RI 8840

Bureau of Mines Report of Investigations/1984

Solid-Liquid Separations in Processing Domestic Laterites

By Gary L. Hundley and R. E. Siemens

UNITED STATES DEPARTMENT OF THE INTERIOR

Report of Investigations 8840

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With an Economic Evaluation by Daniel L. Edelstein

UNITED STATES DEPARTMENT OF THE INTERIOR
William P. Clark, Secretary

BUREAU OF MINES
Robert C. Horton, Director

Library of Congress Cataloging in Publication Data:

Hundley, Gary L

Solid-liquid separations in processing domestic laterites.

(Report of investigations / United States Department of the Interior, Bureau of Mines ; 8840)

Bibliography: p. 22.

Supt. of Docs. no.: I 28.23:8840.

1. Nickel—Metallurgy. 2. Cobalt—Metallurgy. 3. Laterite. 4. Separation (Technology). I. Siemens, R. E. (Richard E.). II. Title. III. Series: Report of investigations (United States. Bureau of Mines) ; 8840.

TN23.U43 [TN799.N6] 669'.733 83-18971

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UNIT OF MEASURE ABBREVIATIONS USED IN THIS REPORT

°C	degree Celsius	L	liter
cu ft	cubic foot	lb	pound
d/wk	day per week	lb/gal	pound per gallon
ft	foot	lb/h	pound per hour
ft^2	square foot	lb/min	pound per minute
ft^2/tpd	square foot per ton per day	lb/ton	pound per ton
g	multiple of gravity	L/min	liter per minute
gal	gallon	min	minute
$gal/ft^2{\cdot}min$	gallon per square foot per minute	mL	milliliter
gal/min	gallon per minute	µm	micrometer
g/L	gram per liter	pct	percent
h	hour	ppm	part per million
h/d	hour per day	rpm	revolution per minute
in	inch	tpd	ton per day
in/yr	inch per year	wt pct	weight percent
kW•h	kilowatt hour	yr	year

SOLID-LIQUID SEPARATIONS IN PROCESSING DOMESTIC LATERITES

By Gary L. Hundley [1] and R. E. Siemens [2]

ABSTRACT

The Bureau of Mines has devised and demonstrated a process for recovering Ni and Co from low-grade domestic laterites. The process consists of four major steps: (1) reduction roast, (2) ammoniacal leach, (3) solvent extraction, and (4) electrowinning. Several solid-liquid separation steps are required that affect the economics of the overall process. In this study, two techniques for solid-liquid separation, centrifugation and thickening, were investigated. This report presents the results of (1) laboratory and pilot plant studies to determine parameters for sizing centrifuges and thickeners and (2) an economic study of the two techniques, based on the requirements of a commercial-size (5,000-tpd) laterite processing plant.

Sizing and other equipment-related recommendations, based on data from the Bureau's studies, were obtained from commercial manufacturers. The costs of thickening and centrifugation, including all the unit operations affected, were determined by the Bureau's process evaluation group. The total operating costs for separation using the two techniques (including depreciation on the capital cost of equipment) were found to be quite close: $17.26 per ton of laterite for centrifugation and $18.15 per ton of laterite for thickening. For centrifugal separation, the greatest cost was found to be the initial capital cost for the separation equipment; using thickeners, the greatest cost was for reagent recovery.

[1]Chemical engineer.
[2]Metallurgist and group supervisor.
Albany Research Center, Bureau of Mines, Albany, OR.

INTRODUCTION

The Bureau of Mines has devised and demonstrated a process for recovering Ni and Co metals from low-grade laterite deposits located in southern Oregon and northern California (5-7).[3] The Ni and Co grades of these deposits, typically 0.7 to 1.2 pct Ni and 0.06 to 0.25 pct Co, are too low to allow economic treatment using present commercial processes. However, these laterites represent a potential resource that could be exploited in an emergency.

A simplified flowsheet of the Bureau's reduction roast, ammonia leach process for treating laterites is shown in figure 1. Because the process is hydrometallurgical, it contains several solid-liquid separation steps that may greatly affect the economics of the process. These separations are difficult to perform because of the small particle size of the laterite. (Approximately 75 pct of the solids is less than 400 mesh). In the initial (or primary) solid-liquid separation, the pregnant leach solution must be separated from the leached solids to provide a crystal-clear, solids-free solution for the solvent extraction step. Other solid-liquid separation steps are necessary in a countercurrent washing circuit in which the leached solids are washed with water to remove residual salts before the residue is returned to the mine site.

Thickening is an established solid-liquid separation method used in other ammonia leach processes such as those used in operations at Nicaro, Cuba, and Greenvale, Australia. A viable technical alternative is to use centrifuges for solid-liquid separation. These two methods--thickening and centrifugal separation--are compared in this report. Early in the research it was determined that filtration would not be practical for the primary solid-liquid separation because of the very slow filtration rates obtained and because of the large amount of free ammonia in the leach solution.

The economic evaluation included as part of this report considers all unit operations in the process that are affected by the solid-liquid separation steps. Among these operations are clarification filtration of the pregnant leach solution after the primary solid-liquid separation step, washing of the filter cakes obtained in the clarification step, and clarification of the final wash water from the countercurrent washing circuit. The reagents in the wash solutions have to be recycled to the leaching step. This is accomplished by steam stripping the free ammonia from the streams followed by evaporation to concentrate the reagents. The steam-stripped free ammonia must then be absorbed back into the concentrated recycle solution.

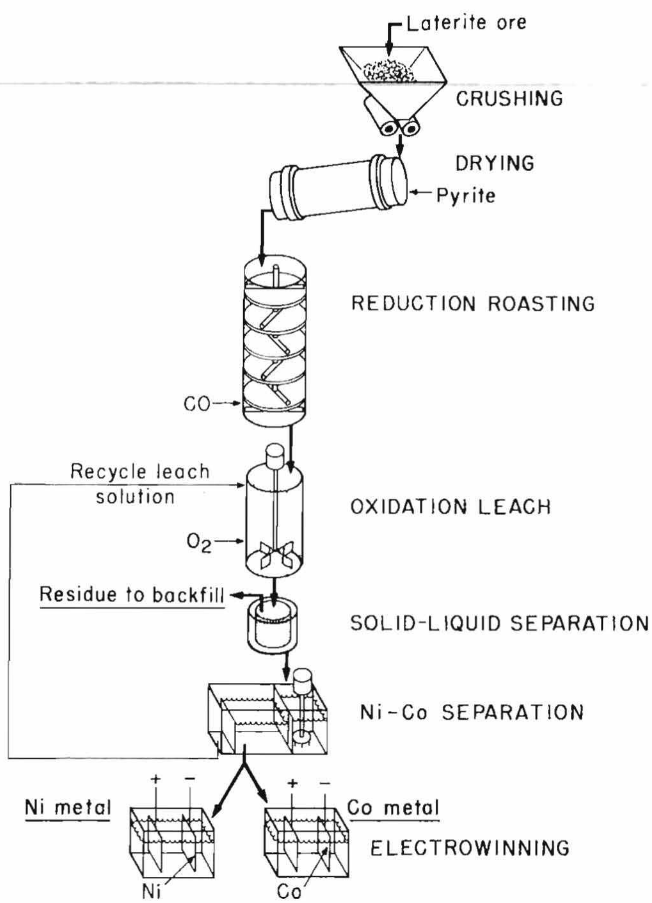

FIGURE 1. - Bureau of Mines reduction roast, ammonia leach process.

[3]Underlined numbers in parentheses refer to items in the list of references preceding the appendixes.

A further consideration is that the final residue must be sufficiently dry to be deposited directly back into the mine site. Dewatering in a tailings pond is not considered to be practical because of excessive rainfall in the southern Oregon-northern California area. Studies have shown that the discharge from a centrifuge is firm enough to be deposited back into the mined area, but further dewatering is necesary before a thickener underflow product can be redeposited. The plant site probably would be located adjacent to the mine site, so it would be possible to haul residue back to the mined area with the same trucks that brought the laterite to the plant.

The Bureau's roast-leach process was operated in a continuous process research unit (PRU) with a capacity of 25 lb of laterite per hour, at the Bureau's Albany (OR) Research Center, and in a 5-tpd pilot plant operated under a Government contract with UOP Inc., at its Tucson, AZ, facility (9). Three types of laterites were tested: A limonitic type from Rough and Ready Creek in southern Oregon, a saprolitic type from Gasquet Mountain in northern California, and a transitional type from Eight Dollar Mountain in southern Oregon.

The studies on centrifuge applications were conducted using leach slurry obtained from the PRU. The thickener results were obtained from settling rate data using leach slurry from the Tucson pilot plant. Preliminary settling rate data were obtained using leach slurry from the PRU.

ACKNOWLEDGMENTS

The cooperation of James K. McGillicuddy, applications engineer, and Jon C. Robbins, district manager, Bird Machine Co., Inc., San Ramon, CA, and Raja G. Ramji, resident manager, Western Region, Dorr-Oliver Inc., Emeryville, CA, is gratefully acknowledged for evaluating the PRU and pilot plant data obtained by the authors and providing estimates of commercial-size equipment requirements and costs.

CENTRIFUGAL SEPARATION

EXPERIMENTAL PROCEDURES

The centrifuge used to obtain scale-up data was a 6-in-diam horizontal-bowl continuous centrifuge manufactured by the Bird Machine Co., Inc.[4] This centrifuge (figs. 2-3) operates in a countercurrent manner with continuous discharges of dewatered solids from one end of the machine and clarified liquid (centrate) from the other end. A scroll conveyor, rotating at a slightly slower speed than the bowl, conveys the solids to one end of the machine, up the tapered end of the bowl, and out of the machine. The centrate discharges from the other end through four orifices. The pool depth of liquid in the bowl can be varied by changing the diameters and radial positions of these orifices. The variables of interest are bowl speed, slurry feed rate, pool depth, and scroll-to-bowl speed differential (4).

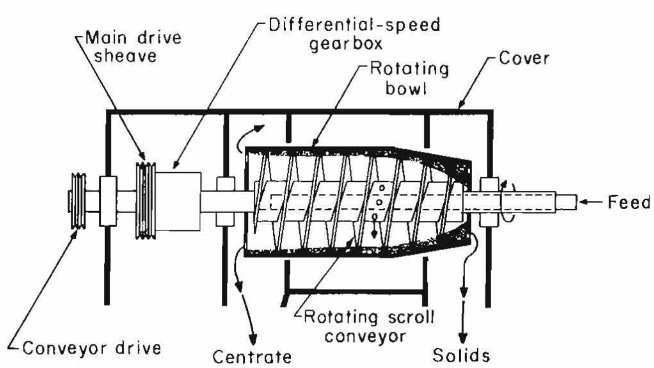

FIGURE 2. - Diagram of continuous centrifuge.

[4]Reference to specific equipment, trade names, or manufacturers is made for identification only and does not imply endorsement by the Bureau of Mines.

FIGURE 3. - Six-inch continuous centrifuge in operation.

The feed slurry to the centrifuge was prepared in batch runs in the PRU and stored in 55-gal drums, each fitted with a mixer to repulp the solids. Most of the centrifuge tests were conducted using the transitional-type laterite; a few were conducted with the saprolitic-type laterite. The results obtained were similar for both laterite types. To insure an adequate supply of slurry, each batch of slurry was reused by combining the centrate from the centrifuge with the dewatered cake and repulping the solids. Each batch of slurry was repulped only once. Particle size analyses of the solids showed that there was no degradation of particles when the solids were reused one time. Repulped solids produced the same results in the centrifuge as fresh solids.

The general operating procedure used to obtain data was to adjust the bowl and speeds to the desired values, start the feed to the centrifuge, and allow the machine to reach steady-state conditions (which required approximately 5 min). The feed slurry to the centrifuge was pumped with a peristaltic tubing pump. The feed rate of slurry to the centrifuge and the centrate flow rate out were then measured for 5 to 20 min, depending on the slurry feed rate; and samples were taken of the slurry feed, centrate, and cake. The solids content of each of these samples and the densities of the liquid samples were determined. The solids content of the slurry was determined by filtering and washing the solids from a weighed portion of the slurry and drying the solids overnight in a drying oven. The solids recovery in the cake was calculated as the difference in mass flow between the solids content of the feed slurry and the solids content in the centrate divided by the solids content of the feed slurry. The mass flow rate of the cake was not determined directly.

Particle size analysis of the solids in the feed slurry, the centrate, and the cake, was performed by a wet screening of the solids larger than 400 mesh and sedimentation analysis of the fraction smaller than 400 mesh. A SediGraph particle size analyzer (Micromeritics Instrument Corp.) was used to measure the particle size distribution of the sub-400-mesh material. For this study, the analyzer was used to measure particle sizes in the range from 0.145 to 40 µm.

Bowl speeds of 3,000, 4,000, and 5,000 rpm, corresponding to centrifugal forces of 767g, 1,363g, and 2,130g, were used with pool depths of 0.19, 0.42, 0.53, and 0.59 in. The slurry feed rates used ranged from 0.5 to 2.5 gal/min. All testing was conducted with the slurry at room temperature. The solids content of the slurry was a nominal 10.0 wt pct.

RESULTS AND DISCUSSION

Results of tests with the batch samples of leach slurry indicated that bowl speed and slurry feed rate were the most important variables. When the feed rate exceeded 1 gal/min, the solids content of the centrate exceeded 2 pct and the solids recovery fell below 80 pct (table 1). As the feed rate was increased beyond 1 gal/min, the solids content of the cake increased slightly, while increases in the solids content of the centrate were proportionally much greater. Increasing the pool depth improved centrate clarity slightly, and the solids content of the cake remained high. The solids content of the cake ranged from 50 to 63 pct, with the majority of the samples having values between 53 and 58 pct. Table 2 shows similar data for separations during the wash stages. The data from tables 1 and 2 were used as the bases for a scale-up to determine the requirements for a commercial-size plant; these requirements are presented and discussed later in the report.

During the PRU operations, two round-the-clock tests of 80 h duration each were conducted in which the centrifuge first separated freshly leached solids from the pregnant leach solution and then separated the solids from the wash water in the washing circuit. By storing re-pulped slurry in surge tanks in the wash circuit, one centrifuge served three functions: the primary solid-liquid separation and the solid-liquid separation in each of two wash stages. The results of these separations are shown in table 3. The cake had a higher solids content than was attained in the batch tests, and the solids content of the centrate was slightly lower. This was probably due to a slightly larger particle size distribution in the feed slurry for the continuous tests. Cake from the primary separation averaged 62 wt pct solids for four samples, while 20 samples of the final residue after a two-stage countercurrent wash averaged 72 wt pct solids. However, as shown in table 3, the solids content of the feed was higher for the wash-stage separations. Figure 4 shows a cake discharge high in solids content; such a discharge can be handled and placed back into the mine site without further dewatering.

TABLE 1. - Primary solid-liquid separations in 6-in-diam centrifuge

(10.0 wt pct solids in feed slurry)

Bowl speed, rpm	Feed rate, gal/min	Scroll-to-bowl differential, rpm	Solids recovery in cake, wt pct	Solids content of centrate, pct	Solids content of cake, pct
		0.19-in POOL DEPTH			
5,000...	0.5	20	89.3	1.3	54.8
	.5	30	88.7	1.4	53.3
	.6	40	89.7	1.3	53.5
	1.9	20	73.9	5.0	56.7
	2.1	30	69.6	5.1	57.9
	2.0	40	66.9	4.2	54.3
4,000...	.7	20	88.7	1.8	54.8
	.6	30	84.4	2.1	56.0
	2.1	20	65.4	4.9	60.6
	1.9	30	ND	5.3	55.0
		0.42-in POOL DEPTH			
5,000...	0.5	10	89.5	1.0	55.7
	.6	20	90.6	1.2	56.2
	2.0	10	ND	2.7	57.1
	2.1	20	73.3	2.5	60.2
4,000...	.6	10	89.8	1.3	57.7
	.6	20	89.8	1.3	57.3
	2.0	10	67.9	3.6	60.6
	2.0	20	67.3	3.8	61.9
3,000...	.5	10	77.1	2.8	55.7
	.5	20	81.5	2.4	52.1
	1.0	10	53.5	4.7	51.9
	1.1	20	59.6	4.7	54.2
	1.0	25	58.8	4.4	50.8
	1.9	10	53.8	5.4	56.8
	1.8	20	56.6	5.8	58.2
	2.2	25	ND	6.0	55.9
	2.6	10	50.2	5.2	63.3
	2.5	20	56.4	5.4	56.4
	2.7	20	54.1	5.2	61.6
		0.53-in POOL DEPTH			
5,000...	0.6	10	88.7	1.2	56.2
	.6	20	89.7	1.1	55.3
	1.1	10	81.0	2.1	57.6
	1.1	20	78.3	2.5	57.4
	1.6	10	72.5	4.3	61.6
	1.5	20	76.7	2.9	59.8
4,000...	.4	10	90.5	1.1	53.9
	.6	20	90.0	1.3	54.1
	.9	10	78.9	2.6	55.2
	1.0	20	66.0	3.6	53.3
	1.2	10	68.6	3.4	56.9
	1.2	20	65.9	4.2	58.6
	1.8	20	69.7	3.6	59.0
		0.59-in POOL DEPTH			
5,000...	0.6	7	90.8	1.5	57.1
	1.0	10	75.2	2.9	56.4

ND Not determined.

TABLE 2. - Wash stage solid-liquid separations in 6-in-diam centrifuge

(Conditions: wash solids pulp density: 30 pct; bowl speed: 5,000 rpm)

Wash stage	Feed rate, gal/min	Scroll-to-bowl differential, rpm	Solids recovery in cake, wt pct	Solids content of centrate, pct	Solids content of cake, pct
		0.42-in POOL DEPTH			
First...	0.5	20	99.4	0.4	61.7
	1.0		91.6	4.7	63.6
	2.0		82.5	9.7	66.8
Second..	.5	20	100.0	.03	65.4
	.8		96.2	1.9	66.8
	1.5		92.9	3.4	66.9
		0.53-in POOL DEPTH			
First...	0.5	20	97.4	1.4	63.3
	1.1	20	84.6	7.3	64.5
	1.0	30	82.9	7.9	62.4
	1.5	20	79.4	9.6	65.0
Second..	.4	10	100.0	.2	62.9
	.5	20	96.5	2.0	65.4
	1.0	20	88.1	6.2	65.7
	1.5	20	79.5	9.1	65.8

TABLE 3. - Solid-liquid separations in 6-in-diam centrifuge, continuous PRU tests

(Conditions: bowl speed: 5,000 rpm; scroll-to-bowl differential: 20 rpm; pool depth: 0.53 in)

	Feed rate, gal/min	Solids content of centrate, pct	Solids content of cake, pct
Primary separations[1]..................	1.2	2.0	61.0
	1.2	.9	67.2
	1.2	.7	59.1
	1.3	2.8	61.0
Wash separations:[2]			
1st wash stage......................	.9	1.7	68.2
	1.3	2.2	70.0
	1.3	.4	66.0
	1.5	.4	68.0
2d wash stage.......................	1.2	1.2	70.0
	1.5	.9	71.1
	1.4	.5	68.7
	1.5	ND	69.8
Final residue, 20-sample average.......	NAp	NAp	72.5

NAp Not applicable. ND Not detected.

[1]14 wt pct solids in feed slurry.

[2]18 wt pct solids in each feed.

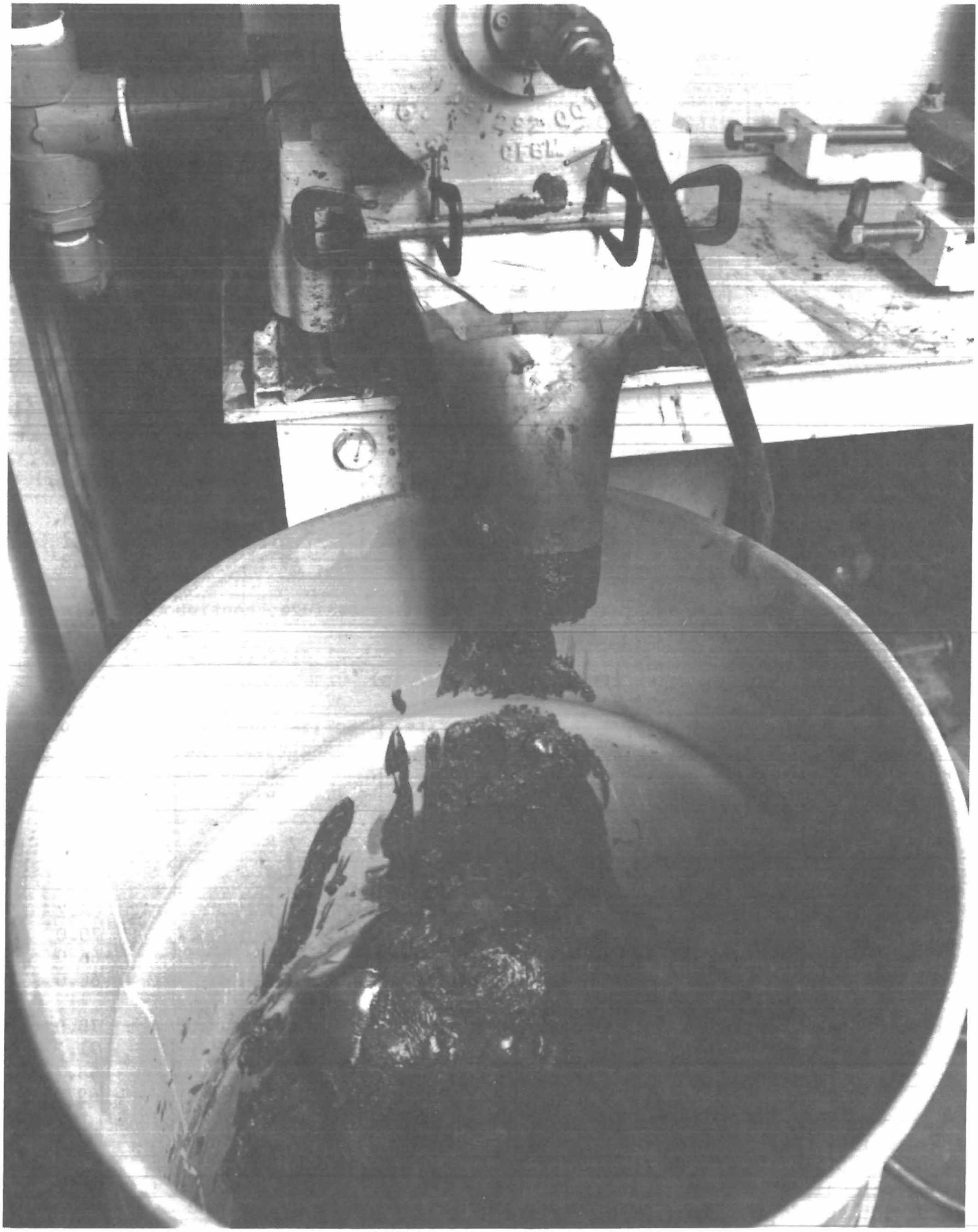

FIGURE 4. - Solids discharge from centrifuge.

PARTICLE SIZE ANALYSIS

To be certain that the centrifuge feed material had a constant size distribution, each batch of leach slurry was sampled and its particle size distribution determined. Some centrate samples and some final solids samples were also analyzed for particle size. A typical screen analysis and particle size distribution for a feed slurry are shown in table 4 and figure 5, respectively. Approximately 75 pct of the solids were less than 400 mesh (38 µm diam) in size. Of the solids in this minus 400-mesh fraction, 50 pct were less than 1.4 µm in diameter. These values were typical of all the feed slurries tested. The centrate from the primary solid-liquid separation step showed a much narrower size distribution (fig. 6). All of the centrate was less than 400 mesh, with 80 pct of the solids in the range of 0.65 to 1.6 µm and 50 pct of the material less than 1.1 µm in diameter. The solids in the centrate must be removed in a clarification step prior to solution purification and solvent extraction.

TABLE 4. - Screen analysis of typical feed slurry[1]

Mesh size	wt pct
Plus 65	10.2
65 by 100	3.0
100 by 150	3.6
150 by 200	2.6
200 by 270	3.1
270 by 400	3.0
Minus 400	74.5
Total	100.0

[1]From transitional type laterite.

Samples of the final cake and final wash water from the wash circuit were also analyzed. The solids in the wash water showed essentially the same size distribution as the solids in the centrate from the primary separation. The average particle size of the solids in the final washed cake was slightly larger than the average solids particle size in the primary separation because a portion of the fines was removed in the centrate.

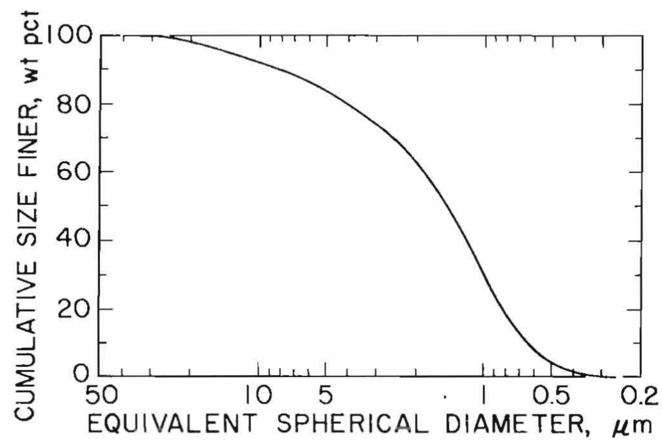

FIGURE 5. - Size distribution of minus 400-mesh fraction of typical slurry feed.

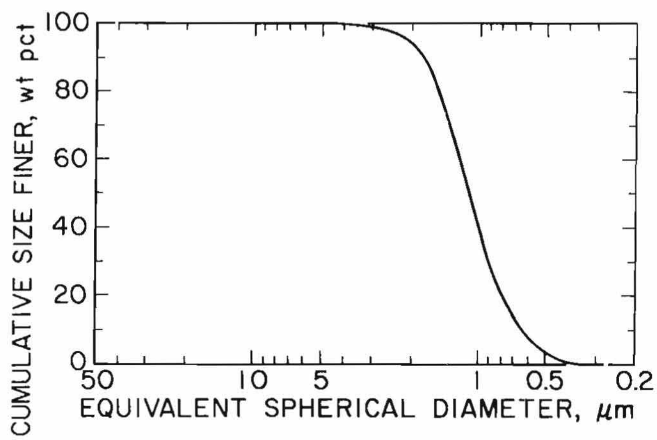

FIGURE 6. - Size distribution of centrate solids from typical primary solid-liquid separation.

Approximately the same amount of material was larger than 400 mesh, but of the minus 400-mesh material, 50 pct was less than 2.0 µm in diameter in the cake versus 50 pct less than 1.4 µm for the primary separation.

COMMERCIAL APPLICATION

Centrifuge equipment sizes and costs for a commercial-size plant were provided by the Bird Machine Co. (1). These recommendations were for a 5,000-tpd laterite processing plant and were based on a scale-up of the PRU data. A 5,000-tpd plant was assumed based on a previous technical and economic evaluation of the Bureau's roast-leach process (8). A detailed flowsheet and material balance for the primary solid-liquid separation step, solids washing circuit, and reagent recovery section of the process are shown in figure 7 and table 5, respectively. The flowsheet shows the basic overall steps required in this portion of the process. (Possible alternate routing of some of the wash water streams was not

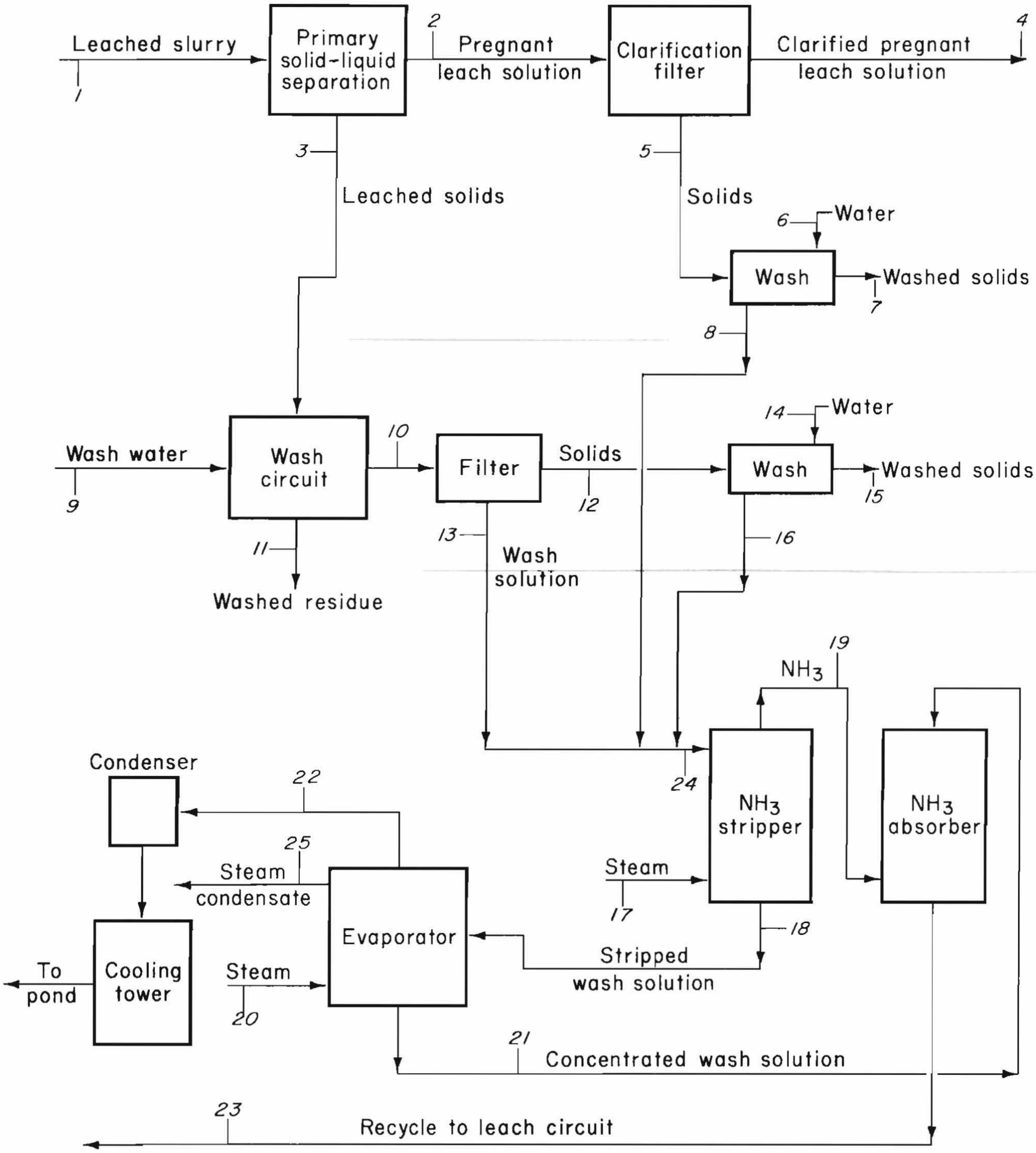

FIGURE 7. - Process flowsheet for centrifugal separations: Primary solid-liquid separation, solids washing, and reagent recovery sections.

TABLE 5. - Material balance for centrifuge circuit

Stream number and name	Solids, lb/min	Liquid, gal/min	Solids, pct	Sp gr of liquid	$(NH_4)_2SO_4$, g/L	NH_4OH, g/L
1. Leached slurry	5,750	3,599	14	1.15	300	100
2. Pregnant leach solution	862	3,325	2.6	1.15	300	100
3. Leached solids	4,888	274	65	1.15	300	100
4. Clarified pregnant leach solution	0	3,265	0	1.15	300	100
5. Clarification filter solids	862	60	60	1.15	300	100
6. Wash water	0	120	0	1.0	0	0
7. Washed solids	862	60	60	1.05	100	33
8. Wash solution	0	120	0	1.05	100	33
9. Primary wash water	0	563	0	1.0	0	0
10. Wash solution	53	589	1	1.05	137	45
11. Washed residue	4,835	248	70	1.0	5.3	1.8
12. Wash solution solids	53	4	60	1.05	137	45
13. Primary wash solution	0	585	0	1.05	137	45
14. Wash water	0	4	0	1.0	0	0
15. Washed solids	53	4	60	1.03	68	23
16. Wash solution	0	4	0	1.03	68	23
17. NH_3 stripper steam	0	[1]355	0	NAp	0	0
18. Stripped wash solution	0	722	0	1.05	130	0
19. NH_3-water vapor	0	30	0	NAp	0	ND
20. Evaporator steam	0	[1]1,813	0	NAp	0	0
21. Concentrated wash solution	0	294	0	1.15	314	0
22. Evaporated water	0	428	0	1.0	0	0
23. Recycle to leach	0	324	0	1.15	285	95
24. Combined wash solution	0	709	0	1.05	130	43
25. Steam condensate	0	[1]1,813	0	1.0	0	0

NAp Not applicable. ND Not determined.
[1]Pounds per minute.

optimized for the economic evaluation.) The material balance was based on the following conditions:

1. The primary solid-liquid separation step has a leached solids feed rate of 5,750 lb/min and a leach liquor feed rate of 3,599 gal/min (solution sp gr = 1.15) for a 14-wt-pct solids content in the slurry feed.

2. Environmental considerations require that the wash circuit remove 98 wt pct of the soluble salts from the laterite residue so that the maximum total nitrogen content in the residue is 0.1 wt pct.

The wash water from the laterite washing must be evaporated in the reagent recovery section of the process and the concentrated solution recycled back to the leaching step. This requires that the amount of wash water and number of wash stages be optimized to provide the required amount of washing while minimizing the costs associated with this section of the process. Table 6 shows the theoretical efficiencies of 2, 3, 4, and 5 wash stages with various wash pulp densities and centrifuge cake (feed) densities. As the wash pulp density is increased (more solids, less wash water) the number of stages required to attain the same washing efficiency increases.

TABLE 6. - Theoretical countercurrent washing efficiency

(Soluble salts left in discard solids, percent)

Solids in wash stage feed	Solids in wash stage underflow				
	70	65	60	55	50
2 WASH STAGES					
30	4.0	6.5	10.3	15.9	24.3
25	2.3	3.8	6.0	9.3	14.3
20	1.3	2.1	3.2	5.0	7.7
15	.6	1.0	1.5	2.4	3.6
3 WASH STAGES					
30	0.9	1.9	3.9	7.9	15.4
25	.39	.8	1.7	3.4	6.7
20	.15	.32	.64	1.3	2.5
15	.05	.10	.21	.40	.77
4 WASH STAGES					
30	0.20	0.57	1.6	4.1	10.4
25	.06	.18	.48	1.25	3.2
20	.02	.05	.13	.33	.83
15	.004	.01	.03	.07	.17
5 WASH STAGES					
30	0.04	0.17	0.62	2.2	7.2
25	.01	.04	.14	.46	1.6
20	.002	.008	.03	.08	.27
15	.0003	.001	.004	.011	.036

This means that if the wash water is minimized, the number of stages will be large, and more centrifuges will be required. However, if the number of stages is minimized, the amount of wash water to be evaporated will be greater.

Based on the material balance, the batch test data, and the continuous PRU test data, Bird Machine Co. (hereafter referred to as the vendor) proposed the use of twelve 44- by 132-in continuous solid-bowl centrifuges handling 300 gal/min each to perform the primary separation step. Results were projected, based on the Bureau's data and the vendor's experience with larger centrifuges, as follows: cake solids, 65 wt pct; solids recovery in the cake, 85 wt pct; and solids in the centrate, 2.5 wt pct. Although the data for the primary separation were obtained with a feed slurry of 10 wt pct solids, and the commercial-size plant was based on a feed slurry of 14 wt pct solids, representatives for the vendor were confident that they could adequately scale the experimental data to the full-size plant. Further work could be done to preconcentrate the solids before the slurry is fed to the primary separation step; this might yield better centrifuge results. The wash circuit, handling approximately 1,200 gal/min, would be operated at 35 to 40 wt pct solids with four countercurrent wash stages. Each stage would incorporate four 44- by 132-in solid-bowl centrifuges, with each centrifuge handling 300 gal/min. The vendor projected that the solids content of the cake would be 70 wt pct and that a solids recovery of 99 pct would be achieved. The wash-circuit data were obtained at a feed solids content of 30 wt pct, whereas the commercial scale plant was sized for 35 to 40 wt pct solids; but again vendor personnel were confident that they could predict the commercial results based on the data.

The same washing efficiency could be attained with only two stages by operating at 20 wt pct solids, but the total number of centrifuges required would be the same because of the increased slurry flow. In addition, the evaporator-stripper section would have to be greatly increased to handle the larger amount of wash water. For these reasons, using fewer wash stages would not be beneficial.

The pregnant leach solution from the primary separation step must be filtered before the solution can be advanced to the solvent extraction step. A filtration rate of 0.5 gal/ft^2·min, which was based on limited data, was used to determine filter sizes for the economic evaluation. It was determined that twenty-two 1,644-ft^2 pressure leaf filters would be required. Further evaluation of filtration requirements are necessary before a final sizing of filters can be determined. The solids are given a double displacement wash, and the resulting filter cake containing 60 pct solids is discarded at the mine site. The water from this wash is combined with the final wash water from the four-stage countercurrent wash circuit and pumped to the reagent recovery section.

The final wash water from the countercurrent wash circuit is also filtered and the resulting solids cake washed with fresh water. The solids are discarded with the other solids, and the wash water is combined with the other wash water streams.

SEPARATION USING THICKENERS

EXPERIMENTAL PROCEDURES

For the thickener studies, settling rate tests were performed on leach slurries obtained both from the PRU and from the UOP 5-tpd pilot plant in Tucson, AZ. Preliminary tests designed to compare flocculants, combinations of flocculants, and different flocculant dosages were performed in 500-mL graduated cylinders; data used to determine thickener sizes for commercial-size applications were obtained in tests with 1-L graduated cylinders. The tests were conducted by pouring the leach slurry into the graduated cylinder and inverting the cylinder several times to insure that the solids were well mixed in the slurry, then adding a measured amount of flocculant, and inverting the cylinder six times to mix the flocculant with the slurry. All tests were conducted at room temperature (approximately 21° C). The position of the interface between the solids and the liquid was recorded as a function of time as the slurry settled. The slurry was allowed to stand for 24 h, and the final settled height was recorded; the supernatant liquor was then decanted off, and the solids were washed, dried, and weighed to determine the settled pulp density of the solids and the overall solids content of the slurry. All tests were conducted without using a final dilution rake to simulate a thickener rake.

Settling rate tests were also conducted on washed solids using the leach material from the pilot plant operations. Samples of the slurry were initially flocculated and allowed to settle for 24 h; the supernatant liquor was then decanted off and replaced with water, and the solids were repulped with a laboratory propeller-type mixer. An additional dosage of flocculant was then added and the settling rate determined. This procedure was repeated for a total of five washing steps. A countercurrent wash was not performed on these slurries.

PRU RESULTS AND DISCUSSION

PRU testing showed that the reducing conditions used in the reduction step of the roast-leach process critically affects the settling rates of laterite leach slurries but that this variable is not important to centrifuge results. Consequently, preliminary settling tests were performed on PRU slurry samples to determine the effect of reducing conditions and pulp density on the settling characteristics of laterite slurries. Slurry samples were produced under different reducing conditions and the settling rates were then determined using two different flocculants, American Cyanamid Co.'s Superfloc 1224, an anionic flocculant, and Superfloc 1128, a nonionic flocculant. One or the other of these flocculants was previously found to work well with PRU slurries.

At a constant laterite feed rate of 11 lb/h in the reduction roaster, carbon monoxide reductant flow rates of 5.0, 7.5, 12.0, and 18.0 L/min were used to produce slurries with different degrees of reduction. It was found that the stronger the reducing conditions were, the more poorly the slurry responded to flocculants. The results using Superfloc 1128 (fig. 8) show that the settling rate decreased as the carbon monoxide flow rate was increased. On aging, the more highly reduced material became easier to settle, as shown in figure 9. The more highly reduced slurries also changed with respect to the type of flocculant required to promote settling; sometimes a nonionic flocculant worked and sometimes an anionic flocculant was required. Unfortunately, the less severe reducing conditions resulted in lower Ni and Co extractions, as shown from the PRU results in table 7. The effect of reducing conditions was not quite as severe at the pilot plant. However, settling problems did occur in the pilot plant operation if the laterite was overreduced. Once the proper reducing conditions were experimentally established, they could be maintained during subsequent operations to obtain consistent settling characteristics.

TABLE 7. - Effect of reduction-step carbon monoxide flow rate on Ni and Co extractions[1]

(Laterite feed rate: 11 lb/h)

CO rate, L/min	Extraction, wt pct	
	Ni	Co
18.0	84.2	80
18.0	81.6	80
12.0	82.9	80
12.0	83.6	80
7.5	65.2	60
7.5	79.0	70
5.0	79.0	70
5.0	67.8	60

[1]From saprolitic-type laterite.

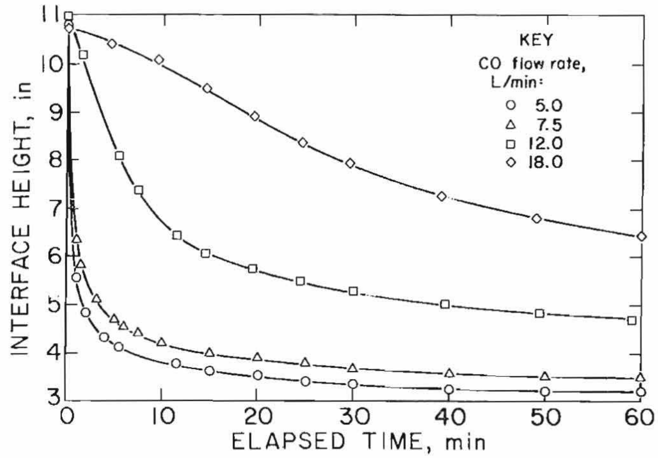

FIGURE 8. - Effect of carbon monoxide flow rate on settling rates. (Conditions: Saprolitic-type laterite. Flocculant: Superfloc 1128 at 0.53 lb/ton of solids. Initial solids content: 10 wt pct.)

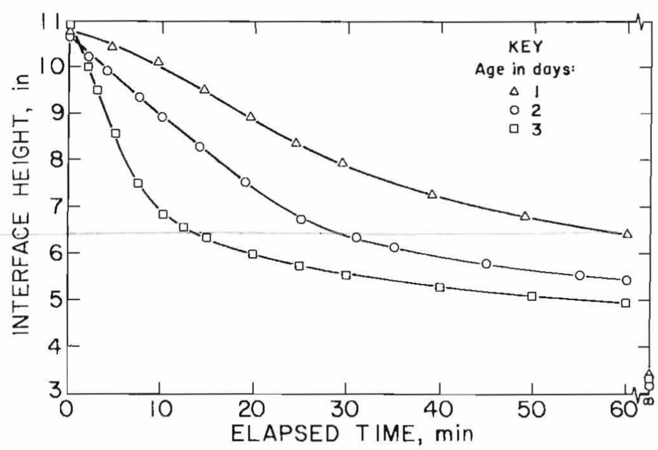

FIGURE 9. - Effect of aging on settling rates. (Conditions: Saprolitic-type laterite; PRU reduced and leached. Reduction temperature: 650° C. Carbon monoxide rate: 18 L/min. Flocculant: Superfloc 1224 at 0.53 lb/ton of solids. Initial solids content: 10.0 wt pct.)

The differences in settling characteristics are believed to be attributable to the nature of the iron oxides in the leached solids. During the reduction step, some of the iron in the laterite is reduced; it is then reoxidized during the leaching step. This reoxidation can result in the formation of iron-oxygen compounds that are different than those present in the original laterite. These iron-oxygen compounds can differ from leach slurry to leach slurry, depending on the reducing conditions used in the reduction step.

The effect of changing the pulp density of the leach slurry was investigated by leaching laterite samples at pulp densities of 10, 15, 20, and 25 wt pct solids. As shown in figure 10, slurries with the three lower pulp densities settled moderately well, but the sample at 25 wt pct solids settled very poorly. The data in figure 10 demonstrate that the pulp density in the primary separation step and in the wash circuit must be kept in the 15-wt-pct range or settling will occur extremely slowly and require a very large thickener area. The effect of changing the pulp density was the same in the pilot plant operations.

PILOT PLANT RESULTS AND DISCUSSION

Settling rate tests in which different flocculants were used with samples of pilot plant leach slurry from both limonitic- and transitional-type laterites were conducted to determine the most effective flocculants for each type of laterite. For both types of laterite, a combination of flocculants was found to produce the best results. The most effective combination for the limonitic-type laterite was made up of equal parts of Guartec 401, a guar gum manufactured by General Mills, Inc., and Polyhall 295, a polyacrylamide polymer manufactured by Celanese Polymer Specialties Co. For the transitional-type laterite, the best combination was 80 pct Guartec 401 and 20 pct Superfloc 1128, a nonionic polyacrylamide emulsion. However, the clarity of the supernatant liquor was poor. A number of combinations of flocculants, including both cationic and anionic types, were tried but failed to improve the liquor clarity. The best clarity resulted from the combinations using the Guartec 401. The settling rate curves obtained for each laterite are shown in figures 11 and 12. The wash stage settling curve for the limonitic-type laterite is also shown in figure 12. The settling rates for the limonitic-type

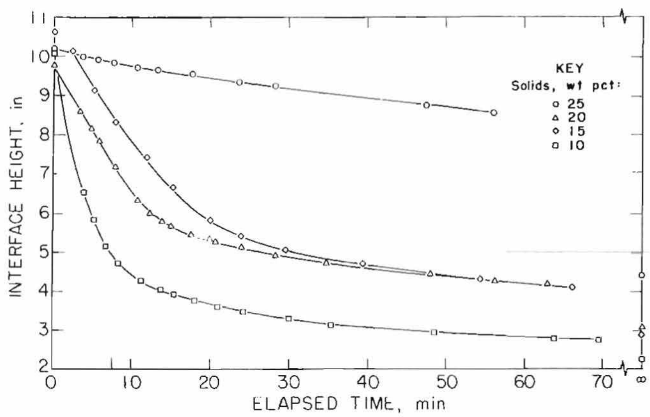

FIGURE 10. - Effect of pulp density on settling rates. (Conditions: Saprolitic-type laterite; laboratory reduced and leached. Flocculants: 50 pct Polyhall 295 and 50 pct Guartec; total flocculant at 0.25 lb/ton of solids.)

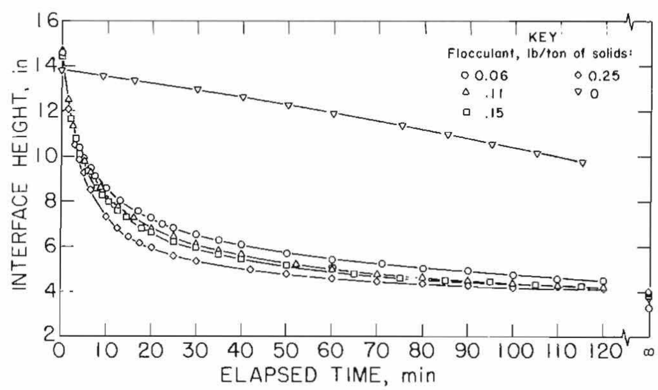

FIGURE 11. - Settling rates for transitional-type laterite. (Conditions: UOP pilot plant leach slurry. Flocculant: 80 pct Guartec and 20 pct Superfloc 1128.)

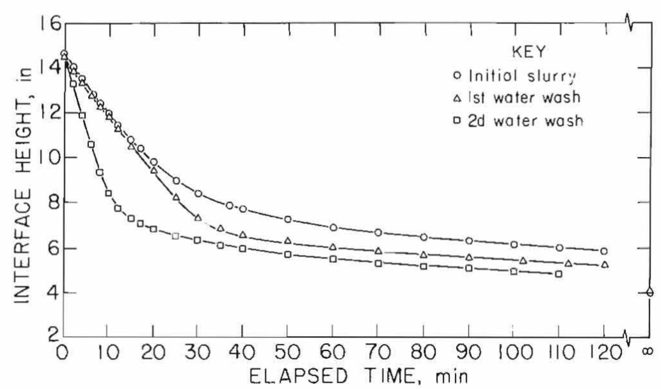

FIGURE 12. - Settling rates for limonitic-type laterite. (Conditions: UOP pilot plant leach slurry. Flocculant: 50 pct Guartec and 50 pct Polyhall 295. Initial flocculant dosage: 0.07 lb/ton of solids; wash stage dosage: 0.04 lb/ton.)

laterite were much slower than those for the transitional-type laterite and are somewhat inconsistent. The thickener unit area requirements that correspond to these curves are shown in table 8. (These requirements were provided by Dorr-Oliver Inc., a manufacturer of thickener units.)

TABLE 8. - Thickener unit area requirements[1] based on UOP pilot plant leach slurries

Sample	Flocculant amount, lb/ton of solids	Unit area, ft^2/tpd
TRANSITIONAL-TYPE LATERITE		
Initial leach slurry.	0	48.5
	.06	3.56
	.11	3.42
	.15	3.36
	.25	2.86
LIMONITIC-TYPE LATERITE		
Initial leach slurry.	0.07	8.92
1st water wash.......	.04	6.31
2d water wash........	.04	3.16

[1]Provided by Dorr-Oliver Inc.

COMMERCIAL APPLICATION

Based on the settling rates shown in figures 11 and 12, Dorr-Oliver recommended the folowing equipment for a 5,000-tpd plant: for the transitional-type laterite, a single 130-ft-diam covered thickener for the primary separation step and a single thickener for each of the wash stages; for the limonitic-type laterite, three 130-ft-diam thickeners for the primary separation step, two thickeners for the first wash stage, and single thickeners for the remainder of the wash stages (2). Due to the high ammonia content of the solutions, the thickeners must be covered to prevent ammonia loss to the atmosphere. Because it becomes increasingly difficult to cover thickeners as the diameters increases, Dorr-Oliver recommended limiting thickener diameters to 130 ft and using multiple thickeners where the settling area requirements call for larger diameters (as in the primary separation step and first wash stage for the limonitic-type laterite, for example). For both laterite types, the countercurent wash circuit would consist of five stages operating at 25 wt pct solids with an underflow of 50 wt pct solids to achieve the required washing efficiency of 98 pct. To avoid settling problems, a portion of the thickener overflow would be recycled back to the incoming slurry to maintain 15 to 20 wt pct solids within each thickener.

Based on pilot plant experience, thickener underflows would contain approximately 50 wt pct solids at best. Additional dewatering of the final washed solids would be necessary before the solids could be deposited back into the mine site. Possible dewatering alternatives would be tailings pond, belt filters, or centrifuges. A tailings pond would probably not be feasible since the annual rainfall in the laterite areas is on the order of 100 in/yr. Significant drying would occur only during the summer months. Filtration on a belt filter would be quite slow, so the use of centrifuges for this additional dewatering was considered in the cost evaluation.

The cost evaluation for thickeners was based on the limonitic-type laterite since it is the worst case and a commercial plant would have to be capable of handling all types of laterite. (In the centrifugation studies, the different laterites all gave essentially the same results.) The flowsheet and material balance used in the thickener cost evaluation are shown in figure 13 and table 9, respectively. The material balance assumes a maximum solids content of 500 ppm in the thickener overflows, so the only filters required on overflow streams would be polishing filters, and no solids washing would be performed on these filters. The pregnant leach solution from the primary thickener overflow is clarified in three 1,025-ft^2 pressure leaf filters, and the spent wash solution from the 5-stage countercurrent decantation circuit is clarified in a 1,311-ft^2 filter. Further pilot plant testing would be required to determine the solids

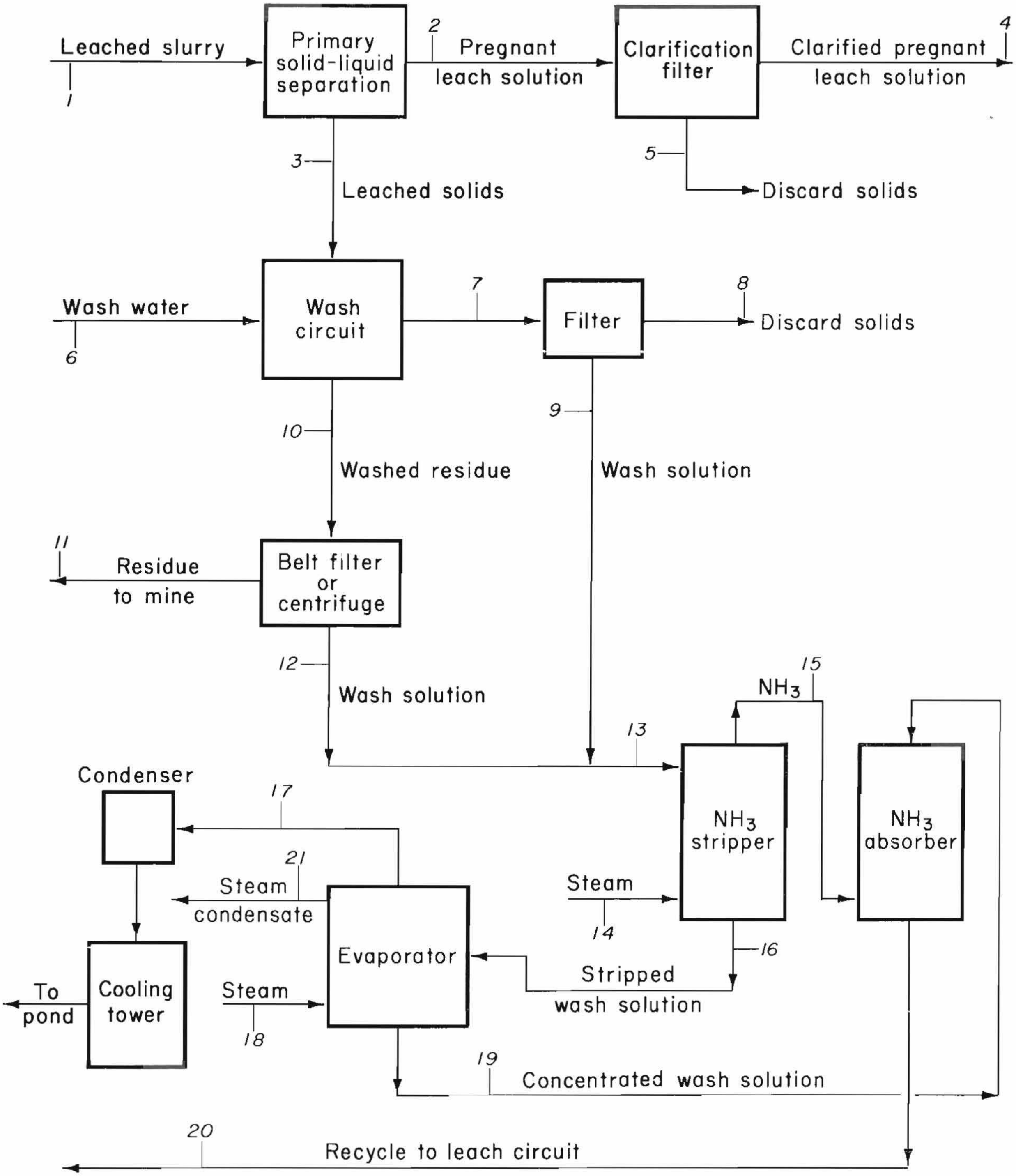

FIGURE 13. - Process flowsheet for thickener separations: Primary solid-liquid separation, solids washing, and reagent recovery sections.

content and filtration requirements of the thickener overflow streams. The material balance also assumes that the underflow from the final washing thickener is dewatered in four 44- by 132-in centrifuges to attain 70 wt pct solids for disposal at the mine site.

TABLE 9. - Material balance for thickener circuit

Stream number and name	Solids, lb/min	Liquid, gal/min	Solids, pct	Sp gr of liquid	$(NH_4)_2SO_4$, g/L	NH_4OH, g/L
1. Leached slurry	5,750	3,599	14	1.15	300	100
2. Pregnant leach solution	2	3,000	[1]500	1.15	300	100
3. Leached solids	5,748	599	50	1.15	300	100
4. Clarified pregnant leach solution	0	3,000	0	1.15	300	100
5. Discard solids	2	.2	50	1.15	300	100
6. Wash water	0	1,377	0	1.0	0	0
7. Wash solution	2	1,287	0	1.05	138	46
8. Discard solids	2	.2	50	1.05	138	46
9. Primary wash solution	0	1,287	0	1.05	138	46
10. Washed residue	5,746	689	50	1.0	4.2	1.4
11. Residue to mine	5,746	295	70	1.0	4.2	1.4
12. Wash solution from centrifuge	0	394	0	1.0	4.2	1.4
13. Combined wash solution	0	1,681	0	1.05	107	36
14. NH_3 stripper stream	0	[2]834	0	NAp	0	0
15. NH_3-water vapor	0	60	0	NAp	0	ND
16. Stripped wash solution	0	1,721	0	1.04	107	0
17. Evaporated water	0	1,182	0	1.0	0	0
18. Evaporator steam	0	[2]5,017	0	1.0	0	0
19. Concentrated wash solution	0	539	0	1.15	297	0
20. Recycle to leach	0	599	0	1.15	297	99
21. Steam condensate	0	[2]5,017	0	1.0	0	0

NAp Not applicable. ND Not determined.
[1]Parts per million.
[2]Pounds per minute.

REAGENT RECOVERY (3)[5]

The reagent recovery sections for both the centrifugation and thickening options are similar, differing primarily in capacity. The first step in reagent recovery is stripping free ammonia from the solution. Combined spent wash solutions are preheated and fed to packed towers. Steam is injected at the bottom of these towers at the rate of 0.5 lb/gal of wash solution. (This stripping steam requirement is based on reported values for producing a concentrated ammonia vapor.) The tower bottom fraction, the stripped wash solution, is pumped to a three-stage multi-effect evaporator where it is concentrated by water evaporation to approximately 300 g/L ammonium sulfate. Vapor from evaporation is used to preheat feed to the stripping towers. Ammonia-water vapor from the stripping towers is reabsorbed in the cooled, concentrated ammonium sulfate solution in a packed absorption tower. The resulting solution is recycled to leaching. Excess water vapor from evaporation not used to preheat the stripping tower feed is condensed, combined with condensate from the evaporator and stripping tower heat exchanger, and pumped to a cooling tower. Cooled condensate can be recycled to process circuits.

[5]This section, the next section ("Economic Evaluation"), and the appendixes to this report were taken from work by Daniel L. Edelstein, geologist, Avondale Research Center, Bureau of Mines, Avondale, MD (now physical scientist, Division of Nonferrous Metals, Bureau of Mines, Washington, DC).

ECONOMIC EVALUATION (3)

CAPITAL COSTS

The capital cost estimates in this report are of the general type called a "study estimate" by Weaver and Bauman (10). This type of estimate, prepared from a flowsheet and a minimum of equipment data can be expected to be within 30 pct of the actual cost for the plant described.

The estimated fixed capital cost on a first-quarter-1982 basis (Marshall and Swift equipment cost index of 739.0) are shown in table 10. The fixed capital costs for the centrifugation and thickening options, for both the solid-liquid separation section and the reagent recovery section, are $65,230,000 and $60,316,200, respectively.

Equipment costs used in this evaluation were based on informal manufacturers' cost quotations and on available capacity-cost data. Costs were updated by use of inflation indexes. Detailed equipment costs and operating cost data are shown in appendix A for the centrifugation option and in appendix B for the thickening option. In developing plant costs, corrosion-resistant construction materials such as stainless steel for centrifuges, rubber linings for thickeners, and fiberglass for tanks were used as appropriate.

Factors for piping, instrumentation, etc., except for the foundation and electrical factors, were assigned to each section using as a basis the effect fluids, solids, or a combination of fluids and solids would have on process costs. (See tables A-4, A-5, B-4, and B-5.) The foundation factor was individually estimated for each piece of equipment, and a factor for the entire section was derived from the total estimated cost. The electrical factor was based on motor horsepower requirements for each section. A factor of 10 pct, referred to as miscellaneous, was added to each section to cover minor equipment and construction costs that are not shown with the equipment listed.

The field indirect cost, which covers field supervision, inspection, temporary construction, equipment rental, and payroll overhead was estimated at 10 pct of the direct cost of each section. Engineering, administration, and overhead were each estimated at 5 pct of the construction cost. A contingency allowance of 10 pct and a contractor's fee of 5 pct were included in the section costs.

The costs of plant facilities and plant utilities (in tables A-1 and B-1) were estimated as 10 and 12 pct, respectively, of the total process section costs and include the same field indirect costs--engineering, administration, overhead, contingency, and contractor's fee--as were included in the section costs. Included under plant facilities were the costs of laboratories, shops, roads, and fences. The costs of water, power, and steam distribution systems were included under plant utilities.

TABLE 10. - Estimated capital costs:[1] Centrifugation versus thickening

	Solid-liquid separation and reagent recovery sections		Reagent recovery section	
	Centrifugation	Thickening	Centrifugation	Thickening
Fixed capital......	$65,230,000	$60,316,200	$12,016,100	$28,792,200
Working capital....	5,604,000	5,898,400	1,616,300	4,227,700
Total.........	70,834,000	66,214,600	13,632,400	33,019,900

[1]First-quarter-1982 basis (Marshall and Swift equipment cost index of 739.0).

Working capital is defined as the funds in addition to fixed capital, land investment, and startup costs that must be provided to operate the plant. Working capital was estimated from the following items: (1) raw material and supplies inventory (cost of raw material and operating supplies for 30 days), (2) product and in-process inventory (total operating cost for 30 days, (3) accounts receivable (total operating cost for 30 days), and (4) available cash (direct expenses for 30 days). Startup and land investment costs were not included in this estimate.

OPERATING COSTS

Estimated operating costs were based on an average of 350 days of operation per year over the life of the plant. This allows 15 days downtime per year for inspection, maintenance, and unscheduled interruptions. The operating costs are divided into direct, indirect, and fixed costs. These costs are summarized in table 11 for both the centrifugation and thickening options. (Annual operating costs are given in tables A-2 and B-2.)

Direct operating costs include raw materials, utilities, direct labor, plant maintenance, payroll overhead, and operating supplies. The raw material costs were estimated as delivered to the plant site. Electricity, water, fuel oil, and coal are purchased utilities. The temperature of the water from the cooling tower was assumed to be 33° C.

TABLE 11. - Estimated operating costs[1] per ton of solids: Centrifugation versus thickening

	Solid-liquid separation and reagent recovery section		Reagent recovery section	
	Centrifugation	Thickening	Centrifugation	Thickening
Direct cost:				
Raw materials:				
Leach reagents	$1.39	$0.13	$0.00	$0.00
Flocculant	.00	.40	.00	.00
Steamplant chemicals	.01	.03	.01	.03
Utilities	4.10	8.02	2.49	7.68
Direct labor	.90	.56	.38	.51
Plant maintenance	2.67	2.02	.43	.99
Payroll overhead	.95	.68	.23	.41
Operating supplies	.53	.40	.09	.20
Total	10.55	12.24	3.63	9.82
Indirect cost	1.43	1.03	.32	.60
Fixed cost[2]	5.28	4.88	.99	2.35
Total operating cost	17.26	18.15	4.94	12.77
Return on investment after taxes[3]	10.11	9.50	NAp	NAp
Total operating cost including return on investment	27.37	27.65	NAp	NAp

NAp Not applicable.
[1]First-quarter-1982 basis.
[2]Includes taxes, insurance, and depreciation.
[3]At 15-pct interest rate.

The direct labor cost was estimated on the basis of assigning 4.2 employees to each position that operates 24 h/d, 7 d/wk. The cost of labor supervision was estimated as 15 pct of the labor costs.

Plant maintenance was separately estimated for each piece of equipment and for the buildings, electrical system, piping, plant utility distribution systems, and plant facilities.

Payroll overhead, estimated as 35 pct of direct labor and maintenance labor, includes vacations, sick leave, social security, and fringe benefits. The cost of operating supplies was estimated as 20 pct of the cost of plant maintenance.

Indirect costs were estimated as 40 pct of the direct labor and maintenance costs. The indirect costs include the expenses of control laboratories, accounting, plant protection and safety, plant administration, marketing, and company overhead. Research and overall company administrative costs outside the plant were not included.

Fixed costs include the cost of taxes (excluding income taxes), insurance, and depreciation. The annual cost of both taxes and insurance were each estimated as 1 pct of the plant construction costs. Depreciation was based on a straight-line, 10-yr period.

DISCUSSION

The costs of the two methods of solid-liquid separation are very close and can be considered to be the same because of the estimating techniques used for some of the equipment sizes and costs. However, the individual items that make up the total costs vary significantly.

In the centrifugation option, the capital costs for the centrifuges and the pressure leaf filters used to clarify the centrate represent the greatest costs. Makeup leach reagent costs (losses associated with the clarification filter solids) and labor and maintenance costs are greater than those for thickening.

In the case of thickening, the reagent recovery section represents approximately 70 pct of the total operating cost. The high cost of this section offsets much of the capital cost advantage of using thickeners. The volume of combined wash solution for the thickening option is approximately 2-1/2 times that of the centrifugation option and results in much higher capital costs for the evaporator and associated equipment. The utility costs associated with steam stripping the ammonia from the wash solution and steam for evaporation are almost three times greater than those for the centrifugation option. The difficulty in consistently flocculating the solids (depending on the reducing conditions, as discussed in the "PRU Results and Discussion" section) could adversely affect the operation of a thickener circuit. This should be considered in technical evaluations of possible separation techniques, even though the economics of the two techniques are similar.

SUMMARY AND CONCLUSIONS

Laboratory studies were conducted to obtain data which were used to determine commercial-scale centrifuge and thickener requirements for performing the solid-liquid separation steps in the Bureau's reduction roast, ammonia leach process for treating laterites. Commercial centrifuge and thickener manufacturers used the Bureau's data to make sizing and equipment recommendations for a plant processing 5,000 tpd of laterite. Based on these recommendations, the costs of both thickening and centrifugation, including all the unit operations affected, were determined by the Bureau's process evaluation group. The overall costs of the two methods were found to be quite close with neither method offering a significant cost advantage. The total operating cost, including depreciation on the capital cost of equipment, was determined to be $17.26 per ton of solids for centrifugation and $18.15 per ton of solids for thickening. In the case of centrifugation, the initial capital cost for the centrifuges and for the pressure leaf filters to clarify the centrate represented the greatest cost, while for the thickening option, the reagent recovery section involving wash-water evaporation represented the greatest cost. Further work on filtration rates and alternate process stream routing is necessary before evaluation of the two methods can be considered final.

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APPENDIX A.--COST DATA: CENTRIFUGATION OPTION (3)

NOTE.--All data in tables A-1 through A-5 are based on 1982 costs; equipment costs are based on a Marshall and Swift equipment cost index of 739.0.

TABLE A-1. - Estimated capital cost, centrifugation option

Fixed capital costs:	
Solid-liquid separation	$37,188,300
Reagent recovery	5,973,600
Steamplant	3,038,600
Subtotal	46,200,500
Plant facilities (10 pct of above subtotal)	4,620,100
Plant utilities (12 pct of above subtotal)	5,544,100
Subtotal	56,364,700
Land cost	0
Subtotal	56,364,700
Interest during construction period	8,865,300
Total	65,230,000
Working capital costs:	
Raw material and supplies	231,100
Product and in-process inventory	2,057,300
Accounts receivable	2,057,300
Available cash	1,258,300
Total	5,604,000
Total capital cost	70,834,000

TABLE A-2. - Estimated operating cost, centrifugation option

	Annual cost	Cost per ton of solids
Direct costs:		
Raw materials:		
$(NH_4)_2SO_4$ at $60/ton	$957,600	$0.66
Aqueous NH_3 at $200/ton	1,064,000	.73
Chemicals for steamplant water treatment	15,200	.01
Total	2,036,800	1.40
Utilities:		
Electric power at $0.055/kW·h	2,642,600	1.82
Process water at $0.25 per thousand gal	71,300	.05
Coal at $31.60/ton	3,231,900	2.23
Total	5,945,800	4.10
Direct labor:		
Labor at $8.50/h	1,131,500	.78
Supervision (15 pct of labor)	169,700	.12
Total	1,301,200	.90
Plant maintenance:		
Labor	1,875,500	1.29
Supervision (40 pct of maintenance labor)	750,200	.52
Materials	1,250,300	.86
Total	3,876,000	2.67
Payroll overhead (35 pct of above labor cost)	1,374,400	.95
Operating supplies (20 pct of plant maintenance)	775,200	.53
Total direct cost	15,309,400	10.55
Indirect cost (40 pct of direct labor and maintenance)	2,070,900	1.43
Fixed costs:		
Taxes (1.0 pct of total plant cost)	563,600	.39
Insurance (1.0 pct of total plant cost)	563,600	.39
Depreciation (over 10-yr life)	6,523,000	4.50
Total operating cost	25,030,500	17.26

TABLE A-3. - Major items of equipment, centrifugation option

Section and item	Quantity	Unit size
Solid-liquid separation:		
Centrifuges	28	44 by 130 in
Clarification filters	22	1,644 ft^2
Reagent recovery:		
Heat exchangers	2	1,504 ft^2
NH_3 strippers	2	7.4 ft diam by 40.0 ft
Evaporator	1	13,563 ft^2 (each stage)
Heat exchanger	1	702 ft^2
Packed tower	1	3.0 ft diam by 15.0 ft
Cooling tower	1	4,947 gal/min

TABLE A-4. - Equipment and related costs summary for solid-liquid separation section, centrifugation option

Item	Cost		
	Equipment	Labor	Total
Centrifuges	$12,687,500	$201,800	$12,889,300
Centrate receivers	99,000	31,700	130,700
Clarification filters	1,763,900	83,600	1,847,500
Discharge chutes	21,900	10,300	32,200
Filtrate receivers	10,500	4,100	14,600
Wash-water receivers	12,200	5,200	17,400
Reslurry tanks	273,100	37,800	310,900
Pressure-leaf filters	801,800	38,000	839,800
Pumps	112,200	26,700	138,900
Total	15,782,100	439,200	16,221,300
Total equipment cost x factor indicated for--			
Foundations, x 0.053			829,100
Buildings, x 0.054			852,900
Structures, x 0.070			1,104,700
Instrumentation, x 0.050			789,100
Electrical, x 0.022			341,900
Piping, x 0.300			4,734,600
Painting, x 0.010			157,800
Miscellaneous, x 0.100			1,578,200
Total			10,388,300
Total direct cost			26,609,600
Field indirect cost (10.0 pct of total direct cost)			2,661,000
Total construction cost			29,270,600
Engineering (5.0 pct of total construction cost)			1,463,500
Administration and overhead (5.0 pct of total construction cost)			1,463,500
Subtotal			32,197,600
Contingency (10.0 pct of above subtotal)			3,219,800
Subtotal			35,417,400
Contractor's fee (5.0 pct of above subtotal)			1,770,900
Total section cost			37,188,300

TABLE A-5. - Equipment and related costs summary for reagent recovery section, centrifugation option

Item	Cost		
	Equipment	Labor	Total
Heat exchangers	$102,800	$3,300	$106,100
NH_3 strippers	183,000	3,000	186,000
Storage tank	13,200	4,700	17,900
Evaporator	1,873,800	8,600	1,882,400
Packed tower	27,800	300	28,100
Pumps	10,100	3,800	13,900
Surge tank	85,400	17,200	102,600
Total	2,296,100	40,900	2,337,000
Cooling tower			[1]263,000
Total equipment cost × factor indicated for--			
Foundations, × 0.061			139,300
Structures, × 0.070			160,700
Insulation, × 0.032			72,900
Instrumentation, × 0.050			114,800
Electrical, × 0.007			15,600
Piping, × 0.400			918,400
Painting, × 0.010			23,000
Miscellaneous, × 0.100			229,600
Total			1,674,300
Total direct cost			4,274,300
Field indirect cost (10.0 pct of total direct cost)			427,400
Total construction cost			4,701,700
Engineering (5.0 pct of total construction cost)			235,100
Administration and overhead (5.0 pct of total construction cost)			235,100
Subtotal			5,171,900
Contingency (10.0 pct of above subtotal)			517,200
Subtotal			5,689,100
Contractor's fee (5.0 pct of above subtotal)			284,500
Total section cost			5,973,600

[1]Installed cost.

APPENDIX B.--COST DATA: THICKENING OPTION (3)

NOTE.--All data in tables B-1 through B-5 are based on 1982 costs; equipment costs are based on a Marshall and Swift equipment cost index of 739.0.

TABLE B-1. - Estimated capital cost, thickening option

Fixed capital costs:	
Solid-liquid separation	$21,740,600
Reagent recovery	13,351,500
Steamplant	7,628,000
Subtotal	42,720,100
Plant facilities (10 pct of above subtotal)	4,272,000
Plant utilities (12 pct of above subtotal)	5,126,400
Subtotal	52,118,500
Land cost	0
Subtotal	52,118,500
Interest during construction period	8,197,700
Total	60,316,200
Working capital costs:	
Raw material and supplies	114,600
Product and in-process inventory	2,162,800
Accounts receivable	2,162,800
Available cash	1,458,200
Total	5,898,400
Total capital cost	66,214,600

TABLE B-2. - Estimated operating cost, thickening option

	Annual cost	Cost per ton of solids
Direct costs:		
Raw materials:		
Guartec 401 at $1.10/lb	$214,800	$0.15
Polyhall 295 at $1.85/lb	361,300	.25
$(NH_4)_2SO_4$ at $60/ton	90,300	.06
Aqueous NH_3 at $200/ton	98,000	.07
Chemicals for steamplant water treatment	43,500	.03
Total	807,900	.56
Utilities:		
Electric power at $0.055/kW·h	2,189,600	1.51
Process water at $0.25 per thousand gal	176,300	.12
Coal at $31.60/ton	9,254,000	6.39
Total	11,619,900	8.02
Direct labor:		
Labor at $8.50/h	707,200	.49
Supervision (15 pct of labor)	106,100	.07
Total	813,300	.56
Plant maintenance:		
Labor	1,419,500	.98
Supervision (40 pct of maintenance labor)	567,800	.39
Materials	946,300	.65
Total	2,933,600	2.02
Payroll overhead (35 pct of above labor cost)	980,200	.68
Operating supplies (20 pct of plant maintenance)	586,700	.40
Total direct cost	17,741,600	12.24
Indirect cost (40 pct of direct labor and maintenance)	1,498,800	1.03
Fixed costs:		
Taxes (1.0 pct of total plant cost)	521,200	.36
Insurance (1.0 pct of total plant cost)	521,200	.36
Depreciation (over 10-yr life)	6,031,600	4.16
Total operating cost	26,314,400	18.15

TABLE B-3. - Major items of equipment, thickening option

Section and item	Quantity	Unit size
Solid-liquid separation:		
Thickeners	9	130 ft
Pressure-leaf filter	1	1,311 ft^2
Centrifuges	4	44 by 132 in
Reagent recovery:		
Heat exchangers	3	2,377 ft^2
NH_3 strippers	3	9.3 ft diam by 40.0 ft
Evaporator	1	3,421 ft^2 (each stage)
Heat exchanger	1	1,363 ft^2
Packed tower	1	4.3 ft diam by 20.0 ft
Cooling tower	1	14,847 gal/min

TABLE B-4. - Equipment and related costs summary for solid-liquid separation section, thickening option

Item	Cost		
	Equipment	Labor	Total
Thickeners	$6,730,500	$671,100	$7,401,600
Sumps	50,800	20,000	70,800
Pressure-leaf filters	327,900	15,600	343,500
Filtrate receivers	9,900	4,000	13,900
Centrifuges	1,812,500	28,800	1,841,300
Centrate receivers	3,700	3,700	7,400
Wash-water receivers	6,600	3,100	9,700
Pumps	129,800	27,800	157,600
Total	9,071,700	774,100	9,845,800
Total equipment cost × factor indicated for--			
Foundations, × 0.078			708,900
Buildings, × 0.10			90,100
Structures, × 0.070			635,000
Instrumentation, × 0.050			453,600
Electrical, × 0.011			103,400
Piping, × 0.300			2,721,500
Painting, × 0.010			90,700
Miscellaneous, × 0.100			907,200
Total			5,710,400
Total direct cost			15,556,200
Field indirect cost (10.0 pct of total direct cost)			1,555,600
Total construction cost			17,111,800
Engineering (5.0 pct of total construction cost)			855,600
Administration and overhead (5.0 pct of total construction cost)			855,600
Subtotal			18,823,000
Contingency (10.0 pct of above subtotal)			1,882,300
Subtotal			20,705,300
Contractor's fee (5.0 pct of above subtotal)			1,035,300
Total section cost			21,740,600

TABLE B-5. - Equipment and related costs summary for reagent recovery section, thickening option

Item	Cost		
	Equipment	Labor	Total
Heat exchangers	$189,500	$5,500	$195,000
NH_3 strippers	335,500	5,900	341,400
Storage tank	41,600	8,800	50,400
Evaporator	4,221,500	12,700	4,234,200
Packed tower	44,400	400	44,800
Pumps	11,800	4,700	16,500
Surge tank	227,600	47,000	274,600
Total	5,071,900	85,000	5,156,900
Cooling tower			[1]778,900
Total equipment cost × factor indicated for--			
Foundations, × 0.052			262,200
Structures, × 0.070			355,000
Insulation, × 0.022			111,700
Instrumentation, × 0.050			253,600
Electrical, × 0.010			48,500
Piping, × 0.400			2,028,800
Painting, × 0.010			50,700
Miscellaneous, × 0.100			507,200
Total			3,617,700
Total direct cost			9,553,500
Field indirect cost (10.0 pct of total direct cost)			955,400
Total construction cost			10,508,900
Engineering (5.0 pct of total construction cost)			525,400
Administration and overhead (5.0 pct of total construction cost)			525,400
Subtotal			11,559,700
Contingency (10.0 pct of above subtotal)			1,156,000
Subtotal			12,715,700
Contractor's fee (5.0 pct of above subtotal)			635,800
Total section cost			13,351,500

[1]Installed cost.