRACTION APPARATUS



Unfortunately, while the extraction of uranium by this system is very efficient, it is not highly selective. Some metal ions present in the rich leach liquor will also extract. Metal ions of vanadium, iron, titanium, and also boron extract to some degree. These contaminant ions would result in an impure uranium product if they were not removed from the process. Removal is accomplished by the scrub section in the solvent-extraction apparatus. Since the distribution coefficient of these ions is not as high as for the uranyl ion, removal can be accomplished in the multi-stage counter-flow apparatus by scrubbing the uranium-containing organic phase with a small stream of water. After scrubbing, the uranium in the extractant (organic phase) is in a pure state and becomes a very suitable product.

A calculation of stage requirements in this apparatus is shown in Fig. 6. The operating lines in this figure are arrived at by taking material balances around the bottom (extraction section) and top (scrub section) of the solvent-extraction apparatus. A uranium distribution ratio organic to aqueous of 30 was assumed in the extraction section. This value appears justified from the data in Refs. 8 and 9. In the scrub section, where no $\text{Ca}(\text{NO}_3)_2$ is present, equilibrium is not constant but varies with uranium concentration. These data were taken directly from the references above. The stages are stepped off on the diagram as dotted lines. The number of stages required in the scrub section is indeterminate since a "pinch" exists in this section. Three scrub stages is well into the pinch and this is believed adequate. In the extraction section, the stages can be easily determined since no pinches exist. The stages required in this section are counted down from the point where the scrub equilibrium and operating lines intersect or $Y = 54.5 \text{ g U/l}$. Six stages are adequate to reduce losses to 0.1%.

It is interesting to note the amount of uranium reflux that occurs in the scrub section and the magnitude of the uranium concentration change between the scrub and extraction changes.

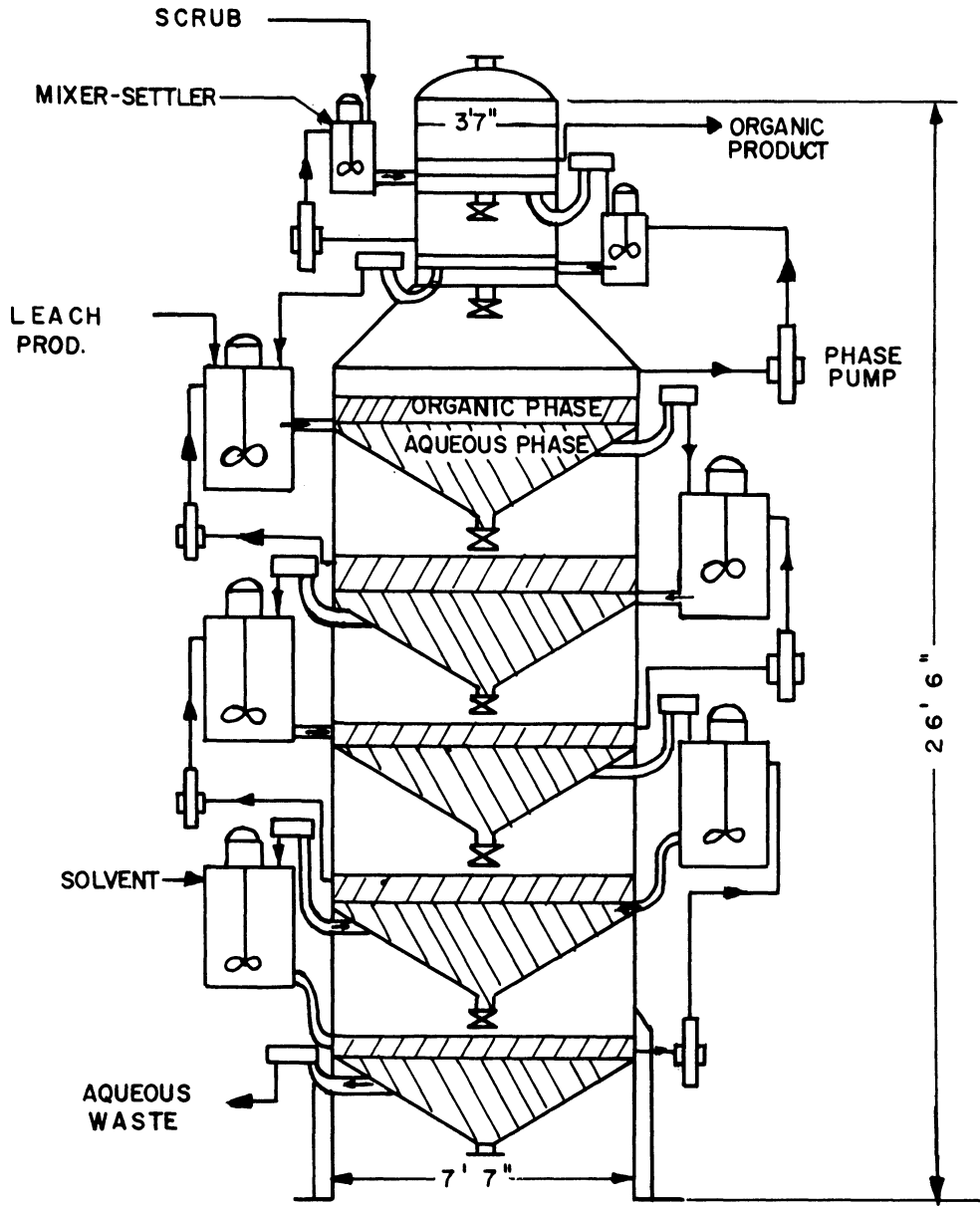
The maximum uranium concentration in the solvent-extraction apparatus occurs at point B, which is just above the feed point. At this point the uranium concentration is 116 g/l in the aqueous phase and 54.5 g/l in the organic phase. At the feed point the uranium concentration is about 6 g/l, which is where the extraction section begins. Thus the reflux in the scrub section varies from 116 to 6 g/l. It is this refluxing action that is important in removing the impurities that are extracted from the uranium. Product and concentrations are represented by point A on the diagram where $Y_p = 31.5 \text{ g/l}$ and is the product stream concentration leaving the apparatus. At the same point but in the aqueous phase, the uranium concentration is 79 g/l. The scrub and extraction section operating lines should intersect at the feed stream concentration if the extraction diagram is drawn correctly. This intersection is shown as point C and the aqueous phase concentration at this point is 3.5 g/l which checks the flowsheet

feed concentration. It is also necessary for the first stage in the extraction section to include this concentration. Clearly stage 1 does this, as shown by the dotted line. Uranium losses from the extraction section are represented by the magnitude of X_w . In Fig. 6, the flowsheet value of .0035 g/l is shown. This concentration would occur somewhere between stages 5 and 6 in the apparatus. Five stages are shown on the flowsheet and in the drawing on the extractor. It would be advantageous in a final design to increase this to 6 stages based upon Fig. 6.

The stripping apparatus stage requirements are much simpler to compute since this apparatus is not complicated by a compound function such as scrubbing and stripping. No computation of stages is included here because of its simplicity. Calculations by the same methods as in Fig. 6 indicate that 3 stripping stages are adequate. It is advisable to include extra stages so 3 to 5 stages are shown on the flowsheet.

The selection of equipment for extraction and stripping has been left open up until now. Among the possible choices are packed columns, spray columns, or pulsed columns. Other types that could be considered are the Schiebel contactors, and the Fenske contactor. Various types of mixer settlers are also available as well as the high-speed centrifugal type of contactors. From the standpoint of reliability, simplicity, and ruggedness, the choice for this application appeared to be between columns of various types and mixer settlers. Most columns operate at maximum rates of 1000 gal/hr ft² sum of both phases which for these flow rates would set the column diameter at 3.2 ft. Since there is no way to predict theoretically the height of a stage in a column, experimental data must be available before a height can be set. The largest column known by the author to have been operated on this system was 2.0 ft in diameter. Extrapolation from 2.0 ft to 3.2 ft was considered unwarranted without pilot plant information. Consequently, a simple mixer-settler system was chosen which requires a minimum of extrapolations and pilot plant information. Tests were made on the equilibrating times, settling times, and maximum agitation for this system, and from these data an average time of agitation of 30 sec and a 5-min settling time were selected as a design basis.

A tentative design for the solvent extractor is shown in Fig. 7. The body of this unit contains conical separators which serve as the settling sections between stages. The bottom of these cones is fitted with valves so that solids being accumulated from leaching (ore extraction) carryover can be flushed from the unit. On the outside of this unit are mounted the interstage mixers, which are small cylindrical vessels with a flange-mounted propeller-type agitator mounted on the top. The aqueous feed (leach product) enters the mixer on the left and is mixed by the agitator with organic solvent being pumped up from the settling section below where the two phases separate. The aqueous phase flows over the weir down into the next mixer, where it is mixed with organic solvent from the third settling section where the sequence is repeated.



ALL DIMENSIONS UNLESS OTHERWISE SPECIFIED MUST BE HELD TO A TOLERANCE - FRACTIONAL $\pm \frac{1}{64}$ " DECIMAL $\pm .005$ " ANGULAR $\pm \frac{1}{2}^\circ$

ENGINEERING RESEARCH INSTITUTE UNIVERSITY OF MICHIGAN ANN ARBOR MICHIGAN		DESIGNED BY M.E.W.	APPROVED BY
		DRAWN BY R.I.K.	SCALE $1/4" = 1'$
		CHECKED BY M.E.W.	DATE 5/27/57
PROJECT 461-1128		TITLE SOLVENT EXTRACTOR	
CLASSIFICATION UNCLASSIFIED		FIGURE-7	
ISSUE	DATE		

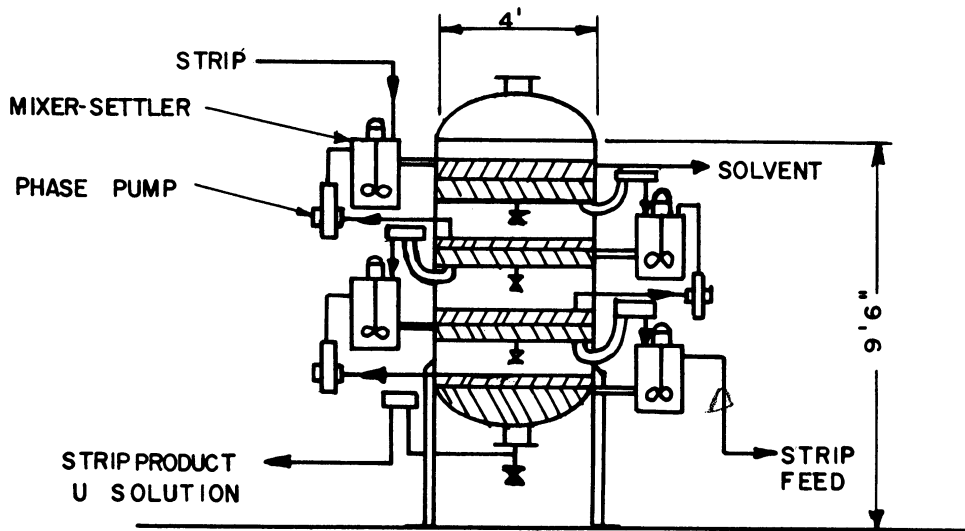
Thus, the aqueous flow proceeds downward and the organic or extractant flow is upwards. Two stages are shown as an illustration for the scrub section, and since the aqueous flow is much smaller here, the over-all size of the equipment is reduced. It is to be noted also that the organic phase is pumped between stages while the aqueous flow is downward by gravity. The solvent extractor is item 19 on the flowsheets D-461-1128-U-2 and U-3.

The organic stream leaving the solvent extractor contains uranium at approximately 31 g/l. It is necessary to transfer this uranium to an aqueous phase before subsequent treatment can take place. This transfer is accomplished in a solvent stripper (23). This piece of equipment is very similar in construction and operation to the solvent extractor except that it is smaller in size and contains fewer stages, as shown in Fig. 8. Since the NO_3 ion concentration is low in this apparatus, the uranium is preferentially extracted into the aqueous phase. The actual equilibrium conditions for this situation are the same as for the scrub section of the solvent extractor except operation is in the more dilute regions. Referring to Fig. 6 for the scrub-section equilibrium line, it is evident that, for uranium concentrations below 30 g/l (aqueous phase), extraction would be preferentially from the organic to the aqueous phase. In the solvent stripper the 19.2 gal/ton of organic phase is contacted in a 3-to-5 stage countercurrent operation with 6.4 gal/ton of distilled water. The uranium-rich aqueous phase flow from the solvent stripper to the product evaporator (26) and the uranium-free organic phase flows to the solvent washing apparatus (27) and (28).

Stage requirements in the stripper were calculated in the same manner as for the solvent extractor, but here the calculations are much simpler since the stripper does not serve a dual purpose. Agitation and settling times within the stages are tentatively the same as for the extractor.

3.3.4 Solvent Treating.—Since tributyl phosphate is an ester of n-butyl alcohol and orthophosphoric acid, it reacts in a manner typical of these compounds. It can decompose by hydrolysis giving mono- and di-butyl phosphates as well as free phosphate ion and alcohol. These decomposition products are undesirable since they interfere with subsequent extraction and stripping of uranium, and so must be reduced down to acceptable levels. In addition, the kerosene diluent has a tendency to nitrate at the double bonds. These nitrate compounds appear to have detrimental effects during the extraction-stripping operations, and so must also be removed.

Removal of these undesirable components is accomplished by mixing the organic phase with a 5-10% solution of sodium hydroxide, settling and decanting the phases, and then repeating the operation using a water wash. These washes are done in the solvent scrubber (27), using the caustic, and the solvent washer (28), using the water. The residence time in the scrubber



ALL DIMENSIONS UNLESS OTHERWISE SPECIFIED MUST BE HELD TO A TOLERANCE - FRACTIONAL $\pm \frac{1}{64}$ " DECIMAL $\pm .005$ " ANGULAR $\pm \frac{1}{2}^\circ$

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		PROJECT	461-1128	TITLE			
		CLASSIFICATION	UNCLASSIFIED	SOLVENT STRIPPER			
ISSUE	DATE			FIGURE-8			

and washer were tentatively set at 5-min agitation and 30-min settling. Flow ratios of 10 organic to 1 wash solution were set by past experience.

Caustic and water washes are combined and discarded with the organic extractant routed to a solvent make-up storage tank (34).

It is pertinent at this time to discuss solvent losses in the solvent extraction section of this process. Cost of replacing lost kerosene will not be significant, but since tributyl phosphate costs 50 cents/lb (\$4.10/gal less freight) in tank car lots at the factory, significant losses of this chemical could be a serious economic factor.

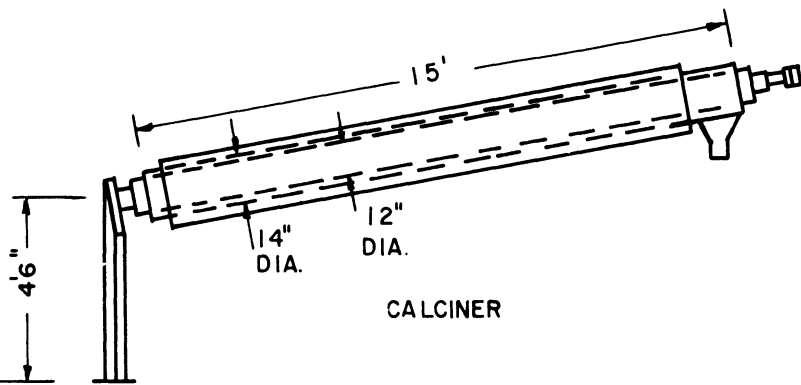
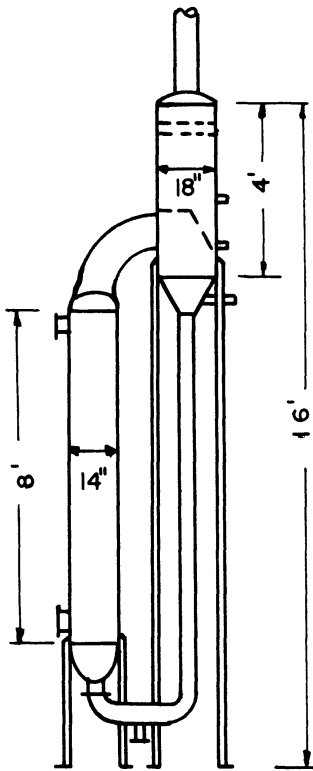
Solvent losses can occur by three different mechanisms: 1) solubility of solvent in aqueous solutions contacting it, 2) hydrolysis or decomposition, and 3) entrainment losses in one of the waste aqueous streams. It has been found by experiment that solubility losses are substantially greater than hydrolysis losses. From an organic phase containing 10 vol % TBP in kerosene and equilibrated with an aqueous phase, the TBP will dissolve to the extent of 0.14 g/l in the aqueous phase. Taking the sum of all aqueous streams contacting the organic phase (183 gal/ton ore), the TBP losses are seen to be about 97 g TBP/ton ore, or about 1/4 lb. This is small enough to be almost negligible; however, solubility losses can be greatly exceeded by losses due to maloperation or poor design of the equipment. At the present time, these losses are not known, but it is definite that TBP losses can approach about 1/4 lb/ton ore under proper conditions. Kerosene losses will amount to only 1-1/2 lb/ton ore as a minimum.

3.3.5 Product Evaporation and Calcining.—The product stream from the solvent stripper contains approximately 94.5 g U/l which must be concentrated and de-nitrated into an oxide to become an acceptable product. This is done in the product evaporator (26), and in the product calciner (32). Sketches of these items are shown in Fig. 9.

The product evaporator receives 2430 lb/hr of solution from the product stripper, and concentrates this solution until its composition is roughly that of $\text{UO}_2(\text{NO}_3)_2 \cdot 6\text{H}_2\text{O}$ with a boiling point at atmospheric pressure of 244°F. Using 60-psig steam, the Δt across the heating surface will be 64°F. Heat load in this unit is 2,220,000 BTU/hr and the area required 140 ft².

A thermosyphon type evaporator was selected since circulation is fairly rapid, and excellent control and uniformity is achieved without external pumps. It is characteristic of this solution that little or no scaling occurs on the heat transfer surfaces. The overhead vapor from the product evaporator is condensed in the product-evaporator overhead condenser (29). This unit is of standard tube shell, single-pass in tube construction.

EVAPORATOR



CALCINER

ALL DIMENSIONS UNLESS OTHERWISE SPECIFIED MUST BE HELD TO A TOLERANCE - FRACTIONAL $\pm \frac{1}{64}$ " DECIMAL $\pm .005$ " ANGULAR $\pm \frac{1}{2}^\circ$

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<p>PROJECT 461-1128</p>		CHECKED BY	M.E.W.	DATE	5/24/57
		<p>TITLE URANIUM PRODUCT EVAPORATOR & CALCINER</p>			
<p>CLASSIFICATION UNCLASSIFIED</p>		<p>FIGURE-9</p>			
ISSUE	DATE				

Entrainment of concentrated liquor in the product-evaporator overhead can be held to 10^{-5} lb liquor/lb overhead by the design of the de-entraining section. Losses of uranium due to entrainment are thus negligible.

Concentrated $\text{UO}_2(\text{NO}_3)_2 \cdot 6\text{H}_2\text{O}$ solution flows from the product evaporator to the product calciner where the uranyl nitrate is converted to the desired UO_3 . This conversion occurs at about 400°F and requires 1650 BTU/lb feed heat input. About 695,000 BTU/hr is required to dewater the feed with the remainder used to decompose the nitrate. Since much of the heat transfer is to a granular solid, and caking or lump formation is undesirable, a steam-heated jacketed screw-type calciner was tentatively selected. Other calciners, such as the vertical tray type common in ore roasting, a fluidized bed unit, a rotary type kiln, and possibly other types could be considered for this service. It would be desirable to reabsorb the NO_2 and O_2 formed from the uranyl nitrate decomposition into water for acid recovery, and for this reason heating with combustion gases appears impractical due to the large dilution and subsequent absorption difficulties of the NO_2 . Dowtherm, or molten salt, and direct fired heating outside a jacket would be worthy of consideration. In this unit, steam at 250 psia was selected for heating since this is the generation pressure from the boilers. Both the screw and the jacket are steam-heated. The screw will make approximately 6.2 revolutions/hr to deliver the 254 lb/hr of UO_3 .

The UO_3 product falls directly from the calciner into a loading hopper equipped with an intermediate gas lock and is charged directly from this hopper into a scale-mounted drum for final weighing, sampling, and packaging.

The evaporated water, NO_2 , and O_2 evolved from the calcining is condensed in the product calciner overhead condenser (33). Item 33 is deliberately oversized from the heat-transfer standpoint so that liquid and gas hold-up times will be long. This gives the NO_2 and O_2 long enough to be reabsorbed into the condensed water giving HNO_3 . About 80% of the NO_2 should be absorbed in the condenser. The remaining gas is discharged out of the stack since it is small in quantity, and it is not considered economical to compress this gas so that it can be totally absorbed in the absorption tower later in the process.

The condensate from the product-evaporator overhead condenser (29) and the product calciner overhead condenser (33) is collected in a tank (30) and is pumped back into the system or can be discarded.

3.3.6 Waste Evaporation and Calcining.—The aqueous waste stream from the solvent extractor (19) contains all the components of the ore that are soluble in nitric acid. These would be largely the lime and to a lesser degree the iron, vanadium, and aluminum. If this stream were to be discarded, large losses of nitric acid would result, particularly when the ore being treated has a high lime content. These large losses would occur

whether the leaching acid is sulfuric, hydrochloric, or nitric. The major advantage of this process lies in its ability to recover the acid in the waste stream, and is the chief justification for using nitric acid for leaching because this acid can be easily recovered by the proposed methods.

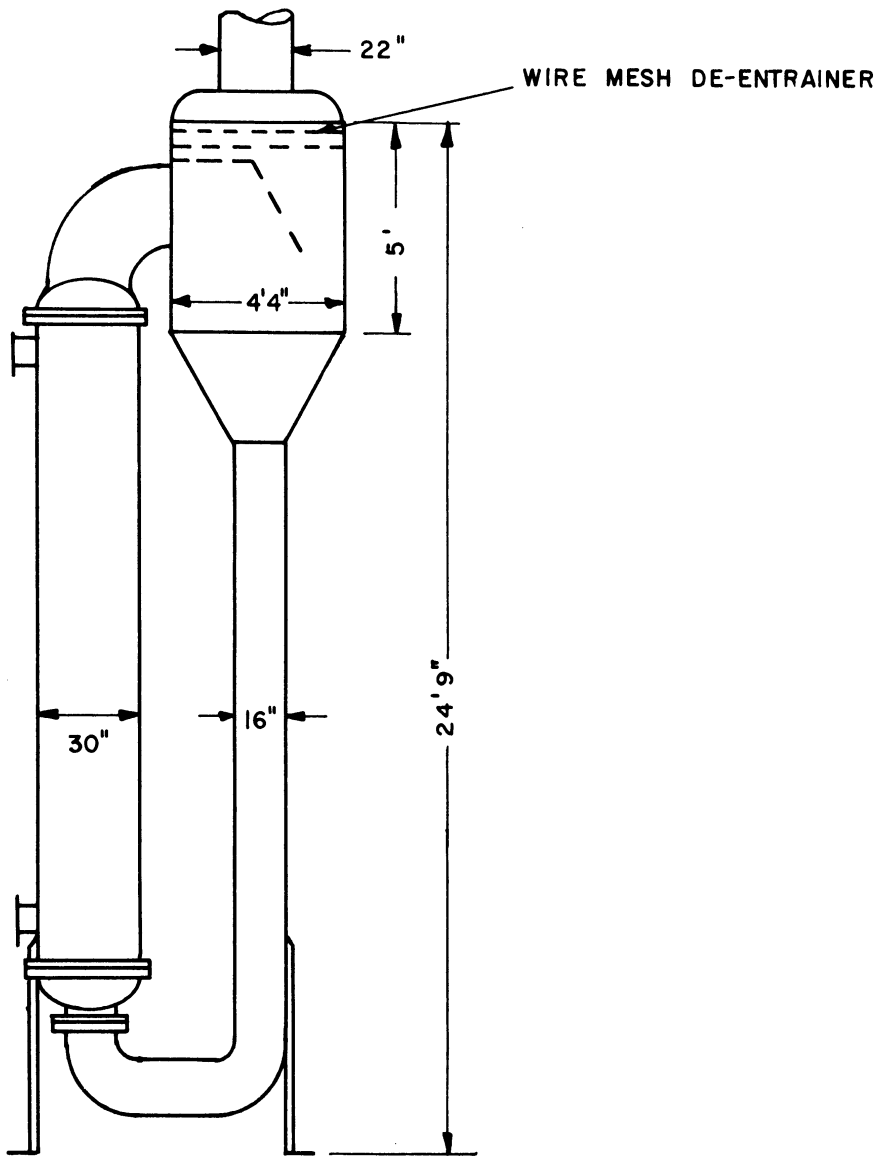
The aqueous waste stream leaves the bottom of the solvent extractor (19) and is pumped by pump (22) to the second waste evaporator overhead to aqueous waste heat exchanger (37). Here part of the second evaporator vapor is condensed, and exchanges its heat with the waste stream heating it from 80°F to 144°F. About 17,090,000 BTU/hr of heat are saved by this interchange. The waste stream then goes to the aqueous waste feed preheater (42) where it is heated to the boiling point and some 5% of the stream is evaporated. The purpose of item 42 is to: 1) volatilize off any Cl^- that may be contained in the stream and thereby decrease corrosion of the subsequent evaporation heat transfer surfaces; 2) volatilize off any organic material that may be dissolved or entrained in the waste stream, and 3) precipitate any SO_4 present as CaSO_4 that has a negative temperature solubility coefficient on the theory that item 42 can be more easily cleaned or spared if necessary than the subsequent evaporators.

Overhead from the aqueous feed preheater (42) is condensed in the aqueous waste feed preheater overhead condenser (43). This condensate is then discarded.

The aqueous waste stream then flows to the waste evaporators at its boiling point, 221°F. The waste stream evaporators are two multi-effect units (items 44 and 45). The first of these is steam-heated with the overhead from this unit supplying the heat for the second. The necessary temperature Δt 's are obtained by a barometric leg (40), pulling through the second waste evaporator overhead condenser (38). Inert gases and the start-up air are removed by a steam ejector and condenser unit (39). The heat-load split between the two evaporators is shown in 461-1128-U-3. Condensate from both evaporator effects is pumped back into the treated water system of the process.

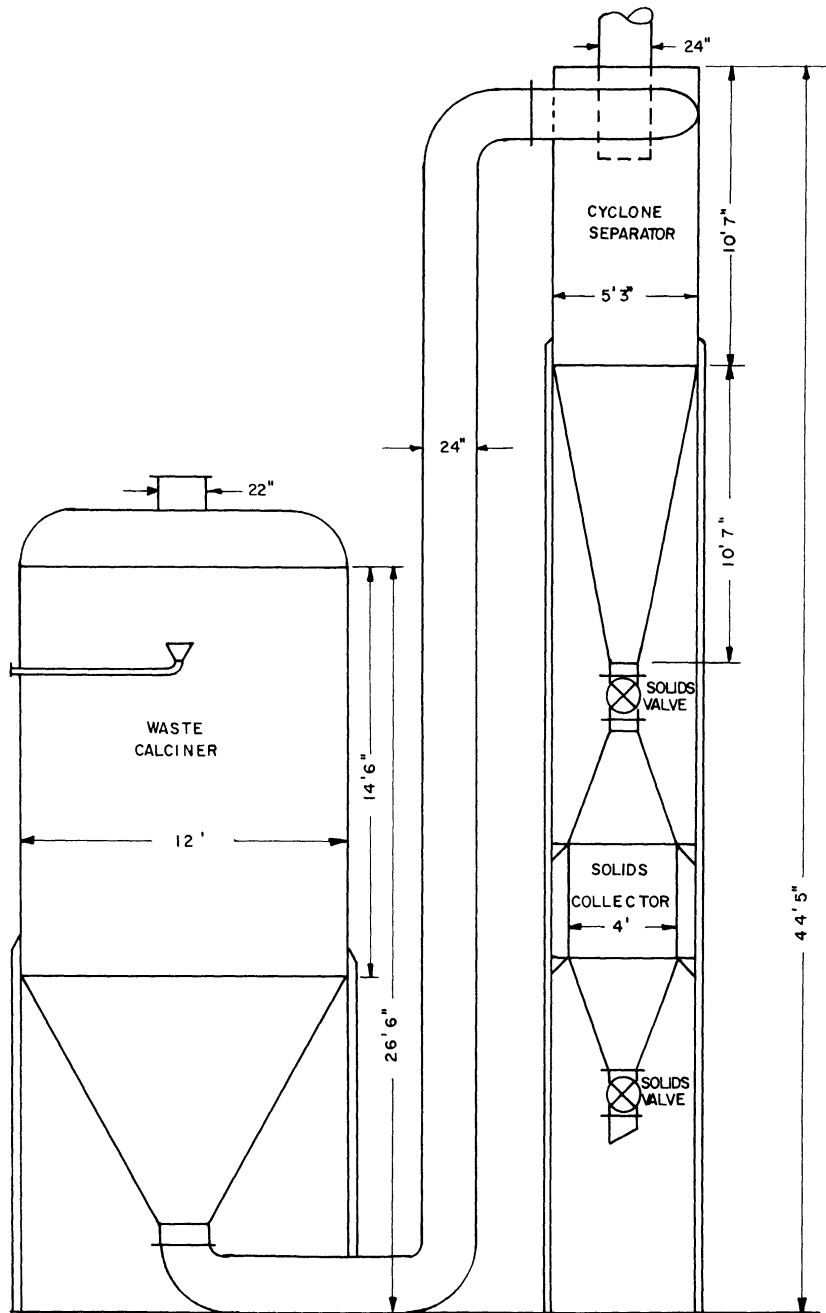
Both evaporators (44 and 45) are similar in design and differ only in the heat-transfer surface contained. The tentative design of these evaporators is shown in Fig. 10. Here again, and for the same reasons given for the product evaporator (26) thermosyphon type of evaporators are used. Entrainment from these evaporators is not as serious as from the product evaporator, so more leeway can be taken in the design of the vapor section. It may be possible to eliminate the mesh de-entraining section as shown in Fig. 10 in the final design.

The bottoms from the second aqueous waste evaporator (45) are pumped to the spray nozzle in the calciner (47). The calciner details are shown in Fig. 11. This calciner operates at 150 psig so that its size is reasonable and is composed of a large chamber into which the concentrated



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		PROJECT	TITLE	
		461-1128	AQUEOUS WASTE EVAPORATOR DESIGN	
		CLASSIFICATION	FIGURE-10	
ISSUE	DATE	UNCLASSIFIED		



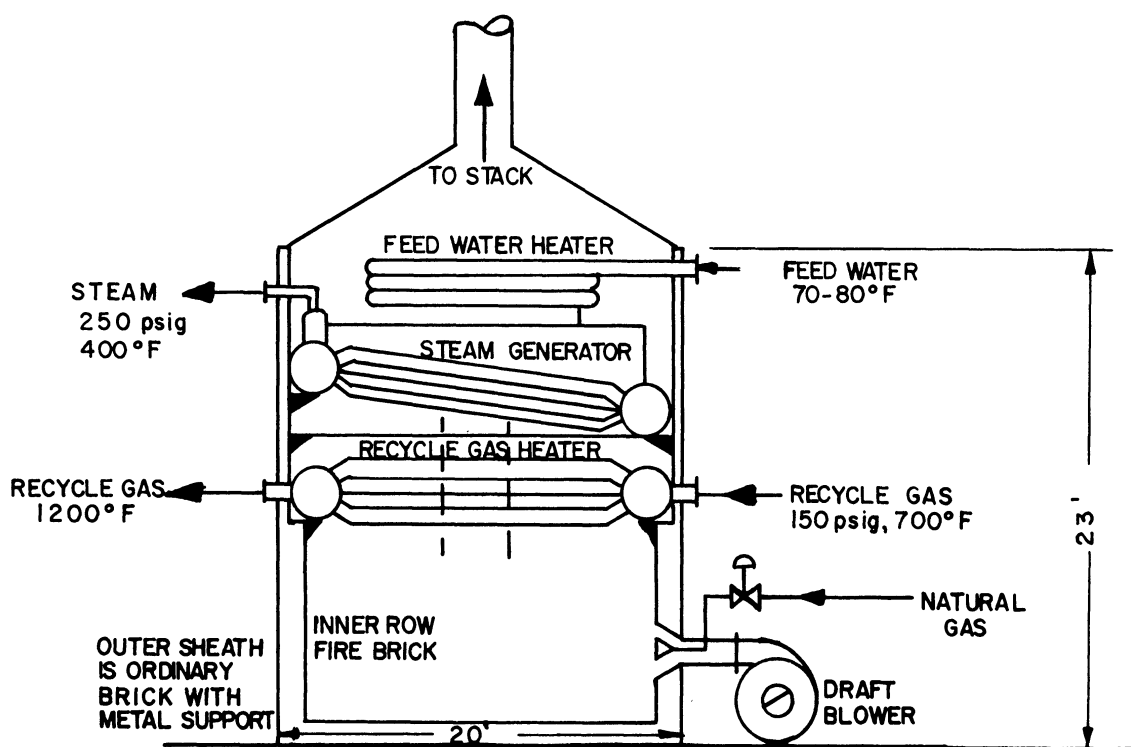
		ENGINEERING RESEARCH INSTITUTE UNIVERSITY OF MICHIGAN ANN ARBOR MICHIGAN		DESIGNED BY	M.E.W.	APPROVED BY	
				DRAWN BY	R.L.K.	SCALE	1/4" = 1'
		PROJECT 461-1128		CHECKED BY	M.E.W.	DATE	5/31/57
				TITLE WASTE CALCINER & CYCLONE SEPARATOR			
ISSUE	DATE	CLASSIFICATION UNCLASSIFIED		FIGURE-II			

aqueous waste salts are sprayed. This spray nozzle breaks the solution up into droplets in the 50-to-20-micron range. As the droplets fall, they are heated and dried by a hot gas stream flowing upward through the calciner temperature. As the particles are dried, they are carried upward by the hot gas stream and pass through a cyclone separator where the solids and gases separate. The solids fall to the conical bottom of the cyclone and are discharged through a solids-collection lock out of the bottom of the unit and are hauled away. The gases are recirculated through a recirculating gas blower (49) and through heating surfaces contained in the recirculating gas heater and steam generator (50). Gas temperatures into the calciner are 1200°F and leave the cyclone at 700°F.

The hydrated salt $\text{Ca}(\text{NO}_3)_2 \cdot 4\text{H}_2\text{O}$ is decomposed at temperatures of 700°F into steam, NO_2 , and O_2 . Residence time in the calciner is approximately 12 sec, which should be adequate for the decomposition to take place. It is these gases that are recirculated through the calciner and thus serve as a heat-transfer medium. A calcining system such as this is discussed in Ref. 10. The NO_2 content of the resultant gas is high and consequently makes an ideal feed to an absorption tower for the formation of nitric acid. Pressure in the calciner unit is controlled by a pressure-regulating valve set at 150 psig; thus the gas discharge from the system is automatic. It is believed that a spray-dryer system as discussed is mandatory for this calciner system. Other types of calciners became prohibitively large in size or result in gases so dilute in NO_2 that difficulties would be encountered in the subsequent absorption step.

The recirculating gas blower (49) poses a design problem, although not an insurmountable one, because it operates at a 700°F temperature and a 150-psig inlet pressure. This blower would have to be constructed of austenitic stainless steel to prevent corrosion and this type of material should also easily meet the temperature conditions. The seal problem would, however, require considerable attention. This blower is rated at 10,000 cfm at 150 psig against a 1.25-psi pressure differential. About 75 hp is required to drive it. A centrifugal type of blower would be necessary to meet this service.

The recirculated gas from the calciner is heated in special heat-transfer surfaces placed in the recirculating gas heater and steam generator (50). This unit is illustrated in Fig. 12. The gas heating surface is placed adjacent to the steam envelope so that the 1200°F outlet temperature can be achieved. This heating surface is formed by rolling and welding the tubes into pipe headers at each end. The whole unit is supported at each end and with two central hanging supports on the tubes. Tubes are approximately 16 ft long and have a 1-in. OD. The total surface of this unit is 2420 ft².



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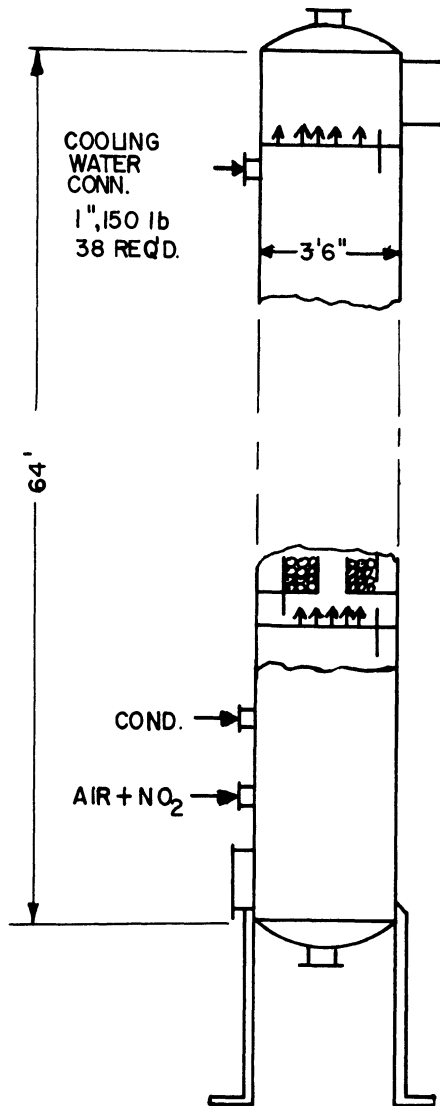
<p>ENGINEERING RESEARCH INSTITUTE UNIVERSITY OF MICHIGAN ANN ARBOR MICHIGAN</p>		DESIGNED BY M.E.W.	APPROVED BY <i>M.E.W.</i>
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		CHECKED BY M.E.W.	DATE 5/27/55
PROJECT		TITLE	
461-1128		RECYCLE GAS HEATER & WASTE STEAM GENERATOR	
CLASSIFICATION		FIGURE - 12	
UNCLASSIFIED			
ISSUE	DATE		

The combustion gases after passing through the gas heating surface are still at about 1500°F. This heat from 1500°F down to 400°F is used to generate steam at 250 psia for use in the process and in building heating. As can be seen in Dwg 461-1128-U-3, there is sufficient heat in the combustion gases after supplying the calciner requirements to supply 10,387 lb/hr of steam more than is required by the process or in heating load. This suggests that some pumps and blowers could be steam-driven; however, no provisions have been made to do so in this study. In Fig. 12, the fuel used is natural gas and this is used at 3000-scfm rate supplied at 25 psig. Oil or some other convenient fuel could be considered.

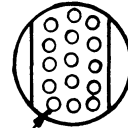
Steam, NO₂, and O₂ must now be converted into nitric acid as it leaves the calciner (47). As the pressure tends to increase in this unit above 150 psig, a pressure-regulating valve bleeds off the gas which then flows to the calciner product cooler and condenser. In this unit the steam is condensed and subcooled to 100°F. Two heat exchangers are used here in series to accomplish the condensing subcooling function. Simultaneously with condensing and subcooling, considerable absorption of the NO₂ and O₂ occurs to form nitric acid in these exchangers. The condensate from the condenser-subcooler is pumped to the second tray from the bottom of the absorber column (53), where the liquid phase acid concentrations are equal. The unabsorbed gas phase goes in below the bottom of the absorber column and passes upward through the bubble trays. Distilled water is pumped by a pump (54) to the top of the absorber column as the absorbing stream. Nitric acid at 56.5% by weight is pumped from the bottom of the column to the nitric acid storage tank (17).

Figure 13 shows the absorption tower (53). This tentative tower design was arrived at from equilibrium data in Ref. 11 and from the reaction kinetics given in Ref. 12 and pilot plant operating data from Ref. 13. Considerable heat is evolved from the NO₂-water absorption reaction and increasing temperature is a disadvantage in further absorption. For this reason every other tray in the absorber column is a cooling tray. Details of these trays are shown in Fig. 13. Twenty bubble trays and 19 cooling trays are thus used. The column dimensions then become 3 ft 6 in. in diameter and 64 ft high, which includes bottom surge capacity. This is a very small column for such a large acid-production capacity; however, the feed is very concentrated in NO₂, and the column is operated under 150-psig pressure which aids absorption, and the downcoming liquor is cooled after each bubble tray. With this provision, it should be readily possible to meet the desired rates. Compressed air is blown into the bottom of the column which aids in absorption since excess oxygen drives the reaction in the desired direction. The air compressor is item 56, and it supplies 10 cfm at 150 psig and 100°F.

3.3.7 Storage and Make-Up Tankage and Auxiliary Equipment.—The process equipment as a whole has been discussed with the exception of the tankage and necessary auxiliary equipment. The tanks provided are the

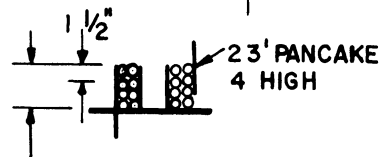
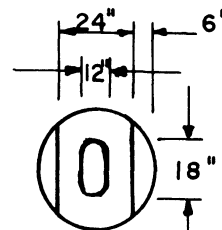


BUBBLE TRAY - 20 REQ'D.



3" PRESSED STEEL SELF SUPPORTING CAPS - 29/TRAY

COOLING TRAY - 19 REQ'D.



ALL DIMENSIONS UNLESS OTHERWISE SPECIFIED MUST BE HELD TO A TOLERANCE - FRACTIONAL $\pm \frac{1}{64}$," DECIMAL $\pm .005$," ANGULAR $\pm \frac{1}{2}^\circ$

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			TITLE	
PROJECT		461-1128	ABSORPTION TOWER	
CLASSIFICATION				
ISSUE	DATE	UNCLASSIFIED	FIGURE-13	

treated water storage tank (16), which holds the clarified water for use in washing stages of the extractor and for general plant use where nonpotable water is required; the nitric acid storage tank (17), which has been discussed; the solvent storage tank (20), which holds up approximately three process hold-up volumes of the TBP-kerosene mixture; the distilled water storage tank, which collects steam drips from the condensers for use in the solvent stripper and solvent washing-scrubbing operations; the solvent make-up tank (34); and solvent wash solution make-up and storage tanks (36 and 35), which serve to make the solvent wash caustic solution to the proper concentration and volume requirements. These tanks are covered further in the cost and specification sheets in a later section.

Also required will be feed-water treatment facilities for the steam generator. This unit is a standard package unit that is in no way unique, so its details are not discussed. Costs for the unit have been allowed for, assuming it to be part of the recirculating gas heater and steam generator.

A water-cooling tower (59) is shown on flowsheet 461-1128-U-3. It was deemed advisable to include this in the plant costs since uranium mills are normally located in arid regions where water is scarce. This unit is a forced-draft package unit that can be furnished by several suppliers. It cools 3294 gpm from 120° to 80°F, and will require 120 gpm of make-up water during normal operation. A cooling-water pump that supplies 3294 gpm cooling water at 60 psig for all plant cooling-water services is also shown.

Provisions for furnishing clarified nonpotable water and potable water are not specifically shown. The specific design varies widely depending upon local water conditions. Costs were allowed in the plant-utilities estimate for average water conditions such as pumping from a well or from a stream. Admittedly, these costs could be in error for any specific situation, but are believed correct for most probable situations.

Other facilities required by the plant would be vehicles, ore loaders, fork-lift loading trucks, and loaders and track cars for calcine disposal. Costs have been allowed for these items, and since use is so specific, no detailed description of their function is required.

3.4 UTILITY REQUIREMENTS

Utility requirements have been discussed in the process discussion; however, the total requirements were not presented. These are given in the paragraphs below.

3.4.1 Fuel Requirements.—Plant vehicles are assumed to be diesel-engine-driven and will consume approximately 10,000 gal/yr of diesel oil. This assumes 24 hr/day operation for 300 days per year.

Natural gas is used for process fuel at the rate of 3000 scfm delivered at 25 psig.

3.4.2 Electrical Power.—The electrical power load of 120-440 volt a-c power, including building and grounds lighting as well as all process and plant loads, will be 825 kw.

3.4.3 Process Water Requirements.—Process water requirements will amount to approximately 200 gpm. This water is clarified, but not biologically purified. This water will supply the boiler feed-water purifiers. Potable water at 30 gal/day per person will be approximately 1.5 gpm.

4.0 EQUIPMENT COSTS

Equipment costs mean very little without an accompanying equipment specification. However, inclusion of a complete specification would make this report very bulky and since most of the major equipment is illustrated in figures together with the throughput and heat-transfer rates, it is believed that the costs given can be related to the equipment without additional specification.

There is another factor that has not been discussed: the materials of construction. Since nitrates or nitric acid is handled all through the process except when raw ore is handled, it is easier to point out those items that are not constructed of stainless steel rather than those that are. The items up to item 16 are of mild or low alloy steel construction. This includes all the ore handling, crushing, and grinding equipment. Item 16, a treated water storage tank, is of galvanized sheet iron construction. It is of course understood that all supports, platforms, walkways, and structures are mild steel. Where heat exchangers are shown with the process fluid in tubes, the shell and baffles may be of mild steel, and where warranted the heads and tube sheets are stainless-steel clad.

Steam and feed water heat-transfer surfaces in item 50, the recirculating gas heater and steam generator, are of low alloy steel. All steam, water, and air lines are mild steel. All items not specifically mentioned above can be assumed to be constructed of either 304, 347, or 316 stainless steel.

4.1 EQUIPMENT PRICING METHODS

The best method of pricing equipment is, of course, by direct quotation from an equipment supplier. Quotations were obtained from suppliers on the ore crushing and grinding equipment, as well as the belt conveyor and the vibratory feeders. However, quotations on all items would be very time-consuming and probably unnecessary. Prices on the other equipment were

obtained using the data in Refs. 14, 15, 16 and 17. Data collected by the author were also used where such data were believed to be more reliable or more recent than data given in the references.

Since the cost information used may be several years old, it is necessary to bring it up to date. This was done by multiplying the cost data given by the ratio of an averaged labor and materials index for 1956 and the year in which the data were prepared. Since the labor and material indexes are not the same, a mean value of the two must be used. Where mild or low alloy steel was used in the equipment, a 50%-labor, 50%-material cost for the item was assumed and the labor-material index averaged accordingly. Where stainless steel equipment is priced, a 40%-labor—60%-material factor was used. These indexes up to May, 1950, are given in Ref. 15. Table III gives the indexes up to December, 1956. These data are from the U. S. Department of Labor, Bureau of Labor Statistics, and from the wholesale price index for metals and metal products.

4.2 FREIGHT CHARGES

The plant pricing methods use as a base the delivered equipment costs which include a freight charge. Freight charges vary with route, nature of shipment, and quantity of shipment and of course the type of shipper. For the purpose in mind it would be impossible to arrive at accurate freight rates without knowing the plant location and the location of the equipment supplier. Freight costs for an average condition are as accurate as can be obtained at this time. If it is assumed that the plant is located in the Salt Lake City, Utah, area, then the equipment would probably be shipped from one of these areas: Chicago, San Francisco, or Los Angeles. With this assumption, freight rates were supplied by the Union Pacific Railroad for various classes of equipment. These data are shown in Table IV. Additional freight cost information is available in Ref. 18. In using this information in this reference, it is necessary to make adjustments for increases occurring since the data were tabulated.

4.3 DELIVERED EQUIPMENT COSTS

Using the methods described above, equipment and freight costs were determined. These costs are given in Table V. This table does not include pump costs. Pump service and ratings are listed separately in Table VI. The price given includes pump, motor and mounts, and freight.

The delivered equipment costs, the sum of totals of Tables V and VI, is then \$826,546.

TABLE III

Material and Labor Index 1950 Through 1956

<u>Year</u> Month	Material Index	Labor Index	<u>Year</u> Month	Material Index	Labor Index
<u>1950*</u>			<u>1951</u>		
Jan.	104.3	148.6	Jan.	124.0	162.7
Feb.	104.9	148.3	Feb.	123.7	163.7
Mar.	104.8	148.6	Mar.	123.2	165.1
Apr.	105.2	149.8	Apr.	123.3	165.9
May	106.6	150.9	May	123.2	166.4
June	108.8	152.4	June	122.7	168.4
July	109.2	153.3	July	122.3	168.2
Aug.	110.8	153.9	Aug.	122.2	168.3
Sept.	113.3	156.1	Sept.	122.1	170.3
Oct.	116.1	157.7	Oct.	122.4	170.3
Nov.	117.7	158.8	Nov.	122.5	170.5
Dec.	121.9	161.5	Dec.	122.5	
<u>1952</u>			<u>1953</u>		
Jan.	122.4	172.5	Jan.	124.0	184.8
Feb.	122.6	172.9	Feb.	124.6	184.8
Mar.	122.6	174.4	Mar.	125.5	185
Apr.	122.5	174.2	Apr.	125.7	185
May	121.8	174.6	May	125.7	186
June	121.8	174.7	June	126.9	187
July	121.9	173.3	July	129.3	188
Aug.	124.1	176.8	Aug.	129.4	188
Sept.	124.6	181.0	Sept.	128.4	189
Oct.	124.1	181.9	Oct.	127.9	189
Nov.	123.9	183.0	Nov.	127.9	189
Dec.	124.0	184.3	Dec.	127.5	190
<u>1954</u>			<u>1955</u>		
Jan.	127.2	191	Jan.	130.1	196
Feb.	126.2	190	Feb.	131.5	196
Mar.	126.3	190	Mar.	131.9	197
Apr.	126.8	190	Apr.	132.9	197
May	127.1	191	May	132.5	199
June	127.1	191	June	132.6	199
July	128.0	191	July	136.7	202
Aug.	128.6	191	Aug.	139.5	201
Sept.	129.1	192	Sept.	141.9	204
Oct.	129.7	193	Oct.	142.3	206
Nov.	129.9	194	Nov.	142.9	206
Dec.	129.8	195	Dec.	144.9	206

TABLE III (Concluded)

<u>Year</u> Month	Material Index	Labor Index	<u>Year</u> Month	Material Index	Labor Index
<u>1956</u>			<u>1956</u> (concluded)		
Jan.	144.9	205	July	144.9	207
Feb.	145.1	205	Aug.	150.2	210
Mar.	146.5	206	Sept.	151.9	213
Apr.	147.7	208	Oct.	152.2	215
May	146.8	208	Nov.	152.1	216
June	145.8	209	Dec.	152.4	218

*The wholesale price index was revised in 1952 to a 1947-49 = 100 basis. These figures are based on that revision.

TABLE IV

Freight Rates on Various Classes of Equipment

Statement of rates in cents per 100 lb to Salt Lake City from Chicago, Los Angeles, and San Francisco on commodities indicated below.				
	From	LCL	CL	M/W for 40-ft Car
Heat Exchangers: (Item 28212) U.F.C. No. 3	Chicago	540	286	24,000
	Los Angeles	385	204	24,000
	San Francisco	385	204	24,000
Grinding or Crushing Equipment: (Item 29662, U.F.C. No. 3)	Chicago	540	272	24,000
	Los Angeles	385	204	24,000
	San Francisco	385	204	24,000
Fabricated Tanks: 1/4 inch or thinner but not thinner than 16 gauge	Chicago	635	(381)	14,000
			(254)	24,000
	Los Angeles)		(272)	14,000
	San Francisco)	454	(182)	24,000
Fabricated Tanks: Thicker than 1/4 inch (Item 41705 U.F.C. No. 3)	Chicago	540		
	Los Angeles)		*(151)	40,000
	San Francisco)	385	*(218)	20,000
Distilling Apparatus, NOIBN: Iron or Steel (Item 28525 U.F.C. No. 3)	Chicago	635	318	24,000
	Los Angeles	454	227	24,000
	San Francisco	454	227	24,000
Bulk Chemicals: in tank cars (Item 1090-E) P.S.F.B. Tariff 260-E	Los Angeles		57	
	San Francisco		57	

* Applies on tanks, plate or sheet, No. 2 gauge or thicker
(Item 8970-E, PSFB Tariff 260-E).

Rates include all ex parte increases

TABLE V

Summary of Equipment and Freight Costs

Item No.	Item	Initial Cost, \$	Wt, lb	Freight	Delivered Cost, \$
1	Ore Screen	150	500	22.70	173
2	Ore Hopper	677	971	37.00	714
3	Grizzly-Scalper-Feeder	3619	7100	380.00	3999
4	Grizzly Underflow Hopper	150	300	14.00	164
5	Jaw Crusher	32760	38000	1030.00	33790
6	Belt Conveyor	12000	15000	810.00	12810
7	Cone Crusher Bin	233	800	36.00	269
8	Cone Crusher	33280	59000	1600.00	34880
9	Cone Crusher Underflow Bin	650	1050	40.00	690
10	Rod Crusher	93100	150000	--	93100
11	Bucket Elevator	3500	6000	320.00	3820
12	Fine Ore Storage Bin	4930	20000	548.00	5478
13	Fine Ore Feeder	1794	3500	189.00	1933
14	Extractor	48900	29760	607.00	49507
16	Treated Water Storage Tank	3100	8500	390.00	3490
17	HNO ₃ Storage Tank	13300	8200	380.00	13680
19	Solvent Extractor	31100	14297	292.00	31392
20	Solvent Storage Tank	1850	1700	77.00	1927
23	Solvent Stripper	9000	3425	155.00	9155
24	Dist. Water Storage Tank	5900	7500	340.00	6240
26	Product Evaporator	2230	1930	74.00	2304
27	Solvent Scrubber	650	241	9.00	659
28	Solvent Washer	650	241	9.00	659
29	Product Evap. Ovhd. Condenser	2230	1500	58.00	2288
30	Prod. Evap. and Prod. Calciner Cond. Collection Tank	300	60	3.00	303
32	Product Calciner	4300	1500	58.00	4358
33	Prod. Calciner Ovhd. Condenser	2730	1200	50.00	2780
34	Solvent Make-Up Tank	925	850	39.00	964
35	Solv. Wash Solution Stor. Tank	1200	900	41.00	1241
36	Solv. Wash Solution Make-Up	950	850	39.00	989
37	Second Waste Evap. Ovhd. to Aqueous Waste Heat Exchanger	7000	6290	240.00	7240
38	2nd Waste Evap. Ovhd. Condenser	31600	14000	540.00	32140
39	Steam Ejector and Condenser	1785	500	27.00	1812
40	Barometric Leg	354	250	10.00	364

TABLE V (Concluded)

Item No.	Item	Initial Cost, \$	Wt, lb	Freight	Delivered Cost, \$
42	Aq. Waste Feed Pre-heater	5920	3500	135.00	6055
43	Aq. Waste Feed Pre-heater Overhead Condenser	6200	2000	77.00	6277
44	1st Aqueous Waste Evap.	21700	8500	330.00	22030
45	2nd Aqueous Waste Evap.	20300	8000	310.00	20610
47	Calciner)				
48	Cyclone Separator) ---	64578	47000	960.00	65538
49	Recirculating Gas Blower)				
50	Recirculating Gas Heater and Steam Generator	137300*	--	--	137300
51	Calc.Prod.Cooler and Cond.	28400	12700	490.00	28890
53	Absorber Column	30900	15900	720.00	31620
56	Air Compressor	2500	4500	123.00	2673
58	Boiler Feedwater Treaters	24110	30000	612.00	24722
59	Cooling Tower	15000	31200	636.00	15636
<u>MISCELLANEOUS</u>					
	Ore Trucks (Hydraulic lift)	6500	--	--	6500
	Prod. Packaging Facilities	2500	1000	54.00	2554
	Fork Lift Truck	1500	1500	81.00	1581
	Calcine Dump Cars and Track	21200	20000	770.00	21970
	Plant Vehicles	11000	--	--	11000
	Power Substation \$20/kw	16000	8000	308.00	16308
			TOTAL		<u>\$ 786,576</u>

*Installed Cost

5.0 PLANT COSTS

Methods of estimating plant costs from delivered equipment costs as well as on other bases are given in Refs. 14, 15, and 16. However, the method that appears best and has the greatest amount of supporting data is Lang's Ref. 16, Part. I. Lang has taken cost data from fourteen plants of various types and broken these costs down into factors times the delivered

TABLE VI

Summary of Pump Costs, Service, and Rating

Item No.	Title	Rating and Function	Materials and Type	Delivered Cost, \$
15	Sand Pump	Pumps 70% solids by vol: 75 gpm, 40-psig head, 2.5 hp	Cast iron body, stainless steel shaft, open centrif- ugal, 1750 rpm	\$ 1,420 (spared)
18	Nitric Acid Feed Pump	66 gpm of 56.5% HNO_3 , $\rho = 1.335$, 1 hp, 25- psig head	304 s.s. casing and impeller and shaft, 1750 rpm	1,462 (spared)
21	Solvent Pump	Pumps 13.2 gpm, solv. $\rho = 0.80$ 25-psig head, 1/2 hp	304 s.s. bowl, im- peller and shaft, 1750 rpm	882 (spared)
22	Aqueous Waste Pump	Solv. of $\text{Ca}(\text{NO}_3)_2$ and HNO_3 , 120 gpm, 60-ft head, $\rho = 1.338$, 2 hp	304 s.s. bowl, im- peller and shaft, 3600 rpm	2,001 (spared)
25	Distilled Water Feed Pump	Pumps combined scrub, strip and solv. wash., 8.5 gpm H_2O , 60-ft head, 1/2 hp	M.S.	859
31	Product Ovhd. Evap. Pump	Pumps 4.5 gpm of 94.59 w/l as nitrates. $\rho = 1.095$, 1/3 hp	304 s.s. bowl, im- peller and shaft, 1750 rpm	442
41	Aqueous Waste Ovhd. Conden- sate Pump	H_2O and HNO_3 solv., $\rho = 1.02$, 35 gpm, 25-psig head, 1/2 hp	Same as No. 31	443
46	Calciner Feed Pump	47.9 gpm at 235°F, $\rho = 1.75$, 233-ft head, 8 hp	s.s. const., 3600 rpm	2,359
52	Calciner Prod. Cooler and Cond. Condensate Pump	65 gpm of 50% HNO_3 , 40-ft hd, $\rho = 1.31$, 3/4 hp	s.s. const., 1750 rpm	442 (spared)
54	Absorber Water Feed Pump	15.2 gpm H_2O , 100-ft head, 1/2 hp	Bronze impeller s.s. shaft, 3600 rpm	1,367 (spared)
55	Absorber Product Pump	66 gpm of 56.5% HNO_3 , $\rho = 1.335$, 25-psig hd, 1 hp	304 s.s. casing, im- peller and shaft, 1750 rpm	1,446 (spared)
57	Boiler Feed Water Pump	97 gpm feed water to 250 psig, 12 hp	Mild steel or bronze s.s. shaft, 3600 rpm	1,243 (spared)
60	Cooling Water Pumps	Fluid cooling, H_2O -3294 gpm, 60-ft hd	M.S. const., 1750 rpm	4,209
61	Steam Conden- sate Return Pump and Well	65 gpm H_2O and Dil HNO_3 , 80-ft hd, $\rho = 1.02$, 2.5 hp	s.s. const., 1750 rpm	562
62	Washed Solv. Pump	Same as No. 21	Same as No. 21	442
63	Product Pump	Same as No. 31	Same as No. 31	442
TOTAL				\$20,021

plant costs or a factor times the total plant investment. The values given seem to be consistent with data collected by the author. Consequently Lang's method of estimating plant costs were largely adhered to.

Unfortunately, the factors to be applied for any item of plant cost can have a rather wide range. A question also arises about just what category a uranium-processing plant fits into. Cost data are given on plants handling only solids, solids and fluids, and only fluids. In these areas the element of judgment enters into the selection of just what should and should not apply. In this estimate the solids (ore crushing and grinding) section was treated as a solids-handling plant, and the remainder of the plant as a fluids-handling plant. Appropriate factors from Lang were then used for each section to arrive at a plant cost.

In Table VII, the various costs that comprise a total plant investment and just what each cost includes, are listed. Plant costs will be \$3,814,000, total cost, \$4,385,000, and physical cost for tax purposes, \$2,311,320, using the basis discussed above.

6.0 PLANT OPERATING COSTS

6.1 FIXED CHARGES

Plant fixed charges are shown in Table VIII. The items in this table are believed self-explanatory. The percentage factors used in arriving at costs are based upon data in Ref. 15.

6.2 VARIABLE CHARGES

A major factor in variable costs will be salaries. Manpower requirements are shown in Table IX. The costs per shift for operating personnel from this table is \$77,430.

All variable charges are shown in Table X, which is believed to be self-explanatory. The variable charges at 1000 tons of ore per day, 300 operating days per year, will be \$950,964. Operating costs for other plant throughput rates can be easily arrived at by using the information presented in Tables VIII and X.

TABLE VII

Plant Cost Summary

Delivered Equipment Cost ⁽¹⁾	\$ 806,597
Installation Costs ⁽²⁾	246,903
Piping ⁽³⁾	523,320
Electrical Eqpt. and Installation ⁽⁴⁾	129,900
Process Control Instruments ⁽⁵⁾	63,200
Service Facilities ⁽⁶⁾	253,900
Buildings ⁽⁷⁾	196,900
Grounds Improvement ⁽⁸⁾	70,600
Land ⁽⁹⁾	20,000
Total Physical Cost	<u>\$2,311,320</u>
Engineering and Construction ⁽¹⁰⁾	694,000
Contingency ⁽¹¹⁾	462,000
Size Factor ⁽¹²⁾	347,000
Total Plant Cost	<u>\$3,814,320</u>
Working Capital ⁽¹³⁾	571,000
TOTAL INVESTMENT	<u>\$4,385,320</u>

- (1) Sum of totals from Tables V and VI.
- (2) The Rod Crusher (10) and the recirculating gas heater and steam generator (50) were quoted on an installed basis, so these items were excluded from installation costs. Installation costs were then taken as 30% of delivered equipment costs less the two items mentioned. Installation includes foundations and supports, ladders, platforms, walkways, and thermal insulation.
- (3) Based on 7.4% of installed equipment costs for solids section, and 62% of same for fluids section. Cost includes all internal process, and auxiliary lines, but not outside lines.
- (4) Based on 13.3% and 12% of installed equipment costs for solids and fluids sections, respectively. Cost includes lines from substation to plant (but not substation itself), switchgear, internal conduit, motor starters, internal lines, lighting, and switches.
- (5) Includes all process control instruments, panels, accessories, spare parts, calibration, and service equipment. Based on 6% of installed equipment costs.

TABLE VII (Concluded)

- (6) Based on 11.1% and 27.8% of installed equipment costs for solids and fluids sections, respectively. Includes plant external piping such as steam and water supply lines, sewers, process drains, cooling water lines, and fire protection apparatus.
- (7) Includes process, laboratory, and office buildings, and office furniture. Based upon 23.5% and 17.3% of installed equipment costs for solids and fluids sections, respectively.
- (8) Includes grading and leveling, fences, roads, sidewalks, parking areas, and landscaping. Based upon 15% and 4.3% of installed equipment costs for solids and fluids sections, respectively.
- (9) Assumes a total of 20 acres required at \$100/acre. This applies to undeveloped arid land in an area where a mill of this type would probably be built.
- (10) Taken as 30% of total physical cost. Includes engineering fees, field and home office expense, contractor fees, and all other construction costs.
- (11) Assumed as 20% of total physical cost.
- (12) Assume as 15% of total physical cost. This factor allows for process changes and capacity adjustments after engineering has been completed.
- (13) Working capital at 15% of total plant costs allowed. This covers normal inventory and other operating burdens encountered during operation.

TABLE VIII

Plant Fixed Charges

Item	Cost, \$/yr
AMORTIZATION—Average life of 15 years assumed = $\frac{2,311,320}{15}$	\$154,000
INTEREST—On plant investment at 6% = $.06 \times \$4,385,000$	263,000
TAXES—State and local at 2% physical cost = $.02 \times 2,311,320$	46,200
INSURANCE—At 1% total physical cost = $.01 \times 2,311,320$	23,100
SALARIES—Of nonoperating personnel (see Table IX)	110,210
LABOR OVERHEAD—At 15% of salaries = $.15 \times 110,210$	16,500
OFFICE OVERHEAD—At 10% of salaries = $.10 \times 110,210$	<u>11,000</u>
TOTAL FIXED CHARGES	\$624,010

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TABLE IX

Plant Manpower Requirements
(Operating 3 shifts/day, 300 days/year)

OPERATING PERSONNEL

No. of People	Location	Job Description	Pay Rate	
			\$/hr	\$/yr
2	Ore Loading	Truck Operators	\$3.70	\$ 7,800
1	Jaw Crusher and Belt Conveyor	Machine Operator	2.28	4,810
1	Cone Crusher and Rod Mill	Machine Operator	2.28	4,810
1	Extractor, Solvent Ext., Stripper, Prod. Evaporator, Prod. Calciner	Chemical Operator	2.48	5,240
1	Product Loader	Chem. Op. Helper	1.98	4,180
1	Product Handling and Storage	Chem. Op. Helper	1.98	4,180
1	Solv. Wash. Solution Make-Up, Tank Farm	Chem. Op. Helper	1.98	4,180
1	Aqueous Waste Preheat and Evapora- tion Tank Farm	Chem. Operator	2.48	5,240
1	Calciner and Abs. Clm.	Chem. Operator	2.48	5,240
1	Recirc. Gas Heater and Steam Generator	Boiler Operator	2.50	5,270
1	Gate	Guard	1.75	3,690
1	Chem. Lab.	Chemist	2.75	5,800
1		Foreman	2.75	5,800
1		Shift Supervisor	3.50	7,390
1	Calcine Disposal	Truck Operator	1.85	3,800
1		Prod. Supervisor	4.00	8,440
TOTAL OPERATING PERSONNEL/SHIFT				\$ 77,430

NONOPERATING PERSONNEL - Maintenance

1	Plant Engineer	4.25	8,960
1	Millwright	3.00	6,330
1	Welder	2.25	4,750
2	Pipefitters at \$2.40	4.80	10,100
2	Maintenance Men at \$2.15	4.30	9,060
2	Janitors at \$1.60	3.20	6,750
			\$ 45,950
	Plant Manager	5.65	12,000
	Accounting (1)	2.50	5,280
	Secretary (1)	1.75	3,690
	Warehouse and Shipping (1)	2.25	4,750
			\$ 25,720

NONOPERATING PERSONNEL - Ore Buying

1	Weighman	2.10	4,430
2	Checkers and Samplers at \$2.10	4.20	8,870
2	Lab Assistants at \$2.00	4.00	8,450
1	Bookkeeper	2.25	4,750
1	Secretary	1.75	3,700
			\$ 30,100

TOTAL NONOPERATING PERSONNEL \$101,770

Salary scales from Oil and Gas Journal, June 9, 1949.

Ratioed up by the labor index to December, 1956.

TABLE X

Plant Variable Charges

(Table assumes 3 shifts/day operating 300 days/yr on a 1000-ton/day rate)

Item	Cost, \$
Salaries: Operating Personnel - 3 x \$77,430	\$232,290
Payroll Overhead - at 15% of payroll	34,800
General Plant Overhead - at 40% of operating salaries	92,800
Repairs and Maintenance - at 5% of total plant physical cost— .05 x 2,311,320	116,000
Chemicals:	
HNO ₃ * Assume 1% loss/ton ore 9.94 lb/ton x 300,000 tons ore x \$.022/lb tech. grade + freight at \$.57/100 lb	82,600
TBP** Losses 270 g/ton ore or \$.312/ton ore \$.312 x 300,000 + freight on 178,600 lb/yr at \$.57/100 lb	94,600
Kerosene Losses 2100 g/ton ore or \$.104/ton at \$.15/gal \$.10 x 300,000 + freight on 1,137,000 lb/yr at \$.57/100 lb	37,680
Na ₂ CO ₃ 0.8 lb Na ₂ CO ₃ /ton ore at \$1.50/100 lb + freight 240,000 lb at \$.57/100 lb	6,210
Fuel and Power:	
Fuel Oil - 10,000 gal at \$.12/gal	1,200
Nat. Gas - 3000 scfm at \$.15/1000 scf	180,000
Power 825 kw at \$.007/kw hr	41,600
Product Shipping Costs:	
1,830,000 lb UO ₃ /yr	
Freight at \$.90/100 lb	16,500
Drums at \$.50/100 lb	2,280
Water Costs:	
Process Water - 200 gpm at \$.14/1000 gal	12,080
Potable Water - 30 gal/person/day at \$.50/1000 gal	324
TOTAL VARIABLE COSTS	\$950,964/yr

*Price quote from DuPont Explosives Division for a 56% technical grade HNO₃.

**Price from Commercial Solvents circular P.S. - No. 38A, January 1, 1957, TBP cost \$.525/lb in drums and carload lots.

7.0 PROBABLE PAY-OUT TIMES AND PROFITS

For comparison purposes it is convenient to have the charges based on a cost-per-ton basis. This is shown in Table XI below. These costs were taken from the data in Tables VIII and X. The total costs for processing a ton of ore is then \$5.25.

TABLE XI

Plant Costs on a Per-Ton-of-Ore Basis
(Table assumes 1000-ton/day plant rate operating 300 days/year)

	<u>Cost, \$/Ton Ore</u>
<u>Fixed Charges</u>	
Amortization	.513
Interest	.875
Taxes (State and local)	.154
Insurance	.077
Salaries (nonoperating personnel)	.366
Labor Overhead	.055
Office Overhead	.037
	<u>\$2.08</u>
<u>Variable Charges</u>	
Salaries (operating personnel)	.774
Payroll Overhead	.116
General Plant Overhead	.310
Repairs and Maintenance	.387
Chemicals	.736
Fuel and Power	.744
Product Shipping	.063
Water Cost	.041
	<u>\$3.17</u>
TOTAL FIXED PLUS VARIABLE COSTS	\$5.25

If costs paid to the miner for the ore are based upon the Atomic Energy Commission schedule, the base cost for U₃O₈ will be approximately \$3.50 per lb. If the bonuses covering higher grades of ore and quantity plus haulage are allowed, the actual cost could approach \$3.75 per lb U₃O₈. On the basis of a 0.3% ore content assumed for this report, the cost of ore to the processor would be \$22.50 per ton for the \$3.75 per lb base. The total cost to the processor would be \$27.75 per ton of ore.

The present price for U_3O_8 from the A.E.C. is approximately \$10 per lb for a product of 75% minimum U_3O_8 . The differential per ton is then \$32.25 per ton of ore. The gross income before Federal income taxes would be \$9,680,000 per year, assuming 1000 tons per day and 300 operating days per year. With income taxes assumed at 52% of profits, the net to the processor would be \$4,650,000 per year. This represents a "pay-out time" of one year.

Other standards could be applied to evaluate the economic feasibility of such a plant. However, by any other standards the process proposed should appear equally good. The major reason for the high profit margin lies in the low chemical costs inherent in the proposed process as compared to processes currently used.

It is possible for a comparatively small investment to add equipment to this process so that a product of much greater purity could be obtained. This would involve a second solvent-extraction-stripping operation similar to the cycle shown on this process but physically smaller. By increasing the purity to meet more rigid specifications, the processor could qualify for a price of \$12.50 per lb of U_3O_8 rather than \$10.00. This addition would be well worth serious consideration if other plant process and design problems are satisfactorily met.

No value has been placed on the CaO from the aqueous waste processing. It has been assumed that its value would offset the disposal costs. Similarly, no provisions have been made for vanadium processing since many ores with high lime content contain little or no vanadium. The ore composition assumed is one with high lime and no vanadium.

8.0 PROCESS AND DESIGN PROBLEMS IN THE PROPOSED PLANT

It should not be construed that the proposed plant is ready to enter a design phase. Rather this study is to serve as a guide in directing a research effort toward a goal that has considerable economic promise. While the steps portrayed in this process are known to be technically sound, the selection of equipment and materials can still be the factors that make or break a plant.

A suggested program to be followed for a successful final plant design is as follows:

- 1) A bench-scale test of each of the proposed steps in the process should be made to confirm or deny the assumptions made here.
- 2) Upon concluding bench-scale tests, a pilot plant of a ton-per-day size should be designed and operated. Every ore that is to be processed should be tested in this pilot plant since wide variations in characteristics can occur in our western ores.

The major questions that pilot plant operation should investigate are:

- 1) required leaching times as a function of particle size, acid strength, temperature, and ore characteristics;
- 2) uranium distribution ratios and solvent-extraction-stripping stage requirements;
- 3) optimum selection of uranium product calcining equipment;
- 4) effects of CaSO_4 present in the ore upon the evaporator and heat-transfer surfaces in the aqueous waste treating section of the process;
- 5) corrosion effects of the Cl^- ion upon heat-transfer surfaces;
- 6) organic solvent losses in the process;
- 7) HNO_3 losses in the process;
- 8) optimum equipment for the waste calcining operation; and
- 9) evaluation of the assumptions made in this report.

It is believed that the questions listed above could be satisfactorily answered in a year's pilot plant operation.

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