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RPC SULPHATION ROAST PROCESS
DEVELOPMENT AND APPLICATION
TO
NEW BRUNSWICK COMPLEX BASE METAL ORES

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THE RPC SULPHATION ROAST PROCESS
DEVELOPMENT AND APPLICATION
TO
NEW BRUNSWICK COMPLEX BASE
METAL ORES

This is a Federal-Provincial Development Project, Funded Under the Minerals and Fuels Subsidiary Agreement of the Canada-New Brunswick General Development Agreement by the Government of New Brunswick and the Department of Regional Economic Expansion of Canada.

Job Number : 8125
Report Number: MDP/78/1

Roy S. Boorman
Head, Mineral Development
and Processing.

research and productivity council
conseil de recherche et de productivité

May 1, 1978

Mr. Dallas W. Davis
Chief of Policy and Planning
Non-Renewable Resources
Department of Natural Resources
P.O. Box 6000
Fredericton, N.B.
E3B 5H1

Dear Mr. Davis,

We take pleasure in submitting this report which details the development of the RPC Sulphation Roast Process and the results of demonstration trials on metal concentrates from New Brunswick complex base metal ores.

The unique physical chemistry of the Sulphation Roast Process makes it ideally suited to the treatment of low zinc-high iron concentrates, which cannot be treated economically in conventional dead roast-electrolytic zinc plants.

The requirement of conventional electrolytic plants for high grade zinc, low iron concentrates is a principal reason for the relatively low value recovery from New Brunswick ores. A New Brunswick sulphation roast-metals recovery plant which would be capable of treating much lower grade zinc concentrates and also recover other contained metals such as lead, copper and silver would effectively remove the constraint on present mine-mill operators to sacrifice recoveries in order to produce concentrates of grades acceptable to existing smelters.

We envision the preferred mode of application for the proposed sulphation roast plant as one in which middlings and scavenger flotation mill products would be treated. This would enable existing and new mine-mill operations to focus on the marketing of higher grade metal concentrates to existing smelters at premium prices while at the same time

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improving overall metal recovery through the sale of lower grade concentrates to the proposed metals recovery plant.

We wish to thank the New Brunswick Department of Natural Resources for accepting our Proposal to undertake this work, and for their trust in our competence. We think that this particular program will demonstrate that sufficient expertise can be generated within the Province to solve even the most complex of technological problems.

Yours sincerely,

Roy S. Boorman
Head, Mineral Development
and Processing.

RSB/bgh

The New Brunswick Research and Productivity Council (RPC) is a Contract Research Organization which carries out research, problem solving and consulting on a cost recovery basis for clients in Canada and countries abroad. This particular project was carried out by a Research and Development Team comprised of the following staff members:

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1.0 INTRODUCTION

Although the Bathurst-Newcastle area of New Brunswick is one of the most important mining camps in Canada, complex metallurgical problems, which result in relatively poor metal recoveries, have kept this region from reaching its full potential. Current flotation practice results in losses of approximately 35 to 45% copper, 30% lead, 15 to 20% zinc and 35 to 50% silver to the tailings. Additional losses occur through the distribution of metals in concentrates of other metals as a result of poor response to flotation. Payment for metals as impurities in other concentrates is normally reduced or eliminated.

Although much work has been done by industry to improve flotation recoveries and some success has recently been reported on the refloating of tailings to recover some of the lost zinc, further improvement is desirable. This will undoubtably have to come as a result of new technology.

One such new technique is the RPC Sulphation Roast Process, a highly selective and efficient technology for separating a range of valuable metals from gangue mineral which might otherwise be difficult to accomplish by more conventional means. Although the range of its application is extensive, from the treatment of the total ore through to tailings, the most promising use for the Process is probably in conjunction with flotation.

The use of a sulphation roast plant as an addition to a flotation concentrator would allow the operator to produce higher grade metal concentrates without concern for recoveries; losses would be recovered in the sulphation roast plant.

Apart from ensuring maximum metal recovery and thereby increasing overall revenues, such an approach would also reduce the unit cost of flotation since much of the present complex flotation circuit, with its heavy recirculating load, could be eliminated.

The implications for this proposed plant are far reaching for the mining economy of the Province. The Bathurst-Newcastle area has many known deposits of the Brunswick-type, several of which could immediately benefit from a solution to their difficult metallurgy.

The preliminary work for the RPC Sulphation Roast Process was carried out in bench scale experiments using a 6-inch diameter fluidized bed reactor. These results, which were incorporated in an earlier report (1), were encouraging and led to the construction of a mini-pilot plant consisting of a 14-inch diameter reactor, a leaching-purification section and electrowinning cells. This plant was used to carry out demonstration trials and collect sufficient technical data for plant scale-up purposes so as to enable the cost of the process to be evaluated.

During the course of the mini-pilot work it was determined that certain assumptions about the mode of operation of the roaster, which were formulated on the basis of the

6-inch roaster trials and earlier work carried out by the St. Joseph Lead Company (pers. comm.) were not technically feasible and had to be abandoned. For example, it was found to be more advantageous to feed dry concentrates to the roaster rather than slurry feed as originally attempted by the St. Joseph Lead Company. This necessitated major changes in approach and a change in emphasis in the program from demonstration trials to process development. Although progress was somewhat hampered by shortage of staff and time, the program from start-up of the roaster in May 1977 to completion of this report in April 1978, was completed on schedule.

In this report the complete results of the experimental work and mini-pilot plant trials are presented which lead to the development of the RPC Sulphation Roast Process (patents applied for) for the treatment of low grade mill products from complex base metals ores of the Brunswick type.

Although some priority was to be given to the testing of a suitable true bulk concentrate, i.e. 32% Zn, 10% Pb, 0.7% Cu and 8.8 oz/ST Ag, in both the preliminary bench scale work which was initiated in November of 1975 and in the mini-pilot work completed in March 1978, this material, which was to be produced by CANMET in a pilot flotation mill was not available throughout both programs. As a result, the development work was carried out for the most part on a second stage tailings refloat product, approximating 30% Zn,3.6% Pb, 0.6% Cu and 4.4 oz/ST Ag, which was obtained from the Brunswick Mining and Smelting Corporation Ltd. Concentrates with higher amounts of zinc, lead and copper were produced by blending appropriate zinc, lead and copper concentrates obtained from Brunswick and Heath Steele Mines Ltd. with the tailings refloat product.

Two methods of feeding the roaster were evaluated: first, a slurry feed (70% solids) using spent zinc electrolyte as the liquid medium, with spraying of the slurry into the reactor freeboard using an air assisted dispersion nozzle, and dry feeding (approximately 2.5% moisture) by pneumatic injection at the bottom of the fluidized bed.

In addition to the tailings refloat product, the roasting characteristics of 30 wt% zinc concentrates containing up to 14.3 wt% lead and up to 6.8 wt% copper were evaluated through the use of concentrate blends.

Leaching, solution purification, electrowinning and neutralization trials were carried out with tailings refloat calcines obtained from the RPC sulphation roaster and dead roast calcines obtained from Canadian Electrolytic Zinc Ltd. These trials were carried out in bench scale tests to establish appropriate leaching conditions and then in the mini-pilot plant with recycle of solutions to the roaster (interlocked mode). Leaching was a two stage operation, a "neutral" leach followed by a hot strong acid leach, with the filtrate from the second stage leach recycled to the reactor for decomposition.

The body of the report deals exclusively with relevant data applicable to two full scale sulphation roast plants based on two feeds, a bulk concentrate and a tailings refloat product. Also included are data and implications for an integrated sulphation roast-dead roast plant which would use "scavenger" products (i.e. tailings refloat and middlings products) in the sulphation roaster and regular or high grade zinc concentrates in a companion dead roaster.

An economic evaluation is given for each plant and comparisons drawn with the Sherritt Gordon Pressure Leach Process treating the same bulk and tailings refloat products, and with a conventional dead roast plant, treating high grade (52%) zinc concentrates.

The application of the RPC Sulphation Roast Process to New Brunswick complex ores is discussed with specific reference to two proposed models; 1) a sulphation roast plant to treat a blend of a modified bulk concentrate from Heath Steele and tailings refloat and zinc middlings products from Brunswick and 2) an integrated sulphation roast-dead roast plant which would use only concentrates from Brunswick, high grade zinc concentrates in the dead roaster and tailings refloat and/or zinc middlings products in the sulphation roaster.

The results of the process development work and implications for plant scale-up, mass and thermal balances for scale-up, mineralogy, calibration data, alternatives to limestone neutralization of spent electrolyte and relevant cost data used in the economic evaluation of the full scale plants are presented in the Appendices.

1.1 ACKNOWLEDGEMENTS

Support of this work on sulphation roasting began with the New Brunswick Community Improvement Corporation and its then director, Mr. Pat Blanchard, which underwrote part of the costs in developing the original concept. Reports prepared for the Corporation (2,3,4) formed the basis of RPC's proposal to the New Brunswick Department of Natural Resources that laboratory scale roasting trials be undertaken to demonstrate the application of sulphation roast technology (5). The initial support of the Community Improvement Corporation and the support of the Mines Branch, Department of Natural Resources for funding the bench scale test program (1) and the mini-pilot trials, which are the subject of this report, is greatly appreciated, as is the cooperation and assistance of individuals with the following organizations:

New Brunswick Department of Commerce and Development
Brunswick Mining and Smelting Corporation
Heath Steele Mines Limited
Anaconda Caribou Limited
St. Joe Minerals Corporation
    (formerly St. Joseph Lead Company)
Hecla Mining Company
Canadian Electrolytic Zinc Limited

We would also like to thank Mrs. Blanche Humble and Mrs. Marlene O'Neill for their careful typing of this report.

2.0 SUMMARY AND CONCLUSIONS

A sulphation roast-leach-electrowinning process has been developed and demonstrated in a mini-pilot plant at the New Brunswick Research and Productivity Council. This new Process recovers high purity zinc, copper metal and minor elements such as cobalt and cadmium, while producing a residue from which a lead oxide-silver product can be recovered by brine leaching.

The RPC Sulphation Roast Process is similar to conventional zinc dead roast-leach-electrowinning practice. Differences occur in the roasting conditions, methods of separating zinc from iron present in the feed and the method of copper recovery. The Process has been successfully tested on zinc products ranging from 20 to 32% zinc, 0.5 to 4.48% copper, 2 to 14.3% lead and 15 to 30% iron, achieving recoveries of 96% zinc and 97.5% copper.

The roasting is performed under sulphating conditions at 685°C and 4.5-5.5% SO_2 to achieve approximately 85% primary metal extraction for zinc and copper in a dilute "neutral" acid leach. The residue, containing the remaining zinc as ferrite ($ZnO.Fe_2O_3$) and remnant unreacted sphalerite, ((Zn,Fe)S), is subjected to a second extraction in a hot acid leach at 40 gpl constant free sulphuric acid, 95°C and two hours retention time at high pulp density. Under these conditions up to 95% of the remaining zinc and copper in the neutral leach residue is solubilized into a concentrated sulphate solution. After solid-liquid separation, the filtrate contains up to 100 gpl zinc and 110 gpl total iron. The solution is continuously recirculated into the sulphation roaster to thermally decompose the iron sulphates to insoluble hematite and recover the zinc and copper as sulphates in the calcine, thereby eliminating the requirement for a complex iron precipitation stage such as the Jarosite Process commonly employed in conventional electrolytic zinc plants. Thermal and mass balances show that in a commercial size reactor the entire hot acid leach solution can be decomposed in the fluidized bed reactor using only the excess heat released during roasting.

This new alternative for effective and simple separation of the zinc from the iron in solution results in virtually all of the zinc and copper reporting to the dilute acid 'neutral' leach solution. Overall extractions of zinc and copper from the calcines are 98.5 and 98%, respectively, with less than 1 gpl soluble iron in the dilute acid "neutral" leach solution. Iron is removed to less than 0.01 gpl by oxidation and precipitation after copper recovery. The precipitated iron is recycled to the neutral leach, redissolved in the hot acid leach, and converted to inert hematite on recycle to the roaster.

Copper is recovered by solvent extraction from the first stage (dilute acid) leach filtrate. Direct electrowinning from the copper sulphate loaded strip liquor produces high purity copper cathodes.

Although no test work was undertaken in this program, previous work by the St. Joseph Lead Company (pers. comm.) showed that up to 97% of the lead and 85% of the silver was recoverable from sulphation roast leach residues of Brunswick Mining and Smelting ores, using a brine leach.

A simplified flowsheet for the RPC Sulphation Roast Process is shown in Figure 1. Two feeds, a tailings refloat product and a bulk concentrate, were used to establish two general full scale models for a sulphation roast plant. The assays for these products, feed rates and overall metal recoveries, are shown in Table 1. Zinc and copper recoveries are based on data from the mini-pilot trials, whereas lead and silver recoveries are derived from mass balance calculations using the St. Joseph Lead Company test results as a guide. The principle roasting and leaching operating conditions and overall extractions for zinc and copper are given in Table 2.

Preliminary leach tests on conventional dead roast calcines, supplied by Canadian Electrolytic Zinc Ltd., showed that no major problems should be encountered in carrying out leaching of combined sulphation and dead roast calcines. Mass and thermal balances indicate that a sulphation roast calcine produced from a 30% zinc product can be combined with a dead roast calcine produced from a 56% zinc concentrate and leached in a single train, two stage

TABLE 1

ROASTER FEED TONNAGE, GRADE AND METAL RECOVERY

	Sulphation Roast (SR) Bulk Concentrate	Sulphation Roast (SR) Tailings Refloat	Integrated Sulphation Roast/Dead Roast	
			SR	DR
Feed Rate (MTPD)	843.96	900.23	389.52	268.63
Wt% Metals				
Zinc	32.00	30.00	30.00	56.00
Lead	10.00	3.60	3.60	1.20
Copper	0.70	0.60	0.60	0.19
Silver (oz/S.T.)	8.80	4.40	4.40	1.83
Recoveries (%)				
Zinc	96.0	96.0	97.0	
Copper	97.5	97.5	97.5	
Cadmium	89.6	89.6	60.6	
Lead	93.9	91.8	92.5	
Silver	85.0	85.0	85.0	

TABLE 2

OPERATING CONDITIONS AND AVERAGE METAL EXTRACTIONS ACHIEVED IN THE RPC SULPHATION ROAST PROCESS

2.1 Roasting:

Temperature = 685 ± 5°C
Retention Time = 5.5 hrs.
Air Factor = 1.8
Gas Velocity = 0.75 m/sec (at 685°C)
Feed Rate = 4.1 T/m^2 - day

2.2 Dilute ("Neutral") Acid Leach:

Temperature = 55 ± 10°C
Retention Time = 1.5 hrs.
pH = 3.6-4.0

Filtrate composition (SR) = 150 gpl Zn, 2 gpl Cu, <1 gpl Fe
Filtrate composition (integrated SR/DR) = 150 gpl Zn, 1 gpl Cu, <0.01 gpl Fe

2.3 Hot Acid Leach:

Temperature = 95 ± 5°C
Retention Time = 2 hrs.
Free H_2SO_4 = 40 gpl
Pulp Density = 45 wt% solids

Filtrate composition (SR) = 100 gpl Zn, 110 gpl Fe, 40 gpl H_2SO_4
Filtrate composition (integrated SR/DR) = 108 gpl Zn, 176 gpl Fe, 40 gpl H_2SO_4

Overall metal extraction (in dilute acid leach) with recycle:

98.5% Zn 98% Cu

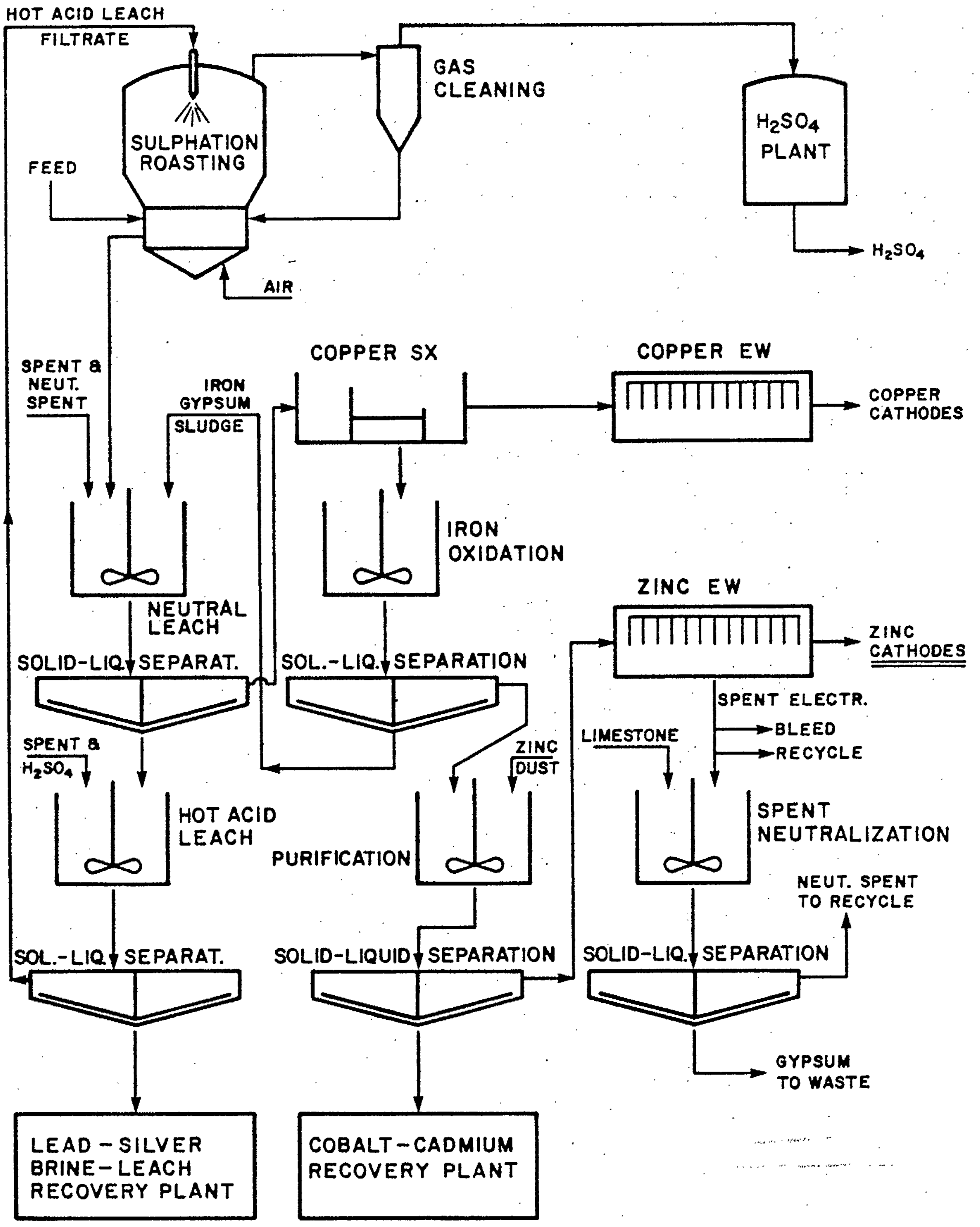

Figure I

CONCEPTUAL FLOWSHEET OF THE R.P.C. SULPHATION-ROAST LEACH-ELECTROWINNING PROCESS APPLIED TO A BULK CONCENTRATE OR TAILINGS REFLOAT PRODUCT leach process. With a roaster feed ratio of 1.43 to 1, sulphation roast feed to dead roast feed, there is sufficient thermal capacity in the sulphation roaster to consume all of the hot acid leach filtrate. As a result, all zinc exits in the neutral leach filtrate with less than 0.01 gpl iron in solution. As such, the requirement for a Jarosite-type iron precipitation stage is eliminated and overall zinc recoveries are slightly improved over what is normally accomplished in a conventional dead roast electrolytic zinc plant, 97% versus 95%. Also, whereas 30% of the copper and silver from the feed entering a conventional electrolytic zinc plant exits with the Jarosite residue, only 2.5% of copper and 15% silver are lost in the integrated sulphation roast-dead roast operation.

A simplified flowsheet for the integrated sulphation roast-dead roast plant is shown in Figure 2, while feeds and metal recoveries are given in Table 1. The main operating conditions and overall zinc and copper extractions are shown in Table 2.

The capital cost estimate for a sulphation roast-metals recovery plant which would treat a bulk concentrate to recover 100,000 STPY slab zinc, 2149 STPY cathode copper, 204 STPY cadmium metal, as well as a lead oxide product (82.9% Pb; 66 oz/ST Ag) containing 30,242 STPY lead and 2,419,851 troy oz silver per year, and 204,772 STPY sulphuric acid (100% H_2SO_4), is 176.3 million dollars (mid 1978, Canadian). The capital cost of the plant using a

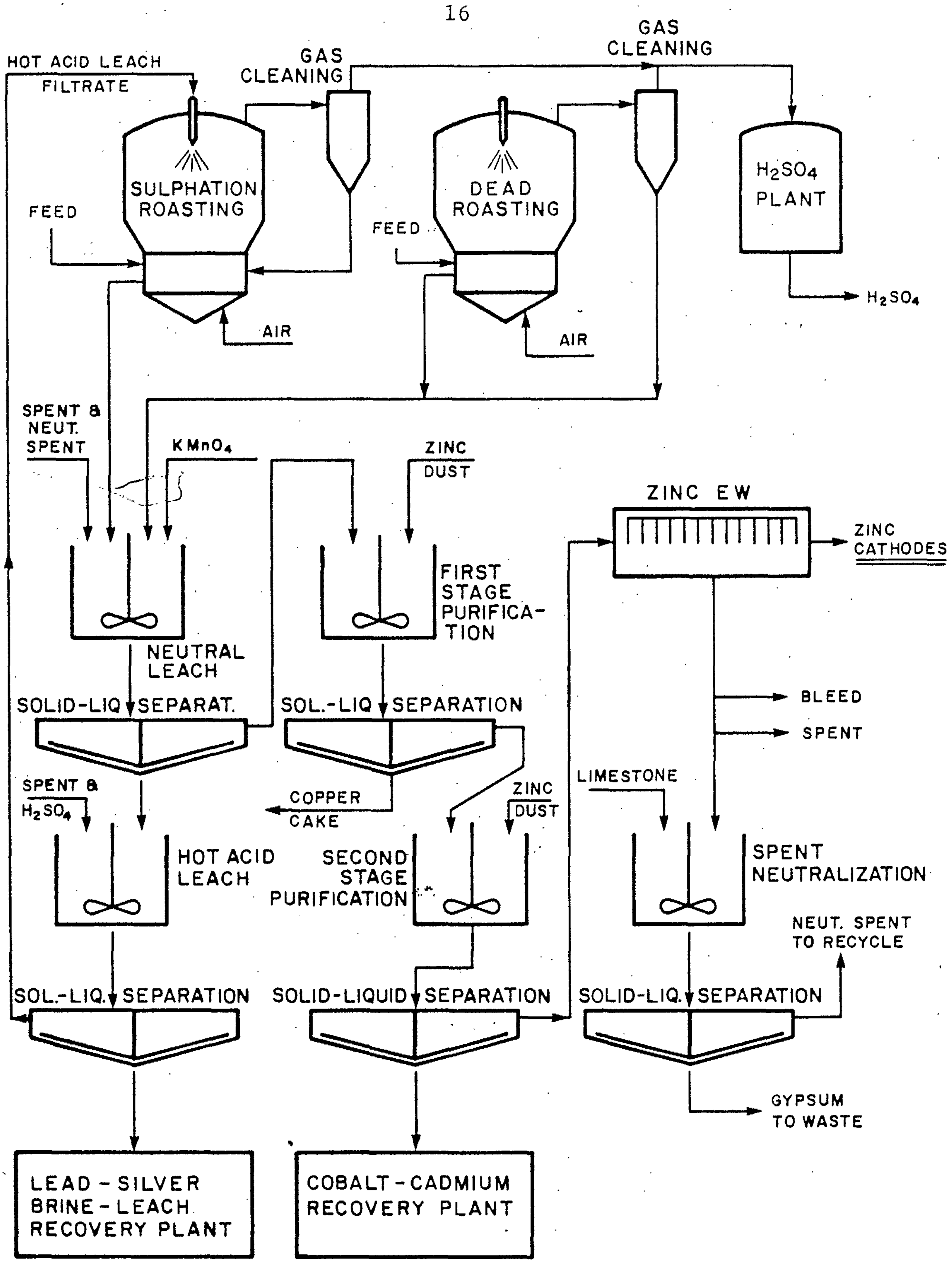

Figure 2
CONCEPTUAL FLOWSHEET OF THE R.P.C. SULPHATION-ROAST-LEACH-ELECTROWINNING PROCESS APPLIED TO AN INTEGRATED SULPHATION ROAST-DEAD ROAST PLANT

tailings refloat product, which would recover 60% less lead and approximately 50% less silver, is slightly higher at 177.4 million dollars. The capital cost estimate for the integrated sulphation roast-dead roast plant is 157.7 million dollars. These costs compare with 143.6 million dollars for a conventional dead roast electrolytic zinc plant.

Although the capital cost for the bulk concentrate sulphation roast plant is approximately 23% higher than for a conventional dead roast plant, this is more than offset by 38% higher revenue for the metals recovery plant. By comparison the tailings refloat and the integrated sulphation roast-dead roast plants generate 19% and 9% more revenue respectively than does the conventional dead roast zinc plant.

Although the capital costs for the sulphation roast plants are estimated to be accurate to ±25%, the mean is possibly high. Comparison with operating conditions in commercial sulphation roast practice would suggest that the size of individual roasters could be considerably smaller than is indicated from the scale-up data generated from the RPC 14-inch fluidized bed reactor. Two variables, retention time and air ratio, are of particular importance in determining roaster capacity. It is believed that, in commercial reactors both of these would be considerably less than was applicable to the RPC 14-inch diameter roaster. As a result, the size of commercial reactors could be 25 to 30% (hearth area) smaller than indicated from the scale-up data. This size reduction would effectively reduce the capital cost of the roasters by approximately 30%.

A reduced air ratio in commercial plant operation, over what was possible in the mini-roaster, would also result in an increased SO_2 concentration in the roaster off-gas, and hence a reduction in the capital required for the acid plant. For example, a reduction in the air ratio from Q = 1.8 to 1.6 would result in a reduction in the capital cost of the acid plant by 7 to 9%.

Scale-up costs, based on data from the mini-pilot plant, particularly from the roasting section, are probably conservative and full pilot plant data may well indicate considerable capital cost reductions for a commercial sulphation roast plant.

Based on the capital costs derived from the mini-pilot plant results, the simple return on investment (before taxes), for the bulk concentrate feed, using the metal prices and payment for concentrate guidelines established by the Advisory Committee, is 15%. This compares favourably with 6% for a conventional dead roast zinc plant.

Although 30% payment for the bulk concentrate was recommended by the Advisory Committee (CANMET), it is felt that 35% payment would be more realistically typical of what would have to be paid, for <u>true</u> bulk concentrates. At this payment, the return on investment for the metals recovery plant is reduced to 12%. This return is based on sale of sulphuric acid at $15/ST. With no payment for acid the return is reduced further to 10.5%. However, with the price of acid increased to the current f.o.b. acid plant price of $54.20/ST, the return on investment increases to 16.5%. $15/ST for acid is considered a reasonable assumption considering that the producer would have tide water access to offshore markets.

The sulphation roast plant based on the tailings refloat product, for which payment is made at 30% of full metal value, is only slightly less profitable than the bulk concentrate plant, i.e. 11.5% versus 12%. However, whereas payment of 35% of full value for a true bulk concentrate may be low, considering that all metals recovered in flotation are in the bulk, the 30% of full value as payment for the tailings refloat may be too high considering it is a 'scavenger' product. Thus the tailings refloat plant may in reality be the more profitable of the two plants. This comparison of two feeds at different payments underlines the sensitivity of plant profitability to the amount paid by the smelter for concentrates.

Also important in the comparison is that, whereas tailings refloat and middlings products are readily available at predictable costs, and in quantities sufficient for the integrated sulphation roast-dead roast plant, if not for a sulphation plant alone, the availability of true bulk concentrates is in question. There is insufficient flotation data to support the basic assumption that metal recoveries in a bulk concentrate can be improved over those previously obtained by Brunswick Mining and Smelting, to 95% Zn, 70% Pb, 70% Cu and 70% Ag, as has been assumed in the guidelines provided.* It might also be questioned whether flotation results from the Brunswick ore would be applicable to other ores in the area.

It is proposed that the best application for a technology such as the RPC Sulphation Roast Process which is capable of treating low zinc-high iron concentrates in the Bathurst-Newcastle area is as a custom type smelter treating scavenger and middlings products from the two present producers and from new mines. This mode of application would enable a reduction in milling costs through the elimination of present complex recirculating load problems, thereby improving the profitability of the mine-mill operation. At the same time, the metals recovery plant would be provided with a lower cost concentrate than would otherwise be possible from a true bulk concentrate.

Two scenarios based on the use of presently available 'scavenger' products were examined:

1) a Heath Steele Model which uses a modified bulk concentrate from Heath Steele and tailings refloat and zinc middlings products from Brunswick Mining and Smelting, and,

2) a Brunswick Model, or integrated sulphation roast-dead roast plant which would employ high grade zinc concentrates in the dead roast section and low grade tailings refloat and zinc middlings or a 'dezincing' product from the copper-lead circuit in the sulphation roaster.

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* Recent laboratory studies at CANMET achieved recoveries of 94% Zn, 84% Pb, 80% Cu, 80% Ag.

The Heath Steele Model would allow the recovery of a copper concentrate for sale to a copper smelter and a bulk concentrate to the proposed metals recovery plant from 4000 TPD of ore. Return on investment, based on a \$0.32/lb zinc price and payment for the bulk at 36% of full value for contained metals is approximately 13%. In addition to enhanced profitability at the mine, achieved through simplification of existing flotation circuits, overall metal recoveries (as value), from the Heath Steele ore would be increased by 23%.

The Brunswick Model would result in a return on investment of approximately 10% compared with 6% for a conventional dead-roast zinc plant at current metal prices. This plant would provide Brunswick Mining and Smelting with a more efficient integrated mine-mill-smelter complex than would be possible with a conventional zinc plant. It would allow the production of higher grade zinc products for sale out of province and result in an additional recovery of 22.0 STPD of zinc, 8.6 STPD lead, 1.4 STPD copper and 105 oz/day silver over what is now achieved in flotation concentrates (including a 50% zinc tailings refloat product). The lead oxide-silver product produced from this plant would also benefit the Brunswick lead smelter operation. As feed to the sinter machine it would help reduce recycles and increase the overall per unit through-put of the lead smelter.

Additional benefits may be possible for the Brunswick Model through the use of stockpiled lead slag, as a replacement for limestone, to neutralize spent zinc electrolyte. Use of lead slag in this manner could be an alternative to fuming as a means of recovering the appreciable quantities of zinc from the slag as well as lead and silver.

In summary, the RPC Sulphation Roast Process appears to be not only technically feasible but also an efficient and economically attractive alternative to a conventional zinc plant for northern New Brunswick. The close similarity to conventional zinc plant practice is a particularly important attribute of the Process which should help to enhance industry interest.

The presence of such a metals recovery plant in New Brunswick would make it possible for companies to produce higher quality concentrates marketable to existing smelters at premium prices and at the same time improve metal recoveries by producing middlings and scavenger products for further processing in the proposed sulphation roast or integrated sulphation roast-dead roast plants.

3.0 RECOMMENDATIONS

As the technical feasibility of the RPC Sulphation Roast Process has been demonstrated in mini-pilot trials and as the preliminary economic evaluation is encouraging, it is recommended that full scale piloting (10 MTPD feed = 3 MTPD slab zinc) of the Process be undertaken. This is required to obtain more accurate process parameters which will be required for commercial plant scale-up. Particularly important areas which can be properly evaluated only in a full pilot are: roasting conditions, gas cleaning, spent electrolyte neutralization, zinc dust purifications and cadmium recovery.

Several areas require additional work in the mini-pilot plant either prior to, or in conjunction with, the pilot plant trials. These are as follows:

1. The lead-silver recovery, brine leach circuit, should be evaluated technically and mass and thermal balances generated.

2. Additional tests in a modified 14-inch roaster are required to assess the influence of varying levels of copper with lead on bed fluidization characteristics. There was some difficulty in the initial trials with bed 'stickiness' with feeds containing both high lead and copper and work is required using higher air velocities, simulating conditions in a commercial scale reactor, to evaluate the effect of composition. As most ore deposits contain a higher copper/zinc ratio than Brunswick ore, it is likely that bulk concentrates or middling products will be higher in copper than the tailings refloat concentrate tested in the mini-pilot program.

3. The use of lead slag from the Belledune Lead Smelter as an option to limestone in neutralization of spent zinc electrolyte, as described in

Appendix A5, requires complete evaluation. Benefits to accrue would be:

i) recovery of zinc from the presently stockpiled slag, a waste product which might not be economically feasible to process by fuming,

ii) increased zinc throughput in the zinc plant,

iii) recovery of lead and silver from the slag,

iv) elimination of limestone for spent electrolyte neutralization,

v) increased SO_2 content from the roaster to the acid plant.

4. The use of an inert fluidized sand bed or the sulphation roaster itself to decompose spent zinc electrolyte using off-gas heat from the dead roaster in the integrated sulphation roast-dead roast plant as an option to limestone neutralization.

5. Solvent extraction tests using the reagents LIX 65N and LIX 34 (General Mills Chemicals) should be carried out to compare their behaviour to that of LIX 64N at 45°C the minimum temperature of the neutral leach. A suitable diluent should be found so that solvent extraction can take place at 60°C and the requisite test work carried out. The higher temperature would eliminate the need for costly solution cooling and heating.

In addition to the above test work, consideration should be given to carrying out a detailed cost/benefit study to establish acceptable payments for various middling and bulk concentrates from both producing mines with perhaps some estimates derived for the Anaconda Caribou ore and other singificant, yet undeveloped, ore bodies in the area.

4.0 BASIC PRINCIPLES OF THE RPC SULPHATION ROAST PROCESS

4.1 THEORY

The RPC sulphation roast process is a combined pyro-hydrometallurgical route to transform relatively insoluble metal sulphides to the soluble sulphate or oxysulphate state. By controlling the roasting conditions, the process can be made selective so that the more valuable metals will form soluble sulphates while the iron, which constitutes the main undesirable element in the feed, is transformed into relatively insoluble hematite. This allows effective separation of valuable metals such as zinc and copper from iron and other gangue material by water or dilute sulphuric acid leaching and further recovery by electrowinning or chemical precipitation.

The overall sulphate formation reaction for a metal M can be expressed as follows:

$$MS + 2O_2 \rightleftarrows MSO_4$$

In some cases, metal sulphates are formed through several intermediate compounds, usually their oxides and oxysulphates, following the general reactions:

$$\begin{aligned} MS + 1.75\ O_2 &\rightleftarrows 0.5\ MO.MSO_4 + 0.5\ SO_2 \\ MO + 0.5SO_2 + 0.25\ O_2 &\rightleftarrows 0.5\ MO.MSO_4 \\ MO.MSO_4 + SO_2 + 0.5\ O_2 &\rightleftarrows 2\ MSO_4 \end{aligned}$$

In addition to these single metal compounds, solid state reactions between condensed phases also take place, usually a metal oxide or oxysulphate with iron oxide, as in the following reactions:

$$MO + Fe_2O_3 \rightarrow MO.Fe_2O_3$$

$$MO.MSO_4 + Fe_2O_3 \rightarrow MO.Fe_2O_3 + MSO_4$$

Such reaction products are classified as ferrites.

The distribution of metals among the above compounds formed during roasting is dependent on the temperature and gas composition as well as the physical conditions existing in the roaster. These variables can be regulated so that only one compound is thermodynamically stable under equilibrium conditions.

Kellogg's stability phase diagrams, which illustrate the stability fields of compounds as a function of the partial pressure of SO_2 and O_2 at a given temperature, are useful in predicting the formation of compounds during roasting. The combined phase diagram calculated at 685°C for the Fe-S-O, Zn-S-O, Cu-S-O and Pb-S-O systems is shown on Figure 3. The cross hatched region ABCD represents the range of gas composition (1-10% SO_2; 1-10% O_2) prevalent inside a fluidized bed roaster operating as a sulphation reactor. Under normal roasting conditions, zinc, copper, and lead will form their respective sulphates, while iron will oxidize to hematite, as shown in Figure 3.

For SO_2 concentrations of approximately 3 to 15%, $ZnSO_4$, $CuSO_4$ and $PbSO_4$ are the only stable sulphates, while

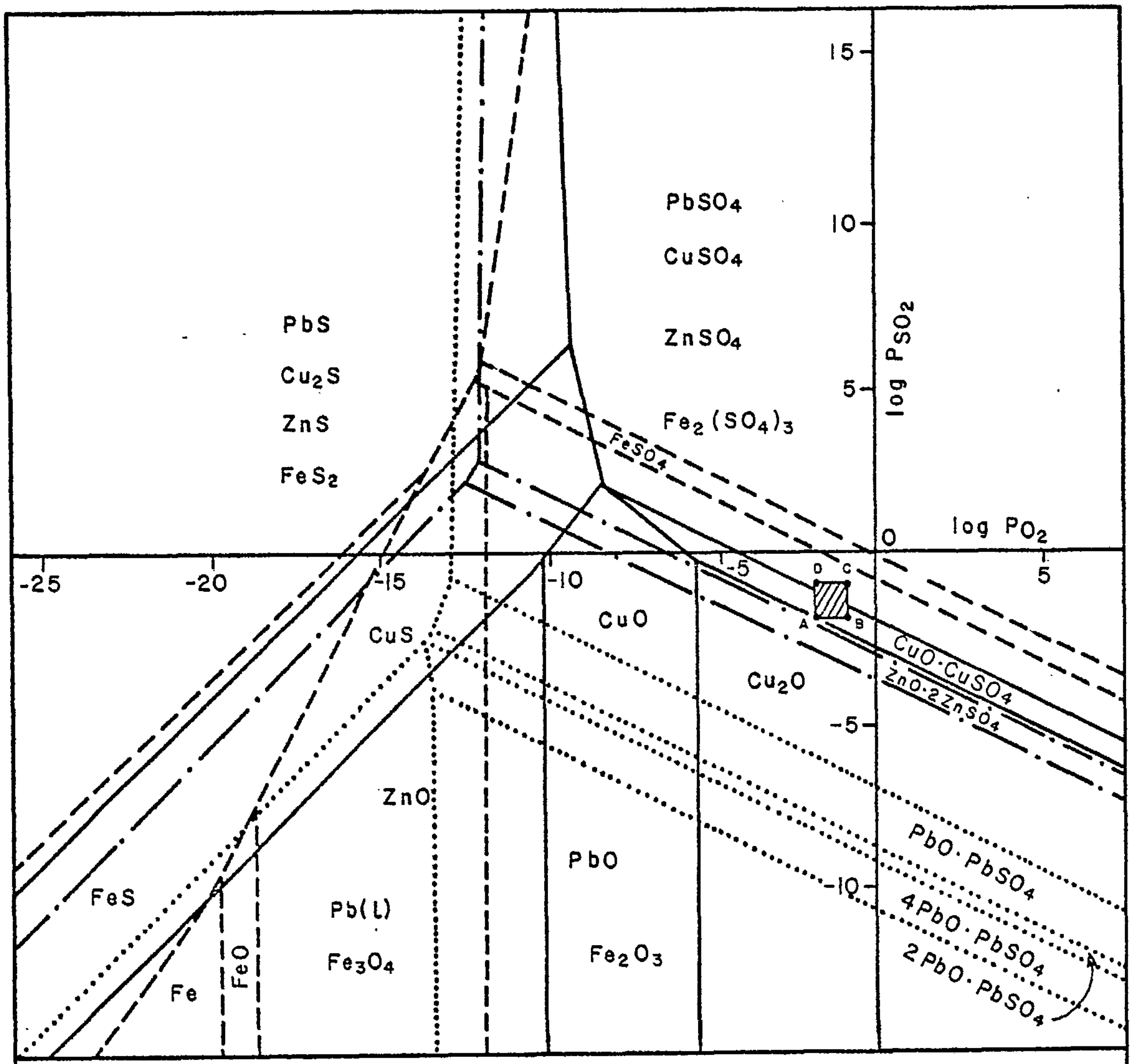

Figure 3

STABILITY PHASE DIAGRAM

—·—·— Zn - S - O system
———— Cu - S - O system
– – – – Fe - S - O system
·········· Pb - S - O system

A = 1 % O_2 + 1 % SO_2 + 98 % N_2
B = 10 % O_2 + 1 % SO_2 + 88 % N_2
C = 10 % O_2 + 10 % SO_2 + 80 % N_2
D = 1 % O_2 + 10 % SO_2 + 88 % N_2
T = 685°C Total Pressure = 1 Atm.

iron sulphates, if present, decompose to Fe_2O_3 in the roaster. For SO_2 concentrations less than 2%, some copper and zinc oxysulphates can be present. These compounds are soluble in dilute sulphuric acid.

Any increase in SO_2 concentration from point C will favour the formation of $FeSO_4$ which is a water or dilute acid soluble compound and hence would result in increased quantities of iron in leach solutions, an undesirable feature in zinc and copper electrolytic plant practice.

Diagonally opposite at point A, where the gas is composed of 1% SO_2 and 1% O_2, $ZnSO_4$, $CuO.CuSO_4$, $PbSO_4$ and Fe_2O_3 are the stable phases. Any further downward trend from A (i.e. lower SO_2 concentration), will favour the formation of $ZnO.2ZnSO_4$. This should also be avoided as $ZnO.2ZnSO_4$ can react with Fe_2O_3 during roasting to form $ZnO.Fe_2O_3$ (ferrite) which is soluble only in strong acid solutions.

At point B, (1% SO_2, 10% O_2) both $CuSO_4$ and $CuO.CuSO_4$ are stable phases, together with $ZnSO_4$, $PbSO_4$ and Fe_2O_3. At point D (10% SO_2, 1% O_2) the compounds formed are the same as at point C.

A decrease in roasting temperature will displace all phase boundaries in Figure 1 towards the lower left. This will effectively increase the stability fields of all sulphates including the iron sulphates, $FeSO_4$ and $Fe_2(SO_4)_3$. As the iron level in leach solutions should be kept to a minimum, it is important that the roasting temperature not be appreciably below 650°C. Conversely, increasing roasting temperatures above 685°C will favour the formation of $ZnO.2ZnSO_4$, and undesirable ferrite. The latter being formed either directly from sphalerite or by reaction of $ZnO.2ZnSO_4$ with Fe_2O_3.

Although the formation of zinc ferrite is normally favoured at roasting temperatures higher than 700°C, its stability field, considering the uncertainty in the available thermodynamic data, could extend to lower temperatures (Figure 4). The reaction between zinc oxide and iron oxide to form ferrite can occur as low as 590°C, while the lowest temperature at which copper ferrite forms by equilibrium reaction of copper oxide with iron oxide is 720±20°C. Although zinc ferrite is a metastable phase under sulphation roasting conditions, its resulphation by SO_2 is a kinetic problem with rates being diffusion controlled. This probably accounts for the presence of ferrite in roaster calcines.

Kinetic factors and the complex gas-solid behaviour within the fluidized bed can cause variations in reaction products from those predicted from phase diagrams, which depict ideal, pure systems. This, plus backmixing conditions in fluidized bed roasters, results in calcines which normally contain one predominant phase (the theoretically stable phase) and several others in minor amounts. For example, in the sulphation roasting of zinc concentrates, the zinc products formed in the calcines are, in order of abundance, $ZnSO_4$, $ZnO.2ZnSO_4$, $ZnO.Fe_2O_3$ and (Zn,Fe)S, the latter being due to the backmixed nature of the roaster.

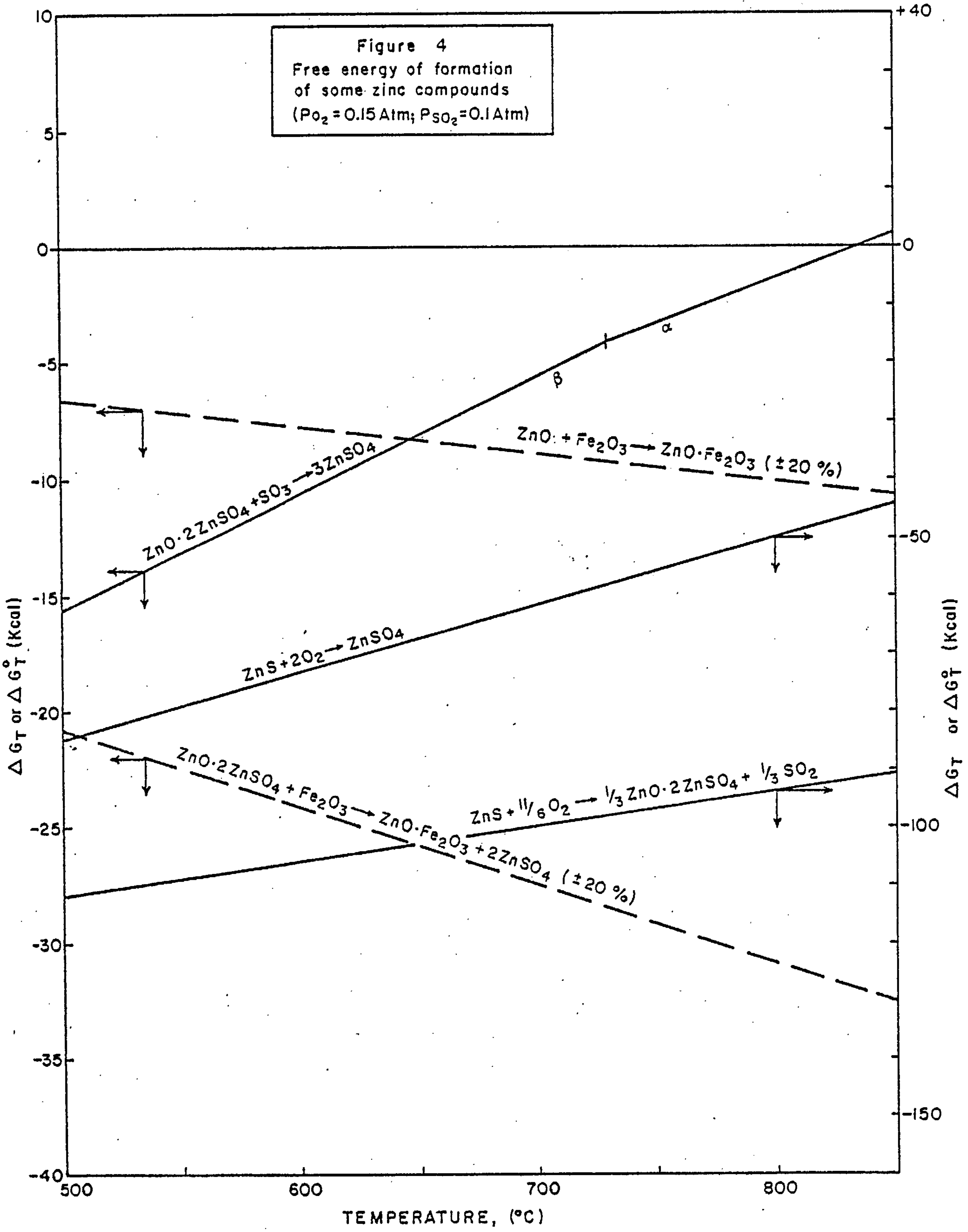

Figure 4
Free energy of formation of some zinc compounds
(P_{O_2} = 0.15 Atm; P_{SO_2} = 0.1 Atm)

4.2 DEVELOPMENT OF THE RPC SULPHATION ROAST PROCESS

421 Previous Work

Two major attempts have been made in the past to use the promising potential of sulphation roasting as a means of recovering valuable metals, principally zinc and copper, from relatively low grade concentrates. The Dowa Mining Company Ltd. at Kosaka, Japan, in 1954 through to 1955, operated a commercial sulphation roast plant on an 80 TPD basis using a sulphide feed averaging 16.7% Zn, 10.6% Cu, 2.4% Pb, 39.0% S and 26.0% Fe. Although extractions and recoveries were high for copper, 94 and 93% respectively, corresponding extractions and recoveries for zinc were relatively low, 88 and 65% respectively. In addition to poor recoveries, problems with recovery of copper and zinc from a single electrolyte were experienced. This work was carried out before the advent of copper solvent extraction technology which might have been used to solve the problem Dowa faced in having to neutralize a spent copper electrolyte carrying 100 gpl zinc prior to zinc electrolysis. At high solution zinc loading considerable zinc is lost to the gypsum residue. This accounts for the spread between zinc extraction 88% and zinc recovery 65%. Roasting was carried out at 700°C, hence, although unstated, zinc ferrite would be present in the roaster calcine. Acid strength in the leach was at a maximum of 5 to 7 grams per litre and the temperature 70°C for a 5 hour retention. Under these conditions only a portion of the ferrite would be leached and this accounts for the relatively low zinc extraction from calcines. However, ferrite leaching did occur such that 28 kg MnO_2 per ton of feed was required to keep iron in solution below 2 gpl, a level which is considered too high for present zinc plant practice.

The St. Joseph Lead Company initiated a sulphation roast test program on New Brunswick ores in 1954. With assistance from the Battelle Memorial Institute in Columbus, Ohio, a program was carried out from bench scale experiments through to full 10 TPD pilot trials which were completed in late 1957.

Best results were obtained from unconcentrated ore which was fed as a slurry with spent electrolyte to the roaster. This ore averaged 8.71% Zn, 3.49% Pb and 0.24% Cu. Over a continuous one month period of operation, the average level of soluble zinc in the calcines was reported as 94.4% while zinc extraction was 90.4%. Similarily 'soluble' copper was 95.8% with only 85.6% extraction. No explanation was offered for the discrepency between 'soluble' metals in calcines and metals extraction, the difference is believed to be due to filtration difficulties possibly resulting from iron hydroxide precipitation.

The iron in the leach solution was high, 5-10 gpl, and some of this might be attributed to the low roasting temperatures (599-643°C) employed. In the present program, it was found that higher temperatures are required if the formation of iron sulphates are to be avoided during roasting. Ferrite in the calcines could also have been a contributor to this high iron. At St. Joseph Lead, the calcines exiting the roaster were cooled to about 300°C and quenched in dilute acid leach solutions which contained 20 gpl sulphuric acid at the outset and was controlled to 10 gpl in the pregnant leach filtrate. Although the acid strength was low, the temperature was high enough (boiling at the outset) and the retention time sufficient to effect dissolution of some ferrite in the calcine.

Attempts by St. Joseph Lead to roast a bulk concentrate (33.7% Zn, 9.51% Pb, and 0.35% Cu) produced from the Brunswick ore were completely unsuccessful. Extreme difficulties were encountered in maintaining bed fluidization and the test program had to be abandoned. St. Joseph Lead used 2 wt% sodium sulphate as an addition to the roaster feed to act as a sulphation promoter and these difficulties may have been a result of relatively low temperature metal-sodium sulphate eutectics which cause bed stickiness and agglomeration. RPC experience with sodium sulphate addition to a feed of similar zinc content was equally bad. Even with very minor additions (<1 wt%) poor bed fluidization was experienced.

In addition to the Dowa and St. Joseph Lead work on sulphation of sulphide concentrates, two sulphation roast plants have been operated as a means of resulphating ferrite from dead roast-neutral leach and hence increasing overall zinc recoveries. These were in operation in 1965 at the

Mitsubishi Metal Mining Company, Akita plant, in Japan and in 1964 at the Det Norske Zinkkompani plant in Norway. In both operations the ferrite residue was roasted together with pyrite concentrates to effect sulphation of ferrite. With the development of the Jarosite Process, sulphation of ferrite in this manner was abandoned.

The principal problems encountered by Dowa and St. Joseph Lead in their respective attempts to develop a sulphation roast process seem to center on the formation of zinc ferrite in the roaster calcine. In both cases, ferrite was probably responsible for the relatively low extraction, by present day zinc plant practice standards, of zinc from calcines and for the unacceptably high levels of iron, 5-10 gpl in the case of the St. Joseph Lead trials, in the leach filtrates. Such high levels of iron would require some iron precipitation procedure. However, both the Dowa and St. Joseph Lead programs predated the Jarosite and similar improvements in the iron removal procedure from zinc plant leach solutions. The resultant iron hydroxide precipitate would have created filtration difficulties and high zinc losses to the residue. This is, in effect, what was experienced in the St. Joseph Lead trials.

The RPC bench scale roaster trials (1) have provided the necessary physical-chemical information on the formation of ferrite in roaster calcines and the conditions that would be required to limit its formation.

Although relatively low levels of ferrite in calcine (approximately 5 wt %) were achieved in the bench scale trials using a 30% zinc concentrate, this is artifically low in that much sphalerite was left unreacted, i.e. greater than 17.3% of zinc in the calcines remains as sphalerite. In comparison, calcines produced from the same feed in the 14-inch reactor contained only 1.5-3.7% zinc as sphalerite. Reduction of unreacted sphalerite in the bench scale calcines to these low levels would result in a corresponding increase in ferrite to approximately 7 to 8%. This is entirely consistent with the lowest level of ferrite achieved in the 14-inch roaster trials (i.e. 8 wt% ferrite BMD 45) at the same roasting temperature. At this level it became obvious that if high zinc extractions were to be achieved then some dissolution of this ferrite would be necessary. With dissolution of ferrite, iron in the leach solutions rises substantially above 1 gpl, the maximum acceptable level for zinc plant practice. Although suitable iron precipitation techniques such as the Jarosite Process are now available to handle high iron solutions, a simpler technique was required for handling this iron problem, if the sulphation roast process was to be feasible. From this evolved the concept of recycling the filtrate from a second stage hot acid leach back to the sulphation roaster as the method of iron control. This concept is the basis of the RPC sulphation roast process described below. (6).

422 The RPC Sulphation Roast Process

The RPC Sulphation Roast Process consists of a fluidized bed sulphation roast followed by a two-stage leach of roaster calcines. The first stage is a dilute sulphuric acid leach entirely similar to the neutral leach employed in conventional zinc plant practice. Approximately 85% of the zinc is leached from calcines with the filtrate, exiting to solution purification, containing less than 1 gpl iron. The balance of the zinc remains in the neutral leach residue as ferrite and remnant sphalerite. This residue is then subjected to a hot strong acid leach and the resultant filtrate, which is very high in iron, is recycled to the sulphation reactor for fixation of iron as insoluble hematite. In this manner virtually all zinc exits through the neutral leach to purification with solutions containing less than 1 gpl iron.

Four basic simple principles are embodied in the filtrate recycle concept:

i) that zinc ferrite and remnant sphalerite can be leached almost completely (95 or 80% respectively) in hot, concentrated sulphuric acid solution to produce, after solid-liquid separation, a concentrated sulphate solution containing zinc, iron and copper and free sulphuric acid.

ii) that given fixed roasting conditions, the mineralogical composition of the final calcine is independent of the form in which the metal(s), zinc, iron and copper in this case, are fed into the reactor, i.e., as solid feed or dissolved in solution.

iii) that because of the endothermic character of the decomposition reactions, solutions containing $ZnSO_4$, $FeSO_4$, $Fe_2(SO_4)_3$ and H_2SO_4 under sulphation roasting conditions will produce only $ZnSO_4$, Fe_2O_3, SO_2, SO_3 and H_2O.

iv) that the heat of reaction released during the sulphation roasting of sulphide concentrates, in a fluidized bed reactor, is sufficient to decompose all of the hot acid leach filtrate generated from calcines containing up to 15 wt% ferrite and 1.5 wt% sphalerite.

A simplified flowsheet of the RPC Sulphation Roast Process using only sulphation roasters is shown in Figure 1. The recirculation of the hot acid leach filtrate to the reactor increases the zinc extraction in the neutral leach to over 98.5% and copper extraction to over 98%. Purification includes solvent extraction-electrowinning of copper, iron oxidation and precipitation, cobalt cementation and cadmium recovery. Zinc is then recovered by electrolysis.

The solid residue from the hot acid leach contains all of the lead as sulphate and all of the silver from the calcine. A brine ($NaCl/CaCl_2$) leach is used to recover lead and silver as a lead oxide product for sale to a lead smelter.

The principles embodied in the RPC Sulphation Roast Process can be used to advantage in an integrated sulphation roast-dead roast plant as well as a sulphation roast plant. Such a zinc plant, instead of the usual two dead roasters, would employ one reactor for sulphation roasting and the other for dead roasting. This integration would not only improve the overall efficiency of an integrated mine-mill-smelter operation in that low grade zinc concentrates could be treated in the sulphation roaster, but would also eliminate the requirement for a Jarosite-type process in conventional zinc plant practice. Given the appropriate distribution of feed to each roaster, there would be sufficient thermal excess in the sulphation reactor to decompose the total hot acid leach filtrate derived from the combined neutral leach residues from both roaster calcines. A simplified flowsheet for this integrated plant is shown in Figure 2.

Copper is recovered as a cement product if amounts are insufficient to warrant solvent extraction and electrowinning.

5.0 RPC SULPHATION ROAST PROCESS: FULL PLANT SCALE-UP

5.1 SULPHATION ROAST PLANT-BULK CONCENTRATE OR TAILINGS REFLOAT PRODUCT

Two general models based on a relatively high lead (bulk concentrate) and low lead (tailings refloat product) 30-32% zinc concentrate to produce 100,000 STPY of slab zinc were considered. Since the basic plant layout is the same for both the bulk concentrate and tailings refloat product, they are analyzed together.

511 Characterization of the Feeds

The tailings refloat product was obtained from the second stage cleaner of the Brunswick Mining and Smelting Corporation Ltd., tailings refloat circuit. Assays and mineralogical compositions for this feed are given in Table 3. The bulk concentrate composition shown in Table 3 was proposed by CANMET as being "typical" of the bulk concentrates from New Brunswick massive sulphide ores. Recoveries in the "bulk" are assumed to be 95% Zn, 70% Pb, 70% Cu and 70% Ag.[*] This bulk, although perhaps typical of what could be produced at Brunswick Mining and Smelting, would not be representative of true bulk concentrates produced from other deposits. Whereas the Brunswick bulk contains low levels of copper, a similar concentrate produced at Heath Steele, for example, would contain 4.27 wt% copper. Similarily an Anaconda bulk would have 2.47 wt% copper. To reproduce a bulk concentrate, similar to that of the proposed model from most New Brunswick

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[*] see footnote p. 20

TABLE 3

CHEMICAL AND MINERALOGICAL COMPOSITION OF FEEDS TO SULPHATION ROAST PLANTS

Bulk Concentrate		Tailings Refloat	
Chemical Composition (wt %)	Mineralogical Composition (wt %)	Chemical Composition (wt %)	Mineralogical Composition (wt %)
zinc = 32.0	sphalerite ((Zn,Fe)S) = 54.1	zinc = 30.0	sphalerite ((Zn,Fe)S) = 50.7
lead = 10.0	galena (PbS) = 11.6	lead = 3.6	galena (PbS) 4.2
copper = 0.7	chalcopyrite ($CuFeS_2$) = 2.0	copper = 0.6	chalcopyrite ($CuFeS_2$) = 1.7
iron = 18.0	pyrite (FeS_2) = 29.7	iron = 22.0	pyrite (FeS_2) = 37.6
sulphur = 36.0	gangue = <3.0	sulphur = 39.0	gangue = <3.0
silver = 8.8 oz/Ton		silver = 4.4 oz/Ton	

deposits, the flotation of a copper concentrate in addition to a bulk concentrate would be required.

512 General Flowsheet

The general flowsheet for a sulphation roast plant to treat either the bulk concentrate or tailings refloat product is shown in Drawing No. 1. Both plants are designed to produce 100,000 STPY of slab zinc, plus quantities of copper, lead, silver, cobalt and cadmium. Yearly production and revenue for all recovered metals and sulphuric acid are given in Section 6 (Tables 10 and 12).

The main plant, for purposes of mass balance calculations, is divided into six sections, each of which has a detailed flowsheet, as follows:

1. Roasting (Drawing No. 2).
2. Leaching and Copper Solvent Extraction (Drawing No. 3).
3. Purification and Electrowinning (Drawing No. 4).
4. Cobalt-Cadmium Plant (Drawing No. 5).
5. Lead-Silver Recovery Plant (Drawing No. 6).
6. Spent Electrolyte-Limestone Neutralization (Drawing No. 7).

Detailed mass balances for each section are given in Appendix A2.

.1 Roasting Section

The detailed flowsheet for the roasting section is shown in Drawing No. 2.

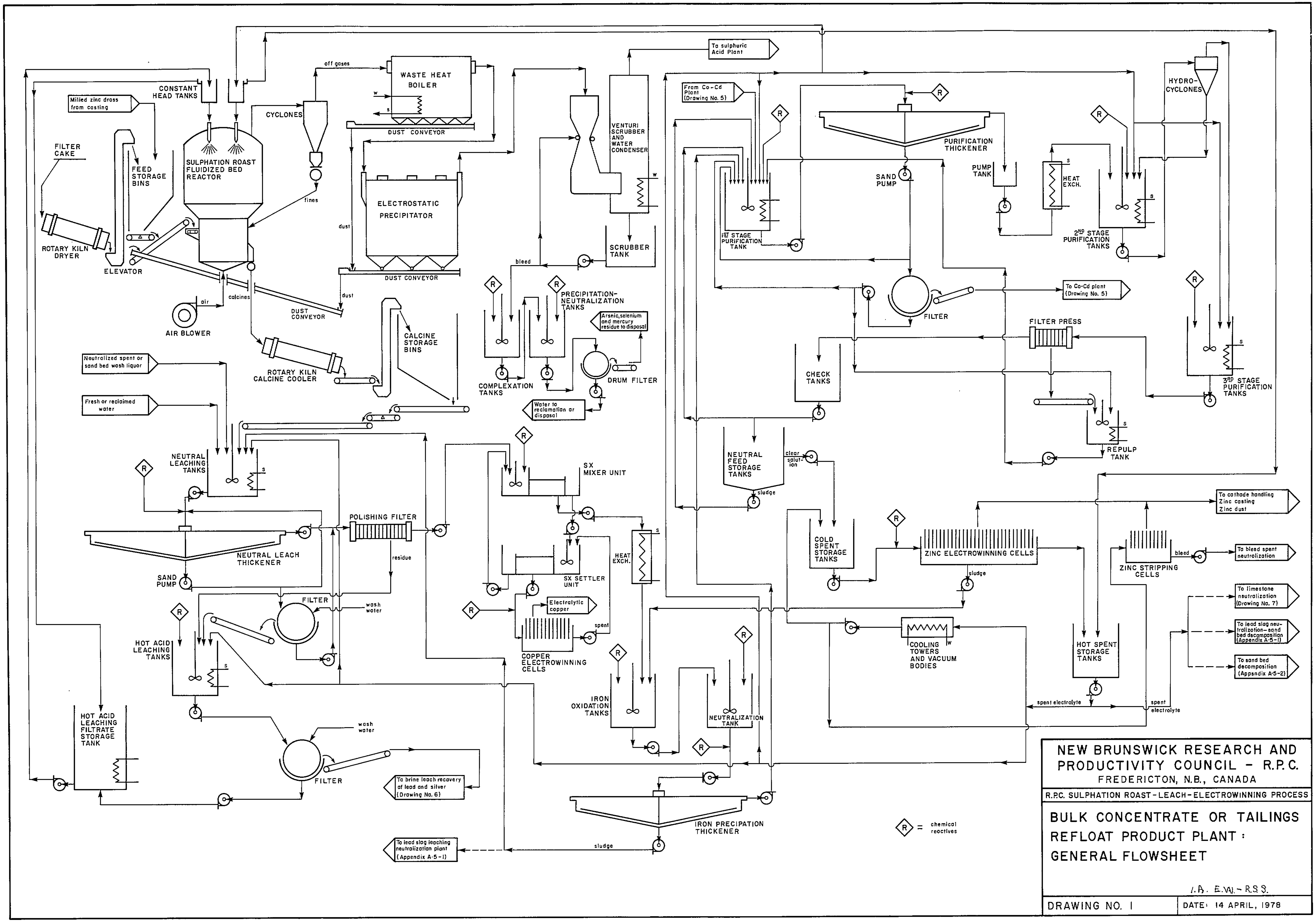
Complete mass and thermal balances for the bulk concentrates in this section are given in Tables A2-1, A2-7, A2-8 and A2-25; and in Tables A2-9, A2-15, A2-16 and A2-25 for the tailings refloat product.

Flotation concentrates, as filter cake containing approximately 8% water, are dried to 2.5% moisture in a rotary drum dryer. The dried product is crushed to 100% minus 100 mesh prior to roasting. Storage for dry concentrate feed is sufficient for 48 hours of operation.

Two Lurgi-type fluid bed roasters are used in each plant, nominal capacity of each roaster is 420 MTPD for the bulk concentrate plant and 450 MTPD for the tailings refloat plant. The reactors are fed by slinger belt feeders, while the calcine is discharged through rotary star valves to a rotary drum calcine cooler. Calcines are stored in 60 day stockpile bins for further treatment in the leaching section.

The strong acid leach filtrate from the ferrite residue dissolution is recirculated to the roaster using an air operated gun inserted through the roof of the reactor. Pressure to the gun is kept constant by using a constant head tank above the reactor. An additional gun, fed with spent electrolyte, is operated to maintain the bed at constant temperature when the reactor has a surplus of heat.

Off-gases are cleaned in a one or two step hot cyclone system. Calcines collected in the cyclones are discharged through gas lock rotary star valves back into the reactors.



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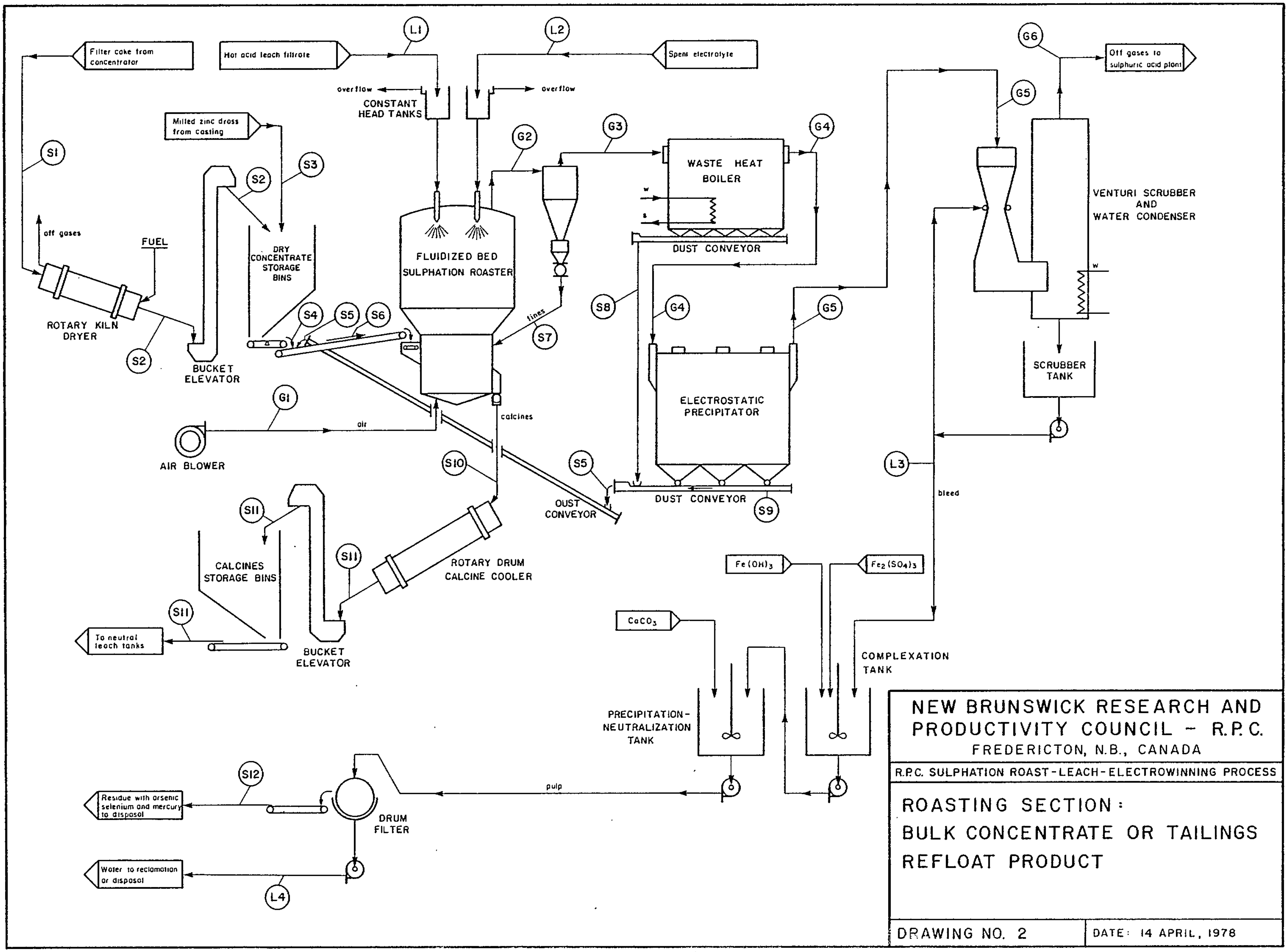
R.P.C. SULPHATION ROAST-LEACH-ELECTROWINNING PROCESS

BULK CONCENTRATE OR TAILINGS REFLOAT PRODUCT PLANT : GENERAL FLOWSHEET

I.B. E.W. - R.S.S.

DRAWING NO. 1

DATE: 14 APRIL, 1978



NEW BRUNSWICK RESEARCH AND PRODUCTIVITY COUNCIL – R.P.C.
FREDERICTON, N.B., CANADA

R.P.C. SULPHATION ROAST-LEACH-ELECTROWINNING PROCESS

ROASTING SECTION:
BULK CONCENTRATE OR TAILINGS REFLOAT PRODUCT

DRAWING NO. 2

DATE: 14 APRIL, 1978

Off-gases from the cyclone system, at 650°C (or lower temperature depending upon the heat recovered by vaporizing the HAL solution and/or spent electrolyte in the freeboard of the roasters) are cooled down to 350°C in two waste heat boilers, one per reactor, and the off-gases finally cleaned in a single electrostatic precipitator.

The proportion of elutriated material collected in the cyclone system, waste heat boilers and electrostatic precipitator are 90%, 2% and 8%, respectively. A mass balance for the solid-gas separation system is given in Table A2-25. Overall collection efficiency is 99.5%.

Dust collected in the waste heat boilers and electrostatic precipitator is continuously recirculated to the reactor as it contains a relatively high proportion of soluble iron and unreacted sphalerite.

The fluidized bed reactor and cyclone systems are well insulated in order to minimize thermal losses, and use the maximum amount of heat available from the roasting reactions to decompose the recycled hot acid leach filtrate and the spent electrolyte.

Since the roaster off-gases are at a relatively low temperature, the waste heat boilers require no radiation section and are similar to those employed by Outokumpu (7) at their cobalt plant in Kokkola (Finland). The steam produced will depend on the temperature of the off-gases entering the boiler. When no latent heat from the off-gases is recovered in the freeboard of the reactor, the temperature of the gases entering the boilers will be approximately

650°C. This temperature has been assumed for the overall thermal balance calculations given in Tables A2-10 and A2-20. A complete detailed thermal balance for the tailings refloat product feed is given in Appendix A3. The excess heat in the fluidized bed using both the tailings refloat product and a bulk concentrate exceeds that required to consume all of the hot acid leach filtrate and wash water, while maintaining the off-gas temperature at 650°C.

After exiting the electrostatic precipitator, the relatively clean gases are cooled to 35°C by vaporizing water in a venturi scrubber. Water vapour from the gases is condensed using cool water which is recirculated to a cooling tower.

The condensed water contains sulphuric acid, produced from the SO_3 contained in the gases, and all of the more volatile impurities in the ore, such as mercury and chlorine, and most of the selenium and arsenic. The bleed from the scrubber-condenser circuit is treated in a conventional manner to precipitate and dispose of the impurities. The final clean gas is then fed to a sulphuric acid plant.

Reactor parameters and operation conditions are given in Table 4.

TABLE 4
REACTOR PARAMETERS AND OPERATION CONDITIONS BULK CONCENTRATE AND TAILINGS REFLOAT PRODUCT

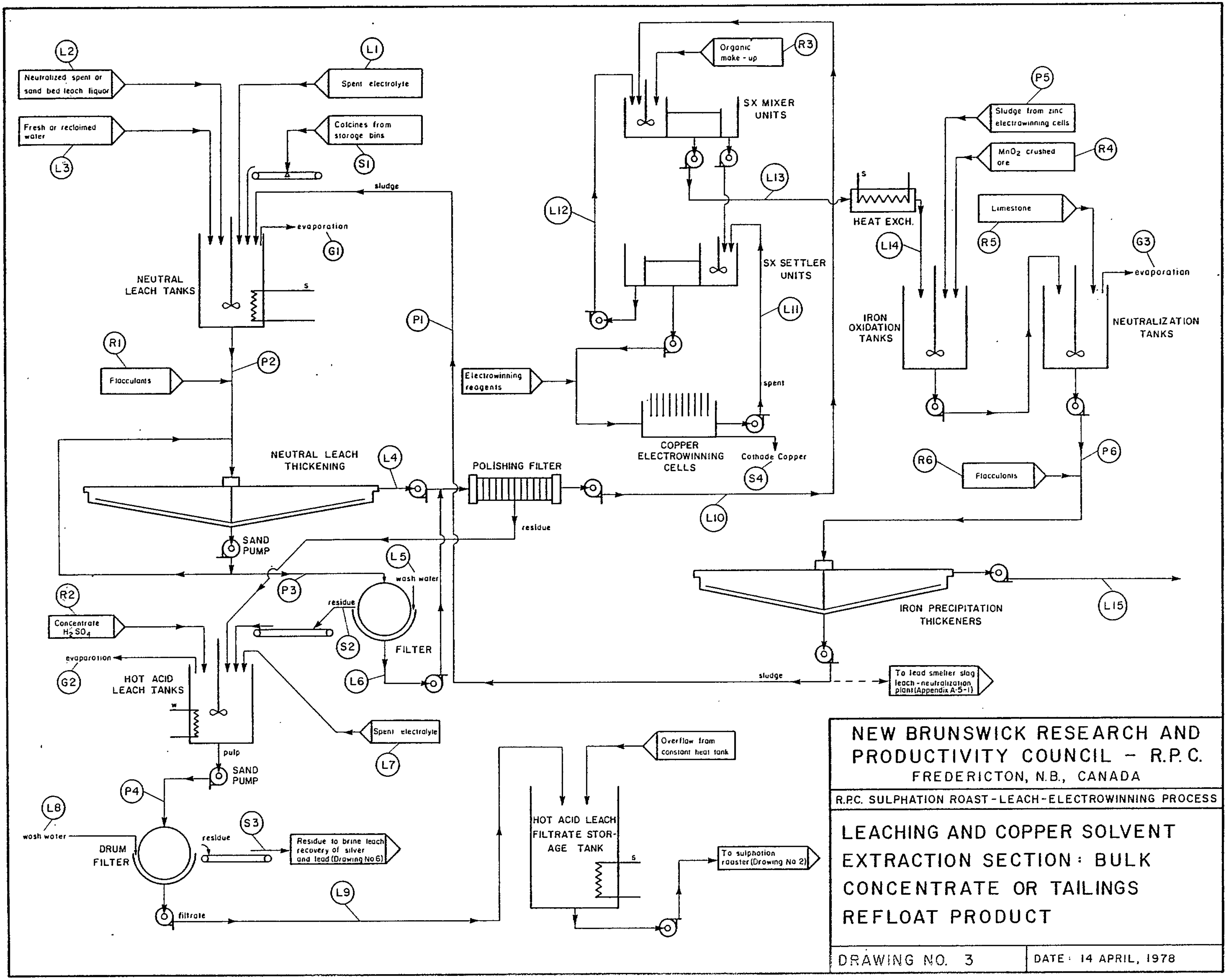
	Bulk Conc.	Tailings Reflt.
No. of reactors	2	2
Nominal capacity (each)(MTPD)	420	450
Reactor diameter (m)	11.4	11.6
Fluidized bed level (m)	1.5	1.5
Feed rate (MT/m^2-day)(each)	4.1	4.3
Roasting Temperature (°C)	685±5°C	685±5°C
Air factor (-)	1.8	1.8
Solid retention time (hrs)	5.5	5.5
Gas velocity (at 685°C)(m/sec)	0.75	0.75

.2 Leach-Copper Solvent Extraction-Iron Removal Section

The flowsheet for the section is shown in Drawing No. 3. Complete mass balances are given in Tables A2-2 and A2-10 and thermal balances in Tables A2-8 and A2-16 for the bulk and tailings refloat products respectively.

.21 Dilute Acid "Neutral" Leaching

Neutral leaching (NL) is similar to that in conventional zinc plant practice. The dissolution of zinc and copper sulphates and zinc oxysulphate is controlled by maintaining the pH at 3.6 to 4.0 for 1.5 hours by the addition of spent electrolyte. Neutralized spent electrolyte and water are also added to the leach. The ambient neutral leach temperature has been calculated to be 66°C assuming no conductive or convective heat losses. Actual plant operation, would be designed to yield a leach temperature of 45 to 52°C during continuous operation without external heating or cooling. Thickness of tank insulation would determine conductive losses.



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R.P.C. SULPHATION ROAST-LEACH-ELECTROWINNING PROCESS

LEACHING AND COPPER SOLVENT EXTRACTION SECTION: BULK CONCENTRATE OR TAILINGS REFLOAT PRODUCT

DRAWING NO. 3

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Oxidant is not directly introducted to the neutral leach, except for the manganese dioxide present in the recycle from the iron oxidation thickener plus permanganate and manganese dioxide present in the spent electrolyte. The leach solution to the neutral leach thickener contains typically 150 gpl zinc, 2 gpl copper, and less than 1 gpl iron.

.22 Neutral Leach Pulp Dewatering

The initial dewatering of the neutral leach residue takes place in a 42 m diameter thickener (based on unflocculated pulp tests). Underflow from the thickener exits at about 30% solids (by weight) and is pumped to vacuum drum filters for final dewatering, while the overflow is pumped to a polishing filter to remove residual suspended solids.

Vacuum drum filters with a total area of about 650m^2 are required for filtration of the NL thickener underflow. Cake washing is carried out on the filter to reduce the recirculating load of zinc. The NL residue filter cake is continuously fed into the hot acid leach (HAL).

The thickener overflow at about 100 mgpl suspended solids is mixed with a 'body feed' filter aid and is pumped to a steam jacketed precoat pressure filter of 140m^2 area. Filtrates from the polishing filter and the vacuum drum filter are combined to give a feed solution for the solvent extraction step having an average suspended solids of 5 mgpl.

.23 Hot Strong Acid Leaching

The hot acid leach differs greatly from a strong acid leach in a conventional dead roast plant. The volume of hot acid leach liquor is a critical parameter as it must not exceed the cooling requirements of the sulphation roaster and accordingly pulp densities in the leach are much higher than in conventional strong acid leaching practice (approximately 45 wt% solids against approximately 15 wt% solids). The recycle of the filtrate to the roaster enables the leach to be operated at very high acid and iron levels in solution. Leach volume is controlled by addition of spent electrolyte. Acid concentration is controlled by addition of concentrated sulphuric acid.

Leaching parameters for continuous leaching are; pH 0.1 to 0.2 (approximately 40 gpl free sulphuric acid), retention time 2.0 hours, and temperature 95°C. Such conditions result in greater than 98 per cent overall extraction of zinc and copper from calcines and yield a solution of about 100 gpl zinc, 100 gpl total iron, 30 gpl acid, 4 gpl copper, and 4 gpl arsenic. The leaching process is autothermal for continuous operation. The high temperature and ferric ion, chloride ion (from large volumes of concentrated acid required as reagent) and high acid concentrations necessitate that tanks must be acid brick, rubber, or plastic lined and all agitators constructed of Hastelloy C, titanium, alloy 20, or rubber-covered steel. The small volume in the hot acid leach and short retention time make the extra expense less prohibitive than similar modifications to a strong acid leach-Jarosite circuit.

.24 Hot Acid Leach Pulp Dewatering

The HAL pulp is filtered and washed in plate and frame filter presses of about $240m^2$ filtering area to give a filtrate for return to the roaster and a leach residue suitable for disposal, or brine leaching for lead and silver recovery.

.25 Copper Solvent Extraction and Electrowinning

Copper solvent extraction takes place in standard mixer settlers with L1X 64N extractant. The incoming neutral leach filtrate must be below 45°C to remain safely below the flash point of the diluent. The only deviation from conventional practice is that the high zinc solution must not be allowed to cool during extraction, thus the mixers will require supplementary heating. Holmes and Narver "low profile" mixer settlers should be suitable for this application.

The concentration of L1X 64N in the organic phase will be about 6.5% by volume. Plant practice for dilute liquors from heap leaching is three minutes retention in each mixer with three stages of extraction and two stage stripping. This should be considered as a 'worst case' since the combination of higher pH and higher temperature results in higher organic copper loadings and faster ion exchange kinetics. The raffinate solution from the SX is at approximately pH 1.2 and contains about 50 mgpl copper.

Spent copper electrolyte enters the SX strip at about 30 gpl copper and 160 gpl sulphuric acid. The pregnant electrolyte leaves the stripping section at about 35 gpl copper and 152 gpl acid.

The copper electrowinning tankhouse is of conventional design. About 22 tanks with 22 anodes and 21 cathodes each are required for a current density of $240 A/m^2$. Three starting sheet tanks are required unless starting sheets can be purchased on an economically favourable basis. If starting sheets are produced, a semi-automatic sheet preparation device will be required.

Entrainment and carry over of manganous ion containing NL solution to the copper electrowinning section may require remedial action. This manganous ion can be oxidized to permanganate at the anode in the copper electrowinning cell and the permanganate can, in turn, oxidize the organic reagents. The normal bleed requirements of the electrolyte may be sufficient to eliminate this problem, or, a small amount of iron (about 1 gpl) can be maintained in the electrolyte to reduce the permanganate to manganous ion.

.26 Iron Oxidation and Neutralization

Solution from copper solvent extraction at pH 1.2, ferrous iron concentration 0.7 gpl, and temperature 40°C is heated to 65°C in a heat exchanger and introduced to the iron oxidation tank(s). The oxidants employed are cellhouse sludge, air, and very small additions of fresh manganese dioxide. A retention time of 1.0 hours is sufficient to reduce the ferrous level to less than 15 mgpl. The solution is then neutralized slowly to between pH 4.0 to 4.5 with ground limestone to precipitate iron as ferric hydroxide. Control of basic zinc sulphate formation and limestone armouring is achieved by slow neutralization over 0.5 to 1.0 hours.

.27 Iron Sludge Thickening

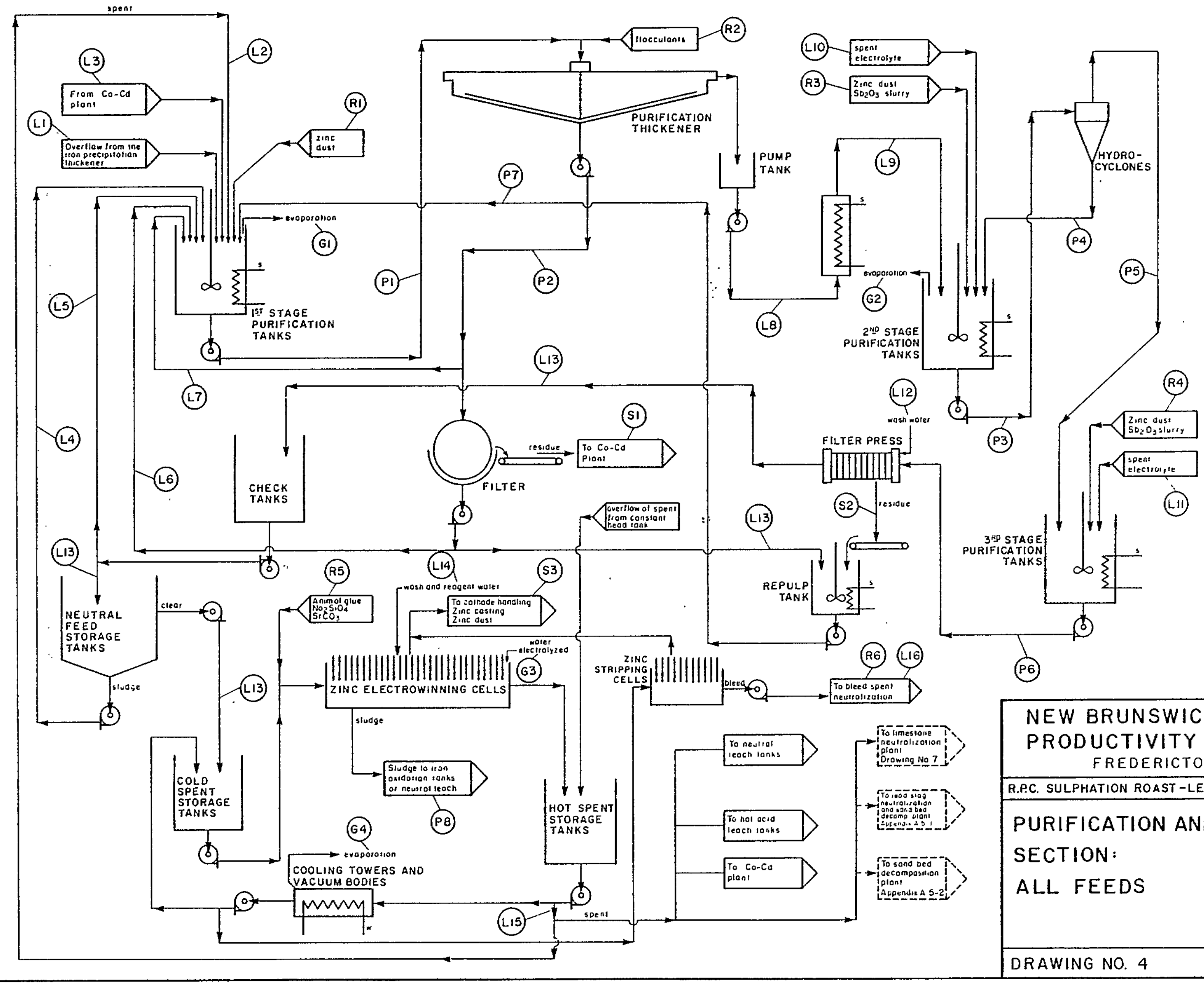
The solution from the iron oxidation and neutralization step is treated with 0.5 mgpl Percol 156 and fed to a 16m diameter clarifier. The overflow, representing about 87% of the feed solution, continues to copper-cadmium precipitation while the underflow is recycled to the neutral leach.

.3 Purification with Zinc Dust, Electrowinning and Casting Section

The flowsheet for this section is shown in Drawing No. 4. Complete mass balances are given in Tables A2-3 and A2-11 and thermal balances in Tables A2-8 and A2-16 for the bulk concentrate and tailings refloat plants respectively.

.31 Purification with Zinc Dust

Cadmium and residual copper are removed from the "impure feed" solution at 60-65°C by cementation on zinc dust (First Stage Purification). Conditions of temperature, zinc dust addition, pH, and retention time are identical to those commonly employed in conventional zinc plant practice. This cementate is separated from the solution in a thickener



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PURIFICATION AND ELECTROWINNING SECTION: ALL FEEDS

DRAWING NO. 4

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(gassing on the cementate is minimal due to the low copper content of the cementate).

Cobalt and nickel are removed by antimony activation on zinc dust at 95°C with a retention time of 3.0 hours (Second Stage Purification). The high cobalt tenor of the solution necessitates higher than normal zinc dust and antimony trioxide addition (8 gpl and 10 mgpl respectively). Provision is made for recycle of the coarser cementation product at the end of the second stage to the first tank of the second stage to reduce the addition of fresh zinc dust and antimony trioxide thereby reducing the recirculating zinc load from casting. The classification of second stage solids is accomplished in a hydrocyclone.

The Third Stage Purification tank is essentially a pump tank for the filter presses. Minor additions of zinc dust may be necessary to control redissolution of cadmium, cobalt, and antimony in the presses.

The press cake would be recycled to the first stage purification to reduce the requirement of fresh zinc dust and also to remove a fraction of the cobalt in this stage, simplifying the second stage operation. A common exit for all cementation products exists via the purification thickener to the cadmium plant. This procedure is presently employed by Canadian Electrolytic Zinc (patent applied for).

It must be emphasized that this plant design is based on liquid-atomized zinc dust of average classification 20 per cent +100 mesh, 45 per cent -100 to +200 mesh, and 35 per cent -200 mesh. Significant departure from this classification would alter design and zinc dust and antimony trioxide mass balances.

.32 Electrowinning and Casting

"Neutral feed" solution to the cellhouse contains 150 gpl zinc, <1 mgpl cadmium, <0.2 mgpl copper, cobalt, and nickel, and <0.02 mgpl antimony. Cellhouse design and operation is standard, with automatic stripping. Production of zinc cathode is 285 MTPD. Spent electrolyte bleed to stripper cells is 100 m^3PD to control magnesium and manganese in the spent electrolyte at 11.5 and 4.3 gpl respectively. Data from a cellhouse operating under similar conditions in a similar climate (Canadian Electrolytic Zinc) indicates that 353 TPD water would be evaporated from the hot spent electrolyte in the cooling towers and vacuum bodies in order to maintain an average electrolyte temperature in the cells of 36-40°C.

Zinc casting is conventional with expected dross production of 5.7 MTPD contained zinc. The high chloride content of the dross necessitates recycling to the roaster after milling as is done in most facilities. Approximately 20 MTPD of zinc dust for use in purification would be produced using off-spec metal and a compressed air gun liquid-atomization technique. The net production of zinc slab is 100,000 STPY (350 days).

.4 Cobalt-Cadmium Plant

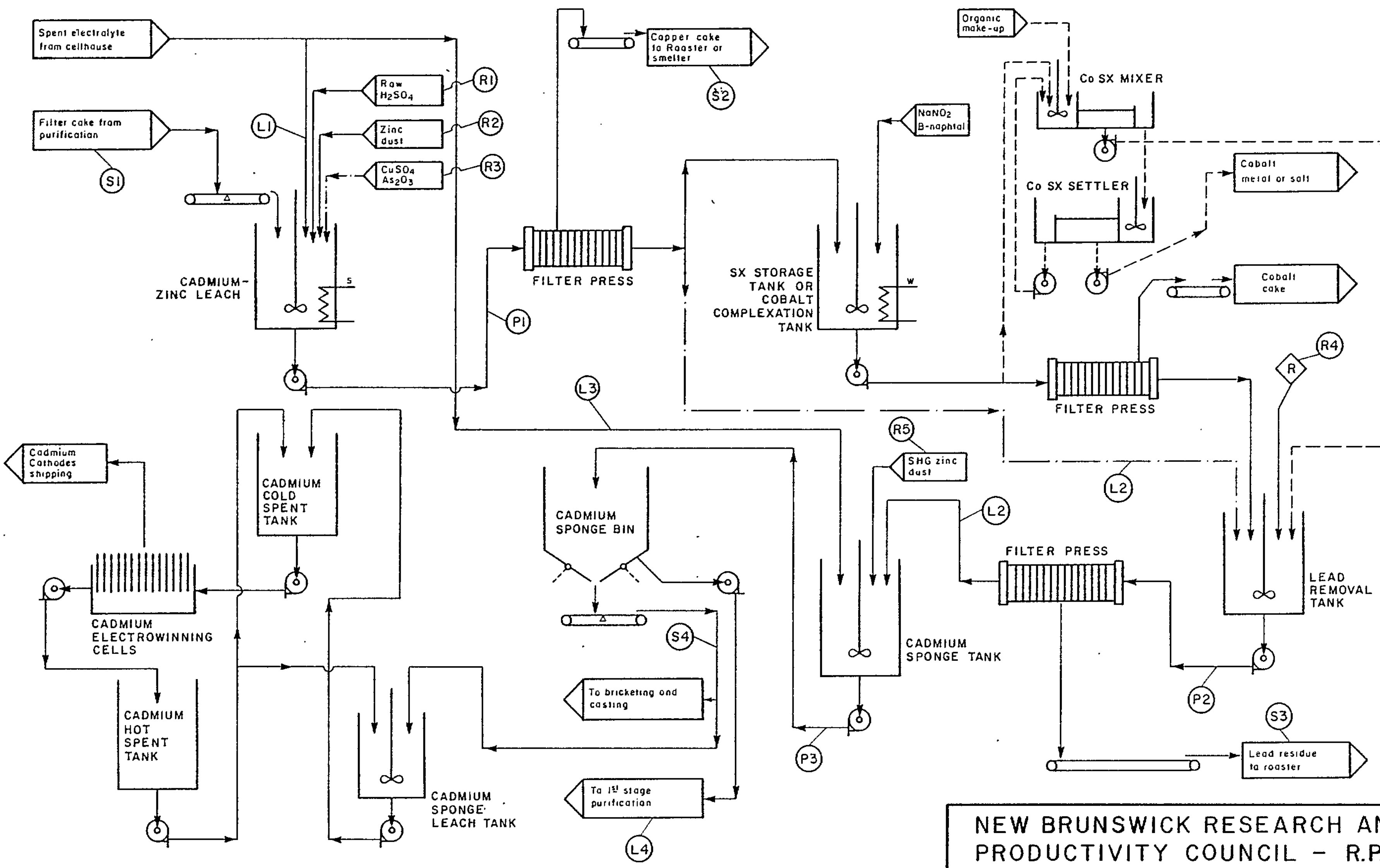
The flowsheet for the plant is shown in Drawing No. 5.

Complete mass balances are given in Tables A2-4 and A2-12 and thermal balances in Tables A2-8 and A2-16 for the bulk and tailings refloat plant respectively.

Three possible designs for a cobalt-cadmium plant are given in the flowsheet. Since no testing was performed on the solvent extraction or complexation models, the mass balances and economic evaluation were performed on a conventional model where the cobalt reports with the copper cake. The purification filter cake is leached in a batch mode in spent electrolyte and raw acid. The final pH is adjusted to 4.0. Zinc dust, copper sulphate and arsenic trioxide are added to precipitate cobalt to less than 50 mgpl and keep the copper level less than 1 mgpl. The high temperature of the leach solution prevents appreciable cadmium cementation. The resultant copper cake is removed in a filter press. Strontium or barium carbonate is added to the solution to precipitate lead which is removed by filtration. Special high grade zinc dust is added to the acidified filtrate to form cadmium sponge. This sponge cadmium after washing is either bricketted and cast or releached and cadmium cathode is electrowon.

.5 Lead-Silver Recovery Plant

The plant flowsheet shown in Drawing No. 6, and the complete mass balances given in Table A2-5 and A2-13, for the bulk and tailings refloat concentrates respectively are based on previous trials carried out by the St. Joseph Lead Company (pers. comm).



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CADMIUM-COBALT PLANT :
ALL FEEDS

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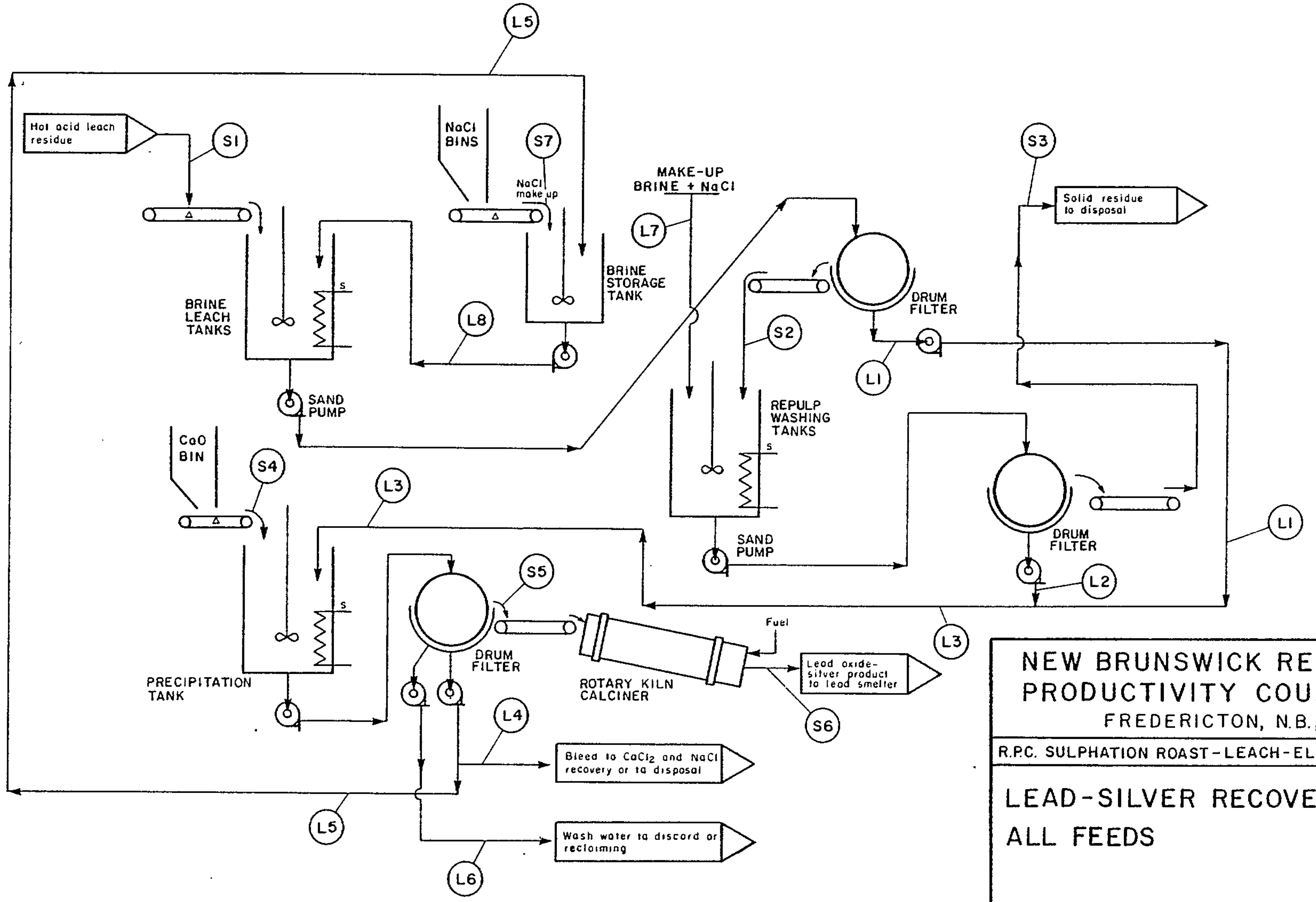
The residue from the hot acid leach dissolution of the ferrite and sphalerite residue, after filtering and washing is treated by a cold, saturated NaCl-$CaCl_2$ brine leach to dissolve the lead sulphate and silver. The leach pulp is filtered and then the lead and silver in the filtrate are precipitated with lime to form a lead-silver hydroxide product which is easily separated from the liquor by filtration. The washed cake is then decomposed by calcination in a rotary kiln at 300-400°C to obtain a lead oxide and silver product. Approximately 104 STPD of lead oxide product assying 83% lead and 66 oz/ST silver and 42 STPD of product assaying 79% lead and 89 oz/ST silver are recovered from the bulk concentrate and tailings refloat products respectively. Recoveries are 85% silver for both cases and 94% and 92% for lead from the two respective concentrates.

.6 Spent Electrolyte Neutralization

The flowsheet for the plant is shown in Drawing No. 7, mass balances are given in Table A2-6 and A2-14. The sulphate balance in the system is maintained by neutralization of excess spent electrolyte with limestone resulting in the precipitation of gypsum. The procedure and conditions chosen are similar to those employed in the fertilizer industry for the sulphuric acid addition to phosphate rock and resultant gypsum precipitation. The control of reaction temperature at 25-30°C, retention time at 1.0-2.0 hours, and pH at 4.5 to 5.2 is essential for producing precipitates of large crystal size, good filtrability, low moisture con-



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LEAD-SILVER RECOVERY PLANT: ALL FEEDS

DRAWING NO. 6

DATE: 14 APRIL, 1978

tents, and low entrained zinc levels.

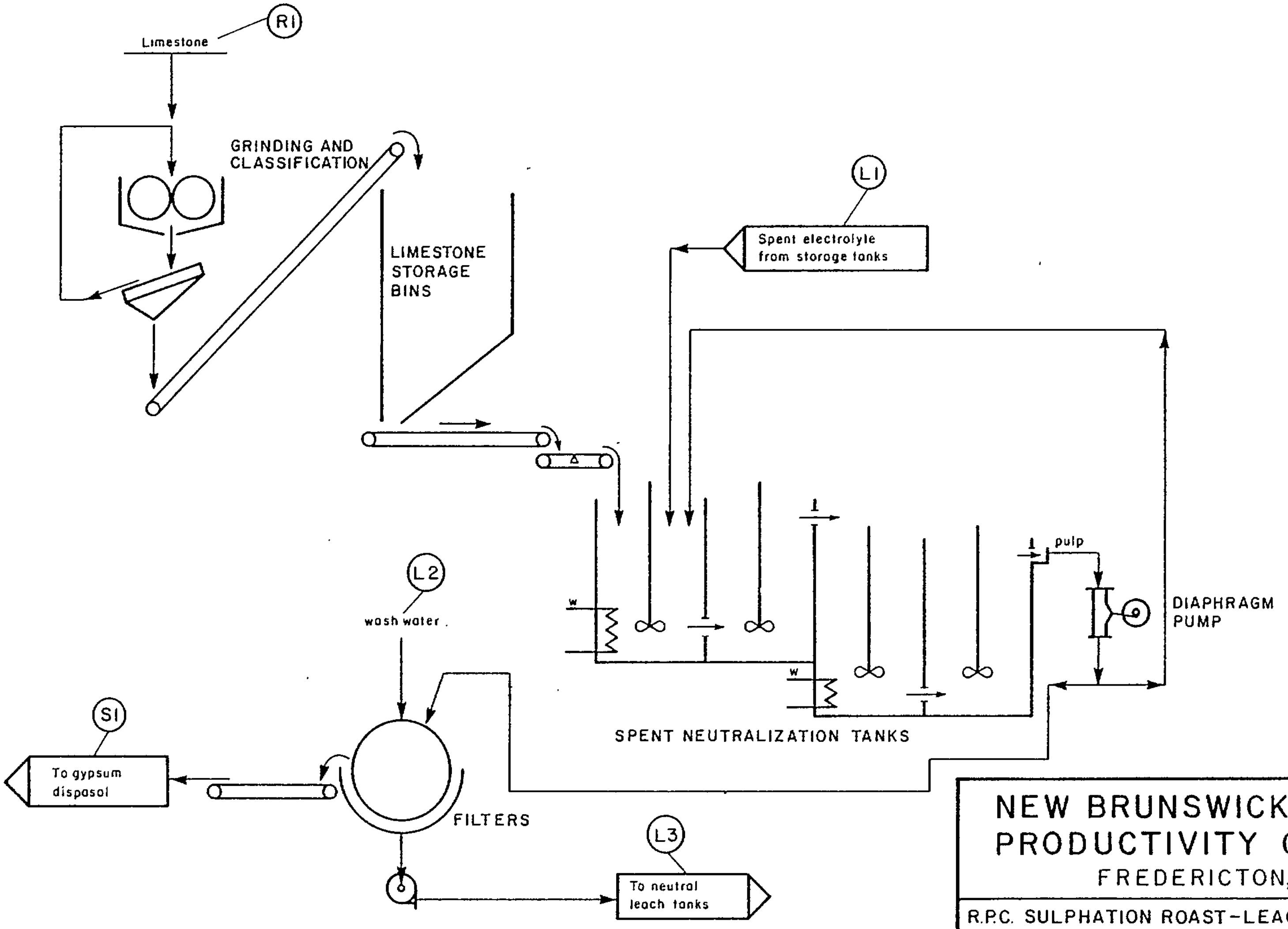
The limestone must meet the following specifications:- 100 mesh size, <0.30% magnesium, <50 ppm chloride, and <1.0% iron. The two New Brunswick limestones tested after milling were well within all specifications. The limestones differed in the $CaCO_3$ content (92 and 97 per cent) and the silica content (3 and 1 per cent).

Total limestone consumption in neutralization is 311 and 313 MTPD for the bulk and tailings refloat feed respectively. The gypsum residue is filtered using a vacuum filter. A Bird Prayon filter, model 18B of $44.6 m^2$ filtration area, is most suited for this application. The washed residue would contain 536 TPD gypsum, 3.2 TPD zinc, and 390 TPD free water. The gypsum residue pond (if required) would be used as a catch and compression basin, surface water returning to the neutral leach as makeup water.

Two other possible alternatives for treating the spent electrolyte, by neutralization with lead slag and by decomposition in an inert fluidized bed roaster, are discussed in Appendix 5.

5.2 INTEGRATED SULPHATION ROAST-DEAD ROAST PLANT

The integrated plant treats feeds of tailings refloat and zinc middlings products by sulphation roasting and high grade zinc concentrate by dead roasting. Combined capacity is 100,000 STPY of zinc slab composed of 43,720 STPY from the sulphation roast feed and 56,280 STPY from the dead roast feed.



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R.P.C. SULPHATION ROAST-LEACH-ELECTROWINNING PROCESS

SPENT ELECTROLYTE NEUTRALIZATION - LIMESTONE: ALL FEEDS

DRAWING NO. 7

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521 Characterization of the Feeds

The tailings refloat product was discussed in the previous section; it is assumed that a zinc middlings product has a similar composition. The assays and mineralogical composition of both the sulphation roaster and dead roaster feeds are given in Table 5.

522 General Flowsheet

The plant treats 390 MTPD of tailings refloat and middlings products and 269 MTPD of high grade zinc concentrate to produce 100,000 STPY of slab zinc, plus recovered secondary metals copper, lead, silver, cobalt and cadmium. The yearly production and revenue for all recovered metals and sulphuric acid are given in Section 6 (Tables 10 and 12).

The general flowsheet for the plant is shown in Drawing No. 8. As in the previous section, the plant, for purposes of mass balance calculations, is divided into six sections each of which has a detailed flowsheet. Only the roasting and leaching sections are changed substantially from the sulphation roast plants in the previous section. Detailed flowsheets are as follows:

1. Roasting (Drawing No. 9).
2. Leach (Drawing No. 10).
3. Purification and Electrowinning (Drawing No. 4).
4. Cobalt and Cadmium Plant (Drawing No. 5).
5. Lead-Silver Recovery Plant (Drawing No. 5).
6. Spent Electrolyte-Limestone Neutralization (Drawing No. 7).

TABLE 5

CHEMICAL AND MINERALOGICAL COMPOSITION OF FEED TO INTEGRATED SULPHATION ROAST-DEAD ROAST PLANT

Sulphation Roast		Dead Roast	
Tailings Refloat or Zinc Middlings		Zinc Concentrate	
Chemical Composition (wt %)	Mineralogical Composition (wt %)	Chemical Composition (wt %)	Mineralogical Composition (wt %)
zinc = 30.0	sphalerite ((Zn,Fe)S) = 50.7	zinc = 56.0	sphalerite ((Zn,Fe)S) = 94.6
lead = 3.6	galena (PbS) = 4.2	lead = 1.20	galena (PbS) = 1.4
copper = 0.6	chalcopyrite ($CuFeS_2$) = 1.7	copper = 0.19	chalcopyrite ($CuFeS_2$) = 0.6
iron = 22.0	pyrite (FeS_2) = 37.6	silver = 1.83 oz/S.T.	gangue = <1.0
sulphur = 39.0	gangue = <3.0		
silver = 4.4 oz/S.T.			

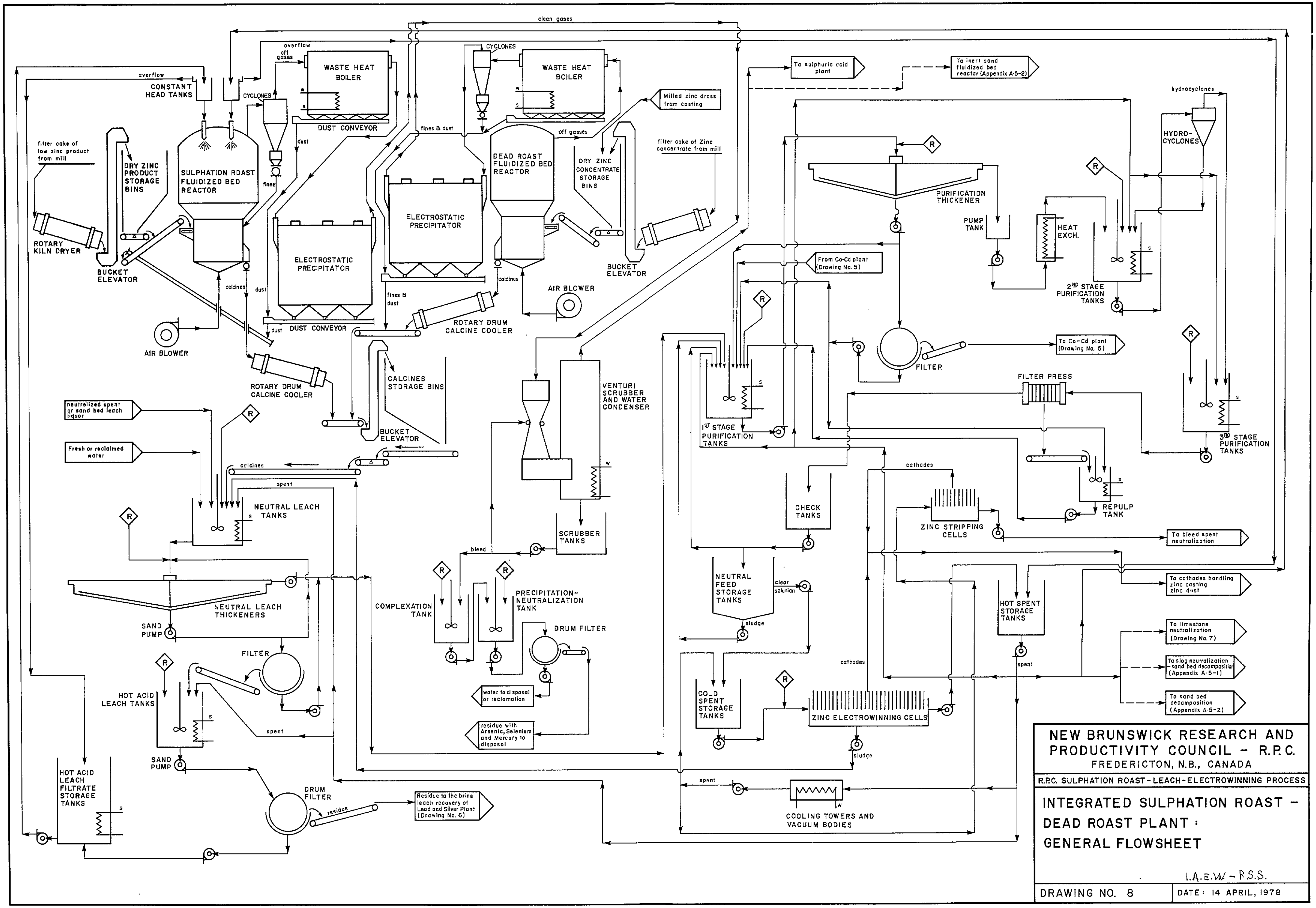
.1 Roasting Section

The flowsheet of the roasting section is given in Drawing No. 9. Complete mass balances for the integrated plant are given in Tables A2-17, A2-23 and A2-25. The thermal balance is given in Table A2-24.

Flotation concentrates, tailings refloat and zinc middlings, containing approximately 8% water, are dried to 2.5% moisture in a rotary drum dryer. The dried product is crushed to 100% -100 mesh prior to roasting. Storage for dry concentrates is sufficient for 48 hours of operation.

High grade zinc concentrates, as filter cake containing 8% water are dried to 6% moisture, also in a drum dryer. Storage is provided for 48 hours of roaster operation.

Two Lurgi-type fluid bed roasters are used. Nominal capacity of the sulphation roaster is 390 TPD and 270 TPD for the dead roaster. In this instance each roaster package is self-contained, each with its own waste heat boiler and electrostatic precipitator. This is necessary as only the dust from the sulphation roaster gas cleaning system, because of its relatively high soluble iron content, need be recycled to the sulphation roaster. The sulphation roaster is fed with a slinger belt feeder, and the calcines discharge through a rotary star valve to a drum rotary cooler and stored in 60 day storage bins. The strong acid leach filtrate producted from the dissolution of both ferrite residues from the sulphation roast and dead roast operations is recirculated



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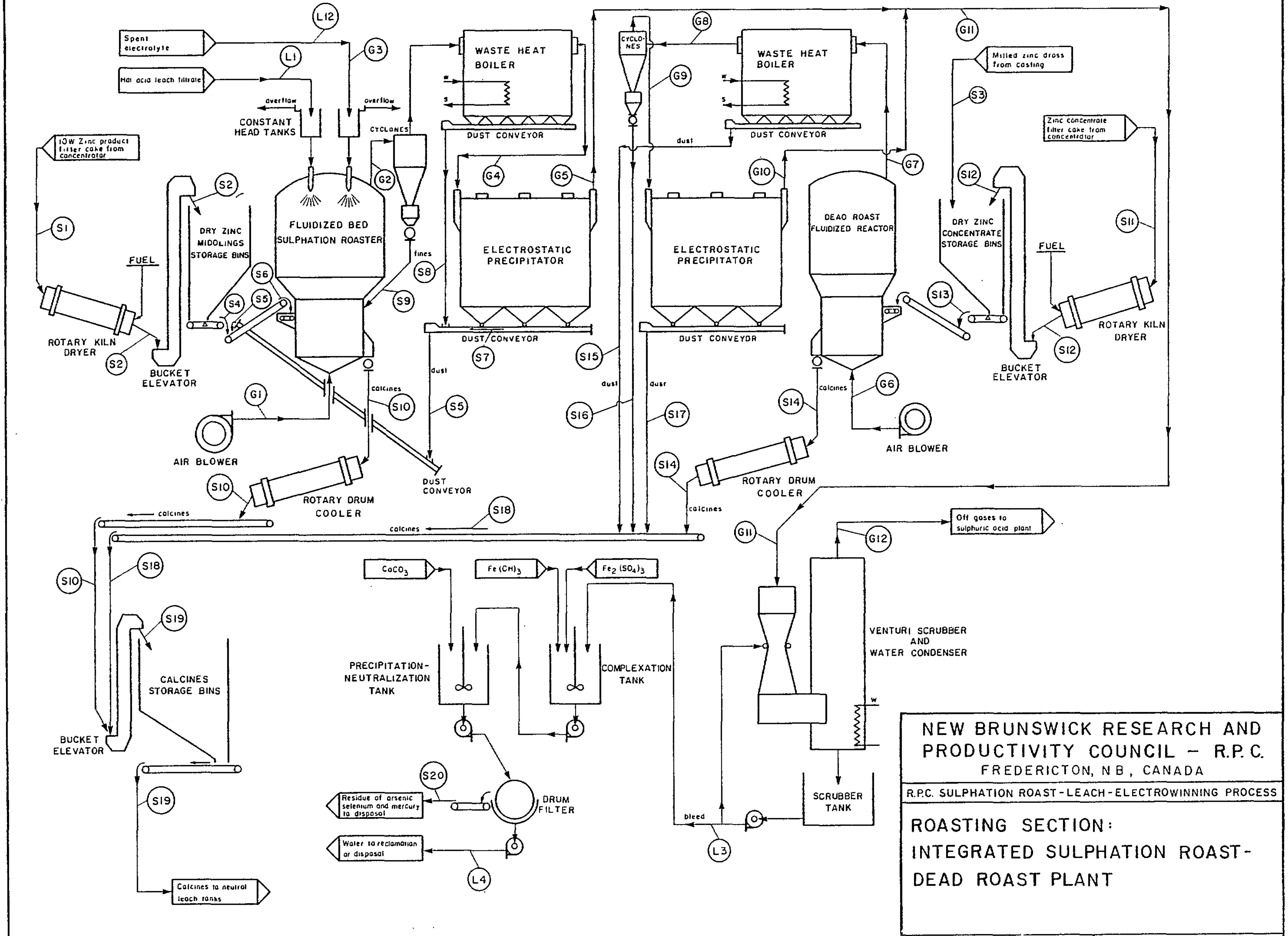
INTEGRATED SULPHATION ROAST -
DEAD ROAST PLANT :
GENERAL FLOWSHEET

I.A.E.W - R.S.S.

DRAWING NO. 8
DATE: 14 APRIL, 1978



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R.P.C. SULPHATION ROAST-LEACH-ELECTROWINNING PROCESS

ROASTING SECTION: INTEGRATED SULPHATION ROAST-DEAD ROAST PLANT

DRAWING NO. 9

DATE · 14 APRIL, 1978

into the sulphation roaster by using air- operated spray guns. Additional guns are provided to spray spent electrolyte, if the thermal capacity of the roaster is such that the available hot acid leach filtrate is insufficient to cool the reactor. Hydrostatic pressure to the guns is kept constant by using constant head tanks above the reactor.

Thermal balance in the sulphation roaster is achieved only by the use of the hot acid leach filtrate and spent electrolyte, if the latter is required, while the amount of heat removed from the hot gases in the freeboard section of reactor is adjusted by the atomization and number of guns in operation. Overall thermal balance of the roasting section is given in Table A2-24. The balance was performed for the minimum amount of heat to be recovered from the fluidized bed with no heat recovered in the freeboard. Latent heat from the off-gases is recovered in the waste heat boilers.

The solid-gas separation system for the sulphation roaster package is basically the same as described previously for the bulk concentrate and tailings refloat sulphation roast plant. Mass balances for the gas-solid separation are given in Table A2-25.

The dead roasting plant is entirely conventional with a nominal capacity of 270 MTPD. The reactor is fed by means of a slinger belt-type feeder, while calcines discharge through a star rotary valve to a rotary drum cooler. Calcines are stored in 60-day storage bins.

The reactor is cooled in the reaction zone by water coils, while the hot gases (930°C) enter a waste heat boiler with a radiation and a convection zone. The waste heat boiler also recovers about 55% of the dust carryover. Gases leaving the waste heat boiler at 350°C are further cleaned in a dry cyclone, which collects 24% of the elutriated calcines and finally in an electrostatic precipitator, collecting 6% of the total elutriated calcines. Overall solid-gas collection efficiency is 99.5%. The thermal balance in the roaster is given in Table A2-24, while the mass balance for the gas-solid separation system is given in Table A2-25.

Clean gases from both roaster systems are further cleaned in a venturi scrubber, as described previously, prior to going to the sulphuric acid plant.

The reactor parameters and operation conditions for both the sulphation roasting and dead roasting plants are given in Table 6.

TABLE 6
REACTOR PARAMETERS AND OPERATING CONDITIONS FOR THE ROASTING SECTION OF THE INTEGRATED SR-DR PLANT

	Sulphation Roast	Dead Roast
Nominal capacity (MTPD)	390	270
Reactor bed diameter (m)	10.9	6.4
Fluidized bed level (m)	1.5	1.5
Feed rate (MT/m^2-day)	4.1	8.4
Roasting temperature (°C)	685	930
Air factor (-)	1.8	1.3
Solid retention time (hrs)	5.5	4.5
Air velocity (at temp. T)(cm/sec)	75.0	185.0

.2 Leaching Section

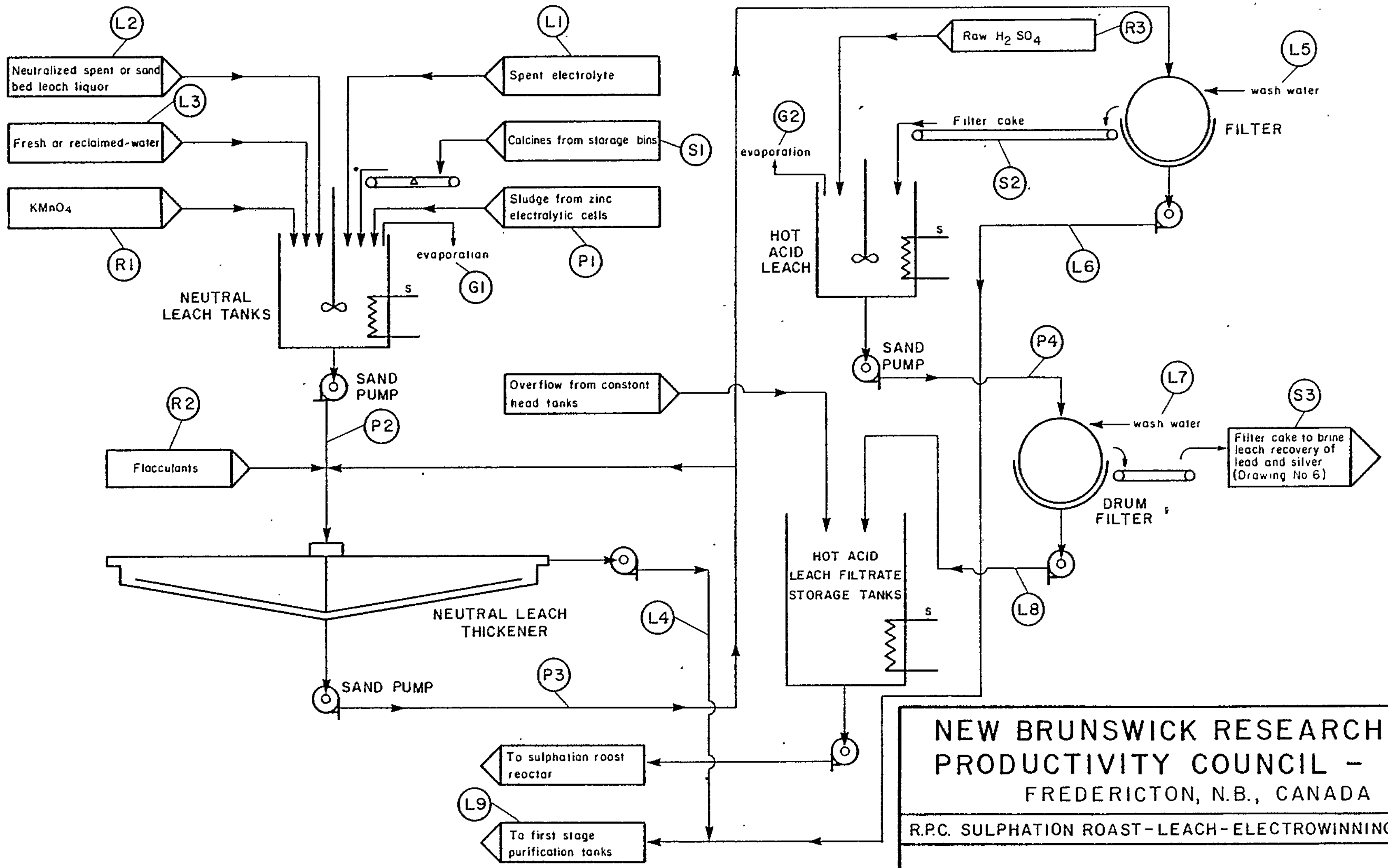
The flowsheet for the section is shown in Drawing No. 10, while mass and thermal balances are given in Tables A2-18 and A2-24.

.21 Dilute Acid "Neutral" Leaching

The sulphation roast and dead roast calcines are leached together under the same conditions as described previously for the sulphation roast calcine alone. The ambient leach temperature has been calculated to be 70°C assuming no conductive or convective heat losses. Actual plant operation would be designed to minimize these losses as low copper content excludes consideration of copper solvent extraction for this plant. It is expected that a leach temperature of 55-65°C could be maintained without external heating. Oxidant is added directly to the neutral leach in order to keep dissolved iron levels below 15 mgpl in the solution to the thickener. It is planned to use air (from air lifts and sparging) and potassium permanganate as oxidants. The leach solution typically contains 148 gpl zinc, 1.1 gpl copper, and <15 mgpl iron.

.22 Neutral Leach Pulp Dewatering

The NL residue is thickened in a 36m diameter thickener (based on unflocculated pulp tests). The thickener underflow exits at about 30% solids and is pumped to vacuum drum filters with a total filtration area of 585m^2 for further dewatering. Thickener overflow and filtrate are combined and are pumped to purification.



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R.P.C. SULPHATION ROAST-LEACH-ELECTROWINNING PROCESS

LEACHING SECTION:
INTEGRATED SULPHATION ROAST-DEAD ROAST PLANT

DRAWING NO. 10

DATE: 14 APRIL, 1978

.23 Hot Strong Acid Leaching

Since ferrite and sphalerite from both calcines are leached together in this operation and the resultant solution must be recycled entirely to the sulphation roaster, control of volume to insure autothermal roaster operation necessitates a leach liquor of higher ionic strength than for the sulphation roast case alone. Accordingly, fresh water is used as makeup and volume control instead of spent electrolyte in order to keep the zinc and sulphate level in the leach liquor below the saturation point. Acid concentration is maintained, as before, with concentrated sulphuric acid. Leaching parameters and extractions remain unchanged but the resultant leach liquor composition will be about 175 gpl iron, 110 gpl zinc, 4 gpl copper, and 8 gpl arsenic.

.24 Hot Acid Leach Pulp Dewatering

The HAL pulp is filtered and washed in plate and frame presses with an effective area of about $73m^2$. The filtrate is returned to the roaster while the residue, which is free of soluble sulphates, is suitable for brine leaching for lead recovery, or for disposal.

.3 Purification with Zinc Dust, Electrowinning and Casting

The flowsheet for the section is given in Drawing No. 4. Mass and thermal balances are given in Tables A2-19 and A2-24.

The proposed flowsheet and operational parameters are identical to the sulphation roast case. Differences occur however in the copper, cadmium and cobalt mass balances. The copper is removed from solution by cementation on zinc dust, either together with the cadmium as shown in the flowsheet, or as a separate product (>80% copper, <2% cadmium, 5-20% zinc) to be cemented by stoichiometric zinc dust addition prior to cadmium cementation. Cobalt cementation will not be as difficult for the integrated plant, however, due to the lower cobalt concentration in the solution (40 versus 80 mgpl).

A spent electrolyte bleed of 70 m^3PD is necessary to control the magnesium and manganese in the spent electrolyte at 8.1 and 8.8 gpl respectively.

.4 Cobalt-Cadmium Plant

The proposed flowsheet (Drawing No. 5) and operational parameters are identical to the sulphation roast case. Mass and thermal balances are in Tables A2-20 and A2-24. The tonnage of copper cake is much larger than in the sulphation roast plant as all the copper in the leach solution reports in this product for sale to a copper smelter.

.5 Lead-Silver Recovery Plant

The plant is identical to the one discussed for the bulk concentrate or tailings refloat product, and is depicted in Drawing No. 6. The mass balance for the integrated plant is given in Table A2-21.

Approximately 25 STPD of lead oxide product, 71% lead and 81 oz/ST silver, is produced with recoveries of 93% for lead and 85% for silver.

.6 Spent Electrolyte Neutralization

Flowsheet of the plant is given in Drawing No. 7, while mass balances are given in Table A2-22.

The increased consumption of spent electrolyte in the neutral leach due to the zinc oxide present in the calcine from the dead roast, significantly reduces the amount of spent electrolyte to be neutralized. Limestone consumption is reduced to 166 MTPD resulting in a residue after filtration of 287 TPD gypsum, 1.7 TPD zinc, and 209 TPD free water. A Bird Prayon filter, model 12C of 250 sq.ft. filtration area would be sufficient for this application.

5.3 ENVIRONMENTAL CONSIDERATIONS

531 Gaseous Process Effluent

Sulphur dioxide off-gas from the roasting operations is converted into sulphuric acid in a high efficiency double catalysis sulphuric acid plant. Dust emissions are controlled by baghouse dust control systems at various material handling transfer points, from pneumatic conveying systems, from casting, melting, zinc dust and dross grinding operations and from rotary dryer exhausts.

532 Solid Process Effluent

The zinc plant solid process residues are:

1. Gypsum filter cake from spent electolyte neutralization.
2. Hematite filter cake from brine leach plant.

The gypsum residue contains approximately 40 percent by weight free water and less than 0.6 per cent zinc as sulphate. Although, the gypsum itself is a stable, environmentally acceptable residue, the inclusion of zinc may necessitate ponding of the residue (cost included under services in the economic analyses). The gypsum pond would probably be designed as a storm water pond to provide deionized process makeup water. Some of the zinc in the gypsum residue would therefore return to the process.

The hematite residue contains only gypsum at 5-10%, NaCl and $CaCl_2$ plus trace quantities of zinc, copper and lead as contaminants. It is expected that moisture contents of less than 15% will be achieved for this residue. Disposal of this stable dry filter cake on a dump should provide no difficulties. A cost for ponding and pond effluent treatment has been included in the economic analyses in case unforeseen problems are evidenced in the brine leach test program.

533 Liquid Process Effluent

The zinc plant liquid process effluents are:

1. Weak acid from acid plant
2. Spent zinc electrolyte bleed from cellhouse (electrolytically stripped to 10 gpl zinc)

3. Blowdowns from waste heat and direct fired boilers and service cooling towers.

4. Feed water demineralizer regeneration effluent.

It is proposed to neutralize the effluents and chemically treat them to precipitate heavy metal contaminants. The effluent water would be separated from the precipitated solids by clarification and filtration, producing a waste solid filter cake. The filtrate may be suitable for use as makeup water after ion exchange for halide removal. Costs for these operations also included under services in the economic analyses.

6.0 ECONOMIC EVALUATION

6.1 CAPITAL COSTS

Capital costs for sulphation roast plants, which would produce 100,000 STPY slab zinc, based on two feeds, 844 MTPD bulk concentrate and 900 MTPD tailings refloat, are summarized in Table 7. Also shown are costs for an integrated sulphation roast-dead roast plant, with the same zinc output, using 390 MTPD combined tailings refloat and zinc middlings products plus 269 MTPD of zinc concentrate. The capital costs for a conventional dead roast zinc plant, which would treat 52% zinc concentrates, are also shown for comparative purposes.

Information on costs for individual packages within a conventional dead roast plant, i.e. roaster section, acid plant, leach-purification-cadmium recovery section and electrolysis section were readily available from a number of sources including equipment manufacturers and industry, and the overall cost estimate for this plant is considered to be accurate to -10% to +20%.

With the exception of the lead/silver and copper recovery section, the components of the sulphation roast plant are similar to those of a conventional dead roast plant. Hence, costing of the sulphation roast plants were made relative to dead roast plant costs. Additional cost details are given in Appendix A7 and a summary of section costs is given in the following sub-sections.

TABLE 7

CAPITAL COSTS
MID 1978 DOLLARS ($,000,000)

	DR+	SR (Bulk)	SR (Tailings Refloat)	SR/DR++
ZINC PLANT				
1. Roasting Section (incl. feed prep., conc. and calcine handling)	20.76	29.18	30.79	24.89
2. Acid Plant (incl. gas cleaning and acid storage)	11.52	17.37	18.76	15.57
3. Leach-Purification Cadmium Recovery	21.25	17.38	17.38	16.42
4. Electrolysis, Melting and Casting	32.78	32.78	32.78	32.78
5. Neutralization	-	3.40	3.40	2.25
6. Auxiliary Facilities	3.75	3.75	3.75	3.75
7. Services	16.42	18.42	18.57	17.42
Direct Zinc Plant Cost	106.48	122.28	125.42	113.08
COPPER SX-ELECTROWIN. PLANT (Turnkey Cost)	-	3.27	3.27	-
LEAD/SILVER RECOVERY PLANT (Turnkey Cost)	-	5.31	2.88	1.88

TABLE 7 (Cont'd)

CAPITAL COST
MID 1978 DOLLARS ($,000,000)
(cont'd)

	DR^{+}	SR (Bulk)	SR (Tailings Refloat)	SR/DR^{++}
INDIRECT COSTS				
1. Engineering (12% Direct Zinc Plant Cost)	12.78	14.67	15.05	13.57
2. Contingencies (10% Direct Zinc Plant Cost)	10.65	12.23	12.54	11.31
*3. Miscellaneous (incl. Start-up training, etc.)	5.00	5.00	5.00	5.00
*4. Licences and Royalties	1.20	1.20	1.20	1.20
5. Pilot Plant and Feasibility Study	-	4.00	4.00	4.00
WORKING CAPITAL	7.45	8.31	8.07	7.64
TOTAL CAPITAL	143.56	176.27	177.43	157.68

* Non depreciable items

SR = Sulphation Roast

$^{+}$ DR = Conventional Dead Roast (52% Zn Conc.)

$^{++}$ DR = Conventional Dead Roast (56% Zn Conc.)

611 Roasting Section

The principal items in the roasting section are the Lurgi roaster package supplied complete with cyclones, electrostatic precipitator, feeding system, etc. (see Table A7-1) and the waste heat boilers. Each plant is equipped with two roasters and two waste heat boilers.

Figure 5 shows the relationship between roaster capacity and reactor hearth area for sulphation roasters for design data derived in the mini-pilot trials. Also shown, for comparative purposes, is the similar relationship for zinc dead roasters from information supplied by Lurgi. It can be seen from these curves that in the range of interest, 390 to 450 MTPD, on the basis of our trial data the hearth area of the required sulphation roaster would be approximately twice that of a dead roaster of similar feed capacity. Information relating the installed cost of a conventional roaster to hearth area is given in Appendix 7 (Figure A7-1). An allowance (+15%) in capital cost was made for additional features associated with a sulphation reactor.

It must be stressed that the data derived from the mini-pilot trials were used for reactor scale-up purposes. This is perhaps too conservative in that it might be expected that a full scale plant would operate at lower retention time and lower air ratio than experienced in the small scale reactor. A 10% reduction in just these two variables would result in a shift of the capacity versus hearth area curve to the projected position shown on Figure 5. With these

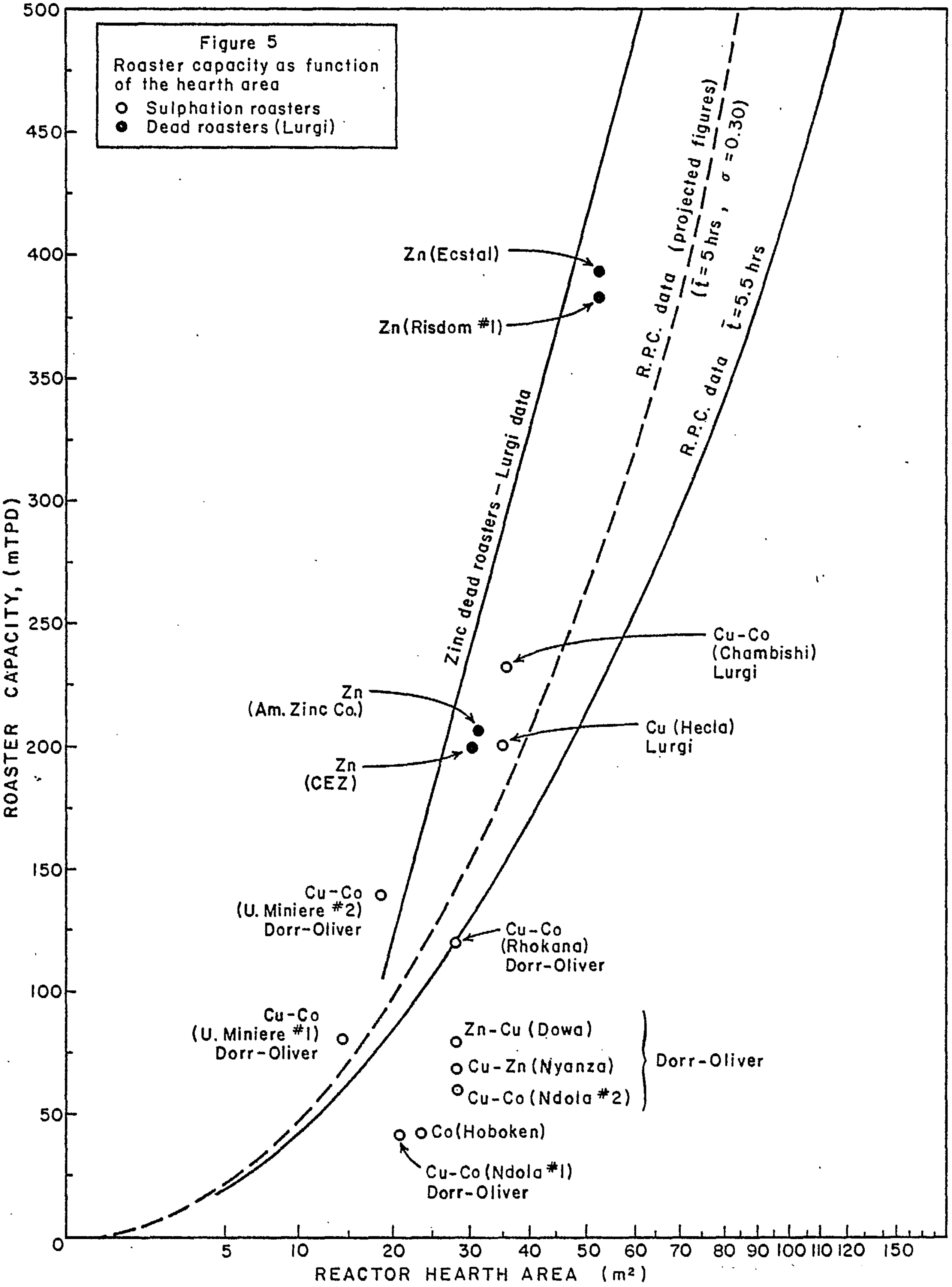

Figure 5
Roaster capacity as function of the hearth area

design criteria, the roaster hearth area would be decreased by 26% and the capital cost of the roasters by approximately 28%.

The waste heat boilers employed are Foster-Wheeler Ltd. forced circulation units with a gas-tight water-cooled "Monowall" casing. Although expensive ($1.4 million per uninstalled unit with 31,400 tph steam rate), the amortized cost is returned in recoverable process steam for use in the leaching section.

There is provision in the costing for drying of the sulphation roaster feed from 8% moisture to 2.5%. The dry concentrate is then crushed to minus 100 mesh. This pretreatment stage was inserted as a conservative precaution to avoid possible agglomerates coarser than 100 mesh, which could enhance ferrite formation. However, in the pilot trials, the minus 100 mesh dry feed disintegrates to finer particles in the calcine, and hence little agglomeration should be experienced in a commercial reactor by injecting feed with 8% moisture directly into the roaster. Although there is sufficient thermal capacity within the roaster to handle this moisture, some lowering of the off-gas temperature may occur. A lower off-gas temperature would result in less steam recovery from the waste heat boiler. Elimination of the pretreatment stage would lower the capital cost of the roasting section by approximately 0.75 million dollars for the two sulphation plants and by approximately 0.5 million for the integrated plant.

If the above changes from the scale-up provisions can be incorporated in a full scale plant the capital cost of the roasting sections shown in Table 7 can be reduced by more than 20% for the two sulphation plants and by more than 10% in the case of the integrated plant.

612 Acid Plant

Costs for this section are based on a double catalysis process. The higher capital costs for the acid plants associated with sulphation roasters reflects the higher acid production and lower SO_2 gas concentrations, 4.42%, 5.18% and 7.41%, for the sulphation roast of the bulk concentrate, tailings refloat product and integrated plants, respectively, compared with the dead roast plant which has an SO_2 off-gas concentration exceeding 9%. These SO_2 concentrations, which are used to calculate the capital costs of the respective sulphation roast plants shown in Table 7, are based on the roasting variables established in the mini-pilot trials. As the strength of the SO_2 off-gas is a direct function of the air ratio employed in the roaster, the air factor is often controlled to purposely low levels to increase the concentration of SO_2 in the off-gas. This facilitates greater efficiency and reduced capital and operating costs in the acid plant. As stated earlier, it is most likely that a commercial zinc sulphation roaster would be run at a considerably lower air ratio (Q) than that employed in the mini-pilot trials. A reduction of Q from 1.8 to 1.6, for example, would decrease the capital cost of the acid plants associated with the sulphation roasters by 7 to 9%. As the

SO_2 gas from the integrated plant is of relatively high concentration, the cost reduction gained by a reduced air factor is less significant.

613 Leach-Purification-Cadmium Recovery Section

The overall costs of this section for each of the sulphation plants and integrated plants is significantly reduced in comparison to a conventional dead roast zinc plant. This is principally a result of the elimination of the Jarosite circuit (Table A7-2). However, the neutral leach and settling stage is considerably more capital intensive for each of the plants relative to a dead roast plant. This results from increased capital costs for filters. The strong acid leach stage is simpler than for a dead roast plant, particularly for the integrated plant, thus, less capital is required.

The cadmium recovery plant is substantially the same as in a conventional zinc plant and the cost is of the same magnitude.

614 Zinc Electrolysis, Melting and Casting

Capital costs for this section are the same as for a conventional electrolytic zinc plant. Included are automatic cathode stripping machines, solution cooling and storage facilities, electrode preparation equipment, and melting and casting facilities.

615 Neutralization

Scale-up design and costing information was obtained from the fertilizer industry and major equipment suppliers.

616 Auxiliary Facilities

Capital costs shown in Table 7 cover building and equipment costs for offices, warehouses, laboratory, garage, maintenance shop, change house, and storage facilities for reagent and operating supplies.

617 Services

This section includes site preparation, water supply system including yard plumbing and in-plant piping, acid plant cooling system, sewage treatment facilities, effluent treatment and ponding, power supply, steam plant, mobile equipment, and fuel storage facilities.

618 Copper Solvent Extraction-Electrowinning Plant

Plant costs as shown in Table 7 are for a turnkey Holmes and Narver package plant (working capital excluded) employing low-profile mixer settlers. The pH of leach solutions is higher than in solutions from heap leaching, and the temperature is also higher. Higher copper loadings in the organic phase are possible at the higher pH and the kinetics of extraction also appear to be faster. The latter could lower the capital cost of the solvent extraction section by allowing shorter retention times and/or two extraction mixer settlers rather than the usual three, since the section cost was based on heap leach solution performance.

Copper SX-electrowinning was incorporated in the two sulphation plants but not in the integrated plants where the copper is less and the capital cost might not be warranted. Although a conventional copper cementation process has been incorporated in this plant based on capital considerations, a careful assessment of the overall cost/benefit of the two copper recovery approaches for this plant is recommended. For example, without the solvent extraction stage, requirements for high cost oxidants to control iron in solution are increased (see Table A7-5). This, and other possible operating advantages, might offset the apparent disadvantages from the stand point of capital.

619 Lead-Silver Recovery Plant

The plant costs shown in Table 7 were determined from estimates given by equipment manufacturers and published data for the component parts of the plant shown in Drawing No. 6. This brine leach plant would treat 439 STPD of dry HAL leach residue assaying approximately 21% lead and 18.63 oz/ST silver and recover 104.27 STPD of lead oxide product assaying 82.87% lead and 66.31 oz/ST silver. The estimate of $5.31 million is similar to the updated estimate by Battelle (2.19 million; 1957 dollars) for a similar product (420 TPD; 21.6% Pb) based on the St. Joseph Lead Company brine leach pilot test work.

620 Working Capital

Working capital consists of accounts receivable such as stockpiled concentrate, metals within the plant, in calcine, leaching solution, etc., and product inventory and product accounts receivable.

6.2 SUMMARY OF CAPITAL COSTS

The similarity of the sulphation roast plants to a conventional dead roast plant greatly simplifies cost estimates for such plants relative to a dead roast electrolytic zinc plant. As such there is more confidence in the accuracy of the capital cost estimates than would otherwise be possible at this stage in the development of the process. The capital costs presented in Table 7 are most probably accurate within ±25% in mid 1978, Canadian dollars. No provision is made for escalation or interest charges during the construction period.

6.3 DIRECT OPERATING COSTS

A summary of operating costs for all plants is shown in Table 8, additional details are given in Appendix A7.

631 Labour

The total number of jobs is estimated at 449 for the two sulphation roast plants and 438 jobs for the integrated sulphation roast-dead roast plant, compared with 398 jobs for a conventional dead roast zinc plant. The overall

TABLE 7

SUMMARY OF OPERATING COSTS
($,000/YR)

		DR	SR (Bulk)	SR (Tailings Refloat)	SR/DR
1.	Labour	7,496	8,457	8,457	8,249
2.	Power	7,274	7,477	7,472	7,325
3.	Reagents	1,313	3,763	3,488	2,659
4.	Operating & Maintenance Supplies				
	Zinc Plant	2,943	3,003	3,003	2,973
	Copper Plant	-	108	108	-
	Lead/Silver Plant	-	175	95	62
5.	Maintenance Capital Expenditures	800	1,000	1,000	900
6.	Fuel (Bunker 'C' and Propane)	1,067	1,263	1,152	910
7.	Taxes and Insurance*	863	1,074	1,084	956
	DIRECT OPERATING COSTS	21,756	26,320	25,859	24,034

* Municipal, Provincial & Business Tax = $3.7623/100 (buildings and land ≃ 15% of depreciable capital)

average cost including basic salaries for operating, maintenance, staff and administration personnel, plus fringe benefits is $18,834 per job in mid 1978 dollars. The labour distribution is shown in Table A7-3.

632 Power

The electrical energy requirements for various plant operations are given in Table A7-4. Total requirements on a kwhr per metric ton of zinc basis are 4417, and 4414 for the bulk and tailings refloat products and 4327 for the integrated plant. These compare with 4297 kwhr/MT of zinc for a conventional dead roast plant. The mid 1978 cost for Class IV power intensive industries in New Brunswick is $0.01866/kwhr.

633 Reagents

The total cost of reagents is shown in Table 8 with details given in Table A7-5. Major additional costs for the sulphation plants relative to those of a conventional zinc plant are limestone for spent electrolyte neutralization and lime, sodium chloride and calcium chloride, for the brine leach recovery of lead and silver. These higher costs are offset somewhat by the elimination of the Jarosite precipitation agent and in the case of the two sulphation roast plants almost total elimination of the cost of iron oxidants (potassium permanganate and manganese dioxide). Potassium permanganate is still required for the integrated plant unless solvent extraction is used for copper recovery.

634 Operating and Maintenance Supplies

This item includes all consumable supplies for complete plant operation.

635 Maintenance Capital Expenditures

This is an allowance for "non-consumable" materials required in maintenance.

636 Fuel

The cost of fuel shown in Table 7 includes the cost for Bunker "C" (No. 6 fuel oil), propane and gasoline. Bunker "C" is used in direct fired boilers and in the roaster plant where it may be replaced by No. 2 fuel oil. Propane is used for process heating, principally in the casting plant and for the start up of the direct fired steam boilers and roasters, and for cellhouse heating.

Fuel requirements for a 100,000 STPY conventional zinc plant are approximately 1.09 million gallons of Bunker C, 820,000 gals of propane, and 200,000 gallons of gasoline per year. Bunker C requirements are approximately 30% higher for the bulk concentrate sulphation plant and 8% higher for the tailings refloat plant. Fuel oil requirements are 42% lower for the integrated plant relative to a dead roast plant. Propane and gasoline requirements are an estimated 9% more for all three proposed plants relative to a dead roast plant. Details on fuel requirements are given in Table A7-6.

The waste heat boilers recover 75%, 80% and 85% of the process heat requirements for the total hydrometallurgy section (leach-purification-electrowinning) for the two sulphation plants and the integrated plants respectively. These compare with 75% for a convention dead roast plant.

637 Taxes and Insurance

Municipal, provincial and business tax in New Brunswick is approximately \$3.76 per \$100 of assessment on buildings and land. Insurance is approximately 0.1% of depreciable capital.

6.4 DEPRECIATION

Depreciation for all plants is based on a 15-year straight line write-off of depreciable capital.

6.5 PAYMENT FOR CONCENTRATES

Guidelines for payment of concentrates were established by CANMET, and were stated as; "the concentrate (whatever its composition) shall be costed at 30% of the value of its contained zinc, lead, copper and silver where these elements have the following values: Zn-\$0.32/lb; Pb-\$0.32/lb; Cu-\$0.62/lb and Ag-\$5.40/ tr. oz". The bulk concentrate contains 32% Zn, 10% Pb, 0.7% Cu, and 300 ppm Ag (8.8 oz/ST) and hence would have a value of \$97.50 per short ton. The tailings refloat product containing 30% Zn, 3.6% Pb, 0.6% Cu and 150 ppm Ag (4.4 oz/ST) at 30% of gross value would be credited with \$73.87 per short ton. 30% of gross value is an acceptable, and possibly high, payment for a true tailings refloat or middlings product which might be produced as a scavenger concentrate in addition to regular concentrates of zinc, copper and lead. However, 30% of value for a true bulk, i.e. one in which all of the recoverable metals report, would appear to be too low. Table 9 shows the composition of 32% zinc bulk concentrates which would be produced from the two producing mines, Brunswick and Heath Steele, and a potential producer, Anaconda Caribou. Recoveries of metals in these bulks are 95% Zn, 70% Pb, 70% Cu and 70% Ag*. At 30% of gross value, payment on a per ton ore basis would be \$25.58, \$15.85 and \$13.62 respectively for bulk concentrates from Brunswick, Heath Steele and Anaconda. The differences are due largely to considerably higher zinc content in the Brunswick ore, approximately twice the grade of the Heath Steele or Anaconda ore. This is reflected in the much higher concentration ratios, i.e. number of tons of ore required per ton of bulk concentrate, for Heath Steele and Anaconda relative to Brunswick ore. The 30% value for payment assigned by CANMET appears to have been arrived at from the Brunswick ore, for which a 30% payment for bulk, which is equivalent to \$25.58 per ton of ore, could be sufficient to return a profit to the mine-mill operation. However, from consideration of possible present net smelter returns to Heath Steele and the estimated break-even point for Anaconda, of approximately \$21 and \$20 per ton or ore respectively, it is clear that returns of \$15.85 and \$13.62 per ton of ore respectively, would not be sufficient, even considering a possible cost benefit which might

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* see p. 20

TABLE 9

VALUE OF BULK CONCENTRATES

	Brunswick			Heath Steele			Anaconda		
	%	lbs Metal	Full Value	%	lbs Metal	Full Value	%	lbs Metal	Full Value
zinc	32.00	640.0	$204.80	32.00	640.0	$204.80	32.00	640.0	$204.80
lead	9.50	190.0	60.80	10.00	200.0	64.00	8.95	179.0	57.28
copper	0.77	15.4	9.55	4.27	85.4	52.95	2.47	49.4	30.63
silver	7.16 oz/ton		38.66	10.66 oz/ton		57.56	9.00 oz/ton		48.60
			$313.81			$379.31			$341.31
30% of full metal value			$ 94.14			$113.79			$102.39
Concentration Ratio (ton ore/ton bulk)			3.681			7.182			7.519
Payment per ton or ore, (bulk at 30% of full value)			$ 25.58			$ 15.85			$ 13.62
Smelter payment per ton of ore, conventional conc.			-			$ 20.86*			$ 20.00†
True Value/ton Bulk conc.			-			$149.82			$150.38
% of Full Value			-			39.5			44.1

* from estimated net smelter returns.
† break-even estimate based on projected mining and milling costs.

accrue from simplified flotation. It would appear that payment for true bulk concentrate must be at a higher rate than 30% of full value. For this reason the economic evaluation of the bulk concentrate was carried out at payments of 35% and 40% in addition to 30%, so that the effect of escalated payment for concentrates can be assessed.

Evaluation using the tailings refloat product was at 30% of full value, which is reasonable for a middlings product. For comparison, an evaluation at 40% of value is also included. Total yearly payment for bulk and tailings refloat concentrates at varying percentages of full metal value and at zinc prices of \$0.24/lb; \$0.32/lb and \$0.40/lb are summarized in Table 13.

Three types of feed are utilized in the integrated sulphation roast plant. The dead roast utilizes 269 MTPD of high grade zinc concentrate (56% zinc) and the sulphation roaster uses approximately 218 MTPD of a tailings refloat product (2nd stage cleaning, 30% Zn, 3.6% Pb, 0.6% Cu and 4.4 Oz/ST Ag) plus 172 MTPD of middlings product of similar metal content to the tailings refloat product. Such a middlings product would be readily available, for example, from existing zinc flotation circuits if zinc concentrates of a grade higher than presently achieved are produced from upgrading from 52% to 56% zinc, or in part from a dezincing concentrate from the lead-copper circuit.

Payment for the 56% zinc concentrate is based on estimated net smelter returns from foreign smelters (Table A7-7).

Payment for the 30% zinc tailings refloat product (50% zinc recovery from tailings) is assigned on the basis of present payment for a 50% zinc concentrate (30% zinc recovery from tailings) as described in Appendix A7. Payment for the middlings product is at 30% of full metal value; this payment correlates well with the value which would be received for this product taken to a 52% zinc concentrate.

On the basis of $0.32/lb for zinc, the 56% zinc concentrate has a value of $164.78/ST, the tailings refloat product $55.42/ST and the middlings at $73.87/ST. Payment per ton of various zinc concentrates at varying zinc prices are given in Table A7-8. Total yearly payment for concentrates consumed by the integrated sulphation roast-dead roast plant at three price levels for zinc are summarized in Table 13.

6.6. REVENUE

The quantities of metals, as pure metal, lead oxide concentrate, cement copper, and sulphuric acid are shown in Table 10. Recoveries for each metal are shown in Table 11. Revenues for each product including zinc at $0.32/lb are shown in Table 12. Copper as cement copper is assigned 65% of market price; lead and silver are considered sold to the Belledune Lead smelter as per smelter schedules given in Table A7-7. Although consideration in this report is for brine leaching of the HAL residue to produce a high grade lead oxide product, the residue from the bulk concentrate product, as is, or with relatively little upgrading could be sold to the Belledune smelter. The HAL residue from bulk

TABLE 10

PRODUCTS
(350 op. days/yr)

	DR	SR (Bulk)	SR (Tailings Refloat)	SR/DR
1. Zinc slab (STPY)	100,000	100,000	100,000	100,000
2. Copper				
i) Electrolytic (STPY)	-	2,149	2,029	-
ii) Cement (STPY)	269*	-	-	1,080
3. Cadmium (STPY)	143	204	204	151
4. Lead (STPY)	2,429†	30,242[1]	11,421[2]	6,124[3]
5. Silver (Troy oz/yr)	259,312†	2,419,581	1,291,100	698,119
6. Sulphuric Acid (STPY)	159,834***	204,772	265,496	249,668

TABLE 11

RECOVERIES
(%)

	DR	SR (Bulk)	SR (Tailings Refloat)	SR/DR
Zinc	95.0	96.0	96.0	97.0
Copper	70.0	97.5	97.5	97.5
Cadmium	<60.0	89.6	89.6	60.6
Lead	100.0**	93.9[1]	91.8[2]	92.5[3]
Silver	70.0	85.0	85.0	85.0

* 202,429 STPY concentrate @ 0.19% Cu (Boorman 1975), 70% recovery remainder lost to jarosite

† 202,429 STPY concentrate @ 1.2% Pb + 1.83 oz/ST Ag (Boorman 1975) assume 100% recovery of lead as residue (8376 Tons @ 29% Pb), 70% recovery of silver.

** recovered as residue (29% Pb)

[1] as lead oxide product (87.87% Pb; 66 oz/S.T. Ag) - 94.59 MTPD
[2] " " " " (78.71% Pb; 89 oz/S.T. Ag) - 37.61 MTPD
[3] " " " " (71.31% Pb; 81 oz/S.T. Ag) - 22.26 MTPD

*** net quantity for sale, remainder used in leaching calcine

TABLE 12

REVENUE
($,000)/yr

		DR	SR (Bulk)	SR (Tailings Refloat)	SR/DR
1.	Zinc (slab)	64,000	64,000	64,000	64,000
2.	Copper				
	i) Electrolytic	-	2,665	2,516	-
	ii) Cement	217	-	-	871
3.	Cadmium	645	920	920	677
4.	Lead	711	12,981	4,861	2,562
5.	Silver	1,293	12,069	6,439	3,482
6.	Sulphuric Acid	2,398	3,072	3,982	3,745
	TOTAL	$69,264	$95,707	$82,718	$75,337

Unit Prices

1.	Zinc	$ 0.32/lb
2.	Copper	$ 0.62/lb
3.	Cadmium	$ 2.25/lb
4.	Lead	$ 0.32/lb
5.	Silver	$ 5.40/Tr. oz.
6.	Sulphuric Acid	$15.00/S.T.

concentrate contains 21.8% Pb as lead sulphate and 19.4 oz/T Ag and is not unlike a lead sulphate product (23% Pb; 23 oz/ton Ag) from Australia which is presently being purchased by the Belledune Lead Smelter. However, some upgrading, possibly via flotation, of the sulphation roast plant product would be required to reduce the amount of hematite. A more important consideration, however, is the finite amount of low grade material that the lead smelter could handle, and hence the decision to incorporate brine leaching to upgrade the lead product. Although the Belledune Lead Smelter could possibly handle the 36,500 short tons per year of high grade lead oxide product which would be generated from a bulk concentrate, their requirements for a lower grade sulphate product would be considerably lower. The difficulty would then be in selling the remainder to other smelters and absorbing high freight costs for the low grade product.

In addition to grade considerations, lead oxide is also a more desirable feed to the sinter machine than lead sulphate would be in assisting the thermal balance in the lead smelter.

Sulphuric acid is given a nominal f.o.b. Bathurst selling price of $15.00/ST.

Summaries of total revenue for each plant at three prices for zinc are given in Table 13.

TABLE 13

ECONOMIC SUMMARY
($,000/YR)

	DR	SR (Bulk)			SR (Tailings Refl.)		SR/DR
Revenue Zinc Price ($/lb)							
0.24	53,449	79,707			66,718		59,337
0.32	69,449	95,707			82,718		75,337
0.40	85,449	111,707			98,718		91,337
Direct Operating Cost	21,756	26,320			25,859		24,034
Depreciation*	8,661	10,784			10,877		9,589
Cost of Concentrates		Percentage of Total Metal Value as Payment					
Zinc Price ($/lb)		30%	35%	40%	30%	40%	
0.24	22,008	26,745	31,203	35,660	20,656	27,541	19,720
0.32	30,559	31,746	37,037	42,329	25,657	34,209	26,629
0.40	39,109	36,748	42,872	48,997	30,658	40,877	33,539
Profit (before tax) Zinc Price ($/lb)							
0.24	1,024	15,858	11,400	6,943	9,324	2,439	5,994
0.32	8,473	26,857	21,566	16,274	20,323	11,771	15,085
0.40	15,923	37,855	31,731	25,606	31,322	21,103	24,175
Total Capital Investment	143,560	176,270			177,430		157,680
Return on Investment (before tax) Zinc Price ($/lb)							
0.24	0.71	9.00	6.47	3.94	5.25	1.37	3.80
0.32	5.90	15.24	12.23	9.23	11.45	6.63	9.57
0.40	11.09	21.46	18.00	14.53	17.65	11.89	15.33

* 15 year straight line

6.7. PROFIT AND RETURN ON INVESTMENT BEFORE TAXES

Profit before taxes is defined as follows:

Profit = Revenue - (Direct Operating Costs + Depreciation).

Return on Investment (R.O.I.) before taxes is calculated as follows:

R.O.I. = Profit/Total Capital Investment,

where total capital investment includes fixed and working capital.

Summaries of profit and return on investment at varying zinc price and payment for concentrates are given in Table 13. This data is also plotted on Figures 6 to 12 as are data for the pressure leach process and a conventional dead roast plant.

The pressure leach data was obtained from Sherritt Gordon summary reports dated January 24, 1978 and March 22, 1978. All capital and operating costs and revenue given in these reports were taken as given without critical analysis, as sufficient details were not provided. However, for comparison purposes their capital costs were escalated by 4.3% (9 months) from second half 1977 to mid 1978 (Table A7-9). Their labour rates were raised to the $18,843/job level, the cost of electric power was raised from $0.0143 to $0.01866/ kwh, and their cost for Bunker C was decreased from $0.54 to $0.45/Imperial gallon, the same basis as that for the various roast plants. Data on capital and operating costs and an economic summary for the Sherritt Gordon Process are given in Tables A7-9 to A7-11.

The pressure leach process, when applied to bulk concentrates, consumes 112,000 tons/year of sulphuric acid.

"Black acid" is specified, and it appears that the bulk of the requirement is for the decomposition of plumbojarosite. While this grade of acid can be used for this purpose, it is unlikely that "black acid" carrying impurities could be used for the make-up in the zinc circuit where impurity build-up, without the exit afforded in pyrometallurgical operations, is already a problem. Hence, even with "black acid" available at $20/ST, make-up acid would be much more costly, i.e. approximately $87.50/ST delivered to Bathurst and therefore the average price of acid consumed by the pressure leach process would be higher than $20/ST.

The availability of "black acid" at $20/T may also be a problem since its production by the Belledune Lead smelter is presently used in total in the Belledune Fertilizer Plant. "Black acid", if available, would have to be imported and present transportation costs alone for unit train shipments from Ontario exceed $15.96/ST. Even with the closure of the Belledune Fertilizer Plant, it is unlikely that the lead smelter acid would be available to a company other than Noranda, for a cost less than $20/ST as present production costs for this acid exceed $17/ST.

As an alternative to purchasing acid for the pressure leach process, acid can be generated from a sulphur-pyrite residue produced in the process. The cost of acid production from this residue either by sulphur recovery and subsequent sulphur burning or by roasting of the residue would be considerably more than $20/ST. Reagent acid is a major cost sensitive item in the pressure leach process and hence, two costs representing possible minimum ($20/ST) and maximum ($40/ST) levels are shown on Figures 6 to 12. It is assumed that if purchased acid is unavailable within this price spread, that sulphuric acid could be produced from the sulphur-pyrite residue for a cost within these boundaries.

Figure 6 indicates that a sulphation roast plant to treat the bulk concentrate, for which payment is made at 35% of full metal value, would generate considerably more before tax profit than would a conventional dead roast plant or a pressure leach plant. Similarily the integrated sulphation plant also appears more profitable than either the pressure leach or dead roast plants. All four plants have different levels of capital expenditure, and this is not truely reflected in Figure 6 as costs of capital other than depreciation are not considered in the calculation of profit. Therefore better comparisons are possible from examination of before tax return on investment where both depreciation and full capital expenditure, including working capital, are considered.

Figure 7 shows the effect on return on investment resulting from different payments for the bulk concentrate at a zinc price of $0.32/lb. At 35% payment, considered minimum for true bulk concentrates, all plants are more profitable than a conventional dead roast plant. Returns are 12.23%, 9.57 and 5.90% respectively for each of the sulphation roast (bulk concentrate), integrated sulphation roast-dead roast and dead roast plants respectively. In compar-

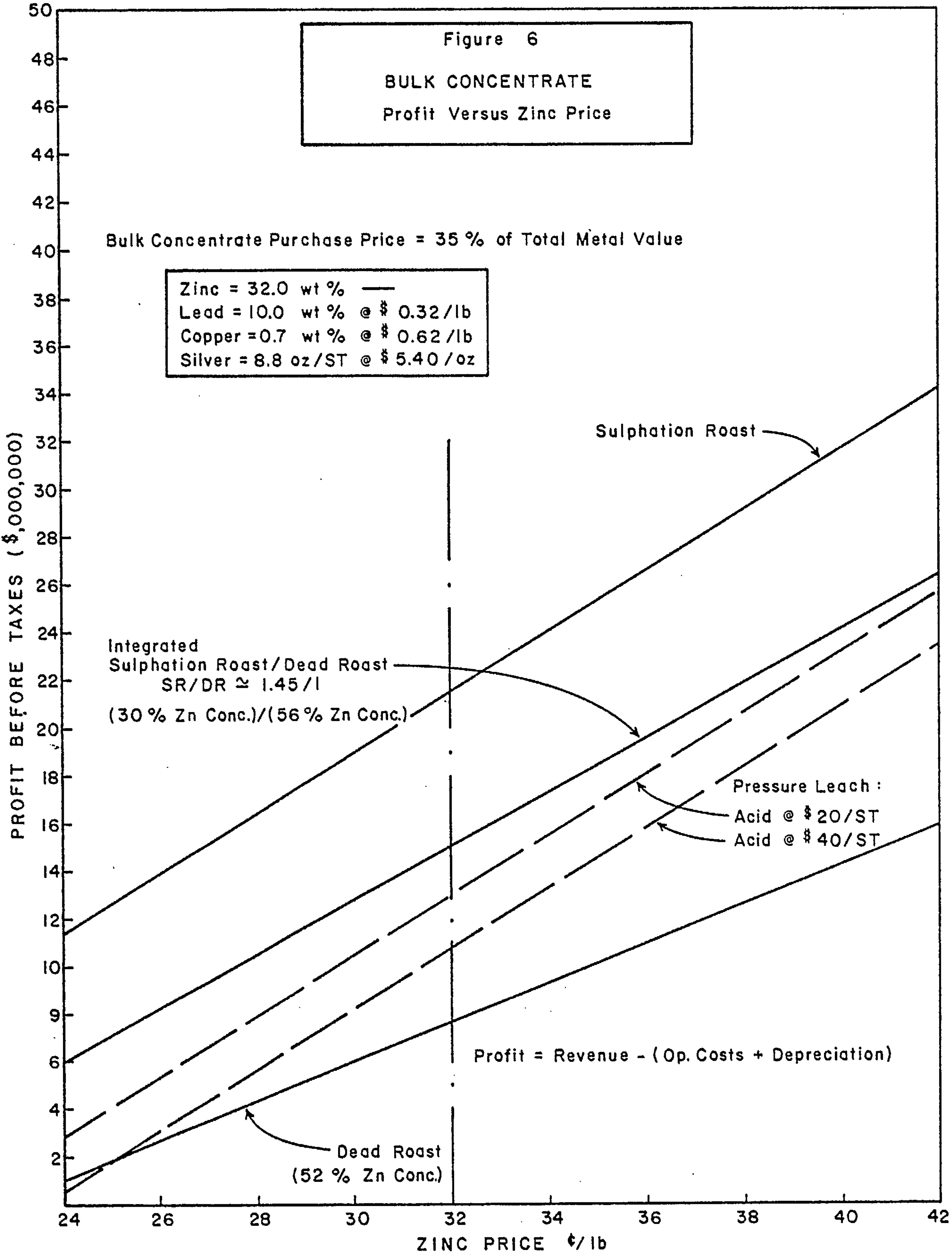

Figure 6

BULK CONCENTRATE

Profit Versus Zinc Price

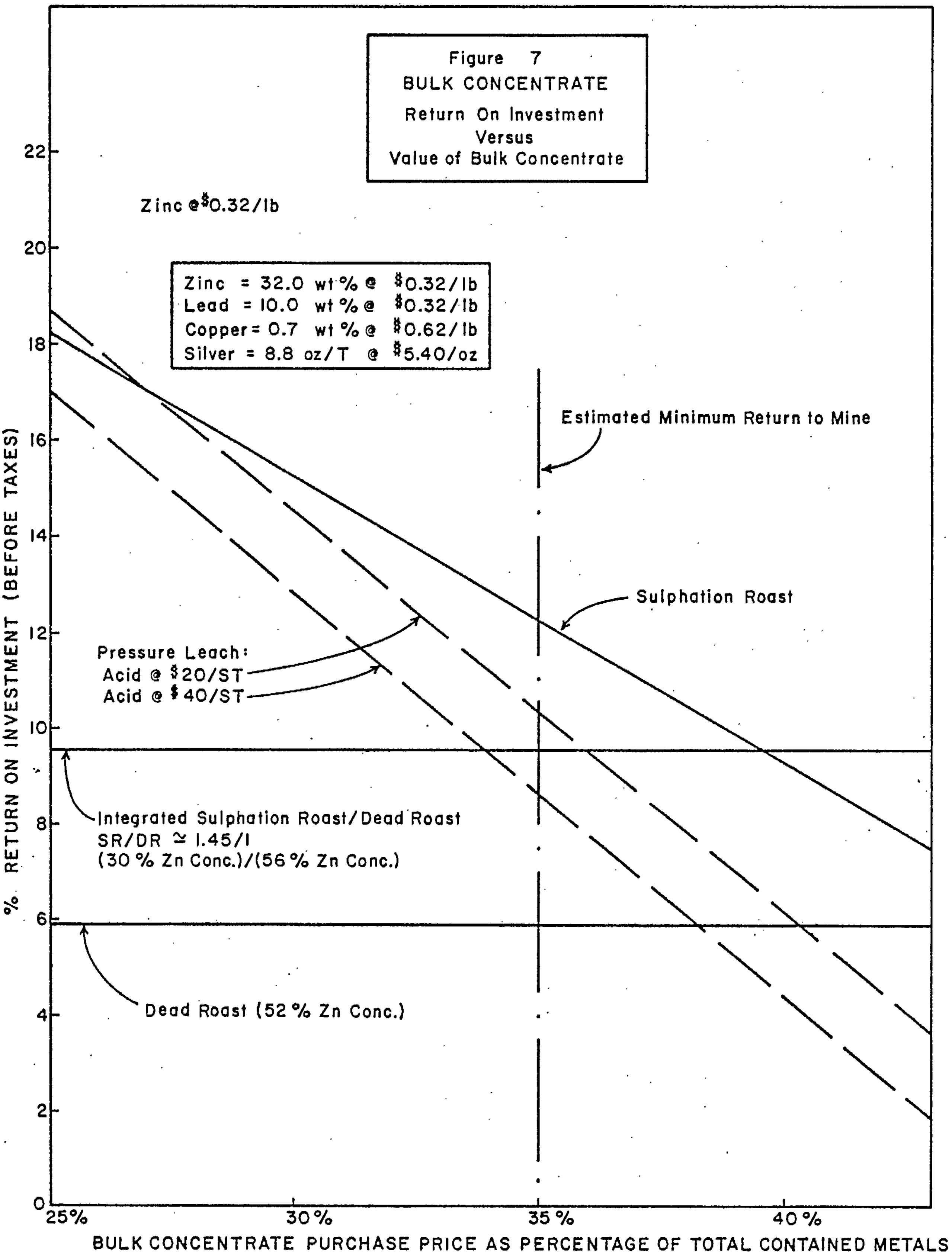

Figure 7
BULK CONCENTRATE
Return On Investment
Versus
Value of Bulk Concentrate
Zinc @$0.32/lb
Zinc = 32.0 wt % @ $0.32/lb
Lead = 10.0 wt % @ $0.32/lb
Copper = 0.7 wt % @ $0.62/lb
Silver = 8.8 oz/T @ $5.40/oz
Estimated Minimum Return to Mine
Sulphation Roast
Pressure Leach:
Acid @ $20/ST
Acid @ $40/ST
Integrated Sulphation Roast/Dead Roast
SR/DR ≃ 1.45/1
(30 % Zn Conc.)/(56 % Zn Conc.)
Dead Roast (52 % Zn Conc.)
% RETURN ON INVESTMENT (BEFORE TAXES)
0
2
4
6
8
10
12
14
16
18
20
22
25%
30%
35%
40%
BULK CONCENTRATE PURCHASE PRICE AS PERCENTAGE OF TOTAL CONTAINED METALS

ison, returns for the pressure leach process are between 8.6% and 10.3% depending on the cost of reagent acid. At 40% payment for concentrates, which might be approached for true bulks (see discussion Sec. 6.5), the profitability of the pressure leach process is equal to or less than that of a conventional dead roast plant depending on the cost for acid whereas the sulphation plants, bulk concentrate and integrated, are approximately 60% more profitable than a corresponding dead roast facility.

The effect of zinc price on profitability for the bulk concentrate plant (35% payment) is shown in Figure 8. This illustrates that all metal recovery plants with the exception of the pressure leach plant at $40/ST acid cost are more profitable than a conventional zinc plant. The profitability of the pressure leach process is much more susceptible to zinc price fluctuations than are the sulphation roast plants. At higher zinc prices, above 40¢/lb, and assuming no escalation in production costs, the pressure leach process becomes competitive with the sulphation roast process.

The effect of acid price, as revenue to the sulphation roast process and as a cost to the pressure leach process, is shown in Figure 9. At 32¢/lb for zinc even with give-away acid, the sulphation roast process is more profitable than the pressure leach process at the minimum purchase price for acid of $20/ST.

Similar economic evaluation plots for the tailings refloat concentrate are given in Figures 10 to 12. The

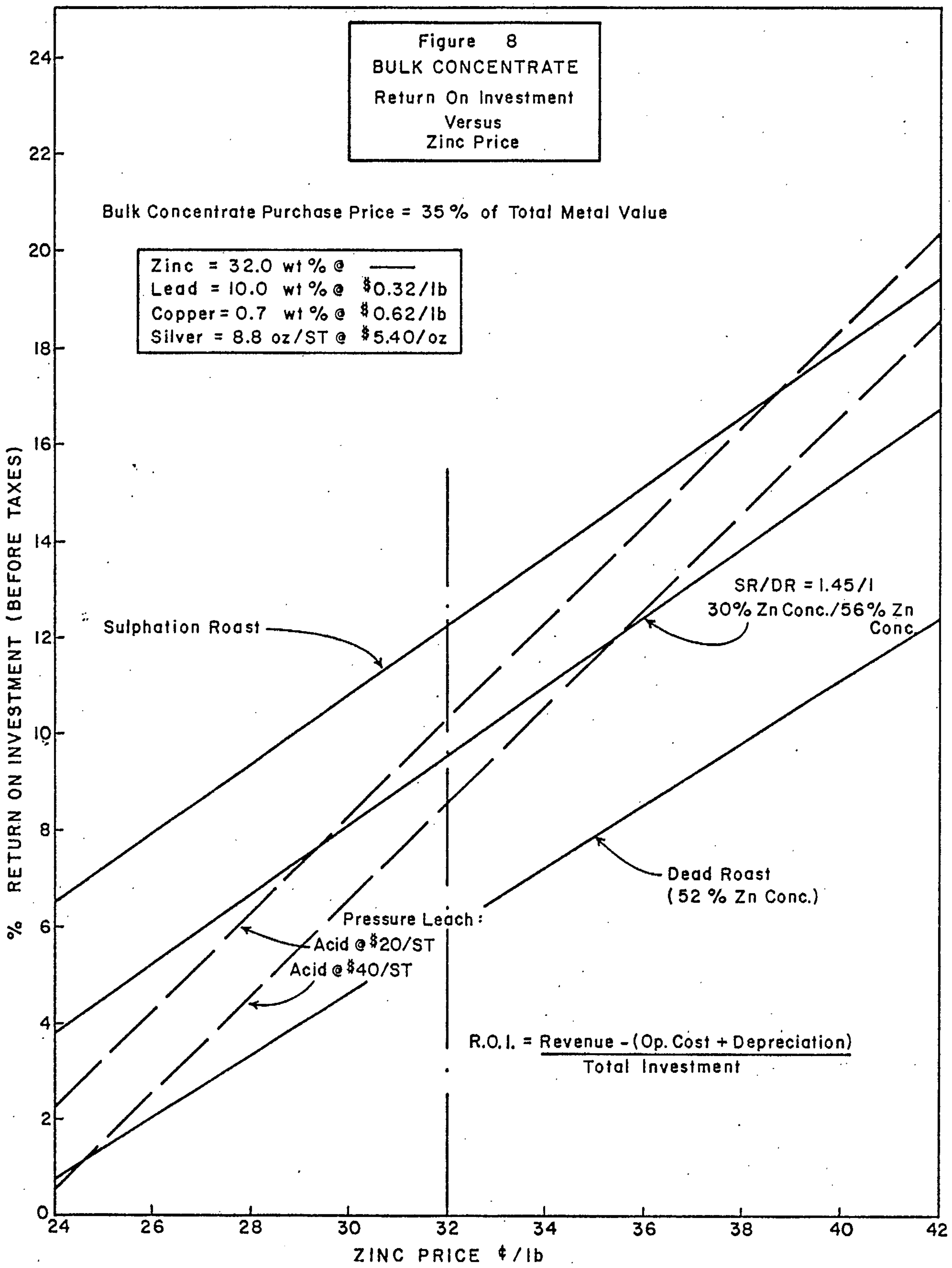

Figure 8
BULK CONCENTRATE
Return On Investment
Versus
Zinc Price

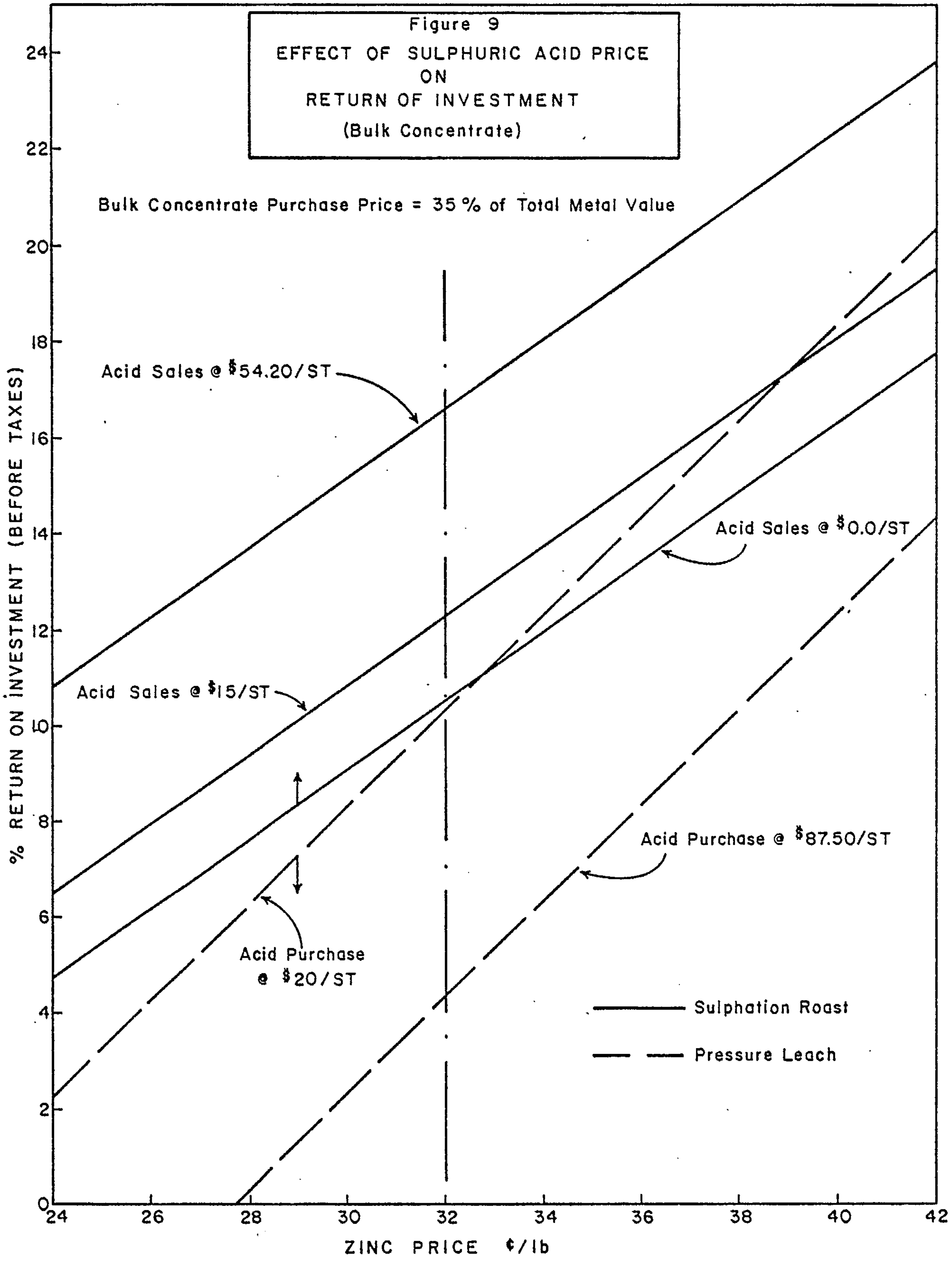

Figure 9
EFFECT OF SULPHURIC ACID PRICE
ON
RETURN OF INVESTMENT
(Bulk Concentrate)

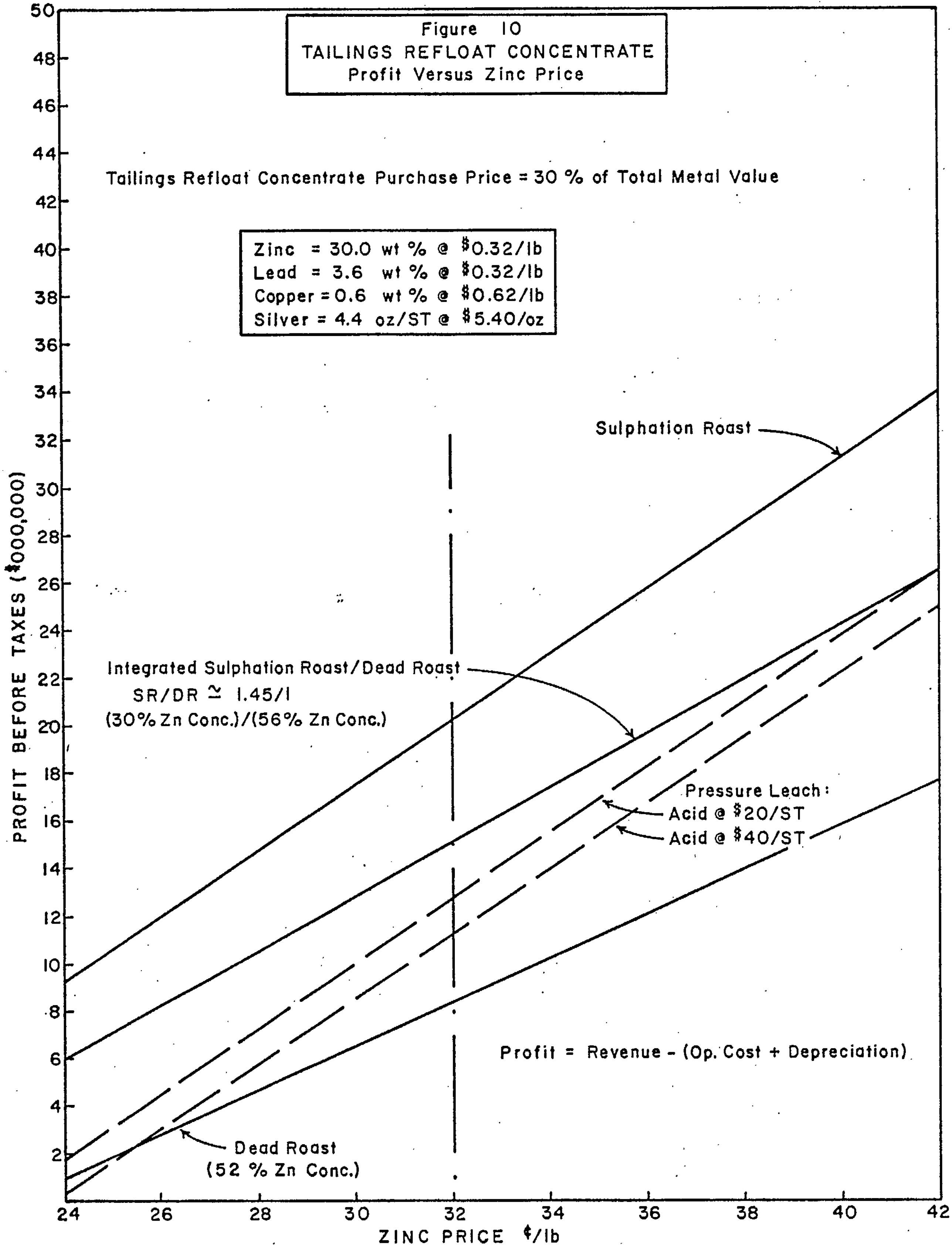

Figure 10
TAILINGS REFLOAT CONCENTRATE
Profit Versus Zinc Price
Tailings Refloat Concentrate Purchase Price = 30 % of Total Metal Value
Zinc = 30.0 wt % @ $0.32/lb
Lead = 3.6 wt % @ $0.32/lb
Copper = 0.6 wt % @ $0.62/lb
Silver = 4.4 oz/ST @ $5.40/oz
Sulphation Roast
Integrated Sulphation Roast/Dead Roast
SR/DR ≃ 1.45/1
(30% Zn Conc.)/(56% Zn Conc.)
Pressure Leach:
Acid @ $20/ST
Acid @ $40/ST
Profit = Revenue - (Op. Cost + Depreciation)
Dead Roast
(52 % Zn Conc.)
PROFIT BEFORE TAXES ($000,000)
50
48
46
44
42
40
38
36
34
32
30
28
26
24
22
20
18
16
14
12
10
8
6
4
2
24
26
28
30
32
34
36
38
40
42
ZINC PRICE ¢/lb

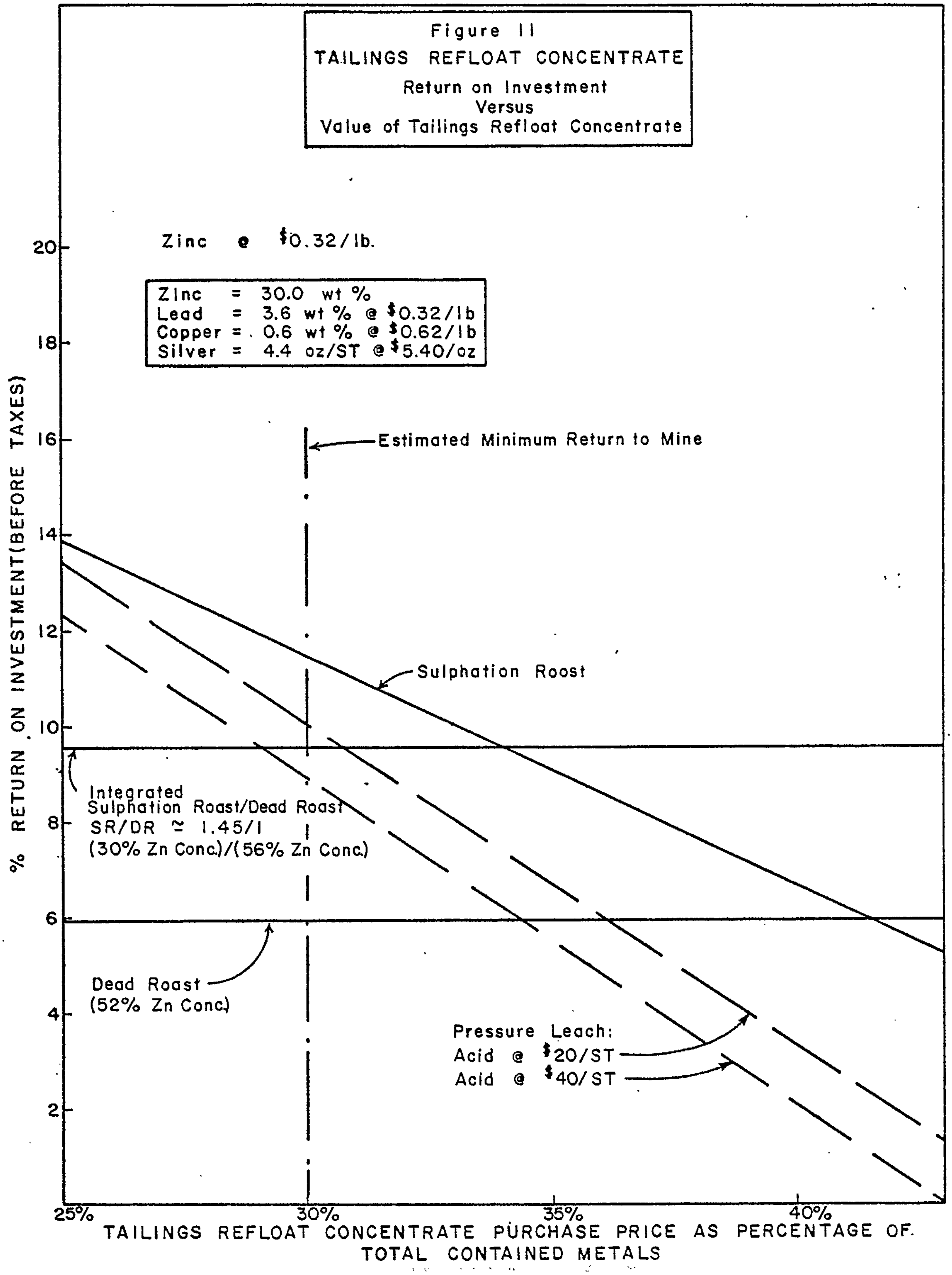

Figure 11
TAILINGS REFLOAT CONCENTRATE
Return on Investment
Versus
Value of Tailings Refloat Concentrate

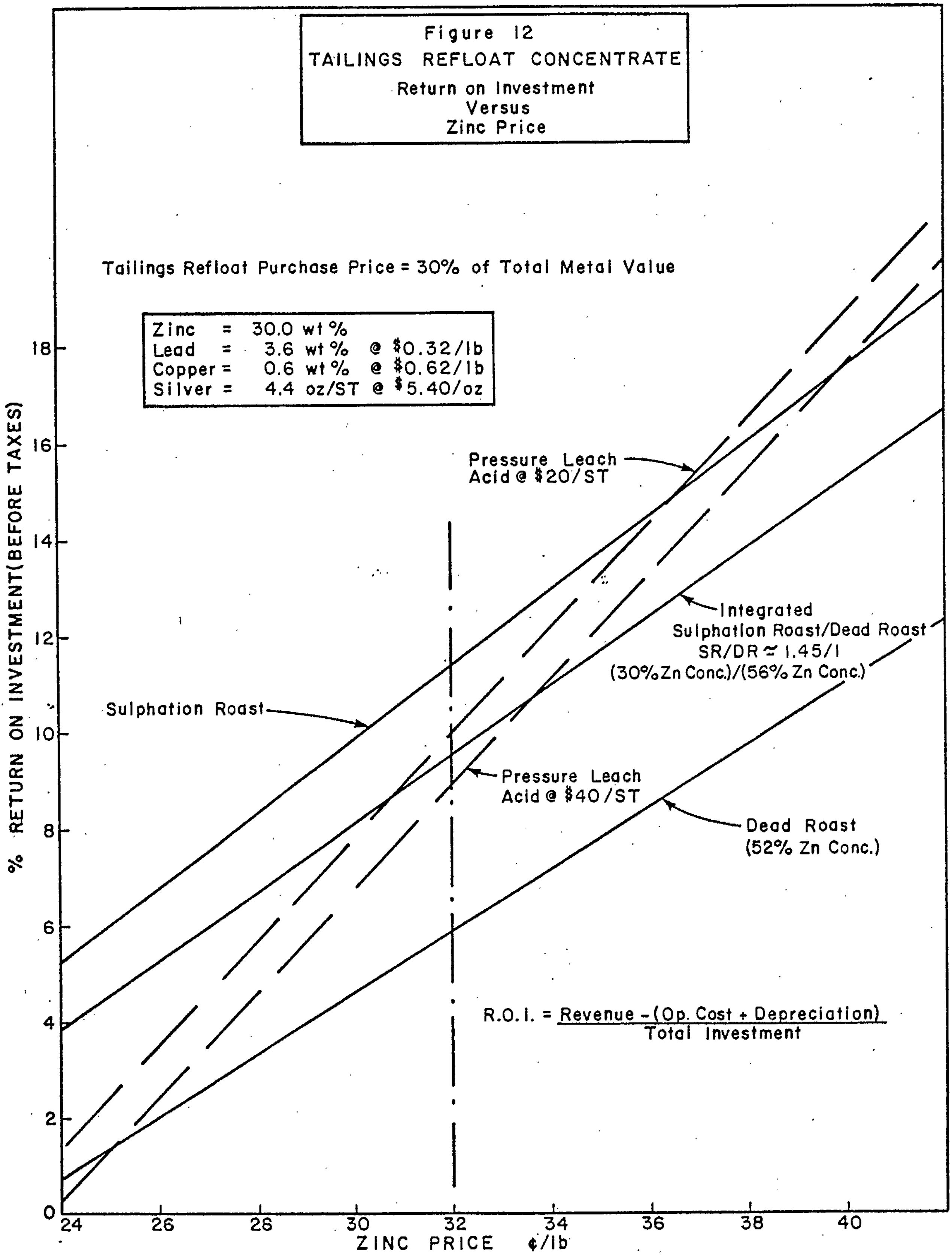

Figure 12
TAILINGS REFLOAT CONCENTRATE
Return on Investment
Versus
Zinc Price

comparative results for all plants are entirely similar to the bulk concentrate results. Thirty per cent payment for tailings refloat concentrates is a reasonable payment, or perhaps slightly high for a "scavenger" concentrate. At 30% payment and a zinc price of $0.32/lb, the profitability of the sulphation roast plant using tailings refloat is only slightly less, 11.45% versus 12.23% R.O.I. than for the bulk concentrate case.

Although the effect of varying lead content on overall plant economics can be assessed by comparison of the bulk concentrate plant with the tailings refloat plant, the effect of zinc grade has not been evaluated.

7.0 APPLICATION OF THE SULPHATION ROAST PROCESS TO NEW BRUNSWICK COMPLEX BASE METAL ORES

Although the return on investment from a sulphation roast plant, treating a true bulk concentrate, i.e. all recoverable metals in a single concentrate, may be higher than for a dead roast plant of similar slab zinc output, there is evidence to suggest that the profitability of both the metals recovery plant and producing mines can be increased through the use of "scavenger" concentrates rather than true bulk concentrate in the metals recovery plant. The mine-mill operation can maximize its return through the sale of higher grade metal concentrates with high recovery maintained through the sale of lower grade middlings products to the proposed metal recovery plant for payment of less than 30% of full metal value. This would ensure enhanced profitability of the metals recovery plant. For example, in the tailings refloat program at Brunswick approximately 30% of the zinc in tailings is recoverable as a 50% zinc concentrate. However, if the refloat process were halted at the second cleaning stage, the product would assay about 30% zinc and contain about 50% of the zinc in the tailings. The annual selling price for such a product would be the same or less than that of the 50% concentrate. This would result in a net payment for this 30% zinc tailings refloat product of less than $55.42/ST or less than 22.5% of full payment for metals in the product (Table A7-8). For this payment the return on investment for a sulphation roast plant, at a price for zinc of $0.32/lb, increases to over 15% (Figure 11). This is higher than could be achieved with a bulk concentrate for which much higher payments are required.

Other "scavenger" concentrates, in addition to the 240 STPD tailings refloat from Brunswick which are currently available or could be produced, are approximately 100 TPD of de-zinc concentrate from the Brunswick copper-lead circuit, and a possible tailings refloat product from Heath Steele. In addition to these products both Brunswick and Heath Steele could find it advantageous to upgrade their present metal concentrates and sell middlings to the proposed sulphation roast plant.

The Heath Steele ore, although relatively rich in copper, is low in zinc in comparison with the Brunswick ore and the total gross metal value of contained metals is only about 60% that of the Brunswick ore. Hence the problem of concentrate grade versus recovery of metals which is common to these types of ores is especially acute here. Zinc and lead concentrates in particular are of relatively low grade and metal losses to tailings are substantial, 19.83%, 29.66%, 34.51% and 39.83% for zinc, lead, copper and silver respectively (Table 14).

In addition to lower smelter returns for low grade concentrates, the additional freight charges for the lower grade concentrates are an added burden which leads to a marginal or submarginal operation, especially during depressed market conditions. The presence of a metals re-

TABLE 14

HEATH STEELE
CONVENTIONAL FLOTATION MODEL

	STPD	Wt % of Feed	Assays (wt % or oz/T)				% Distribution				Metal Distribution (STPD)			
			Zn	Pb	Cu	Ag	Zn	Pb	Cu	Ag	Zn	Pb	Cu	Ag
Feed	4000.00	100.00	4.69	1.99	0.85	2.12	100.00	100.00	100.00	100.00	187.60	79.60	34.00	8480
Zinc Conc.	295.43	7.39	48.05	1.96	0.78	2.98	75.67	7.27	6.76	10.38	141.95	5.79	2.30	880
Lead Conc.	158.02	3.95	3.52	29.57	1.24	16.90	2.96	58.71	5.76	31.50	5.56	46.73	1.96	2671
Copper Conc.	82.92	2.07	3.49	4.18	21.72	18.70	1.54	4.36	52.97	18.29	2.89	3.47	18.01	1551
Tailings	3463.63	86.59	1.07	0.68	0.33	0.98	19.83	29.66	34.51	39.83	37.20	23.61	11.73	3378

covery plant which would be capable of treating low grade products would be of particular benefit to Heath Steele or the Anaconda Caribou deposit as well as other yet undeveloped deposits in the province.

As stated previously, this facility would allow these mine-mill operations to aim at the production of relatively high grade metal concentrates for sale out of province at higher returns than are presently possible, while high recoveries would be ensured through the sale of middlings products to the custom sulphation roast metals recovery plant.

A custom smelter for the treatment of scavenger flotation products from a number of mines in the area is perhaps the best approach to efficient utilization of Northern New Brunswick mineral resources. However, sufficient middlings products to supply the majority of feed to a 100,000 STPY zinc-metals recovery plant could not be generated from a single producer, with the possible exception of Brunswick, and hence a consortium of companies might be required to realize such a plant.

The alternative to consortium ownership is possible through two majority ownership scenarios, neither of which would be as attractive as the custom smelter concept. In the first case, the Heath Steele model, a modified bulk concentrate produced from Heath Steele ore could be blended with the tailings refloat product from Brunswick and a middlings product also from Brunswick as feed for a sulphation roast plant. In the second case, the Brunswick model, the blend of tailings refloat and zinc middlings product is used as feed to a sulphation roaster and high grade zinc concentrate (56% Zn) to a dead roaster. This model is the integrated sulphation roast-dead roast plant discussed in the previous section.

7.1 HEATH STEELE MODEL

In this model a bulk concentrate of zinc, lead, copper and silver, is recovered from 4000 TPD of Heath Steele ore after the separation of a copper concentrate. The prefloating of a copper concentrate may or may not be necessary depending on the levels of copper plus lead which can be tolerated for roasting to operate satisfactorily. Table 15 illustrates the distribution of metals in the feed, copper concentrate, bulk concentrate and tailings for the proposed Heath Steele flotation model. The principal assumption in this distribution being that 95% of the zinc, 70% lead, 70% lead, 70% copper and 70% silver are recoverable in the two concentrates*; the copper concentrate contains the majority of the copper and the bulk concentrate contains most of the recoverable zinc, lead and silver.

In addition to the 547.91 STPD Heath Steele bulk concentrate, 240 STPD of Brunswick tailings refloat product and 167.63 STPD of Brunswick zinc middlings are required to bring zinc slab production up to 100,000 STPY. A summary of the economic evaluation for this model, based on the capital

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* see footnote p. 20

TABLE 15

HEATH STEELE BULK CONCENTRATE FLOTATION MODEL

	STPD	WT % of Feed	Assays (WT % or oz/T)				% Distribution				Metal Distribution (STPD)			(oz/ Day)
			Zn	Pb	Cu	Ag	Zn	Pb	Cu	Ag	Zn	Pb	Cu	Ag
Heath Steele														
Feed	4,000.00	100.00	4.69	1.99	0.85	2.12	100.00	100.00	100.00	100.00	187.60	79.60	34.00	8480
Copper Conc.	82.92	2.07	3.49	4.18	21.72	18.70	1.54	4.36	52.97	18.29	2.89	3.47	18.01	1551
Bulk Conc.	547.91	13.70	32.00	9.54	1.06	8.00	93.46	65.64	17.03	51.71	175.33	52.25	5.79	4385
Tailings	3,369.17	84.23	0.28	0.71	0.30	0.76	5.00	30.00	30.00	30.00	9.38	23.88	10.20	2544
Brunswick														
1. Tailings Refloat	240.00	-	30.00	3.60	0.60	4.37	-	-	-	-	72.00	8.64	1.44	1049
2. Middlings	167.63	-	30.00	3.60	0.60	4.37	-	-	-	-	50.29	6.03	1.01	733
TOTAL FEED TO SR	955.54	-	31.15	7.00	0.86	6.45	-	-	-	-	297.62	66.92	8.24	6167

and operating costs for a general bulk concentrate as given earlier, is presented in Table 16. At a zinc price of \$0.32/lb the return on investment for the sulphation roast plant would be greater than 12.5% before taxes. Payment for Heath Steele concentrates, i.e. the bulk concentrate to the sulphation roast plant and the copper concentrates to copper smelters, would return to the mine the same value per ton of ore, approximately \$20.90/ST, now received through the sale of separate metal concentrates. Advantages to the mine-mill operation in the proposed model are in savings achieved through simplification of flotation circuits.

In addition to enhanced profitability at the mine, overall recoveries, as paid for metals in concentrates, increases by approximately 23% (Tables 17 and 18). Also important is the added value generated within the province. For example, whereas net smelter returns to Heath Steele for sale of concentrates from the conventional flotation model would approximate \$84,719/day, the revenue from the sale of the copper concentrate and products produced within the sulphation roast plant from the Heath Steele bulk concentrate alone would approximate a return of \$171,805. This represents a 103% added value increase to a Heath Steele owned smelter from Heath Steele concentrates.

7.2. BRUNSWICK MODEL

This is the integrated sulphation roast-dead roast plant reviewed in Section 6. The hot acid leaching of

TABLE 16

HEATH STEELE MODEL
ECONOMIC EVALUATION

	SR ($,000)
Revenue	
zinc[†]	64,000
copper	3,360
lead	9,431
silver	9,150
cadmium	920
sulphuric acid	3,072
Total Revenue	89,933
Operating Cost	26,320*
Cost of Concentrates	30,800**
Depreciation	10,784
Net Profit (before taxes)	22,029
Total Capital	176,270
R.O.I. % (before taxes)	12.50

[†] Zinc @ $0.32/lb

* No reduction for quantities of CaO, NaCl, $CaCl_2$ for SR - Assume same operating costs as for general model

** Heath Steele bulk concentrate based on present value received for concentrate at the mine minus present value received for copper concentrates = $114.40/ST bulk conc.
Brunswick tailings refloat @ $55.42/ST } Table A7-8
Brunswick middlings @ $73.87/ST }

TABLE 17

METAL RECOVERY AND REVENUE
CONVENTIONAL FLOTATION
(4000 TPD Heath Steele Ore)

	Metals in Feed (4000 STPD)		Metals Recovered As Concentrates		Net Smelter Return (Foreign Smelters)
	Amount	Gross Value (Dollars/Day)	Amount	Gross Value (Dollars/Day)	(Dollars/Day)
Zinc (STPD)	187.60	120,064			
1. Zinc Conc.			141.95	90,848	38,343
Lead (STPD)	79.60	50,944			
1. Lead Conc.			46.73	29,907	11,016
Copper (STPD)	34.00	42,160			
1. Copper Conc.			18.01	22,332	13,491
2. Lead Conc.			1.96	2,430	1,215
Silver (oz/Day)	8480	45,792			
1. Copper Conc.			1551	8,375	7,336
2. Lead Conc.			2671	14,423	13,322
		$258,960		$168,315	$ 84,723
Value Recovery (As % of Gross Value in Feed)				65.0	32.7

TABLE 18

METAL RECOVERY AND REVENUE
SR-L-E PLANT
(4000 STPD Heath Steele Ore)[†]

	Metals in Feed (4000 STPD)		Metals Recovered As Concentrates		Metals Recovered in N. B. SR-L-E Plant	
	Amount	Gross Value (Dollars/Day)	Amount	Gross Value (Dollars/Day)	Amount	Gross Value (Dollars/Day)
Zinc (STPD)	187.60	120,064				
1. Bulk Conc.			175.33	112,211	168.32	107,725
Lead (STPD)	79.60	50,944				
1. Bulk Conc.*			52.25	33,440	49.04	21,049
Copper (STPD)	34.00	42,160				
1. Copper Conc.			18.01	22,332	18.01	13,491
2. Bulk Conc.			5.79	7,180	5.65	7,006
Silver (oz/Day)	8480	45,792				
1. Copper Conc.			1551	8,375	1551	7,334
2. Bulk Conc.*			4385	23,679	3727	18,588
		$258,960		$207,217		$171,195
Recovery (As % Gross Value in Feed)				80.0		67.7

* Recovery as a PbO product assaying ≃ 82.87% Pb, 62.98 oz/T Ag.

[†] Includes production of 82.92 TPD Copper Concentrate and 547.91 TPD Bulk Conc. (see Table 15)

neutral leach residues from dead roast operations was tested successfully in an integrated mode with sulphation roast products in the mini pilot trials (Appendix A1).

This integrated sulphation roast-dead roast plant, although seeming less profitable than either of the sulphation roast plants utilizing bulk concentrates or tailings refloat, is perhaps a less risky venture since in terms of return on investment the integrated plant is still 62% more profitable than a conventional dead roast plant. In addition, the integrated plant holds some distinct advantages over sulphation roasting alone as follows:

i) the nature of the roaster feed is not encumbered by major assumptions as to recoveries. This is a serious limitation in the bulk concentrate approach where the economic evaluation is based on recoveries of 95% Zn, 70% Pb, 70% Cu and 70% Ag as suggested by CANMET. This is probably a reasonable assumption for the Brunswick ore, and requires only a slight improvement over what has already been achieved by Brunswick for a 36% Zn, 12.5% Pb bulk with +92% Zn, 68% Pb, 50% Cu and 60% Ag recoveries (1967 company tests).* However, considerable test work is required to establish that these recoveries would be attainable from other deposits.

Feed for the sulphation roaster in the integrated plant is presently readily available from scavenger products in the mill, i.e. tailings refloat, middlings products from the zinc circuit and the de-zincing product from the copper-lead circuit.

ii) the use of zinc middlings products in the sulphation roaster would allow higher grade zinc concentrates to be produced in the mill for sale at enhanced revenue per ton of concentrate. Also, high grade (56%) zinc concentrates, when employed in the dead roaster segment of the integrated plant, will effectively result in reduction in the capital cost of the dead roaster section compared with costs in a conventional dead-roast plant roasting run-of-mill zinc concentrates at 52% zinc.

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* see footnote p. 20

iii) as spent electrolyte is required for the neutral leach of the dead roast calcine, the total amount of spent to be neutralized from the integrated plant is reduced by approximately 45% relative to the bulk concentrate sulphation roast plant.

Recovery of a 30% zinc concentrate from the Brunswick tailings refloat program rather than the 50% zinc concentrate presently required, would increase zinc metal recoveries for tailings from 30% to 50%. This would result in an additional 22 tons/day of recoverable zinc, 8.64 TPD lead, 1.44 TPD copper, and 104.9 oz/day silver (assuming that payment for lead, copper and silver is presently not received in a 50% zinc concentrate).

Utilization of this plant by Brunswick as a replacement for current, but apparently on the shelf, plans for a conventional zinc plant in northern New Brunswick would provide the company with much greater flexibility in its integrated mine-mill-zinc smelter-lead smelter operation. Of particular interest is the proposed lead oxide product produced by brine leaching of the HAL residue from the integrated plant. This product, although relatively minor tonnage, approximately 8500 STPY, compared with that from a sulphation roast of a bulk concentrate which would yield in excess of 36,000 STPY, would be most suitable as an addition to the sinter machine at the Belledune Lead Smelter. The use of this lead oxide product, which contains approximately 71% Pb and 81 oz/ton silver, in this manner would serve to offset sulphur and thermal imbalances in the sinter machine and would effectively increase the lead smelter output. Although low lead

(3.6%) concentrates were used in the model, the flexibility does exist for increasing the lead content, perhaps by addition of middlings products from the copper-lead flotation circuit, to increase the production of lead oxide for consumption in the lead smelter.

Further integration and improved utilization of Brunswick resources might be possible using the present stockpiled lead slag at Belledune as a replacement for limestone in the neutralization of spent zinc electrolyte as detailed in Appendix A5.1. The use of the slag (10.5% Zn) in this manner could result in recovery of a substantial proportion of the zinc in the slag i.e. a replacement process for zinc fuming plus possible recovery of some lead and silver.

Figure 13 is a schematic illustration of how the proposed integrated sulphation roast-dead roast plant could be completely integrated with other Brunswick operations.

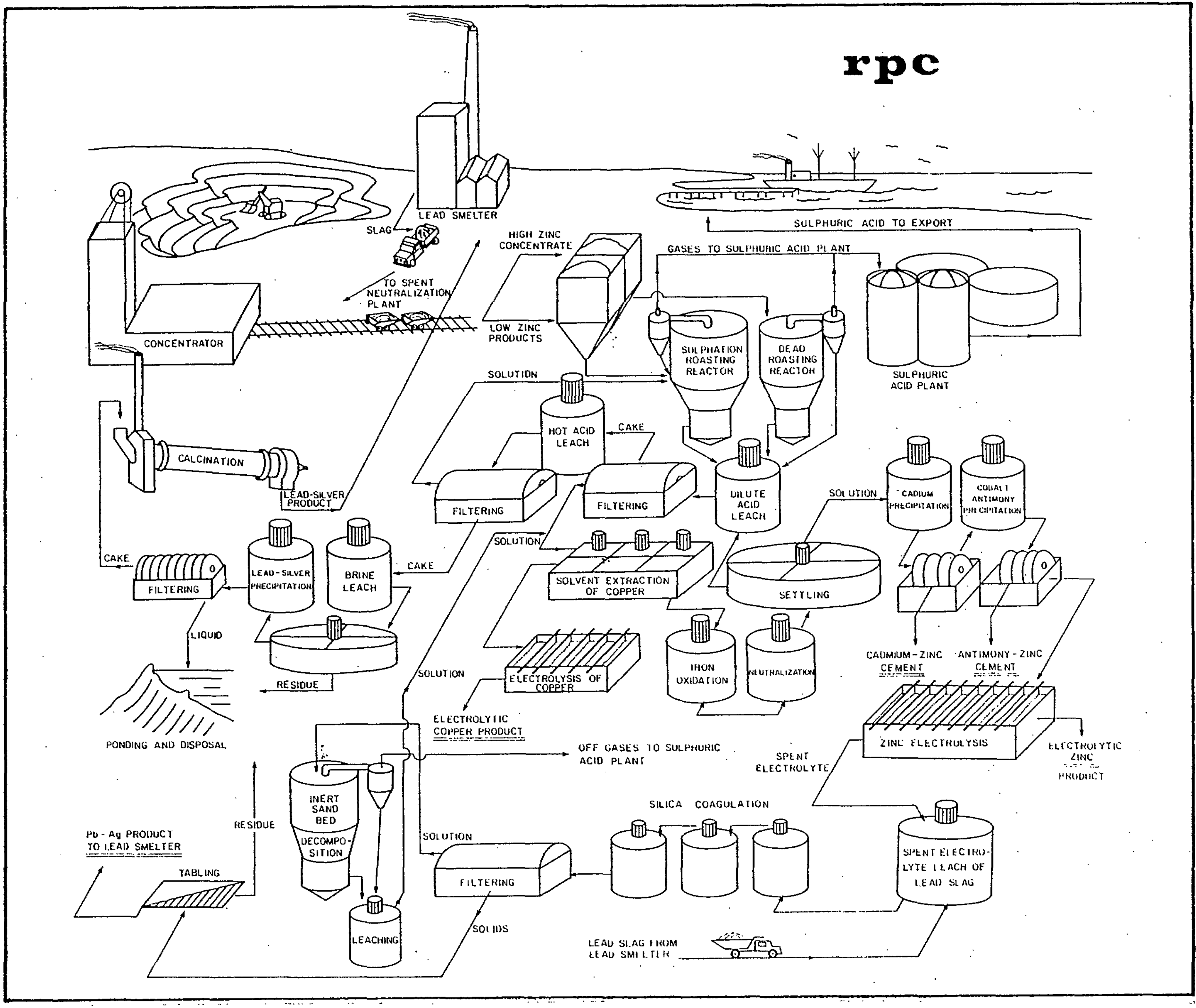

FIGURE 13 SCHEMATIC VIEW OF A SULPHATION ROAST/DEAD ROAST PLANT

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APPENDIX A1

RPC SULPHATION ROAST-LEACH-ELECTROWINNING PROCESS-DEVELOPMENT PROGRAM

APPENDIX A1

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APPENDIX A1

RPC SULPHATION ROAST-LEACH-ELECTROWINNING PROCESS DEVELOPMENT PROGRAM

A1.1 DESCRIPTION OF SEMI-PILOT PLANT

The roasting capacity of the plant was approximately 120-150 kg feed per day, on a three shift basis. Actual operation was limited to two consecutive shifts per day. The roasting equipment is shown in Plate 1 and illustrated schematically in Figure A1-1. The fluidized reactor, which was constructed of 316 stainless steel, measures 30 cm (14 in.) in diameter by 120 cm (47 in.) in height with an adjustable fluidized bed height of 40-55 cm. The gas distributor at the base of the reactor consisted of 19 mushroom-type tuyeres, each with six, 1 mm diameter holes and a total pressure drop across the distributor of 45 cm H_2O. The gas cleaning system includes a 20.3 cm (8 in.) diameter primary cyclone in series with a 15.2 cm (6 in.) diameter secondary cyclone and a dry single electrode electrostatic precipitator (cottrell).

Dust collected in the gas cleaning system discharged through individual rotary star valves rotating at 45 and 35 RPM and was continuously recirculated back into the reactor by means of a 3-inch diameter slanted screw feeder rotating at 25 RPM, as shown in Figure A1-1.

Slurry feed was sprayed from the top of the reactor by means of a compressed air operated gun, cooled with a

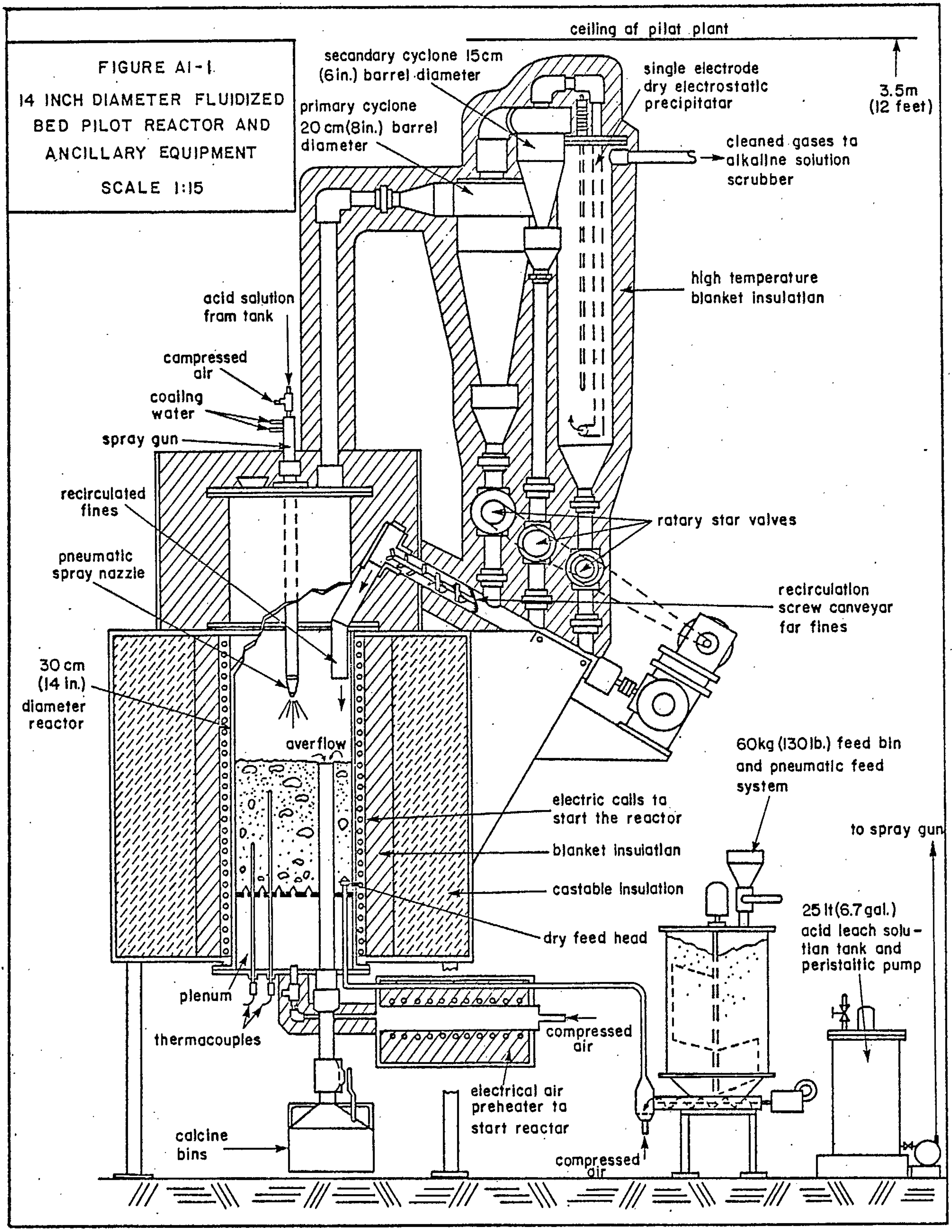


FIGURE A1-1.
14 INCH DIAMETER FLUIDIZED BED PILOT REACTOR AND ANCILLARY EQUIPMENT
SCALE 1:15

water jacket. Provision was made for varying the distance of the gun nozzle from the fluidized bed surface. Calibration data for this feed system is given in Figure A6-1. Dry feed was measured to the pneumatic injection system by means of a screw feeder, while slurry was pumped to the gun with a variable speed peristaltic rotary pump. The recirculated hot acid leach filtrate was sprayed into the reactor from the top, using the same gun and pump used for slurry feeding.

The reactor was started by means of electric heating coils wrapped around its body and with an electric air preheater. During operation, the unit was autothermally maintained, with no external power normally required.

External power, sufficient to compensate for the relatively high conductive heat losses associated with a reactor of this comparatively small size, was required to maintain thermal balance in the interlocked trials where hot acid leach filtrate was decomposed in the reactor.

The hydrometallurgical section of the RPC semi-pilot plant is illustrated in Figure A1-2. Design capacity was sufficient to handle 150 kg calcine per day on a three shift basis, however, actual operation was adjusted according to daily roaster production. Each calcine was leached and treated in each unit as a batch due to equipment and manpower limitations. Several operations were internally fully continuous such as copper solvent extraction and zinc electtrowinning and the leaching operations were semi-continuous

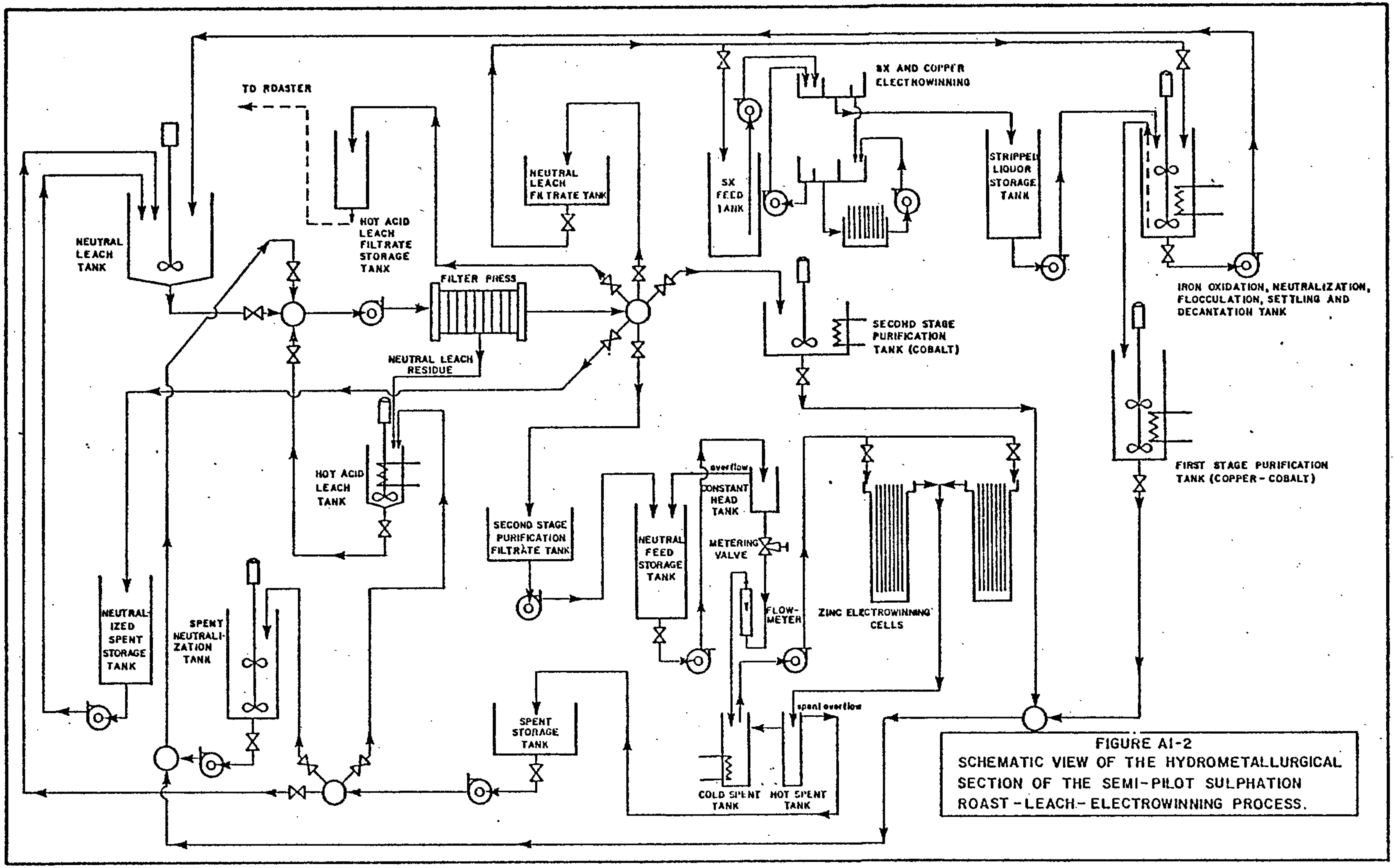


FIGURE A1-2
SCHEMATIC VIEW OF THE HYDROMETALLURGICAL SECTION OF THE SEMI-PILOT SULPHATION ROAST-LEACH-ELECTROWINNING PROCESS.

PLATE 1
ROASTING SECTION

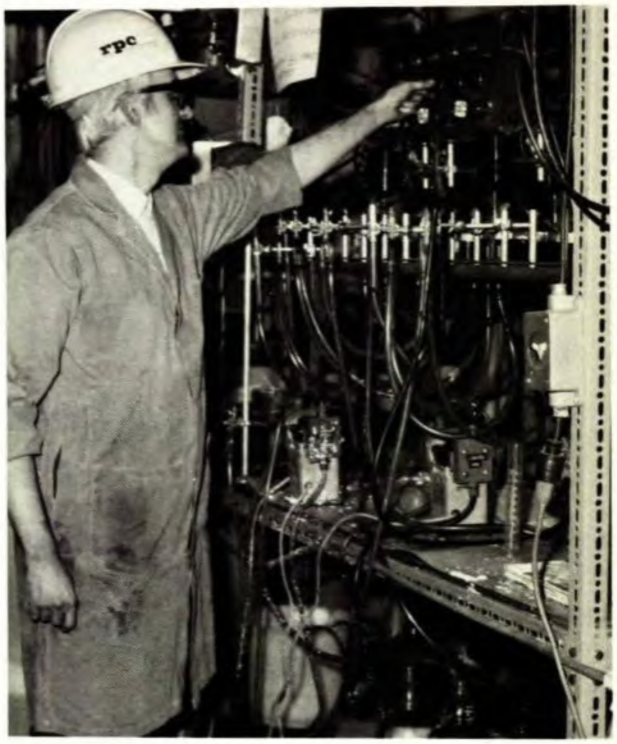
14" Diameter Pilot Fluidized Bed Reactor
And Ancillary Equipment

Adjusting Compressed Air Operated
Liquor Spray Gun in Reactor

Calcine Discharge From Reactor

PLATE 2
HYDROMETALLURGY

Leaching and Filtration



Cu Solvent Extraction

Purification

Zn Electrowinning

rpc

in that acidity was kept constant by continuous addition of acid. Plate 2 shows typical procedures during operation. Detailed discussion of equipment and operation is given in Section A1.3.

A1.2 ROASTING SECTION

121 Feeds Tested

Five different products were tested in the roasting circuit. Three products are representative of low grade zinc concentrates with zinc contents from 22 to 32% and varying amounts of lead and copper. Also tested was a 51% zinc concentrate and a blend of a tailings refloat (30% Zn) product and a ferrite residue from a conventional dead roast zinc plant.

The ranges of chemical compositions of the samples tested are given in Table A1-1. The detailed chemical analyses of feeds for each run are given in Table A1-2.

For the slurry feed tests, feed with approximately 10% moisture was slurried with spent electrolyte. For the dry feed tests the sulphide material, which was originally milled to 85% minus 325 mesh, was dried to less than 2% moisture and screened to -100 mesh.

The process development was carried out for the most part using tailings refloat products from Brunswick Mining and Smelting. Concentrates with higher lead and copper levels were prepared by blending lead concentrates from Brunswick Mining and Smelting and copper concentrates from Heath Steele Mines Ltd.

TABLE A1-1

RANGES OF COMPOSITION OF SAMPLES TESTED IN SULPHATION ROASTING (wt %)

	(1) Tailings Refloat Conc.	(2) Bulk Conc.	(3) High Cu Bulk Conc.	(4) Zinc Conc.	(5) Tailings Refloat and Ferrite Res.
Zn	22-32	27-31	30.5	51.2	32
Cu	0.4-0.9	0.7-0.9	4.5	0.5	0.4
Pb	2-4	10-14	4.8	2.0	2
Fe	22-26	18-20	21.5	9.9	26
S	36-40	34-35			
Co	0.015-0.025	0.020-0.023	0.053	0.006	0.002
Cd	0.060-0.065	0.053-0.065	0.060	0.097	0.055
As	0.3-0.5	0.3-0.5	0.27	0.75	0.4

(1)	Series	B2C, BMS, BMD	(Tailings refloat from Brunswick Mining and Smelting)
(2)	"	BHL	(Bulk concentrate, low in copper)
(3)	"	BHB, BHC, BHD	(Bulk concentrate, high in copper)
(4)	"	BZC	(Low-grade zinc concentrate)
(5)	"	BF	(Tailings refloat and ferrite residue)

TABLE A1-2

CHEMICAL ANALYSES OF FEEDS TESTED IN SULPHATION ROASTING

CONCENTRATES	RUN #	Zn	Cu	Pb	Fe	S	Co (ppm)	Cd (ppm)	As	Ag (ppm)	Sn	Sb (ppm)	In (ppm)	Ni (ppm)	CaO	Al_2O_3	SiO_2	MgO	Ge (ppm)	Ga (ppm)	Cl	Hg
Zn tailings relfoat II-2	B2C 1-22	22.80	0.22	1.82	28.50	42.76	-	-	-	-	-	-	-	-	0.10	0.67		0.03	-	-	-	-
" " " I	" 23-29	22.30	0.28	2.70	28.00	41.28	-	-	-	-	-	-	-	-	0.14	1.76		0.03	-	-	-	-
" " " II-1	" 30-40	22.70	0.22	2.46	28.50	42.20	-	-	-	-	-	-	-	-	0.28	0.94		0.08	-	-	-	-
" " " III	BHS 1-16	18.40	0.37	2.70	31.40	42.90	-	-	-	114	-	-	-	-	0.31	0.15	1.66	0.10	-	-	-	-
" " " VI	BMD. 1-22	29.80	0.98	3.13	23.41	40.05	-	-	-	136	-	-	-	-	0.28	0.21	0.58	0.10	-	-	-	-
" " " VI	" 23-36	29.00	1.16	3.30	23.70	39.50	-	-	-	-	-	-	-	-	0.36	0.19	0.82	0.11	-	-	-	-
" " " VII	" 37-45	29.50	1.28	3.70	22.50	39.10	230	650	0.44	94	-	-	-	-	0.35	0.23	0.86	0.11	-	-	-	-
" " " VIII (Blend 25% Zn + 51% Zn)	" 46-51	30.50	0.54	2.60	24.38	39.60	205	615	0.20	-	-	-	-	-	-	-	-	-	-	-	<0.01	trace
" " " IX (Blend 24% Zn + 51% Zn)	" 52	30.50	0.48	2.65	23.75	37.59	170	650	0.35	91	-	-	-	-	0.18	-	0.90	0.10	<3	8	"	"
" " " VII	" 53	29.50	1.28	3.70	22.50	39.10	230	650	0.44	94	-	-	-	-	0.35	0.23	0.86	0.11	-	-	"	"
" " " IX (Blend 24% Zn + 51% Zn)	" 54-61	30.50	0.48	2.65	23.75	37.59	170	650	0.35	91	-	-	-	-	0.24	-	0.90	0.10	3	8	"	"
" " " X (Blend 23% Zn + 51% Zn)	" 62-64	32.62	0.46	2.30	22.38	37.82	225	645	0.45	115	<0.01	<25	290	<100	0.18	-	0.68	0.10	3	16	"	"
Blend Zn tailings refloat X + Ferrite residue	BF 1-2	31.79	0.39	1.96	25.97	32.15	200	550	0.38	98	<0.01	<21	247	<90	0.15	-	0.58	0.09	-	-	"	"
Zn conc. (low grade)	BZC 1	51.20	0.50	2.00	15.90*	27.60*	60	970	0.20	75	<0.01	100	-	-	-	-	-	0.12	-	-	-	-
Blend Zn tailings refloat VII + Zn conc. + Pb conc.	BHL 1-4	30.80	0.74	10.64	18.40	34.39	200	625	0.33	240	<0.01	-	-	-	0.22	-	-	-	-	-	<0.01	trace
" " conc. + Pb conc.	" 5	26.75	0.92	14.30	19.40	34.14	230	530	0.45	240	<0.01	-	-	-	-	-	-	-	-	-	"	"
Blend Zn conc. + Cu-Pb conc.	BHB 1-2	28.75	5.12	9.05	19.38	33.00	500	560	0.25	250	<0.01	-	-	-	-	-	-	-	-	-	"	"
" " tailings refloat VII + Zn conc. + Cu-Pb conc.	BHC 1	30.50	6.80	3.00	19.00	32.56	520	580	0.28	260	<0.01	-	-	-	-	-	-	-	-	-	"	"
" " " " " " " " "	BHD 1	30.50	4.48	4.80	21.50	35.15	530	600	0.30	270*	<0.01	-	-	-	0.89*	0.20*	0.49*	0.34*	-	-	"	"

(*) Calculated Values

122 Process Development

In the early stages in the development of the Process, consideration was given to the use of a slurry feeding system for the reactor. The reasoning being that spent zinc electrolyte could be used as the liquid medium, and reactor cooling agent, and hence the quantity of electrolyte required to be neutralized would be reduced. This was the approach followed earlier by the St. Joseph Lead Company in their attempt to develop a sulphation-roast process for zinc (1) as well as by the Dowa Co. in Japan in their commercial zinc-copper sulphation plant (2). The principal problems encountered by Dowa and in the St. Joseph Lead program, i.e. high iron levels in leach solutions going to purification and poor zinc recoveries, were found to be principally due to an enhancement of ferrite formation which was identified as a problem inherent to slurry feeding. As a result, the slurry feed approach was eventually abandoned in favour of dry feeding.

The main responses of the roasting operation were the ferrite formation, measured by x-ray diffraction, and the zinc extraction in both water and dilute acid, monitored by means of small, 50 g calcine sample control leaches in water (16% solid, 25°C, 1 hr) and in dilute acid, (16% solid, 25 gpl H_2SO_4, 25°C, 1 hr). Under these conditions the control leaches produce solutions with 30 to 35 gpl Zn.

The following summarizes the experimental programs for both feed systems.

.1 Slurry Feed System

A constant 70% solid slurry, made with synthetic spent electrolyte of 140 gpl H_2SO_4 and 65 gpl Zn as $ZnSO_4$ as the liquid medium and a Brunswick Mining and Smelting tailings refloat product (22-32% Zn) as the solids fraction, was pumped by a positive displacement pump through a specially designed air-operated gun. Pumping the corrosive slurry proved to be difficult in the early trials, and a Moyno pump, commonly assumed to be adequate to this kind of task, failed continually. Later, a plastic hose "finger pump" was used which performed satisfactorily and, finally, a rotary peristaltic pump with a Tygon plastic hose performed well.

The design and construction of the slurry gun required a considerable trial and error effort to obtain an adequate and adjustable pattern of droplet size and at the same time avoiding buildup of a dried crust on the tip of the gun or defluidization of the bed caused by "streaming" rather than spraying. An attempt to use the Falconbridge-type slurry gun was not successful since an acceptable dispersion pattern of droplets was not easy to achieve at low flow rates for the slurry and crust formation was a continuous problem. Corrosion and erosion were serious problems at the gun nozzle tip. Tips made of SS316 lasted only three to four days in most cases. Later, Hastelloy C-276 was found to be the best material of construction for the gun tip and nozzle.

Up to five weeks of continuous work, with little erosion and no chemical attack were experienced with the Hastelloy tips (Figure A6-1).

The experimental conditions and leaching results for the roasting work are shown in Table A1-3.

In previous bench scale work (RPC Report M/77/209), the formation of ferrite and the presence of unreacted sphalerite were identified as the main problems to be solved if high zinc extraction with acceptably low levels of iron are to be achieved in acid leach solutions. Results from all tests using slurry feed proved that sphalerite levels could be maintained below 1.5 wt%, however, ferrite formation was found to be a serious problem, reaching levels from 15 up to 40 wt% in the calcines (Table A1-4).

The operating conditions and design parameters of the slurry gun were found to influence the formation of ferrite to some extent. The nozzle tip diameter of the gun and the air flow rate determines the pattern and droplet size. When a rather large tip diameter of 4 mm was used, calcines were coarse (10% plus 10 mesh) and pellets not well formed, while calcines produced with a 1.5 mm diameter tip were finer (100% minus 40 mesh) and pellets round and well formed. The extent of ferrite formation is related to the size of the pellet and hence to the tip diameter of the gun. The ferrite content in calcines decreased by approximately 20% when the tip diameter was reduced from 4 to 1.5 mm (Figure A1-3).

TABLE A1-3(a)

ROASTING CONDITIONS AND CONTROL LEACH RESULTS FOR TAILINGS REFLOAT-SERIES B2C

Run No.	Temp.	Av. Ret. Time	Air Ratio	Feed Rate	Gas Velocity	Slurry Feeder System						% Metal Extraction*						Promoter %
						Spent Electrolyte, (gpl)		Pulp % solids	Slurry gun			water			dil. acid			
	T (°C)	$\bar{t}$ (hrs)	Q (-)	Fo (kg/hr)	u'_o (cm/sec)	H_2SO_4	Zn		Tip Ø (mm)	Air (lt/min)	Level Lg (cm)	Cu	Zn	Fe	Cu	Zn	Fe	
B2C Series																		
(1)																		
B2C-10	650	6.72	1.72	6.87	22.1	140	65	70	4	37.5	30	7.3	50.0	0.0	59.5	62.7	1.7	-
B2C-11	650	6.70	1.72	6.87	22.1	140	65	70	4.5	65	40	8.2	49.1	0.0	59.7	68.0	2.1	-
12	685	6.71	1.90	6.80	24.7	140	65	70	4	56	40	10.5	48.0	0.0	69.5	71.1	1.5	-
13	650	6.72	1.90	6.80	24.2	140	65	70	4	56	40	22.1	52.3	0.0	66.8	70.3	2.2	-
14	680	6.73	2.00	5.0	22.6	140	65	70	4	65	40	4.7	65.1	0.0	88.2	71.0	0.7	-
(2)																		
B2C-15	660	6.57	1.80	5.0	20.3	140	65	70	3.5	72	40	1.3	63.4	0.0	86.8	69.5	0.6	-
16	675	6.65	1.80	5.0	20.6	140	65	70	2.3	72	40	3.4	65.0	0.1	74.0	72.7	2.4	-
16	-60 mesh fraction			✓	✓	✓	✓	✓	✓	✓	✓	2.1	65.2	0.2	73.8	73.6	2.6	-
16	-35/+60 mesh fraction			✓	✓	✓	✓	✓	✓	✓	✓	-	-	-	75.3	71.3	1.6	-
16	+35 mesh fraction			✓	✓	✓	✓	✓	✓	✓	✓	-	-	-	75.0	71.1	1.3	-
17	700	6.46	1.80	5.0	20.8	140	65	70	2.3	72	40	1.5	50.0	0.0	72.8	68.2	0.6	-
18	650	6.70	1.80	5.0	19.2	140	65	70	2.3	72	40	2.1	58.2	0.0	71.7	71.5	2.8	-
19	675	7.16	1.40	5.0	15.3	140	65	70	2.3	72	40	1.2	46.2	0.0	71.0	68.3	2.5	-
20	675	6.26	2.05	5.0	23.2	140	65	70	2.3	72	40	-	41.1	-	-	66.3	-	-
21	675	6.66	1.60	5.0	18.3	140	65	70	2.3	72	40	-	46.8	-	-	68.1	-	-
22	675	6.65	1.80	5.0	20.6	140	65	70	2.3	72	60	-	45.9	-	-	68.2	-	-
23	675	6.70	1.80	5.0	20.6	140	65	70	2.3	72	30	-	52.7	-	-	70.5	-	-
24	675	6.70	1.80	5.0	20.6	140	65	70	2.3	72	20	-	55.0	-	-	72.5	-	-
25	685	6.57	1.80	5.0	21.1	140	65	70	2.3	72	50	1.9	49.0	0.0	70.6	71.7	1.1	-
26	665	6.73	1.80	5.0	20.2	140	65	70	2.3	72	20	2.4	58.4	0.0	73.4	73.0	1.8	-
27	675	6.33	1.85	5.0	20.9	140	65	70	2.3	72	20	3.1	61.5	0.0	73.3	75.1	1.5	-
28	675	6.85	1.75	5.0	20.5	140	65	70	2.3	72	20	0.6	47.9	0.0	72.3	73.4	1.2	-
(3)																		
B2C-29	675	6.57	1.80	5.0	20.6	140	65	70	2.3	72	20	0.0	44.0	0.0	71.7	72.3	0.8	-
30	675	6.62	1.80	5.0	20.6	-	-	70	2.3	72	20	-	31.1	-	-	61.1	-	-
31	675	6.58	1.80	5.0	20.6	-	-	-	2.3	72	20	-	13.7	-	-	43.3	-	1.0% Na_2CO_3
32	675	6.55	1.80	5.0	20.6	-	-	-	2.3	72	20	-	23.5	-	-	51.7	-	1.0% Na_2CO_3
33	685	6.57	1.80	5.0	20.2	140	65	70	2.3	72	20	1.7	49.1	0.0	70.7	71.0	0.9	-
34	675	6.55	1.80	5.0	20.6	140	65	70	2.3	72	20	1.7	56.1	0.0	66.5	67.9	1.7	0.5% Na_2SO_4
35	675	6.55	1.80	5.0	20.6	140	65	70	2.3	72	20	0.0	27.7	0.0	45.4	50.6	0.9	1.0% Na_2SO_4
36	675	6.56	1.80	5.0	20.6	-	-	-	2.3	72	20	3.3	51.2	0.0	51.2	60.6	1.5	-
37	675	6.56	1.80	5.0	20.6	-	-	-	2.3	72	20	-	34.5	-	-	59.7	-	0.3% KNO_3
38	675	6.53	1.80	5.0	20.6	-	-	-	2.3	72	20	-	21.2	-	-	36.4	-	1.0% NaCl
39	675	6.33	1.85	5.0	20.9	140	65	70	2.3	72	20	-	46.3	-	-	62.0	-	-
40	675	7.74	1.85	5.0	20.9	140	65	70	1.5	72	10	-	70.5	-	-	72.9	-	-

* Metal extractions based on control leaches of 50 gm. calcine samples.
Water leach: 16% solid, 25°C, 1 hr.
Acid leach : 25 gpl H_2SO_4, 16% solid, 25°C, 1 hr.

✓ Same conditions as calcine.

TABLE A1-3(b)

ROASTING CONDITIONS AND CONTROL LEACH RESULTS FOR TAILINGS REFLOAT-SERIES BMS

Run No.	Temp.	Av. Reten. Time	Air Ratio	Feed Rate	Gas Velocity	Slurry Feeder System						% Metal Extraction						Promoter %
						Spent Electrolyte, (gpl)		Pulp % solids	Slurry gun			water			dil. acid			
	T (°C)	$\bar{t}$ (hrs)	Q (-)	Fo (kg/hr)	u'_o (cm/sec)	H_2SO_4	Zn		Tip Ø (mm)	Air (lt/min)	Level Lg (cm)	Cu	Zn	Fe	Cu	Zn	Fe	
BMS Series																		
BMS- 1	675	6.8	1.88	5.0	20.6	140	65	70	1.5	72	10	-	43.8	-	-	70.0	-	-
2	675	6.8	1.80	5.0	19.8	140	65	70	1.5	72	10	-	64.8	-	-	69.5	-	-
3	660	6.8	1.80	5.0	19.8	140	65	70	1.5	72	10	-	64.3	-	-	73.1	-	-
4	675	6.8	1.80	5.0	19.8	140	65	70	1.5	72	10	-	64.2	-	-	72.4	-	2% Bent.
5	675	6.8	1.80	5.0	19.8	140	65	70	1.5	72	10	-	69.9	-	-	79.9	-	2% Bent. 0.5% Na_2SO_4
6	675	6.8	1.80	5.0	19.8	140	65	70	1.5	72	10	-	68.5	-	-	69.8	-	2% Bent. 0.5% Na_2SO_4
7	675	6.8	1.80	5.0	19.8	140	65	70	1.5	72	10	-	78.5	-	-	76.0	-	1% Bent. 0.75% Na_2SO_4
8	675	6.8	1.80	5.0	19.8	140	65	70	1.5	72	10	-	70.4	-	-	73.8	-	1% Bent. 1% Na_2SO_4
9	675	6.8	1.80	5.0	19.8	140	65	70	1.5	72	10	-	75.2	-	-	76.7	-	1% Na_2SO_4
9	1st cyclone		✓	✓	✓	✓	✓	✓	✓	✓	✓	-	-	-	-	61.6	-	✓
9	2nd cyclone		✓	✓	✓	✓	✓	✓	✓	✓	✓	-	-	-	-	69.9	-	✓
9	cottrell		✓	✓	✓	✓	✓	✓	✓	✓	✓	-	-	-	-	59.0	-	✓
10	675	6.8	1.80	5.0	19.6	140	65	70	1.5	72	0	-	75.5	-	-	78.0	-	-
11	675	6.0	1.20	8.0	19.8	140	65	70	1.5	72	10	-	66.4	-	-	75.3	-	-
12	675	6.8	1.80	5.0	19.8	-	-	-	1.5	72	10	-	62.7	-	-	73.6	-	-
13	675	6.8	1.80	2.0	19.8	-	-	-	-	-	-	-	67.3	-	-	79.8	-	-
13 -	250 mesh fraction		✓	✓	✓	-	-	-	-	-	-	-	70.0	-	-	83.1	-	-
14	675	6.8	1.80	2.0	19.8	-	-	-	-	-	-	-	64.3	-	-	79.9	-	-
14 -	250 mesh fraction		✓	✓	✓	-	-	-	-	-	-	-	61.6	-	-	82.2	-	-
15	675	6.7	2.0	2.0	20.9	-	-	-	-	-	-	-	66.0	-	-	76.6	-	-
15 -	250 mesh fraction		✓	✓	✓	-	-	-	-	-	-	-	69.7	-	-	83.7	-	-
16	675	6.7	1.60	2.0	18.3	-	-	-	-	-	-	-	74.4	-	-	82.7	-	-
✓	✓	✓	✓	✓	✓	-	-	-	-	-	-	-	78.6	-	-	85.0	-	-

TABLE A1-3(c)

ROASTING CONDITIONS AND CONTROL LEACH RESULTS* FOR TAILINGS REFLOAT-SERIES BMD

Run No.	Temp.	Av Ret. Time	Air Ratio	Feed Rate	Velocity	Slurry & Liquor Feed									% Metal Extraction					
	T (°C)	t (hr)	Q (–)	Fo (kg/hr)	u'o (cm/sec)	Spent Electrolyte, (gpl)		Pulp % solids	Slurry Gun			Cooling		Dry Feed (mesh)	water, 25°C			acid, 25°C, 25gpl		
						H_2SO_4	Zn		Tip Ø (mm)	Air (lt/min)	Level Lg (cm)	water (cc/min)	liquor (cc/min)		Cu	Zn	Fe	Cu	Zn	Fe (gpl)
BMD Series																				
BMD – 1	675	7.1	1.80	5.0	19.8	140	65	70	1.5	72	20	–	–	–	–	67.48	–	–	80.36	–
2	675	7.0	2.00	5.0	22.8	140	65	70	1.5	72	20	–	–	–	–	69.70	–	–	78.21	–
3	675	7.0	2.00	5.0	22.8	140	65	70	1.5	72	20	–	–	–	–	–	–	–	–	–
4	675	7.3	1.60	5.0	17.2	140	65	70	1.5	72	20	–	–	–	–	69.40	–	–	75.59	–
5	675	7.1	1.90	5.0	20.8	140	65	70	1.5	72	20	–	–	–	–	72.21	–	–	74.48	–
6	675	7.2	1.90	5.0	20.8	140	65	70	1.5	72	20	–	–	–	–	–	–	–	–	–
7	675	7.1	1.80	5.0	19.8	140	65	70	1.5	72	-5	–	–	–	–	70.90	–	–	74.70	–
8	675	7.1	1.80	5.0	19.8	140	65	70	1.5	72	20	–	–	–	–	73.60	–	–	26.20	–
9	675	7.1	1.80	5.0	19.8	140	65	70	1.5	72	20	–	–	–	–	73.34	–	–	76.21	–
10	675	7.1	1.80	5.0	19.8	140	65	70	1.5	72	20	–	–	–	–	75.61	–	–	75.61	–
11	675	7.0	1.80	5.0	19.8	140	65	70	1.5	72	20	–	–	–	–	–	–	–	–	–
12	675	7.0	2.40	5.0	26.5	140	65	70	1.5	72	25	–	–	–	–	72.55	–	–	75.82	–
13	675	7.1	1.85	5.0	20.3	140	65	70	1.5	72	20	–	–	–	–	71.02	–	–	78.92	–
14	675	7.1	1.60	5.0	17.5	140	65	70	1.5	72	20	–	–	–	–	–	–	–	–	–
15	675	7.1	1.60	5.0	17.5	140	65	70	1.5	72	20	–	–	–	–	40.90	–	–	69.90	–
16	675	6.9	2.20	5.0	23.2	140	65	70	1.5	72	20	–	–	–	–	76.36	–	–	79.20	–
17	630	7.1	2.20	5.0	23.0	140	65	70	1.5	72	20	–	–	–	–	77.58	–	–	80.39	–
18	660	7.1	2.00	5.0	21.9	–	–	–	1.5	72	20	–	–	-100	–	74.29	0.21	–	79.76	4.41 (0.53)
19	660	6.9	2.00	5.0	21.3	–	–	–	1.5	72	20	43	–	-100	–	80.03	0.35	–	83.40	3.75 (0.43)
20	660	6.9	2.00	5.0	21.9	–	–	–	1.5	57	20	–	38	-100	–	85.65	2.44	–	88.57	5.81 (0.63)
20 – 14 mesh		✓	✓	✓	✓	–	–	–	✓	✓	✓	✓	✓	✓	–	–	–	–	88.00	6.10 (0.68)
20 + 14 mesh		✓	✓	✓	✓	–	–	–	✓	✓	✓	✓	✓	✓	–	–	–	–	91.16	6.51 (0.37)
21	660	7.0	2.00	5.0	21.9	–	–	–	1.5	50	20	29	–	-100	–	89.38	1.74	–	89.38	5.72 (0.66)
22	660	6.9	1.80	5.0	18.9	–	–	–	1.5	50	35	49	–	-100	–	82.48	0.21	–	86.71	3.70 (0.40)
(1)																				
23	660	6.9	1.80	5.0	18.9	–	–	–	1.5	58	35	37	–	-100	–	79.41	0.49	–	84.10	6.59 (8.74)
24	668	6.3	2.28	5.0	24.2	–	–	–	1.5	65	35	17	–	-100	–	76.74	0.32	–	86.73	5.25 (0.57)
25	660	6.0	2.40	5.0	26.1	–	–	–	1.5	68	35	16	–	-100	–	62.46	0.02	–	84.81	4.57 (0.54)
26	700	6.8	2.00	5.0	22.0	–	–	–	1.5	33	35	–	–	-100	–	37.88	0.02	–	76.45	1.23 (0.16)
27	680	6.9	2.00	5.0	21.9	–	–	–	1.5	42	35	–	–	-108	–	40.45	0.82	–	72.38	0.31 (0.04)
28	660	6.9	2.00	5.0	21.8	–	–	–	1.5	44	35	33	–	-100	–	60.25	0.02	72.33	82.68	2.50 (0.30)
29	675	6.8	1.80	5.0	19.2	–	–	–	1.5	50	35	25	–	-100	–	49.20	0.02	51.27	80.52	2.92 (0.08)
38	675	6.4	1.60	5.0	15.3	–	–	–	1.5	57	35	45	–	-100	–	45.15	0.02	–	78.49	1.23 (0.15)
31	675	6.8	1.80	5.0	19.2	–	–	–	1.5	57	35	13	–	-100	–	53.45	0.02	70.81	86.02	1.33 (0.16)
32	675	6.7	1.70	5.0	18.8	–	–	–	1.5	57	35	19	–	-100	–	56.64	0.01	83.72	85.57	0.88 (0.10)
33	675	6.7	1.70	5.0	18.8	–	–	–	1.5	57	35	25	–	-100	–	56.15	0.01	81.82	80.82	2.14 (0.52)
34	675	6.9	1.90	5.0	21.0	–	–	–	1.5	57	35	37	–	-100	–	65.15	0.01	84.47	84.62	1.42 (0.16)
35	675	6.8	1.80	5.0	19.2	–	–	–	1.5	57	35	35	–	-108	–	53.55	0.01	82.96	84.46	0.65 (0.08)
36	675	6.8	1.80	5.0	19.2	–	–	–	1.5	57	45	46	–	-100	–	45.11	0.01	–	80.25	0.18 (0.02)
(2)																				
37	675	6.8	1.80	5.0	19.2	–	–	–	1.5	57	35	29	–	-100	–	59.62	–	–	83.88	–
38	675	6.0	1.60	5.7	19.2	–	–	–	1.5	72	35	18	–	-100	–	49.10	–	–	80.01	–
39	675	4.7	1.30	7.3	19.2	–	–	–	1.5	72	35	93	–	-100	–	35.67	–	–	71.34	–
40	675	4.7	1.60	7.3	24.6	–	–	–	1.5	72	35	79	–	-100	–	42.60	–	–	77.94	–
41	675	6.8	1.80	5.0	19.2	–	–	–	1.5	43	20	–	36	-100	–	71.89	0.00	–	83.79	2.76 (0.36)
42	675	6.8	1.80	5.0	19.2	–	–	–	1.5	22	20	–	32	-100	–	77.55	0.02	–	85.27	2.62 (0.35)
42 + 45 mesh		✓	✓	✓	✓	–	–	–	✓	✓	✓	–	✓	✓	–	–	–	–	86.70	3.20 (8.43)
42 – 45 mesh		✓	✓	✓	✓	–	–	–	✓	✓	✓	–	✓	✓	–	–	–	–	83.63	3.24 (0.44)
43	685	6.7	2.00	5.0	24.8	–	–	–	1.5	22	20	–	30	-100	–	81.19	0.19	–	87.94	3.40 (0.47)
44	685	6.7	2.00	5.0	24.8	–	–	–	1.5	22	28	–	24	-100	–	82.80	0.03	–	88.78	3.23 (0.44)
45	675	6.5	2.20	5.0	26.6	–	–	–	1.5	22	20	–	10	-100	–	80.93	0.04	–	89.30	3.50 (0.48)
(3)																				
46	685	6.8	2.00	5.0	24.8	–	–	–	1.5	22	20	–	31	-100	–	75.81	3.52	–	87.87	6.57 (0.78)
47	685	6.8	1.88	5.0	19.2	–	–	–	1.5	43	20	–	32	-100	–	63.60	0.02	–	85.51	1.59 (0.18)
48	685	6.8	1.80	5.0	19.2	–	–	–	1.5	43	20	–	28	-100	–	50.38	0.02	–	81.88	0.25 (0.03)
49	685	6.8	1.80	5.0	19.2	–	–	–	1.5	22	20	–	30	-100	–	54.95	0.02	–	83.69	1.34 (0.16)
50	685	6.8	1.80	5.0	19.2	–	–	–	1.5	36	20	–	40	-100	–	58.78	0.02	–	83.59	1.45 (0.17)
51	685	6.9	2.00	5.0	24.5	–	–	–	1.5	36	20	–	22	-100	–	56.68	0.02	–	85.23	1.45 (0.17)
(4)																				
52	685	6.8	1.80	5.0	19.8	–	–	–	1.5	43	20	–	14	-100	24.17	64.92	0.02	85.25	85.70	1.78 (0.20)
(3)																				
53	685	6.8	1.80	5.0	19.8	–	–	–	1.5	36	20	–	31	-100	18.39	73.37	0.04	82.94	85.90	2.07 (0.26)
(5)																				
54	685	6.8	1.80	5.0	19.8	–	–	–	1.5	36	20	–	38	-100	38.50	80.07	0.13	87.66	85.66	3.35 (0.37)
55	685	6.8	1.80	5.0	19.8	–	–	–	1.5	36	20	–	38	-100	5.58	70.74	0.01	79.40	80.26	2.48 (0.28)
56	685	6.8	1.80	5.0	19.8	–	–	–	1.5	36	20	–	38	-100	26.50	75.32	0.14	82.70	83.55	2.89 (0.33)
57	685	6.6	1.85	5.0	20.8	–	–	–	1.5	57	35	–	38	-100	5.88	72.94	0.01	83.00	82.33	2.49 (0.28)
58	685	6.6	1.85	5.0	20.8	–	–	–	1.5	57	35	–	32	-100	15.48	74.98	0.14	84.50	86.20	2.51 (0.28)
59	685	6.6	1.85	5.0	20.8	–	–	–	1.5	43	35	–	28	-100	5.90	71.72	0.01	83.90	84.91	2.39 (8.26)
60	685	6.6	1.85	5.0	20.8	–	–	–	1.5	58	35	–	32	-100	4.40	68.60	0.01	82.10	83.36	2.21 (0.25)
61	685	6.6	1.85	5.0	20.8	–	–	–	1.5	50	35	–	32	-100	17.50	72.28	0.04	84.20	83.72	2.76 (0.33)
62	685	6.6	1.85	5.0	20.8	–	–	–	1.5	43	35	–	32	-100	18.10	75.13	0.01	83.50	85.60	2.96 (0.34)
63	685	4.9	1.85	5.5	22.5	–	–	–	1.5	50	35	–	31	-100	3.34	64.83	0.01	81.25	83.91	2.18 (0.24)
64	685	3.8	1.85	7.0	27.8	–	–	–	1.5	50	35	–	44	-100	1.71	61.88	0.01	88.00	82.91	2.81 (0.31)
(6)																				

(1) BMD 1-22 Zn middlings V
(2) BMD 23-36 Zn middlings VI
(3) BMD 37-45 Zn middlings VII
(4) BMD 46-51 Zn middlings VIII (Blend 25% Zn + 51% Zn)
(5) BMD 54-61 Zn middlings IX (Blend 24% Zn + 51% Zn)
(6) BMD 62-64 Zn middlings X (Blend 23% Zn + 51% Zn)

* (results from BMD 41 to 64 represent primary metal extraction only)

TABLE A1-4(a)

ZINC COMPOUNDS IN BRUNSWICK TAILINGS REFLOAT CALCINES
SERIES B2C

	Weight %				Zinc Distribution (%)			
	sphalerite	ferrite	oxysulphate	sulphate	sphalerite	ferrite	oxysulphate	sulphate
B2C-1	Rejected due to operational problems							
2	" " " " "							
3	" " " " "							
4	" " " " "							
5	" " " " "							
6	" " " " "							
7	" " " " "							
8	" " " " "							
9	" " " " "							
10	3.94	27.42	12.62	27.80	8.42	26.84	22.10	40.64
11	3.88	29.45	18.58	24.14	8.31	28.91	32.63	35.38
12	1.17	25.99	22.04	20.62	2.59	26.32	39.93	31.18
13	1.40	26.88	17.55	26.41	3.03	26.68	31.16	39.14
14	0.48	26.61	5.56	38.69	1.10	27.91	10.43	60.59
15	0.97	27.31	5.58	38.13	2.20	28.36	10.37	59.12
16	2.26	21.19	7.06	38.03	5.17	22.19	13.22	59.46
17	1.50	28.81	17.77	28.82	4.87	28.55	31.51	42.65
18	3.22	22.37	13.20	35.46	6.83	21.73	22.94	48.54
19	1.01	26.96	19.76	18.54	2.39	29.24	38.34	30.03
20	1.57	30.57	24.53	15.36	3.40	30.33	43.54	22.76

TABLE A1-4(a) (Cont'd)

ZINC COMPOUNDS IN BRUNSWICK TAILINGS REFLOAT CALCINES
SERIES B2C

	Weight %				Zinc Distribution (%)			
	sphalerite	ferrite	oxysulphate	sulphate	sphalerite	ferrite	oxysulphate	sulphate
B2C-21	2.08	24.75	18.92	19.28	4.94	26.92	36.82	31.32
22	1.43	26.45	20.04	18.42	3.36	28.47	38.59	29.60
23	2.62	21.76	15.98	24.80	6.14	23.38	30.71	39.79
24	3.09	18.69	15.72	26.22	7.27	20.16	30.33	42.24
25	1.31	23.90	20.69	24.80	3.03	25.31	39.20	39.22
26	1.58	21.20	12.92	29.25	3.77	23.19	25.28	47.77
27	1.38	19.87	12.05	31.70	3.28	21.66	23.50	51.60
28	2.54	20.32	24.02	19.08	5.70	20.88	44.17	29.28
29	1.62	23.60	26.73	15.32	3.62	24.13	48.89	23.39
30	1.01	29.30	28.26	6.98	2.21	29.35	50.64	10.44
31	1.97	46.25	25.76	-	4.81	51.72	51.53	-
32	2.52	45.54	28.88	2.00	5.20	43.09	48.89	2.83
33	1.76	23.07	19.75	20.57	4.13	24.82	38.01	33.05
34	2.35	25.37	12.52	28.18	5.40	26.72	23.59	44.33
35	3.85	44.29	24.10	7.73	7.83	41.26	40.16	10.75
36	0.92	37.15	9.09	29.50	2.02	37.35	16.35	44.30
37	0.87	34.42	22.03	9.71	2.10	38.15	43.69	16.07
38	1.90	57.09	14.35	6.67	4.24	58.34	26.23	10.18
39	1.94	35.28	15.81	24.21	4.07	33.94	27.21	34.78
40	1.51	23.34	2.22	45.24	3.36	23.78	4.05	68.83

TABLE A1-4(b)

ZINC COMPOUNDS IN BRUNSWICK TAILINGS REFLOAT CALCINES
SERIES BMS

	Weight %				Zinc Distribution (%)			
	sphalerite	ferrite	oxysulphate	sulphate	sphalerite	ferrite	oxysulphate	sulphate
BMS-1	1.99	25.84	25.60	16.58	4.31	25.64	45.45	24.57
2	1.66	23.13	4.02	36.01	4.13	26.37	8.20	61.32
3	1.67	19.72	7.43	33.69	4.19	22.68	15.29	57.86
4	1.86	22.26	7.41	36.30	4.36	24.26	14.23	58.19
5	0.76	19.76	5.19	39.04	1.86	22.21	10.44	65.52
6	0.80	24.71	1.08	41.64	1.90	26.90	2.10	67.70
7	0.48	18.51	1.97	40.40	1.29	22.76	4.33	74.18
8	0.95	20.47	2.79	39.16	2.41	23.81	5.81	68.02
9	0.75	18.47	1.25	42.85	1.26	22.11	2.59	74.09
10	1.07	18.53	2.08	43.85	2.63	20.88	4.19	73.79
11	1.05	20.70	3.89	43.48	2.86	25.79	8.67	80.90
12	1.62	21.78	6.47	40.64	3.58	22.02	11.70	61.36
13	1.02	13.91	9.71	31.23	2.78	17.36	21.68	58.21
14	0.12	15.78	9.88	33.85	0.31	18.73	20.98	60.00
15	0.63	15.79	7.52	28.59	1.88	21.55	18.36	58.27
16	0.39	12.79	6.81	35.48	1.03	15.55	14.81	64.41

TABLE A1-4(c)

ZINC COMPOUNDS IN BRUNSWICK TAILINGS REFLOAT CALCINES
SERIES BMD

	Weight %				Zinc Distribution (%)			
	sphalerite	ferrite	oxysulphate	sulphate	sphalerite	ferrite	oxysulphate	sulphate
BMD-1	1.78	15.63	12.36	38.63	3.91	15.75	22.28	58.12
2	1.22	19.12	8.20	42.50	2.67	19.14	14.69	63.55
3	0.64	27.66	13.47	31.53	1.42	28.10	24.48	47.83
4	2.12	19.76	5.95	43.43	4.59	19.78	10.48	64.96
5	0.85	22.91	6.40	41.15	1.90	23.42	11.70	62.81
7	2.10	23.12	3.71	45.84	4.56	23.01	6.61	68.14
8	1.43	23.12	2.58	48.09	3.14	22.96	4.61	71.66
9	0.48	25.42	0.36	42.15	1.18	28.56	0.72	70.33
10	1.01	22.69	-	49.71	2.25	23.12	-	75.64
11	1.45	21.26	8.92	43.36	3.01	20.23	15.19	61.62
12	1.97	20.30	3.12	46.29	4.37	20.61	5.67	70.20
13	1.21	23.47	8.02	45.97	2.51	22.32	13.64	65.30
14	1.00	52.63	27.11	7.66	1.90	45.77	42.18	9.95
15	0.78	30.04	27.83	13.97	1.66	29.34	48.63	20.38
16	0.86	18.47	2.67	48.67	1.92	18.89	4.89	74.33
17	5.75	13.85	2.64	49.95	12.73	14.04	4.79	75.64
18	1.66	16.84	5.35	47.63	3.58	16.65	9.46	70.33
19	0.82	14.10	3.30	49.47	1.87	14.75	6.18	77.27
20	1.24	8.13	2.67	53.00	2.86	8.58	5.04	83.57
21	1.58	6.36	-	55.28	3.74	6.89	-	89.42
22	0.99	10.83	4.01	52.01	2.21	11.07	7.34	79.41

TABLE A1-4(c) (Cont'd)

ZINC COMPOUNDS IN BRUNSWICK TAILINGS REFLOAT CALCINES
SERIES BMD

	Weight %					Zinc Distribution (%)			
	sphalerite		ferrite	oxysulphate	sulphate	sphalerite	ferrite	oxysulphate	sulphate
	(1)	(2)							
BMD-23	1.02	-	13.37	4.46	49.93	2.27	13.63	8.13	76.01
24	1.83	-	9.47	9.81	47.22	3.94	9.33	17.29	69.47
25	2.07	-	10.84	21.85	31.25	4.47	10.72	38.67	46.17
26	0.91	-	23.86	41.01	7.05	1.81	21.74	66.87	9.59
27	2.03	-	25.46	33.43	12.43	4.09	23.52	55.25	17.15
28	2.00	-	14.89	22.74	6.23	4.17	14.21	38.82	8.88
29	1.66	1.50	16.44	31.20	18.18	3.52	15.96	54.18	26.35
30	1.39	1.20	20.46	35.21	15.36	2.78	18.73	57.67	48.14
31	1.02	0.75	12.60	33.43	21.08	2.10	11.88	56.37	29.67
32	0.91	1.35	14.57	32.10	27.31	1.73	12.70	50.06	35.55
33	1.79	1.56	16.88	25.86	27.69	3.61	15.58	42.70	38.16
34	0.96	1.05	13.57	19.17	34.77	2.06	13.32	33.67	50.97
35	0.91	0.90	14.46	31.65	21.98	1.88	13.66	53.48	31.00
36	1.72	1.20	17.37	36.11	13.85	3.53	16.35	60.80	19.47
37	0.92	0.90	14.31	24.07	27.97	1.99	14.15	42.59	41.31
38	1.36	1.50	17.31	30.31	18.02	2.93	17.07	53.49	26.54
39	1.54	2.05	27.06	36.55	6.83	3.17	25.54	61.71	9.63
40	1.34	1.50	19.47	34.77	11.44	2.88	19.14	61.16	16.80
41	1.88	1.20	11.28	10.92	40.18	4.32	11.89	20.59	63.24
42	1.20	1.05	10.92	6.68	44.49	2.85	11.88	13.01	72.30

TABLE A1-4(c) (Cont'd)

ZINC COMPOUNDS IN BRUNSWICK TAILINGS REFLOAT CALCINES
SERIES BMD

	Weight %					Zinc Distribution (%)			
	sphalerite		ferrite	oxysulphate	sulphate	sphalerite	ferrite	oxysulphate	sulphate
	(1)	(2)							
BMD-43	0.72	1.35	9.56	6.02	47.13	1.70	10.36	11.68	76.30
44	0.68	0.75	8.88	6.24	47.49	1.61	9.61	12.08	76.75
45	0.95	1.35	7.99	7.60	47.10	2.20	8.50	14.47	74.85
46	0.92	0.90	10.40	11.93	45.92	1.96	10.17	20.87	67.04
47	0.71	1.20	13.98	22.74	34.20	1.44	13.03	37.93	47.61
48	0.81	0.75	16.02	33.44	20.14	1.61	14.59	54.50	27.40
49	0.81	0.75	15.60	30.31	24.81	1.62	14.31	49.73	33.98
50	1.23	1.20	13.54	25.40	28.84	2.54	12.79	42.92	40.68
51	1.15	1.50	17.40	30.31	26.36	2.29	15.85	49.39	35.86
52	0.91	1.20	13.25	21.40	35.50	1.87	12.44	35.95	49.78
53	0.99	1.50	11.84	12.03	42.70	2.18	11.93	21.68	62.24
54	1.62	(1.05)	10.66	5.35	50.41	3.57	10.77	9.67	76.03
55	0.69	(0.60)	18.56	9.36	43.44	1.48	18.26	16.47	63.82
56	0.82	(1.20)	14.82	8.03	46.89	1.77	14.68	14.23	69.35
57	1.63	(2.10)	14.60	9.36	45.60	3.45	14.17	16.26	66.11
58	0.90	1.05	12.20	11.14	45.94	1.91	11.80	19.41	66.83
59	1.00	0.60	13.60	13.37	43.58	2.08	12.98	22.83	62.10
60	1.16	1.62	15.16	15.16	41.12	2.38	14.28	25.54	57.83
61	1.12	1.05	13.95	11.15	43.15	2.43	13.85	19.80	63.96
62	1.53	0.90	11.22	10.26	45.76	3.30	11.08	18.13	67.51
63	0.95	1.35	15.77	20.47	37.82	1.87	14.22	33.02	50.92
64	2.40	2.40	13.08	21.83	33.43	4.88	12.19	36.41	46.54

() values not considered due to stickiness of the bed
(1) values calculated from chemical analyses
(2) values determined from linear relationship of 111 sp reflection peak intensities to wt% sp in calcine (from chemical analyses)

Zinc extraction in a dilute acid leach (25 gpl H_2SO_4, 25°C and one hour retention) increased from 68% to 76% when the tip diameter was reduced from 4 to 1.5 mm (Figure A1-4).

The <u>distance of the gun nozzle</u> to the fluidized bed surface may also have some effect, although minor, on ferrite formation. Ferrite formation, on average, was found to be slightly reduced when the gun nozzle was positioned from 10 to 20 cm from the bed surface. However, the large scatter of the data made it difficult to establish a clear trend, either in zinc extraction in dilute acid leach or in ferrite formation (Figure A1-5).

As expected, the <u>roasting temperature</u> influences the formation of ferrite in calcines and zinc extraction in dilute acid, as shown in Figure A1-6 and Tables A1-3 and A1-4. Ferrite content in calcines dropped from 29.5 wt% at 700°C down to 22% at 650°C. However, this still represents only 72-73% Zn extracted in a dilute acid leach. Roasting at lower temperatures would achieve still higher zinc extractions but only at the expense of higher iron in solution, a condition to be avoided.

Although the <u>air factor</u>, defined arbitrarily as the theoretical air necessary to roast the concentrate to a desired composition over the actual air used, did influence the formation of ferrite, the minimum levels achieved were still unacceptably high, i.e. 22.4 wt% ferrite in calcines roasted at an air ratio of 1.75-1.85. This corresponds to 71.3% zinc extraction in a dilute acid leach (Figure A1-7).

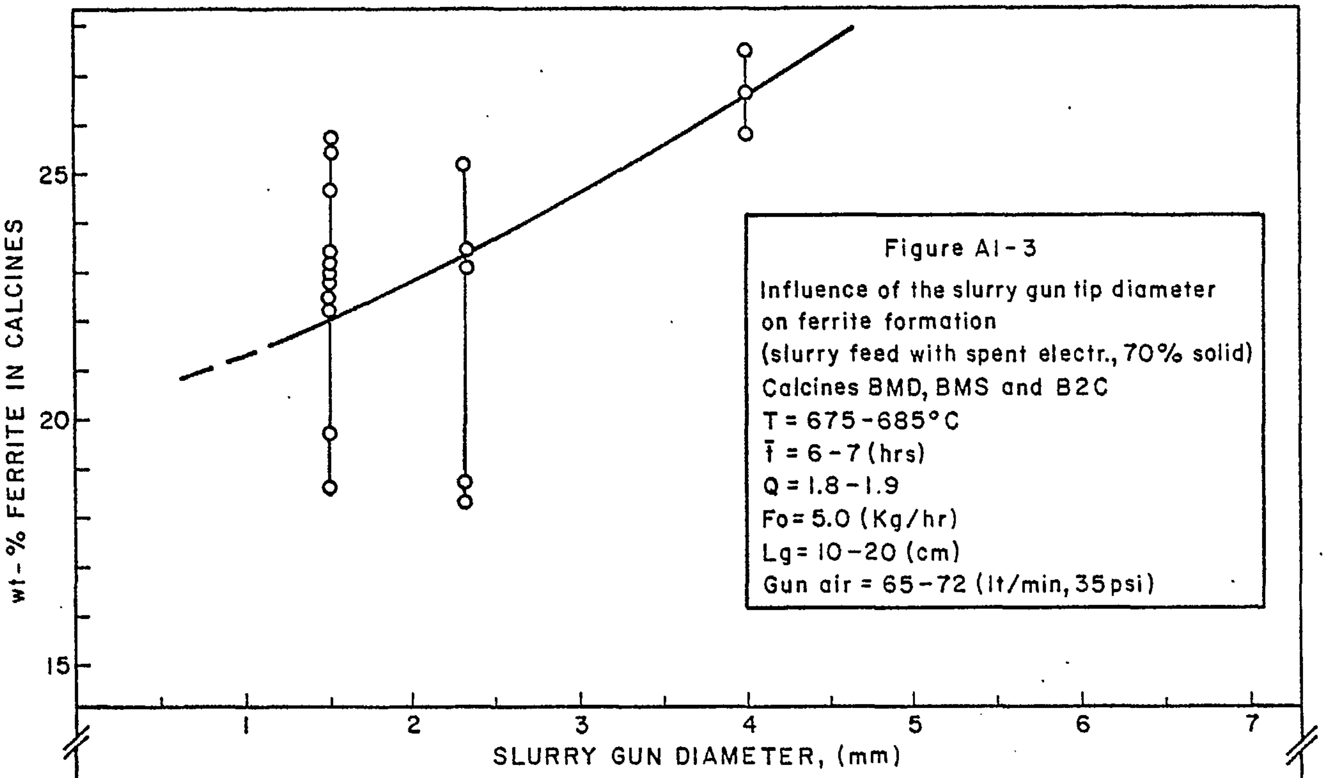


Figure A1-3

Influence of the slurry gun tip diameter on ferrite formation

(slurry feed with spent electr., 70% solid)

Calcines BMD, BMS and B2C

T = 675-685°C

$\bar{t}$ = 6-7 (hrs)

Q = 1.8-1.9

Fo = 5.0 (Kg/hr)

Lg = 10-20 (cm)

Gun air = 65-72 (lt/min, 35 psi)

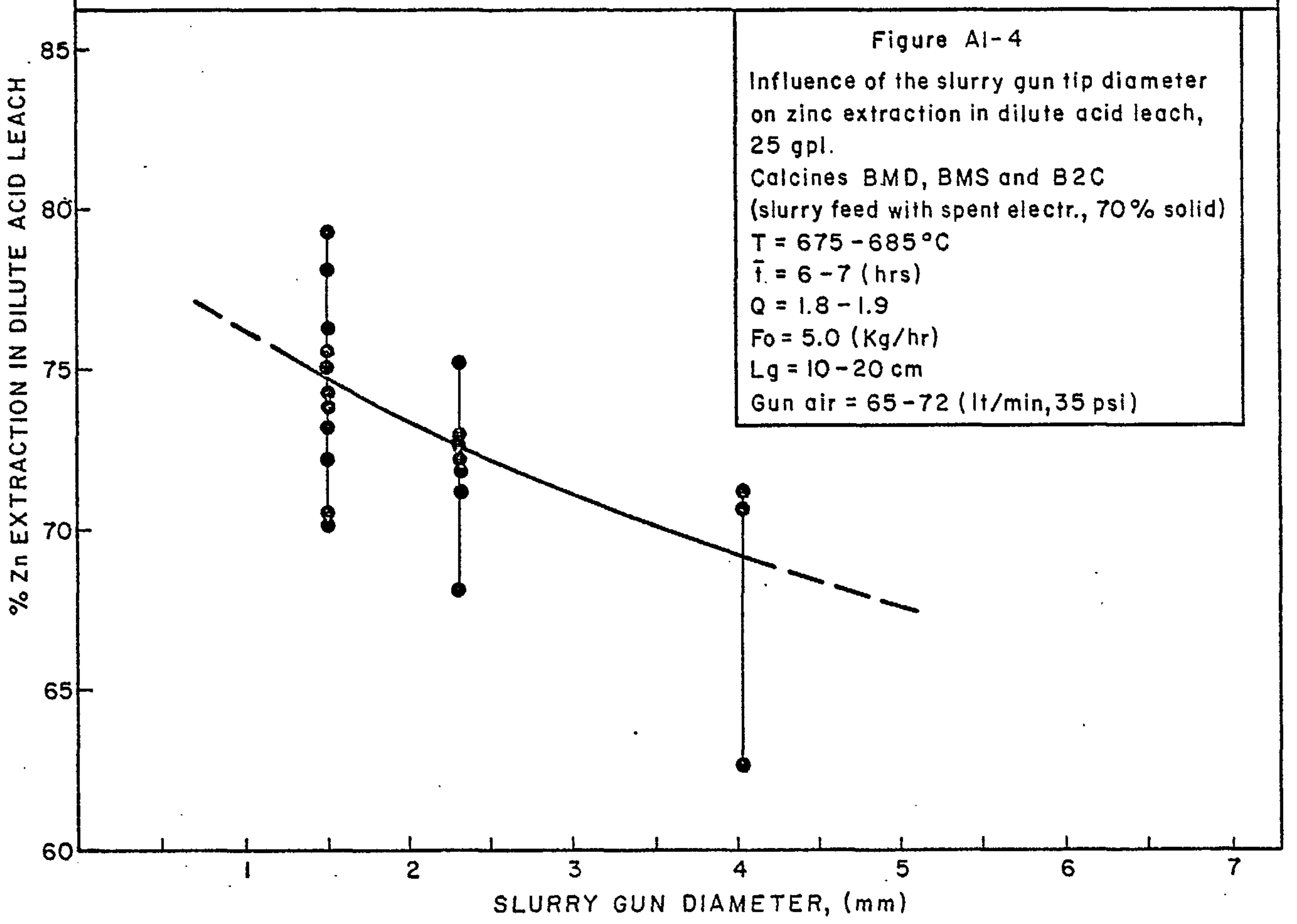


Figure A1-4

Influence of the slurry gun tip diameter on zinc extraction in dilute acid leach, 25 gpl.

Calcines BMD, BMS and B2C

(slurry feed with spent electr., 70% solid)

T = 675-685°C

$\bar{t}$ = 6-7 (hrs)

Q = 1.8-1.9

Fo = 5.0 (Kg/hr)

Lg = 10-20 cm

Gun air = 65-72 (lt/min, 35 psi)

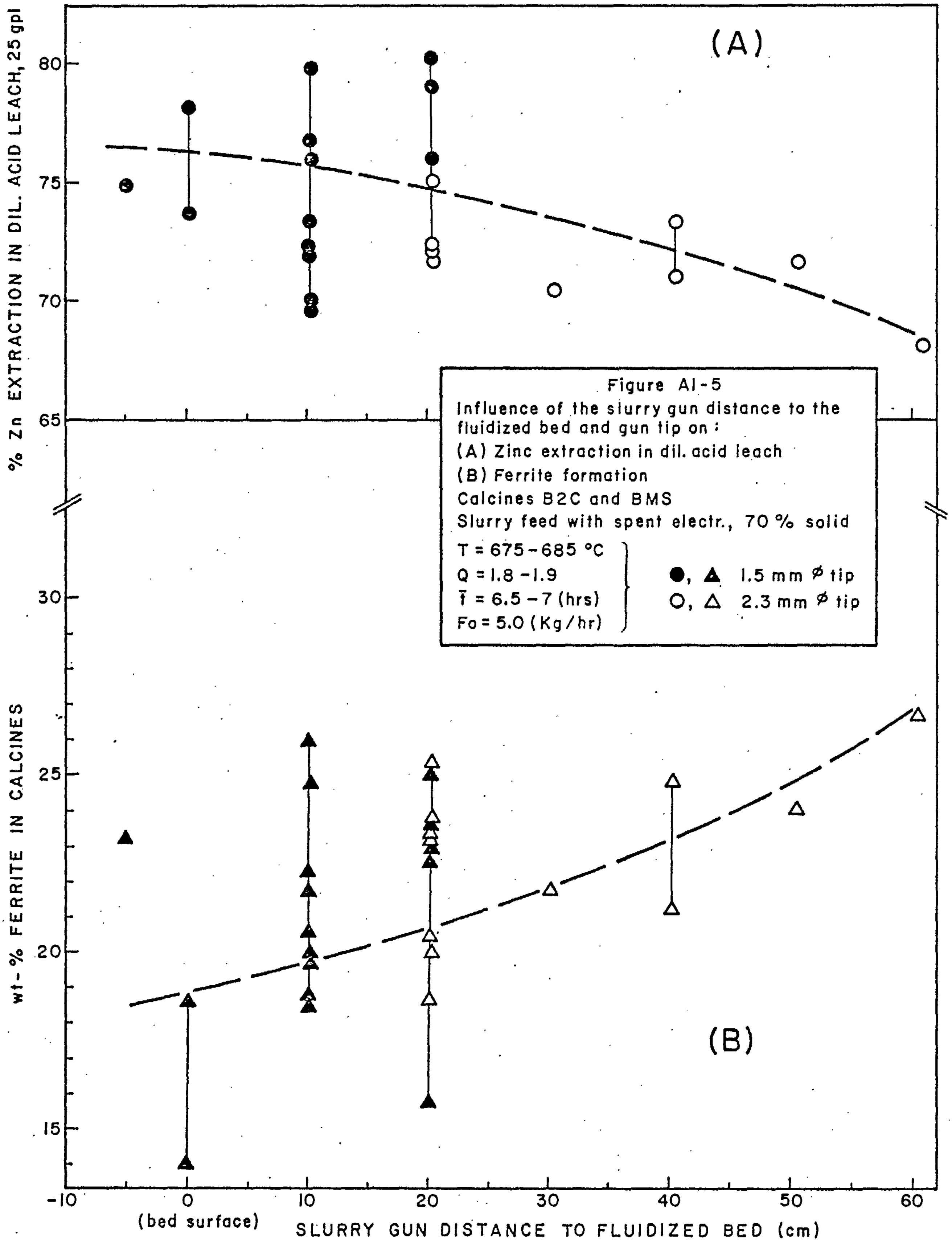


Figure A1-5

Influence of the slurry gun distance to the fluidized bed and gun tip on :
(A) Zinc extraction in dil. acid leach
(B) Ferrite formation
Calcines B2C and BMS
Slurry feed with spent electr., 70 % solid

T = 675 - 685 °C
Q = 1.8 - 1.9
$\bar{t}$ = 6.5 - 7 (hrs)
Fo = 5.0 (Kg/hr)

●, ▲ 1.5 mm ø tip
○, △ 2.3 mm ø tip

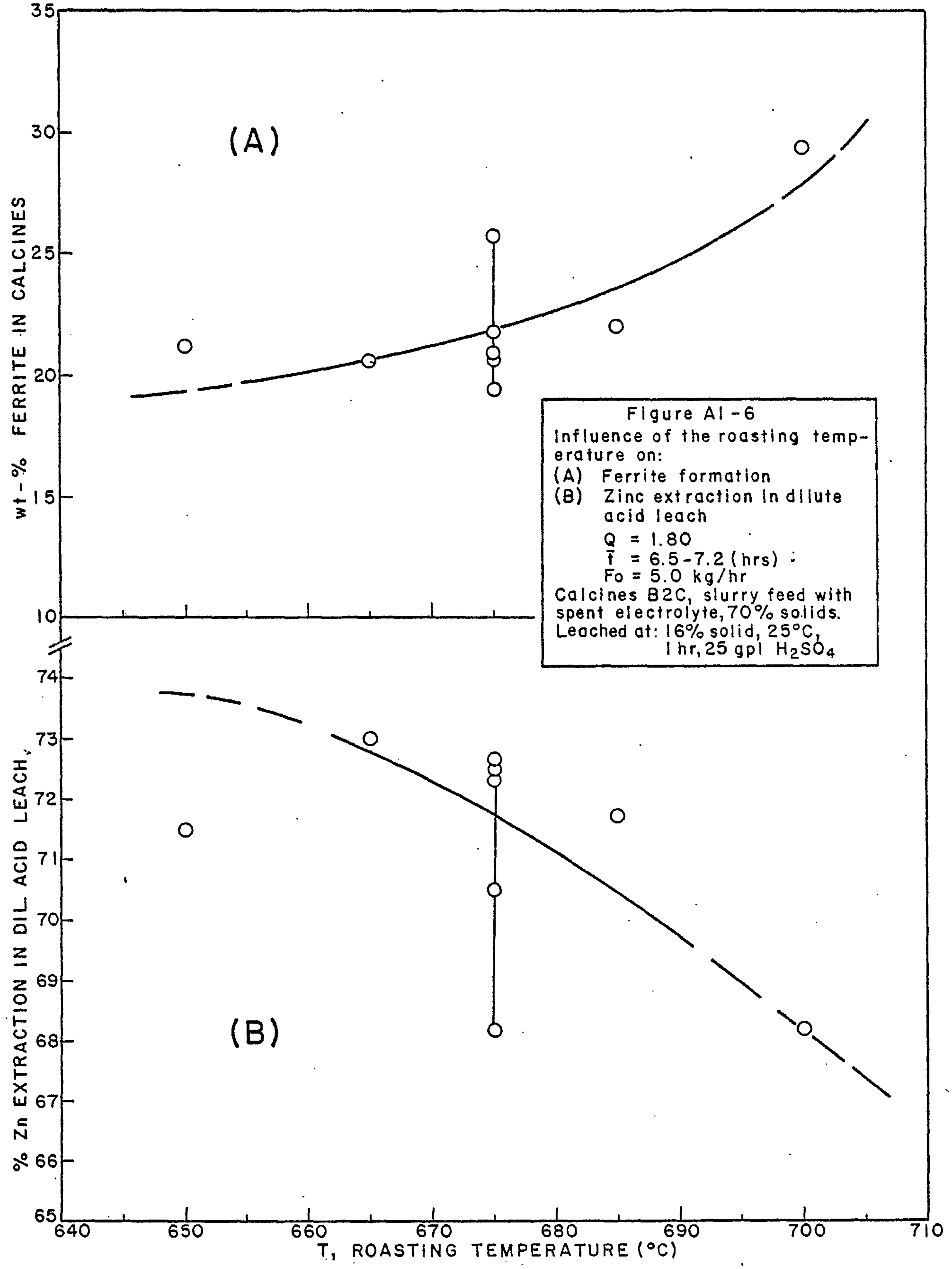


Figure A1-6
Influence of the roasting temperature on:
(A) Ferrite formation
(B) Zinc extraction in dilute acid leach
$Q = 1.80$
$\bar{t} = 6.5-7.2$ (hrs)
$F_o = 5.0$ kg/hr
Calcines B2C, slurry feed with spent electrolyte, 70% solids.
Leached at: 16% solid, 25°C, 1 hr, 25 gpl H_2SO_4

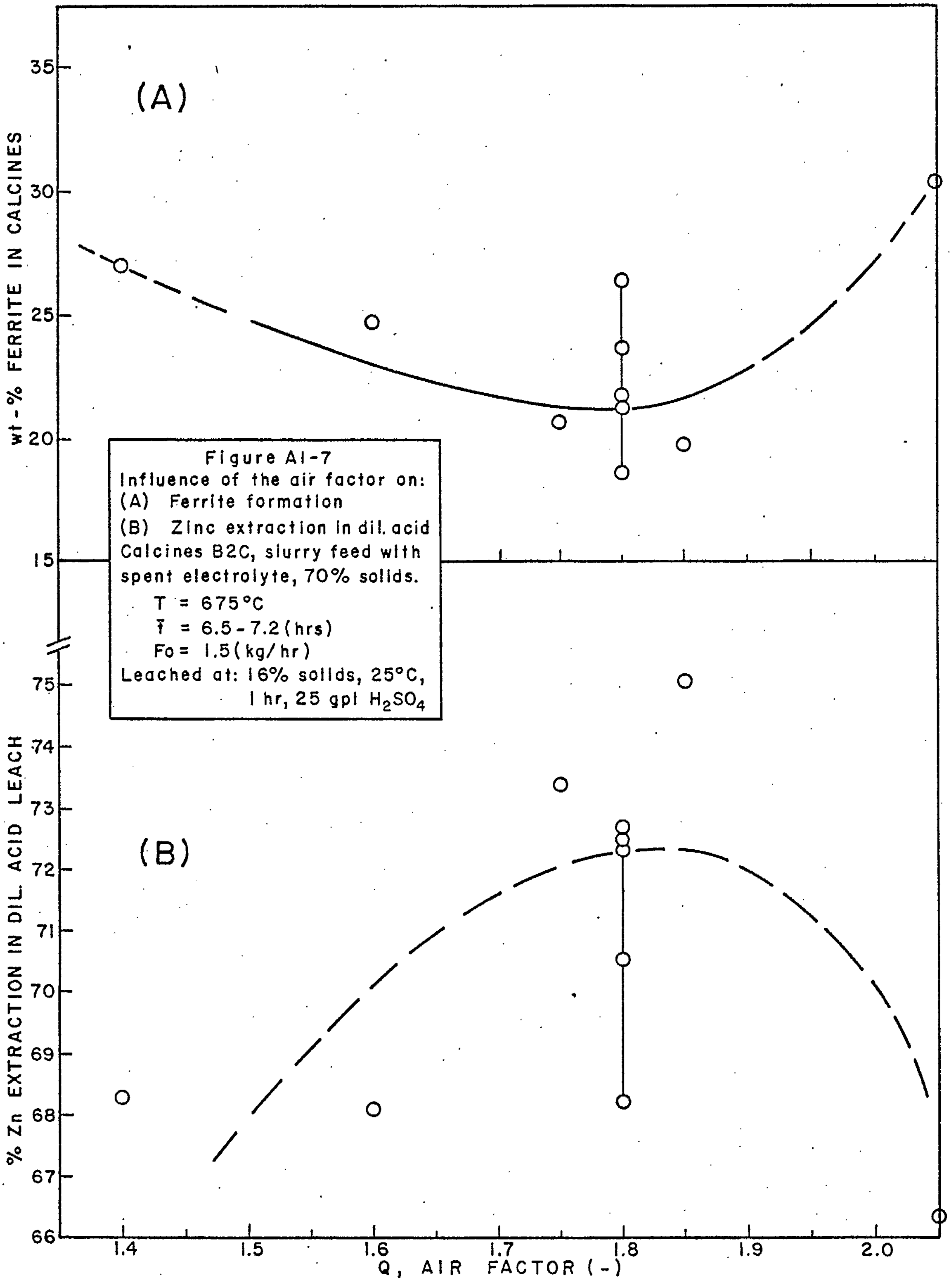


Figure A1-7
Influence of the air factor on:
(A) Ferrite formation
(B) Zinc extraction in dil. acid
Calcines B2C, slurry feed with
spent electrolyte, 70% solids.
T = 675°C
$\bar{\tau}$ = 6.5 - 7.2 (hrs)
Fo = 1.5 (kg/hr)
Leached at: 16% solids, 25°C,
1 hr, 25 gpl H_2SO_4

The addition of so-called "sulphation promoters" created stickiness problems and bed defluidization, without any apparent advantage in increased zinc extraction.

Five promoters were tested: Na_2SO_4, Na_2SO_4 plus bentonite, Na_2CO_3, KNO_3 and NaCl. In all cases, with the exception of the Na_2SO_4 and bentonite mixture, stickiness occurred which resulted in agglomeration of the bed. Na_2SO_4 added at 1 wt%, caused the bed to sinter and defluidize completely. The only encouraging results were achieved with the Na_2SO_4 and bentonite mixture in the range 0.5-0.75% Na_2SO_4 and 0.75%-2% bentonite (Figure A1-8). In this case zinc extraction in dilute acid leach approached 80%. Unfortunately, the bed did not fluidize well with these additives and the settling and filtering of the pulp was difficult due to the fine bentonite in suspension.

With all factors considered it was apparent that zinc extractions exceeding 75% could not be achieved consistently with a dilute acid leach of calcines produced from a slurry fed roaster. The problem was ferrite formation. The enhancement of ferrite was found to be related directly to the size of pellets formed from the slurry spray. Screened fractions of several calcines show a distinct correlation between pellet size and ferrite content as shown in Figure A1-9. Pellets over 2 mm in diameter contain from 25 to 35% ferrite, whereas the smaller pellets, which approach the mean particle size of individual compounds in the calcine (≃85% minus 325 mesh), contain from 12 to 15 wt% ferrite.

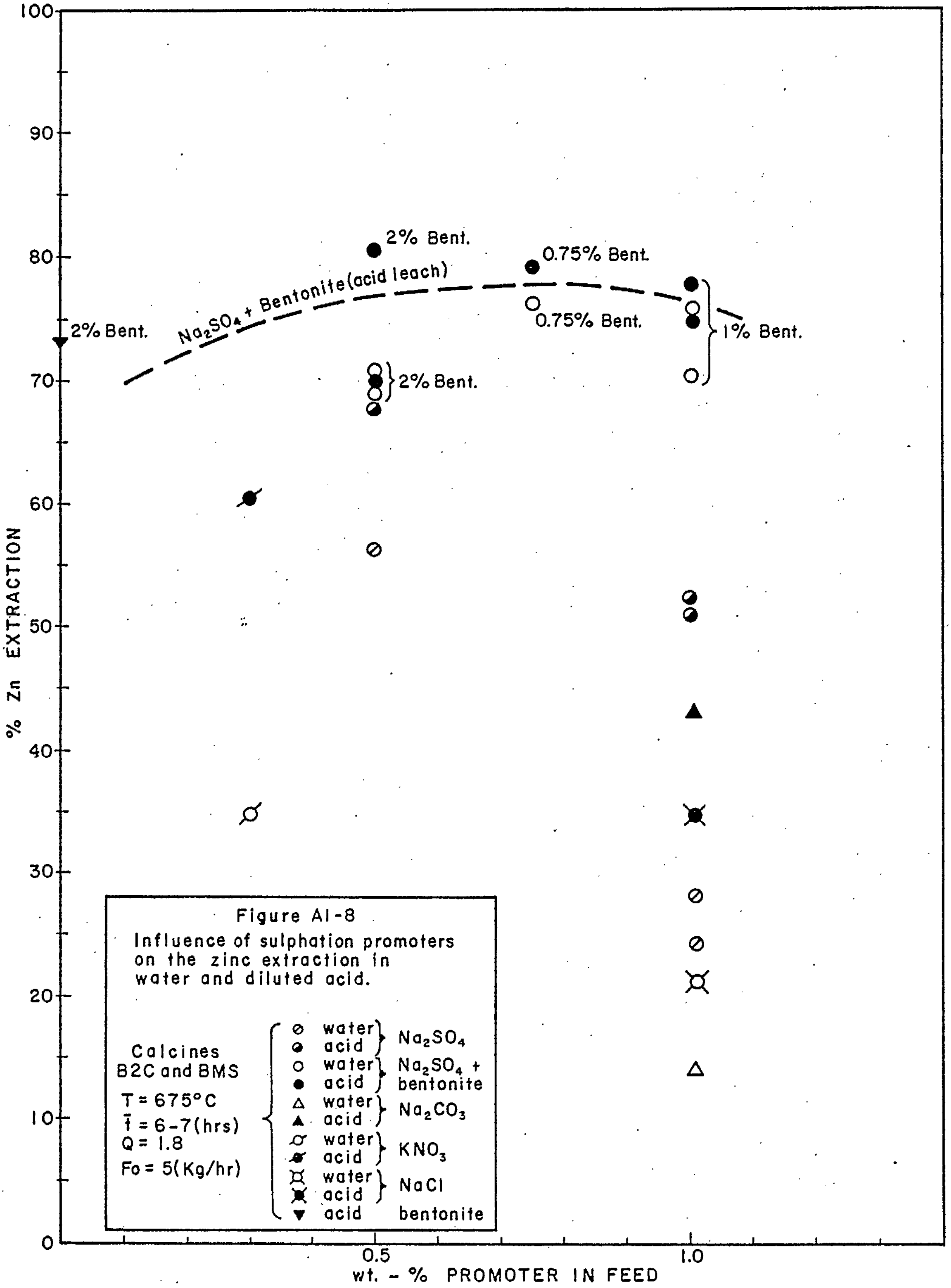


Figure AI-8
Influence of sulphation promoters on the zinc extraction in water and diluted acid.

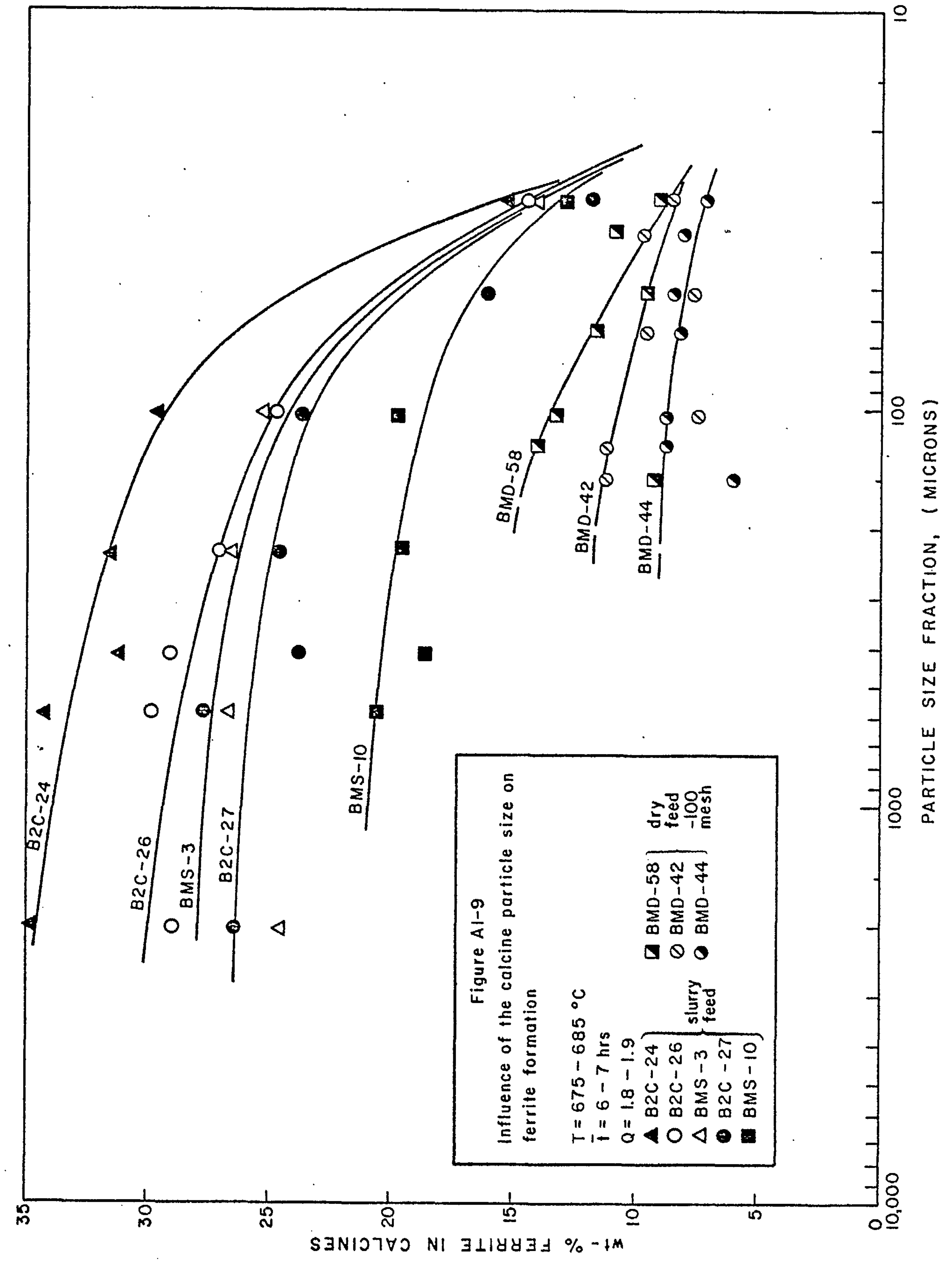


Figure AI-9

Influence of the calcine particle size on ferrite formation

The relationship between pellet size and ferrite content is most probably a function of the thermal cycle of the slurry droplets. Since all sulphation and iron oxidation reactions are highly exothermic, the temperatures within pellets can be significantly higher than the average bed temperature. Examples of these exothermic reactions for sphalerite and pyrite, the two principal minerals in the bed are as follows:

Sphalerite

$$(0.87\ ZnS + 0.13\ FeS) + 1.97\ O_2 \rightarrow 0.87\ ZnSO_4 + 0.06\ Fe_2O_3 + 0.12\ SO_2$$

$$\Delta H^\circ_{685^\circ C} = -177.72\ \text{Kcal/mole sphalerite}$$

Pyrite

$$FeS_2 + 2.5\ O_2 \rightarrow 0.5\ Fe_2O_3 + 2\ SO_2$$

$$\Delta H^\circ_{685^\circ C} = -197.29\ \text{Kcal/mole pyrite}$$

Reactions of sphalerite to yield oxysulphate also generate large amounts of heat ($\Delta H^\circ_{685^\circ C} = -136.46$ Kcal/mole sphalerite). Decomposition of sulphuric acid, on the other hand, is an endothermic reaction as follows:

$$H_2SO_4 \rightarrow H_2O + SO_2 + 0.5\ O_2$$

$$\Delta H^\circ_{685^\circ C} = +55.3\ \text{Kcal/mole sulphuric acid}$$

Since the molar ratio in the feed was (Zn,Fe)S: FeS_2: H_2SO_4 = 1:0.657:0.063, the overall balance is strongly exothermic, with a theoretical value of -303.86 Kcal of heat per mole of sphalerite present in the feed. In addition to this, other sulphides such as $CuFeS_2$ and PbS, although in minor amounts in the feed, also generate heat during sulphation.

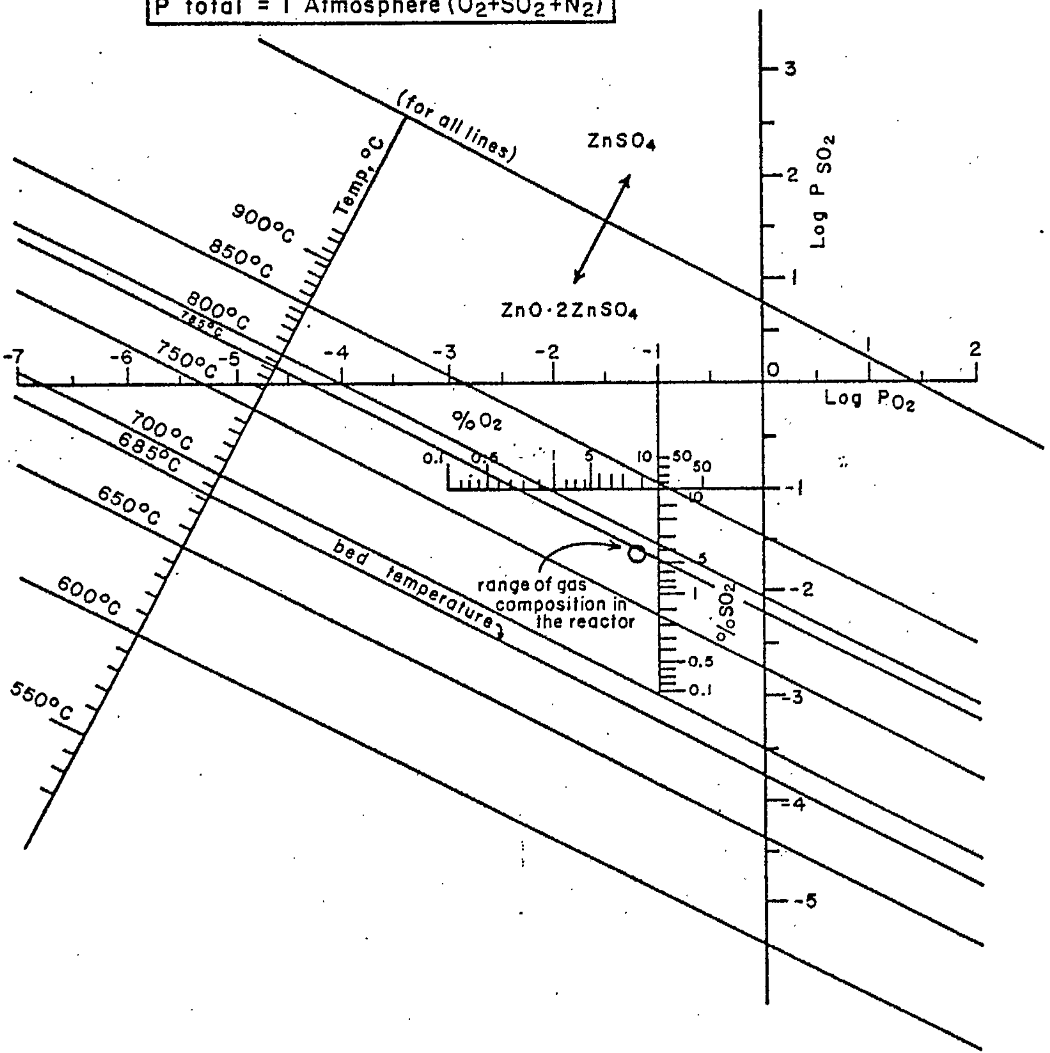
Since heat dissipation per unit surface area of pellet is inversely proportional to the pellet volume, large pellets will dissipate proportionally less heat per unit surface area than small pellets and their temperature therefore can be up to 100°C higher than the bed temperature (3). This temperature increase displaces the stability phase field from $ZnSO_4$ into the oxysulphate range (Figure A1-10): at the bed temperature (685°C) only $ZnSO_4$ is the stable phase, while a 100°C increase will form only $ZnO.2\,ZnSO_4$. Since the pellets should follow a non-isothermal "S" diagram of transformation, due to the thermal instability (during the early stages of oxidation in particular) the overall composition of the pellets will contain a mixture, in different proportions, depending upon the pellet size, of $ZnSO_4$ and $ZnO.2\,ZnSO_4$.

The presence of zinc oxysulphate which, although soluble in dilute acid, is undesirable since it reacts with hematite to form ferrite, following the solid state reaction:

$$ZnO.2\,ZnSO_4 + Fe_2O_3 \rightarrow ZnO.Fe_2O_3 + 2\,ZnSO_4$$

which has a free energy of formation of -27±5 Kcal at 685°C. The hard pellets formed using slurry feed, contain both $ZnO.2\,ZnSO_4$ and Fe_2O_3 in intimate contact, which greatly enhances the formation of ferrite.

Figure A1-10
Stability phase diagram for the system Zinc - Sulphur - Oxygen in the region of $ZnSO_4$ - $ZnO \cdot 2ZnSO_4$
P total = 1 Atmosphere (O_2+SO_2+N_2)



Since the pellet size could not be reduced below approximately 100% -40 mesh, even with the smallest gun tip of 1.5 mm, overall ferrite levels in the calcine could not be consistently held below 20 wt%. Zinc extractions exceeding 80% could only be achieved from these calcines with a hot acid leach to dissolve ferrite. However, this dissolution would also result in high iron in the leach solution and hence a requirement for a complex iron precipitation procedure such as the Jarosite Process prior to solution purification and zinc electrolysis. Since particle size analyses versus ferrite content indicated that a marked decrease in ferrite with the decrease in the particle size takes place, achieving 12 to 15 wt% ferrite for the size range approximating the mean particle size (≈85% minus 325 mesh) of the feed itself, it was decided to abandon the slurry feed approach in favour of dry feeding. This method of feeding, providing the feed is finely divided (minus 100 mesh), avoids the two problems of slurry feeding which contribute to high ferrite formation: 1) exothermic reactions in relatively large, hard pellets which result in the formation of $ZnO.2\ ZnSO_4$, and 2) intimate solid state contact of $ZnO.2\ ZnSO_4$ and hematite and reaction to form ferrite.

.2 Dry Feed System

Dry concentrates, containing 26-30% Zn, were screened to -100 mesh and fed to the reactor using a variable speed screw feeder and a pneumatic injection device. Since no liquid was pumped with the feed, the excess heat from the reactor was consumed by spraying water with the same gun previously used for slurry feeding.

Results of these tests are given in Table A1-3(b) and A1-4(b); (BMS 13 to 16), Tables A1-3(c) and A1-4(c); (BMD 18 to 64). A marked decrease in ferrite content is evident in these calcines compared with the previous ones produced by slurry feeding. In Figure A1-11 the effect of slurry feed and dry feed on ferrite formation between 660 to 700°C is illustrated. In all cases, calcines produced from dry feed contain, on average, about 50% less ferrite than those produced using the slurry feed. This represents an approximate 12% increase in zinc extraction in the dilute acid leach.

The zinc oxysulphate/ferrite ratios in calcines produced by dry feeding are higher than in calcines produced by slurry feeding (Figure A1-12). This is due to the more intimate association of zinc oxysulphate with hematite in the slurry pellets and hence the increased reaction of ZnO.2 ZnO.2 $ZnSO_4$ with Fe_2O_3 to form $ZnO.Fe_2O_3$ (ferrite). The large scatter for the slurry data on Figure A1-12 reflects the wide range of factors that affect the formation of ferrite when slurry feeding is used. The amount of ferrite in the calcines produced by dry feeding was less in all size fractions than it was in corresponding size fractions from slurry fed calcines. The fact that the slope (ferrite content/size fraction) of the curves in Figure A1-9 for the

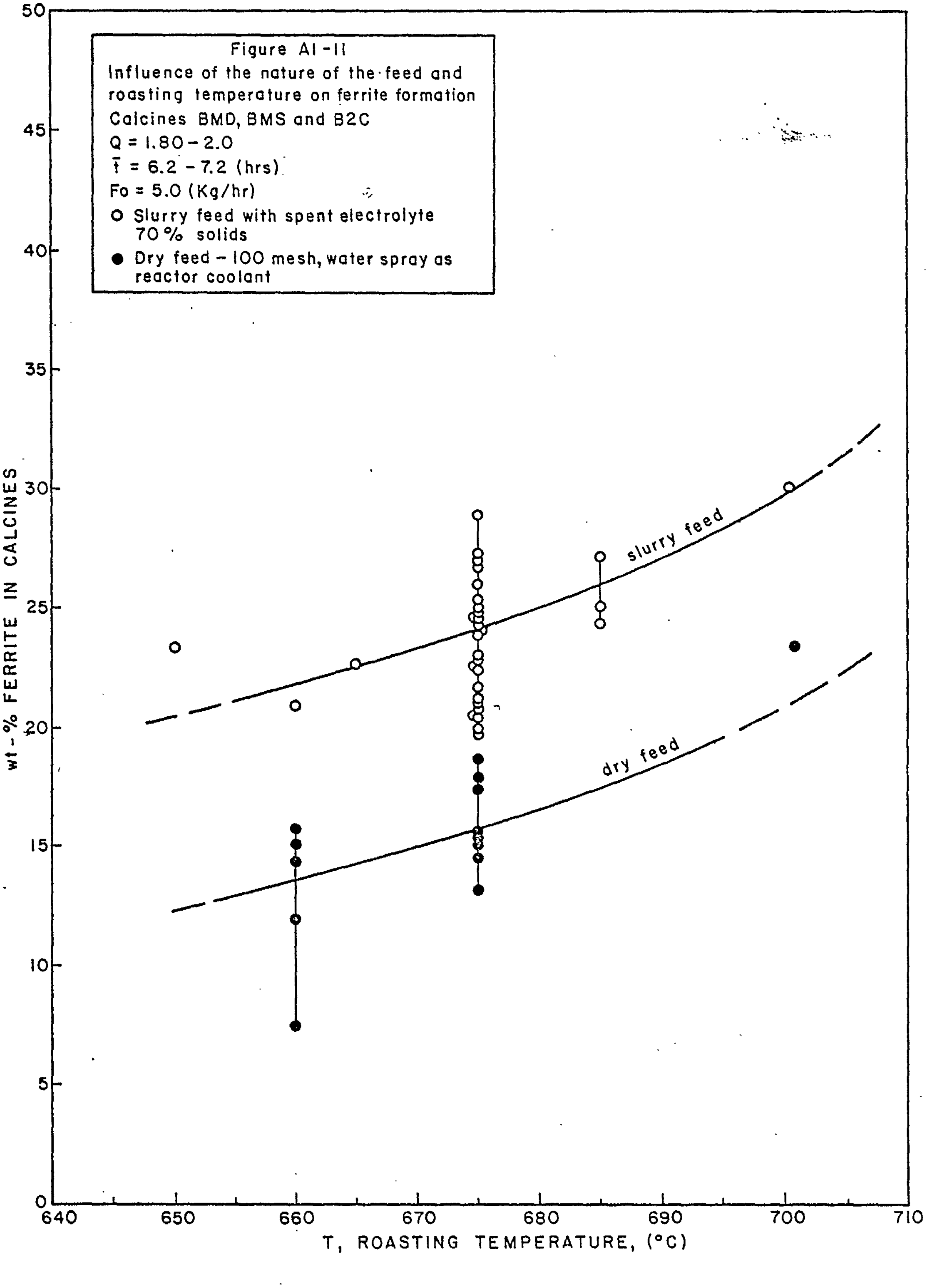


Figure AI-II
Influence of the nature of the feed and roasting temperature on ferrite formation
Calcines BMD, BMS and B2C
Q = 1.80 - 2.0
t̄ = 6.2 - 7.2 (hrs)
Fo = 5.0 (Kg/hr)
○ Slurry feed with spent electrolyte 70% solids
● Dry feed - 100 mesh, water spray as reactor coolant
slurry feed
dry feed
wt - % FERRITE IN CALCINES
T, ROASTING TEMPERATURE, (°C)
0
5
10
15
20
25
30
35
40
45
50
640
650
660
670
680
690
700
710

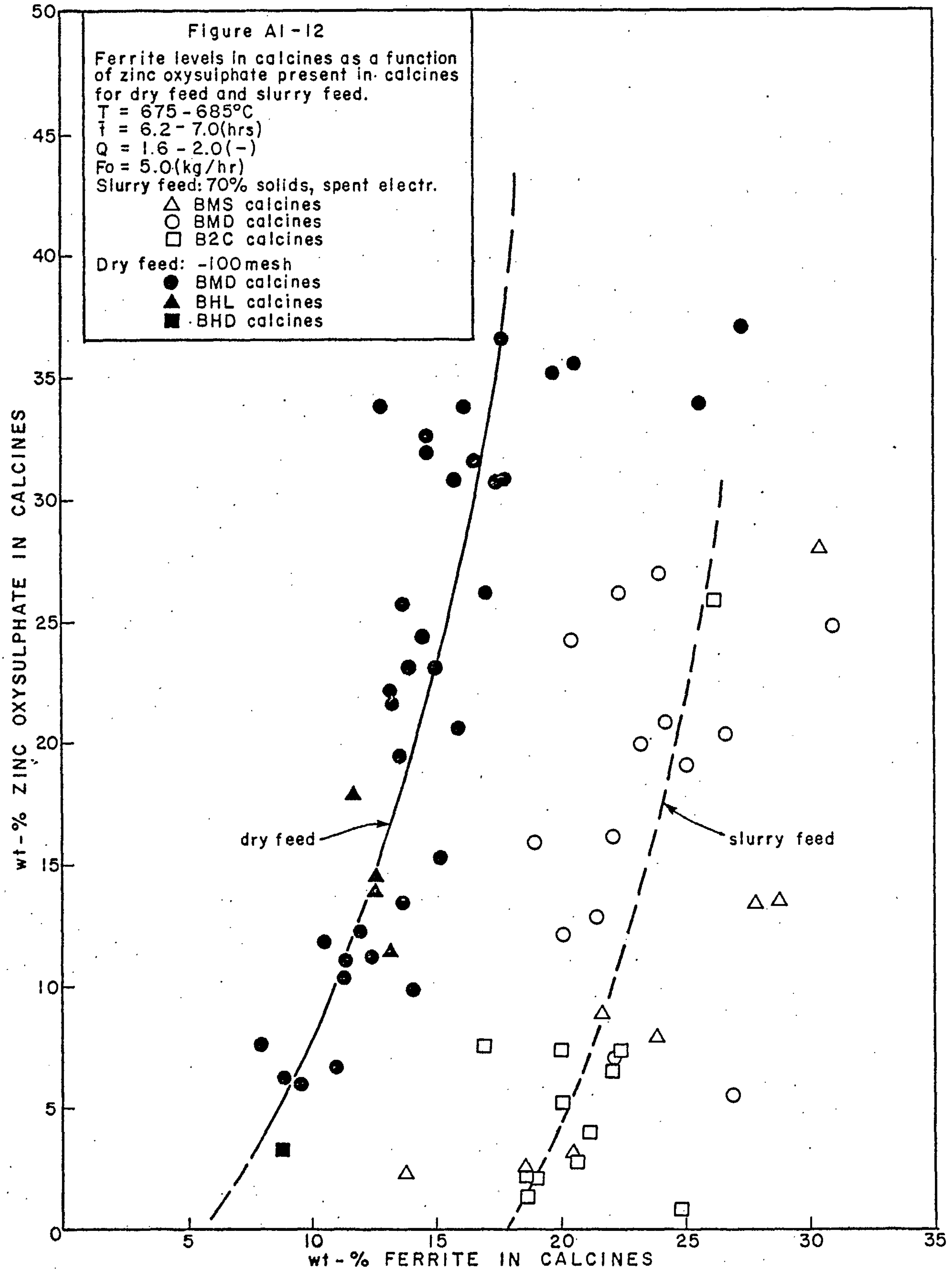


Figure A1-12

Ferrite levels in calcines as a function of zinc oxysulphate present in calcines for dry feed and slurry feed.

dry fed calcines is smaller than for the slurry calcines may indicate a limiting factor to which ferrite content can be reduced as a function of feed size. Even very fine fractions might contain as much as 5% ferrite. That dry fed calcines contain less ferrite than slurry fed calcines of the same size fraction might be a reflection of the relative compactness of the agglomerated particles. In the slurry feed case the bonding due to sulphate formation within slurry droplets is strong and the pellets were hard and dense, whereas the bonding of agglomerated particles in the dry feed is a product of drying prior to entering the reactor. As a result, this bonding is weaker and the agglomerates were less dense. The denser the pellet or agglomerate the better the chance for solid state reaction of zinc oxysulphate with hematite to form ferrite. On average, dry feed calcines contain about 68-70 wt% minus 100 microns, while the distribution of slurry feed calcines in the -100 micron fraction is only 15-30 wt% (Figure A1-13). The calcines produced from dry feed concentrates were also finer than the feed itself. This is due to the desintegration of the loose bonded agglomerates of -100 mesh feed in the reactor. Although the dry feed calcines were finer than the screened feed, they were nevertheless still coarser than the originally milled 100% minus 325 mesh concentrates, as shown in Figure A1-13.

Zinc extractions in a dilute acid leach versus <u>temperature</u> for the dry feed trials are compared with the slurry feed results in Figure A1-14. Zinc extractions using dry

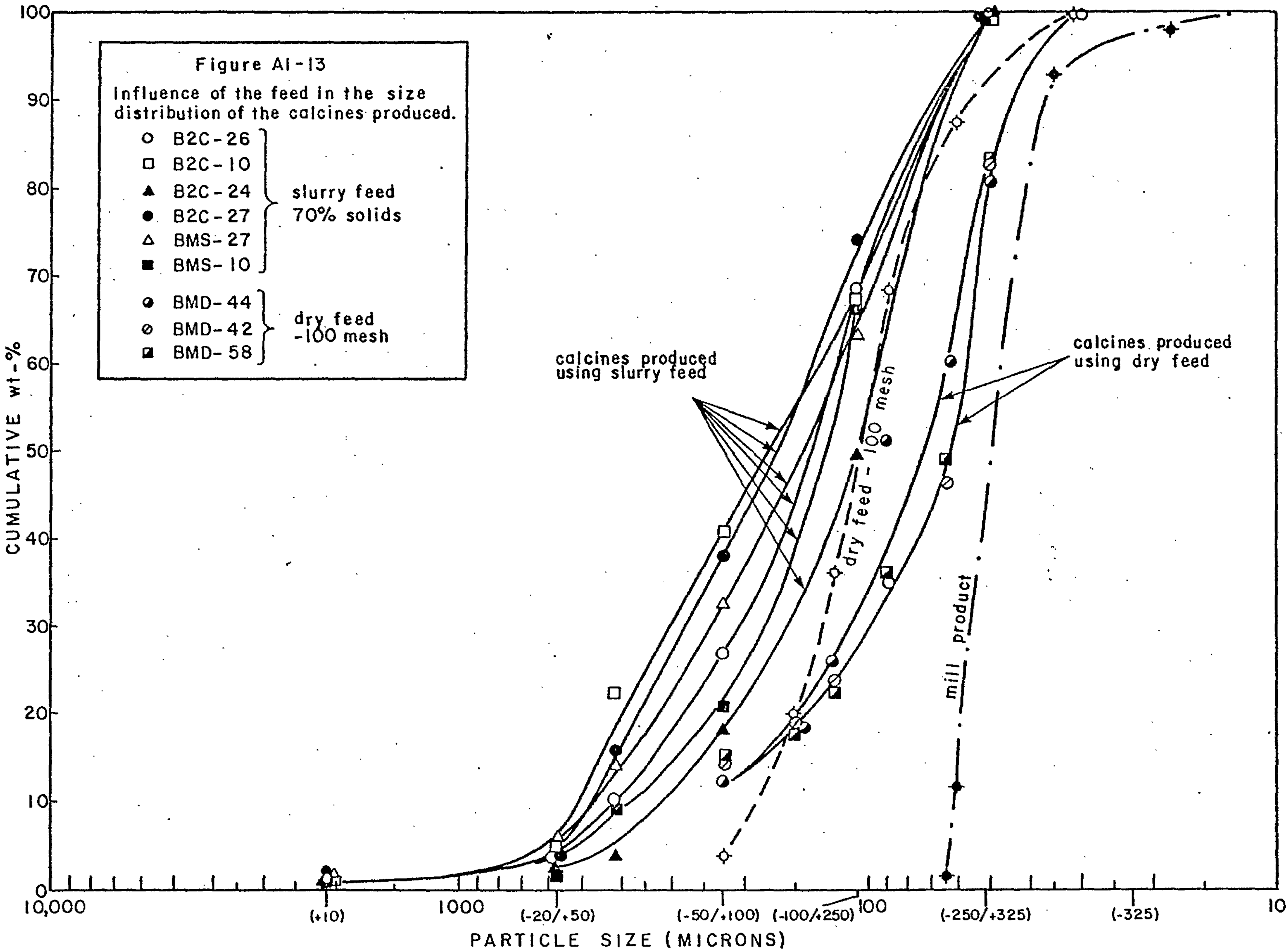


Figure A1-13

Influence of the feed in the size distribution of the calcines produced.

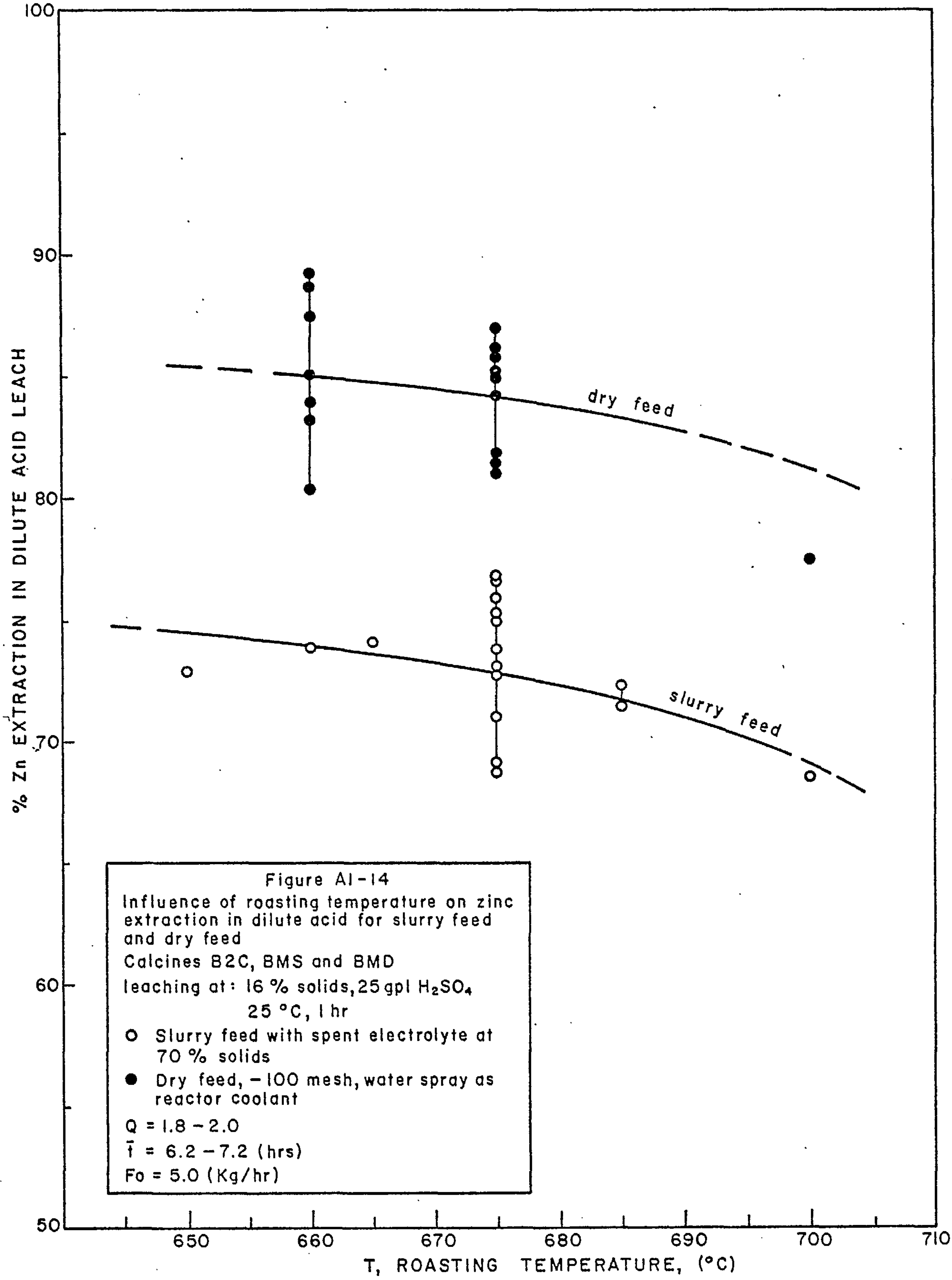


Figure A1-14

Influence of roasting temperature on zinc extraction in dilute acid for slurry feed and dry feed

Calcines B2C, BMS and BMD

leaching at: 16 % solids, 25 gpl H_2SO_4

25 °C, 1 hr

- ○ Slurry feed with spent electrolyte at 70 % solids
- ● Dry feed, −100 mesh, water spray as reactor coolant

Q = 1.8 − 2.0

$\bar{t}$ = 6.2 − 7.2 (hrs)

Fo = 5.0 (Kg/hr)

feeds were consistently 10 to 20% higher than with slurry feeds.

The air factor had a marked influence on ferrite formation in the dry feed system as shown on Figure A1-15. At Q = 1.3, calcines contained up to 27 wt% ferrite, whereas for Q = 1.9, the ferrite content dropped to approximately 13 wt%. Although the air factor does influence the SO_2/O_2 equilibrium inside the fluidized bed, it is suspected that the rather large effect noted may be due mainly to improved fluidization as Q increases. This effect is therefore largely a function of the size of the reactor: for example, the small 6-inch reactor trials (RPC Report M/77/209), using the same feed material, proper bed fluidization and minimum ferrite formation were achieved only at a relatively high air factor (Q = 3.2).

The influence of the air factor on zinc extraction in the dilute acid leach is shown in Figure A1-15B. For the range of air factor from Q = 1.7 to Q = 1.9, zinc extraction reached 83 to 85%.

The levels of ferrite achieved using dry feed, although significantly less than that achieved with the slurry feed system, are nevertheless still too high: in the best case, approximately 87% zinc was extracted in the dilute acid leach. For the process to be economically viable, a second stage strong acid leach must be employed to effect higher overall zinc extractions.

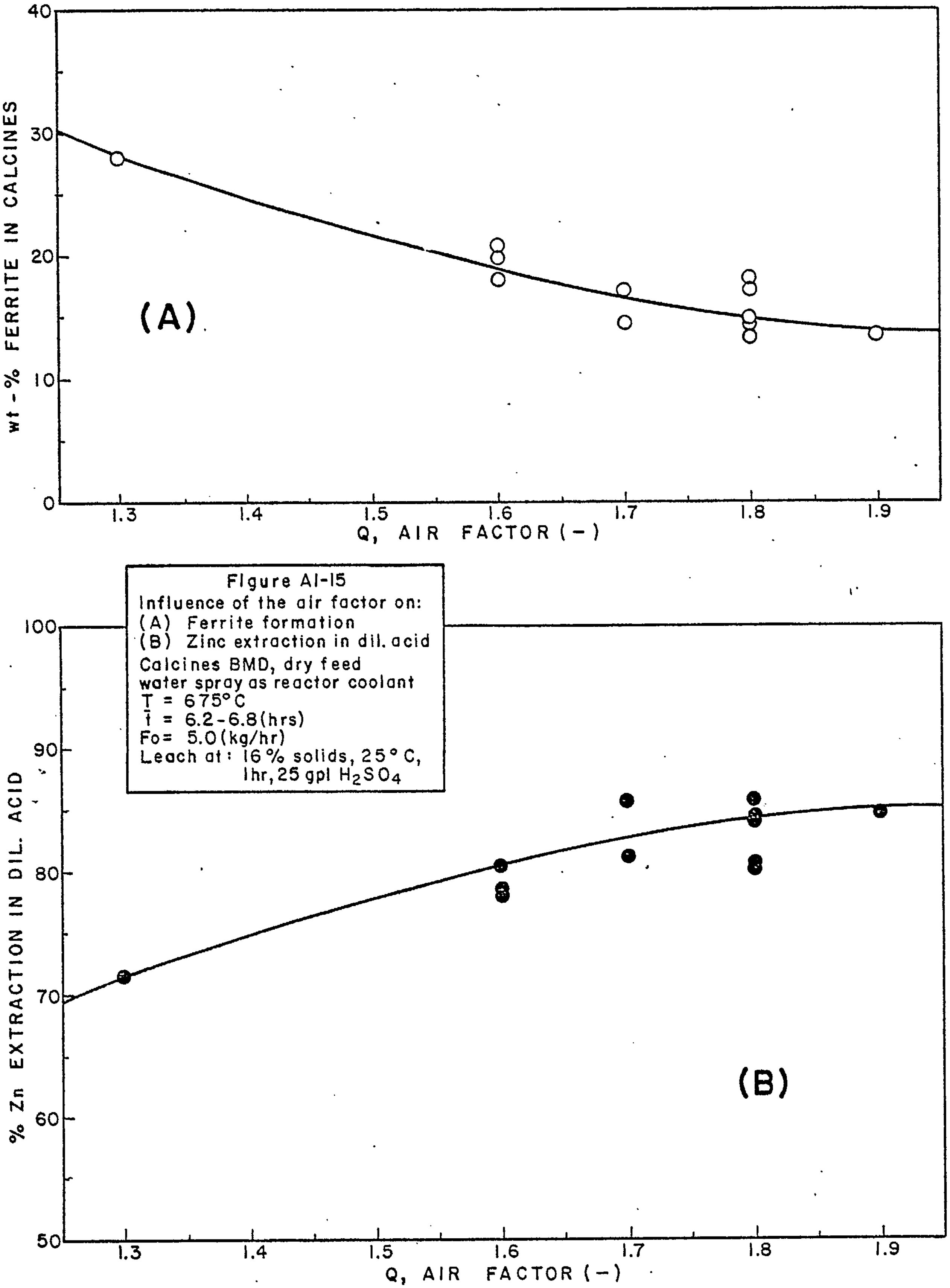


Figure A1-15
Influence of the air factor on:
(A) Ferrite formation
(B) Zinc extraction in dil. acid
Calcines BMD, dry feed
water spray as reactor coolant
$\underline{T}$ = 675°C
$\bar{t}$ = 6.2-6.8 (hrs)
Fo= 5.0 (kg/hr)
Leach at: 16% solids, 25°C,
1hr, 25 gpl H_2SO_4

.3 Hot Acid Leach Filtrate Recycle

The necessity of a second stage hot strong acid leach to dissolve the ferrite in the dilute (neutral) leach residue, creates the problem common to all zinc plants, i.e. high iron in the leach filtrate and hence the requirement for an iron elimination stage prior to solution purification and zinc electrolysis.

As a means of overcoming this iron problem, the possibility of decomposing the total hot acid leach filtrate in the fluid bed roaster was tested. As in the case of the spent electrolyte slurry trials, the hot acid filtrate forms pellets when sprayed as the coolant in the reactor. However, unlike the case with the slurry feed, the sum total of the decomposition reactions, i.e. iron sulphates to Fe_2O_3 and sulphuric acid to SO_2 are endothermic and therefore it was reasoned that the formation of zinc oxysulphate and therefore ferrite via the solid state reaction of oxysulphate with hematite would be minimal.

In the test program the residue from the dilute acid leach containing the ferrite, remnant sphalerite and the insoluble hematite was subjected to a strong acid leach at 95°C and constant 40 gpl H_2SO_4 for two hours. This was required to dissolve most of the sphalerite and virtually all of the ferrite in the residue. The resulting pulp was filtered and washed, and the final filtrate, with 70-120 gpl zinc and 80-130 gpl iron and 40 gpl free H_2SO_4 was thermally decomposed by spraying the solution inside the fluidized bed reactor consuming the excess thermal energy released from the oxidation and sulphation reactions.

At the roasting temperature of 685°C, water from the solution vaporizes readily, followed by the thermal decompoition of iron sulphates to Fe_2O_3 and SO_2+SO_3, and H_2SO_4 to SO_2+SO_3 and H_2O. These endothermic reactions are as follows:

$$Fe_2(SO_4)_3 \xrightarrow{\text{heat}} Fe_2O_3 + 3\ SO_2 + 1.5\ O_2$$

$$FeSO_4 \xrightarrow{\text{heat}} 0.5\ Fe_2O_3 + SO_2 + 0.25\ O_2$$

$$H_2SO_4 \xrightarrow{\text{heat}} SO_2 + H_2O + 0.5\ O_2$$

$ZnSO_4$ and $CuSO_4$, which are thermodynamically stable at 685°C under the gas composition (4.5-5% SO_2; 9-12% O_2) inside the roaster remain as such in the small porous pellets formed during the spraying of the solution, and act as binder for the hematite formed from the decomposition of the iron sulphates. The pellets formed by decomposition of the hot acid leach solution (HAL soln.) are continuously mixed with the calcine inside the fluidized bed thereby increasing the size distribution in the total bed material and hence considerably improving the fluidization properties of the calcine.

Results of 24 runs using the integrated HAL filtrate recycle into the reactor are given in Table A1-3(c). The average zinc extraction values are plotted as a function of the roasting temperature in Figure A1-16. For both cases, using water or the HAL solution as coolant, primary zinc extractions based on fresh feed are at a maximum at 660° (lowest temperature tested). Zinc extractions decrease

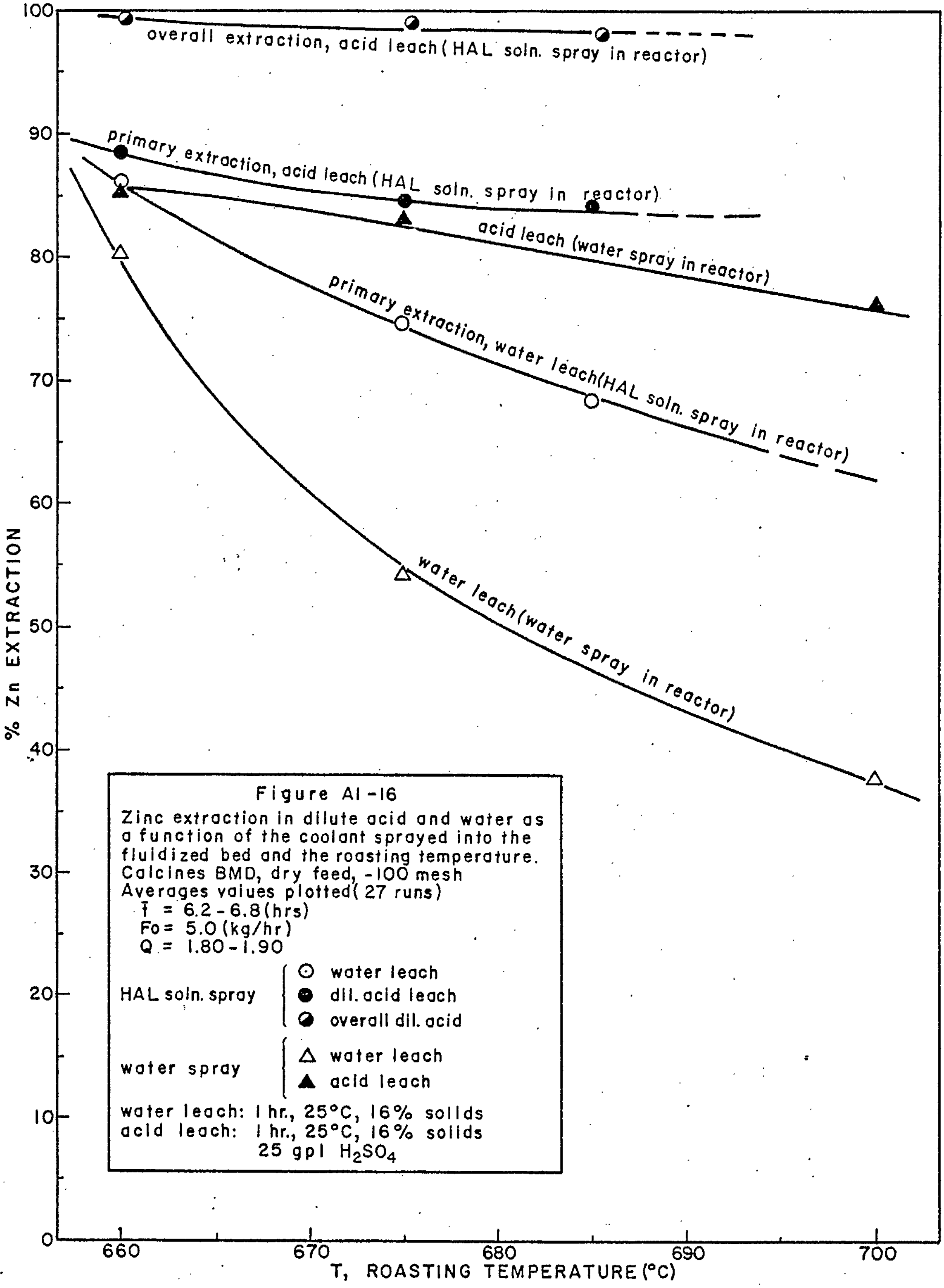


Figure AI-16

Zinc extraction in dilute acid and water as a function of the coolant sprayed into the fluidized bed and the roasting temperature.
Calcines BMD, dry feed, -100 mesh
Averages values plotted (27 runs)
$\bar{t}$ = 6.2 - 6.8 (hrs)
Fo = 5.0 (kg/hr)
Q = 1.80 - 1.90

water leach: 1 hr., 25°C, 16% solids
acid leach: 1 hr., 25°C, 16% solids
25 gpl H_2SO_4

slightly as the temperature increased up to 700°C. As an average, it was found that when the HAL solution was sprayed, dilute acid leach primary extraction of zinc was from 1.5 to 3% higher than when water was used as the coolant. Water soluble zinc ($ZnSO_4$) in the calcine was from 6 up to 23% higher for the HAL solution than for the water coolant. This increase in water soluble zinc ($ZnSO_4$) is due to the 15 to 30% higher SO_2+SO_3 concentration achieved inside the bed when the HAL solution was recirculated.

The overall zinc extraction, including the zinc recovered from the HAL solution after decomposition, was in all cases over 96%, being as high as 99% at 660°C.

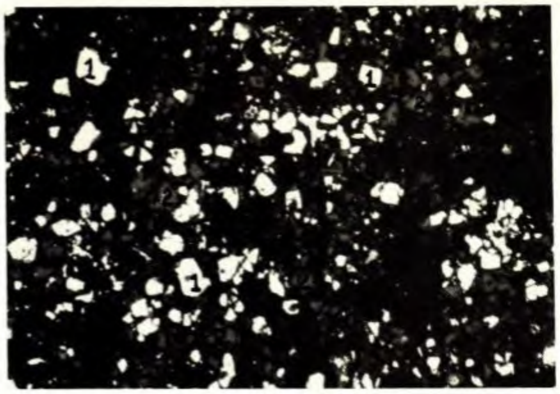
The pellets formed from the decomposition of the HAL solution were, in general, light weight and porous with a reddish hematite color. Their resistance to attrition in the bed is good which is presumably due to the strong binding effect of the $ZnSO_4$. Photomicrographs of feed, calcine, neutral leach residue, strong acid leach residue and pellets formed from the hot acid leach filtrate recycle are shown in Plate 3.

The pellets contain up to 20 wt % ferrite principally as irregular coatings on the outer surfaces or in vugs in the porous pellets. As the relatively low temperatures within the pellets, a result of the endothermic decomposition reactions, would apparently preclude the formation of zinc oxysulphate and hence ferrite, and as spray tests of the filtrate in a 1" diameter reaction tube at 685°C failed to produce either the oxysulphate or ferrite, an explanation of the presence of ferrite in the pellets is required.

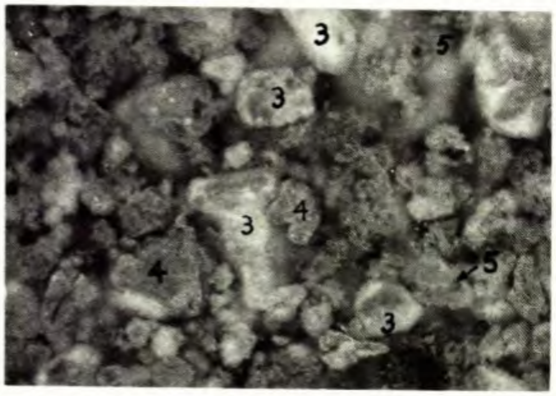
The amount of ferrite in pellets, as indicated by ferrite/hematite x-ray diffraction peak ratios from x-ray analyses of pellets, increases proportionally with the amount of zinc oxysulphate in the calcine (Figure A1-17). This fact and the observation that the small but denser pellets formed when pure water was sprayed as a coolant also contained ferrite with the same correlatable relationship with oxysulphate content in the calcine (open circles, Figure A1-17), leads to the conclusion that ferrite does not form directly from the HAL solution. Instead, the evidence would indicate that ferrite in pellets results from a pick-up of dust containing oxysulphate and hematite from the bed.

The dust pick-up is due to the short period of free-fall of the HAL solution droplets from the gun tip down to the fluidized bed surface. The water vaporization rate from the HAL solution droplet may follow a decaying exponential function (4), where at the beginning, steam evolves at a constant fast rate from the surface of the droplet (at an equivalent linear velocity between 10 to 20 m/sec) which precludes any contact of the droplet with solids from the bed or elutriated dust present in the freeboard of the reactor during the free-falling period. However, once the water content in the droplets has fallen below about 10% of the initial content, steam evolution decreases rapidly which then permits dust adhesion to the droplet surface. This will produce, at the end of the period, a hardened crust

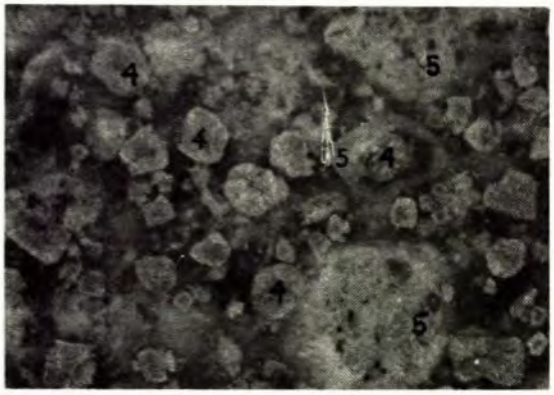
PLATE 3
ROASTER AND LEACH PRODUCTS



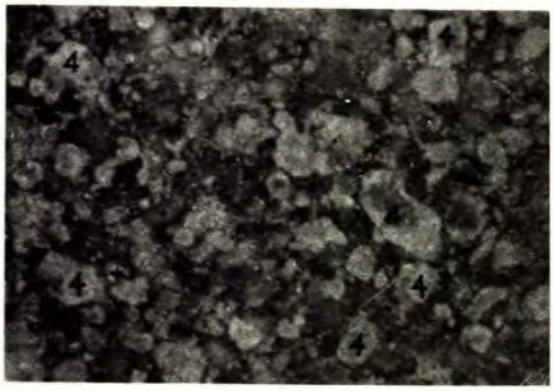
Feed, 1 cm = 63μ



Calcine, 1 cm = 63μ

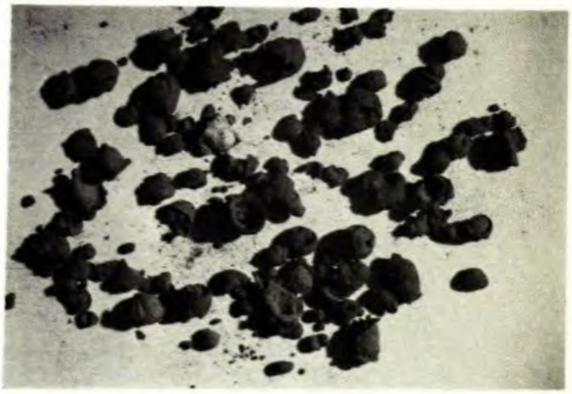


Neutral Leach Residue, 1 cm = 63μ



Strong Acid Leach Residue, 1 cm = 63μ

Calcine, 1 cm = 7 mm



Pellets, 1 cm = 7 mm

1 - pyrite

2 - sphalerite

3 - Zn sulphate-oxysulphate

4 - hematite

5 - ferrite

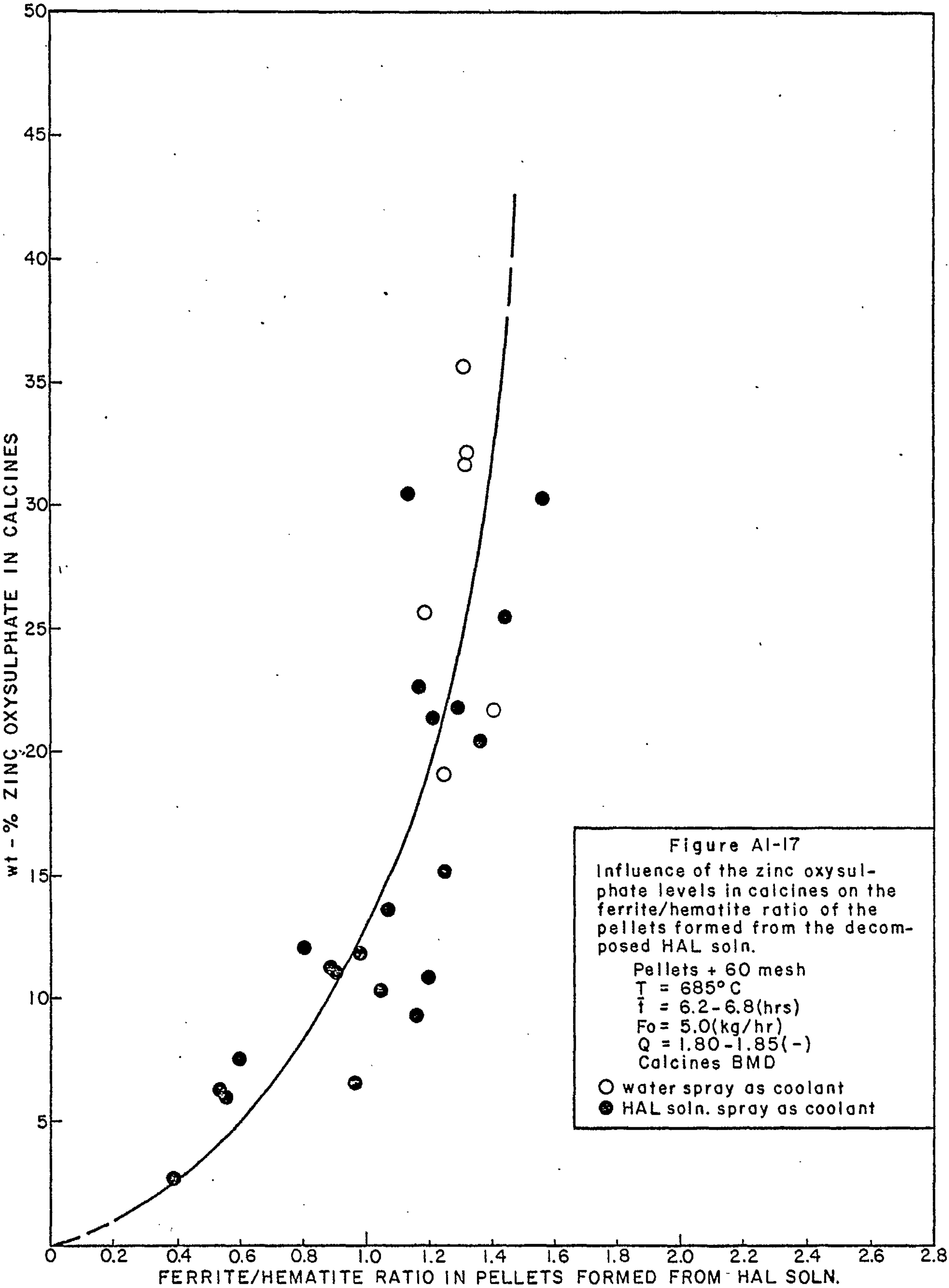


Figure AI-17
Influence of the zinc oxysulphate levels in calcines on the ferrite/hematite ratio of the pellets formed from the decomposed HAL soln.

Pellets + 60 mesh
$\bar{T}$ = 685°C
$\bar{t}$ = 6.2-6.8(hrs)
Fo= 5.0(kg/hr)
Q = 1.80-1.85(-)
Calcines BMD

○ water spray as coolant
● HAL soln. spray as coolant

(due to the $ZnSO_4$ and $ZnO.2\ ZnSO_4$ present in the calcine) which greatly enhances the reaction between $ZnO.2\ ZnSO_4$ picked up from the dust with Fe_2O_3, to form ferrite.

In a commercial roaster, with the gun for the HAL solution installed at the top or sides of the reactor from 3 to 12 meters from the bed surface (instead of 20 cm as in the pilot unit), it would be expected that the HAL solution droplets will dry completely in the freeboard where dust concentrations are minimal before reaching the bed and therefore avoid the ferrite formation due to dust pick-up from the bed.

The <u>air factor</u>, as noted for the water cooled dry feed trials, influences ferrite formation and hence primary (from the fresh feed, excluding recycle) zinc extraction. Primary zinc extraction in dilute acid and the ferrite content of calcines as a function of the air factor are shown in Figures A1-18 and A1-19 respectively. An increase in the air factor from Q = 1.8 to Q = 2.2 resulted in an increase in primary zinc extraction of approximately 5% (from 84 to 89%). This corresponds to a decrease in ferrite content in the calcines from 14.5 wt% to 8 wt%. Again it is stressed that this air factor effect appears to be mainly a reflection of improved fluidization properties of the bed and is in reality a function of the gas velocity in the reactor rather than the arbitrary value Q.

An increase in Q also has a negative affect on the SO_2+SO_3 strength in the off-gases. Figure A1-20 is a plot of Q versus measured values of SO_2+SO_3 determined with a

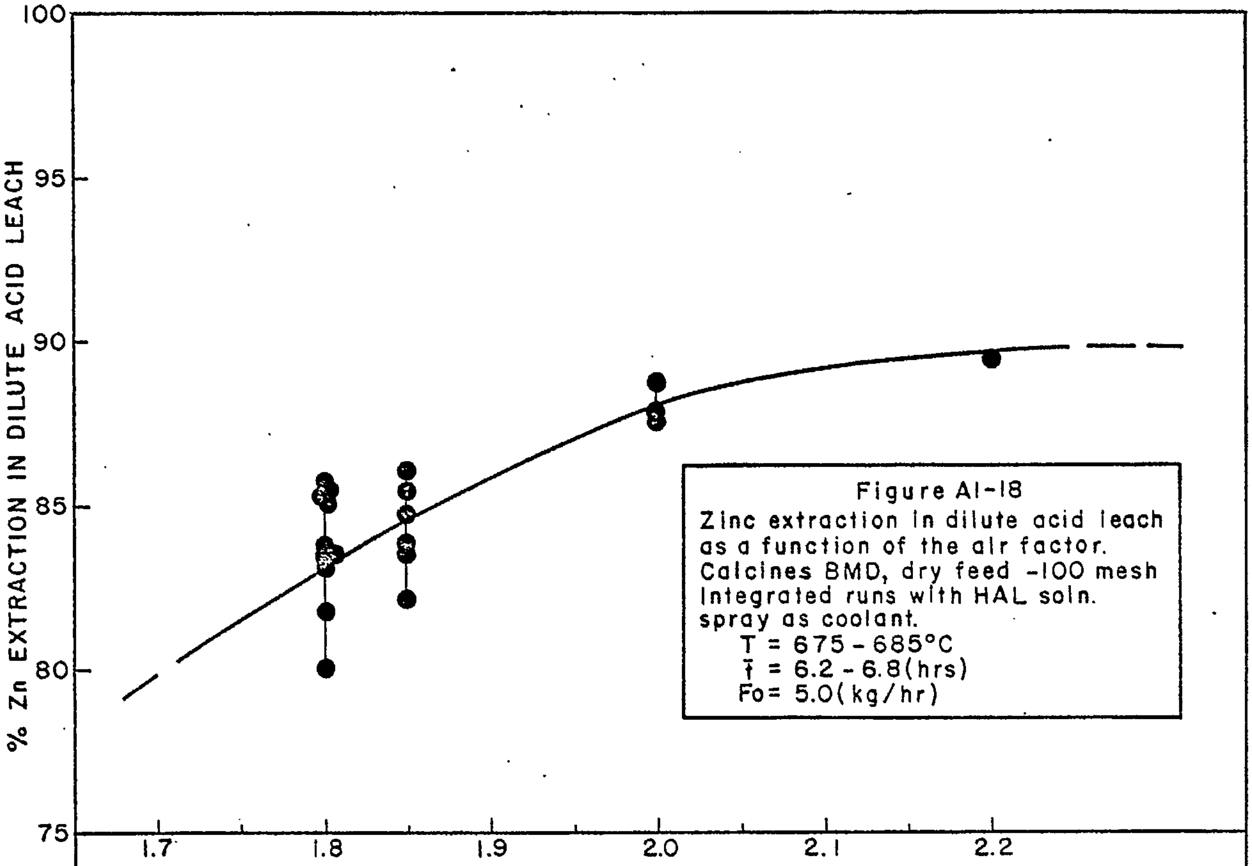


Figure A1-18
Zinc extraction in dilute acid leach as a function of the air factor.
Calcines BMD, dry feed -100 mesh
Integrated runs with HAL soln. spray as coolant.
T = 675 - 685°C
$\bar{t}$ = 6.2 - 6.8 (hrs)
Fo = 5.0 (kg/hr)

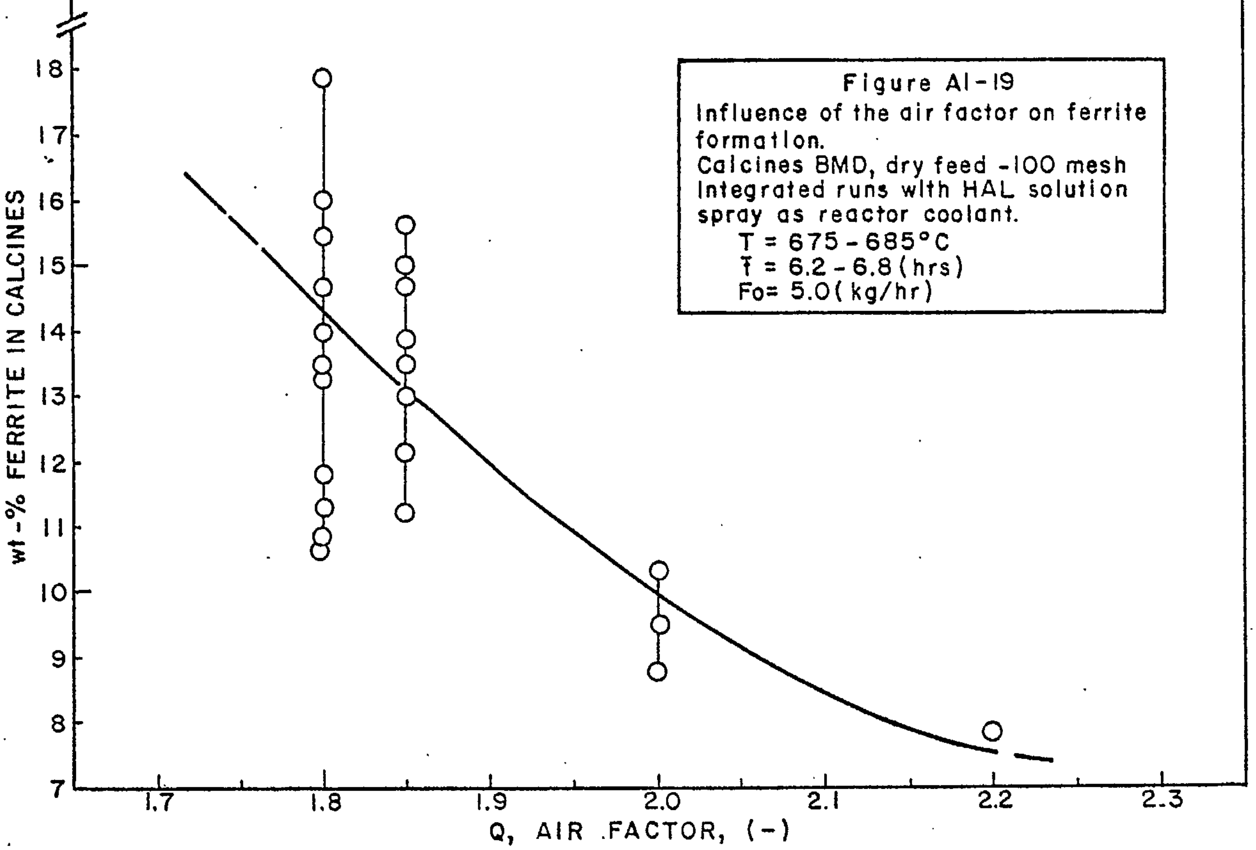


Figure A1-19
Influence of the air factor on ferrite formation.
Calcines BMD, dry feed -100 mesh
Integrated runs with HAL solution spray as reactor coolant.
T = 675 - 685°C
$\bar{t}$ = 6.2 - 6.8 (hrs)
Fo = 5.0 (kg/hr)

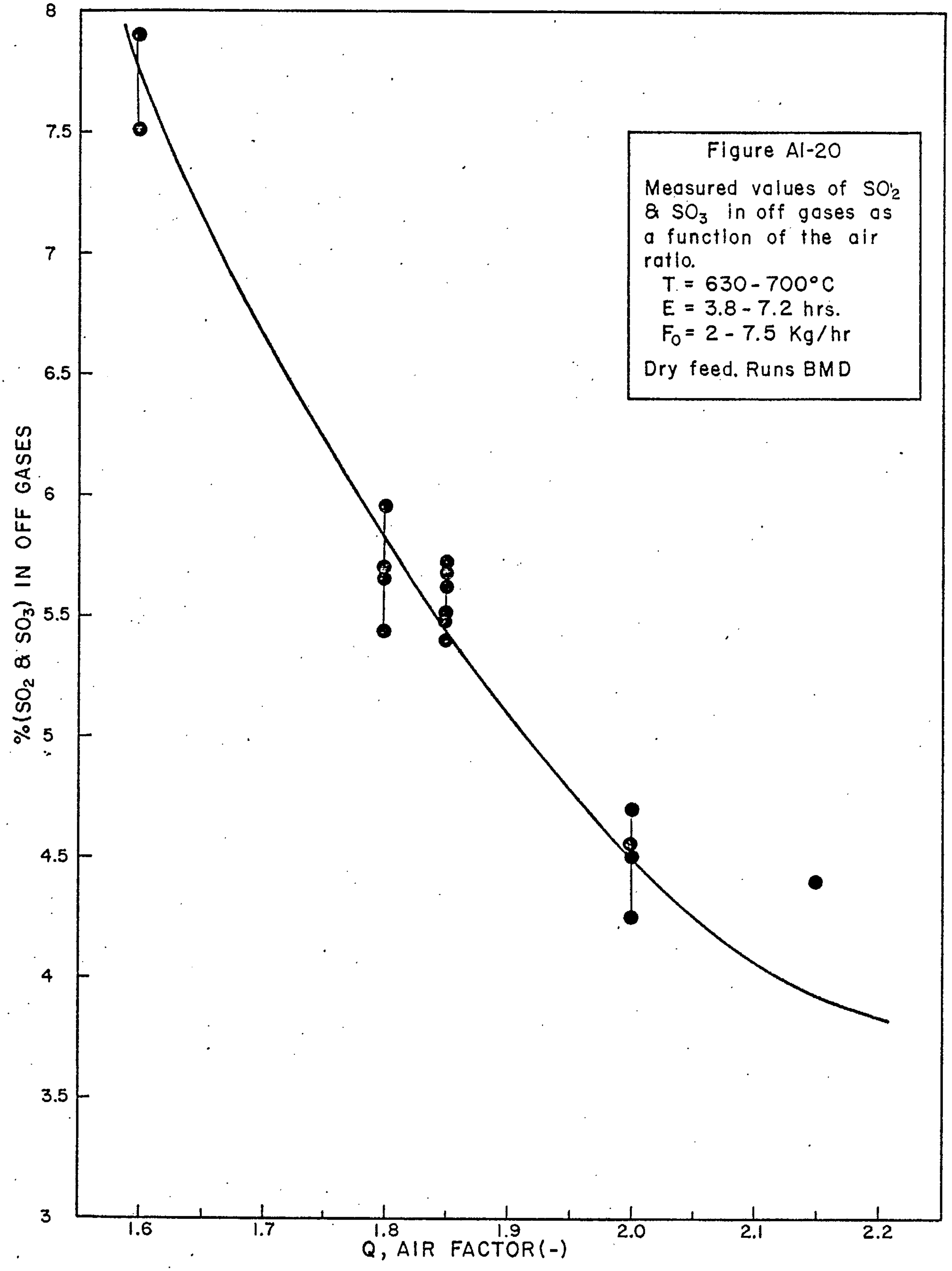


Figure A1-20

Measured values of SO_2 & SO_3 in off gases as a function of the air ratio.
T = 630-700°C
E = 3.8-7.2 hrs.
F_0 = 2-7.5 Kg/hr

Dry feed. Runs BMD

continuous infrared analyzer connected to the off-gases from the pilot reactor. At Q = 1.6 the off-gases contain approximately 7.5% (SO_2+SO_3), while at Q = 2 this value drops to 4.2-4.7%. A more dilute gas increases the capital and operating costs of the sulphuric acid plant and also carries over a larger dust load due to the resulting higher elutriation rate. These two undesirable effects indicate that in a commercial unit, Q should be kept at the lowest possible value in order to achieve a more concentrated SO_2+SO_3 off-gas and less dust carryover. For a commercial reactor, with a bed depth of 1.5 m, the gas velocity at Q = 1.6 could be 65 cm/sec (at 685°C) sufficient to achieve a near perfect fluidization, while in the 14 inch pilot unit with a bed depth of 0.55 m and Q = 1.85, at 685°C the gas velocity was only 22 cm/sec. For example, at the copper-cobalt sulphation plant at Chambishi (Zambia)(5) the air factor is kept at 1.31, for a bed depth of 1.3 m and a reactor diameter of 6.9 m operating at 680°C.

The Hecla Mining Company copper sulphation reactor in Arizona (6), while in operation in 1976/77 purposely employed a low air ratio (Q = 1.1-1.2) so as to achieve as high an SO_2 concentration as possible for their sulphuric acid plant. The two reactors in this plant have a diameter of 6.9 m with a bed depth of 1.35 m.

It should be noted also that an excessive decrease in Q could produce a near complete depletion of oxygen in the off-gases, resulting in an SO_2+SO_3 content of over 10% that could effectively displace the stability field from Fe_2O_3 to $FeSO_4$ to such an extent that the soluble iron level in the "neutral" leach solution would be raised to unacceptably high levels.

Off-gases from the pilot roaster contain from 4.5 to 7.5% of (SO_2+SO_3) (dry basis) depending upon the air factor used and the amount of HAL solution recirculated to the reactor (Figure A1-20). The SO_3 levels in the off-gases were measured at about 1%. Values from the St. Joe Lead Co. zinc sulphation roasting pilot trials indicated a much lower value, less than 0.1% (1). Reported values from the Kokkola cobalt sulphation plant of Outokumpu in Finland are approximately 0.8% SO_3, with a total of 5.6% (SO_2 + SO_3) (7).

Theoretical calculations, assuming that equilibrium can be reached at 40% conversion at 685°C in the presence of Fe_2O_3 (best conversion value achieved by older plants using hematite as a catalyst) give a theoretical value of 1.2%.

In commercial operation, the SO_3 level will be influenced by the air ratio, temperature, and particle size of hematite present. It is possible that the actual level under these conditions will approximate 0.5-0.7%, similar to that experienced at the Outokumpu operation.

The distribution of the zinc compounds in the calcines was determined by x-ray diffraction. Average values of remnant sphalerite, ferrite, zinc sulphate and oxysulphate for 10 runs using dry feed concentrate (series BMD) and water as coolant for the reactor and 17 runs using HAL solution as coolant are plotted in Figure A1-21. These

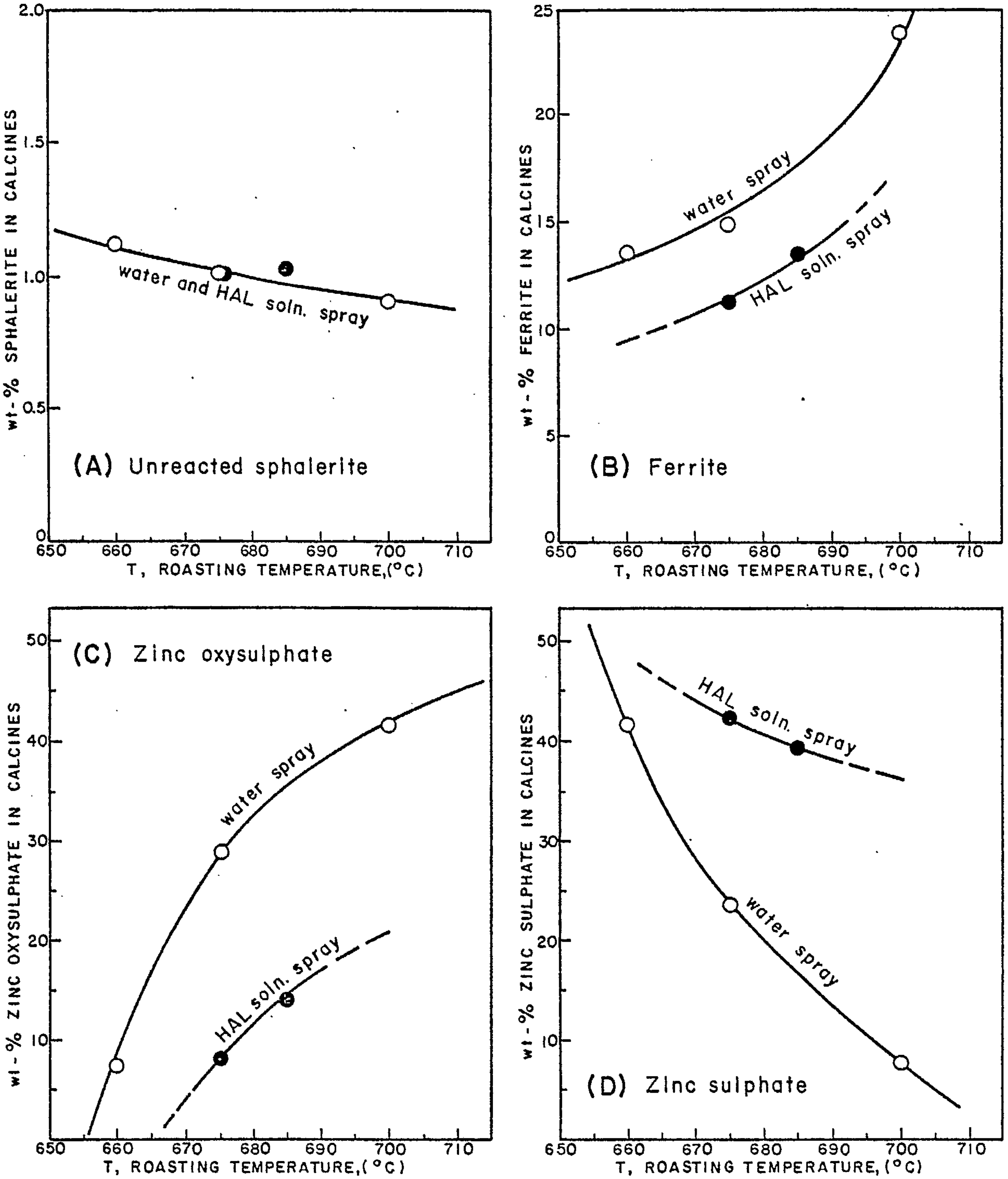


Figure A1-21

Distribution of zinc compounds in calcines as function of the roasting temperature for water and HAL solution sprayed into the roaster as coolant. (Calcines BMD)

Q = 1.8 - 2.0; $\bar{t}$ = 6.2 - 6.8(hr); Fo = 5.0(kg/hr)

○ water spray as coolant

● HAL soln. spray as coolant

Averages of 10 runs for water and 17 runs for HAL soln.

results are in agreement with previous data obtained in earlier bench scale tests (RPC Report M/77/209).

The level of remnant sphalerite in the calcine is essentially the same for water or HAL coolant sprays. The amount of sphalerite decreases on average with increasing temperature from 1.25 wt% at 660°C to 0.92 wt% at 700°C (Figure A1-21A).

The amount of ferrite in calcines is a function of the roasting temperature and the nature of the coolant. For water spray at 680°C the ferrite level was 13.8 wt% which increased to 24.0 wt% at 700°C. The ferrite content of the calcine when the HAL coolant used was approximately 25% less than for the water coolant produced calcine (Figure A1-21B).

Oxysulphate levels show the expected increase with increasing roasting temperature for both the water and HAL coolant solutions. However, the level produced with the HAL spray is very much less than with the water spray. For example at 675°C the oxysulphate levels were 8.8 wt% for the HAL solution and 29.3 wt% for the water spray (Figure A1-21C).

Zinc sulphate levels decrease with increasing roasting temperature. The decrease is particularly marked for the water spray case, from 42 wt% sulphate at 660°C to approximately 7 wt% at 700°C. The trend to lower sulphate levels with increasing roasting temperatures is much less pronounced using the HAL coolant. Also at any given temperature, the sulphate level in calcines produced with the HAL coolant is considerably higher than for those produced with the water spray. At 675°C HAL calcines contain approximately 42 wt% zinc sulphate compared with approximately 27 wt% for the water spray calcines (Figure A1-21D).

The decrease in oxysulphate and ferrite and the increase in zinc sulphate in calcines produced with the HAL coolant compared with the water coolant reflect the marked influence of increased SO_2 partial pressure resulting from the decomposition of the HAL solution in the roaster. Of particular importance is the effect on the sulphate/oxysulphate equilibrium: any reduction in the oxysulphate level will reduce the effect of the oxysulphate-hematite reaction to form ferrite, and hence will enhance zinc extraction in the dilute (neutral) acid leach.

The influence of both roasting temperature and water or HAL solution spray on soluble iron in the dilute acid leach is shown in Figure A1-22. A rapid decrease in the iron in solution takes place when the temperature increases from 660°C to 685°C, from an average value of 0.55 down to 0.1 gpl when water was sprayed as the coolant media, while for HAL solution, iron levels were higher than for water, decreasing from 0.65 to 0.26 gpl iron from 660 to 685°C. The higher levels of iron in solution found when the HAL solution was sprayed may be due to remnant and/or shortcircuited pellets in the discharged calcine in which some iron sulphates remain undecomposed.

Although roasting temperature over the range 675-685°C did not appreciably affect the level of ferrite in

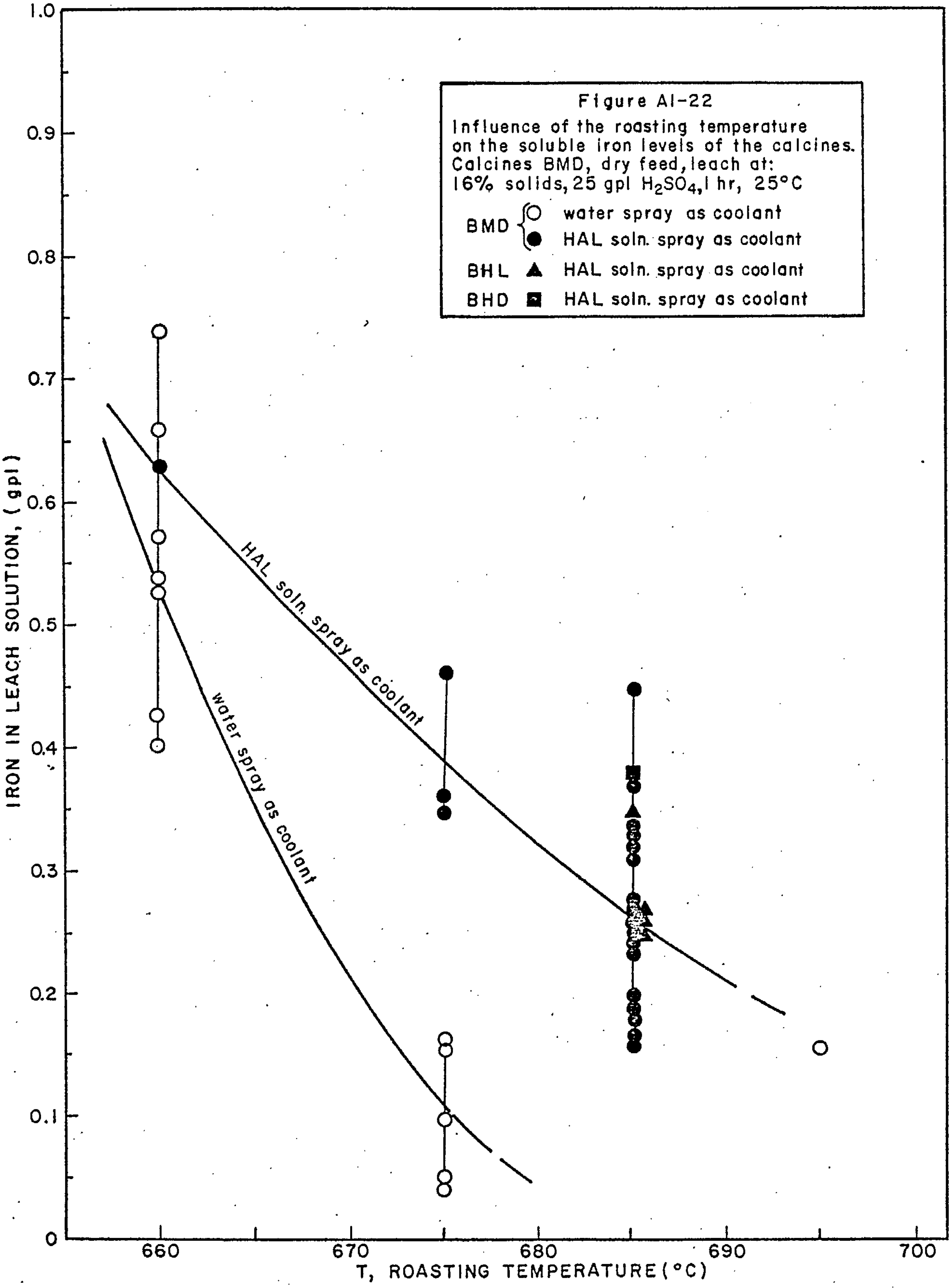


Figure A1-22

Influence of the roasting temperature on the soluble iron levels of the calcines. Calcines BMD, dry feed, leach at: 16% solids, 25 gpl H_2SO_4, 1 hr, 25°C calcines (Figure A1-11) or zinc extraction in the dilute acid leach (Figure A1-14) the influence on soluble iron is significant.

The dilute acid leach filtrate from calcines roasted at 685°C contained 0.26 gpl iron (for the control leaches at 16% solid), whereas in calcines roasted at 675°C, the iron level in the filtrate increases to 0.40 gpl. Since iron levels in the neutral leach in commercial practice must be minimized to avoid filtration problems and excessive requirements for oxidants, 685°C is considered the preferred roasting temperature to keep ferrite formation low and to achieve acceptable low iron levels in solution.

To fully evaluate the advantages of the HAL solution recirculation in continuous operation, a series of interlocked runs were made using the tailings refloat product as the roaster feed with the strong acid leach filtrate of the neutral leach residue continually recycled to the roaster as a coolant.

.4 Interlocked Runs

Calcines from BMD-52 up to 62 were produced and leached in an interlocked manner with the hydrometallurgy section, the coolant spray for each run being the hot acid leach filtrate from the calcines of the previous run. These trials were carried out to evaluate the reproducibility of data and to follow the possible buildup of metals and minor elements in the calcine over an extended period of time.

Results from these interlocked runs for the roasting section are given in Table A1-3(c) and are fully interpreted later in the leaching-purification-electrowinning section of this report.

The behaviour of zinc in the interlocked runs is shown in Figure A1-23B. Zinc extractions in water increased rapidly from 65% to an average of 72-75% after the third consecutive interlocked run, while primary (refers to fresh feed) zinc extraction in dilute acid remained constant at an average of 85%. The overall zinc extraction, including extraction from the HAL solution, in the dilute (neutral) acid leach varied from 95 up to 99.4% in the first six interlocked runs. This variation reflects initial operational difficulties in the hot acid leach stage (discussed in Section 131.6). In the last four interlocked runs it was shown that under optimized HAL conditions, over 98% overall zinc extraction could be consistently achieved.

The calcines exhibit a rather stable composition, with a slight drop in zinc content at the beginning of the interlocked runs (Figure A1-23A). This is due to the proportionally higher iron in the pellets formed from the HAL solution compared with the calcine from the bulk of the dry feed.

Copper levels in the feed material were variable in the range 0.47-0.60 wt% and this variation is also reflected in the calcine (Figure A1-24A). The calcine, on average,

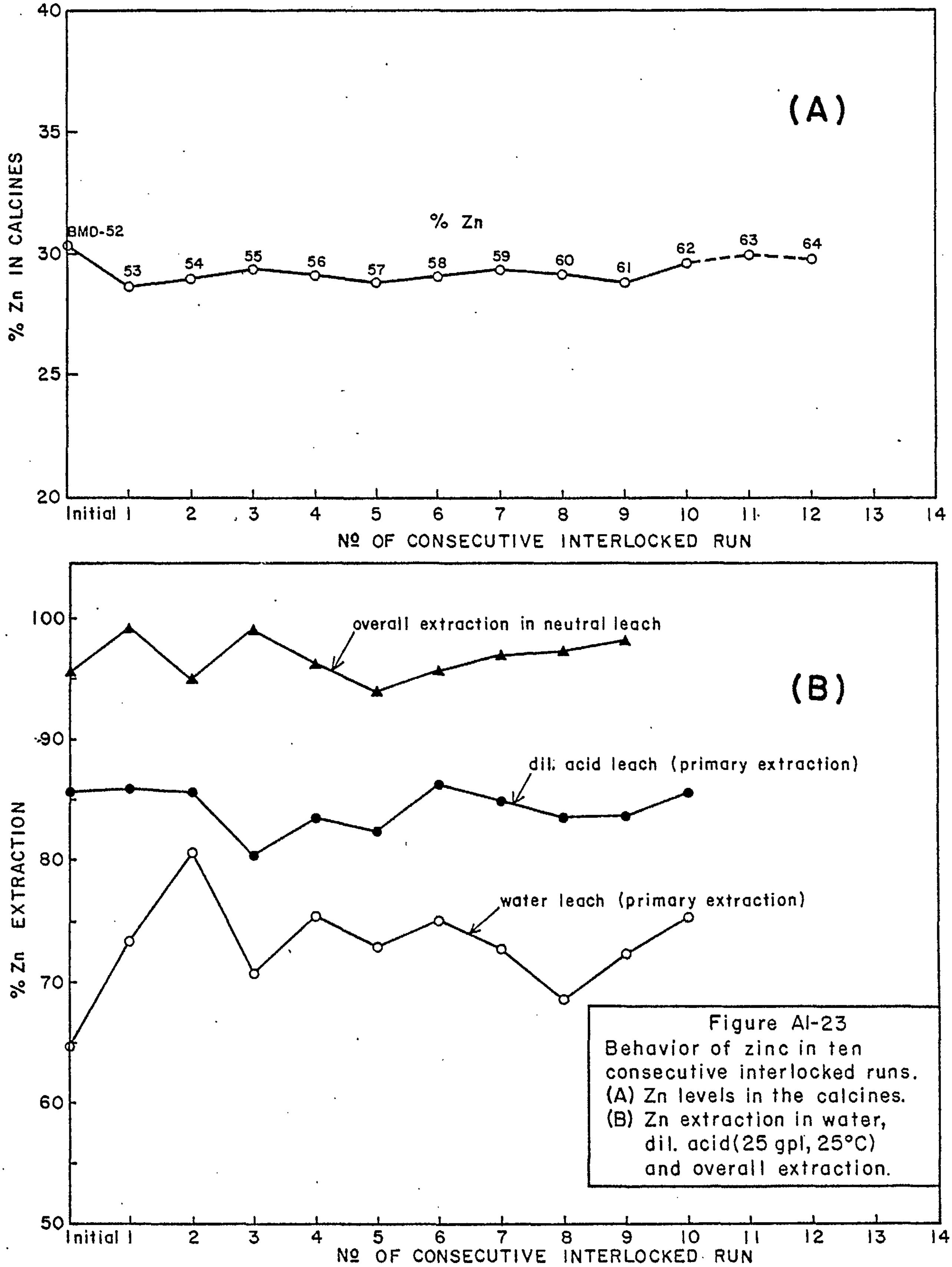


Figure AI-23
Behavior of zinc in ten consecutive interlocked runs.
(A) Zn levels in the calcines.
(B) Zn extraction in water, dil. acid(25 gpl, 25°C) and overall extraction.

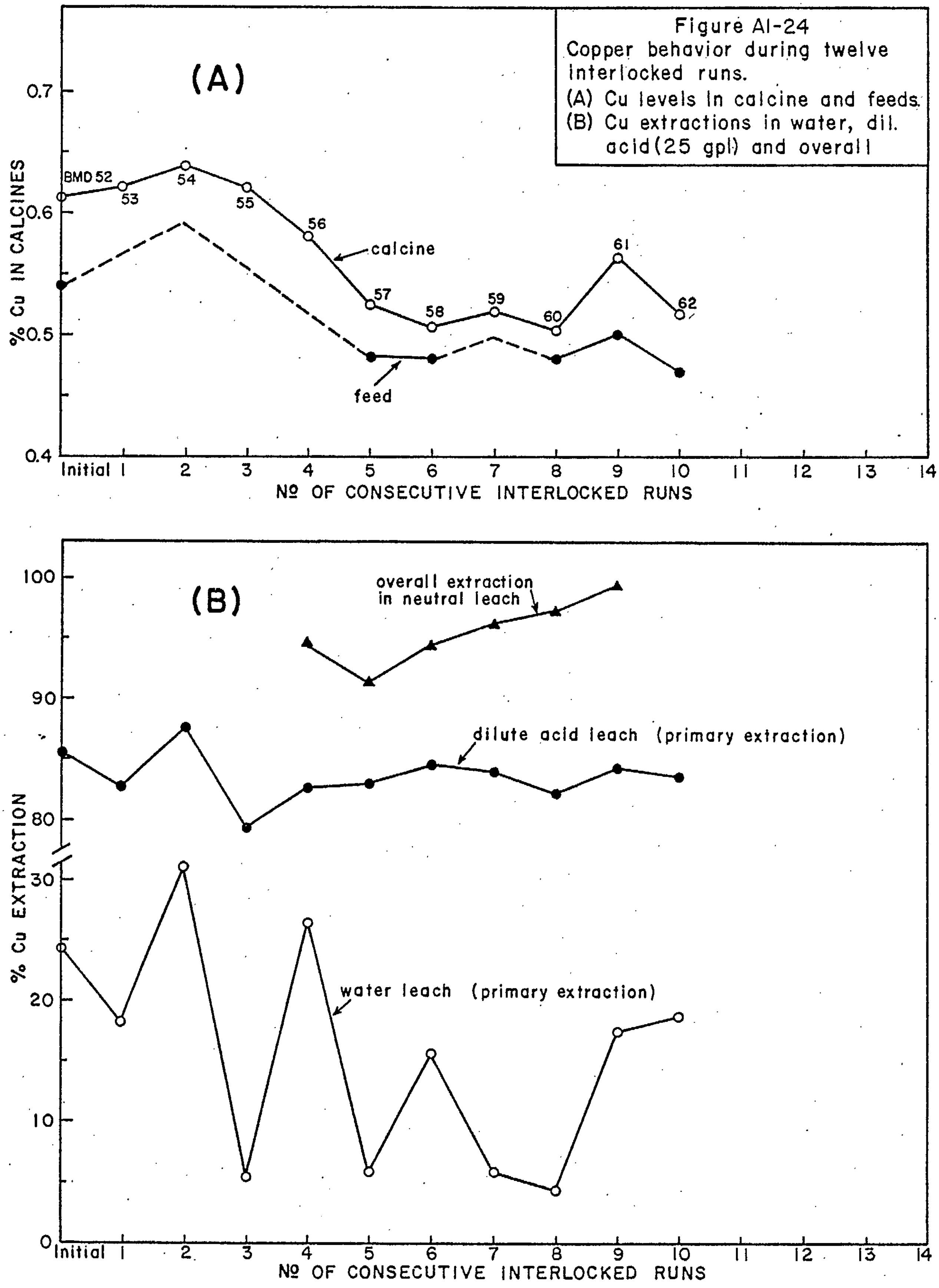


Figure AI-24
Copper behavior during twelve interlocked runs.
(A) Cu levels in calcine and feeds.
(B) Cu extractions in water, dil. acid (25 gpl) and overall

contains 5 to 10% higher copper than the feed. This is due to precipitation of some copper oxysulphate in the neutral leach which is subsequently recovered in the hot acid leach stage, effectively increasing the recirculating load of copper in calcines.

The overall copper extraction was high, from 92 up to 99.5% in the last runs where optimization of the leaching conditions was achieved.

Copper extraction in water (Figure A1-24B) shows a rather erratic pattern, paralleling the zinc extraction in water (Figure A1-23B). This may be due to the oxidation of the $CuSO_4$ by the $ZnO.2\ ZnSO_4$ by a solid state reaction in the roaster as follows:

$$2CuSO_4 + ZnO.2ZnSO_4 \rightarrow CuO.CuSO_4 + 3ZnSO_4$$

This reaction has a negative free energy of -3.5 Kcal at 685°C and an inversion temperature of 620°C, as shown in Figure A1-25. This figure also illustrates a relationship between the level of copper in calcines and the amount of oxysulphate formed. For calcines containing greater than 0.7 wt% copper, the zinc oxysulphate content decreases markedly to less than 5 wt%. Conversely for copper levels approaching 0.45 wt%, the oxysulphate level increases to over 30 wt%. This relationship of copper to zinc oxysulphate requires more study as it could be particularly important as a limiting factor on the formation of ferrite, i.e. less zinc oxysulphate would result in less ferrite and hence greater zinc extraction in the neutral leach.

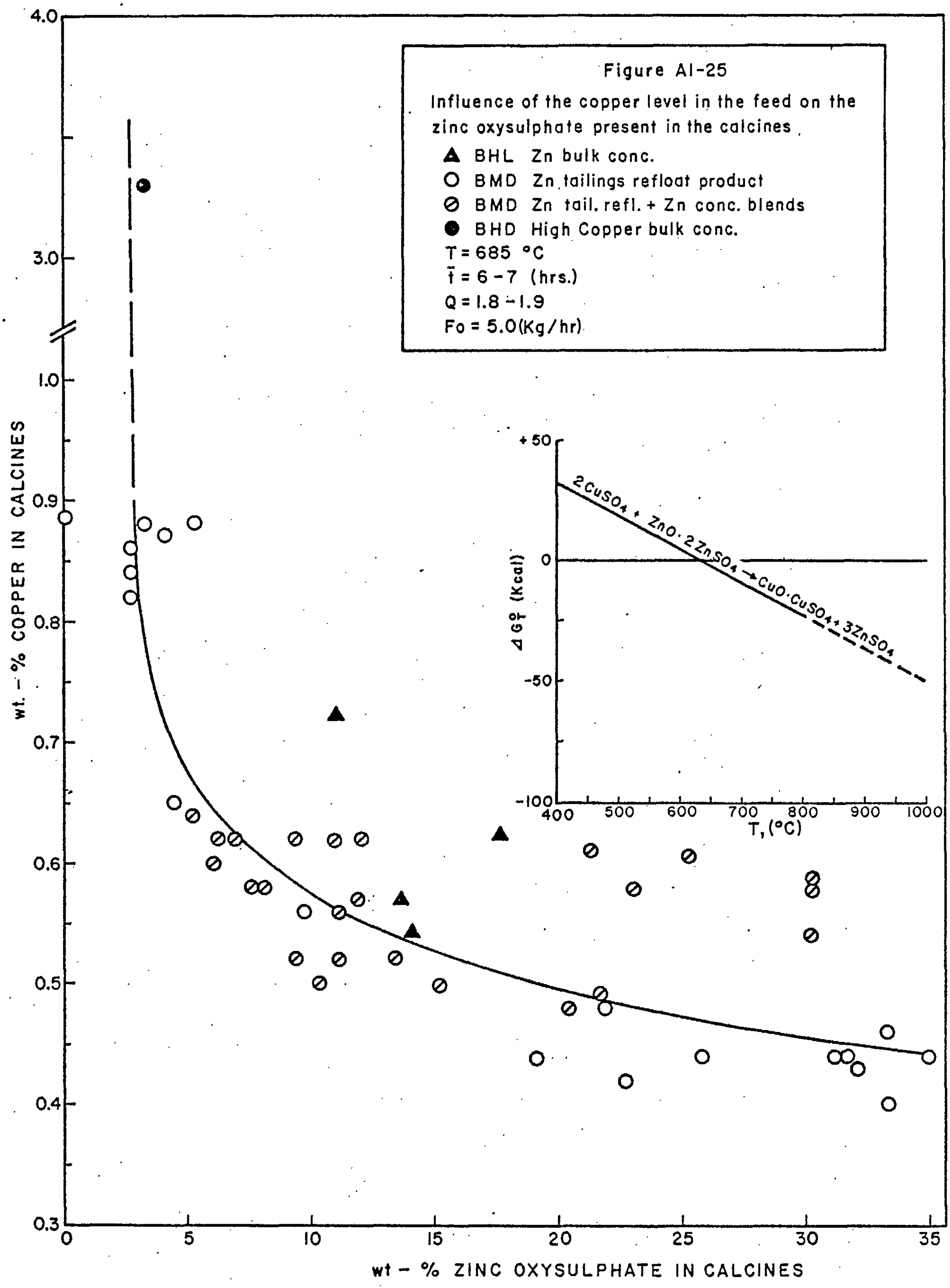


Figure A1-25

Influence of the copper level in the feed on the zinc oxysulphate present in the calcines.

Arsenic in the calcines showed an initial increase in concentration in the interlocked runs from 0.20 wt% at the start, up to 0.42-0.45 wt% at the end of the program. The initial content in the feed was 0.35 wt%, of which approximately 68% was volatilized with the off-gases and condensed in the scrubber. After the hot acid leach step, most of the arsenic is solubilized and recirculated to the reactor where it builds up to 0.40-0.45% in the calcines, which represents an increase of 30-40% over the initial feed. The HAL residues contain about 32% of the arsenic from the feed. The arsenic content in calcine is shown in Figure A1-26, together with the levels of cobalt and cadmium both of which remain essentially unchanged at less than 0.05 wt% for cadmium and 0.02 wt% for cobalt. Other impurities such as chlorine, selenium and mercury were virtually eliminated by volatilization, with final average levels in the calcines of 10 ppm for mercury, less than 0.1 ppm for selenium and less than 50 ppm for chlorine. The percentage change in arsenic content in the calcines is shown in Figure A1-27. The trend would indicate that after about the 8th consecutive interlocked run, the arsenic content levels off at approximately 0.45 wt%. Values for calcines BMD-63 and 64 were much lower (0.25 wt%) and are not plotted on Figures A1-26 and A1-27 since HAL solutions sprayed were not from interlocked calcines.

As an average, less than 1.5 wt% sphalerite was noted in calcines from the interlocked runs. The principle leach-

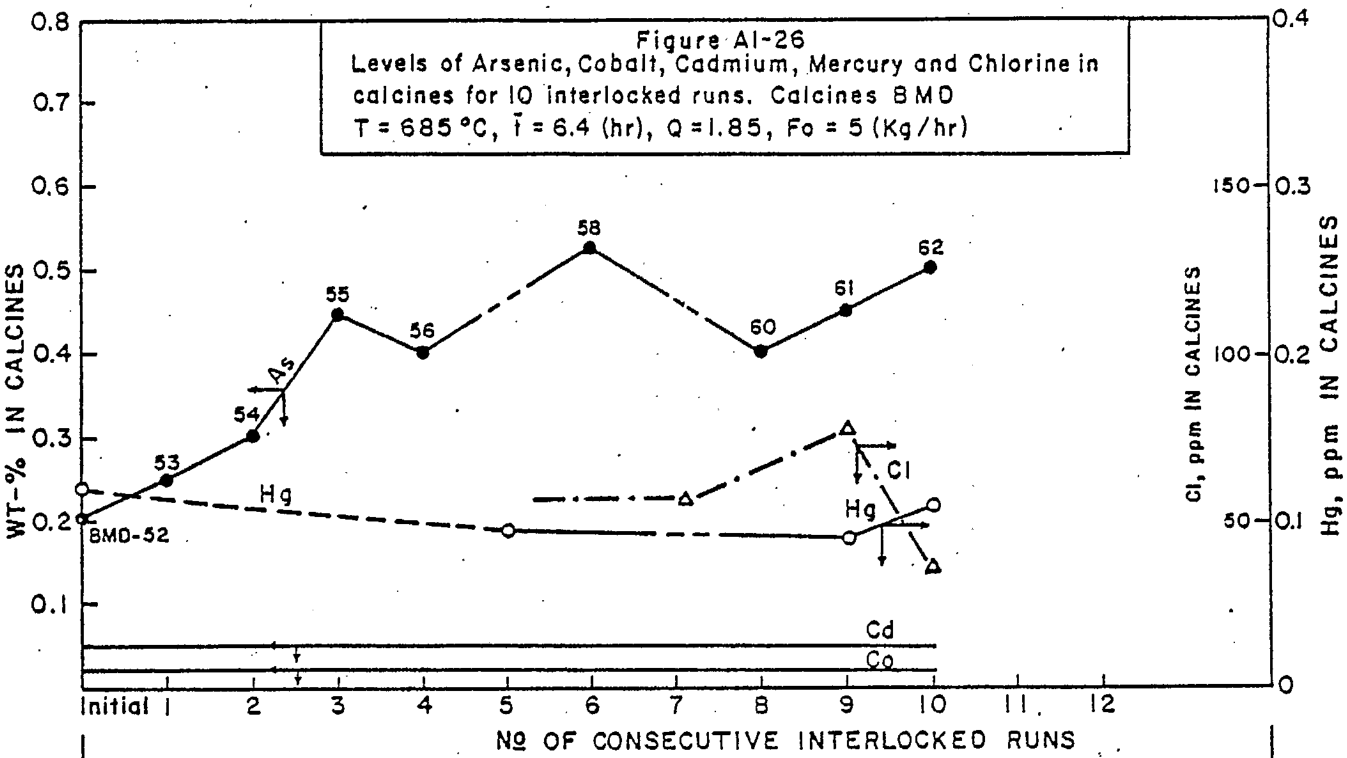


Figure A1-26
Levels of Arsenic, Cobalt, Cadmium, Mercury and Chlorine in calcines for 10 interlocked runs. Calcines BMD
T = 685 °C, $\bar{t}$ = 6.4 (hr), Q = 1.85, Fo = 5 (Kg/hr)

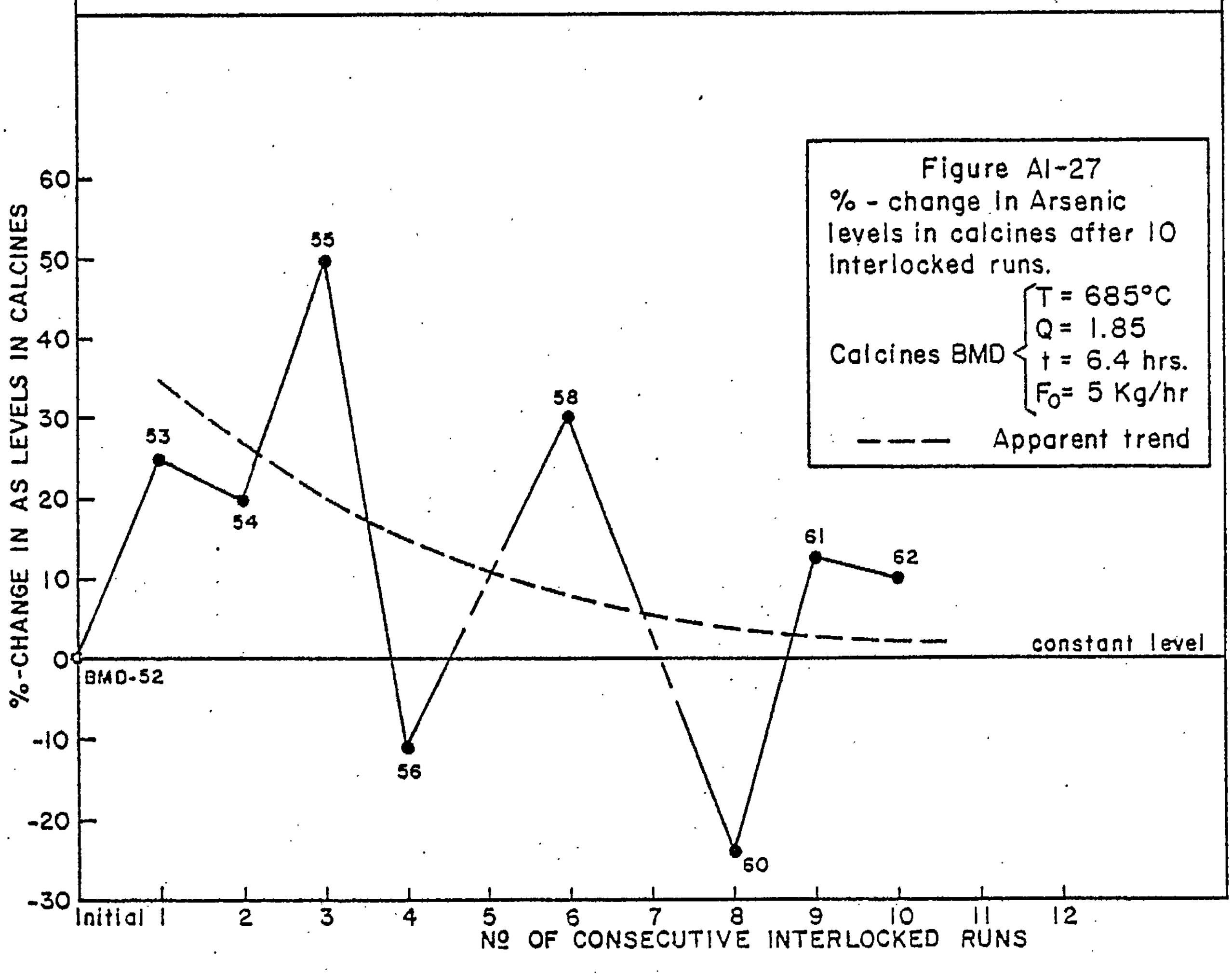


Figure A1-27
% - change in Arsenic levels in calcines after 10 interlocked runs.
Calcines BMD { T = 685°C, Q = 1.85, t = 6.4 hrs., F_0 = 5 Kg/hr
— — — Apparent trend

ing agent for sphalerite in the hot acid leach is ferric ion derived from the dissolution of ferrite. It was determined that a minimum ratio of 1/4 for sphalerite/ferrite in calcine was required to achieve high sphalerite dissolution (more details are given in Section 131.1).

In order to determine the minimum retention time of solids inside the reactor to achieve 1.5 wt% remnant sphalerite, and hence establish the reactor capacity, several roasting tests were carried out at different retention times. The results of these trials are given in Table A1-3(c) and plotted in Figure A1-28. Although there is considerable scatter in the data, there is enough correlation to suggest that a retention time of as little as 5.5 hours is sufficient to yield calcines with 1.5 wt% remnant sphalerite. From previous experience with the 6-inch roaster trials (RPC Report M/77/209), it was found that remnant sphalerite in calcine is to some extent associated with the size of the reactor employed. For example, the 6-inch roaster trials, using the same feed material and the same roasting temperatures, resulted in 12-15 wt% unreacted sphalerite in calcine. Hence it is reasonable to assume that less than 1.5 wt% remnant sphalerite might be achieved with 5.5 hours retention, or alternatively at 1.5 wt% sphalerite the retention time might be reduced in a commercial reactor. This would effectively increase the reactor capacity and thereby reduce the capital cost of the roasting section.

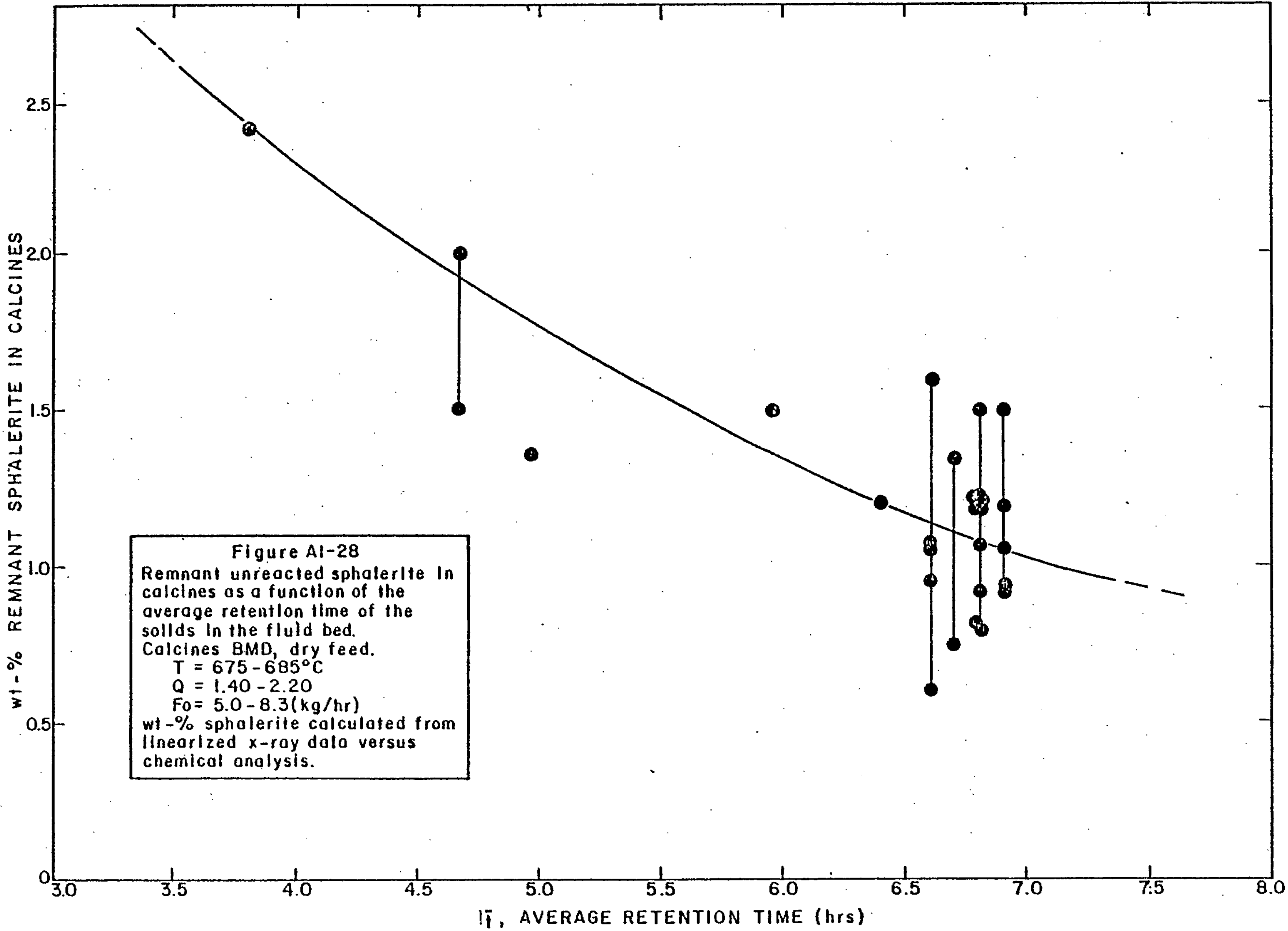


Figure Al-28

Remnant unreacted sphalerite in calcines as a function of the average retention time of the solids in the fluid bed.
Calcines BMD, dry feed.
T = 675 - 685°C
Q = 1.40 - 2.20
Fo = 5.0 - 8.3(kg/hr)
wt-% sphalerite calculated from linearized x-ray data versus chemical analysis.

TABLE A1-5

CALCINE PHYSICAL PROPERTIES

BMD SERIES - DRY FEED, HOT ACID LEACH (HAL) FILTRATE RECYCLE
(Interlocked Runs)

Calcine No.	HAL Feed (lt/kg)*	Calcine Wt %		Calcine Vol. %		Calcine Bulk Density (gr/cc)		Calcine Bulk Density, (gr/cc)
		-60mesh**	+60mesh***	-60mesh	+60mesh	-60mesh	+60mesh	(all calcine)
BMD-55	0.456	80.2	19.8	63.9	36.1	1.19	0.62	1.07
56	0.456	85.9	14.1	74.1	25.9	1.24	0.58	1.15
57	0.456	95.9	4.1	92.3	7.7	1.26	0.65	1.23
58	0.336	87.6	12.3	76.5	23.5	1.27	0.59	1.19
59	0.384	90.0	10.0	82.9	17.1	1.25	0.68	1.19
60	0.384	91.8	8.2	83.8	16.2	1.21	0.56	1.16
61	0.384	87.9	12.1	68.9	31.1	1.13	0.34	1.03
62	0.338	87.1	12.9	71.4	28.6	1.20	0.45	1.10

* lt filtrate per kg of dry feed

** calcine produced from dry feed

*** pellets produced from decomposed HAL filtrate

The physical properties of calcines obtained with the HAL solution recycle differ from those obtained using water as coolant for the reactor or using slurry feed. Overall bulk density decreased from 1.25 T/m^3 for dry feed with water spray down to 1.15 T/m^3 for dry feed with the HAL solution as the coolant (Table A1-5). The lower bulk density of the latter is accounted for by the low density (0.56 T/m^3) porous pellets from the decomposition of the HAL solution. The pellets in the interlocked runs approximate 12.3 wt% of total calcine bed material, this represents 23.8 volume %. The increase in total weight of the calcines due to the HAL solution decomposition was taken as constant in calculations for the reactor scale-up. The factor being 0.123 F_o, where F_o is the feed rate. This factor is representative of the highest tolerated levels of ferrite and sphalerite (15 wt% ferrite and 1.5 wt% sphalerite).

The average physical properties of the calcines produced in the interlocked runs are given in Table A1-5 and Table A1-6.

TABLE A1-6

AVERAGE PHYSICAL PROPERTIES OF CALCINES FOR INTERLOCKED RUNS

	Calcine Distribution				Calcine Bulk Density*	
F_o/HAL Soln.	wt%		vol%		(T/m^3)	
(kg/lt)	-60 mesh	+60 mesh	-60 mesh	+60 mesh	-60 mesh	+60 mesh
2.506	87.7	12.3	76.2	23.8	1.23	0.56

-60 mesh represents calcines produced from fresh feed;
+60 mesh represents pellets formed from HAL solution decomposition.
* calcine bulk density total calcine = 1.15 T/m^3.

The recirculation of the fines collected in the cyclones and electrostatic precipitator (cottrell) was necessary since the fines contain a higher proportion of unreacted sphalerite and soluble iron than does the bulk of the calcine. Soluble iron is due to resulphation of the fine hematite particles. This problem was especially acute for the cottrell where the temperature of the off-gases dropped below 400°C. Levels of remnant sphalerite and ferrite in calcine and dust products are shown in Figure A1-29 and Table A1-7. Ferrite shows lower levels in dust than in the calcines, probably because some resulphation of ferrite takes place, or kinetic phenomena related to the particle size is involved in the ferrite formation. Sphalerite levels were, on average, similar for calcine and cyclone products but higher in the cottrell dust. Soluble iron, which in the calcines produced control leaches of iron level varying between 0.2-0.4 gpl (Figure A1-21) shows an increase of 20-30% in cyclone products and a 150-250% increase for cottrell dust, with values as high as 1.8 gpl iron in solution shown in Figure A1-30. Zinc sulphate and oxysulphate did not show any particular trend in the dust, as shown in Table A1-7.

The overall collection efficiency of the duplex cyclone system and cottrell was, under normal conditions of operation, 98.98% on average (Table A1-8). The main difficulty was the shortcircuiting of the cottrell electrode due to caking by iron sulphate. In commercial size electrostatic

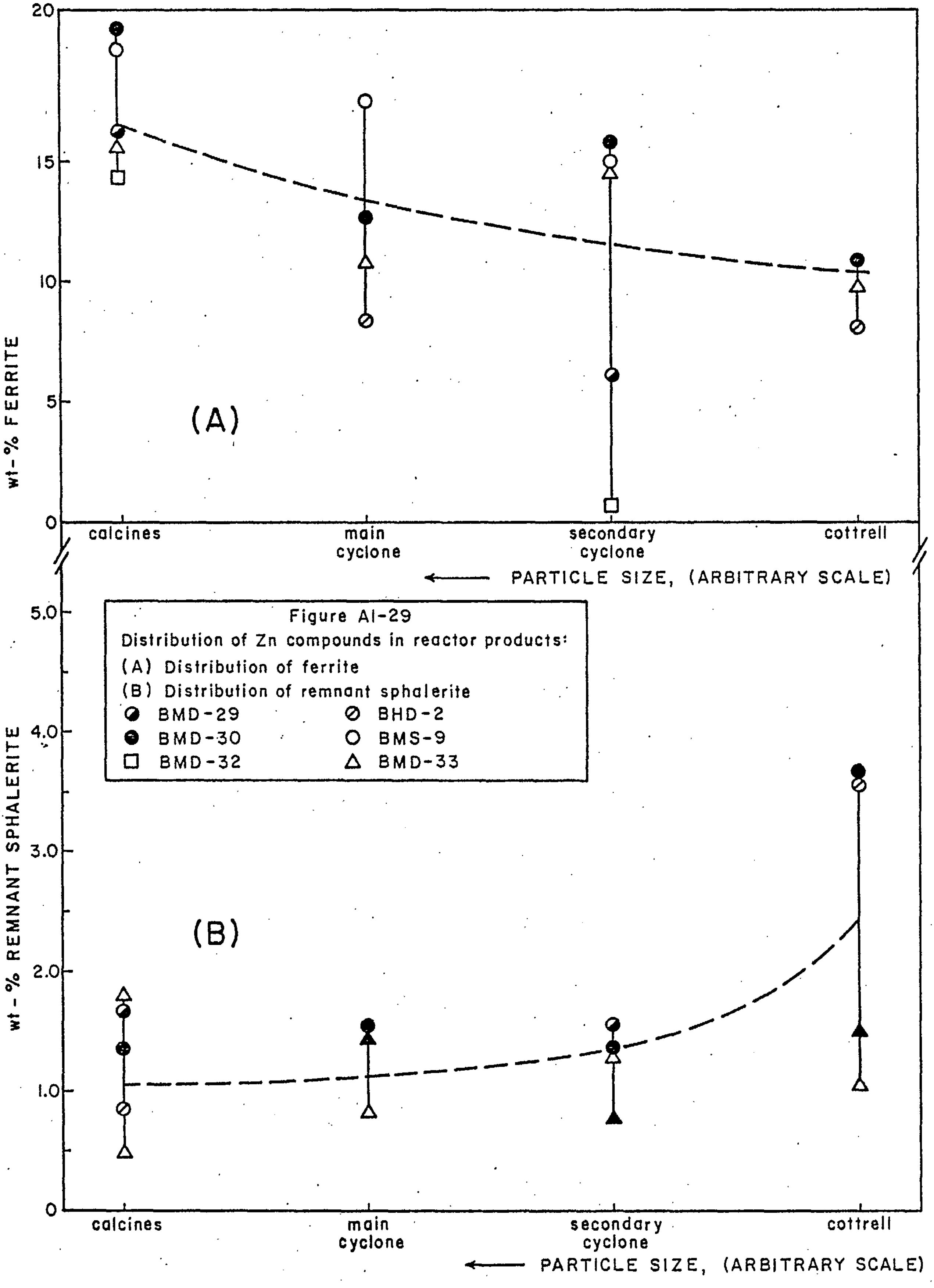


Figure A1-29

Distribution of Zn compounds in reactor products:

(A) Distribution of ferrite

(B) Distribution of remnant sphalerite

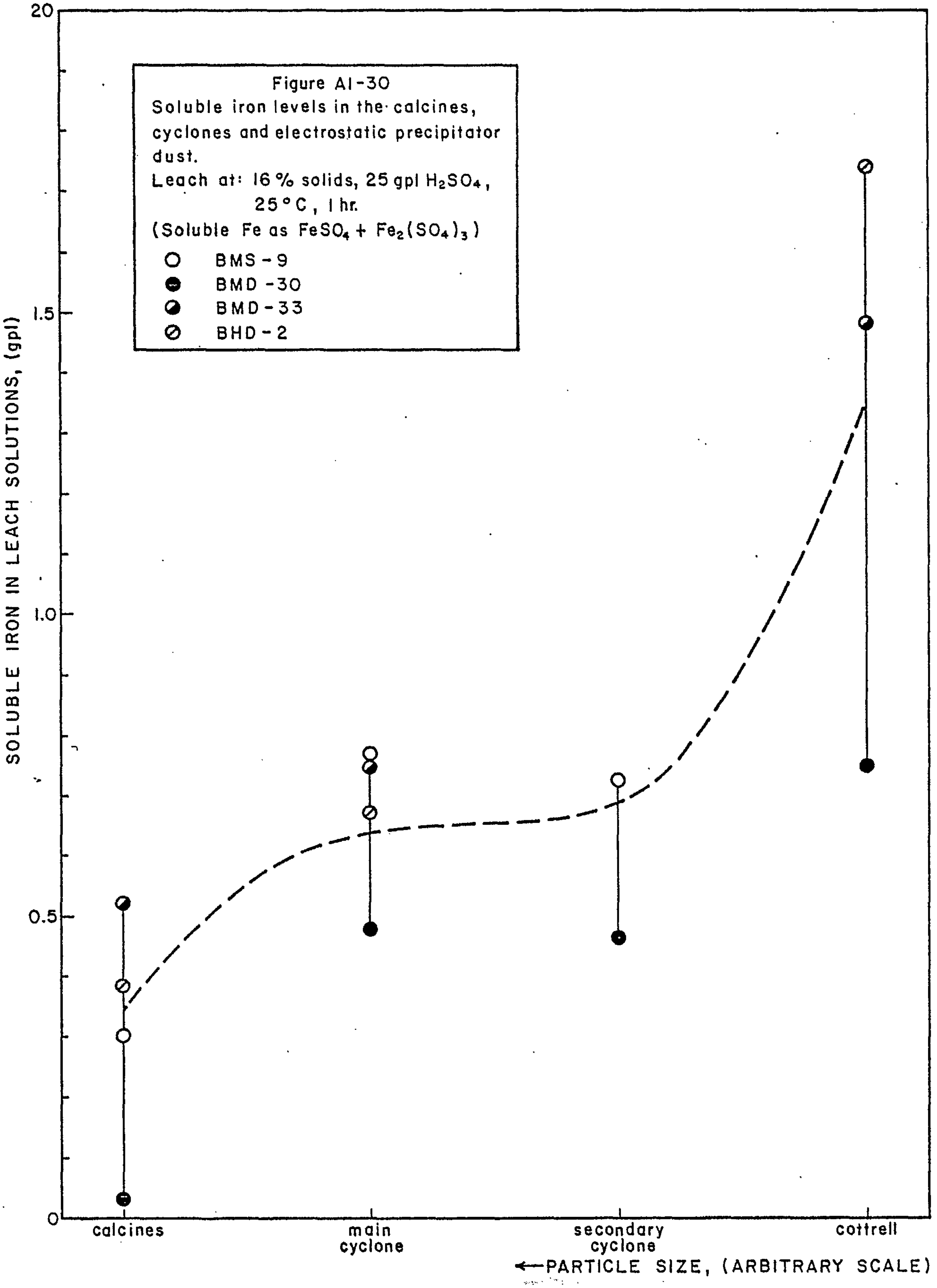


Figure A1-30

Soluble iron levels in the calcines, cyclones and electrostatic precipitator dust.

Leach at: 16 % solids, 25 gpl H_2SO_4, 25 °C, 1 hr.

(Soluble Fe as $FeSO_4 + Fe_2(SO_4)_3$)

TABLE A1-7

DISTRIBUTION OF ZINC COMPOUNDS AND SOLUBLE IRON IN THE CALCINES, CYCLONES AND ELECTROSTATIC PRECIPITATOR DUSTS

Run No.	Product	Zinc Compound Distribution (wt %)				Iron in solution (gpl) (*)
		sphalerite	ferrite	oxysulphate	sulphate	
BMD-29	calcine	1.66	16.44	31.20	18.18	0.04
	2nd cyclone	1.55	6.36	11.09	49.11	-
BMD-30	calcine	1.39	20.46	35.21	15.36	0.05
	1st cyclone	1.56	12.89	35.49	19.67	0.49
	2nd cyclone	1.39	16.37	30.16	23.97	0.46
	Cottrell	3.60	10.93	41.07	10.97	0.74
BMD-32	calcine	0.91	14.57	32.10	27.31	0.10
	2nd cyclone	-	0.97	13.23	60.20	-
BMD-33	calcine	1.79	16.88	25.86	27.69	0.52
	1st cyclone	0.89	10.99	14.50	46.43	0.76
	2nd cyclone	1.39	14.52	23.58	32.49	-
	Cottrell	1.08	9.11	16.20	45.38	1.48
BMS-9	calcine	0.50	19.10	1.25	42.85	0.31
	1st cyclone	1.50	17.23	10.04	32.01	0.77
	2nd cyclone	0.75	14.99	8.03	37.42	0.71
	Cottrell	1.50	28.01	10.13	24.71	-
BHD-2	calcine	0.87	8.86	3.12	53.07	0.38
	1st cyclone	3.40	8.34	2.15	48.66	0.67
	Cottrell	3.60	8.04	2.18	47.29	1.74

(*) Leaching at: 16 wt % solids, 25°C, 1 hr, 25 gpl H_2SO_4

TABLE A1-8

OVERALL PARTICLE COLLECTION EFFICIENCY AND CALCINE LOSSES TO THE SCRUBBER

Run No.*	Cal. Total Elutriated (kg)	Total Calcines Produced (kg)	Total Solid In Scrubber (kg)	Overall Elutriated Collection Eff. (%)	Dust Losses to Scrubber (%)
BMD 13	230	30.1	0.36	98.8	1.2
15	236	32.9	0.52	98.4	1.6
17	301	38.3	0.77	98.0	2.0
18	559	75.3	1.33	98.3	1.8
21	306	39.7	0.89	97.8	2.2
28	429	56.1	0.50	99.1	0.9
30	163	35.6	0.13	99.6	0.4
32	179	35.5	0.18	99.5	0.5
33	161	24.3	0.12	99.5	0.5
34	232	30.1	0.13	99.6	0.4
35	153	26.0	0.26	99.0	1.0
36	193	32.9	0.53	98.4	1.6
42	209	35.6	0.50	98.6	1.4
43	485	63.0	0.77	98.8	1.2
47	242	41.1	0.19	99.5	0.5
49	243	41.0	0.22	99.5	0.5
50	274	46.5	0.28	99.4	0.6
51	232	30.1	0.43	98.6	1.4
52	250	41.2	0.45	98.9	1.1
54	437	73.9	0.57	99.2	0.8
55	485	82.1	0.18	99.8	0.2
57	250	42.4	0.40	99.1	0.9

* Runs BMD-12,16, 19, 20, 23-26, 37, 41, 44, 45, 53, 56, 58 and 61-63 rejected due to Cottrell short circuit and/or secondary cyclone leg plugging.

Runs BMD-1-11, 14, 22, 27, 29, 31, 38-40, 46, 48, 59-60: no samples taken.

precipitators this problem is overcome with continuous electrode rappers, such as those used at the cobalt sulphation roast plant of Outokumpu in Finland (4).

Limitations in the roaster height and second cyclone design also created some problems with occasional plugging of the second cyclone vortex discharge. This problem was found to be associated with the design and size of the equipment and is not related to the physical characteristics of the dust. No stickiness, other than that noted when sulphation promoters such as Na_2SO_4 were employed, was noted in the fluidized bed, cyclones or cottrell and no deleterious effects such as corrosion were noted with the spraying of various coolants including the HAL solution.

.5 Effect of Increased Lead-Application to a Bulk Concentrate

The roasting process with the HAL solution recirculation to the roaster was tested on a synthetic bulk concentrate which was prepared by blending a lead concentrate with the tailings refloat product. The bulk concentrate sent to RPC by CANMET (also prepared by blending) was highly oxidized with much of the lead sulphide already converted to lead sulphate. As the bulk concentrate roasting program was in part to assess the sulphation of lead, the oxidized CANMET bulk was considered unacceptable and was not used in the test program. The composition of the synthetic bulk concentrate tested in this program is given in Table A1-1 and Table A1-2.

TABLE A1-9

ROASTING CONDITIONS AND CONTROL LEACH RESULTS* FOR ZINC PRODUCTS OF VARYING COMPOSITION

Run No.	Temp.	Av Ret. Time	Air Ratio	Feed Rate	Velocity	Slurry & Liquor Feed									% Metal Extraction					
						Electrolyte, (gpl)		Pulp	Slurry Gun			Cooling		Dry Feed	water, 25°C			acid, 25°C, 25 gpl		
	T (°C)	$\bar{t}$ (hr)	Q (-)	Fo (kg/hr)	u_o' (cm/sec)	H_2SO_4	Zn	% Solids	Tip Ø (mm)	Air (lt/min)	Level Lg (cm)	water (cc/min)	liquor (cc/min)	(mesh)	Cu	Zn	Fe	Cu	Zn	Fe (gpl)
BF Series																				
BF -1	685	6.8	1.80	5.0	20.8	-	-	-	1.5	50	35	-	5	-100	1.59	55.17	0.01	79.54	79.41	1.09 (0.14)
2	685	6.8	1.80	5.0	20.8	180	80	31	1.5	43	35	-	-	-100	0.56	7.88	0.01	63.73	81.84	0.29 (0.04)
BHL Series																				
BHL-1	685	6.6	1.85	5.0	20.8	-	-	-	1.5	50	35	-	12	-100	3.07	69.68	0.02	80.35	83.70	3.63 (0.35)
2	685	6.6	1.85	5.0	20.8	-	-	-	1.5	50	35	-	16	-100	1.80	66.32	0.01	80.69	84.00	2.99 (0.27)
3	Rejected due to operational problems																			
4	685	6.8	1.85	5.0	20.8	-	-	-	1.5	50	35	-	32	-100	2.12	70.09	0.00	80.98	80.38	3.21 (0.28)
5	685	6.6	1.85	5.0	20.8	-	-	-	1.5	50	35	-	32	-100	3.12	70.77	0.02	80.53	82.79	2.99 (0.27)
BZC Series																				
BZC-1	685	6.6	1.85	5.0	19.4	-	-	-	1.5	50	35	-	31	-100	0.62	39.11	0.02	1.17	65.96	0.01(<0.1)
BHB Series																				
BHB-1	Rejected due to operational problems																			
2	Rejected due to operational problems																			
BHC Series																				
BHC-1	Rejected due to operational problems																			
BHD Series																				
BHD-1 (1 sample)	685	6.4	1.85	5.0	19.6	-	-	-	1.5	50	35	-	28	-100 }	67.15	85.36	0.72	89.72	88.74	4.37 (0.39)
2	685	6.4	1.85	5.0	19.6	-	-	-	1.5	50	35	-	28	-100						

* (values for BHD series represent primary metal extractions)

TABLE A1-10

ZINC COMPOUNDS FOR ZINC PRODUCTS OF VARYING COMPOSITIONS

ZINC COMPOUNDS IN BRUNSWICK TAILINGS REFLOAT PRODUCT WITH FERRITE RESIDUE

	Weight %				Zinc Distribution (%)			
	sphalerite	ferrite	oxysulphate	sulphate	sphalerite	ferrite	oxysulphate	sulphate
BF -1	1.11	20.29	25.86	27.68	2.19	18.40	41.95	37.48
2	1.18	17.07	29.42	24.03	2.38	15.79	48.68	33.19

ZINC COMPOUNDS IN BRUNSWICK ZINC BULK CONCENTRATE

	Weight %				Zinc Distribution (%)			
	sphalerite	ferrite	oxysulphate	sulphate	sphalerite	ferrite	oxysulphate	sulphate
BHL-1	1.84	12.58	13.82	40.57	3.94	12.35	24.27	59.48
2	2.31	11.64	17.83	37.30	4.84	11.16	30.60	53.43
3	Rejected due to operational problems							
4	1.64	12.55	14.27	41.26	3.47	12.16	24.73	59.69
5	1.54	13.11	11.13	39.78	3.51	13.69	20.79	62.04

ZINC COMPOUNDS IN BRUNSWICK ZINC CONCENTRATE

	Weight %				Zinc Distribution (%)			
	sphalerite	ferrite	oxysulphate	sulphate	sphalerite	ferrite	oxysulphate	sulphate
BZC-1	3.61	28.97	44.01	18.12	5.53	20.32	55.22	18.98

ZINC COMPOUNDS IN BRUNSWICK HIGH COPPER BULK CONCENTRATE

	Weight %				Zinc Distribution (%)			
	sphalerite	ferrite	oxysulphate	sulphate	sphalerite	ferrite	oxysulphate	sulphate
BHB-1	Rejected due to operational problems							
2	Rejected due to operational problems							
BHC-1	Rejected due to operational problems							
BHD-2	0.87	8.86	3.12	53.07	1.98	11.07	5.85	82.93

Four continuous, interlocked runs with HAL solutions were performed. Lead in the concentrates ranged from 10 up to 14.3 wt%, while the levels of zinc and copper remained at the same approximate values as in the previously tested tailings refloat product.

The roasting variables established for the tailings refloat product were also employed for the bulk, these were:

temperature = 685°C
air factor = 1.85
retention time = 6.6 (hrs)
feed rate = 5.0 (kg/hr)

The experimental conditions and results of these runs are given in Tables A1-9, A1-10 and A1-11.

TABLE A1-11
RESULTS OF ROASTING TESTS WITH BULK CONCENTRATE

Run	Feed, wt%			%Metal Extraction				
				Dil. acid leach(1)		Overall(2)		
	Zn	Cu	Pb	Zn	Cu	Zn	Cu	Fe(gpl)
BHL-1	29.95	0.72	10.85	83.70	80.35	98.70	98.56	0.35
2	29.95	0.72	10.85	84.00	80.69	98.72	98.59	0.27
3	30.20	0.74	10.50	----- rejected due to operational problems ----				
4	30.20	0.74	10.50	84.39	80.98	98.75	98.61	0.28
5	26.75	0.92	14.30	82.79	80.53	98.63	98.58	0.27

(1) Primary extraction. Leach at 25 gpl H_2SO_4, 25°C, 1 hour, 16% solids
(2) Including HAL solution decomposition in dilute acid leach

The HAL solutions used as coolant spray in runs BHL-1 to 5 were derived from calcine BMD-61 plus approximately 30% of the total volume derived from a synthetic mix of chemicals to approximate the composition of the HAL solution from the BMD-61 neutral leach residue.

The results show that both primary extraction in dilute acid leach from fresh feed and overall extraction including the HAL solution recycle were high for both zinc and copper, with an average of 98.70% for zinc and 98.59% for copper, with 0.30 gpl iron in solution for the dilute acid leach. The behavior of the fluidized bed was reasonably good, with only minor initial problems. Run BHL-3 was rejected due to erratic temperature control. Fluidization was adequate even at 14.3 wt% lead in the feed. These runs proved that, even at the relatively low air velocity used (20.8 cm/sec at 685°C), the bed was agitated to a sufficient extent to avoid defluidization. This is an important finding since at the actual gas velocity in a commercial reactor (70-80 cm/sec) agitation is more violent and defluidization should not be a problem.

The cyclones and cottrell performed well and the overall performance during the roasting was similar to that experienced using the tailings refloat products.

.6 Effect of Increased Copper

The roasting process was further tested with higher copper products in order to determine the effect of high copper and lead on fluidization behavior.

Blends of tailings refloat products, zinc concentrate and copper-lead bulk concentrate were prepared to obtain feeds with the chemical composition given in Table A1-12.

TABLE A1-12
HIGH COPPER BULK CONCENTRATES CHEMICAL COMPOSITION

Feed	Wt % Zn	Cu	Pb	Fe
BHB	28.75	5.12	9.05	19.38
BHC	30.50	6.80	3.00	19.00
BHD	30.50	4.48	4.00	21.50

Roasting conditions used in all three tests were the same as those used for the tailings refloat products and also for the bulk concentrates, as follows:

temperature = 685°C
air factor = 1.85
feed rate = 5.0 (kg/hr)
retention time = 6.6 (hrs)

Defluidization problems were experienced with run BHB with 5.12 wt% copper and 9.05 wt% lead. Although the bed did not sinter or harden, the 'stickiness' was sufficient to avoid proper fluidization with an air velocity of 19.2 cm/sec. A similar, although reduced effect, was experienced with run BHC with copper at 6.80 wt% and lead at 3.00 wt%. With the copper level lowered to 4.5 wt% and lead at 4.8 wt%, as in run BHD, the stickiness problem disappeared. This particular feed was run continuously, on a two shift basis, for two days without difficulty. The calcine obtained was low in zinc oxysulphate and ferrite, 3.1 wt% and 8.9 wt% respectively, which is consistent with the observations from the tailings refloat tests, i.e., that high copper content could reduce the level of zinc oxysulphate in calcines and as a consequence limit the formation of ferrite.

The stickiness problem associated with 'relatively' high copper with lead is most probably a result of a low melting eutectic involving both copper and lead and this may not be a problem in commercial reactors where air velocities are greater. However, information on the lead-copper-sulphate system could not be found in the literature and hence more test work should be undertaken to establish the working range for copper and lead in the sulphation process.

Results of interlocked tests with BMD material using HAL solution equivalent to that which would be obtained from a neutral leach residue from a calcine containing 15 wt% ferrite and 1.5 wt% sphalerite are given in Tables A1-9, A1-10 and A1-13.

TABLE A1-13
RESULTS OF ROASTING TESTS WITH HIGH COPPER BULK CONCENTRATE

Run	Metal Extraction %				
	dil. acid leach(1)		overall(2)		
	Zn	Cu	Zn	Cu	Fe(gpl)
BHB-1	--------- rejected due to defluidization of bed --------				
BHC-1	--------- rejected due to defluidization of bed --------				
BHD-1 & 2	87.39	86.28	98.99	98.90	0.40

(1) Primary extraction from feed, leach at 25 gpl H_2SO_4, 1 hour, 25°C, 16% solids.
(2) Calculated assuming 92% extraction Zn and Cu in HAL.

.7 Effect of Increased Zinc

A zinc concentrate containing 51 wt% zinc was subjected to one preliminary test to determine the effect of increased zinc. The results, given in Tables A1-9 and A1-10 and summarized in Table A1-14 were rather poor, with less than 66% primary extraction of zinc in the dilute acid leach. This is an artificially low result due to insufficient acid in the control leach for efficient oxysulphate leaching. The oxysulphate level was high as was the quantity of ferrite at 44.0 wt% and 29.0 wt% respectively, possibly due to the low SO_2 partial pressure in the bed resulting from the low pyrite content of the feed. More test work would obviously be required to re-establish optimum roasting conditions for zinc-rich concentrates.

TABLE A1-14
RESULTS OF ZINC CONCENTRATE ROASTING TEST

Run	Feed			Metal Extraction %				
	%Zn	%Cu	%Pb	water(1)		acid(2)		
				Zn	Cu	Zn	Cu	Fe(gpl)
BZC-1	51.2	0.5	2.0	39.11	0.61	65.96	1.17	0.00

(1) Primary extraction in water. Leach: 1 hr, 25°C, 16% solids.
(2) Primary extraction in acid. Leach: 25 gpl H_2SO_4, 25°C, 1 hr, 16% solid.

.8 Ferrite Resulphation Test

Two roasting tests were carried out to gain information on the resulphation of ferrite. In the one case, BF-1, a mixture of 30% zinc tailings refloat product and -100 mesh synthetic ferrite was prepared to give a blend containing 15 wt% ferrite. Synthetic HAL solution (approximating the composition which would be derived from a calcine containing

15 wt% ferrite and 1.5 wt% sphalerite) was used as the coolant. A second test, BF-2, was carried out using the same proportion of ferrite to tailings refloat product, but with the feed introduced to the roaster as a slurry with the HAL solution.

The leaching results indicate that, without the use of promoters, resulphation of ferrite is only partially achieved, i.e. approximately 47.8% resulphation with the dry feed, and 66.7% with the slurry feed (Tables A1-9, A1-10 and A1-15). The higher resulphation of ferrite in the slurry feed system is probably a result of prior dissolution of ferrite in the HAL solution. Higher sulphation of ferrite values using ferrite residues and pyrite concentrate blends, plus Na_2SO_4 promoters were reportedly achieved by Det Norske Zinkkompani in Norway (8) and by the Akita Zinc Co. in Japan (9) both in commercial operations. This resulphation process has apparently been abandoned at both plants.

TABLE A1-15
RESULTS OF FERRITE RESULPHATION TESTS

Run	Feeds (%)			Metal Extractions %		
	Tailings Refloat		Ferrite	From Ferrite Res.	From Tailings Refl.(3)	
	Zn	Cu	Zn	Zn	Zn	Cu
BF-1	36.62	0.46	27.11(1)	47.84	85.60	79.54
BF-2	32.62	0.46	27.11(2)	67.68	85.61	63.73

(1) Blend of dry -100 mesh tailings refloat (85 wt%) + ferrite (15 wt%).
(2) Tailings refloat dry -100 mesh (85 wt%) + ferrite slurried with HAL solution (15 wt%).
(3) Primary metal extraction in acid. Leach: 25 gpl H_2SO_4, 1 hr, 25°C, 16% solid.

123 Considerations for Plant Scale-up

Based on the 14-inch diameter reactor system, the following roasting conditions were chosen for scale-up to a commercial unit:

roasting temperature = 685°C
retention time of solids = 5.5 (hrs)
air factor = 1.8
feed rate = 4.1 (MT/m^2-day)

.1 Roasting Temperature

The roasting temperature was established at 685°C to limit the amount of soluble iron in the neutral leach. However, since at least part of the iron solubilized appears to be due to incomplete decomposition of the HAL solution in shortcircuited pellets, it is possible that in a commercial unit, where shortcircuiting would be less of a problem, the roasting temperature could be reduced, perhaps to 675°C or lower. The lower roasting temperature would also help to decrease the amount of ferrite in the calcine.

.2 Retention Time

Retention time is the limiting factor determining the amount of unreacted sphalerite in the calcine. However, as retention time is to some extent a function of the size of the reactor, it is possible that less than the specified 5.5 hours retention will be possible in commercial reactors without increasing the amount of unreacted sphalerite in the calcine above 1.5 wt%. Lower retention time means increased reactor capacity per unit area and hence lower capital costs per unit of recovered zinc.

.3 Air Factor

Again, the air ratio to some extent is a function of reactor size, i.e. the smaller the reactor the greater the required air velocity to maintain proper bed fluidization. This, plus consideration of existing commercial sulphation roasters, would indicate that the proposed zinc sulphation roaster might operate at an air factor of 1.8 or lower.

The air factor is also an important variable in determining the (SO_2+SO_3) strength in the roaster off-gas and the dust load. Also, as heat is required to preheat the air inside the reactor, any reduction in the air factor will effectively increase the capacity of the roaster for consuming either HAL solution or spent electrolyte.

An air ratio of 1.8 was selected for scale-up purposes. At a bed depth of 1.5 m at 685°C this corresponds to a gas velocity of 75 cm/sec compared with 21 cm/sec for the 14-inch pilot reactor under the same roasting conditions.

.4 Feed Rate

For a given reactor diameter, the feed rate is determined by the fluidized bed height. Although bed depths can extend up to 3 m, a bed height of 1.5 m was chosen for calculation purposes. This corresponds to the bed levels commonly employed in conventional dead roasting of zinc concentrates.

.5 Thermal Balance

The consumption of all HAL solution and wash water from the ferrite residue leach in the roaster is an essential part of the RPC process. Thermal balances performed on all zinc products tested show that for 15 wt% ferrite in the calcines, the heat excess generated in the bed alone, not considering the sensible heat of the hot gases inside the reactor, is sufficient to consume all HAL solution and wash water (Appendix A3). For example, for the tailings refloat product the capacity of the reactor for decomposing HAL solution is 0.814 ton of HAL solution per ton of sulphide feed. The basis of this solution is 100 gpl Fe and 100 gpl Zn (Figure A1-31), with a specific gravity of 1.53 g/cm^3, containing 0.53 tons H_20 per ton feed. This represents a 15 % excess capacity in the reactor over that required to consume all HAL solution from a calcine containing 15 wt% ferrite and 1.5 wt% sphalerite.

The above thermal calculations are in agreement with published data for the Chambishi copper-cobalt sulphation plant (5) and with the experimental results achieved in the St. Joseph Lead pilot trials with zinc concentrates (1). At Chambishi, spent electrolyte is consumed in the reactor at the rate of 1.07 tons of water per ton of feed, that is accomplished with a temperature drop in the freeboard from 680°C to 530°C. At the Hecla copper sulphation plant in Arizona (6), gases exit from the reactor at 650°C. The slurry gun is only 0.95 m from the fluidized bed surface and there is virtually no cooling of the gases in the freeboard of the reactor. (Roasting is done at 680°C.) In this plant, the water vaporized to feed ratio is 0.61.

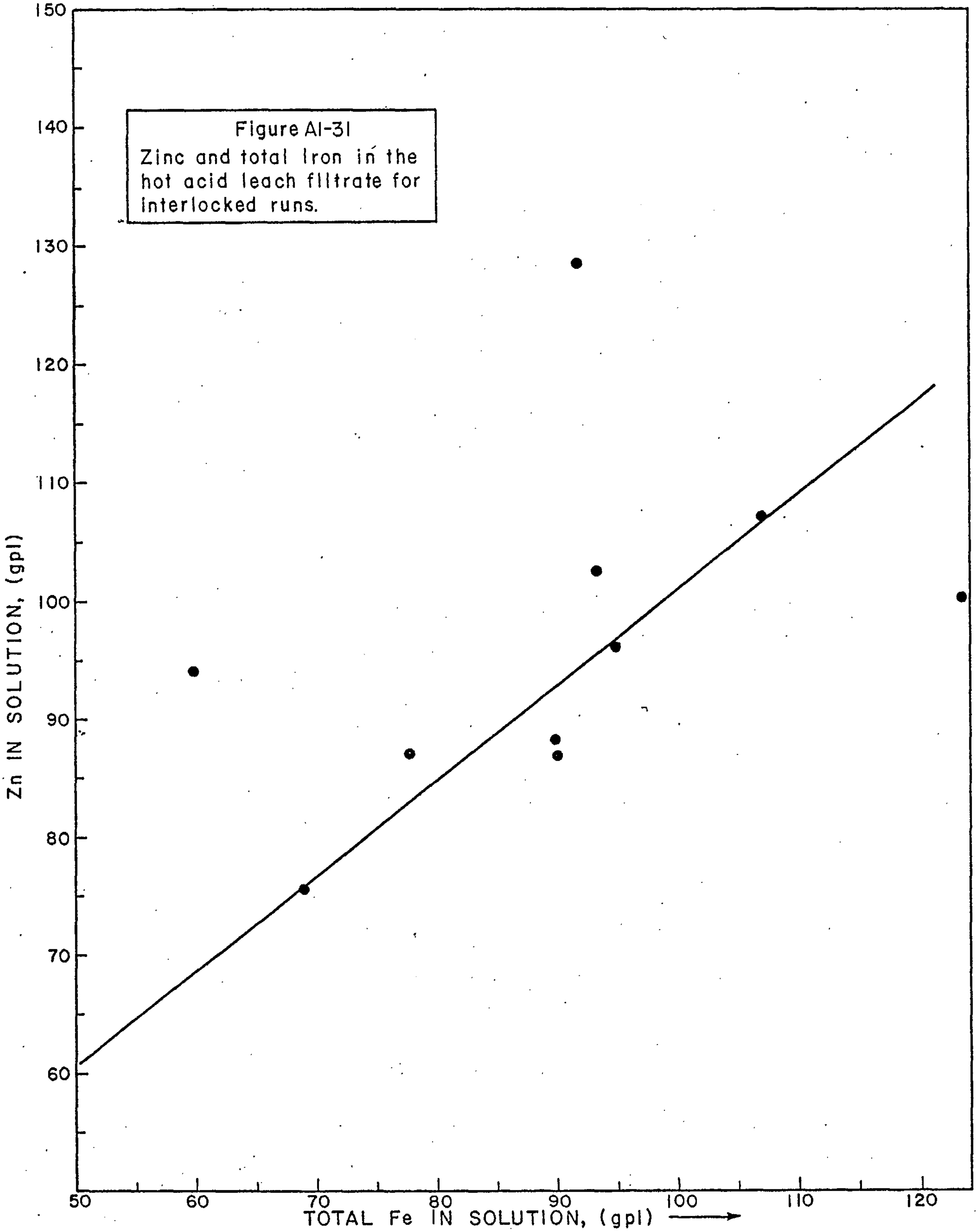


Figure A1-31
Zinc and total Iron in the hot acid leach filtrate for interlocked runs.

Sulphation and oxidation reactions for copper concentrates (chalcopyrite) generate approximately 85% of the heat evolved from the roasting of tailings refloat products and 90% of the heat generated for the bulk concentrate on a per ton of feed basis. In the St. Joe Lead work spent electrolyte was consumed in the reactor at the equivalent rate of 0.8 to greater than 2 tons of water per ton of feed, depending on the feed and operating conditions employed.

Additional heat capacity in the roaster would also be derived with a decrease in the air factor. A decrease in Q from 1.8 to 1.6 would increase the capacity of the reactor to consume HAL solution by approximately 20%. The relationship of air factor, freeboard gas temperature, and HAL decomposition capacity in commercial reactors is shown in Figure A1-32. The calculations are based on a tailings refloat product, which, when roasted, yields a calcine containing 15 wt% ferrite and 1.5 wt% sphalerite.

It can be seen from Figure A1-32 that the reactor's capacity to consume the HAL solution increases rapidly with the decrease of the freeboard gas temperature. For example, at Q = 1.6, and allowing the off-gases to cool down from 685°C to 500°C, the capacity of the roaster to decompose the HAL solution increases by 83%, this represents an increase in thermal capacity of 28 kg HAL solution/°C-ton feed. However, excess consumption of heat in the freeboard to consume HAL solution will decrease the temperature of the exit off-gas and thereby reduce the production of steam in the waste heat boilers as shown in Figure A1-32.

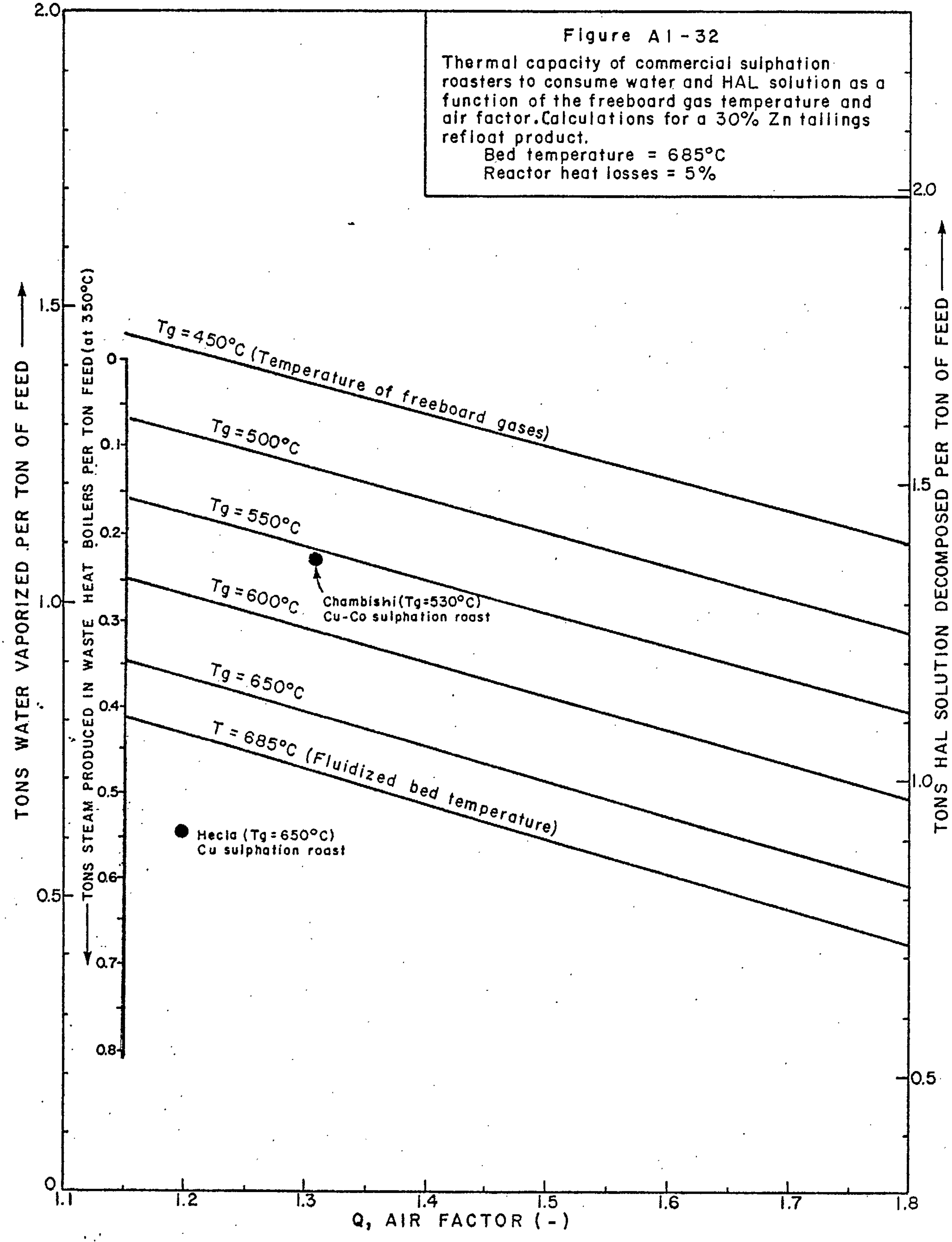


Figure A1-32

Thermal capacity of commercial sulphation roasters to consume water and HAL solution as a function of the freeboard gas temperature and air factor. Calculations for a 30% Zn tailings refloat product.

Bed temperature = 685°C
Reactor heat losses = 5%

The roasting of the bulk concentrate generates about 7% less heat due to the lower levels of pyrite in the feed and higher proportion of galena present.

The thermal losses are estimated as 5% of the reaction heat, a figure usually achieved in commercial units that recover the heat generated in the reactor. In the small 14-inch pilot unit, measured thermal losses were approximately 13-17%. The increase in heat losses from the roaster rapidly decreases the excess heat available (from the bed), for example, for a 15% heat loss, the capacity for consuming the HAL solution decreases by about 38%. During the interlocked runs, external heat was added to compensate for the difference in heat losses from an equivalent commercial unit and achieve the same ratio between feed and HAL solution consumed.

.6 Calculation of Reactor Diameter

The reactor diameter for a roaster is calculated from the following relationship:

$$Dr = 2\left[\left(\frac{1}{\pi}\right)\left(\frac{1}{1-\sigma}\right)\left\{\frac{\bar{t}(F_o\gamma+R)}{L_f\ \rho_c}\right\}\right]^{0.5}$$

The variables are defined as follows:

Dr	=	reactor diameter		(m)
F_o	=	feed rate, (dry basis)	(variable)	(TPH)
R	=	dissolved solids in HAL solution	= $0.123F_o$	(TPH)
$\bar{t}$	=	average retention time of solids	= 5.5	(hrs)
L_f	=	fluidized bed depth	= 1.5	(m)
ρ_c	=	average bulk density of calcines	= 1.15	(T/m^3)
σ	=	gas fraction in fluid bed	= 0.35	(-)
γ	=	weight transformation coefficient	= 1.095	(-)

The gas fraction in the fluidized bed (σ) was calculated using a theoretical model while the experimental values for σ determined in the pilot reactor are given in Figure A6-2.

Values for roaster diameter (Dr) as a function of roaster capacity for retention times ($\bar{t}$) of 5.5 and 6 hours are shown in Figure A1-33. The calculated capacity is approximately 80% greater than for the Dowa (2) zinc-copper sulphation roaster which operated commercially for a short period during the late 1950's, and 30% smaller than the Hecla copper sulphation plant.

Since the scale-up calculation is based on conservative estimates for $\bar{t}$, L_f and Q, it is possible that for a given reactor diameter, the calculated capacity for a commercial plant may be increased by as much as 25-30%.

The capacity of a fluidized bed sulphation reactor, in addition to relating to the variables outlined above, also appears to be largely a function of independent design parameters. For example, comparison of reactors designed by two prominent manufacturers, Dorr-Oliver and Lurgi, show a large difference in capacity per unit area of bed as shown in Figure A1-34. Copper-cobalt sulphation roasters designed by Dorr-Oliver, i.e. Rhokana, U. Miniere, Ndola 2, Ndola 1 and Hoboken all have capacities of less than 4 MT/m^2-day, whereas similar plants built by Lurgi, i.e. Chambishi, and Hecla have capacities greater than 6 MT/m^2-day. Similarly the roaster capacity (2.6 MT/m^2-day) for

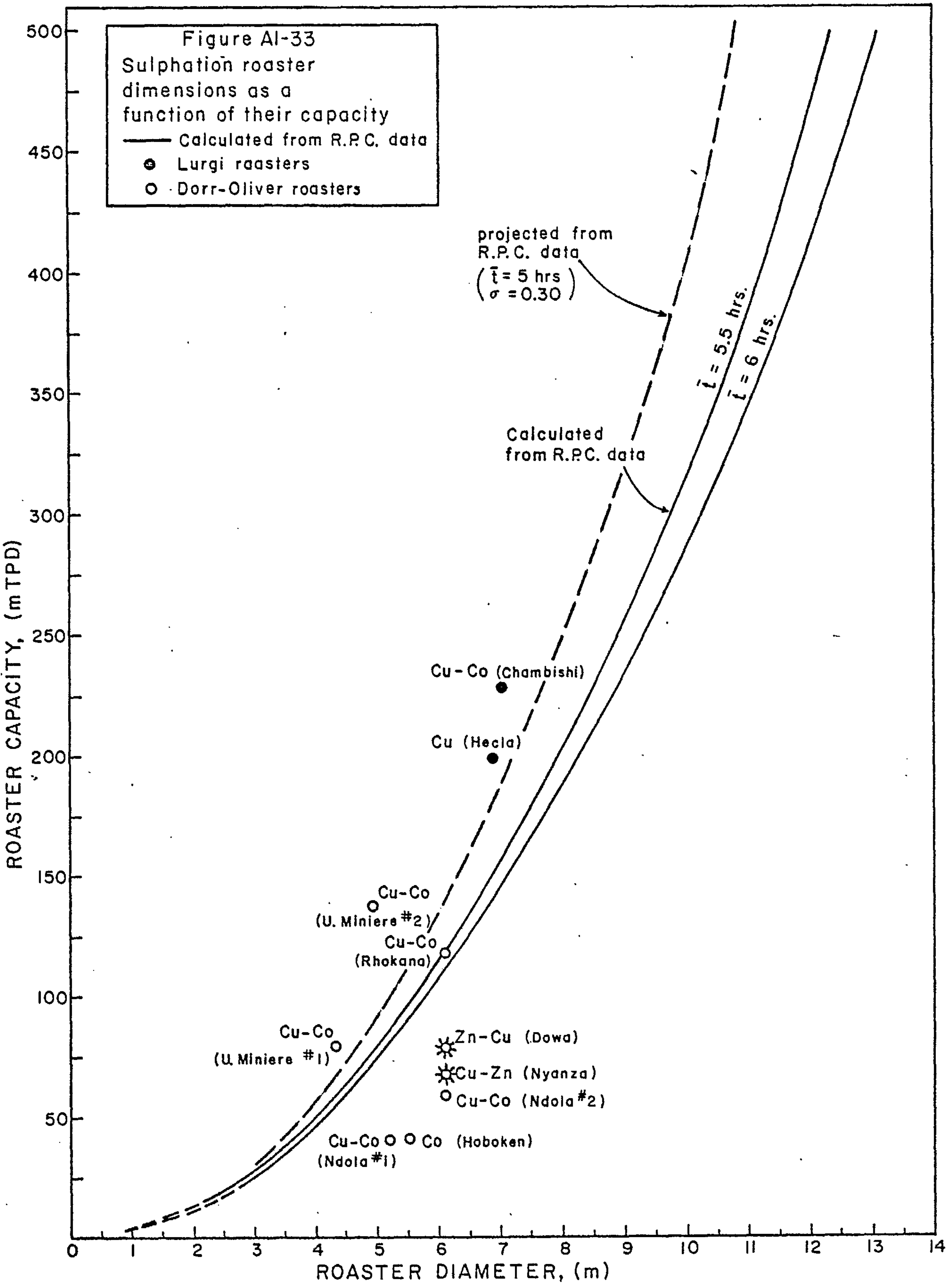


Figure AI-33
Sulphation roaster dimensions as a function of their capacity

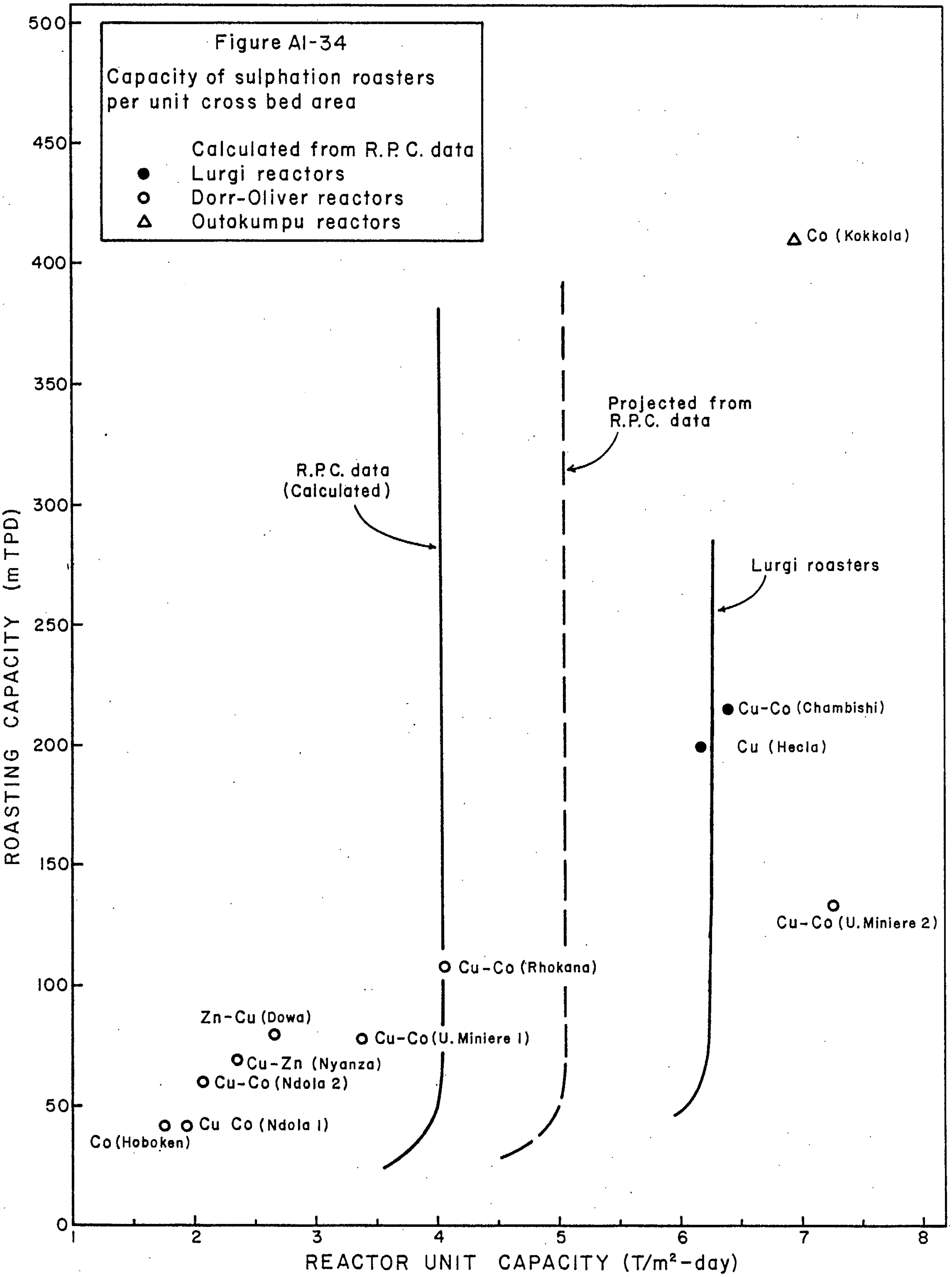


Figure AI-34

Capacity of sulphation roasters per unit cross bed area the Dowa Zn-Cu plant, which used Dorr-Oliver roasters, is small in comparison to what would be required for a sulphation roast zinc plant as determined using the present experimental data (4.1 MT/m^2-day).

.7 Elutriated Solids Collection and Gas Cleaning

The elutriated solids collection equipment used in a commercial sulphation roaster would be similar to that used in conventional dead roast systems.

A one or two stage hot cyclone system with a combined collection efficiency of 90% would be employed with discharge of elutriated material back to the bed through rotary star valves. The off-gases exiting the cyclones at 650°C or less (depending on the heat recovered in the freeboard for decomposition of the HAL solution) are further cooled to 350°C in a waste heat boiler. Since the entrance temperature of the waste heat boilers is low (relative to a dead roast system), no radiation section is required in the waste heat boiler. The waste heat boiler with no radiation zone would be similar to that presently employed by Outokumpu in their cobalt sulphation plant (7). The waste heat boiler would recover from 0.4 to 0.6 tons of steam per ton of feed; depending upon the composition of the feed and the gas temperature. This satisfies a sizable amount of the total steam requirements in the hydrometallurgy section (Appendix A4). The waste heat boiler would also collect about 2% of the total dust elutriated from the reactor.

An electrostatic precipitator operating at 350 to 300°C, and similar to that employed at the cobalt sulphation plant of Outokumpu, is the final gas cleaning stage. Approximately 8% of the elutriated dust is collected in the electrostatic precipitator. This material and that derived from the cyclones and waste heat boilers represents an overall collection efficiency of 99.5%. Dust collected from the cyclones, waste heat boilers and electrostatic precipitators is relatively high in soluble iron and hence, in all probability, must be recycled to the roaster.

Operation of the electrostatic precipitator at 350°C versus the 400 to 450°C employed in the mini-pilot runs may result in a buildup in arsenic in the calcine due to the recycle of precipitator dust to the roaster. It is proposed that, if further testing indicates such a buildup, the electrostatic precipitator dust be "neutral" leached separately to avoid dissolution of the As_2O_3 and flocculated and thickened. The residue would be incorporated with the scrubber acid, and the clarified liquor recycled to the roaster with the HAL solution or used as a base solution for the hot acid leach step.

Off-gases exiting from the electrostatic precipitator are cooled in a scrubber by vaporizing water and condensing the resulting acid water before entering the acid plant. Virtually all impurities, such as arsenic, selenium, chlorine and mercury, which are volatilized during roasting are removed in the acid water bleed.

A1.3 HYDROMETALLURGY SECTION

131 Development of Process-Hydrometallurgy Section

.1 Preliminary Leach Tests

The purpose of these trials was to examine the leaching characteristics of sulphation roast calcines under conditions similar to those employed in conventional zinc plant practice. The pilot plant experience of St. Joe Lead had indicated that problems in obtaining simultaneously high zinc and copper extraction, and low iron and arsenic extractions may be experienced. The low roaster temperatures (630°-660°C) and high leaching temperatures (>70°C) and acid tenors (5-20 gpl) employed at St. Joe resulted in zinc and copper extractions in dilute acid leaching of greater than 90 per cent but iron levels in solution above 7 gpl resulted in very poor filtration and high zinc losses. Accordingly, a leaching program was drawn up to determine a set of operating roaster and leach parameters suitable for sulphation roast conditions.

.11 Cold, Dilute Acid Leaching Tests at Constant pH

Samples of various calcines were leached at constant pH and at initial pulp densities estimated to produce a leach filtrate of 150 gpl zinc. Batch leaches ranged in volume from 10 to 20 litres. Agitation was maximized to the first indications of vortexing. Spent electrolyte was added continuously from a buret to control the pH at the desired value. All solutions were filtered in a Denver Pressure Filter at 40 psi with a Propex 2 filter cloth. The effects of temperature, retention time, pH, calcine roasting temperature, and calcine composition on zinc, copper, arsenic, cadmium, cobalt and iron extraction were studied. The results are given in Table A1-16.

.111 Effects of pH, Temperature, and Retention Time

The mean neutral leach temperature experienced in leaches conducted in insulated vessels with no external heating, using calcines and solutions at room temperature, was 45°C. The extraction of copper and arsenic between 45° and 70°C increased dramatically when the pH was decreased from 4.7 to 1.0 as shown in Figures A1-35 and A1-36. Iron extractions were only slightly dependent on pH over this range at 45°C, however, at 70°C the effect on iron extraction (Figure A1-37) was similar to copper. Zinc extractions were basically independent of pH below 4.0 at 45°C and only slightly dependent at 70°C. It appears the rise in zinc and iron extractions with temperature and pH are correlatable and due primarily to zinc ferrite extraction.

Copper extraction showed a strong positive correlation with retention time (Figure A1-35). At 45°C, zinc and iron extractions were independent of retention time for retention times over 1.25 hours and pH below 4.0. At 70°C, zinc and iron extractions exhibited increasingly stronger correlation with retention time as the acid tenor in the leach was in-

TABLE A1-16

CONDITIONS AND RESULTS OF PRELIMINARY LEACH TESTS AT CONSTANT pH

Test #	Calcine	Background Solution	Acidifying Reagent	Average Temp.(°C)	Ret. Time (hr)	pH	Filtrate Composition (gpl)					
							Zn	Cu	Cd	Co	As(mgpl)	Fe
1	B2C-12	water	spent	45	1.5	4.0	119	0.10	0.21	0.12	<1.0	0.80
2	B2C-27	neutralized spent:water (4:1)	"	45	1.5	4.5	145	0.08	0.22	0.10	<0.1	0.67
3	"	"	"	45	3.0	4.5	157	0.58	0.24	0.11	0.2	0.68
4	"	"	"	45	3.0	3.8	162	0.78	0.23	0.10	0.2	0.68
5	"	"	"	45	3.0	3.8	160	0.78	0.23	0.10	0.2	0.82
6	"	"	"	45	3.0	3.0	166	1.26	0.23	0.10	0.2	1.48
7	"	"	"	45	3.0	1.8	166	1.53	0.23	0.10	40.0	1.42
8	"	"	"	45	3.0	4.5	159	0.58	0.23	0.10	0.4	0.68
9	B2C-24	"	"	45	1.5	4.7	137	0.10	0.21	-	-	0.80
10	B2C-26	"	"	45	1.5	3.8	147	-	-	-	-	0.92
11	BMD-23	water	"	45	1.5	3.9	147	2.06	-	-	-	2.30
12	BMD-30	"	"	63	1.5	3.9	120	0.72	-	-	-	0.11
13	BMD-17	"	"	45	1.5	4.0	149	2.87	-	-	-	5.80
14	B2C-14	partially neutralized spent (50 gpl acid)	-	70	3.0	1.0	131	0.80	0.17	0.08	188.0	6.25
15	B2C-22	"	-	70	3.0	3.3	145	0.24	0.16	0.08	0.5	2.10
16	BMS-13	"	-	70	1.5	1.3	131	1.58	0.16	0.13	625.0	6.00
17	BMS-13	"	-	70	3.0	1.3	141	1.89	0.16	0.15	375.0	5.12

TABLE A1-16 (Cont'd)

CONDITIONS AND RESULTS OF PRELIMINARY LEACH TESTS AT CONSTANT pH

Test #	Calcine	Residue Composition (wt %)				Extraction (wt %)					
		Zn	Cu	Co	Cd	Total*	Zn	Cu	Cd	Co	As
1	B2C-12	15.0	0.45	–	–	41	69	7.0	>95	>95	<1
2	B2C-27	15.5	0.43	0.02	0.02	41	68	4.3	>95	>95	<1
3	"	–	–	–	–	–	74	33.0	–	–	–
4	"	–	–	–	–	–	76	44.0	–	–	–
5	"	–	–	–	–	–	76	44.0	–	–	–
6	"	–	–	–	–	–	79	67.0	–	–	–
7	"	–	–	–	–	–	79	87.0	–	–	–
8	"	–	–	–	–	–	75	33.0	–	–	–
9	B2C-24	15.0	0.44	0.02	<0.01	40	67	5.7	>95	>95	<1
10	B2C-26	13.0	0.36	–	–	44	73	10.0	–	–	–
11	BMD-23	–	–	–	–	–	–	36.0	–	–	–
12	BMD-30	–	–	–	–	–	–	13.0	–	–	–
13	BMD-17	–	–	–	–	–	–	62.0	–	–	–
14	B2C-14	13.5	–	–	–	–	75	59.0	–	–	20
15	B2C-22	13.0	0.40	–	–	–	67	18.0	–	–	<1
16	BMS-13	9.3	0.16	–	–	–	85	73.0	–	–	75
17	BMS-13	9.3	0.17	–	–	–	86	88.0	–	–	50

$$* \quad \text{Total} = \frac{\text{wt original calcine} - \text{wt. residue}}{\text{wt original calcine}} \times 100$$

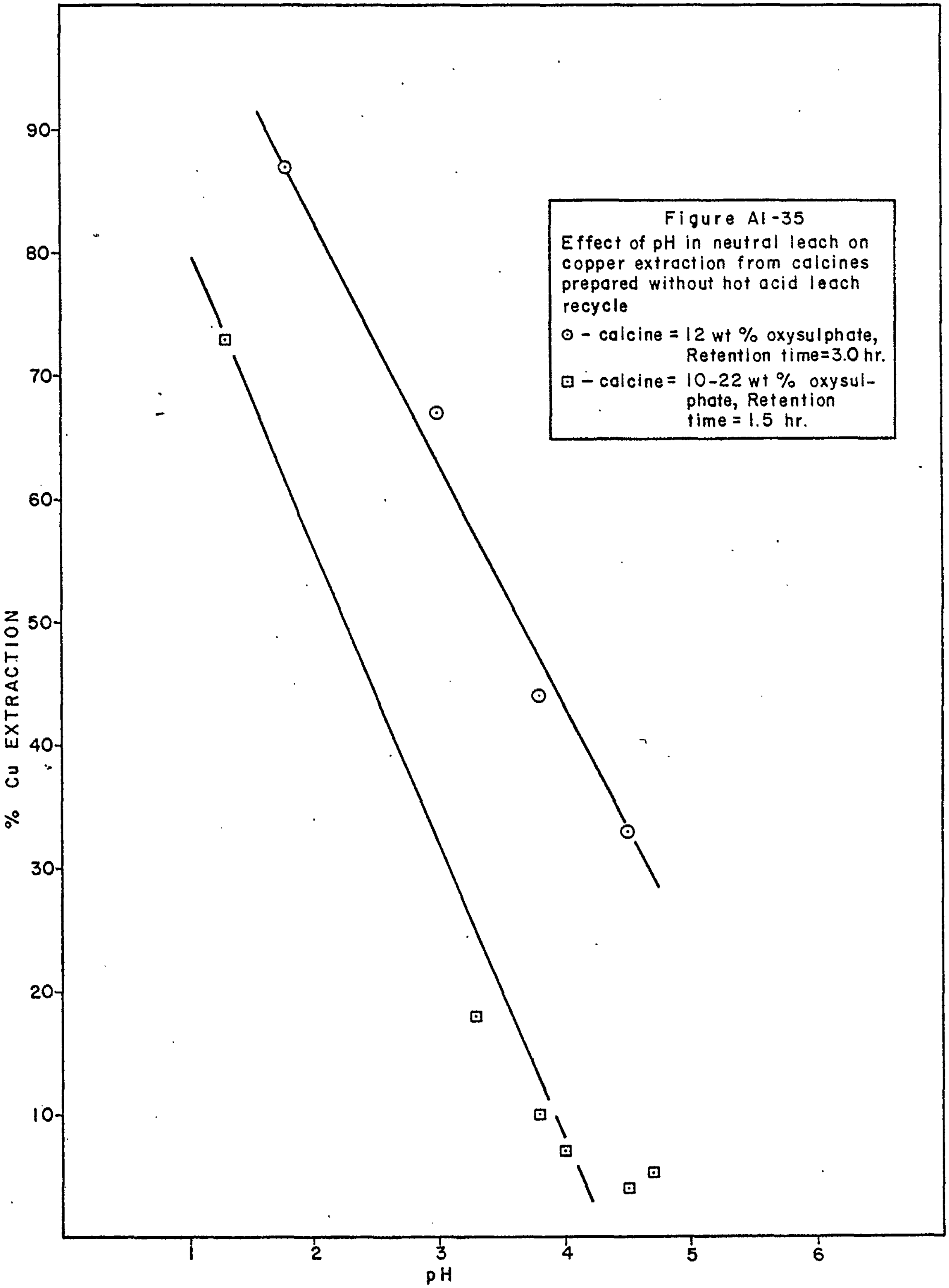


Figure AI-35
Effect of pH in neutral leach on copper extraction from calcines prepared without hot acid leach recycle

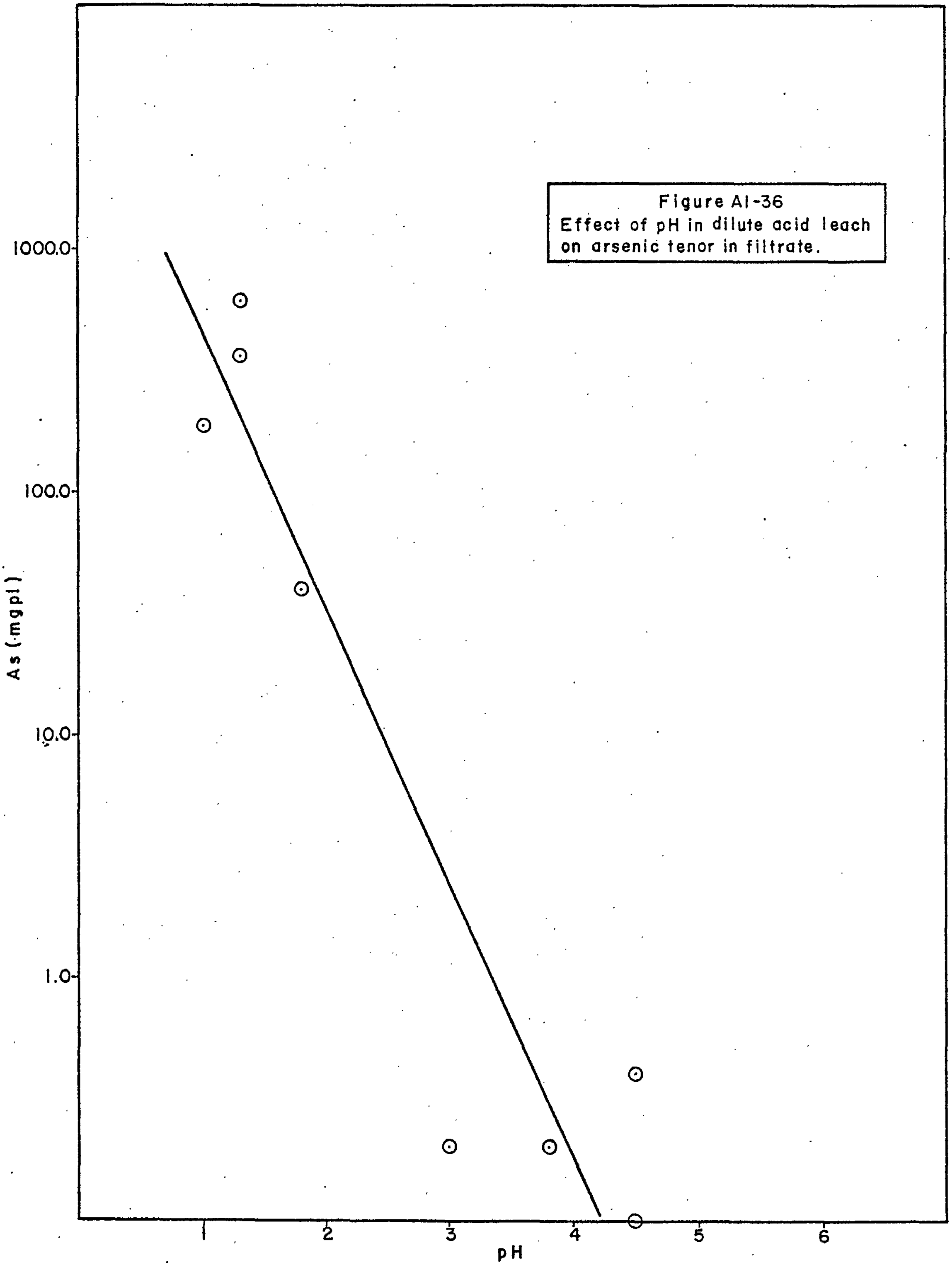


Figure A1-36
Effect of pH in dilute acid leach on arsenic tenor in filtrate.

creased. This is again indicative of zinc ferrite dissolution.

Cadmium and cobalt exhibited greater than 95 per cent dissolution, independent of pH, temperature, or retention time over the ranges studied.

.112 Effects of Calcine Composition and Roasting Temperature

Zinc extraction, as expected, was found to be extremely dependent on zinc distribution in the calcine. Virtually all zinc as sulphate and oxysulphate was extracted at all temperatures studied, when the pH was controlled below 4.0 and the retention time longer than 1.25 hours. Zinc as ferrite and sphalerite was found to be unleachable except under conditions of high temperature, low pH, and long retention time. Copper extraction at constant pH and retention time was found to be an inverse function of the amount of zinc oxysulphate in the calcine as shown in Figure A1-38. It is proposed that the following solid state reaction in the calcine is responsible for this effect.

$$ZnO.2ZnSO_4 + 2CuSO_4 \rightarrow CuO.CuSO_4 + 3ZnSO_4$$

Zinc and copper extractions were found to increase significantly when calcines were produced at low roaster temperatures. This was attributed to a shift in the distribution of zinc in the calcine as the temperature was decreased,

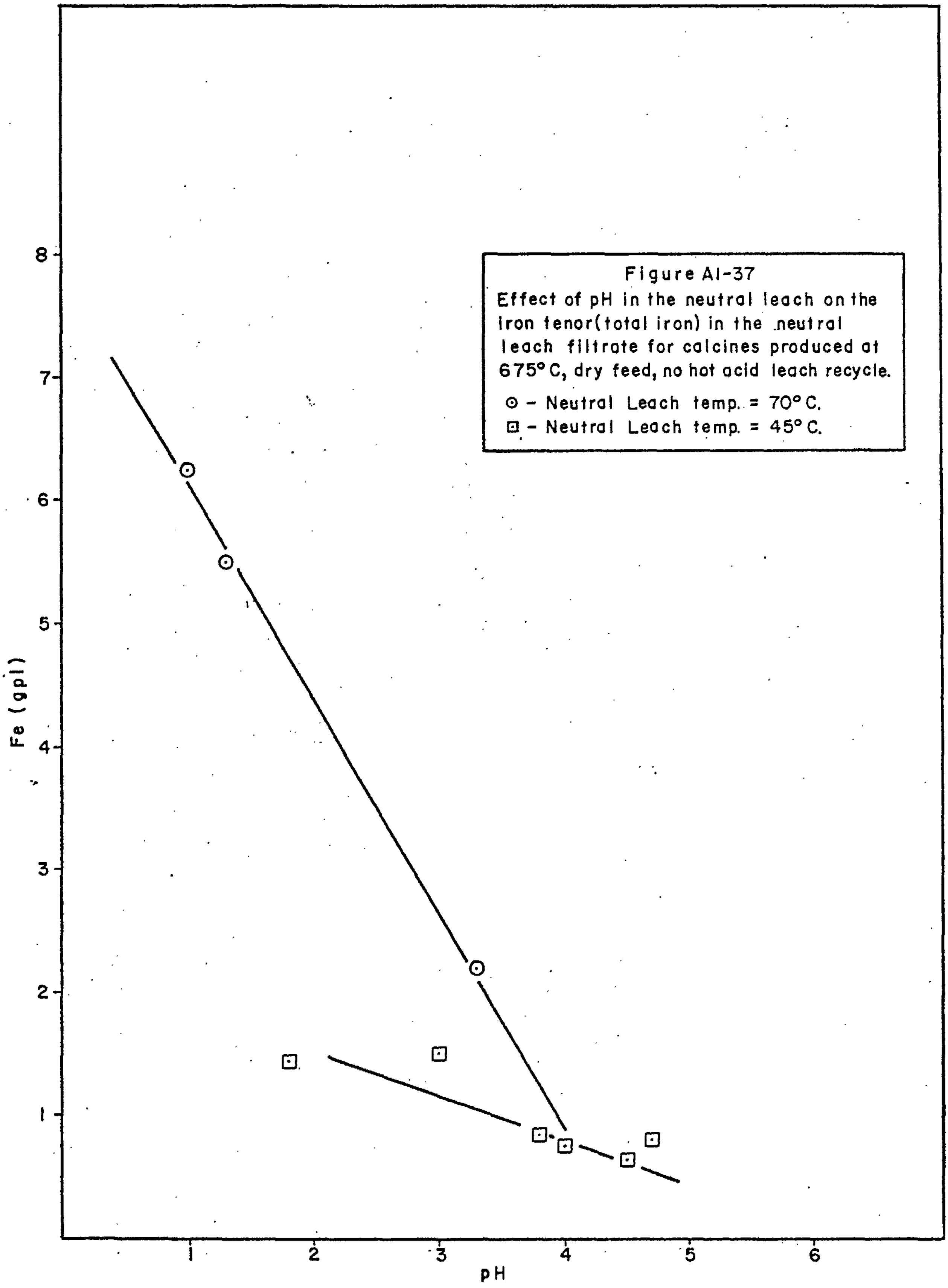


Figure Al-37

Effect of pH in the neutral leach on the iron tenor(total iron) in the neutral leach filtrate for calcines produced at 675° C, dry feed, no hot acid leach recycle.

to a larger fraction of zinc sulphate relative to oxysulphate and ferrite. However, a sharp increase in iron extraction in the dilute acid leach was evident as the roaster temperature was decreased (Figure A1-39). This was attributed to the increased stability of iron sulphates at the lower roaster temperatures.

Cadmium and cobalt extractions were found to be independent of calcine composition.

.12 Hot Acid Leaching of Dilute Acid Leach Residues

It became obvious from the dilute acid leach experiments that no single set of conditions in a single stage leach would produce the desired results of high zinc and copper extraction and very low iron and arsenic extractions unless a calcine of less than 5 per cent ferrite could be consistently produced at a roaster temperature above 675°C. The minimum ferrite level obtained in the 14-inch roaster appeared to be approximately 7-15 wt%.

With ferrite levels in calcines above 7 wt% and sphalerite at 1.5 wt%, it was necessary to consider a second stage hot acid leach of the 'neutral' leach residue to increase zinc extraction. However, dissolution of ferrite results in high iron levels in the leach filtrate and hence the requirement for a removal of iron from solution procedure. Conventional approaches such as the Jarosite process; or autoclave precipitation etc., tend to be both capital and energy intensive procedures, the incorporation of which

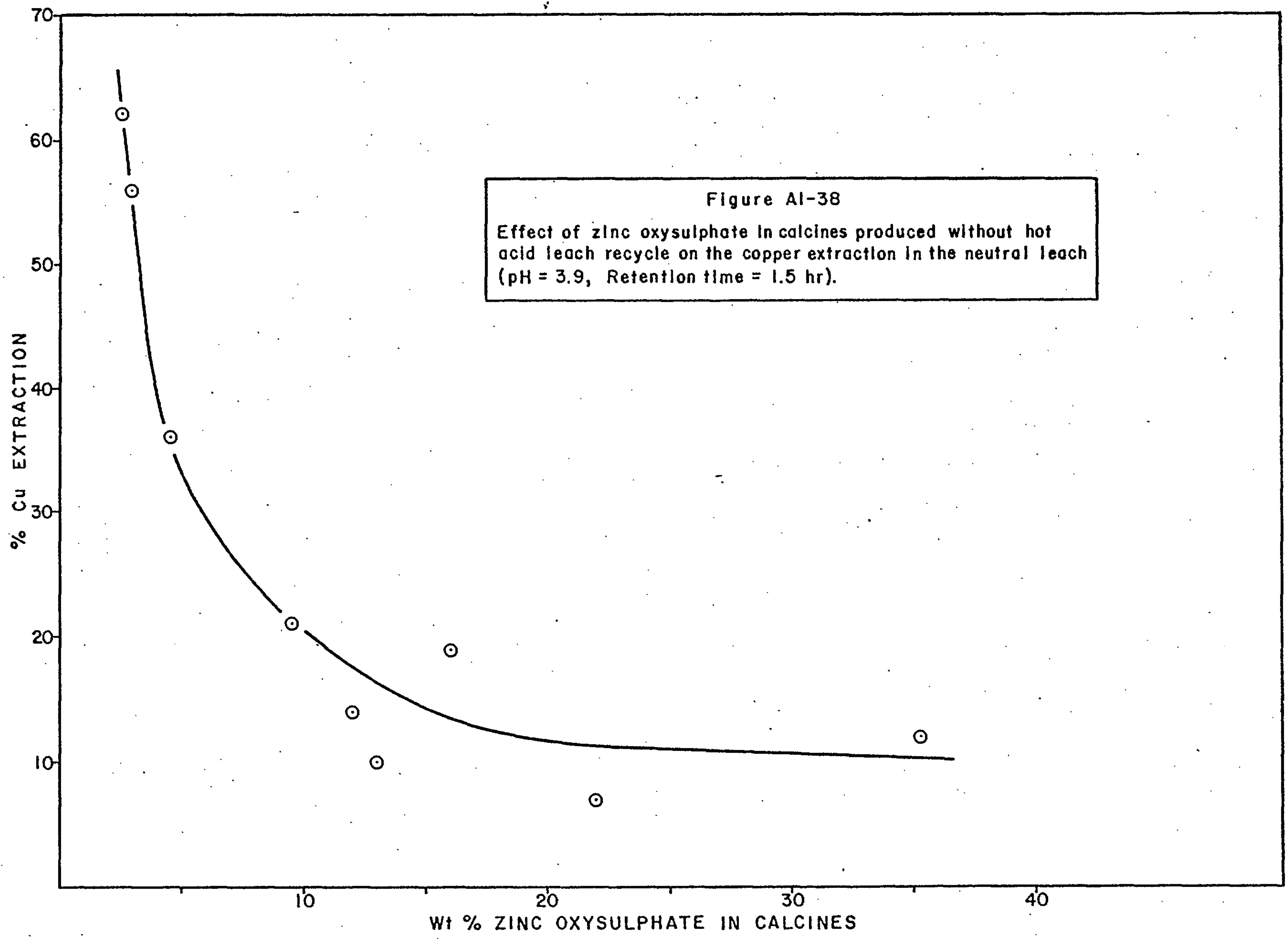


Figure Al-38

Effect of zinc oxysulphate in calcines produced without hot acid leach recycle on the copper extraction in the neutral leach (pH = 3.9, Retention time = 1.5 hr).

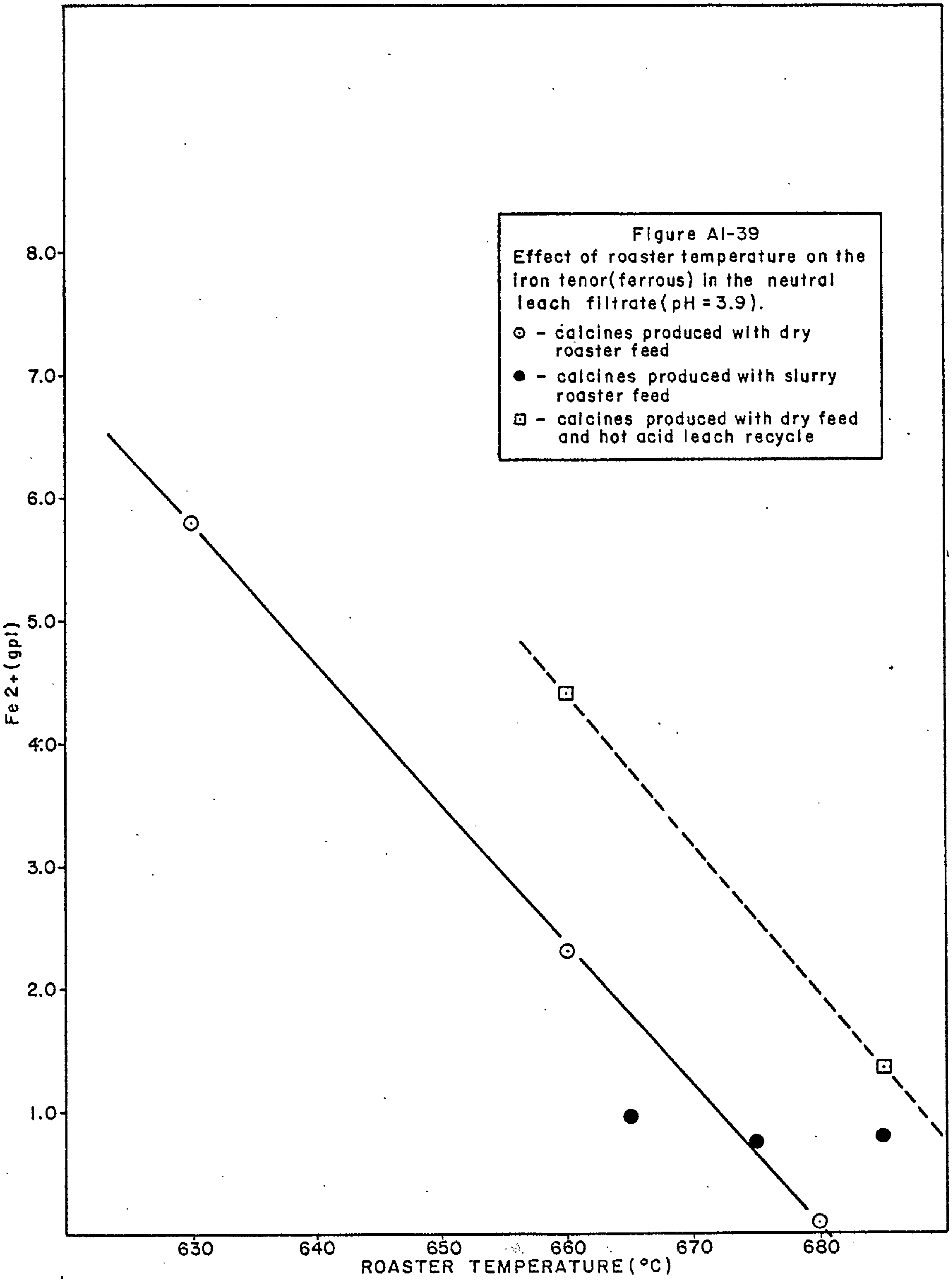


Figure AI-39
Effect of roaster temperature on the iron tenor(ferrous) in the neutral leach filtrate(pH = 3.9).

would limit the competitiveness of the Sulphation Roast Process. As an alternative, the possibility of decomposing the strong acid leach filtrate, recycled as a coolant, in the sulphation roaster was studied, as discussed in Section A1-2. Under the relatively low temperature conditions peculiar to the sulphation process, zinc and copper sulphates in solution are dehydrated and remain as sulphates in the roaster calcine. The iron sulphates in the filtrate decompose to inert hematite. Accordingly, little iron (less than 1 gpl) enters solution when the calcine is subjected to a neutral leach. Filtering of the hot acid leach solution is facilitated by the large amount of hematite in the residue.

Decomposition of the hot acid leach liquor in the reactor enables the leach to be run at a much higher iron tenor. The volume of solution, relative to that required in conventional zinc plant practice, is reduced and hence the energy input necessary to bring the solution to the reaction temperature can be kept to a minimum.

A volume balance can be established whereby the total leach filtrate can be decomposed in the roaster with the reactor operated autothermally. Some preliminary leaching tests with filtered and washed 'neutral' leach residues were undertaken in solution volumes which were calculated so as to maintain thermal equilibrium in the roaster on recycling. The effect of temperature, acid concentration, and retention time on zinc extractions were studied. It was determined that a leaching temperature of 95°C, a controlled pH of 0.2 (approximately 40 gpl free acid) throughout the leach, and a retention time of 3.0 hours were necessary to insure greater than 90 per cent extraction of zinc from the neutral leach residues. Over 98% of copper was extracted over all conditions studied. Results are shown in Figure A1-40 and Table A1-17.

.121 Hot Acid Leach Tests on Synthetic Ferrite and Sphalerite

The co-leaching behaviour of ferrite and sphalerite was examined as a function of acid and retention time in a 140 gpl zinc background solution. Results are given in Table A1-18 and Figures A1-41 and A1-42. The leaching of ferrite was shown to be relatively independent of acid concentration as long as acid tenors exceeded 40 gpl throughout the experiment. Effectively 95% of the ferrite was leached in two hours under these conditions. Sphalerite dissolution was strongly dependent on the ferric ion concentration in the solution and to a very minor extent on the acid tenor. The rate of reaction was much slower than for ferrite dissolution.

The dissolution of zinc ferrite as a function of iron concentration in solution at 95°C, greater than 50 gpl acid in solution, 140 gpl zinc in solution, and a two hour retention time was studied. The results are given in Table A1-19. Virtually 100 per cent extraction of zinc and iron from ferrite was observed for iron backgrounds between 25 and 120 gpl.

TABLE A1-17

CONDITIONS AND RESULTS OF PRELIMINARY HOT ACID LEACH TESTS
ON DILUTE ACID LEACH RESIDUES

Test #	Residue Leached[a]	Repulp Solution (background)	Acidifying Reagent	Temp. (°C)	Ret. Time(hr)	pH	Filtrate Composition(gpl)			Zn Extraction %	
							Zn	Acid	Fe	Hot Acid Leach	Overall (dilute acid leach + hot acid leach)
1	B2C-26	spent electrolyte[b]	raw acid	87	1.5	0.5	89	16	-	12.2	76.6
2	B2C-26	water	" "	89	1.5	0.5	49	17	-	30.0	81.4
3	B2C-26	"	" "	93	3.0	0.2	157	35	-	80.5	94.8
4	B2C-26	"	" "	96	3.0	0.2	183	42	-	92.9	98.1
5	BMD-23	spent + wash solution[c] ratio 1:2	" "	95	3.0	0.2	159	40	199	100.0	100.0

a - residues were washed thoroughly with water and contained approximately 30% by weight water.
b - spent electrolyte contains 50 gpl zinc and 170 gpl acid.
c - wash solution contains 200 gpl iron, 160 gpl zinc, and 80 gpl acid.

TABLE A1-18

CONDITIONS AND RESULTS OF SIMULTANEOUS SPHALERITE-SYNTHETIC FERRITE BATCH LEACH TESTS IN 140 gpl ZINC SOLUTION

Test #	Vol	Additions (gm)			Final Residue Composition (gm)		Ferrite Extraction %			Sphalerite Extraction %		
	(lt)	Ferrite	Sphalerite*	Acid	Total Mass	S°	1 hr	2 hr	3 hr	1 hr	2 hr	3 hr
1	0.550	10	3.07	20	4.95	0.50	55	68	75	28	41	50
2	0.550	10	3.07	40	2.71	0.64	92	94	96	44	54	63
3	0.550	10	3.07	60	2.35	0.61	92	96	97	44	55	61
4	0.525	10	3.07	80	2.39	0.64	93	99	99	47	62	64
5	0.525	10	3.07	120	2.30	0.63	98	98	98	50	62	63
6	0.500	5	3.07	120	2.64	0.45	87	91	97	23	31	45
7	0.575	15	3.07	120	2.67	0.57	100	96	100	37	39	56
8	0.575	20	3.07	120	1.36	0.36	92	93	100	32	43	63
9	0.550	30	3.07	120	2.22	0.69	85	100	99	34	60	72
10	0.550	50	3.07	120	2.48	0.87	100	96	100	53	70	82

* 4.04 gm of high grade zinc concentrate was added as the source of sphalerite.

Note: all reagents were added initially by batch; samples were taken every 30 minutes for analyses of Fe^{3+}, Fe^{2+}, and elemental S. Temperature was controlled at 95±3°C; background solution was 140 gpl zinc and 4 gpl copper.

TABLE A1-19

CONDITIONS AND RESULTS OF SYNTHETIC FERRITE AND SPHALERITE BATCH LEACH TESTS IN HIGH ZINC-HIGH IRON SOLUTIONS

Test #	Vol	Additions(gm)		Initial Background Sol'n(gpl)				Final Residue		Ferrite Extraction %	Sphalerite Extraction %
		ferrite	sphalerite	Zn^{2+}	Fe^{3+}	Cu^{2+}	Acid	Total Mass (gm)	Water Insoluble Zinc(gm)		
1	0.55	50	0	140	25	4	200	9.12*	0.04	99.7	-
2	0.55	50	0	140	50	4	200	13.00*	0.04	99.7	-
3	0.55	50	0	140	80	4	200	24.00*	0.04	99.7	-
4	0.55	50	0	140	120	4	200	25.93*	0.03	99.8	-
5	0.55	50	0	140	120	4	200	25.07*	0.03	99.8	-
6	0.55	100	0	140	120	4	364	25.93*	0.05	99.6	0
7	0.55	0	15	110	110	4	45	2.85	1.91	-	81

* solutions were supersaturated and contained precipitated zinc and iron sulphate; cooling during vacuum filtration increased the sulphate precipitation.

Note - all experiments lasted two hours at 95± 3°C;
all reagents were added initially by batch.

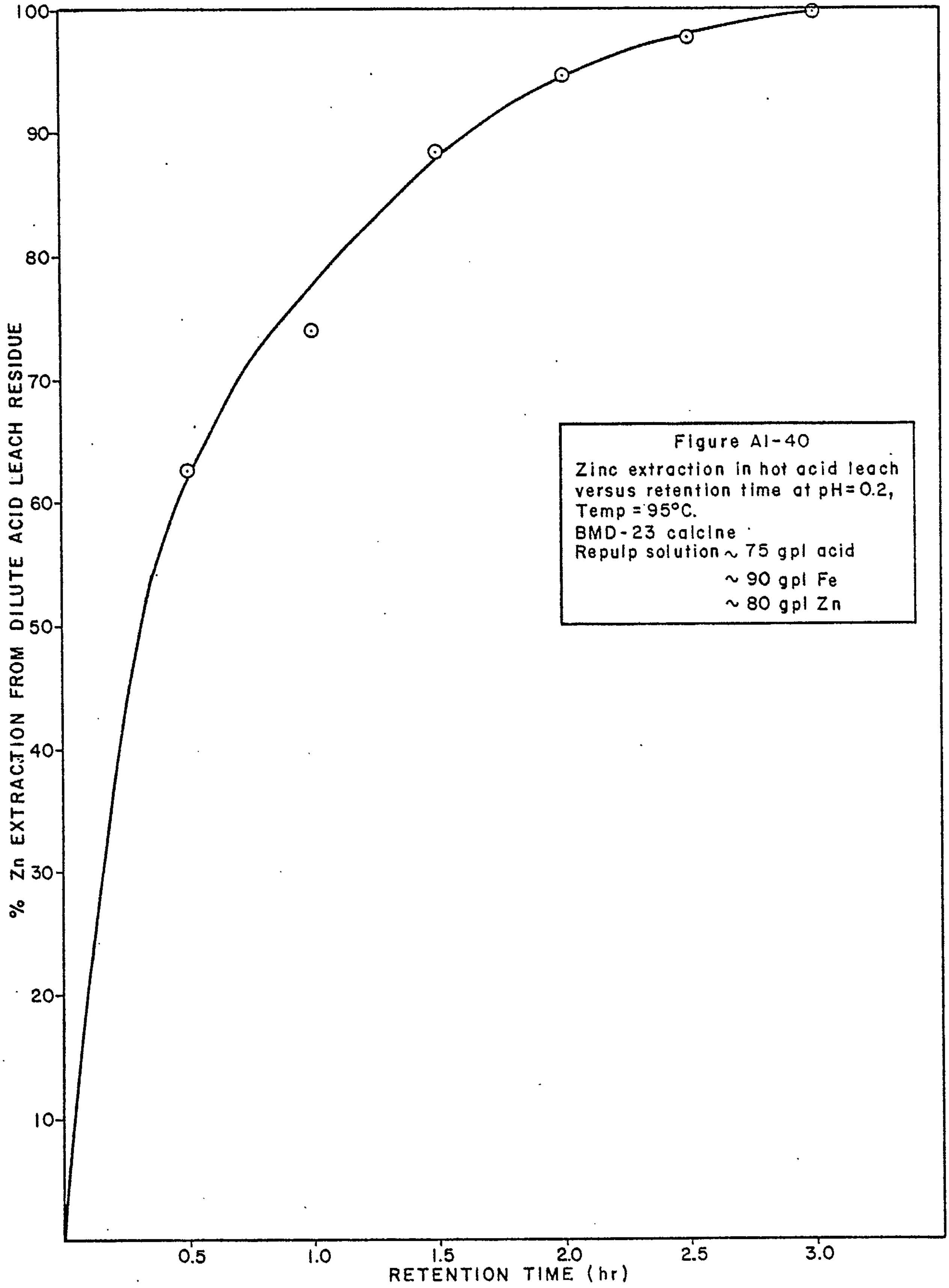


Figure A1-40

Zinc extraction in hot acid leach versus retention time at pH = 0.2, Temp = 95°C.

BMD-23 calcine

Repulp solution ~ 75 gpl acid
~ 90 gpl Fe
~ 80 gpl Zn

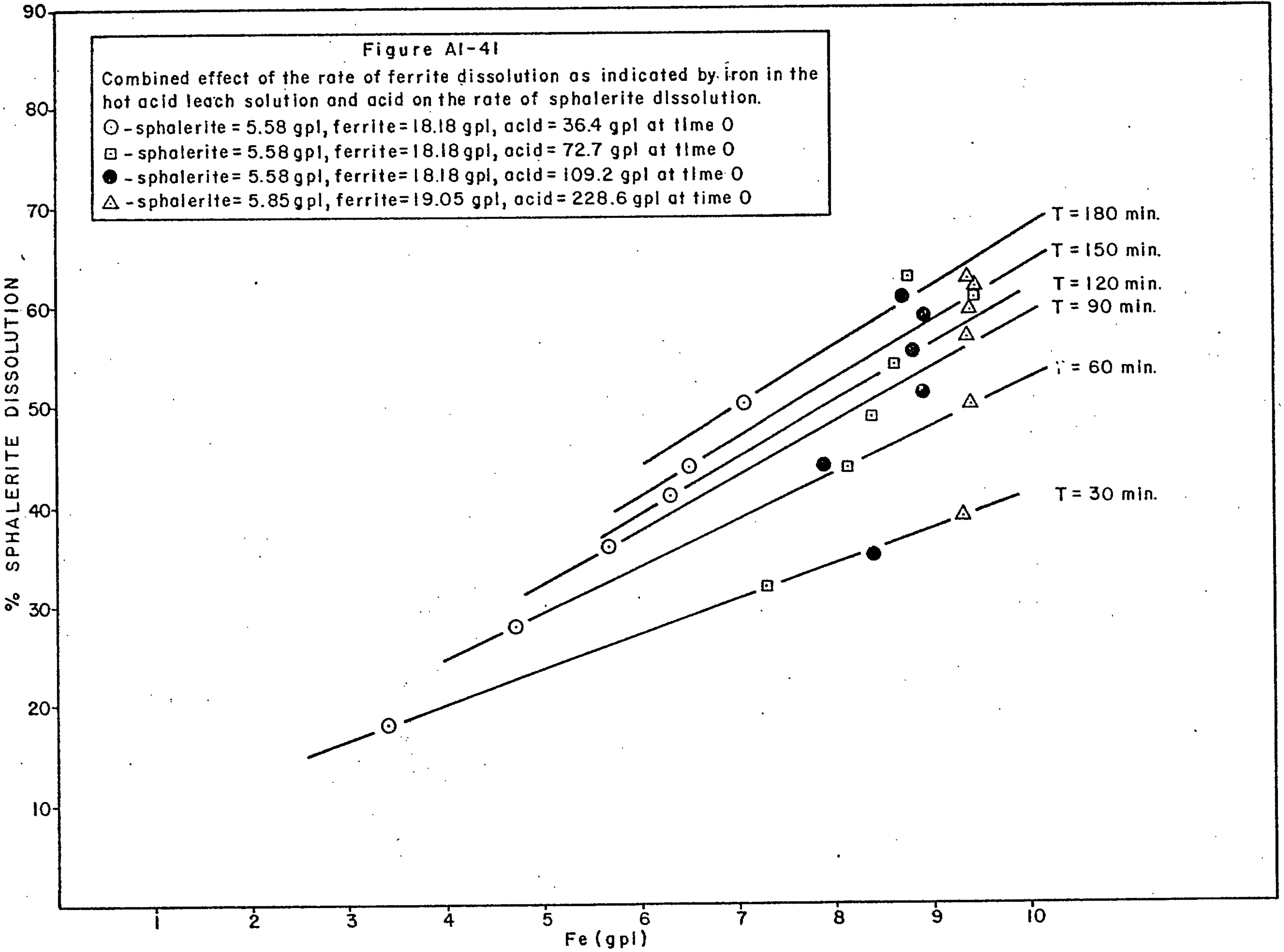


Figure AI-41

Combined effect of the rate of ferrite dissolution as indicated by iron in the hot acid leach solution and acid on the rate of sphalerite dissolution.

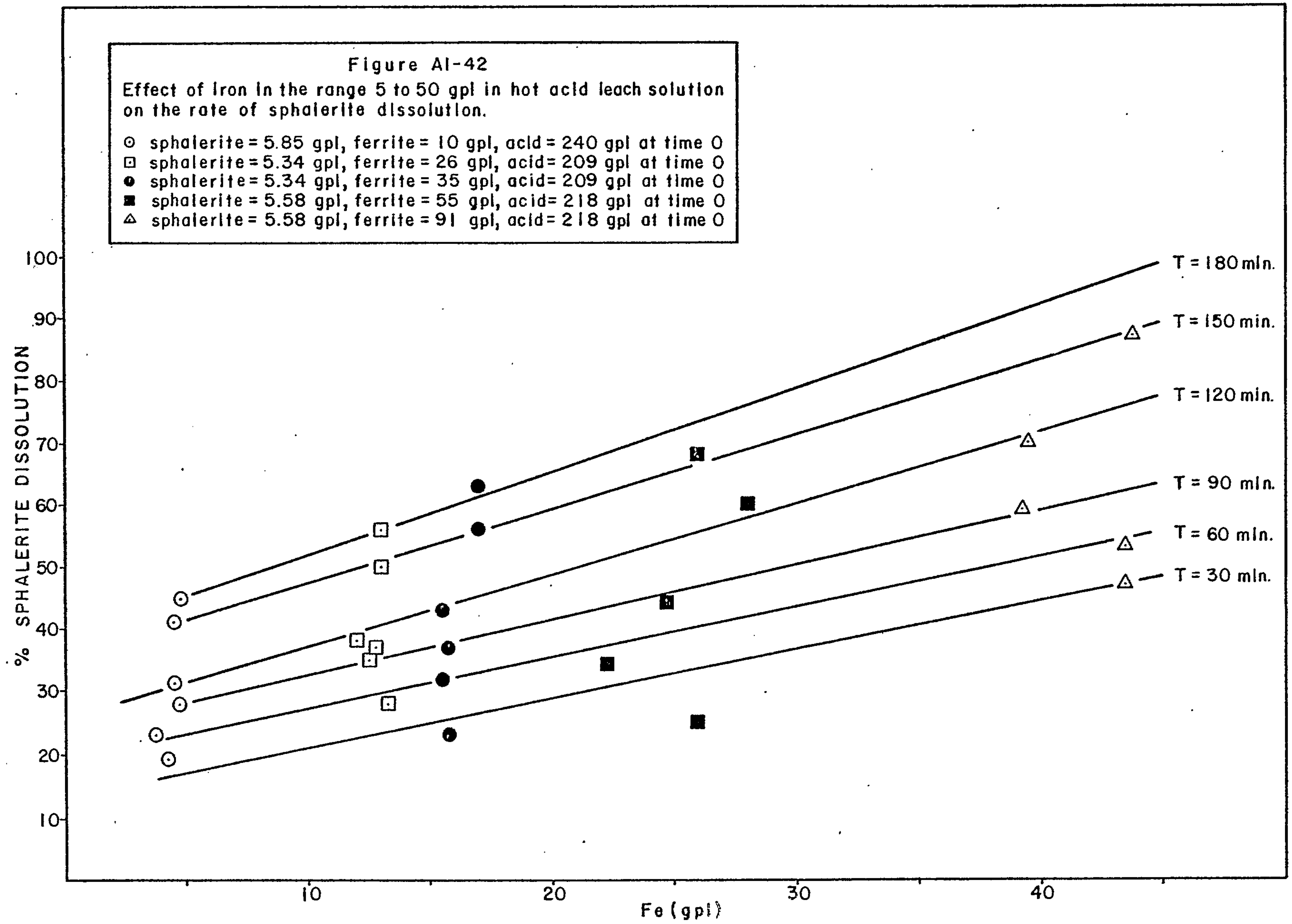


Figure AI-42

Effect of iron in the range 5 to 50 gpl in hot acid leach solution on the rate of sphalerite dissolution.

The extraction of zinc from sphalerite in solution containing 110 gpl zinc, 110 gpl iron, and 45 gpl acid was measured after two hours retention time at 95°C. The percentage sphalerite leached was determined to be 81 per cent.

.13 Dilute Acid Leach Tests on Calcines Prepared with Hot Acid Leach Recycle

Several calcines were prepared using a synthetic hot acid leach recycle as coolant in the roaster and their behaviour during dilute acid leaching examined. Zinc extractions increased due to the reduced ferrite and oxysulphate in the calcine. Copper extractions increased markedly to greater than 50 per cent. Iron extraction, however, more than doubled from previous tests as shown in Figure A1-39. A sample of one calcine was screened and the larger fraction (+100 mesh) which was attributed mostly to the pellet formation of the hot acid leach recycle was leached under the same conditions as the total calcine. Iron extraction nearly doubled and copper extraction was over 90 per cent indicating that the increase in iron in solution was due to the hot acid leach recycle and that the zinc in the recycle was forming primarily zinc sulphate with little or no oxysulphate. Results are given in Table A1-20. Filtration rates of these neutral leach solutions were quite acceptable at 0.8 $lt/min/ft^2$ to a 2-inch thick filter cake.

TABLE A1-20

CONDITIONS AND RESULTS OF PRELIMINARY DILUTE ACID LEACH TESTS (CONSTANT pH) ON CALCINES PREPARED WITH HOT ACID LEACH RECYCLE

Calcine Leached	pH	Ret. Time(hr)	Background Solution Composition	Temp (°C)	Filtrate Composition (gpl)			Filtration[a] Rate (lt/min/ft^2)	Leach Recycle Extraction (%)	
					Zn	Cu	Fe		Zn	Cu
BMD-20	3.9	1.5	water	60	150	2.58	4.40	0.8	88.0	56
BMD-41	3.8	1.5	"	55	146	-	1.38	-	-	-
BMD-41 (+100 mesh fraction)	3.8	1.5	"	55	151	-	2.08	-	-	91

a - filtration done on Denver pressure filter at 40 psi.

.2 Iron Oxidation and Precipitation on Solutions from Preliminary Leach Tests

Prior to electrowinning zinc from a leach solution, all metallic impurities in the solution which are more electropositive than zinc must be removed to acceptably low levels. Most impurities may be removed from solution by cementation on zinc dust. Iron is one impurity which will not cement effectively, in fact the presence of iron in solution during the cementation of other impurities, hinders these reactions and increases the consumption of zinc dust due to forward and reverse ferrous-ferric oxidation-reduction reactions. In order to electrowin consistently high purity zinc at high current efficiency iron should be kept below 25 mgpl in the electrolyte.

The removal of iron from solution in most zinc plants is accomplished by first oxidizing all soluble iron to the ferric state which immediately precipitates as colloidal iron hydroxide at pH 3.6 to 5.2, the pH of most solutions in 'neutral leach', leaving less than 10 mgpl iron in solution. The oxidation of iron may be accomplished by sparging the solution with air, oxygen, or ozone or adding a chemical oxidant such as hydrogen peroxide, manganese dioxide or potassium permanganate. Oxygen and ozone sparging, although extremely effective, require a cost of production and increases the corrosion rate of equipment. Hydrogen peroxide has a tendency to flash and produce large amounts of basic zinc sulphate. Air sparging with no associated reagent costs and imparting no loading factor to the solution would appear to be most suitable but long retention times and difficulty in control make its implementation as a sole oxidant most unreliable. Potassium permanganate, although quite expensive, is the most reliable oxidant especially in near neutral solution. Consumption of potassium permanganate necessary for iron removal, however, is very dependent on the sphalerite content of the calcine. One advantage of permanganate use is that manganese in solution is desirable for decreasing lead anode corrosion, thereby increasing zinc purity and reducing anode replacement costs. Manganese dioxide, although relatively inexpensive, is a slow oxidant for iron unless very finely divided due to its insolubility and lower reduction potential. At higher acidities, however, it is an excellent oxidizing agent for ferrous ion. Accordingly, the combination of air sparging and potassium permanganate addition would appear to be the logical choice for iron oxidation in neutral leach solutions.

A series of experiments were run to study the effects of these oxidants on filtered neutral leach solutions from the preliminary trials. The results of these tests are given in Table A1-21.

Before any conclusions can be drawn on an oxidation system for use, several other aspects must be considered. The high copper levels in a bulk concentrate and the zinc tailings refloat lends the system to Cu removal by solvent extraction. This process must precede zinc dust purification and the solution temperature cannot exceed 50°C. The acidity of the solution increases to around pH 1.5 during

TABLE A1-21

CONDITIONS AND RESULTS OF IRON OXIDATION AND SETTLING TESTS ON PRELIMINARY LEACH SOLUTIONS

Test #	Neutral Leach Filtrate from Calcine #	Initial Fe (gpl)	Temp. (°C)	Ret. Time(hr)	pH	Oxidant Addition(gpl)		Flocculant Addition(mgpl)		Final Fe (mgpl) of Decant Sol'n
						MnO_2	$KMnO_4$	Percol 156	Percol E-24	
1	B2C-12(1)	0.80	45	0.5	4.0	–	0.56	–	–	450[a]
2	" (1)	0.80	45	0.5	4.0	–	2.24	–	1.0	170[a]
3	" (1)	0.80	45	0.5	4.0	–	2.24	1.0	–	110[a]
4	B2C-27(2)	0.67	45	1.0	4.5	–	0.95	1.0	–	6[a]
5	B2C-24(9)	0.80	45	"	4.0[b]	c	–	1.0	–	<1[a]
6	B2C-24 + 27(2 & 9) combined	0.67	42	"	4.2[d]	–	0.92	–	–	<1[g]
7	"	0.67	70	"	4.2[b]	–	0.92	–	–	<1[g]
8	"	0.67	42	"	4.2[d]	–	0.61	–	–	2[g]
9	"	0.67	70	"	4.2[b]	–	0.61	–	–	<1[g]
10	"	0.67	42	"	4.2[d]	–	0.40	–	–	100[g]
11	"	0.67	70	"	4.2[b]	–	0.40	–	–	1[g]
12	"	0.67	42	"	4.2[b]	0.92[e]	–	–	–	500[g]
13	"	0.67	70	"	4.2[b]	0.92[e]	–	–	–	100[g]
14	"	0.67	70	"	4.2[b]	0.92[f]	–	–	–	10[g]
15	"	0.67	70	"	2.5[b]	0.92[e]	–	–	–	<1[h]

a – 50% of solution decanted after 3 hours
b – pH controlled by adding spent continuously
c – MnO_2 sludge from Test #4 was added; Test #4 and Test #5 were on equal solution volumes
d – originally pH fell to 2.8, then climbed slowly back to 4.2

e – electrolytic MnO_2
f – reagent produced from a $KMnO_4$ oxidation test
g – filtered sample
h – filtered sample after limestone neutralization to 4.0
( ) – bracketed numbers indicate leach test numbers (see table A1-16)

this copper-hydrogen exchange and must be increased to a minimum of pH 4 before zinc dust purification. This involves a neutralization step. Test #15 showed that manganese dioxide was a very effective oxidant at pH 2.5 and should be even more effective at pH 1.5. Perhaps the iron oxidation should best take place between solvent extraction and neutralization. Using this system consumption of expensive permanganate by sphalerite* oxidation in the neutral leach and excessive retention time in the neutral leach due to the lower temperature and slower oxidation rate are avoided, and small amounts of manganese are introduced to the system from inexpensive manganese dioxide ore (the majority of the manganese dioxide would come from the cellhouse sludge). After neutralization with limestone, the flocculated and settled MnO_2, gypsum and iron hydroxide sludge could be recycled to the neutral leach thickener to decrease zinc losses, and avoid filtration except when a very large excess of easily filtered hematite was present. The advantages being obvious, the iron oxidation following solvent extraction route was chosen.

.3 Copper, Cadmium, and Cobalt Removal by Cementation on Zinc Dust

The applicability of a conventional two stage zinc dust purification (Vieille Montagne, Canadian Electrolytic Zinc, or Cominco Process) to iron purified sulphation roast calcine,

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* sulphation roast calcines may contain more sphalerite than dead roast calcine.

neutral leach liquors was investigated using batch techniques. All zinc dust employed was obtained from Canadian Electrolytic Zinc and was produced by compressed air atomization of liquid zinc. No further information will be given in this report on its composition or sizing, other than it was a representative sample of purification zinc dust. Results and experimental conditions are given in Table A1-22.

It can be concluded from these experiments that a conventional two stage purification is applicable to these leach liquors but excessive antimony trioxide and zinc dust is required in the second stage. A small recirculating load of coarser cementation product may decrease these additions substantially but further tests are necessary. This optimization would be best achieved in a full pilot scale unit under continuous conditions.

Cementation cake leaching and cadmium, cobalt and copper recovery should be examined thoroughly, simultaneous to purification studies as optimization of purification and optimization of minor metal recovery are completely interdependent. No work was performed on this aspect under the present contract.

.4 Electrowinning Trials on Preliminary Leach Solutions after Purification

A small (7 litre) electrowinning unit was built on a Noranda Research Centre design. Neutral feed solution was mixed with recirculating spent electrolyte in a small cold spent tank and pumped to the cell. The overflow from the

TABLE A1-22

CONDITIONS AND RESULTS OF PRELIMINARY PURIFICATION WITH ZINC DUST OF LEACH SOLUTIONS (B2C-24, B2C-27) AFTER IRON REMOVAL

Test #	Initial Solution Composition (mgpl)				Temp. (°C)	Ret. Time (hr)	pH	Zinc Dust		Sb_2O_3 Addition (mgpl)	Final Solution Composition (mgpl)					Impurities Exhibiting Redissolution With Time
	Zn (gpl)	Cu	Cd	Co				Addition (gpl)	Sizing (mesh)		Zn (gpl)	Cn	Cd	Co	Sb	
1	142	74	213	100	70	0.66	4.2	0.10	+100	-	141	55	180	101	-	
2	"	74	213	100	70	1.33	4.7	0.10	+100	-	142	37	191	101	-	Cd
3	"	37	191	101	60	0.66	5.2	0.40	-100	-	141	33	168	101	-	
4	"	"	"	"	60	1.33	5.2	0.40	-100	-	-	24	164	98	-	
5	"	"	"	"	65	0.66	5.1	0.40	-100	-	-	28	178	95	-	
6	"	"	"	"	65	1.33	5.1	0.40	-100	-	-	17	182	95	-	Cd
7	"	"	"	"	70	0.66	5.2	0.40	-100	-	-	29	178	98	-	
8	"	"	"	"	70	1.33	5.2	0.40	-100	-	-	17	183	92	-	Cd
9	"	"	"	"	60	0.66	5.2	0.40	+100	-	-	33	187	98	-	
10	"	"	"	"	60	1.33	5.2	0.40	+100	-	-	26	179	98	-	
11	"	"	"	"	70	0.66	5.2	0.40	+100	-	-	21	160	92	-	
12	"	"	"	"	70	1.33	5.2	0.40	+100	-	-	10	173	92	-	Cd
13	"	"	"	"	65	0.66	4.2	0.85	-100	-	-	0.15	1.56	44	-	
14	"	"	"	"	65	1.33	4.2	0.85	-100	-	-	<0.05	1.10	88	-	Co
15	143	<0.05	1.10	88	95	5.0	4.7	4.00	-100	1.2	-	"	<0.10	64	<0.05	
16	"	"	"	"	"	"	"	4.00	-100	2.4	-	"	"	41	"	
17	"	"	"	"	"	"	"	4.00	-100	5.0	-	"	"	35	"	
18	"	"	"	"	"	"	"	8.00	-100	2.4	-	"	"	29	"	
19	"	"	"	"	"	"	"	8.00	-100	5.0	-	"	"	12	"	
20	"	"	"	"	"	"	"	8.00	-100	10.0	145	"	"	0.08	"	

cell went into a small hot spent tank where excess spent overflowed to a storage reservoir for analysis and use in leaching and neutralization tests. The majority of the spent flowed from the hot spent tank into the cold spent tank for mixing with neutral feed. The cell was constructed of polyethylene. The anodes, cathode blanks, and edge strips were obtained from Canadian Electrolytic Zinc. These materials had been in use for at least six months at CEZ before shipment. This ensured low lead contamination of deposits. The cell consisted of four anodes and three cathodes in series-parallel for a total cathode deposition area of 0.88 square feet. Control of neutral feed addition rate was critical and was achieved with a Williams Oscillomatic Pump. Neutral feed solutions prepared from B2C-24 and B2C-27 calcines yielded zinc at very high current efficiency. Surface roughness of the deposits was attributed to the extreme purity of the solution (i.e. absence of inerts such as Mn, Mg, Na, ammonia, and organics) and the resulting poor throwing power. The level of these inerts in a system would gradually increase with time until equilibrium levels are reached. Conditions and results for these preliminary experiments are given in Table A1-23.

.5 Spent Electrolyte Neutralization with Limestone

The electrowinning of zinc from a sulphate system is essentially a replacement reaction of hydrogen ion for zinc in the system. In normal zinc plant practice this spent electrolyte product is consumed in the leaching of zinc oxide calcine. However, when a zinc sulphate calcine

TABLE A1-23

CONDITIONS AND RESULTS OF PRELIMINARY ELECTROWINNING TESTS IN SEVEN LITRE CELL

Test #	Neutral Feed Composition(gpl)			Spent in Cell at Time = 0 Composition (gpl)			Current Density (Asf)	Feed Rates (lt/hr)		Electrolyte Mean Temp. (°C)	Stripping Cycle(hr)	Additives(mgpl)	
	Zn^{2+}	Acid	Mn^{2+}	Zn^{2+}	Acid	Mn		N.F.	Spent Overflow			Glue	Na_2SiO_3
1	150	<5	1.5	50	180	1.5	50.8	0.48	0.45	33	20.25	18	40
2	146	<5	0.6	50	175	1.5	50.8	0.48	0.45	36	21.00	18	40
3	146	<5	0.6	46	182	-	50.8	0.50	0.48	38	24.00	18	40

Note: Neutral Feed #1 is synthetic, N.B. #2 and #3 were B2C-24, B2C-27 dilute acid leach solutions purified as in purification test #20. Co <0.1 mgpl, Cd <1.0 mgpl, Cu <0.1 mgpl, Sb <0.05 mgpl, Fe <10 mgpl.

TABLE A1-23 (Cont'd)

CONDITIONS AND RESULTS OF PRELIMINARY ELECTROWINNING TESTS IN SEVEN LITRE CELL

Test #	Spent Composition (gpl) at Stripping Zn^{2+}	Acid	Mn	Deposit Weight (g)	Current Efficiency (%)	Cell Voltage (V)	Deposit Appearance
1	50	175	0.9	1012	91.0	3.6	few nodules, some pinholes, surface very rough, no wet-backing
2	46	182	-	1072	93.0	3.4	no nodules, no pinholes, surface very rough but deposit dense, with no wet-backing (typical of zinc electrowon from very low Mg, Na, NH_4^+ solution)
3	-	-	-	-	92.3	-	similar to #2 but a little less rough

is leached, only a small percentage of the sulphuric acid in the spent electrolyte is necessary for the leaching process, accordingly there is an excess of sulphate in the system equivalent to the sulphated zinc in the calcine. The spent electrolyte contains about 25 per cent of the total zinc leached. Therefore, the primary requirement of any process designed to remove excess sulphate must be low zinc losses.

The common approach to sulphate removal in chemical industries is the production of gypsum. It appeared that our system was ideally suited for the use of low cost limestone as the neutralizing, precipitating agent. Unlike most conventional zinc plants, no water balance problems would exist in a sulphation roast operation, in fact, fairly large amounts of fresh water would have to be added to the leach even if all the water in the spent was recycled. Most processes are not amenable to limestone neutralization, due to volume balance problems associated with the fresh water which must be introduced to the system for effective washing of the gypsum residue.

The feasibility of limestone neutralization of spent electrolyte, gypsum precipitation, filtration, and washing was investigated on a laboratory scale. Studies were performed originally as batch tests in beakers, and acceptable zinc losses (gypsum residue <1-2% zinc) were obtained. These conditions were followed during the interlocked roaster-leach trials. Zinc losses were much higher (>3% of residue). This was found to be due to lack of control in the large scale batch conditions, primarily retention time and armouring (gypsum forming around a clump of limestone, preventing reaction of the interior). Also, the Perrin filter press used had no capability for displacement washing.

Accordingly, an apparatus was designed and constructed based on fertilizer industry gypsum precipitation techniques, to undertake experiments on a continuous basis. Also, filtration was performed with a vacuum system, enabling efficient displacement washing.

This system was capable of producing residues of <0.6% zinc, a filtration rate of >3 lt/min/ft^2, a moisture content of ∿42% and a gypsum crystal size of up to 200 microns, using the optimum conditions.

.51 Experimental Program

.511 Preliminary Beaker Tests

Ten litre batches of synthetic spent electrolyte (60 gpl Zn, 1.5 gpl Mn, 169 gpl H_2SO_4) were neutralized to the desired pH by the continuous addition of limestone (Havelock: 95% $CaCO_3$, 0.25% Mg max.) The resultant slurry was filtered in a Denver laboratory batch pressure filter using a polypropylene filter cloth (Porritts & Spencer Filtration Ltd., #2 Propex, 2-4 cfm) and a filtration pressure of 40 psig. Slurry addition to the filter was controlled to produce a 5.08 cm (2 in.) thick filter cake.

Washing tests were conducted both by repulping and by displacement washing in the filter.

Results of the beaker tests are presented in Table A1-24. Washed cakes provided zinc levels of <0.80% of the residue for slurries neutralized to pH 4.0, and <1.70% for slurries neutralized to pH 1.8. Doubling the wash water volume from 50 to 100% of the neutralized spent volume decreased the residual zinc levels by about 40%. In situ washing produced lower residual zinc levels than repulping in all tests. The filter cakes produced were 49-50% solids.

.512 Interlocked Runs-Batch Tests

Limestone was added slowly to 100 litre (22 gal.) batches of spent electrolyte until the desired pH was attained. The slurry was filtered in the Perrin filter press. Due to armouring of the limestone lumps, and the inefficient washing of the filter, residual zinc levels were 6.6 and 4.6% of the residue weight. The cakes were 50 and 45% solids respectively. Conditions and results are given in Table A1-42.

.513 Interlocked Runs-Continuous Testing

Because of the failure of the batch tests above to produce an easily washed gypsum residue, it was decided to attempt continuous neutralization. For this purpose, a rectangular container 48 x 19 x 22 cm high (18.9" x 7.4" x 8.7" high) was constructed from polyethylene. Two internal partitions were installed to make three mixing compartments. Each compartment was equipped with a single, two bladed,

TABLE A1-24

RESULTS OF SPENT NEUTRALIZATION RUNS FROM BEAKER TRIALS

Neutralization to pH	% Moisture	% Zn (dry)	% Zn (dry) After Washing[a]
4.0	54	5.82 (cake a1)	3.90 (cake a2)
1.8	57	4.85 (cake b1)	3.70 (cake b2)

RESULTS OF SPENT NEUTRALIZATION RUNS FROM BEAKER TRIALS-WASHING TESTS

Cake #	Wash Water (lt)	% Moisture	% Zn[1]	% Zn[2]
a1	2.5[a]	35	0.80	0.75
a2	5.0[b]	34	0.48	0.46
b1	2.5[a]	51	1.70	0.90
b2	5.0[b]	38	1.00	0.70

a - 50% of neutralized spent volume
b - 100% of neutralized spent volume

1 - washing: repulping for 5 minutes
2 - washing: in situ in pressure filter

low shear stirrer, driven by a variable speed motor. The first cell was also equipped with a water cooled, metal container for temperature control.

Spent electrolyte (50 gpl Zn, 3.5 gpl Mn, 180-190 gpl H_2SO_4) was fed to the first cell using a small metering pump. Limestone (Havelock, 100% -100 mesh, 95% $CaCO_3$ 0.25% Mg max.) was allowed to drop on the surface of the first cell from a variable speed screw feeder.

Initially, it was intended that slurry would overflow the first partition to the second cell, then flow under the second partition into the third cell and finally overflow to exit the system. However, it was found that incoming limestone was carried out of the first cell on the foam produced by the limestone-acid reaction, causing considerable shortcircuiting. The first partition was modified to have a slot at mid-depth. This improved the performance of the cell, although shortcircuiting was not eliminated.

Various acid:limestone ratios and retention times were tested. Samples of the third cell overflow were test filtered in a 195 mm Buchner funnel using Watman #1 filter paper and 500 mm Hg vacuum. Unfortunately, the filter paper in the funnel usually lifted near the end of the filtration, greatly affecting the results. The test results are given in Table A1-25.

The test conditions that produced the best results were repeated with the addition of gypsum seed crystals (Belledune Fertilizer) in the spent electrolyte. Feed additions

TABLE A1-25

CONDITIONS AND RESULTS OF CONTINUOUS SPENT NEUTRALIZATION RUNS

Run #	$CaCO_3:H_2SO_4$ (mole:mole)	Ret. Time (hr)	pH final	pH filter	Slurry & Wash Water (ml)	Cake Depth (cm)	Filtrate (ml)	Time (min) to dry	Time (min) to crack	Zn in Cake	Gypsum (size-μ	% Rhomb.)	Filter Rate (lt/min/ft^2)	% Solid in Cake	Residual $CaCO_3$ (wt%)		Comments
F8A	1:1	1.47	1.15	5.25	1500 + 500	4.0	1150	5.2	0.7	0.56	30-40	10.0	1.36	34.2	<1	a	a-when gassing pH <1.2
F8B	1:1	1.47	1.15	5.00	750 + 250	1.7	750	2.0	0.2	1.15	<30	20.0	2.38	~33	<1	a	pH up to 5, then fell to 1
F8C	1:1	1.47	1.15	5.00	175 + 60	1.9	160	1.4	0.3	----	--	----	2.77	~33	--	a	
F9A	.95:1	1.50	.95	1.50	750 + 250	0.8	800	1.4	0.3	0.24	5x40	----	3.29	65.8	<1	b	b-some gassing, pH always <1
F9B	.95:1	1.50	.95	1.50	1500 + 500	3.8	1050	3.5	0.3	0.23	<50	15.0	1.86	30.5	1	b	
F9C	.95:1	1.50	.95	1.50	1000 + 325	1.8	900	2.4	0.3	----	--	----	2.33	33.8	--	c	c-pH rose to 2.9, then feeder rate erratic; .. pH fell to 1.5
F9D	.95:1	2.10	1.55	2.20	1000 + 325	2.0	700	3.7	0.2	----	--	----	1.26	32.7	--	c	
F9E	.95:1	2.10	1.55	2.20	750 + 250	1.5	850	2.3	--	0.68	<50	10.0	2.58	54.3	<1	d	d-a bit of gassing in last cell
F9F	.95:1	3.50	2.40	4.75	750 + 250	1.7	550	2.7	0.3	1.70	>50	----	1.28	34.1	<1	e	e-some gassing at first; pH fell to 1 from 5, than rose to 5 again
F10A	.975:1	1.50	4.95	5.20	1500 + 500	2.0	750	1.6	1.0	2.52	>50	35.0	3.28	35.8	5	e	
F10B	.975:1	1.50	4.95	5.20	1500 + 500	3.0	1700	2.5	--	0.26	>50	40.0	4.76	57.3	<1	e	
F13A	.90:1	1.50	1.15	4.85	750 + 250	4.0	1250	5.3	1.0	1.06	>50	30.0	1.65	40.7	1	f	f-pH fell to 1.15 - lots of gassing
F13B	.90:1	1.50	1.15	4.85	1500 + 250	1.5	800	2.3	--	0.83	>60	35.0	2.43	64.2	<1	f	
F13C	.90:1	2.90	.85	.85	1500 + 500	3.5	1300	3.5	--	2.64	<30	10.0	2.60	45.7	<1	g	g-pH never >1
F13D	.90:1	2.90	.85	.85	750 + 250	1.7	550	1.7	0.2	1.47	<30	5.0	2.02	33.3	<1	g	

were controlled to 2.5% of the final gypsum mass. Results of these tests are presented in Table A1-26.

The best conditions occurred with limestone:acid = .975:1, and retention time of 1.5 hr. The two samples (F10A and F10B) under these conditions produced:

i) high filtration rates (3.28 and 4.76 lt/min/ft^2)
ii) low zinc in cake in F10B (0.26%)
iii) maximum pH of 5.0
iv) one dry cake for F10B (57.3% solids)
v) 35-40% of cake as large rhombic gypsum crystals

Other sets of conditions produced some of these desirable characteristics, but these conditions alone produced all five.

The seeding produced cakes with a much higher content of large gypsum rhombs. This is illustrated in Plate 4. The photo in the lower right, F17H (with seeding), contains a greater ratio of large rhombic crystals than that in the lower left, F10B (without seeding) for the same conditions. Also, a photo of a residue which was from a run with a low final pH (upper right), and a photo of Belledune Fertilizer gypsum (upper left), are shown for comparison.

The larger grain size for the seeding runs produced cakes of much lower moisture content (for cakes where the filter paper lifted, the seeded cakes averaged 40-45% solids as compared to 30-36% for the unseeded). Presumably, this should have produced much better zinc washing and higher filtration rates. However, this was true in only a few

TABLE A1-26

CONDITIONS AND RESULTS OF CONTINUOUS SPENT NEUTRALIZATION RUNS WITH GYPSUM SEEDING

Run #	Filter pH	Cake Depth (cm)	Filtrate		Filtrate & Water		Cracking (min)	Zn in Cake (wt %)	Gypsum		Filtrate Total (lt/min/ft^2)		% Solid in cake	Residual $CaCO_3$ (wt %)	Comments
			(ml)	(min)	(ml)	(min)			(size-μ)	(% rhombs)					
F17A	1.40	3.4	700	2.2	550	1.4	0.6	0.66	<100	70	2.23	2.08	45.9	<1	gassing in 3rd cell lasted 2 hours
F17B	4.95	3.5	550	1.3	450	1.5	0.4	0.82	100-200	70	2.96	2.19	41.6	1	
F17C	5.35	4.0	750	1.6	550	1.8	0.4	2.55	100	50	3.28	2.39	42.4	1	
F17D	5.25	4.0	500	1.8	500	2.6	0.4	1.62	<100	35	1.94	1.46	-	1	
F17E	5.05	4.0	350	1.3	550	1.8	0.3	1.40	<200	50	1.88	1.85	-	5	
F17F	4.85	4.0	700	1.2	500	1.4	0.0	1.82	200	45	4.08	3.23	40.8	5	
F17G	5.10	3.3	600	2.4	700	1.8	1.1	1.88	100-200	50	1.63	2.04	49.6	1	
F17H	5.45	3.6	650	1.6	750	2.6	3.8*	0.64	200	70	2.84	2.33	58.1	5	* never cracked; pressure down to 400 mm Hg after 3.8 min.

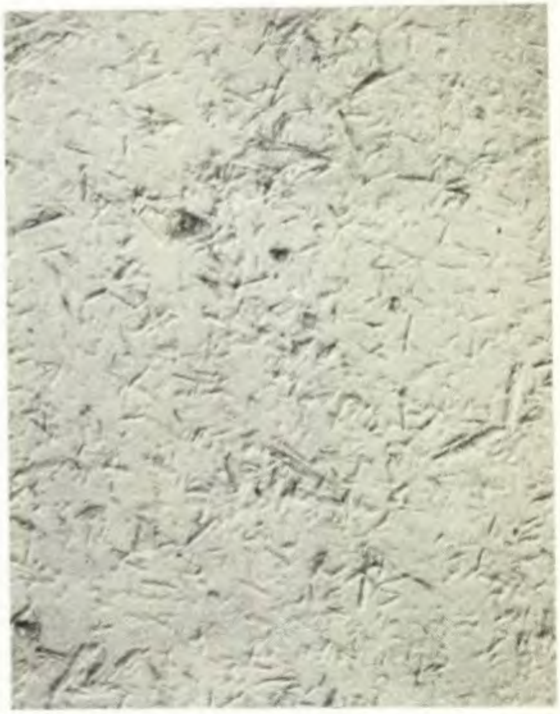
CONDITIONS: $CaCO_3:H_2SO_4$ = .975:1 (mole:mole)

Retention time = 1.5 hr

Seeds : 30-40μ rhombs
7.5 gpl

PLATE 4
NEUTRALIZATION: GYPSUM CRYSTALLIZATION

Agricultural Gypsum, 1 cm = 100μ



Gypsum Ppt'n: Uncontrolled, 1 cm = 100μ

Gypsum Ppt'n: Controlled, 1 cm = 100μ

Gypsum Ppt'n: Controlled
2% Seeding, 1 cm = 100μ

cases, due to the problem with the lifting of the filter paper. An example of this is seen for runs F10A and F10B. The feeds to the filter were identical. However the filter paper for F10A lifted immediately, while that for F10B never lifted, therefore, the % zinc and moisture content in each cake was quite different.

As in the interlocked runs, armouring was a problem, although to a much lesser extent. Although in the majority of tests residual $CaCO_3$ was less than 1%, in some of the seeded runs the filter cakes were up to 5% residual $CaCO_3$. A possible solution to this would be to pneumatically inject the limestone underneath the slurry surface. The resulting fine particles would then have a chance of being thoroughly wetted. It is expected that residual $CaCO_3$ would be significantly lower in a full pilot or commercial scale plant due to decreased shortcircuiting.

.6 Interlocked Roast-Leach Trials

The purposes of these trials were as follows:

1. To test the conditions for the individual process steps determined in the preliminary testing program on a semi-continuous basis on a larger scale.
2. To determine the effects of minor fluctuations experienced in continuous operation on the efficiency of individual process steps.
3. To determine the interractions between individual process stages and to examine the feasibility of the overall process.

4. To trace minor chemical components and adjust parameters for their control.

5. To obtain information necessary for detailed mass balances and economic assessment for extrapolation to commercial scale plant.

6. To obtain sufficient information for full pilot design and start-up.

.61 Experimental Program

The experimental program proceeded according to the following outline:

1. At start-up a synthetic hot acid leach solution was produced, the composition of which was based on projected hot acid leach equilibrium conditions. This solution was employed as coolant spray in the production of calcine from a BM&S zinc tailings refloat product.

2. Calcine prepared with synthetic hot acid leach recycle was leached under dilute acid ("neutral") leach conditions estimated to produce a solution containing 150 gpl zinc, filtered, and the residue was hot acid leached and filtered. The volume of hot acid leach liquor was controlled to insure thermal balance in the roaster. The hot acid leach liquor was then used as coolant for another calcine. Synthetic spent electrolyte was used as the acidifying agent in the neutral leach.

3. The neutral leach filtrate was then purified of copper by solvent extraction.

4. While the solvent extraction was proceeding, another calcine was leached as in Step 2.

5. Leach solution #1 was advanced from solvent extraction to iron oxidation and neutralization.

6. Leach solution #2 was advanced to solvent extraction.

7. Another calcine #3 was leached and all solutions advanced.

8. The iron oxidation sludge from leach solution #1 was recycled to the neutral leach of the 4th calcine.

9. The above process was continued as leach solutions advanced from the leach → solvent extraction → iron oxidation and neutralization → 1st stage purification → 2nd stage purification → electrowinning stepwise as new calcines were leached.

The sludge from each iron oxidation and neutralization step was recycled to the next neutral leach.

10. As soon as spent from the electrowinning cells was available, it was used to replace synthetic spent in leaching and the remainder neutralized with limestone and sent to the neutral leach for concentrating with zinc.

Several unit operations in the pilot plant are shown in Plate 2. A complete schematic of the pilot plant has been shown in Figure A1-2.

.611 "Neutral" Leaching

The neutral leach was conducted in a large 316 SS tank with vigorous stirring but no vortexing permitted. The background solutions of combinations of water, neutralized spent electrolyte and iron oxidation sludge were mixed and to this the calcines were added as a batch. The pH was controlled at pH 3.8 to 4.0 by continuous addition of spent electrolyte for 1.5 hours. Small samples of the 100 to 140 litre batches were taken for flocculating, settling, filtering and extraction tests. The fine pulp was filtered in a Perrin plate and frame filter press. The residue was washed in the press by pumping 10 litres of wash water from the previous leach filtration through the press. This solution was combined with the filtrate. Ten litres of fresh water were then pumped through the press and the wash solution saved for the first washing of the next neutral leach residue. Washing efficiency varied between 50 and 85 per cent due to the fact that the Perrin press was not equipped for displacement back washing or compression. Considerable channelling was evident in the washed cakes. Neutral leach test conditions and results are given in Tables A1-27, A1-28, and A1-29.

Extraction of zinc from sulphate and oxysulphate in the calcine was virtually 100 per cent efficient as shown in Figure A1-43. Copper extractions varied considerably from run to run as indicated by Figure A1-44. The fluctuations in copper level in the neutral leach filtrate correlate well with variations in zinc oxysulphate level in the calcines (Figure A1-44 and Table A1-27). If the calcine leached contained a larger amount of zinc oxysulphate than the previous run, then the copper level in solution fell and

TABLE A1-27

COMPOSITION AND PRIMARY EXTRACTIONS OF CALCINES LEACHED IN INTERLOCKED TRIALS

Neutral Leach #	Calcines Leached (BMD)	Overall Composition (wt%)			Distribution of Zinc in Calcine (wt%)				Primary Extractions (%)			
		Zn	Cu	Pb	$ZnSO_4$	$ZnO.2ZnSO_4$	$ZnO.Fe_2O_3$	(Zn,Fe)S	water 1 hr, 25°C		sulphuric acid 1 hr, 25°C, 25gpl	
									Zn	Cu	Zn	Cu
1	41,42,43	25.2	0.61	2.81	70.61	15.09	11.38	2.96	76.9	–	85.7	–
2	37,38,39	27.9	0.56	2.63	25.83	52.60	18.92	2.70	48.0	–	78.4	–
3	46	27.7	0.57	2.08	67.04	20.87	10.17	1.96	75.8	–	87.9	–
4	47,48,49	29.5	0.58	2.28	36.33	47.39	13.98	1.56	56.5	–	83.7	–
5	43,47	27.1	0.59	2.22	61.96	24.81	11.70	1.57	72.4	–	88.0	–
6	29,30,31	28.8	0.44	2.53	34.72	56.07	15.52	2.80	49.3	–	82.0	–
7	54	26.9	0.58	2.10	76.03	9.67	10.77	3.57	80.1	38.5	85.7	87.6
8	56	27.4	0.58	2.25	69.35	14.23	14.68	1.77	75.3	26.5	83.6	82.7
9	58	27.8	0.52	2.38	66.83	19.41	11.80	1.91	75.0	15.4	86.2	84.5
10	59,60	28.6	0.51	2.38	59.97	24.19	13.63	2.23	70.2	5.1	84.2	83.0
11	61	27.3	0.56	2.30	63.96	19.80	13.85	2.43	72.3	17.5	83.7	84.2
12	63,64	29.6	0.46	2.55	48.73	34.72	13.21	3.38	63.4	–	83.4	80.6
13	50,51,52,53	28.8	0.61	2.30	47.14	37.34	13.25	2.22	63.4	–	85.1	–
14	62,CEZ	39.4	0.54	1.38	–	–	–	–	–	–	–	–
15[a]	62	29.25	0.50	1.95	67.51	18.13	11.08	3.30	75.1	18.1	85.6	83.5

a – Beaker test – No hot acid leach

TABLE A1-28

CONDITIONS OF NEUTRAL LEACH RUNS FROM INTERLOCKED TRIALS

Neutral Leach #	Calcine Wt (kg)	Background Solution					Spent Electrolyte Addition		
		Water (lt)	Neutralized Spent (lt)	Iron Oxidation Underflow (lt)	NL Res. Wash Water (lt)	HAL Res. Wash Water (lt)	Composition(gpl) Zn	Composition(gpl) Acid	Vol. (lt)
1	64	103	0	0	10	0	50	170[a]	9.5
2	64	103	0	0	10(1)	0	50	170[a]	21.0
3	64	103	0	0	10(2)	0	50	170[a]	14.0
4	64	103	0	20(1)	10(3)	0	50	170[a]	18.0
5	64	103	0	11(2)	10(4)	0	50	170[a]	14.0
6	64	103	0	20(3)	10(5)	0	50	170[a]	21.0
7	64	103	0	10(4)	10(6)	0	50	170[a]	8.5
8	50	103	0	14(5)	10(7)	0	50	180[a]	8.5
9	50	103	0	17(6)	10(8)	0	50	180[a]	6.0
10	50	53	50(1)	12(7)	10(9)	0	42(2)	180[b]	13.0
11	50	53	50(2)	14(8)	10(10)	0	62(4)	180[b]	12.0
12	46	28	75[a]	13(9)	0	21(5)[c]	0	180[a]	12.0
13	50	28	75[a]	15(10)	0	0	50	180[a]	24.8
14	31[d]	18	35	0	0	0	46	177	35.0
15	-	-	-	-	-	-	-	-	-

Note - Retention time of all leaches was 1.5 hr., Temperature was ambient 40-47°C.

() - indicates solution number.

a - synthetic solution, synthetic neutralized spent = 50 gpl Zn.

b - acid added to increase acidity.

c - Zn, Fe and Cu added to HAL WW 11 to simulate efficient washing.

d - 21.6 kg BMD 62, 10.6 kg CEZ calcine

TABLE A1-29

RESULTS OF NEUTRAL LEACH RUNS FROM INTERLOCKED TRIALS

Neutral Leach #	Neutral Leach Filtrate Composition								Neutral Leach Residue Composition					
	Filtrate #	Vol. (lt)	gpl			mgpl			Residue #	Wet Wt (kg)	H_2O (%)	Dry Wt (kg)	Zn (%)	Cu (%)
			Zn	Cu	Fe^{2+}	Cd	Co	As						
1	1	110	130	1.72	0.98	204	67	<1	1	–	–	–	–	–
2	2	125	121	1.56	0.38	228	58	<1	2	–	–	–	–	–
3	3	115	144	2.40	1.96	236	73	<1	3	–	–	–	–	–
4	4	135	147	1.95	0.99	226	63	<1	4	–	–	–	–	–
5	5	120	143	2.25	1.12	212	69	<1	5	–	–	–	–	–
6	6	140	130	1.23	0.47	206	63	<1	6	–	–	–	–	–
7	7	115	132	2.00	1.09	238	70	<1	7	–	–	–	–	–
8	8	–	119	2.40	1.28	244	72	<1	8	–	–	–	–	–
9	9	125	113	1.34	0.22	238	62	<1	9	36.8	32.0	25.0	–	–
10	10	130	122	1.13	0.48	210	54	<1	10	39.5	32.5	26.7	–	–
11	11	130	125	1.54	0.77	190	48	<1	11	38.0	31.3	26.1	–	–
12	12	139	140	1.28	1.70	–	–	<1	12	–	–	–	–	–
13	13	127	129	1.62	0.74	185	130	<1	13	37.8	28.4	27.1	–	–
14	14	95	132	0.88	<0.01	190	–	<1	14a	21.9	22.3	15.7	13.8	–
15	15	–	–	–	–	–	–	–	15	–	–	–	–	–

a – residue not washed

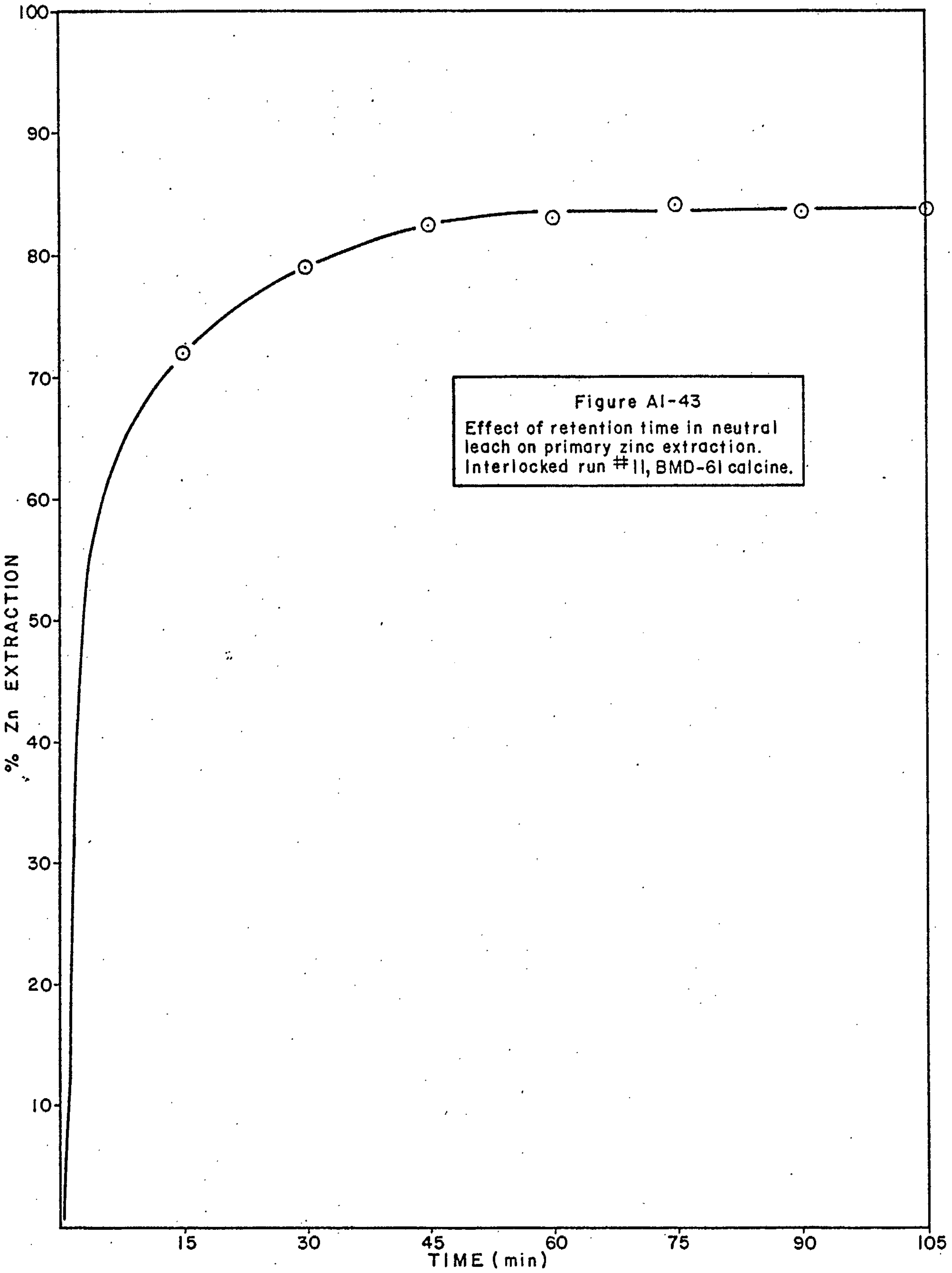


Figure A1-43

Effect of retention time in neutral leach on primary zinc extraction. Interlocked run #11, BMD-61 calcine.

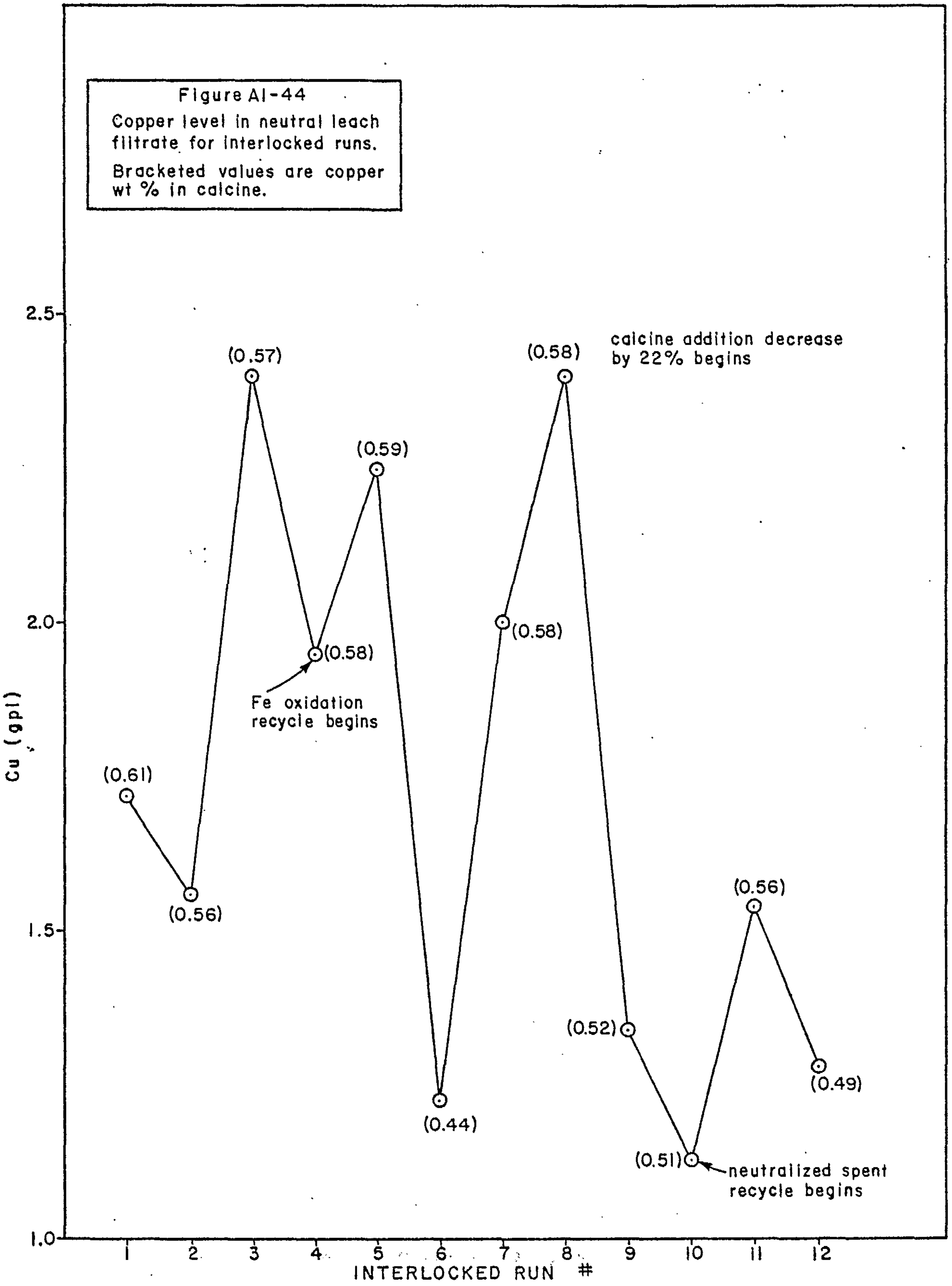


Figure A1-44

Copper level in neutral leach filtrate for interlocked runs.

Bracketed values are copper wt % in calcine.

correspondingly when the zinc oxysulphate content was decreased, the copper level increased. Generally copper extractions in the neutral leach were very much greater than what would have been expected from the preliminary tests at the high oxysulphate levels experienced in the interlocked runs. The fact that, up to 40 per cent of the copper in all the calcines was there due to the hot acid leach recycle and this copper is easily leachable due to lack of physical contact with zinc oxysulphate; and bed fluidization was improved due to pellet formation must be the basis of this increase in extraction.

Ferrous iron levels in the neutral leach filtrate showed variations between runs nearly duplicating those observed for copper (Figure A1-45).* Whether these fluctuations are due to the oxysulphate levels in the calcine or to the same cause as the oxysulphate fluctuations is not known. In all cases, when the neutral leach was conducted at 45-70°C, and the pH controlled between 3.6 and 4.0, the ferric ion level in the neutral leach filtrate did not exceed 25 mgpl. It appears that in a neutral leach of calcines prepared at 685°C with hot acid leach recycle, the total iron level in the liquor can be controlled at less than 1 gpl without air sparging or permanganate addition to the neutral leach.

Extraction of cobalt and cadmium in the neutral leach trials exceeded 95 per cent in all cases. Arsenic tenor in the neutral leach filtrate did not exceed 1 mgpl.

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* Sharp increase in iron in Run #12 due to recycle of HAL wash water to the NL in this run only. Roaster control leaches (Table A1-3(c)) indicate that iron extraction from calcine should be < Runs 11 and 13, similar to 9 and 10.

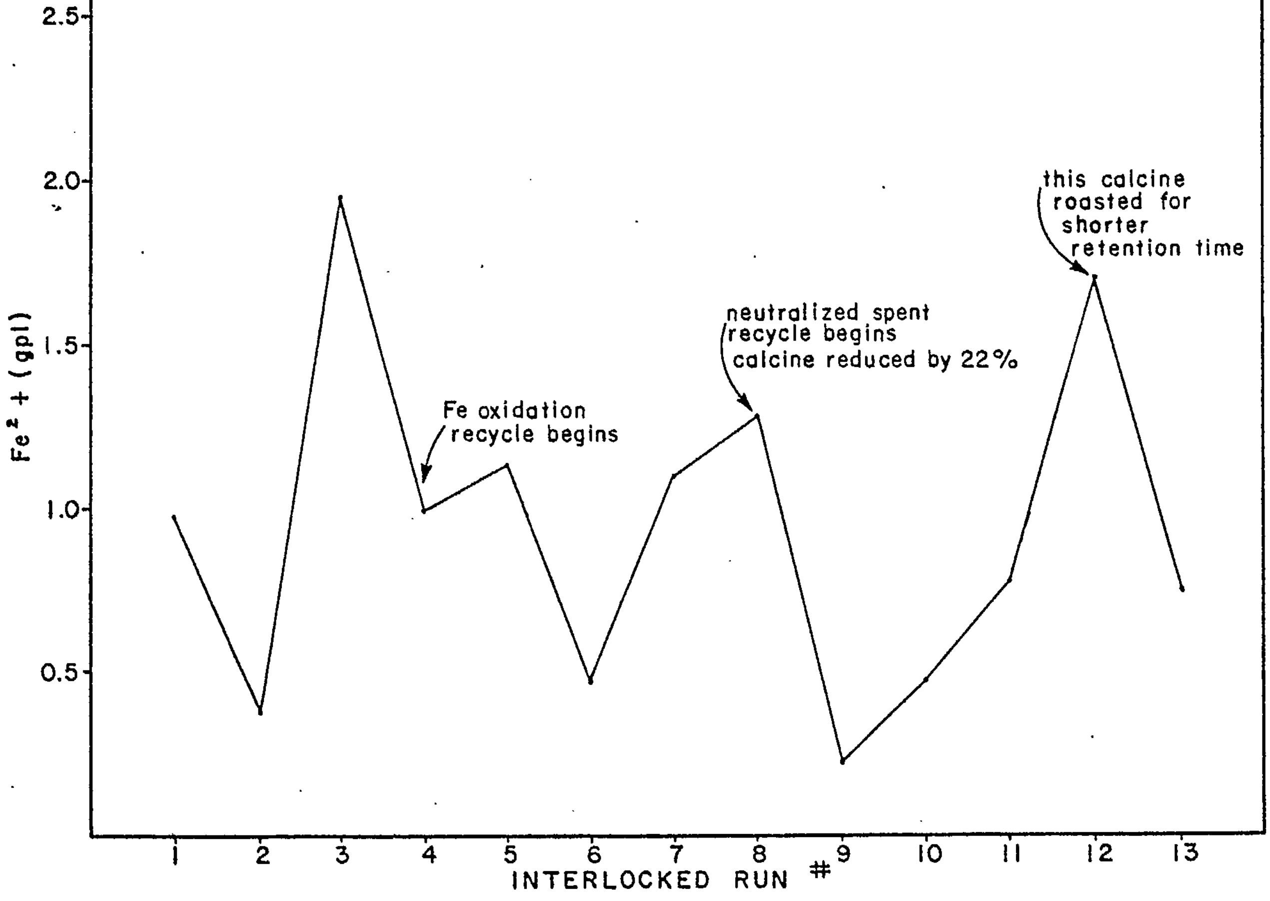


Figure AI-45
Iron tenor in the neutral leach filtrate in the interlocked runs.

Neutral leach #12 on BMD 63, 64 calcines was conducted using the hot acid leach wash water from BMD-61 calcine as part of the background solution. Although the leach proceeded as in previous cases without complication, the pulp was found to be unfilterable due to filter cloth blinding and cake sliming with ferric hydroxide. The recycle of hot acid leach wash water to the neutral leach increased the amount of iron hydroxide in the neutral leach residue by a factor of 10 (from 1 to 10% of the total residue). No further experiments were conducted on this solution and the residue was discarded.

One leach, (#14), was conducted on a combined sulphation roast (BMD-62) -dead roast (supplied by Canadian Electrolytic Zinc) calcine. The ratio of calcines used was 2.03 to 1.00 sulphation roast to dead roast. This is equivalent to a feed ratio in an integrated sulphation roast-dead roast plant operating at equilibrium of 1.43 (tailings refloat) to 1.00 (high grade zinc concentrate). Difficulty was experienced in leaching this combined calcine at constant pH under batch conditions, due to the small amount of background neutralized spent and water to be added initially and the large amount of spent electrolyte added during the leach. This problem would not be encountered in a fully continuous operation. Due to the low copper in the combined calcine, solvent extraction was not considered for copper removal, and hence potassium permanganate was added to the leach after 1.0 hours for iron oxidation ($KMnO_4$ addition 0.36

grams per litre of leach solution). Iron tenor in the neutral leach filtrate was <0.01 gpl.

.612 Neutral Leach Thickening and Filtration

Thickening tests were carried out in a 1 litre graduate cylinder which was immersed in a constant temperature bath held at 50°C, a temperature chosen as mid-way between 40°C, the temperature of a neutral leach followed by solvent extraction (SX), and 60°C, the expected temperature of a continuous neutral leach with no SX. The sample was stirred with a plunger and the height of the pulp-solution interface was recorded at various times in the usual manner.

The neutral leach pulp proved too dilute (11.3% solids) to exhibit line settling so the slurry density was increased to 15.7% where line settling was observed. This concentration of the pulp was achieved by settling a large sample of pulp, held at 50°C, decanting the supernate and mixing varying amounts of supernate and slurry.

The settling data obtained is shown in Table A1-30 and plotted in Figure A1-46 as a Roberts plot(10). This representation shows the compression point more clearly than a plot on a linear scale. In this case, the constant, H_∞, equals zero.

An analysis of the data, using the Talmage and Fitch method (10,11) gives the solids flux that could be passed through a thickener (G_t) equal to 0.500 TPD/m^2 (0.0512 short TPD/ft^2). A compression depth (P_d) of 0.0189 m (0.062 ft)

TABLE A1-30

SETTLING TEST

NEUTRAL LEACH SLURRY, NO FLOCCULANT ADDITION
BMD 50, 51, 52, 53 CALCINE BLEND

Time (min)	*Height H of Interface (cc)	Time (min)	Height H of Interface (cc)
0.0	990	147.2	630
7.0	980	152.3	610
13.5	970	158.5	600
30.9	940	165.0	570
42.7	920	175.0	540
57.1	890	181.7	520
66.2	870	191.0	500
74.7	850	197.5	480
83.4	830	206.3	460
91.4	810	212.0	450
101.9	780	220.5	440
108.5	760	228.8	430
115.3	740	239.0	420
121.7	720	244.8	410
130.4	690	255.1	400
136.4	670	277.5	380
141.4	650	288.5	370

Overnight interface height	320 cc
Solution s.g.	1.29
Solids s.g.	5.01
Weight of solids in test	227.25 g
Supernate clarity (141.4 min)	114 mgpl

* 864 cc/ft on graduate

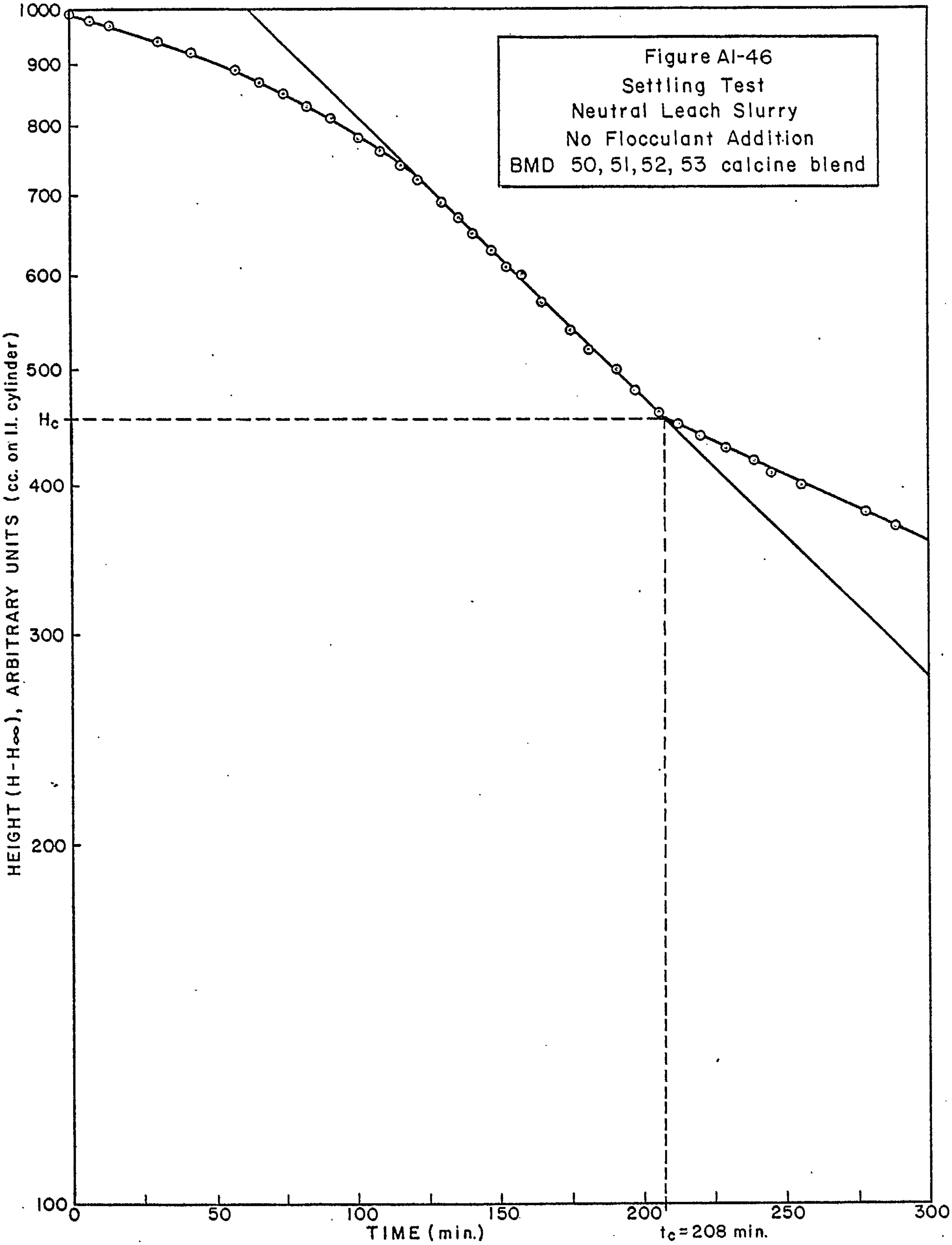


Figure A1-46
Settling Test
Neutral Leach Slurry
No Flocculant Addition
BMD 50, 51, 52, 53 calcine blend
HEIGHT (H-H∞), ARBITRARY UNITS (cc. on l.l. cylinder)
1000
900
800
700
600
500
Hc
400
300
200
100
0
50
100
150
200
250
300
TIME (min.)
tc = 208 min.

would be required to achieve 33% solids in the thickener underflow. This underflow density was chosen arbitrarily and represents a slurry height of 400 cc (overnight height was 320 cc without stirring).

No settling tests were performed on the sulphation roast/dead roast neutral leach pulp because both the solids and the liquor were identical in composition although lower in solids content.

Several synthetic flocculants, marketed under the Percol trade name by Allied Colloids (Canada) Ltd. were tested qualitatively for their effect on the neutral leach pulp since an effective flocculant can considerably reduce the thickener area required. These tests were qualitative only because of the long time required for a full series of quantitative tests and current equipment limitations.

These tests were carried out by mixing various doses of the different flocculants with 500 ml of the pulp in 1000 ml beakers. The behaviour of the solids was observed as well as the turbidity of the supernate. In most cases, fresh pulp was generated by doing 5 litre beaker leaches on BMD-61 calcine under the same conditions as the standard neutral leach. The initial tests were performed on pulp samples saved from a full scale neutral leach.

Of the products tested, the anionic Percol 156, a flocculant used in the zinc industry in flocculating ferric hydroxide, had no effect on this pulp. However, Percol 155, another anionic flocculant as well as nonionic Percol 351 and cationic 352 all flocculated the bulk of the solids immediately. Further tests indicated that Percol 351 gives the best results, in terms of suspended solids. Dosage rates as low as 16 g/tonne of solids (0.04 lb/short ton) produced a supernate with about 200 mgpl suspended solids after four hours of settling at 50°C.

The tests showed that dispersion of the flocculant will be critical for any plant scale application. The flocculation and settling of the bulk of the solids regardless of doseage, leaves a cloudy supernate which clears slowly as the particles flocculate and settle. X-ray analysis shows that these slow to flocculate solids are largely hematite and zinc ferrite with no detectable colloidal silica or basic zinc sulphate. No flocculation tests were performed on the sulphation roast/dead roast pulp. The flocculation tests for the bulk concentrate sulphation roast should apply equally well to the sulphation roast/dead roast pulp since the solids and the solution are almost identical.

All filtrations for this part of the project were done on a Perrin 18" x 18" plate and frame filter press which contains five chambers, each with a filtration area of 0.321 m^2 (3.46 ft^2). The filtration medium was polypropylene cloth (Porritts & Spencer Filtration Ltd. #2 Propex, 2-4 cfm) dressed with kraft paper (DD 40 natural kraft MF). The filter was supplied by the manufacturer with a Warren Rupp model ST1-A, V.I.P. Sandpiper air driven teflon diaphragm pump. For all the products filtered, the filter and media

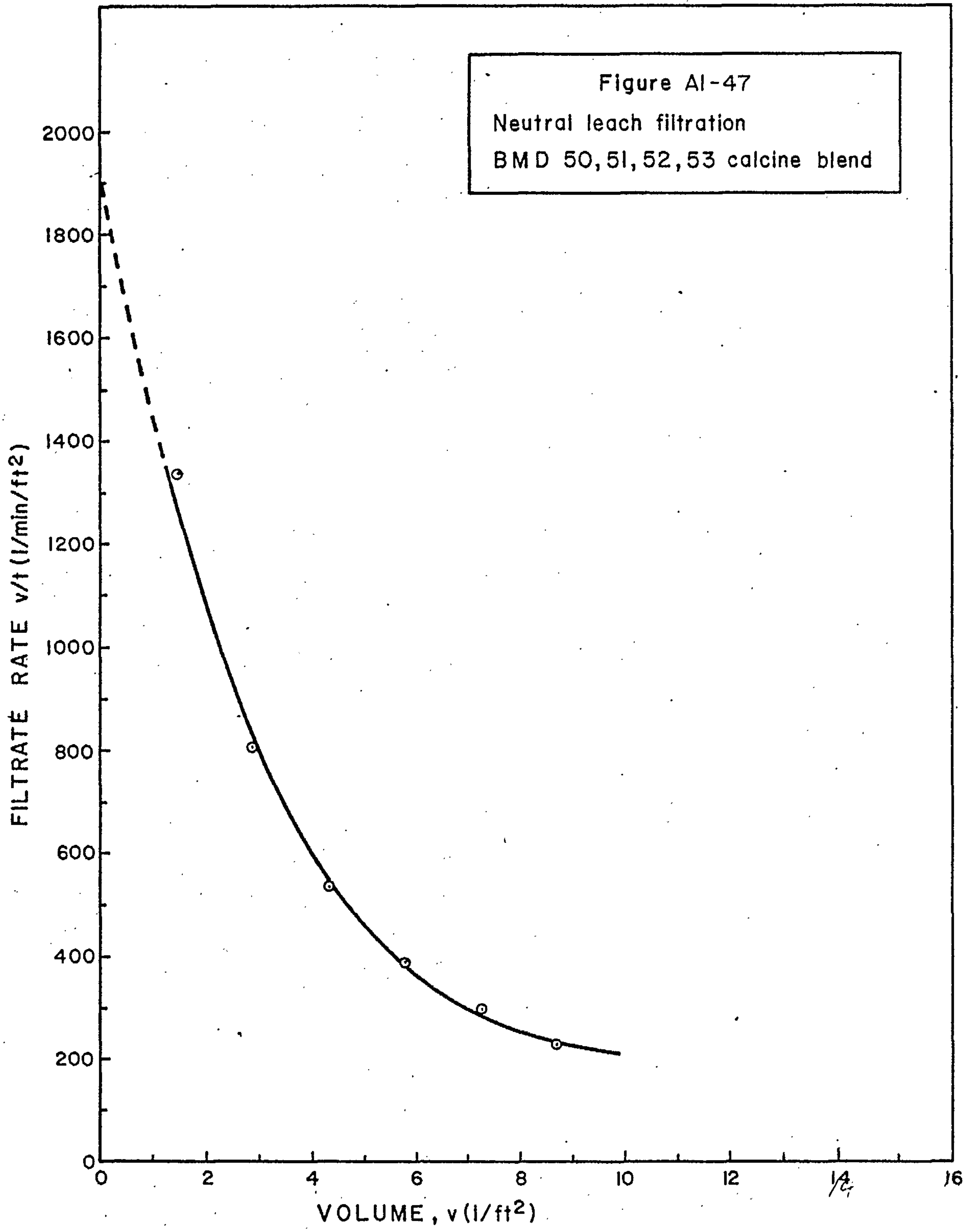


Figure AI-47
Neutral leach filtration
BMD 50,51,52,53 calcine blend
FILTRATE RATE v/t (l/min/ft2)
0
200
400
600
800
1000
1200
1400
1600
1800
2000
0
2
4
6
8
10
12
14
16
VOLUME, v(l/ft2)

TABLE A1-31

NEUTRAL LEACH FILTRATION
BMD 50, 51, 52, 53 CALCINE BLEND

Total Vol (lt)	Total Time (min)	Vol (cc)	Time (min)	Rate (ml/min/ft^2)
0	0.0	-	-	-
5	0.8	370	0.08	1337
10	2.0	445	0.16	804
15	4.0	390	0.21	537
20	7.5	425	0.32	384
25	11.5	410	0.40	296
30	17.0	370	0.47	228

Total filtrate	33 lt
Pressure	60 psig
Filter area	3.46 sq.ft.
Cake thickness	0.5 in.
Dry wt. filter cake	7.67 kg
Cake mositure	28.3%
Solids in feed	14.4%

performed satisfactorily and the same cloths were used throughout the test program.

A constant pressure filtration test, using one chamber was carried out on the pulp from the neutral leach of a blend of BMD 50, 51, 52, 53 calcines. The results are given in Table A1-31 and are plotted on Figure A1-47.

These results indicate that an average filtration rate of 6.05 lt/min/m^2 (0.15 US gal/min/ft^2) at 5.27 kg/cm^2 (60 psig) is possible for this pulp, producing a cake with 18.4% moisture content.

Another constant pressure filtration test, using the same equipment and conditions, was carried out on the BMD 62-C.E.Z. neutral leach pulp. The results of this test are presented on Table A1-32 and are plotted in Figure A1-48. Examination of these results shows that there is no significant difference in filtration characteristics between this pulp and the sulphation roast pulp above, within the limits of the test method.

Further tests should be carried out to determine if these pulps can be effectively filtered on a continuous vacuum filter, at an acceptable rate, without precipitation of basic zinc sulphate.

.613 Hot Acid Leaching

Neutral leach residues for hot acid leaching were repulped in water or spent electrolyte and heated to 95°C. Raw acid was added continuously to this pulp to maintain the pH

TABLE A1-32

NEUTRAL LEACH FILTRATION
BMD 6.2-CEZ CALCINE BLEND

Total Vol. (lt)	Total Time (min)	Rate		
		Vol (cc)	Time (min)	Rate (ml/min/ft^2)
0.0	0.0	400	0.06	1927
5.0	0.8	415	0.09	1333
10.0	2.0	365	0.10	1055
12.5	3.0	385	0.18	618
15.0	4.3	365	0.22	480
17.5	5.9	355	0.24	428
20.0	8.0	410	0.36	329
22.5	10.2	400	0.39	296
25.0	12.7	435	0.47	267
27.5	15.6	390	0.46	245
30.0	18.7	370	0.48	223
35.0	26.0	375	0.63	172
37.5	30.0	365	0.63	167
40.0	35.8	275	0.57	139

Total filtrate	42 lt
Pressure	60 psig
Filter area	3.46 sq.ft.
Cake thickness	0.5 in.
Dry wt. filter cake	6.54 kg
Cake moisture	31.5%
Solids in feed	10.3%

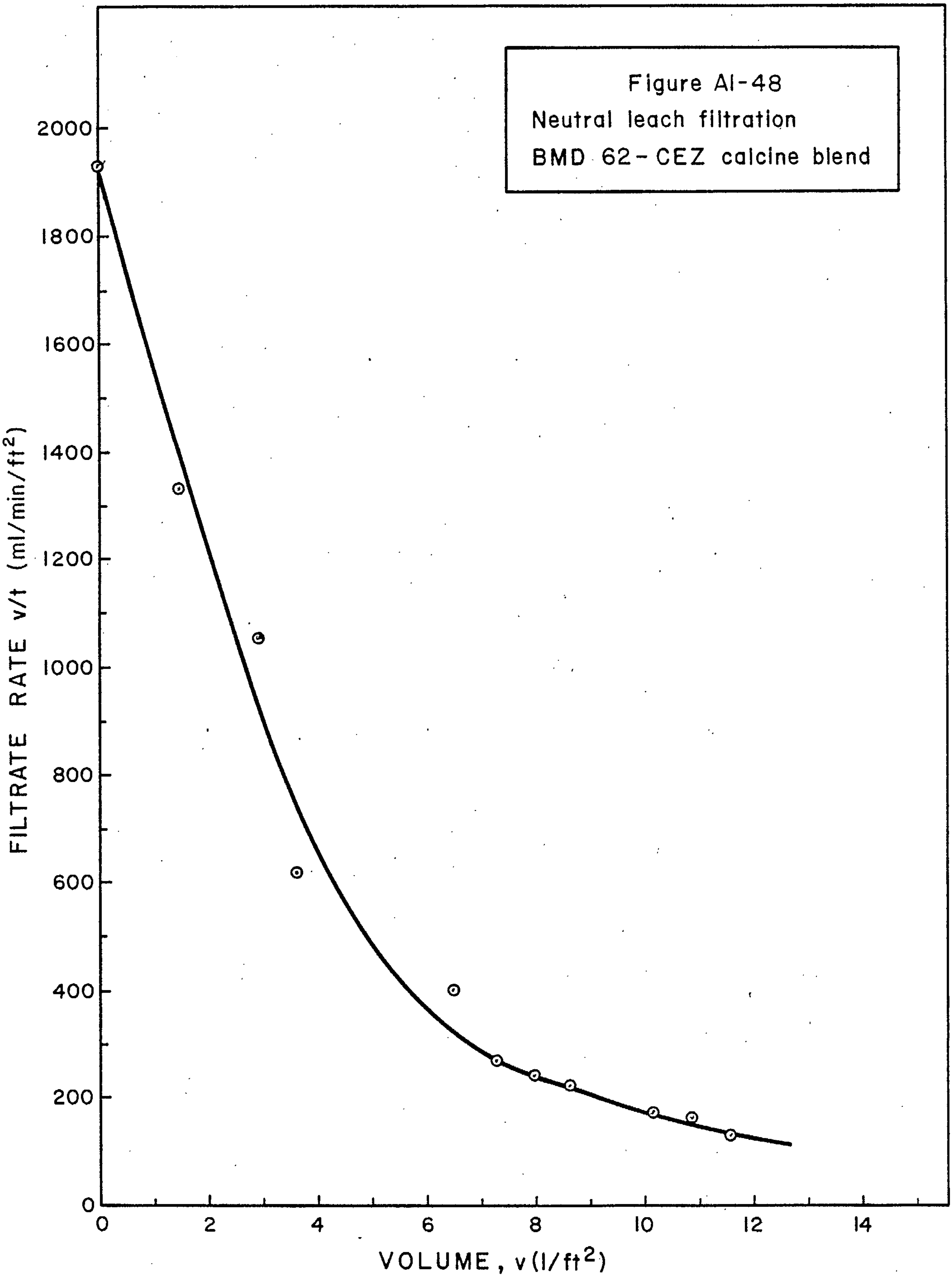


Figure AI-48
Neutral leach filtration
BMD 62-CEZ calcine blend
FILTRATE RATE v/t (ml/min/ft²)
2000
1800
1600
1400
1200
1000
800
600
400
200
0
0
2
4
6
8
10
12
14
VOLUME, v (l/ft²)

at 0.2 ± 0.1 for 1.5 to 2.0 hours. The pulp was then filtered in the Perrin filter press. The residue was washed in situ by pumping warm water through the residue. The filtrate was sent to the roaster for use as coolant spray. Hot acid leach filtrates from tailings refloat products contain 100 ± 30 gpl Zn, 100 ± 30 gpl total iron, 10 ± 5 gpl ferrous iron, 5 ± 4 gpl arsenic, 2.5 ± 1.5 gpl Cu and 40 ± 15 gpl acid. Corrosion problems were not experienced during the hot acid leach trials as the mild steel tank was lined with polyethylene and the agitator turbine was constructed of Hastelloy C. Conditions and results of the hot acid leach trials are given in Table A1-33.

Difficulties were experienced in efficiently washing the sulphates from the leach residues in the filter press. This was entirely due to the design of the press employed (see Section 131.611). Accordingly samples of each hot acid leach residue were repulped with 30% by weight hot water and filtered quickly in a buchner funnel. The filter cake was then displacement washed with a volume of hot water equal to the repulp water. The percents water soluble zinc, iron and copper in the original residue were determined from the analyses of the washed and unwashed residues. Values for masses and compositions of wash water and residue were then calculated assuming equivalent washing efficiency for a displacement wash (volume equal to the moisture content of the residue) in a filter press. The calculated composition of the wash solution includes the combination from the original inefficient, in situ wash. Results are given in

TABLE A1-33

CONDITIONS AND RESULTS OF HOT ACID LEACH RUNS FROM INTERLOCKED TRIALS

Neutral Leach Residue #	Repulp and Makeup		HAL Conditions				HAL Filtrate							Filtrate #
	H_2O (lt)	Spent (lt)	pH	Temp (°C)	Ret. Time (hr)	Raw Acid (lt)	Vol (lt)	S.G. (g/cc)	Composition(gpl)					
									Zn	Cu	As	Fe^{2+}	Fe (total)	
1	12	7[a]	0.3	95	1.5	7.5	25	1.28	94	2.8	2.0	4.65	60.8	1
2	12	7[a]	0.3	95	1.5	7.0	23	1.44	87	3.5	3.0	8.38	90.0	2
3	16	13[a]	0.3	95	1.5	7.0	23	1.44	128	4.0	5.0	15.21	92.0	3
4	–	0	0.5	95	1.5	6.5	21	1.39	87	3.5	3.0	7.32	78.0	4
5	–	0	0.2	95	1.5	6.5	-------- discarded due to line blockage --------							
6	–	0	0.2	95	1.5	7.5	21	1.48	102	4.2	4.0	18.37	93.5	5
7	–	0	0.2	87	1.5	6.0	21	1.38	76	4.1	1.0	5.19	69.0	6
8	–	0	0.2	92	1.5	7.0	21	1.44	88	3.5	2.0	9.05	90.0	7
9	–	0	0.2	95	1.5	7.5	19	1.60	100	2.6	9.0	8.71	125.0	8
10	14	0	0.2	95	1.8	8.0	19	1.53	96	1.5	9.0	7.76	95.0	9
11	17	0	0.2	95	2.0	8.5	21	1.54	107	1.2	6.0	15.30	107.0	10
12	–	–	0.1	97	2.0	–	-------- discarded --------							
13	0	23[a]	0.2	95	2.0	2.5	19	1.53	–	–	–	–	–	11
14	0	17	0.2	95	2.0	2.5	10	1.65	144	3.8	3.0	3.05	85.5	14
15	-------- residue discarded --------													

[a] – synthetic solution, synthetic neutralized spent = 50 gpl Zn

TABLE A1-33 (Cont'd)

CONDITIONS AND RESULTS OF HOT ACID LEACH RUNS FROM INTERLOCKED TRIALS

Neutral Leach Residue #	HAL Residue Wash Water								HAL Residue												
			Composition (gpl)										Composition %								
	Vol (lt)	S.G. (g/cc)	Zn	Cu	Fe^{2+}	Fe (total)	As	Wash Water #	Res. #	Wet Wt (kg)	Solids (%)	Dry Wt (kg)	Zn	Cu	Fe	Pb	Ag (ppm)	Ca	S^o	Total S	As
1	———— residue not washed ————								1	38.7	79.2	30.7	6.5	0.25	46.5	4.7	146	0.07	0.45	6.83	0.22
2	———— " " " ————								2	36.0	64.7	23.3	5.0	0.18	52.8	6.1	84	0.11	0.16	3.32	0.20
3	———— " " " ————								3	26.0	76.6	19.9	1.6	0.06	58.8	6.9	250	0.07	0.42	2.78	0.20
4	———— " " " ————								4	32.6	75.8	24.7	7.0	0.23	48.3	5.3	183	0.49	0.36	5.33	0.18
5	———— " " " ————								5	-	71.4	-	2.0	0.08	55.0	5.5	136	1.02	0.31	4.02	0.20
6	25	1.04	12.4	0.6	2.5	13.0	0.5	1	6	32.0	87.6	28.0	4.2	0.14	51.5	7.1	150	0.30	0.23	3.74	0.20
7	25	1.09	18.3	1.0	1.9	17.0	1.0	2	7	31.2	72.1	22.5	5.1	0.16	54.5	6.3	220	0.82	0.17	2.45	0.30
8	25	1.06	13.1	0.5	1.5	16.0	2.4	3	8	29.6	69.0	20.4	4.2	0.14	47.8	6.1	202	1.68	0.38	4.49	0.25
9	———— residue not washed ————								9	21.8	75.4	16.4	3.1	0.08	47.8	6.8	220	2.90	0.28	4.63	0.25
10	25	1.15	29.4	1.0	2.2	29.3	0.8	4	10	23.9	70.0	16.7	3.4	0.07	49.5	7.0	203	2.12	0.31	4.93	0.25
11	21	1.09	18.0	0.5	2.1	17.6	0.8	5	11	25.6	70.6	18.1	3.4	0.08	49.5	6.4	196	2.14	0.44	5.91	0.10
12	———— discarded ————								12	-	-	-	-	-	-	-	-	-	-	-	-
13	———— residue not washed ————								13	-	-	-	-	-	-	-	-	-	-	-	-
14	———— residue not washed ————								14	10.3	79.9	8.2	13.8	0.25	33.8	3.1	-	0.40	0.38	9.97	0.20
15	————————								15	———— residue discarded ————											

Table A1-34. Since hot acid leach liquor volumes calculated to retain thermal equilibrium in the roaster include the volume of liquor remaining in the filtered residue, the inclusion of a volume of wash water with the filtrate equal to the moisture content of the residue cake will not effect the roaster thermal balance.

Overall extractions of zinc, copper and silver were calculated by comparing the elemental content of the calculated residue with the elemental content of the calcine. For hot acid leach conditions of 2.0 hours, at 95°C, and pH 0.2, zinc and copper extractions can be controlled at 98.5 and 98 percent respectively. Virtually all the silver and lead (>99 per cent) report to the hot acid leach residue. Variations in overall copper and zinc extraction from the calcines are shown in Figure A1-49.

Hot acid leaching tests on neutral leach residue from a combined sulphation roast-dead roast calcine (run #14) resulted in an overall zinc extraction of 98.8% and copper extraction of 94.3 per cent based on zinc and copper in residue versus zinc and copper in calcine.

The leaching of zinc ferrite and sphalerite was monitored by X-ray diffraction. Figures A1-50 and A1-51 show the effect of leaching conditions on ferrite and sphalerite extraction. Since the mass of hematite does not vary from calcine to hot acid leach residue, an approximation of the relative abundance of ferrite and sphalerite in the calcine and residues may be obtained by comparing the peak heights to the hematite peak height (see Appendix A4). It can be

TABLE A1-34

CALCULATIONS OF HOT ACID LEACH RESIDUE AND WASH WATER COMPOSITIONS ASSUMING EFFICIENT DISPLACEMENT WASH

HAL Res.#	% Water Soluble Res.			Calculated Displacement Wash Water					Thoroughly Washed Residue								
	Zn	Cu	Fe	Vol. (lt)	Composition (gpl)				Dry wt (kg)	Composition (%)							
					Zn	Cu	Fe (total)	As		Zn	Cu	Fe	Pb	Ag (ppm)	Ca	As	S^o
1	4.5	0.20	4.3	8.1	172	7.6	164	-	18.1	3.4	0.08	71.6†	8.0	248	0.12	0.37	0.76
2	1.7	0.07	2.1	12.7	32	1.3	39	-	19.1	4.0	0.13	61.8	7.4	102†	0.13	0.24	0.20
3	0.9	0.02	0.8	6.1	29	0.7	26	-	18.3	0.0	0.04	63.0	7.5	272	0.08	0.22	0.46
4	3.4	0.08	2.4	7.9	106	2.5	75	-	18.1	4.9	0.20	62.6	7.2	250	0.67	0.25	0.49
5	1.5	0.06	1.6	-	-	-	-	-	-	-	-	-	-	-	-	-	-
6	1.7	0.07	1.9	3.5	225	10.0	245	4.3	23.3	3.0	0.08	59.6	8.5	180	0.36	0.24	0.28
7	0.5	0.02	0.6	8.7	66	3.4	64	2.9	21.3	4.9	0.15	56.9	6.7	232	0.87	0.32	0.18
8	1.5	0.06	1.5	9.2	69	4.8	77	6.5	17.6	3.1	0.09	53.7	7.1	234	1.95	0.30	0.44
9	0.5	0.02	0.5	5.4	17	0.4	15	-	15.6	2.7	0.06	49.7	7.1	231	3.00	0.26	0.29
10	1.0	0.03	0.9	7.2	126	4.2	123	7.0	15.2	2.6	0.04	53.4	7.7	223	2.30	0.27	0.34
11	2.0	0.06	2.0	7.5	99	2.9	98	4.2	14.7	1.8	0.02	58.5	7.8	241	2.60	0.12	0.54
12	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
13	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
14	-	-	-	-	-	-	-	-	5.3	2.7	0.18	54.4	4.8	-	0.62	0.08	0.59
15	------residue discarded------																

† uncertainty in analysis

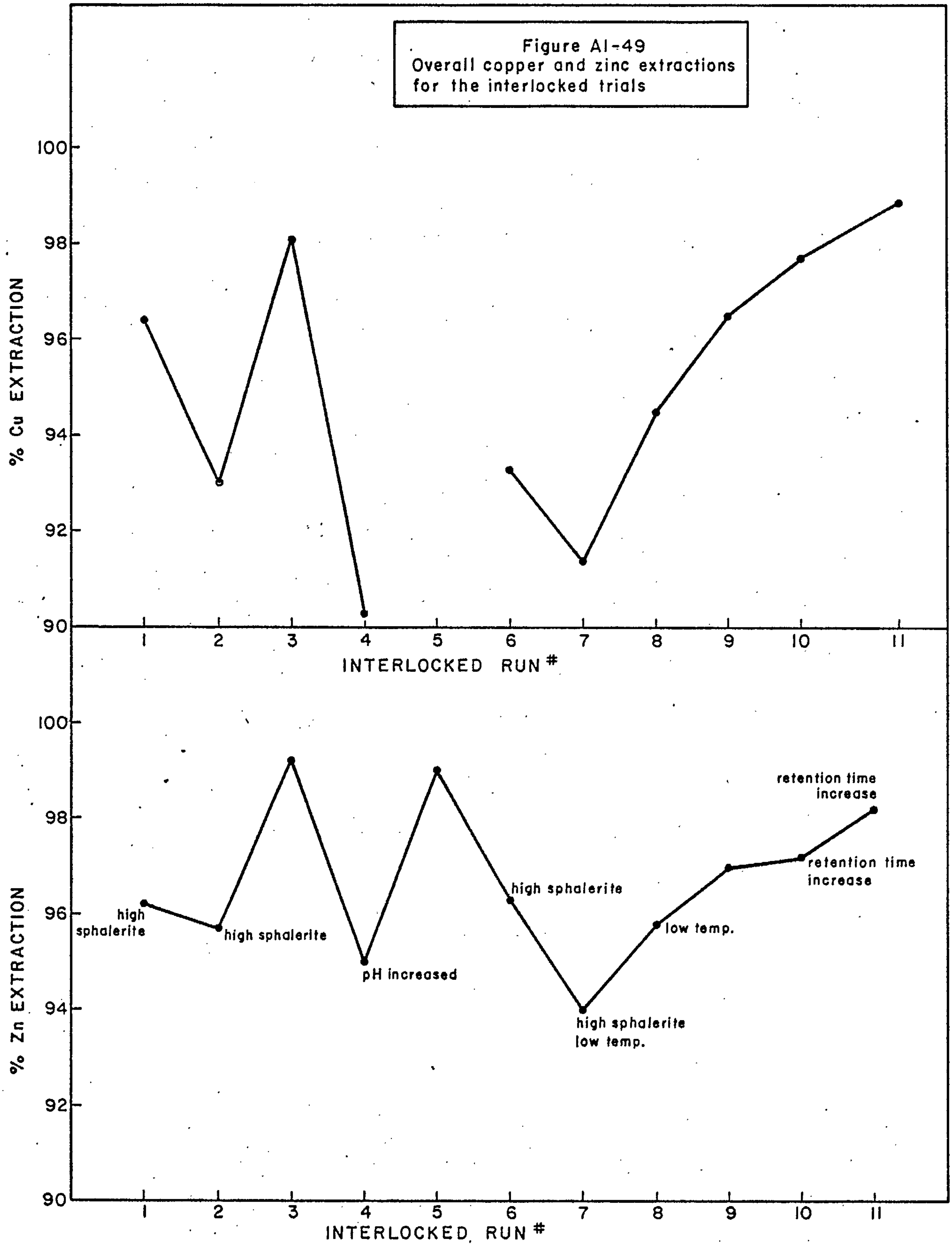


Figure AI-49
Overall copper and zinc extractions
for the interlocked trials

seen that negligible leaching of ferrite and sphalerite occurs under neutral leach conditions. Hot acid leaching, dissolves greater than 95% of the ferrite and greater than 60% of the sphalerite. Calculations of absolute extractions are not valid by this x-ray technique due to changes in the matrix. The neutral leach residue does not contain any of the zinc oxides, oxysulphates, or sulphates present in the calcine. For the sulphation roast calcine alone, when the iron oxidation recycle to the neutral leach was employed (Figure A1-50) significant quantities of gypsum appear in the hot acid leach diffraction trace. Although a larger ratio of ferrite to hematite was found in the hot acid leach residue for the combined calcine than for tailings refloat calcine residue alone, there is no conflict with chemical analysis regarding zinc extraction since the mass of hematite in the calcine differ by 1:2.5. The lower copper extraction in the combined leach is due to the formation of copper ferrite in the dead roast. The hot acid leach filtrate volume was controlled to ensure thermal balance on its recycle to the sulphation roaster.

Extensive recrystallization of zinc and iron sulphates were experienced during the filtration of the hot acid leach pulp for the combined calcine. This was attributed to the decrease in solution temperature on filtration from 95°C to 55°C and the fact that the solution was oversaturated in sulphate due to the use of spent electrolyte as the repulp and makeup media. Oversaturation can be avoided by using

Figure A1-50
X-RAY DIFFRACTION TRACES
TAILINGS REFLOAT PRODUCT
(Effect of Leaching Conditions on Zinc Compounds)

BMD-61
Air Ratio, Q = 1·85
Roasting Temperature, T = 695° C

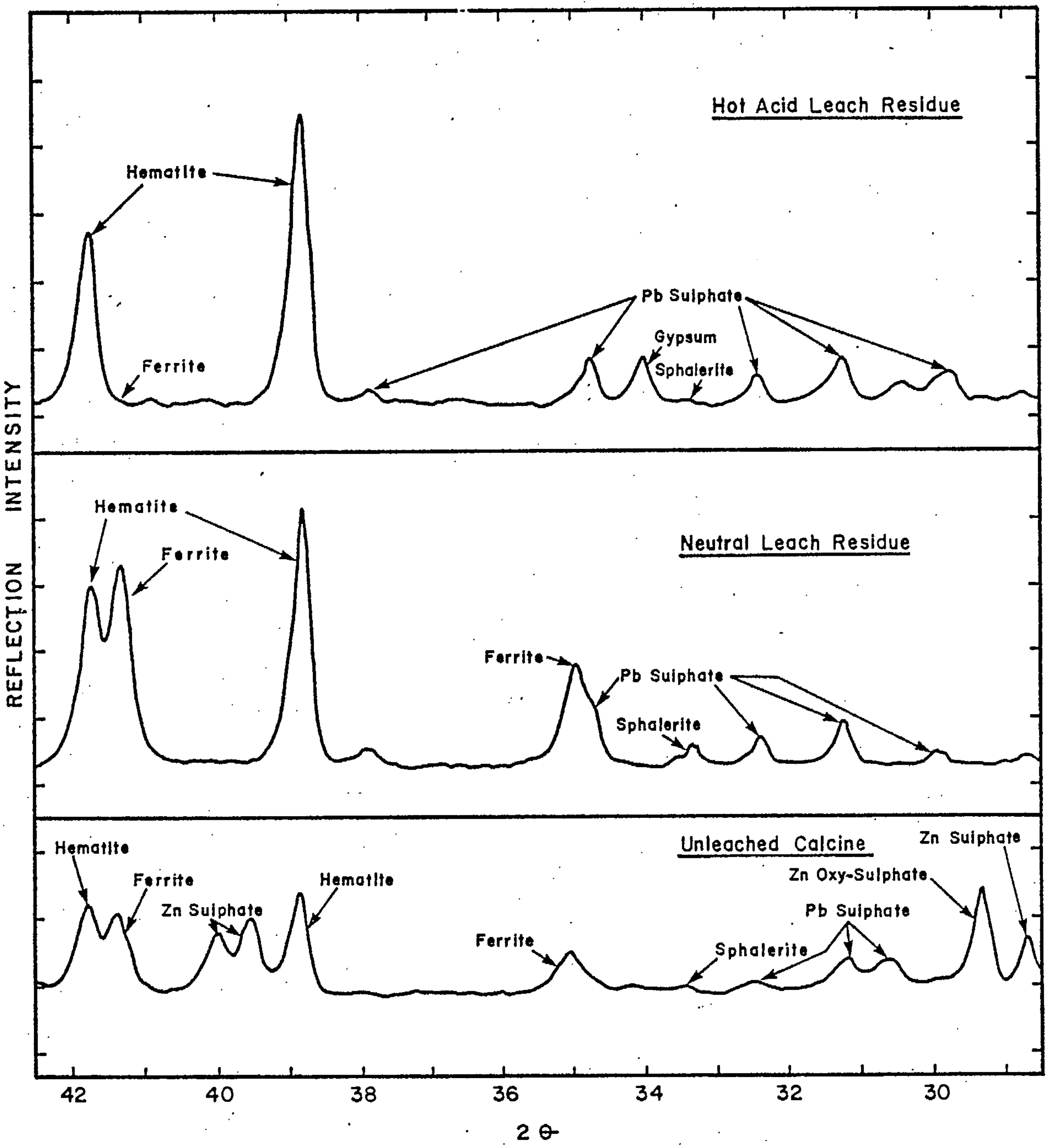
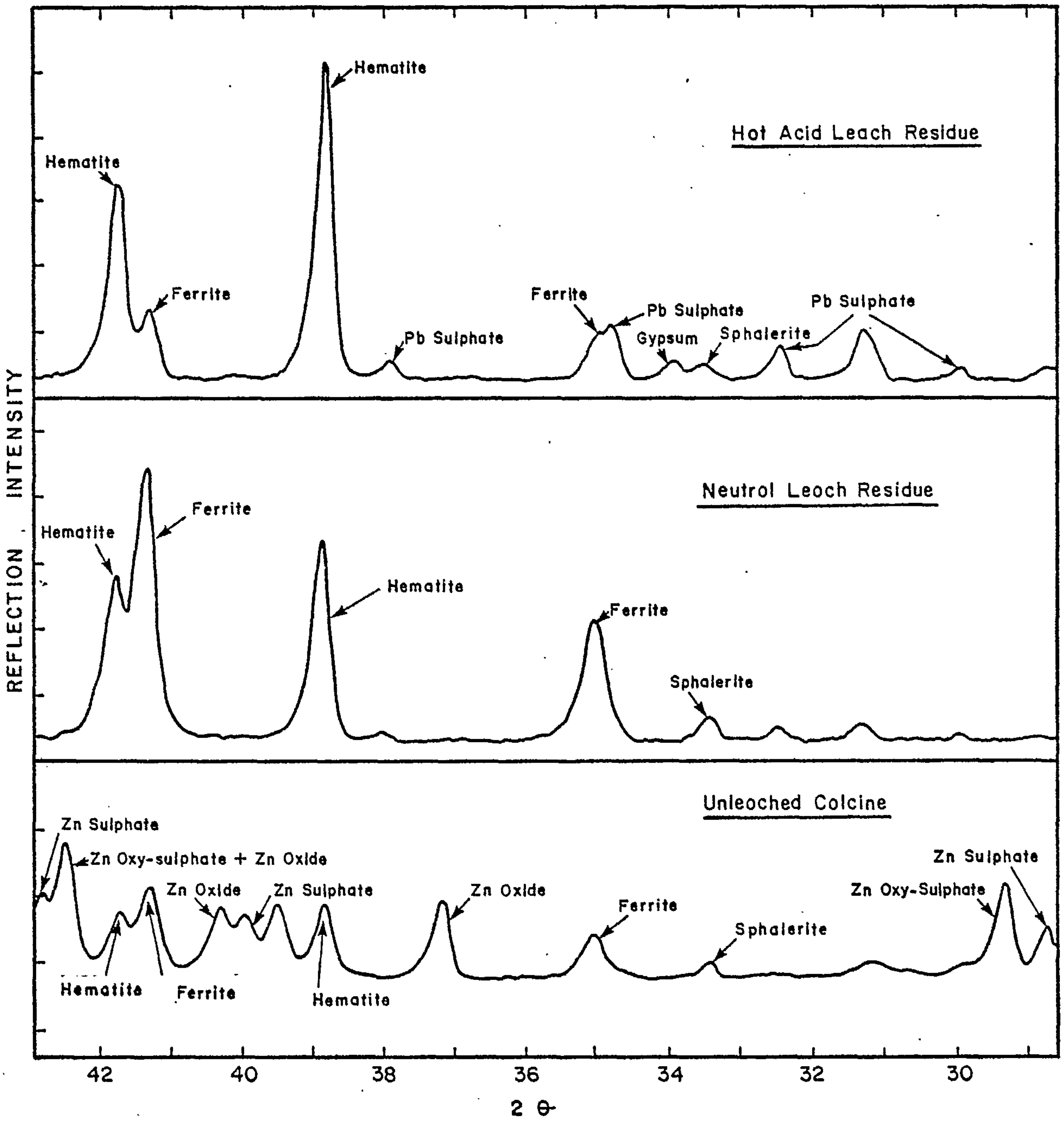


Figure A1-51
X-RAY DIFFRACTION TRACES
INTEGRATED SULPHATION ROAST – DEAD ROAST CALCINE
(Effect of Leaching Conditions on Zinc Compounds)



fresh water for repulp and makeup, decreasing the zinc and sulphate tenors by approximately 50 and 70 gpl respectively.

.614 Hot Acid Leach Filtration

The solids in this pulp are almost entirely hematite and lead sulphate (see Table A1-33). Although the slurry does have a high solids content, filtration was rapid throughout the test program.

A constant pressure filtration was carried out as for the neutral leach filtration above. The results are given in Table A1-35 and plotted in Figure A1-52. The constant initial filtrate flowrate suggests that pump starvation was occurring due to the small diameter of the suction line (1/4 in. i.d.). Over the test, the average filtrate flow rate was 15.56 $lt/min/m^2$ (0.38 US $gal/min/ft^2$) or about 2.5 times the neutral leach filtrate rate.

A similar test was attempted for the hot acid leach pulp from the BMD62-CEZ material. This test had to be abandoned because of line plugging and solution cooling.

.615 Solvent Extraction of Copper

The solvent extraction of copper was carried out in a Bell Engineering continuous mixer-settler laboratory unit (see Plate 2) with three stages of extraction and two stages of stripping, with the temperature in the extraction mixers maintained at 40-45°C using minature immersion heaters and a mixer retention time of 1 min. per stage. The O/A ratio in both extraction and stripping was held at 1.0 with no re-

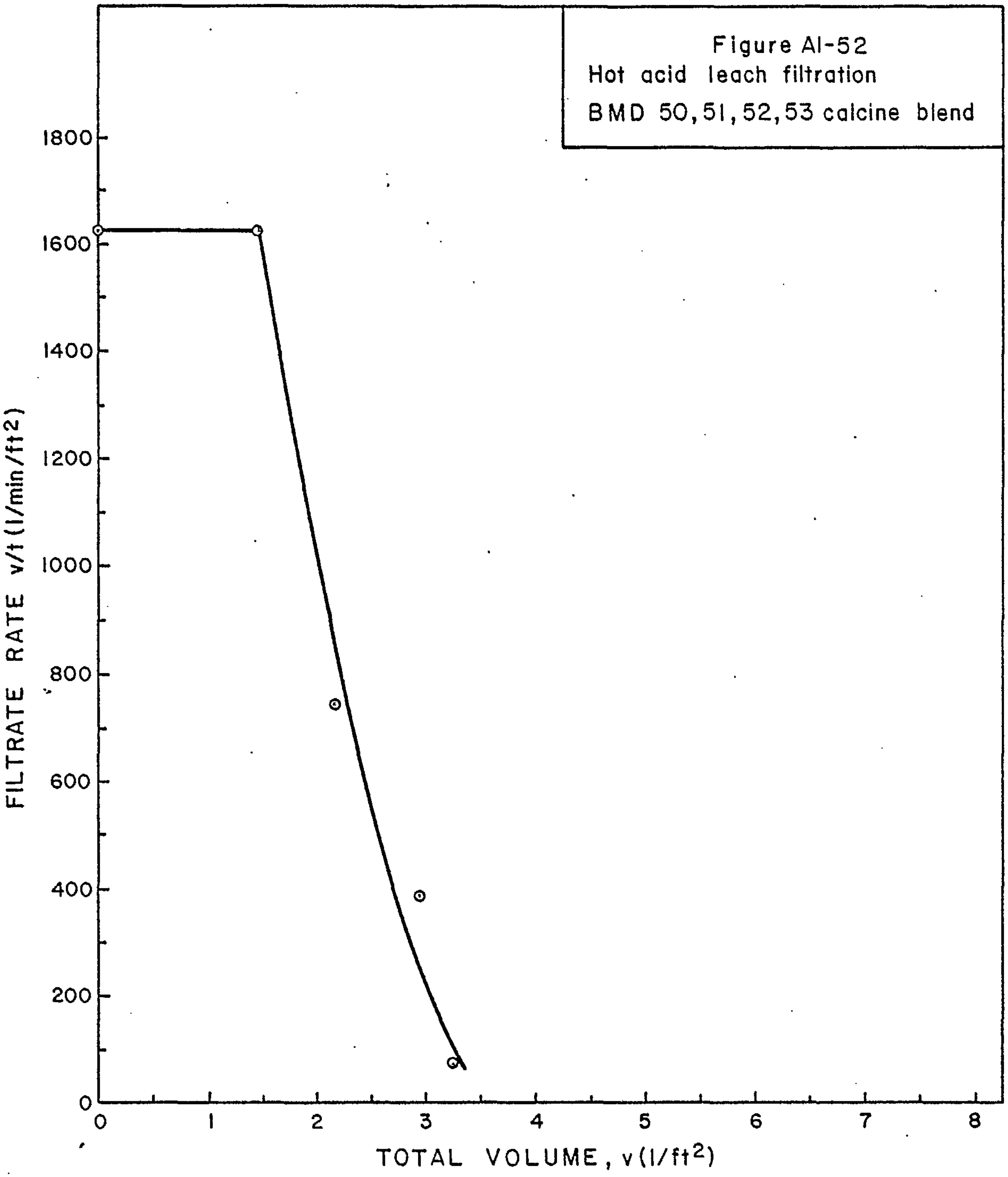


Figure AI-52
Hot acid leach filtration
BMD 50,51,52,53 calcine blend

TABLE A1-35

HOT ACID LEACH FILTRATION
BMD 50, 51, 52, 53 CALCINE BLEND

Total Vol. (lt)	Total Time (min)	Vol. (cc)	Time (min)	Rate (ml/min/ft^3)
0.0	0.0	450	0.08	1626
2.5	0.5	-	-	-
5.0	0.9	450	0.08	1626
7.5	1.4	410	0.16	741
10.0	1.9	385	0.29	384
11.3	2.5	100	0.38	76

Total filtrate	12.5 lt
Pressure	60 psig
Filter area	3.46 sq.ft.
Cake thickness	0.5 in.
Dry wt. filter cake	8.37 kg
Cake moisture	18.4%
Solids in feed	28.8%

cycles. Mixer volume in the unit is approximately 180 ml and the settler area is about 61.5 cm^2.

LIX 64N (General Mills Chemicals) was used as the extractant based on General Mills' recommendation although the temperature of extraction makes its use borderline. Kermac 470B (Kerr McGee Crop.) was the diluent. A LIX concentration of 10% (by volume) was chosen on the basis of some preliminary shaking tests and the estimated copper concentration of the neutral leach filtrate. However, the projected level of copper was not achieved because of a change in the feed. The organic was stripped with a spent electrolyte (150 gpl H_2SO_4, 30 gpl Cu) which was continuously regenerated by electrowinning copper.

The first batch of solution passed through the SX unit performed very well but subsequent runs proved disastrous in terms of phase separation in the extraction stages. Experimentation with the unit showed that in order to prevent emulsion problems:

1. The feed solution must be as fresh as possible.
2. The feed solution must not be allowed to cool between the neutral leach and SX or in the copper extraction stages.
3. All the extraction and stripping mixers should be operated in the aqueous continuous condition.

If the above conditions were met, the unit ran indefinitely with little or no adjustment or attention required.

Ageing of solutions prior to SX is a common phenomenon. In this case it is probably caused by the oxidation of ferrous ion with subsequent precipitation of ferric hydroxide. This results in crud formation in the extraction stages.

Any cooling of the feed solution, which contains 150 gpl zinc at pH 4, results in the precipitation of basic zinc sulphate. This in turn causes the formation of crud in the SX. No precipitation was observed in the unheated raffinate storage tank, even after long periods of cooling, but this solution is at a lower pH. During the later stages of the test program, it was demonstrated that the heat input to the second extraction stage could be reduced and could be eliminated entirely in the third stage due to decreasing pH.

No explanation can be effered for the need to have all the extraction and stripping mixers operate in the aqueous continuous condition. Initially, an attempt was made to operate extraction stage 3 and strip stage 2 in the organic continuous mode. Observations indicated that organic continuous in any of the extraction mixers soon resulted in the corresponding settler filling with emulsion. Problems with organic continuous operation were less severe in the strip but there was a definite increase in the emulsion band thickness in the settler during organic continuous operation relative to the thickness during aqueous continuous operation.

As expected, SX removed copper with no detectable change in any of the other elements in solution. Copper extraction was variable because of the variable copper content of the feed solutions. The levels of copper in the raffinate solution are shown in Table A1-36, note that in most cases, some solution was bypassed because the laboratory SX unit could not keep up with the process requirements for solution. Also, few samples were taken when serious phase separation problems occurred and in the later stage of the project, the analytical laboratory was overloaded, thus data is minimal.

Toward the end of the test program, permanganate colour appeared in the copper electrowinning section, caused by the carryover of entrained feed solution into the electrowinning section. This permanganate could oxidize the organic reagents, although minimal oxidation was detected by General Mills on samples sent to them after completion of the test program.* If this phenomenon is a problem in full scale plant, the copper electrolyte can be dosed with iron. This would result in reduction of the permanganate by ferrous ion, which would be regenerated in the electrolytic cells.

The SX tests demonstrate that with proper temperature control and control of the continuous phase in the mixers, solvent extraction can remove over 95% of the copper from the neutral leach solution (or copper can be removed to less than 50 mgpl) with a retention time of one minute. Some analyses suggest that it may be possible to extract sufficient copper with two extraction stages. The one minute retention time and/or two extraction stages would result in a considerable capital cost saving over a conventional SX plant operating on a dump leach solution.

.616 Iron Oxidation and Neutralization

Oxidation of ferrous iron in neutral leach filtrate with manganese dioxide, neutralization with limestone to precipitate ferric hydroxide, flocculation of the suspension and settling was performed in a single 130 litre tank

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* see p. 24

TABLE A1-36

CONDITIONS AND RESULTS OF IRON OXIDATION AND NEUTRALIZATION RUNS
FROM INTERLOCKED TRIALS

Iron Oxidation Run #	Neutral Leach Filtrate #	%SX[a]	Initial Solution					Iron Oxidation			Neutralization		
			Composition (gpl) Cu	Fe^{2+}	Vol. (lt)	pH	Temp. (°C)	MnO$_2$ Add'n (gpl)	Ret. Time (hr)	Final Fe^{2+} (mgpl)	Limestone Add'n (gpl)	Ret. Time (hr)	Final pH
1	1	100	0.24	0.96	100	2.4	60	2.3	1.0	<10	3.5	0.5	4.2
2	2 + 3	5	2.00	1.20	115	2.9	60	2.7	1.0	<10	3.0	0.5	4.1
3	5	0	2.25	1.00	120	3.8	60	2.3	1.0	260	0.0	0.0	4.0
4	6	–	–	0.47	120	2.8	62	1.3	1.0	<10	3.0	0.5	4.3
5	4	100	0.03	1.07	120	2.5	60	1.0	1.0	<10	4.0	0.5	4.1
6	7	50	1.09	1.09	120	2.4	65	0.9	1.0	<10	4.2	0.5	4.2
7	8	100	0.09	1.17	120	1.4	60	0.9	0.7	15	7.4	1.5	4.2
8	9	70	0.64	0.60	120	2.3	60	0.5	1.0	<10	3.5	1.5	4.3
9	10	95	0.13	0.44	120	2.4	60	0.5	1.0	<10	4.2	0.8	4.2
10	11	0	–	0.77	120	2.5	65	0.7	1.0	<10	4.0	0.5	4.4
11	13	0	1.62	0.74	–	2.5[b]	62	0.6	1.0	<10	3.0	0.5	4.2

a – approximate % of solution which was passed through the SX unit

b – spent added to decrease pH

TABLE A1-37

CONDITIONS AND RESULTS OF IRON HYDROXIDE, MANGANESE DIOXIDE GYPSUM FLOCCULATION AND SETTLING RUNS FROM INTERLOCKED TRIALS

Iron Oxidation Run #	Conditions		Decant Solution to 1st Stage Purif.							Residue Underflow			
	Percol 156 Add'n (mgpl)	Ret. Time (hr)	Solution #	Vol. (lt)	S.G. (g/cc)	Composition (mgpl)				Res. #	To N.L.	S.G. (g/cc)	Vol. (lt)
						Cu	Cd	Co	Fe (total)				
1	1.0	6	1	80	–	118	–	–	–	1	4	–	20
2	0.9	20	2	104	–	1130	–	–	–	2	5	–	11
3	0.8	8	3	100	–	1460	174	70	299	3	6	–	20
4	0.8	12	4	110	–	1412	206	57	9	4	7	–	10
5	0.5	10	5	106	–	–	242	70	–	5	8	–	14
6	0.3	12	6	103	–	510	160	64	20	6	9	–	17
7	0.8	36	7	100	–	10	236	70		7	10	–	12
8	0.8	10	8	106	–	84	192	70	27	8	11	–	14
9	0.8	12	9	107	1.28	–	–	–	–	9	12	1.31	13
10	0.8	12	10	105	–	–	–	–	–	10	13	–	15
11	0.5	4	11	–	–	1100	–	–	<10	11	–	–	–

equipped with an underflow pump and siphon for decantation. The tank was 32 inches deep by 17 inches in diameter. Agitation was controlled and a two turbine system was employed. Results are given in Tables A1-36 and A1-37.

A 50 per cent excess of manganese dioxide to stoichiometeric was sufficient in all cases (except at pH 3.8) for oxidation of iron to less than 15 mgpl ferrous in one hour. The pH of the solution from solvent extraction (SX) varied from 1.4 to 2.9 and the temperature during oxidation controlled at 60-65°C.

After the time allotted for effective oxidation of ferrous to ferric, milled New Brunswick limestone (-100 mesh, <0.30 per cent Mg) was added gradually until the solution was neutralized to pH 4.0 to 4.4. The quantity of limestone necessary for neutralization ranged from 3.0 to 7.4 gpl (average 4.1) and the time necessary for neutralization was 0.5 hours.

The fine suspension produced from the oxidation and neutralization was flocculated with Percol 156, supplied by Allied Colloids. Additions were varied from 0.3 to 1.0 mgpl. Attempts at filtering this flocculated slurry were unsuccessful due to the slimy cake, filter cloth blinding, and high zinc losses in the cake. Accordingly, after flocculation, the solids were allowed to settle undisturbed until the tank was required for the next solution. The clear solution was decanted to the first stage purification and the wet residue pumped to the neutral leach tank.

The small size of the solvent extraction unit resulted in organic carryover visible in the solutions to iron oxidation. After flocculation, however, no evidence of organic remained in the decant solution. Although the mechanism for this 'cleaning' (i.e., adsorption, absorption, physical entrapment in the ferric hydroxide-manganese dioxide-gypsum sludge or oxidation, or simple volatilization) was not determined, the result was nevertheless most significant. In a plant scale operation, organic carryover, although much less than observed in the small pilot tests, would probably be purged from the solution in the same manner.

The recycling of the iron oxidation sludge to the neutral leach greatly affected the filtration of the neutral leach pulp. Filtration experiments carried out before this recycle gave average filtration rates of 0.8 lt/min/ft^2 for filtration to a 2" thick cake. This rate was reduced to 0.56 lt/ min/ft^2 to a 0.5" thick cake and some sliming became evident in the cake. Although this negative effect is significant, it is felt that the sludge must be handled by this route in order to avoid high zinc and copper losses and the very poor filtration characteristics of the sludge alone.

The small quantity of solids in this solution plus the high proportion of ferric hydroxide in the solids makes flocculation and settling the most practical way of treating this solution. Percol 156 was chosen as the flocculant based on zinc plant experience with the flocculation of ferric hydroxide and this choice was confirmed during the bench scale phase of hydrometallurgical testing. Flocculation of the suspended solids appears to be immediate upon dispersion of Percol 156 in the solution; this product was used throughout the semi-pilot phase of the project with satisfactory results.

Beaker tests similar to those described in Section 132.612 showed that doseages in excess of 0.5 mgpl of Percol 156 produced no improvement in floc size, flocculation rate or settling rate. This dosage rate was employed in settling tests at 60°C, using the same apparatus and method described above, (131.612). The results are given in Table A1-38 and presented as a Roberts plot in Figure A1-53 with H_∞ equal to 110 cc, which shows the compression point clearly.

The Talmage and Fitch method of data analysis (10,11) gives the solids flux that can be passed through a clarifier (G_t) equal to 0.143 TPD/m^2 (0.0146 TPD/ft^2). If the clarifier underflow is 15% of the original volume, then a compression depth (Pd) of 0.272 m (0.510 ft) is required.

.617 First Stage Purification - Copper and Cadmium Removal

Conventional zinc dust cementation techniques were employed for the removal of cadmium and residual copper from iron oxidation decant solution. A single 130 litre tank (17" dia. x 32" depth), two turbine variable agitation, and the Perrin filter press constituted the apparatus for this process. Results obtained were comparable to those experienced in commercial application of this technique. At conditions of 1.3 hours retention time, pH 4.0 to 4.3, and temperature 60-65°C, first stage purification filtrates

TABLE A1-38

SETTLING TEST

IRON OXIDATION-NEUTRALIZATION SOLUTION, 0.5 mgpl PERCOL 156 BMD 50, 51, 52, 53 CALCINE BLEND

Time (min)	*Height of Interface (cc)	$H-H_\infty$ (cc)
0.0	1000	890
2.5	770	660
4.1	640	530
6.1	520	410
8.4	430	320
10.5	390	280
13.9	350	240
15.5	330	220
19.7	300	190
23.3	280	170
29.3	260	150
32.7	250	140
39.8	230	120
47.9	210	100
54.3	200	90
66.6	180	70
74.9	170	60
82.2	160	50
93.2	150	40
104.5	140	30
125.0	130	20

Overnight interface height = 110 cc

H_∞ = 110 cc

Solution s.g. = 1.29

Solids s.g. = 2.71

Weight of Solids In Test = 9.51 g

Supernate clarity (125 min)= 97 mgpl

* 864 cc/ft on graduate

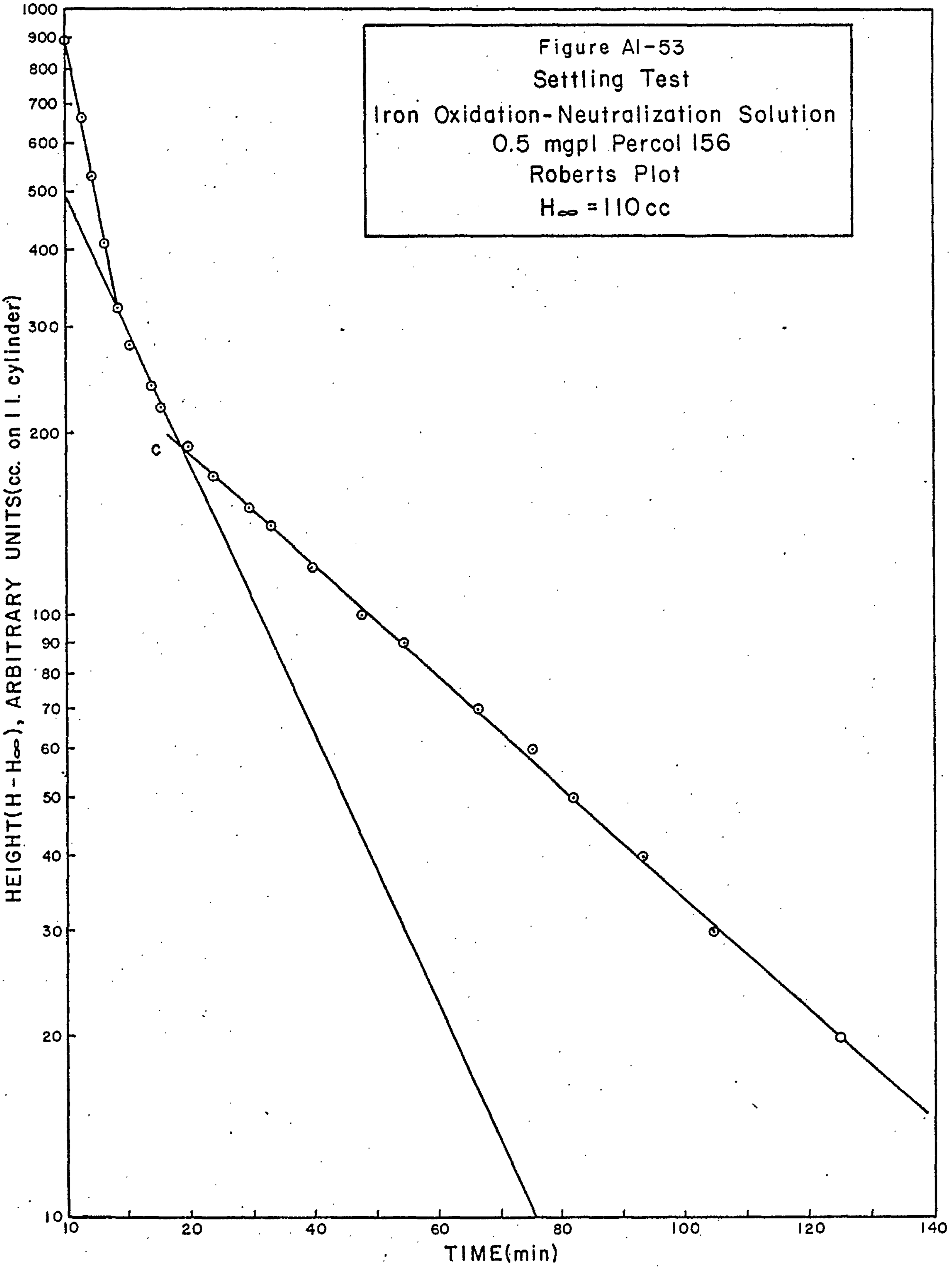
Figure AI-53
Settling Test
Iron Oxidation-Neutralization Solution
0.5 mgpl Percol 156
Roberts Plot
H∞ = 110 cc
HEIGHT(H - H∞), ARBITRARY UNITS(cc. on 1 l. cylinder)
TIME(min)
C
1000
900
800
700
600
500
400
300
200
100
90
80
70
60
50
40
30
20
10
10
20
40
60
80
100
120
140

containing less than 1 mgpl copper and cadmium were consistantly produced. Minimum zinc dust additions were calculated at 1.25 times stoichiometric addition for copper removal plus 3.00 times stoichiometric addition for cadmium removal. Some contamination of solution in the filter press due to incomplete purging of lines and pumps was experienced (see Section 131.618). Results and conditions are given in Table A1-39.

.618 Second Stage Purification - Cobalt Removal

Antimony activated cementation on zinc dust was the technique employed for this purification. The apparatus consisted of a 140 litre tank (26" dia. x 18" depth), a large flat bladed turbine to reduce vortexing, an 18 kw 316 SS immersion heater, and the Perrin filter press. Conditions of 3.0 hour retention time, 95±2°C, zinc dust addition of 8 gpl, and Sb_2O_3 addition of 10 mgpl were employed. Although these conditions were sufficient in the preliminary tests to reduce the cobalt and antimony to less than 0.1 mgpl, difficulties were encountered in reducing cobalt to this level in the large scale operation as shown in Table A1-40. A comparison of the cobalt in solution with time for a preliminary test and for an interlocked test is shown in Figure A1-54. Although some initial recontamination with cobalt occurs in the filter press due to incomplete purging of leach and 1st stage purification filtrates, the major cause of high cobalt was redissolution.

TABLE A1-39

CONDITIONS AND RESULTS OF FIRST STAGE PURIFICATION FROM INTERLOCKED TRIALS

First Stage Pur. Run #	Decant Solution #	Conditions								Filtrate							
		Ret. Time (hr)	pH	Temp. (°C)	Vol. (lt)	Zn Dust Add'n.				Filtrate #	Vol. (lt)	pH	Composition (mgpl)				
						gpl	% of Stoichiometric						Zn (gpl)	Cu	Co	Cd	Fe (Total)
							for Cu		for Cd								
1	1	1.3	4.2	65	80	1.55	125	+	400	1	80	4.4	133	<0.5	51	0.9	63
2	2	1.3	4.1	65	105	4.10	125	+	400	2	100	4.5	135	1.0	47	0.5	60
3	3	1.3	4.0	65	100	2.70	125	+	232	3	100	4.4	148	0.8	57	0.9	260
4	4	1.0	4.3	65	110	1.42	0	+	400	4	110	4.5	130	<1.0	51	<<1.0	28
5	5	1.3	4.1	60	106	1.25	0	+	300	5	110	4.4	122	1.6	62	<<1.0	16
6	6	1.0	4.2	65	103	1.19	125	+	200	6	105	4.4	120	0.7	--	2.5	22
7	7	1.3	4.2	65	100	1.22	0	+	300	7	100	4.5	130	<<1.0	--	<1.0	4
8	8+9	1.3	4.2	65	125	1.14	0	+	300	8	120	---	141	<1.0	--	<1.0	22
9	9+10	1.3	4.3	65	125	1.38	0	+	400	9	125	---	---	0.1	52	0.1	--

TABLE A1-40

CONDITIONS AND RESULTS OF SECOND STAGE PURIFICATION FROM INTERLOCKED TRIALS

Second Stage Pur. Run #	First Stage Filtrate #	Conditions							Filtrate									
		Vol. (lt)	Ret. Time (hr)	pH	Temp. (°C)	Sb_2O_3 Add'n (mgpl) at time 0	Zn Dust Add'n (gpl)		Filtrate #	Vol. (lt)	Composition							
							at time 0	at time 20			gpl			mgpl				
											Zn	Mn	Mg	Cu	Cd	Co	Sb	Fe
1	1	80	3.0	4.5	95	10	8.0	0	1	80	137	-	-	<<0.1	0.6	3.9	<<0.1	-
2	2	100	3.0	4.5	95	10	8.0	0	2	100	140	-	-	0.7	1.3	2.1	0.1	22
3	3	100	3.0	4.4	95	10	8.0	0	3	100	151	-	-	<1.0	1.0	8.0	<0.1	195
4	4	110	3.0	4.5	97	10	8.0	2	4	110	130	-	-	0.2	<<1.0	<1.0	<<0.1	6
5	*3 + 5	135	3.0	4.5	93	10	8.0	0	5	135	127	-	-	0.4	<<1.0	1.2	<0.1	12
6	*3 + 6	135	3.0	4.5	97	10	8.0	0	6	135	123	-	-	0.7	<<1.0	1.9	<0.1	20
7	*3 + 7	135	3.0	4.5	95	10	8.0	0	7	135	134	3.3	-	-	-	-	-	-
8	8	120	3.0	-	95	10	8.0	0	8	120	141	3.6	0.57	0.6	0.5	2.4	-	-
9	9	125	3.0	-	95	10	8.0	0	9	125	-	-	0.68	0.2	0.3	2.6	-	-

* Co filtrate from 3 was oxidized with $KMnO_4$ and decanted into runs 5, 6 and 7

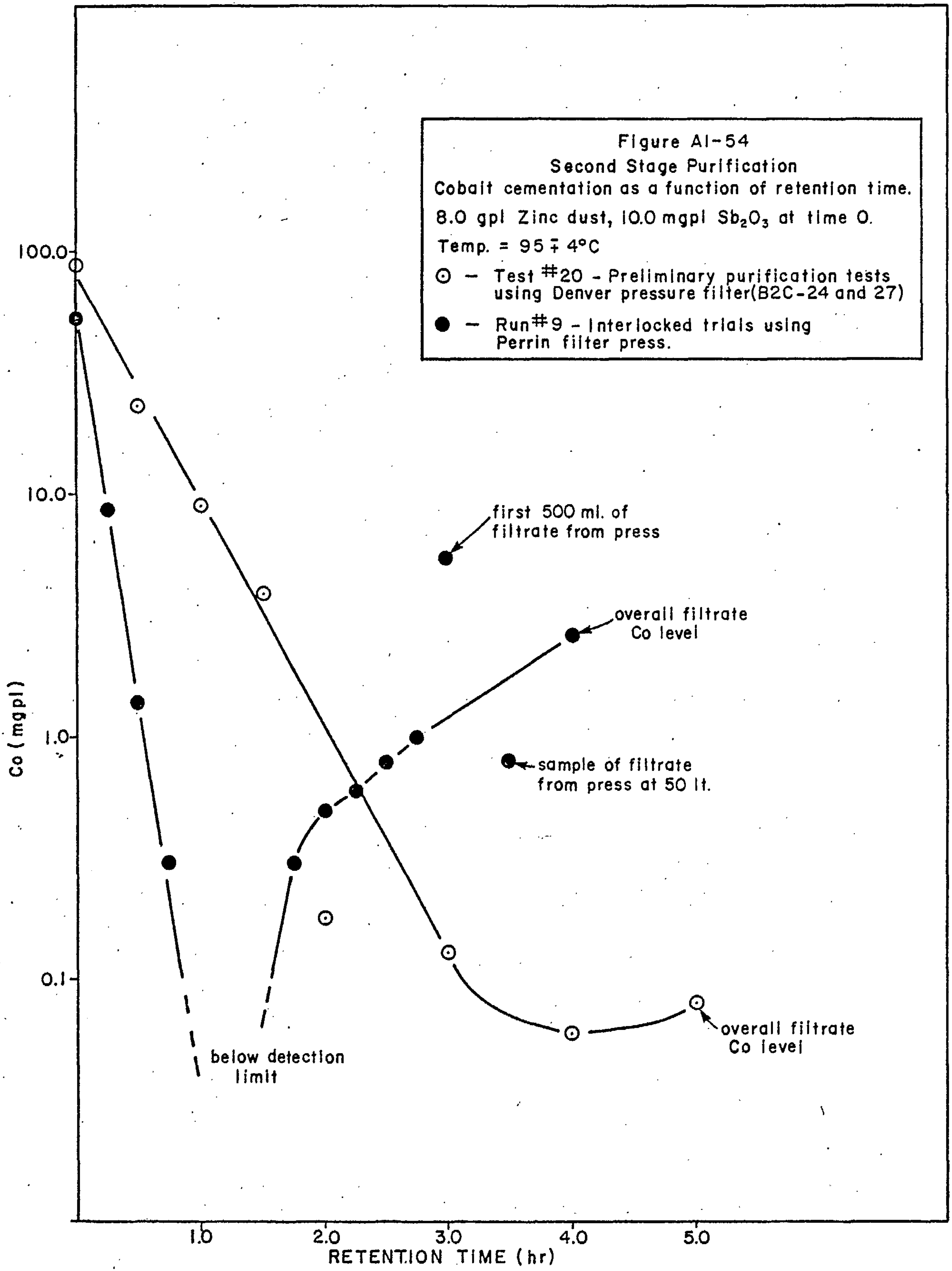

Figure AI-54
Second Stage Purification
Cobalt cementation as a function of retention time.
8.0 gpl Zinc dust, 10.0 mgpl Sb_2O_3 at time 0.
Temp. = 95 ∓ 4°C

The semi-continuous mode of operation in the interlocked operation is believed to be the cause of the observed redissolution. The design of the tank and heating coils resulted in nearly 25% of the cementation products remaining in the tank after filtration. Exposure to air, resulted in partial oxidation. When the tank was filled with the next solution (usually 5-10 hours later), heated, and fresh zinc dust and antimony added, the initial rate of cementation increased due to the presence of the old cementate. With time, this old product oxidized further and redissolution became evident. Only when fresh zinc dust was added at the first signs of redissolution was solution produced of acceptable cobalt concentration. The high cobalt in Run #1 was attributed to contamination from the filter press and excessive vortexing resulting in high oxygen levels in the solution. The impeller design was altered to reduce vortexing after Run #3.

It is believed that oxidation due to air exposure would be minimal in a fully continuous operation. Also filter presses would be devoted entirely to this operation and the continuous operation would reduce oxidation and contamination during filtration. A partial recycle of the coarser fraction of cementation product (no exposure to air) may considerably reduce the amount of fresh Sb_2O_3 and zinc dust necessary for cobalt removal to less than 0.2 mgpl as indicated by the initial slopes in Figure A1-54. These hypotheses are supported by previous work carried out by one member of the research team (Dr. R.S. Salter) and work by Vieille Montagne (Can. Pat. #983,726; 17 Feb., 1976), Outokumpu Oy (U.S. Pat. #3,979,266; 7 Sept. 1976), and Cominco (Can. Pat. #895,794; 21 Mar. 1972).

.619 Zinc Electrowinning

The two zinc electrowinning cells were similar to the laboratory cell employed in the preliminary tests. The "neutral feed" was added to the recirculating spent from a constant head tank and the resultant upgraded spent was pumped to the two cells which were operated in parallel with respect to solution flow. The electrical feed to the cells was in series (cathode header bar on cell 1 to anode header bar on cell 2). The cells were designed according to the Noranda Research Centre design for pilot cells constituting a gas evolving electrode. Each cell contained four anodes (immersed area 3.75" wide by 34" long) and three cathodes (deposition area 4" wide by 35" long). Electrolyte cooling was effected by polyethylene cooling coils in the cold spent tank. Additional coils were added to each cell after Run #2.

The difficulties experienced in cobalt removal during the interlocked trials and tight scheduling not permitting further purification resulted in neutral feed solutions which were not suited for efficient electrowinning. It was decided, however, to continue the electrowinning trials to refine the apparatus and to produce spent electrolyte suitable for neutralization trials and as acidifying agent in the leach. The results of these tests are given in

TABLE A1-41

CONDITIONS AND RESULTS OF ZINC ELECTROWINNING RUNS FROM INTERLOCKED TRIALS

Run #	Sol'n in cell at time = 0	Neutral Feed Sol'n (2nd Stage Filtrate #)	# of cells in use	Temp. (°C)	Stripping Cycle (hr)	N.F. Rate (ml/min)	Current (Amps)	C.D. (Asf)	Deposit Wt. (kg)	C.E. (%)	Deposit Appearance	Comments
1	#1 2nd Stage Pur. Filtrate	---	2	36-51	15	0	303	55	8.2	74	very nodular many pinholes	Co too high, more cooling necessary
2	#1 Spent	2	2	36-47	19	156	303	55	5.0	68	very nodular many pinholes cobwebbed dendritic growth*	Co too high, more cooling necessary, power breakdown & feed blockage for 9 hr. during run
3	#2 Spent	4	2	36-42	12	156	303	55	7.2	85	smooth, few pinholes or nodules	
4	#3 Spent	5	2	36-42	15	156	303	55	7.0	63	very nodular many pinholes cobwebbed dendritic growth*	feed blockage for 10 hr. during run, Co too high
5	#4 Spent	6+7	2	36-41	24	156	303	55	14.2	80	few nodules badly pinholed	Co, Cu too high
6	#5 Spent	7+8	1	36-41	48	60	230	42	9.2	68	many pinholes dendritic growth*	feed blockage for 5-10 hr. during run, Co, Cu too high
7	#6 Spent	8+9	1	36-43	48	60	230	42	10.5	78	some pinholes many nodules	Co too high

Note - glue and sodium silicate added in all runs.
* - evidence of shorting at bottom of cathode.

Table A1-41.

.620 Spent Electrolyte Neutralization

A detailed discussion of the test work performed on spent neutralization has been given in Section 131.5. The batch techniques employed in the interlocked runs were unsuccessful in producing gypsum residues of low moisture and zinc content. Batch additions of limestone to high acidity solution attributed to the armouring of unreacted limestone with precipitated gypsum. High zinc losses in the gypsum residue were expected because the residue could not be washed in the Perrin filter press. Sufficient neutralized spent was produced from the two large batch operations and the continuous operations which followed to provide the necessary volumes required for neutral leach makeup. Results for the batch trials are given in Table A1-42.

.621 Trace Element Analyses

Nickel	-	Neutral leach solutions contained 1.8±0.8 mgpl with no buildup evident. Ni was effectively removed to less than 0.2 mgpl in the neutral feed solution during zinc dust purification. The hot acid leach liquors contained between 8.9±2.8 mgpl Ni with no buildup. Some of the nickel entered the system due to oxidation of 316 SS in the roasting and leach section.
Indium	-	Neutral leach solutions contained less than 0.5 mgpl (except in one anomalous case at 9.6 mgpl). All neutral feeds contained less than 0.5 mgpl. Indium in hot acid leach liquors reached and maintained an equilibrium value of 230±50 mgpl by the second recycle. The hot acid leach residues contained 45±5% of the In in the feed. The remainder reported in the roaster off-gas. The hot acid leach residues contained 160±20 ppm In.

Tin - Neutral leach and neutral feed solutions contained less than 0.5 mgpl, the detection limit. Greater than 95 per cent of the tin in the feed reported to the hot acid leach residue. These residues contained 925±100 ppm Sn.

Germanium - Neutral leach solutions contained less than 3 mgpl, the detection limit.

Gallium - Neutral leach solutions contained less than 2 mgpl, the detection limit. Hot acid leach liquors and residues contained about 9 ppm.

Mercury - Neutral leach solutions contained less than 0.001 mgpl. Hot acid leach residues contained about 0.25 ppm.

Chloride - Chloride in neutral leach solutions averaged 78 mgpl over Runs #8 to 14. Most of the chloride entered the hydrometallurgical system from makeup and wash water which was untreated Fredericton city water (20-30 ppm) and the leaching of PVC tubing by concentrated acid. Most zinc electrowinning plants average about 60-90 mgpl chloride in the electrolyte. The chloride level in the hot acid leach liquors was 1000±500 mgpl depending on the amount of concentrated sulphuric acid required for the leach. Hot acid leach residues contained 100±50 ppm chloride.

Fluoride - Neutral leach solutions contained less than 0.1 mgpl. Hot acid leach liquors contained between 8 and 56 mgpl (random fluctuations, no buildup experienced.).

Antimony - Neutral leach solutions contained less than 0.1 mgpl antimony. The antimony concentration in the hot acid leach filtrate gradually built up to an equilibrium level of 27.5 mgpl which was maintained through the last four runs.

Bismuth - 40±10% of the bismuth in the feed reported to the hot acid leach residue, the remainder exiting in the roaster off-gas. The hot acid leach recycle contained 60±30 mgpl Bi with no buildup evident.

Selenium - Neutral leach solutions contained less than 0.01 mgpl. Levels in the hot acid leach liquor were less than 1 mgpl.

TABLE A1-42

RESULTS OF BATCH SPENT NEUTRALIZATION RUNS FROM INTERLOCKED TRIALS

Run #	Spent from Electrowinning (Run #)	Spent (#)	Vol. (lt)	Spent Composition (gpl)				Neutralized Spent+						Gypsum Residue*				
				Zn	Acid	Mn	Mg	N.S. (#)	Vol. (lt)	pH	Composition (gpl) Zn	Mn	Mg	Wet Wt. (kg)	Solids (%)	Composition (%; dry) Zn	Mn	Mg
1	2	2	90	42	63	2.47	0.48	1	50	4.3	46	2.2	--	46.0	50	6.6	0.07	0.1
2	4	4	60	62	47	2.47	0.29	2	50	4.5	60	2.47	--	29.5	45	4.6	0.04	0.1

+ - reaction very slow due to difficulties with limestone employed (high Mg, Fe; very unreactive), excess limestone always added, solution filtered at pH 4.0.

* - residue not washed.

.622 High Copper and/or High Lead Feeds

Although concentrates containing copper and lead as high as 6 and 15 wt% of the concentrate respectively were tested in the roasting section and small control leaches performed on the resultant calcines, no mini-pilot scale hydrometallurgical testing was performed on these calcines due to time restrictions. Since copper levels in the feeds tested in the hydrometallurgical pilot were approximately 0.6 wt% and the copper level in the CANMET bulk concentrate is 0.7 wt%, copper behaviour experienced in the mini-pilot can be extrapolated to the CANMET bulk. The highest lead content of any material tested in the hydrometallurgical mini-pilot was 3.5 wt% as compared to a projected 10 wt% for the CANMET bulk. Roaster trials on feed material containing 3 to 15 wt% lead showed complete lead sulphation. Accordingly, since lead sulphate is inert in the leaching section, reporting entirely to the hot acid leach residue, and since it possesses excellent filtration characteristics, mini-pilot results on low lead calcines should be applicable to high lead calcines.

Bulk concentrates containing copper as high as 6 wt% could be handled in the hydrometallurgical pilot providing >97.5% recovery with only minor alteration. For copper levels between 1 and 3%, the copper SX would probably require three extraction stages. For copper levels between 3 and 6%, the copper SX unit would have three stages and an additional leach (50°C, pH 1.5 to 2.5) would be performed on the neutral leach residue before the hot acid leach. This leach would extract greater than 50% (experimental result) of the copper for direct electrowinning after partial neutralization for iron and arsenic removal. The iron-arsenic thickener sludge would be recycled to the hot acid leach. Alternatively, all the copper could be removed in the neutral leach (after hot acid leach recycle to the roaster) without resorting to an intermediate leach. The SX unit, operating on three stages, would effectively recover all the copper from the calcine. The resulting high acidity (30 gpl H_2SO_4) and low zinc (less than 130 gpl due to copper loading in the leach) in the solution from SX would necessitate neutralization with a high zinc neutralization agent such as a dead roast calcine instead of limestone.

132 Considerations for Full Plant Scale-up

.1 Leaching

.11 Neutral Leach Settling and Filtration

Thermal balance calculations on a fully continuous operation were necessary to compute the ambient temperature in a full plant neutral leach which differs only marginally from the ambient temperatures in the interlocked runs. A continuous leach usually requires longer retention time due to higher zinc backgrounds than batch leaches, however, tests at constant pH with varying zinc background showed that the 1.5 hour retention time was more than adequate for a continuous full-scale operation.

Data on filter sizing and residue washing were obtained from major equipment suppliers. The thickener size developed is conservative since it was based on unflocculated pulp tests (see section A131.612, p A1-151 and Figure A1-46, P A1-153). Qualitative tests indicate that polymeric flocculants will substantially increase settling rates. The extrapolations were based on thickening and filtration but it may be possible to use countercurrent washing while retaining thermal and volume balances in the leach-roast recycle if less than 10% ferrite in the calcine and greater than 38% solids in the thickener underflow can be achieved (note the effect of Cu on ferrite formation, section A122.6, p A1-82).

.2 Hot Acid Leaching and Filtration

Sufficient data were obtained from the preliminary and interlocked runs for direct extrapolation to full plant scale. Filter manufacturers supplied data for sizing and residue washing. Plant scale-up and economic analyses were based on the use of filter presses equipped for back washing since all experimental data is applicable to this equipment. Several manufacturers, however, were optimistic about the possibility of filtering and washing the hot acid leach pulp on a continuous vacuum filter, particularly a moving belt device. A low vacuum would have to be used to prevent flashing of the filtrate and subsequent crystallization. High efficiency countercurrent washing is possible on such machines.

.3 Solvent Extraction and Copper Electrowinning

It is expected that the more efficient settlers used in a plant operation will result in less neutral leach filtrate and organic carryover to copper electrowinning resulting in good quality copper cathodes. The zinc solution, stripped of copper, should also contain less organic carryover. In a plant scale continuous operation, the retention time between the neutral leach and the SX unit will be decreased from hours to minutes resulting in less crud formation in the SX unit due to iron hydroxide formation. Fully continuous roasting and leaching will result in constant copper levels in solution, better control of electrowinning parameters, and good copper deposits.

Scale-up sizing and costing data were obtained from equipment manufactures and by extrapolation from several different operating plants. The plant size was based on three minute retention in mixing with three stages of extraction and two stages of stripping as in plants currently handling heap leach solutions. It should be possible to reduce the retention time and/or the number of extraction stages due to improved kinetics at the higher solution temperature in a Sulphation Roast plant. (Note also that mini-pilot trials using one minute retention time per mixer removed copper to satisfactory limits).

.4 Iron Oxidation, Neutralization, Flocculation and Settling

The manganese dioxide-lead sludge from the cellhouse would be the primary oxidant employed in this operation on a fully continuous plant scale. The input to the system of fresh manganese dioxide would be minimal. Better control of neutralization should also be achieved as was experienced in spent electrolyte neutralization in the interlocked runs in going from batch to continuous.

.5 Zinc Dust Purification, Electrowinning and Cadmium Plant

Plants have been operating in continuous mode for nearly 10 years now using these procedures on similar solutions. Accordingly, scale-up data was based on design, thermal and mass balance data from existing electrolytic zinc plants and operational parameters determined in the preliminary and interlocked trials.

.6 Spent Electrolyte Neutralization

Since the continuous unit operated in the interlocked runs was based on units presently in use throughout the fertilizer industry, scale-up design and costing information was obtained from fertilizer industry representatives and major equipment suppliers. Operational parameters and results were projected directly from the test results.

.7 Extractions and Recoveries

The mass balances for leaching and spent electrolyte neutralization calculated for the full scale plants were based on extraction and metal loss data obtained in the interlocked runs. Losses in purification cakes were extrapolated from plant operating values from existing electrolytic zinc plant operations with minor modifications for the differences in cementate composition and tonnage. Zinc losses in the stripped spent electrolyte bleed were based on equilibrium electrolyte concentrations of <15 gpl magnesium

TABLE A1-43
METAL RECOVERIES FROM FEEDS IN THE SULPHATION ROAST AND INTEGRATED SULPHATION ROAST-DEAD ROAST PLANTS

Plant	Feed	Recovery (%)					
		Zinc as slab	Copper as cathode and cake	Lead in hot acid leach residue	Ag in hot acid leach residue	Cadmium as as cathode or bricket	Cobalt in copper cake
Sulphation Roast	1. Bulk Concentrate	96.0	97.5	99.5	99.5	89.6*	99.5
	2. Zinc Tailings Refloat	96.0	97.5	99.5	99.5	89.6*	99.5
Integrated Sulphation Roast-Dead Roast	Tailings Refloat & Zinc Concentrate	97.0	97.5	99.5	99.5	60.6*	99.5

* dependent on mass of copper cake

and <10 gpl manganese. The following assumptions based on the interlocked tests and discussions with equipment manufacturers were used in calculating mass balances:

1. 0.5% of the calcine lost to the wet scrubber.
2. 98.5% of the zinc and 98% of the copper in the calcine is extracted overall.
3. 0.6% of the zinc in the feed is lost as zinc sulphate in the hot acid leach residue due to incomplete washing.
4. 100% of the zinc sulphate and oxysulphate in the calcine is extracted in the neutral leach.
5. 95% of the zinc ferrite in the calcine is extracted in the strong acid leach.

Overall metal recoveries, based on the above discussions and used in the mass balance calculations for the full scale plants are given in Table A1-43.

A1.4 LEAD/SILVER RECOVERY

Essentially all of the lead and silver in the roaster feed remains as acid leach insoluble compounds in the strong acid leach residue and can be recovered by a simple brine leach. Results from the St. Joseph Lead pilot plant study showed that 90-98% of the lead and 80-85% of the silver are recoverable as a lead oxide product from a mixed $NaCl-CaCl_2$ brine leach (per. comm.-St. Joseph Lead Company reports). $CaCl_2$ is added to remove sulphate as gypsum.

Lead is present in the residue as lead sulphate which is readily soluble in NaCl brine as shown in Figure A1-55. Silver is also soluble in a brine leach (per. comm. - St. Joseph Lead Company reports). However, whereas lead is present as lead sulphate in the roaster calcine and is carried through the acid leaching stages, silver may follow a more complex path. The silver content of the calcine is small and hence difficult to locate as discrete particles for microprobe analyses. However, there is an indication (from a single silver containing grain located with the microprobe on a sample of zinc tailings refloat calcine, and from grains from the previous bench scale work on calcines from more silver-rich copper concentrates) that silver in roaster calcines may, in part, be present as sulphides; i.e. Ag_2S.

With acid leaching of the calcines, especially in the hot acid leach stage where there is abundant ferric iron to act as a leaching agent, the silver sulphides are solubilized given sufficient retention time, and since there is sufficient chloride ion available in solution, the silver is precipitated as silver chloride and remains in the residue. This was evidenced by analyses of neutral and strong acid leach filtrates, in which silver was not detected. Silver chloride is soluble in chloride solution through complex formation.

The proposed flowsheet for the brine leach lead-silver recovery plant is shown in Drawing No. 6. The hot acid leach residue after thorough washing is subjected to a three stage leach with the lixiviant being a 27% brine ($NaCl + CaCl_2$) solution. The slurry density is determined by $PbCl_2$ solubility in the brine solution at room temperature. As the solubility of lead chloride in the brine solution is only 2-3% at room temperature (Figure A1-55), the volume of brine solution to hot acid leach residue must be kept at relatively high levels. For example, for the case of the bulk concentrate, approximately 3292.5 MTPD of brine are required to leach 543.77 MTPD (27.6% moisture) of hot acid leach residue. The slurry is filtered with the resultant filtrate going to the lime precipitation

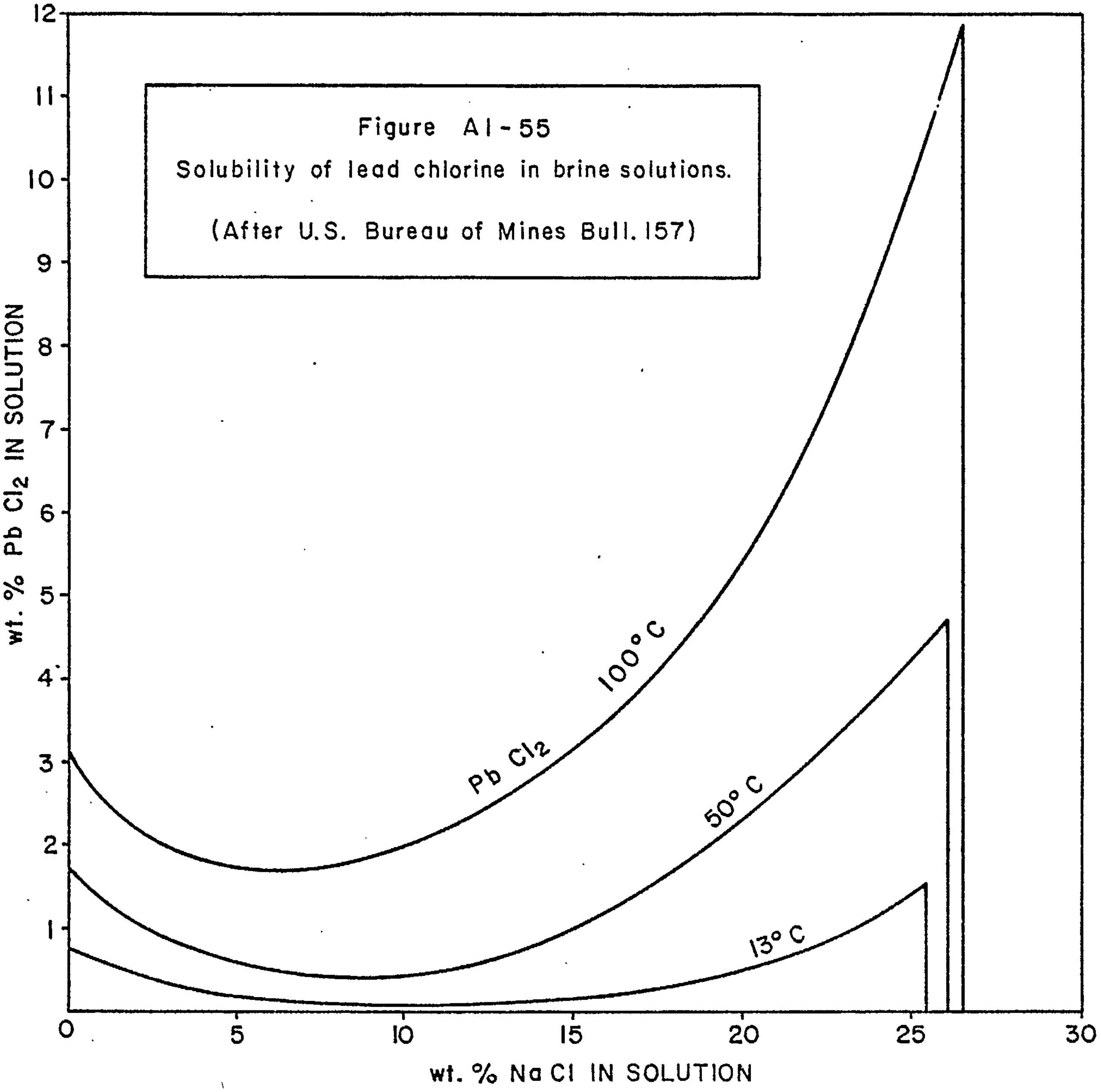

Figure A1-55
Solubility of lead chlorine in brine solutions.

(After U.S. Bureau of Mines Bull. 157)

tanks and the solids repulped, washed and filtered to recover entrapped lead chloride and silver chlorides; this wash filtrate also goes to the lime precipitation tanks. The solids, which contain all of the hematite and the majority of sulphate fixed as gypsum, exit to disposal ponds.

In the lime precipitation tanks, the filtrate containing the bulk of the lead and silver is further treated with lime (CaO) to precipitate lead and silver as a lead hydrate product. The pulp is filtered and washed and the recovered solution returned to the circuit. A small bleed is required to effect a water balance in the circuit. This bleed can be evaporated to recover NaCl and $CaCl_2$ or discarded depending on the relative cost of evaporation compared with the price of the makeup reagents $CaCl_2$ and NaCl.

The precipitate is calcined in an oil-fired rotary kiln at temperatures above 300°C to form a lead oxide product containing silver and traces of minor amounts of other metals.

Calculated mass balances (Appendix A2) for the proposed brine leach process, based on the St. Joseph Lead pilot plant results, applied to each of the bulk concentrate, tailings refloat and integrated SR/DR options yielded lead oxide products with the characteristics shown in Table A1-44.

TABLE A1-44
LEAD OXIDE PRODUCT CHARACTERISTICS

	Bulk Concentrate	Tailings Refloat	SR/DR
Lead Oxide Product (STPD)	104.27	41.46	24.54
%Pb	82.87	78.71	71.31
%Zn	1.80	1.67	7.50
%Cu	0.13	0.11	0.25
oz/T Ag	66.31	88.00	81.29
Lead Recovery (%)	93.85	91.79	92.48
Silver Recovery (%)	85.00	85.00	85.00

These lead oxide feeds are a highly acceptable product for the Brunswick Mining and Smelting lead smelter at Belledune. The low sulphur content makes this type of product particularly suitable for addition to the lead smelter sinter machine as a means of balancing sulphur and thermal requirements.

A1.5 REFERENCES

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APPENDIX A2

MASS AND THERMAL BALANCES FOR THE BULK CONCENTRATE, TAILINGS REFLOAT PRODUCT AND INTEGRATED SULPHATION ROAST-DEAD ROAST PLANTS

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TABLE A2-1

ROASTING SECTION: BULK CONCENTRATE MATERIAL BALANCE

	SOLIDS (TPD)												GASES (TPD)						LIQUIDS (TPD)			
	S1	S2	S3	S4	S5	S6	S7	S8	S9	S10	S11	S12	G1	G2	G3	G4	G5	G6	L1	L2	L3	L4
FEED (SOLIDS)																						
(Zn,Fe)S	456.16	456.16	-	456.16	-	456.16	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
PbS	96.93	96.93	-	96.93	-	96.93	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$CuFeS_2$	16.97	16.97	-	16.97	-	16.97	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
FeS_2	240.65	240.65	-	240.65	-	240.65	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SiO_2	5.82	5.82	-	5.82	-	5.82	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$CaCO_3$	2.96	2.96	-	2.96	-	2.96	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$MgSiO_4$	1.77	1.77	-	1.77	-	1.77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ag	0.25	0.25	-	0.25	-	0.25	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CoS	0.23	0.23	-	0.23	-	0.23	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CdS	0.76	0.76	-	0.76	-	0.76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MnS	0.50	0.50	-	0.50	-	0.50	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
AsFeS	8.78	8.78	-	8.78	-	8.78	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Al_2O_3	1.61	1.61	-	1.61	-	1.61	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Others	9.84	9.84	-	9.84	-	9.84	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H_2O (as moisture)	58.58	16.88	-	16.88	-	16.88	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
DROSS (SOLIDS)																						
Zn	-	-	5.70	5.70	-	5.70	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CALCINE (SOLIDS)																						
$ZnSO_4$	-	-	-	-	48.99	48.99	417.77	9.21	36.60	515.77	515.77	-	-	464.19	51.58	41.26	2.58	-	-	-	2.58	-
$ZnO.2ZnSO_4$	-	-	-	-	13.59	13.59	115.83	2.26	9.71	143.00	143.00	-	-	128.70	14.30	11.44	0.72	-	-	-	0.72	-
$ZnO.Fe_2O_3$	-	-	-	-	11.58	11.58	98.71	2.40	9.14	121.87	121.87	-	-	109.68	12.19	9.75	0.61	-	-	-	0.61	-
(Zn,Fe)S (remnant)	-	-	-	-	1.70	1.70	14.45	0.26	1.34	17.84	17.84	-	-	16.06	1.78	1.43	0.09	-	-	-	0.09	-
$PbSO_4$	-	-	-	-	11.62	11.62	99.02	2.05	9.03	122.25	122.25	-	-	110.03	12.23	9.78	0.62	-	-	-	0.62	-
$CuSO_4$	-	-	-	-	1.60	1.60	13.61	0.34	1.26	16.80	16.80	-	-	15.12	1.68	1.34	0.08	-	-	-	0.08	-
Fe_2O_3	-	-	-	-	19.86	19.86	169.45	4.07	15.34	209.20	209.20	-	-	188.28	20.92	16.74	1.05	-	-	-	1.05	-
$MgSO_4$	-	-	-	-	1.24	1.24	10.59	0.26	0.98	13.08	13.08	-	-	11.77	1.31	1.05	0.07	-	-	-	0.07	-
Na_2SO_4	-	-	-	-	0.20	0.24	1.68	0.04	0.16	2.07	2.07	-	-	1.86	0.21	0.17	0.01	-	-	-	0.01	-
As_2O_3	-	-	-	-	0.58	0.58	4.93	0.12	0.46	6.09	6.09	3.28	-	5.48	0.61	0.49	0.03	-	-	-	0.03	-
$CaSO_4$	-	-	-	-	0.38	0.38	3.25	0.08	0.30	4.01	4.01	-	-	3.61	0.40	0.32	0.02	-	-	-	0.02	-
$CdSO_4$	-	-	-	-	0.10	0.10	0.87	0.02	0.08	1.08	1.08	-	-	0.97	0.11	0.09	0.03	-	-	-	0.01	-
$CoSO_4$	-	-	-	-	0.04	0.04	0.32	0.01	0.03	0.39	0.39	-	-	0.35	0.04	0.03	0.00	-	-	-	0.00	-
$MnSO_4$	-	-	-	-	0.32	0.32	2.66	0.07	0.25	3.29	3.29	-	-	2.96	0.33	0.26	0.02	-	-	-	0.02	-
SiO_2	-	-	-	-	0.69	0.69	5.88	0.15	0.54	7.26	7.26	-	-	6.53	0.73	0.58	0.04	-	-	-	0.04	-
Ag	-	-	-	-	0.03	0.03	0.20	0.01	0.02	0.25	0.25	-	-	0.23	0.03	0.02	0.00	-	-	-	0.00	-
Others	-	-	-	-	0.16	0.16	1.35	0.03	0.13	1.67	1.67	5.69	-	1.50	0.17	0.13	0.01	-	-	-	0.01	-
GASES																						
O_2	-	-	-	-	-	-	-	-	-	-	-	-	865.20	397.59	397.59	397.59	397.59	397.59	-	-	-	-
N_2	-	-	-	-	-	-	-	-	-	-	-	-	2849.32	2849.32	2849.32	2849.32	2849.32	2849.32	-	-	-	-
SO_2	-	-	-	-	-	-	-	-	-	-	-	-	-	402.02	402.02	402.02	402.02	355.60	-	-	-	-
H_2O (vapour)	-	-	-	-	-	-	-	-	-	-	-	-	-	455.37	455.37	455.37	455.37	97.59	-	-	-	-
As_2O_3(gaseous)	-	-	-	-	-	-	-	-	-	-	-	-	3.28	3.28	3.28	3.28	3.28	-	-	-	-	-
Hg,Cl	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
HAL SOLUTION & LIQUIDS																						
$FeSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	38.11	-	-	-
$Fe_2(SO_4)_3$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	140.19	-	-	-
$ZnSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	118.21	(-)	-	-
As (as AsO_4^{-3})	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	2.89	-	2.49	-
$CuSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	2.34	-	-	-
$MgSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	11.39	-	-	-
Na_2SO_4	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	4.14	-	-	-
$MnSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	9.01	-	-	-
H_2SO_4	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	19.71	(-)	69.80	-
H_2O	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	438.58	(-)	357.78	304.11
Total, TPD (dry) (solids)	843.96	843.96	5.70	849.66	112.68	962.34	960.59	21.35	85.39	1185.92	1185.92	8.97	3717.80	1067.32	118.62	94.88	5.96	-	345.99	-	72.29	-
Total, TPD (wet) (solids)	970.14	860.84	-	866.54	-	979.22	-	-	-	-	-	(-)	-	-	-	-	-	-	784.57	(-)	430.07	304.11
Total gases, (m^3/day) x 10^6	-	-	-	-	-	-	-	-	-	-	-	-	2.884	3.2431	3.2431	3.2431	3.2431	2.802	-	-	-	-

TABLE A2-2

LEACHING AND COPPER EXTRACTION CIRCUIT: BULK CONCENTRATE MATERIAL BALANCE

	SOLIDS (TPD)				LIQUIDS (TPD)														
	S1	S2	S3	S4	L1	L2	L3	L4	L5	L6	L7	L8	L9	L10	L11	L12	L13	L14	L15
$ZnSO_4$	515.77	-	4.27	-	25.00	211.46	-	-	-	-	24.70	-	118.24	1062.54	-	-	-	-	923.95
$Fe_2(SO_4)_3$	-	-	6.19	-	-	-	-	-	-	-	-	-	140.19	-	-	-	-	-	-
$FeSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	38.11	5.58	-	-	-	-	-
Total Fe															-	-	-	-	0.02
$CoSO_4$	0.39	-	-	-	-	-	-	-	-	-	-	-	-	0.45	-	-	-	-	0.39
$CdSO_4$	1.08	-	-	-	-	-	-	-	-	-	-	-	-	1.24	-	-	-	-	1.07
$CuSO_4$	16.80	2.33	0.30	-	-	-	-	-	-	-	-	-	2.33	14.51	220.63	-	0.37	-	0.35
$PbSO_4$	122.25	122.25	122.25	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$MnSO_4$	3.29	-	-	-	2.39	20.24	-	-	-	-	2.37	-	2.37	30.44	-	-	-	-	30.14
$MgSO_4$	13.08	-	-	-	11.53	101.33	-	-	-	-	11.39	-	11.39	144.84	-	-	-	-	126.08
Na_2SO_4	2.07	-	-	-	-	-	-	-	-	-	-	-	2.07	-	-	-	-	-	-
$CaSO_4$	4.01	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$CaSO_4 \cdot 2H_2O$	-	26.57	26.57	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$ZnO.2ZnSO_4$	143.00	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Fe_2O_3	209.20	209.20	209.20	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$ZnO.Fe_2O_3$	121.87	121.87	6.09	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SiO_2	7.26	7.26	7.26	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
As_2O_3	6.09	6.09	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MnO_2	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$Fe(OH)_3$	-	2.39	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
(Zn,Fe)S	17.84	17.84	3.99	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
As	-	-	.86	-	-	-	-	-	-	-	-	-	2.89	<1 mgpl	-	-	-	-	-
S°	-	-	4.01	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ag (kg/day)	253	253	253	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Gangue (e.g. Al_2O_3)	1.67	1.67	1.67	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H_2SO_4	-	-	-	-	34.71	-	-	-	-	-	34.41	-	19.71	-	468.81	-	-	-	-
H_2O	-	232.49	144.99	-	180.94	1662.56	539.39	-	232.49	-	178.84	144.99	438.58	2823.39	2675.25	-	-	-	-
Cathode Copper	-	-	-	5.57	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Percol 351 (kg/day)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Percol 156 (kg/day)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Lix 64N (kg/day)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Kerosene (lc/day)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MnO_2 Ore	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Milled Limestone	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Suspended Solids (gpl)	-	-	-	-	-	-	-	0.10	-	-	-	-	-	-	-	-	-	-	-
Specific Gravity (g/cc)	-	4.95	-	-	1.26	1.17	-	-	-	-	1.26	-	1.57	1.39	1.15	-	-	-	1.39
Pulp Density (% Sol.)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
pH	-	-	-	-	-	-	-	-	-	-	-	-	-	4.0-5.2	-	-	1.22	0.82	4.2
Organic Sp. Grav. @ 15°C	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	0.80 g/cc	-	-	-
Temperature (°C)	-	-	-	-	40	30	25	-	25	-	40	-	95	40	40	40	40	60	60
VOLUME (m^3/day)	-	-	-	-	202.34	1712.21	-	2124.20	-	805.84	200.00	-	492.79	2930.04	2930.04	2930.04	2930.04	2930.04	2547.86
DRY WEIGHT	-	517.72	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
TOTAL WEIGHT	1185.67	750.21	543.03	14.14	254.57	1995.59	539.39	-	232.49	-	251.70	144.99	775.88	4082.98	3364.69	-	-	-	3537.15

TABLE A2-2 (Cont'd)

LEACHING AND COPPER EXTRACTION CIRCUIT BULK CONCENTRATE MATERIAL BALANCE

	PULPS (TPD)						GASES (TPD)			REAGENTS (TPD)					
	P1	P2	P3	P4	P5	P6	G1	G2	G3	R1	R2	R3	R4	R5	R6
$ZnSO_4$	138.59	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$Fe_2(SO_4)_3$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$FeSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Total Fe	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$CoSO_4$	0.05	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$CdSO_4$	0.17	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$CuSO_4$	0.05	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$PbSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$MnSO_4$	4.51	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$MgSO_4$	18.91	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Na_2SO_4	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$CaSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$CaSO_4 2H_2O$	21.50	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$ZnO.2ZnSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Fe_2O_3	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$ZnO.Fe_2O_3$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SiO_2	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
As_2O_3	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MnO_2	0.51	-	-	-	1.44	-	-	-	-	-	-	-	-	-	-
$Fe(OH)_3$	3.92	-	-	-	-	-	-	-	-	-	-	-	-	-	-
(Zn,Fe)S	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
As	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
S^o	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ag (kg/day)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Gangue (e.g. Al_2O_3)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H_2SO_4	-	-	-	-	-	-	-	-	-	-	177.12	-	-	-	-
H_2O	368.27	-	-	-	-	-	2.00	8.50	10.90	80.00	-	-	-	-	1.25
Cathode Copper	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Percol 351 (kg/day)	-	-	-	-	-	-	-	-	-	87.89	-	-	-	-	-
Percol 156 (kg/day)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	1.25
Lix 64N (kg/day)	-	-	-	-	-	-	-	-	-	-	-	17.00	-	-	-
Kerosene (lt/day)	-	-	-	-	-	-	-	-	-	-	-	280.00	-	-	-
MnO_2 Ore	-	-	-	-	-	-	-	-	-	-	-	-	0.09	-	-
Milled Limestone	-	-	-	-	-	-	-	-	-	-	-	-	-	12.50	-
Suspended Solids (gpl)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Specific Gravity (g/cc)	1.46	1.51	-	-	-	-	-	-	-	-	-	-	-	-	-
Pulp Density (% Sol.)	5.5	11.25	31.6	29.8	-	-	0.7	-	-	-	-	-	-	-	-
pH	4.2	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Organic Sp. Grav. @ 15°C	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Temperature (°C)	60	40	40	95	-	40	-	-	-	-	-	-	-	-	-
VOLUME (m^3/day)	382.18	3034.57	910.37	713.31	-	2930.04	-	-	-	-	-	-	-	-	-
DRY WEIGHT	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
TOTAL WEIGHT	556.48	-	-	-	-	-	-	-	-	-	-	-	-	-	-

TABLE A2-3

PURIFICATION AND ELECTROWINNING CIRCUIT BULK CONCENTRATE MATERIAL BALANCE

	SOLIDS (TPD)			LIQUIDS (TPD)															
	S1	S2	S3	L1	L2	L3	L4	L5	L6	L7	L8	L9	L10	L11	L12	L13	L14	L15	L16
Zn	11.60	12.09	–	374.07	–	17.73	–	–	–	–	–	–	–	–	–	403.95	–	117.73	1.00
Co	0.15	< 0.15	–	0.15	–	–	–	–	–	–	–	<0.15	–	–	–	< 0.2mgpl	–	–	–
Cd	0.58	–	–	0.58	–	–	–	–	–	–	–	<1.0mgpl	–	–	–	< 1.0mgpl	–	–	–
Cu	0.14	–	–	0.14	–	–	–	–	–	–	–	<0.02mgpl	–	–	–	< 0.2mgpl	–	–	–
Mn	–	–	–	10.96	–	–	–	–	–	–	–	–	–	–	–	11.72	–	10.12	0.43
Mg	–	–	–	25.47	–	–	–	–	–	–	–	–	–	–	–	<27.49	–	27.08	1.15
Sb	<0.02	<0.02	–	–	–	–	–	–	–	–	–	–	–	–	–	0.02mgpl	–	–	–
Zn°	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Co°	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Cd°	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Cu°	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Zn as dross	–	–	5.70	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Zn as dust	–	–	20.32	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Zn as slab	–	–	259.20	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Zn as cathode	–	–	285.22	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
H_2SO_4	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	403.91	24.02
H_2O	–	–	–	2455.12	–	96.31	–	–	–	–	–	–	–	–	–	2594.98	39.20	2105.52	89.42
MnO_2	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Sb_2O_3 (kg/day)	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Limestone	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Percol 156 (kg/day)	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Animal glue	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Na silicate	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Sr carbonate	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Specific gravity (g/cc)	–	–	–	1.39	–	1.40	–	–	–	–	–	–	–	–	–	1.40	–	1.26	–
Pulp density (% Sol.)	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
pH	–	–	–	4.2	–	–	–	–	–	–	–	–	–	–	–	5.00	–	–	–
Temperature (°C)	–	–	–	60	–	95	–	–	–	–	65	95	–	–	–	–	–	40	–
VOLUME (m^3/day)	–	–	–	2547.86	<1.00	100.95	0.00	0.00	0.00	0.00	–	2647.82	≈64.99	<1.00	5.00	2693.01	–	2354.64	100.00
WEIGHT OF SOLIDS	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
TOTAL WEIGHT	12.49	12.26	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–

TABLE A2- 3 (Cont'd)

PURIFICATION AND ELECTROWINNING CIRCUIT: BULK CONCENTRATE MATERIAL BALANCE

	PULPS (TPD)								GASES (TPD)				REAGENTS (TPD)					
	P1	P2	P3	P4	P5	P6	P7	P8	G1	G2	G3	G4	R1	R2	R3	R4	R5	R6
Zn	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Co	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Cd	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Cu	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Mn	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Mg	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sb	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Zn°	11.60	11.60	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Co°	0.15	0.15	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Cd°	0.58	0.58	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Cu°	0.14	0.14	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Zn as dross	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Zn as dust	-	-	-	-	-	-	-	-	-	-	-	-	0.00	-	17.52	≃2.0	-	-
Zn as slab	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Zn as cathode	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H_2SO_4	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H_2O	-	-	-	-	-	-	-	-	8.70	22.30	86.20	353.00	-	4.50	<1.50	0.00	-	-
MnO_2	-	-	-	-	-	-	-	1.44	-	-	-	-	-	-	-	-	-	-
Sb_2O_3 (kg/day)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	<27.00	0.00	-	-
Limestone	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	24.50
Percol 156 (kg/day)	-	-	-	-	-	-	-	-	-	-	-	-	-	4.50	-	-	-	-
Animal glue	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	≃0.05	-
Na silicate	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	≃0.13	-
Sr carbonate	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	N.A.	-
Specific gravity (g/cc)	1.40	1.56	-	-	-	-	1.55	-	-	-	-	-	-	-	-	-	-	-
Pulp density (% Sol.)	0.3	10.0	0.8	19.2	0.3	0.3	9.9	-	-	-	-	-	-	-	-	-	-	-
pH	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Temperature (°C)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
VOLUME (m^3/day)	2723.32	80.00	2752.01	60.00	2692.01	2693.01	80.00	-	-	-	-	-	-	-	-	-	-	-
WEIGHT OF SOLIDS	12.47	12.47	30.26	20.00	10.26	12.25	12.25	-	-	-	-	-	-	-	-	-	-	-
TOTAL WEIGHT	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-

TABLE A2-4

CADMIUM PLANT
BULK CONCENTRATE MATERIAL BALANCE

	SOLIDS (TPD)					LIQUIDS (TPD)			PULPS (TPD)			REAGENTS (TPD)				
	S1	S2	S3	S4	L1	L2	L3	L4	P1	P2	P3	R1	R2	R3	R4	R5
Zn	11.60	0.10	-	-	5.38	17.33	-	17.73	-	-	-	-	-	-	-	-
Cu	0.14	<0.16	-	-	-	< 1mgpl	-	<1mgpl	-	-	-	-	-	-	-	-
Cd	0.58	0.05	-	-	-	0.53	-	<20mgpl	-	-	-	-	-	-	-	-
Co	0.15	0.15	-	-	-	<50mgpl	-	<50mgpl	-	-	-	-	-	-	-	-
Sb	<0.02	<0.02	-	-	-	-	-	-	-	-	-	-	-	-	-	-
As	-	<0.04	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Product Cd	-	-	-	0.53	-	-	-	-	-	-	-	-	-	-	-	-
H_2SO_4	-	-	-	-	18.46	-	-	-	-	-	-	<1.00	-	-	-	-
H_2O	-	-	-	-	96.31	96.31	-	96.31	-	-	-	-	-	-	-	-
Zn dust	-	-	-	-	-	-	-	-	-	-	-	-	0.45	-	-	0.35
$CuSO_4 \cdot 5H_2O$	-	-	-	-	-	-	-	-	-	-	-	-	-	0.05	-	-
As_2O_3	-	-	-	-	-	-	-	-	-	-	-	-	-	0.05	-	-
$SrCO_3$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	N.A.	-
Specific gravity (g/cc)	-	-	-	-	-	1.40	-	-	-	-	-	-	-	-	-	-
Pulp density (% Sol.)	-	-	-	-	-	-	-	-	0.4	N.A.	0.4	-	-	-	-	-
VOLUME (m^3/day)	-	-	-	-	107.64	99.95	<1.00	100.95	99.95	99.95	100.95	-	-	-	-	-
WEIGHT OF SOLIDS	12.49	<0.52	N.A.	-	-	-	-	-	-	-	-	-	-	-	-	-

TABLE A2-5

LEAD/SILVER RECOVERY PLANT: BULK CONCENTRATE MATERIAL BALANCE

	SOLIDS (TPD)							LIQUIDS (TPD)							
	S1	S2	S3	S4	S5	S6	S7	L1	L2	L3	L4	L5	L6	L7	L8
$ZnFe_2O_4$	6.09	6.09	6.09	-	-	-	-	-	-	-	-	-	-	-	-
(Zn,Fe)S	3.99	3.99	3.99	-	-	-	-	-	-	-	-	-	-	-	-
Fe_2O_3	209.20	209.20	209.20	-	-	-	-	-	-	-	-	-	-	-	-
$PbSO_4$	122.25	6.11	6.11	-	-	-	-	-	-	-	-	-	-	-	-
$CaSO_4.2H_2O$	26.57	98.73	98.73	-	6.56	-	-	-	-	-	-	-	-	-	-
SiO_2	7.26	7.26	7.26	-	-	-	-	-	-	-	-	-	-	-	-
As	0.86	0.86	0.86	-	-	-	-	-	-	-	-	-	-	-	-
S°	4.55	4.55	4.55	-	-	-	-	-	-	-	-	-	-	-	-
Ag (kg/day)	253	≃ 21	≃ 21	-	≃240	≃240	-	≃240	<10	≃240	-	-	-	-	-
Gangue	1.67	1.67	1.67	-	-	-	-	-	-	-	-	-	-	-	-
$ZnSO_4$	4.27	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$CuSO_4$	0.30	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$Fe_2(SO_4)_3$	6.19	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$PbO.H_2O$	-	-	-	-	91.25	-	-	-	-	-	-	-	-	-	-
$ZnO.H_2O$	-	-	-	-	2.60	-	-	-	-	-	-	-	-	-	-
$CuO.H_2O$	-	-	-	-	0.18	-	-	-	-	-	-	-	-	-	-
$Fe(OH)_3$	-	-	-	-	3.27	-	-	-	-	-	-	-	-	-	-
PbO	-	-	-	-	-	84.44	-	-	-	-	-	-	-	-	-
ZnO	-	-	-	-	-	2.13	-	-	-	-	-	-	-	-	-
CuO	-	-	-	-	-	0.15	-	-	-	-	-	-	-	-	-
$CaSO_4$	-	-	-	-	-	5.19	-	-	-	-	-	-	-	-	-
Fe_2O_3	-	-	-	-	-	2.44	-	-	-	-	-	-	-	-	-
CaO	-	-	-	25.35	-	-	-	-	-	-	-	-	-	-	-
H_2O(Free)	145.19	80.55	60.67	-	9.05	-	-	2418.36	101.11	2519.47	132.82	2368.68	73.48	81.23	2363.68
H_2O(Non Free)	5.13	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$PbCl_2$	-	3.44	1.29	-	-	-	-	103.07	2.15	105.22	-	-	-	-	-
NaCl	-	23.88	20.80	-	-	-	33.20	715.33	34.67	750.00	39.92	710.51	4.02	31.59	743.71
$CaCl_2$	-	4.50	1.69	-	-	-	12.96	135.09	2.81	137.90	9.78	173.16	0.98	-	186.12
$ZnCl_2$	-	0.12	0.04	-	-	-	--	3.49	0.07	3.57	-	-	-	-	-
$FeCl_3$	-	0.16	0.06	-	-	-	-	4.86	0.10	4.96	-	-	-	-	-
$CuCl_2$	-	0.01	<0.01	-	-	-	-	0.24	<0.01	<0.25	-	-	-	-	-
Na_2SO_4	-	0.18	0.07	-	-	-	-	5.29	0.11	5.40	-	-	-	-	-
TOTAL TONS/DAY	543.77	451.32	423.11	25.35	113.15	94.59	46.16	3385.73	141.04	3527.02	182.52	3252.35	78.48	112.82	3293.51
MOISTURE (%)	27.6	25.0	20.0	0.0	8.0	-	-	-	-	-	-	-	-	-	-

PRODUCT = 94.59 MTPD; 82.87% Pb, 66.31 oz/ sh.T.

TABLE A2-6

SPENT ELECTROLYTE NEUTRALIZATION
BULK CONCENTRATE MATERIAL BALANCE

	SOLIDS (TPD)	LIQUIDS (TPD)			REAGENTS (TPD)
	S1	L1	L2	L3	R1
Zn	3.21	88.88	-	85.61	-
Mg	-	19.69	-	20.47	0.78
Mn	-	7.36	-	7.36	-
$CaSO_4.2H_2O$	534.63	-	-	-	-
H_2SO_4	-	304.72	-	-	-
H_2O	389.47	1589.47	519.47	1662.56	-
Limestone	-	-	-	-	310.97
pH	-	-	-	≈5.0	-
Temperature(°C)	-	40	-	30	-
VOLUME(m^3/day)	-	1776.42	-	1712.21	-
TOTAL WEIGHT	927.31	-	-	-	-

TABLE A2-7

BULK CONCENTRATE SULPHUR BALANCE IN ROASTING SECTION

	S (TPD)	SO_2 (TPD)	H_2SO_4 (TPD)
Feed to roasters (32.0% Zn)	268.67	-	-
Calcines: (11.67% S)			
from feed	114.87	-	-
from HAL solution	23.48	-	-
Off gases from roasters:			
from feed	153.80	307.20(1)	-
from HAL solution	48.21	96.40(1)	-
Off gases after scrubber	173.55	346.65	-
Bleed from scrubber(4)	28.46	58.31(2)	71.42
H_2SO_4 production from SO_2(3)	168.34	336.25	530.76
Losses	5.21	10.40	-

	Gases Composition (%)	
	After Roaster	After Scrubber
O_2	8.57	9.93
N_2	70.30	81.30
SO_2	4.35	4.42
H_2O	16.80	4.33

(1) $SO_2 + SO_3$

(2) As SO_3

(3) 97% efficiency in acid plant

(4) Bleed contains 199.6 gpl H_2SO_4

TABLE A2-8

BULK CONCENTRATE THERMAL BALANCE

	Heat (Kcal/day) x 10^6	
	generated or heat content	consumed
1.0 Roasting Section		
1.1 Feed Dryer		
Preheat feed (dry base)	-	+ 13.59
Vaporize water (8 to 2.5%)	-	+ 35.70
Net balance	-	+ 49.29
1.2 Roaster		
Preheat fluidizing air	-	+625.74
Preheat feed (dry)	-	+ 76.08
Vaporize moisture of feed	-	+ 15.31
Reheat recirculated dust	-	+ 16.12
Preheat solids in HAL solution:		
$FeSO_4$	-	+ 3.28
$Fe_2(SO_4)_3$	-	+ 13.81
$ZnSO_4$	-	+ 18.38
H_2SO_4	-	+ 2.42
Decompose HAL solution:		
$FeSO_4$	-	+ 14.88
$Fe_2(SO_4)_3$	-	+ 76.38
H_2SO_4	-	+ 9.99
Vaporize water from HAL solution	-	+397.79
Heats of reaction:		
(Zn,Fe)S	-872.18	-
FeS_2	-395.63	-
$CuFeS_2$	- 28.09	-
PbS	- 63.96	-
Heat losses through shell	-	+ 67.99
Spent to balance heat surplus (Equiv. 22.15 TPD spent)	-	+ 21.79
Net Balance	-1359.86	+1359.86

TABLE A2-8

BULK CONCENTRATE THERMAL BALANCE (cont'd)

	Heat (Kcal/day) x 10^6 generated or heat content	Heat (Kcal/day) x 10^6 consumed
1.3 Waste heat boilers		
Gases to boilers (650°C)	-675.50	-
Gases off boilers (350°C)	-322.70	-
Steam generation (350°C) (Equiv. to 477.4 TPD steam)	-	+352.80
Net balance (at exit)	-322.70	+352.80
1.4 Electrostatic Precipitator		
Gases to precipitator (350°C)	-322.70	-
Gases off precipitator (300°C)	-272.30	-
Heat losses through shell	-	+ 50.40
Net balance (at exit)	-272.30	+ 50.40
1.5 Water scrubber		
Gases to scrubber (300°C)	-272.30	-
Gases off scrubber (35°C)	- 10.28	-
Gases cooling (300 to 35°C)	-	+256.08
Water condensation (to 35°C)	-	+262.02
Net balance (at exit)	- 10.28	(+518.10)*
1.6 Overall balance		
Net heat recovered or required:		
Feed dryer	-	+ 49.29
Waste heat boilers	-352.80	-
Scrubber	-	(+518.10)*

(*) Heat dissipated in cooling towers

TABLE A2-8

BULK CONCENTRATE-THERMAL BALANCE (cont'd)

	Heat (Kcal/day) x 10^6	
	generated or heat content	consumed
2.0 Hydrometallurgical Section*		
2.1 Neutral Leach		
Preheat calcine	-	+ 5.71
Preheat water	-	+ 34.98
Preheat neutral solution	-	+ 60.73
Preheat spent electrolyte	-	+ 4.98
Preheat neutral leach filtrate recycle	-	+ 2.27
Neutral solution (30°C)	- 8.53	-
Spent electrolyte (40°C)	- 2.95	-
Neutral leach filtrate recycle (60°C)	- 13.80	-
Heats of dissolution and reaction	- 84.54	-
Heat of evaporation	-	+ 1.15
Net balance	- 109.82	+ 109.82
2.2 Solution Cooling for Solvent Extraction (65.7-40°C) **	- 83.73	-
Net heat required or recovered	0	0
2.3 Hot Acid Leach		
Preheat neutral leach residue	-	+ 21.64
Preheat acid	-	+ 4.20
Preheat spent electrolyte	-	+ 11.28
Spent electrolyte (40°C)	- 3.07	-
Heat of reaction of sphalerite	- 1.45	-
Heat of reaction of zinc ferrite	- 33.08	-
Heat of solution of acid	- 30.66	-
Heat of evaporation	-	+ 4.68
Net balance **	- 68.26	+ 41.80
Net heat required or recovered	0	0

TABLE A2-8

BULK CONCENTRATE-THERMAL BALANCE (cont'd)

	Heat (Kcal/day) x 10^6 generated or heat content	consumed
2. 4 Heat exchanger between solvent extraction and iron oxidation	-	+ 59.48
2. 5 Heat of evaporation in iron oxidation	-	+ 6.44
2. 6 Heat exchanger between iron oxidation and 1st purification	-	+ 12.93
2. 7 Heat of evaporation in 1st purification	-	+ 5.06
2. 8 Heat exchanger between 1st and 2nd purification	-	+ 82.95
2. 9 Heat of evaporation in 2nd purification	-	+ 12.29
2.10 Heat of evaporation in cooling towers	-	+ 213.57
Net balance	0	+ 392.72
Net heat required or recovered	0	+ 392.72

* conductive and convective heat losses not included in these balances. 20 percent overall heat losses for hydrometallurgical section assumed in economic analyses for fuel requirements.

** excess heat dissipated by conduction and convection

TABLE A2-9

ROASTING SECTION: TAILINGS REFLOAT PRODUCT MATERIAL BALANCE

	SOLIDS (TPD)												GASES (TPD)						LIQUIDS (TPD)			
	S1	S2	S3	S4	S5	S6	S7	S8	S9	S10	S11	S12	G1	G2	G3	G4	G5	G6	L1	L2	L3	L4
FEED (SOLIDS)																						
(Zn,Fe)S	456.23	456.23	-	456.23	-	456.23	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
PbS	37.43	37.43	-	37.43	-	37.43	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$CuFeS_2$	15.60	15.60	-	15.60	-	15.60	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
FeS_2	352.16	352.16	-	352.16	-	352.16	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SiO_2	6.21	6.21	-	6.21	-	6.21	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$CaCO_3$	3.15	3.15	-	3.15	-	3.15	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$MgSiO_4$	1.89	1.89	-	1.89	-	1.89	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ag	0.14	0.14	-	0.14	-	0.14	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CoS	0.28	0.28	-	0.28	-	0.28	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CdS	0.25	0.25	-	0.25	-	0.25	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MnS	0.54	0.54	-	0.54	-	0.54	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
AsFeS	9.36	9.36	-	9.36	-	9.36	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Al_2O_3	1.72	1.72	-	1.71	-	1.72	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Others	14.50	14.50	-	14.50	-	14.50	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H_2O (as moisture)	63.02	22.51	-	22.51	-	22.51	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
DROSS (SOLID)																						
Zn	-	-	5.70	5.70	-	5.70	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CALCINES (SOLID)																						
$ZnSO_4$	-	-	-	-	46.51	46.51	418.59	9.30	37.21	565.09	565.09	-	-	467.94	49.35	40.05	2.84	-	-	-	-	-
$ZnO.2ZnSO_4$	-	-	-	-	8.73	8.73	78.56	1.75	6.98	106.06	106.06	-	-	87.83	9.26	7.52	0.53	-	-	-	-	-
$ZnO.Fe_2O_3$	-	-	-	-	10.66	10.66	95.93	2.13	8.53	129.50	129.50	-	-	107.24	11.31	9.18	0.65	-	-	-	-	-
(Zn,Fe)S (remnant)	-	-	-	-	0.87	0.87	7.87	0.18	0.70	10.63	10.63	-	-	8.80	0.93	0.75	0.05	-	-	-	-	-
$PbSO_4$	-	-	-	-	3.88	3.88	34.96	0.78	4.69	47.20	47.20	-	-	39.09	4.12	3.35	0.24	-	-	-	-	-
$CuSO_4$	-	-	-	-	1.26	1.26	11.38	0.25	1.01	15.36	15.36	-	-	12.72	1.34	1.09	0.08	-	-	-	-	-
Fe_2O_3	-	-	-	-	22.60	22.60	203.41	4.52	18.08	274.60	274.60	-	-	227.39	23.98	12.37	0.88	-	-	-	-	-
$MgSO_4$	-	-	-	-	1.24	1.24	11.36	0.25	1.01	15.34	15.34	-	-	12.70	1.34	1.09	0.08	-	-	-	-	-
Na_2SO_4	-	-	-	-	0.20	0.20	1.84	0.04	0.16	2.48	2.48	-	-	2.05	0.22	0.18	0.01	-	-	-	-	-
As_2O_3	-	-	-	-	0.50	0.50	4.51	0.10	0.40	6.09	6.09	3.52	-	5.04	0.53	0.43	0.03	-	-	-	-	-
$CaSO_4$	-	-	-	-	0.35	0.35	3.18	0.07	0.28	4.29	4.29	-	-	3.55	0.37	0.30	0.02	-	-	-	-	-
$CdSO_4$	-	-	-	-	0.09	0.09	0.80	0.02	0.07	1.08	1.08	-	-	0.89	0.09	0.08	0.01	-	-	-	-	-
$CoSO_4$	-	-	-	-	0.04	0.04	0.36	0.01	0.03	0.48	0.48	-	-	0.40	0.04	0.03	0.00	-	-	-	-	-
$MnSO_4$	-	-	-	-	0.31	0.31	2.79	0.06	0.25	3.76	3.76	-	-	3.11	0.33	0.27	0.02	-	-	-	-	-
SiO_2	-	-	-	-	0.64	0.64	5.73	0.13	0.51	7.74	7.74	-	-	6.38	0.68	0.55	0.04	-	-	-	-	-
Ag	-	-	-	-	0.01	0.01	0.10	0.00	0.01	0.14	0.14	-	-	0.12	0.01	0.01	0.00	-	-	-	-	-
Others	-	-	-	-	0.15	0.15	1.32	0.02	0.12	1.78	1.78	5.98	-	1.47	0.16	0.13	0.01	-	-	-	-	-
GASES																						
O_2	-	-	-	-	-	-	-	-	-	-	-	-	989.46	454.60	454.60	454.60	454.60	454.60	-	-	-	-
N_2	-	-	-	-	-	-	-	-	-	-	-	-	3256.97	3256.97	3256.97	3256.97	3256.97	3256.97	-	-	-	-
SO_2	-	-	-	-	-	-	-	-	-	-	-	-	-	532.33	532.33	532.33	532.33	477.45	-	-	-	-
H_2O (vapour)	-	-	-	-	-	-	-	-	-	-	-	-	-	495.27	495.27	495.27	495.27	89.51	-	-	-	-
As_2O_3 (gaseous)	-	-	-	-	-	-	-	-	-	-	-	-	-	7.04	7.04	7.04	7.04	-	-	-	-	-
Hg.Cl	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SAL. SOLN. & LIQUIDS																						
$FeSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	18.74	-	-	-
$Fe_2(SO_4)_3$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	177.71	-	-	-
$ZnSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	112.78	(-)	-	-
As (as AsO_4^{-3})	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	2.89	-	2.66	-
$CuSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	1.86	-	-	-
$MgSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	13.66	-	-	-
Na_2SO_4	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	2.47	-	-	-
$MnSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	2.83	-	-	-
H_2SO_4	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	21.06	(-)	84.01	-
H_2O	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	472.76	(-)	477.36	405.76
Others	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	5.99	-
Total TPD (dry)(solids)	900.23	900.23	5.70	905.93	98.08	1004.01	882.69	19.62	78.46	1191.63	1191.63	9.51	-	986.76	104.07	84.45	5.99	-	354.00	-	92.66	-
Total TPD (wet)(solids)	963.25	922.74	-	928.44	-	1026.52	-	-	-	-	-	(-)	-	-	-	-	-	-	826.76	(-)	570.25	405.76
Total gases, (m^3/day x 10^6) (gases)	-	-	-	-	-	-	-	-	-	-	-	-	3.48	3.76	3.76	3.76	3.76	3.22	-	-	-	-

TABLE A2-10

LEACHING AND COPPER EXTRACTION CIRCUIT: TAILINGS REFLOAT PRODUCT MATERIAL BALANCE

	SOLIDS (TPD)				LIQUIDS (TPD)														
	S1	S2	S3	S4	L1	L2	L3	L4	L5	L6	L7	L8	L9	L10	L11	L12	L13	L14	L15
$ZnSO_4$	565.09	-	4.27	-	18.53	212.27	-	-	-	-	29.64	-	112.78	1061.85	-	-	-	-	923.36
$Fe_2(SO_4)_3$	-	-	6.19	-	-	-	-	-	-	-	-	-	177.71	-	-	-	-	-	-
$FeSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	18.74	5.58	-	-	-	-	-
Total Fe	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	0.02
$CoSO_4$	0.48	-	-	-	-	-	-	-	-	-	-	-	-	0.55	-	-	-	-	0.47
$CdSO_4$	1.08	-	-	-	-	-	-	-	-	-	-	-	-	1.24	-	-	-	-	1.07
$CuSO_4$	15.36	2.13	0.27	-	-	-	-	-	-	-	-	-	1.86	13.58	222.62	-	0.38	-	0.35
$PbSO_4$	47.20	47.20	47.20	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$MnSO_4$	3.76	-	-	-	1.79	20.32	-	-	-	-	2.86	-	2.83	30.44	-	-	-	-	30.14
$MgSO_4$	15.34	-	-	-	8.56	101.72	-	-	-	-	13.66	-	13.66	144.84	-	-	-	-	126.08
Na_2SO_4	2.48	-	-	-	-	-	-	-	-	-	-	-	2.47	-	-	-	-	-	-
$CaSO_4$	4.29	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$CaSO_4 \cdot 2H_2O$	-	26.92	26.92	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$ZnO.2ZnSO_4$	106.06	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Fe_2O_3	274.60	274.60	274.60	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$ZnO.Fe_2O_3$	129.50	129.50	6.47	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SiO_2	7.74	7.74	7.74	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
As_2O_3	6.09	6.09	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MnO_2	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$Fe(OH)_3$	-	2.39	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
(Zn,Fe)S	10.63	10.63	3.82	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
As	-	-	0.86	-	-	-	-	-	-	-	-	-	2.89	<1 mgpl	-	-	-	-	-
S°	-	-	1.98	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ag (kg/day)	135	135	135	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Gangue (e.g. Al_2O_3)	1.78	1.78	1.78	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H_2SO_4	-	-	-	-	25.74	-	-	-	-	-	41.29	-	21.06	-	473.02	-	-	-	-
H_2O	-	228.74	146.22	-	134.19	1668.81	581.54	-	228.74	-	214.61	141.10	472.76	2823.39	2699.30	-	-	-	2455.12
Cathode Copper	-	-	-	5.26	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Percol 351 (kg/day)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Percol 156 (kg/day)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Lix 64N (kg/day)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Kerosene (lt/day)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MnO_2 Ore	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Milled Limestone	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Suspended Solids (gpl)	-	-	-	-	-	-	-	0.10	-	-	-	-	-	-	-	-	-	-	-
Specific Gravity (g/cc)	-	4.85	5.16	-	1.26	1.17	-	-	-	-	1.26	-	1.57	1.38	1.15	-	-	-	1.39
Pulp Density (% Sol.)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
pH	-	-	-	-	-	-	-	-	-	-	-	-	-	4.0-5.2	-	-	0.8	1.25	4.2
Organic Sp. Grav. @ 15 C	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	0.80 g/cc	-	-	-
Temperature (°C)	-	-	-	-	40	30	25	-	25	-	40	-	95	40	40	40	40	60	60
VOLUME (m^3/day)	-	-	-	-	150.07	1718.76	-	2142.88	-	813.51	240.00	-	526.45	2956.38	2956.38	2956.38	2956.38	2956.38	2547.86
DRY WEIGHT	-	509.12	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
TOTAL WEIGHT	1191.63	737.86	528.46	5.26	188.81	2003.13	581.54	-	228.74	-	302.03	141.10	826.76	4081.47	3394.94	-	-	-	3536.59

TABLE A2-10 (Cont'd)

LEACHING AND COPPER EXTRACTION CIRCUIT TAILINGS REFLOAT PRODUCT MATERIAL BALANCE

	PULPS (TPD)						GASES (TPD)			REAGENTS (TPD)					
	P1	P2	P3	P4	P5	P6	G1	G2	G3	R1	R2	R3	R4	R5	R6
$ZnSO_4$	138.59	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$Fe_2(SO_4)_3$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$FeSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Total Fe	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$CoSO_4$	0.08	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$CdSO_4$	0.17	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$CuSO_4$	0.05	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$PbSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$MnSO_4$	4.51	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$MgSO_4$	18.91	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Na_2SO_4	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$CaSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$CaSO_4.2H_2O$	21.50	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$ZnO.2ZnSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Fe_2O_3	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$ZnO.Fe_2O_3$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SiO_2	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
As_2O_3	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MnO_2	0.51	-	-	-	1.44	-	-	-	-	-	-	-	-	-	-
$Fe(OH)_3$	3.92	-	-	-	-	-	-	-	-	-	-	-	-	-	-
(Zn.Fe)S	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
As (kg/day)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
S°	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ag	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Gangue (e.g. Al_2O_3)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H_2SO_4	-	-	-	-	-	-	-	-	-	-	183.31	-	-	-	-
H_2O	368.27	-	-	-	-	-	2.00	8.50	10.90	80.00	-	-	-	-	1.25
Cathode Copper	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Percol 351 (kg/day)	-	-	-	-	-	-	-	-	-	87.89	-	-	-	-	-
Percol 156 (kg/day)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	1.25
Lix 64N (kg/day)	-	-	-	-	-	-	-	-	-	-	-	17.00	-	-	-
Kerosene (lt/day)	-	-	-	-	-	-	-	-	-	-	-	280.00	-	-	-
MnO_2 Ore	-	-	-	-	-	-	-	-	-	-	-	-	0.09	-	-
Milled Limestone	-	-	-	-	-	-	-	-	-	-	-	-	-	12.50	-
Suspended Solids (gpl)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Specific Gravity (g/cc)	1.46	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Pulp Density (% Sol.)	33.8	11.1	31.0	28.2	-	0.6	-	-	-	-	-	-	-	-	-
pH	4.2	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Organic Sp. Grav. @ 15°C	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Temperature (°C)	60	40	40	95	-	40	-	-	-	-	-	-	-	-	-
VOLUME (m^3/day)	382.18	3061.23	918.38	600.55	-	2930.04	-	-	-	-	-	-	-	-	-
DRY WEIGHT	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
TOTAL WEIGHT	556.51	-	-	-	-	-	-	-	-	-	-	-	-	-	-

TABLE A2-11

PURIFICATION AND ELECTROWINNING CIRCUIT TAILINGS REFLOAT PRODUCT MATERIAL BALANCE

	SOLIDS (TPD)			LIQUIDS (TPD)															
	S1	S2	S3	L1	L2	L3	L4	L5	L6	L7	L8	L9	L10	L11	L12	L13	L14	L15	L16
Zn	11.60	12.09	-	373.83	-	17.73	-	-	-	-	-	-	-	-	-	403.95	-	117.73	1.00
Co	0.18	<0.18	-	0.18	-	-	-	-	-	-	-	<0.18	-	-	-	< 0.2mgpl	-	-	-
Cd	0.58	-	-	0.58	-	-	-	-	-	-	-	<1.0mgpl	-	-	-	< 1.0mgpl	-	-	-
Cu	0.14	-	-	0.14	-	-	-	-	-	-	-	<0.02mgpl	-	-	-	< 0.2mgpl	-	-	-
Mn	-	-	-	10.96	-	-	-	-	-	-	-	-	-	-	-	11.72	-	10.12	0.43
Mg	-	-	-	25.47	-	-	-	-	-	-	-	-	-	-	-	<27.49	-	27.08	1.15
Sb	<0.02	<0.02	-	-	-	-	-	-	-	-	-	-	-	-	-	0.02mgpl	-	-	-
Zn°	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Co°	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Cd°	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Cu°	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Zn as dross	-	-	5.70	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Zn as dust	-	-	20.32	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Zn as slab	-	-	259.20	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Zn as cathode	-	-	285.22	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H_2SO_4	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	403.91	24.02
H_2O	-	-	-	2455.12	-	96.31	-	-	-	-	-	-	-	-	-	2594.98	39.20	2105.52	89.42
MnO_2	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sb_2O_3(kg/day)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Limestone	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Percol 156 (kg/day)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Animal glue	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Na silicate	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sr carbonate	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Specific gravity (g/cc)	-	-	-	1.39	-	1.40	-	-	-	-	-	-	-	-	-	1.40	-	1.26	-
Pulp density (% Sol.)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
pH	-	-	-	4.2	-	-	-	-	-	-	-	-	-	-	-	5.0	-	-	-
Temperature (°C)	-	-	-	60	-	95	-	-	-	-	65	95	-	-	-	-	-	40	-
VOLUME (m^3/day)	-	-	-	2547.86	<1.00	100.95	0.00	0.00	0.00	0.00	-	2647.82	≈64.99	<1.00	5.00	2693.01	-	2354.64	100.00
WEIGHT OF SOLIDS	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
TOTAL WEIGHT	12.52	12.29	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-

TABLE A2-11 (Cont'd)

PURIFICATION AND ELECTROWINNING CIRCUIT: TAILINGS REFLOAT PRODUCT MATERIAL BALANCE

	PULPS (TPD)								GASES (TPD)				REAGENTS (TPD)					
	P1	P2	P3	P4	P5	P6	P7	P8	G1	G2	G3	G4	R1	R2	R3	R4	R5	R6
Zn	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Co	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Cd	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Cu	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Mn	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Mg	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Sb	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Zn°	11.60	11.60	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Co°	0.18	0.18	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Cd°	0.58	0.58	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Cu°	0.14	0.14	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Zn as dross	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Zn as dust	–	–	–	–	–	–	–	–	–	–	–	–	0.00	–	17.52	≃2.00	–	–
Zn as slab	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Zn as cathode	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
H_2SO_4	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
H_2O	–	–	–	–	–	–	–	–	8.70	22.30	86.20	353.00	–	4.50	<1.50	0.00	–	–
MnO_2	–	–	–	–	–	–	–	1.44	–	–	–	–	–	–	–	–	–	–
Sb_2O_3 (kg/day)	–	–	–	–	–	–	–	–	–	–	–	–	–	–	<27.00	0.00	–	–
Limestone	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	24.50
Percol 156 (kg/day)	–	–	–	–	–	–	–	–	–	–	–	–	–	4.50	–	–	–	–
Animal glue	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	≃0.05	–
Na silicate	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	≃0.13	–
Sr carbonate	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	N.A.	–
Specific gravity (g/cc)	1.40	1.56	–	–	–	–	1.55	–	–	–	–	–	–	–	–	–	–	–
Pulp density (% Sol.)	0.3	10.0	0.8	19.2	0.3	0.3	9.9	–	–	–	–	–	–	–	–	–	–	–
pH	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
Temperature (°C)	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
VOLUME (m^3/day)	2723.32	80.00	2752.01	60.00	2692.01	2693.01	80.00	–	–	–	–	–	–	–	–	–	–	–
WEIGHT OF SOLIDS	12.50	12.50	30.29	20.00	10.29	12.29	12.25	–	–	–	–	–	–	–	–	–	–	–
TOTAL WEIGHT	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–

TABLE A2-12

CADMIUM PLANT
TAILINGS REFLOAT PRODUCT MATERIAL BALANCE

	SOLIDS (TPD)				LIQUIDS (TPD)				PULPS (TPD)			REAGENTS (TPD)				
	S1	S2	S3	S4	L1	L2	L3	L4	P1	P2	P3	R1	R2	R3	R4	R5
Zn	11.60	0.10	-	-	5.38	17.33	-	17.73	-	-	-	-	-	-	-	-
Cu	0.14	0.16	-	-	-	< 1mgpl	-	<1mgpl	-	-	-	-	-	-	-	-
Cd	0.58	0.05	-	-	-	0.53	-	<20mgpl	-	-	-	-	-	-	-	-
Co	0.18	0.18	-	-	-	<50mgpl	-	<50mgpl	-	-	-	-	-	-	-	-
Sb	<0.02	<0.02	-	-	-	-	-	-	-	-	-	-	-	-	-	-
As	-	<0.04	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Product Cd	-	-	-	0.53	-	-	-	-	-	-	-	-	-	-	-	-
H_2SO_4	-	-	-	-	18.46	-	-	-	-	-	-	<1.00	-	-	-	-
H_2O	-	-	-	-	96.31	96.31	-	96.31	-	-	-	-	-	-	-	-
Zn dust	-	-	-	-	-	-	-	-	-	-	-	-	0.45	-	-	0.35
$CuSO_4.5H_2O$	-	-	-	-	-	-	-	-	-	-	-	-	-	0.05	-	-
As_2O_3	-	-	-	-	-	-	-	-	-	-	-	-	-	0.05	-	-
$SrCO_3$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	N.A.	-
Specific gravity (g/cc)	-	-	-	-	-	1.40	-	-	-	-	-	-	-	-	-	-
Pulp density (% Sol.)	-	-	-	-	-	-	-	-	0.4	N.A.	0.4	-	-	-	-	-
VOLUME (m^3/day)	-	-	-	-	107.64	99.95	<1.00	100.95	99.95	99.95	100.95	-	-	-	-	-
WEIGHT OF SOLIDS	12.52	<0.55	N.A.	-	-	-	-	-	-	-	-	-	-	-	-	-

TABLE A2-13

LEAD/SILVER RECOVERY PLANT: TAILINGS REFLOAT PRODUCT MATERIAL BALANCE

	SOLIDS (TPD)							LIQUIDS (TPD)							
	S1	S2	S3	S4	S5	S6	S7	L1	L2	L3	L4	L5	L6	L7	L8
$ZnFe_2O_4$	6.47	6.47	6.47	-	-	-	-	-	-	-	-	-	-	-	-
(Zn,Fe)S	3.82	3.82	3.82	-	-	-	-	-	-	-	-	-	-	-	-
Fe_2O_3	274.60	274.60	274.60	-	-	-	-	-	-	-	-	-	-	-	-
$PbSO_4$	47.20	2.36	2.36	-	-	-	-	-	-	-	-	-	-	-	-
$CaSO_4 \cdot 2H_2O$	26.92	62.64	62.64	-	1.25	-	-	-	-	-	-	-	-	-	-
SiO_2	7.74	7.74	7.74	-	-	-	-	-	-	-	-	-	-	-	-
As	0.86	0.86	0.86	-	-	-	-	-	-	-	-	-	-	-	-
S^o	1.98	1.98	1.98	-	-	-	-	-	-	-	-	-	-	-	-
Ag (kg/day)	135	≈ 18	≈ 22	-	≈124	≈124	-	≈117	≈ 7	≈124	-	-	-	-	-
Gangue	1.78	1.78	1.78	-	-	-	-	-	-	-	-	-	-	-	-
$ZnSO_4$	4.27	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$CuSO_4$	0.28	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$Fe_2(SO_4)_3$	6.19	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$PbO.H_2O$	-	-	-	-	34.46	-	-	-	-	-	-	-	-	-	-
$ZnO.H_2O$	-	-	-	-	2.54	-	-	-	-	-	-	-	-	-	-
$CuO.H_2O$	-	-	-	-	0.17	-	-	-	-	-	-	-	-	-	-
$Fe(OH)_3$	-	-	-	-	3.20	-	-	-	-	-	-	-	-	-	-
PbO	-	-	-	-	-	31.89	-	-	-	-	-	-	-	-	-
ZnO	-	-	-	-	-	2.08	-	-	-	-	-	-	-	-	-
CuO	-	-	-	-	-	0.14	-	-	-	-	-	-	-	-	-
$CaSO_4$	-	-	-	-	-	0.99	-	-	-	-	-	-	-	-	-
Fe_2O_3	-	-	-	-	-	2.39	-	-	-	-	-	-	-	-	-
CaO	-	-	-	12.06	-	-	-	-	-	-	-	-	-	-	-
H_2O(Free)	141.10	86.68	64.94	-	3.63	-	-	876.42	109.12	985.54	151.68	824.37	29.48	87.38	824.37
H_2O(Non-Free)	5.11	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$PbCl_2$	-	3.70	1.38	-	-	-	-	37.42	2.32	39.74	-	-	-	-	-
NaCl	-	25.93	22.35	-	-	-	37.38	262.20	37.56	299.76	10.45	252.51	1.64	33.98	289.89
$CaCl_2$	-	4.15	1.55	-	-	-	12.32	41.91	2.60	44.51	46.46	56.77	0.37	-	69.09
$ZnCl_2$	-	0.32	0.12	-	-	-	-	3.29	0.20	3.49	-	-	-	-	-
$FeCl_3$	-	0.45	0.16	-	-	-	-	4.57	0.28	4.85	-	-	-	-	-
$CuCl_2$	-	0.02	0.01	-	-	-	-	0.22	0.01	0.23	-	-	-	-	-
Na_2SO_4	-	0.10	0.03	-	-	-	-	0.96	0.07	1.03	-	-	-	-	-
TOTAL TONS/DAY	528.46	483.62	452.81	12.06	45.36	37.61	49.70	1227.11	152.17	1379.15	208.59	1133.65	31.49	121.36	1183.35
MOISTURE (%)	27.7	25.0	20.0	0.0	8.0	0.0	0.0	-	-	-	-	-	-	-	-

PRODUCT = 37.61 MTPD; 78.71% Pb, 88.98 oz/sh.T.

TABLE A2-14

SPENT ELECTROLYTE NEUTRALIZATION
TAILINGS REFLOAT PRODUCT MATERIAL BALANCE

	SOLIDS (TPD)	LIQUIDS (TPD)			REAGENTS (TPD)
	S1	L1	L2	L3	R1
Zn	3.23	89.17	-	85.94	-
Mg	-	19.75	-	20.53	0.78
Mn	-	7.38	-	7.38	-
$CaSO_4 \cdot 2H_2O$	538.33	-	-	-	-
H_2SO_4	-	305.92	-	-	-
H_2O	392.16	1599.45	517.83	1668.81	-
Limestone	-	-	-	-	313.12
pH	-	-	-	≈5.0	-
Temperature(°C)	-	40	-	30	-
VOLUME(m^3/day)	-	1788.69	-	1718.76	-
TOTAL WEIGHT	933.72	-	-	-	-

TABLE A2-15

TAILINGS REFLOAT PRODUCT SULPHUR BALANCE IN ROASTING SECTION

	S (TPD)	SO_2 (TPD)	H_2SO_4 (TPD)
Feed to roasters (30.0% Zn)	270.07	-	-
Calcines: (6.99% S)			
from feed	57.16	-	-
from HAL solution	26.11	-	-
Off gases from roasters:			
from feed	212.91	425.26 (1)	-
from HAL solution	53.61	107.07 (1)	-
Off gases after scrubber	239.03	477.45	-
Bleed from scrubber (4)	27.49	51.88 (2)	84.01
H_2SO_4 production from SO_2 (3)	231.86	463.13	688.16
Losses	7.17	14.32	-

	Gases Composition (%)	
	After roaster	After scrubber
O_2	8.71	9.87
N_2	69.88	80.79
SO_2	4.88	5.18
H_2SO_4	16.53	4.16

(1) $SO_2 + SO_3$

(2) As SO

(3) 97% efficiency in acid plant

(4) Bleed contains 176.0gpl H_2SO_4

TABLE A2-16

TAILINGS REFLOAT PRODUCT - THERMAL BALANCE

	Heat (Kcal/day) x 10^6 generated or heat content	Heat (Kcal/day) x 10^6 consumed
1.0 Roasting Section		
1.1 Feeder Dryer		
Preheat feed (dry base)	-	+ 14.57
Vaporize water (8 to 2.5%)	-	+ 38.48
Net balance	-	+ 53.05
1.2 Roasters		
Preheat fluidizing air	-	+ 683.46
Preheat feed (dry)	-	+ 81.16
Vaporize moisture of feed	-	+ 20.41
Reheat recirculated dust	-	+ 18.77
Preheat solids in HAL solution:		
$FeSO_4$	-	+ 1.62
$Fe_2(SO_4)_3$	-	+ 17.50
$ZnSO_4$	-	+ 14.10
H_2SO_4	-	+ 2.59
Decompose HAL solution		
$FeSO_4$	-	+ 7.32
$Fe_2(SO_4)_3$	-	+ 96.82
H_2SO_4	-	+ 10.67
Vaporize water from HAL solution	-	+ 428.89
Heats of reaction:		
(Zn,Fe)S	-871.76	-
FeS_2	-579.05	-
$CuFeS_2$	- 26.29	-
PbS	- 24.77	-
Heats losses through shell	-	+ 75.07
Spent to balance heat surplus (Equiv. 44.20 TPD spent)	-	+ 43.49
Net balance	-1501.87	+1501.87

TABLE A2-16

TAILINGS REFLOAT PRODUCT - THERMAL BALANCE (cont'd)

	Heat (Kcal/day) x 10^6	
	generated or heat content	consumed
1.3 Waste heat boilers		
Gases to boilers (650°C)	-724.29	-
Gases off boilers (350°C)	-346.01	-
Steam generation (350°C) (Equiv. 511.88 TPD steam)	-	+ 378.28
Net balance (at exit)	-346.01	+ 378.28
1.4 Electrostatic precipitators		
Gases to precipitator (350°C)	-346.01	-
Gases off precipitator (300°C)	-291.97	-
Heat losses through shell	-	+ 54 04
Net balance (at exit)	-291.97	+ 54.04
1.5 Water scrubber		
Gases to scrubber (300°C)	-291.97	-
Gases off scrubber (35°C)	- 11.02	-
Gases cooling (300 to 35°C)	-	(+ 274.58)
Water condensation (to 35°C)	-	(+ 280.94)
Net balance (at exit)	- 11.02	(+ 555.52) (*)
1.6 Overall balance		
Net heat recovered or required:		
Feed dryer	-	+ 53.05
Waste heat boilers	- 378.28	-
Scrubber	-	(+ 555.52) (*)

(*) Heat dissipated in cooling towers.

TABLE A2-16

TAILINGS REFLOAT PRODUCT - THERMAL BALANCE (cont'd)

	Heat (Kcal/day) x 10^6	
	generated or heat content	consumed
2.0 Hydrometallurgical Section*		
2.1 Neutral Leach		
Preheat calcine	-	+ 6.11
Preheat water	-	+ 35.00
Preheat neutral solution	-	+ 60.69
Preheat spent electrolyte	-	+ 3.65
Preheat neutral leach filtrate recycle	-	+ 2.22
Neutral solution (30°C)	- 8.47	-
Spent electrolyte (40°C)	- 2.22	-
Neutral leach filtrate recycle (60°C)	- 13.63	-
Heats of dissolution and reaction	- 84.50	-
Heat of evaporation	-	+ 1.15
Net balance	- 108.82	+ 108.82
2.2 Solution Cooling for Solvent Extraction (65.7-40°C) **	- 76.43	-
Net heat required or recovered in neutral leach	0	0
2.3 Hot Acid Leach		
Preheat neutral leach residue	-	+ 21.88
Preheat acid	-	+ 4.51
Preheat spent electrolyte	-	+ 12.88
Spent electrolyte (40°C)	- 3.50	-
Heat of reaction of sphalerite	- .54	-
Heat of reaction of zinc ferrite	- 35.14	-
Heat of solution of acid	- 32.95	-
Heat of evaporation	-	+ 4.68
Net balance **	- 72.13	+ 43.95
Net heat required or recovered	0	0

TABLE A2-16

TAILINGS REFLOAT PRODUCT - THERMAL BALANCE (cont'd)

		Heat (Kcal/day) x 10^6	
		generated or heat content	consumed
2. 4	Heat exchanger between solvent extraction and iron oxidation	-	+ 59.48
2. 5	Heat of evaporation in iron oxidation	-	+ 6.44
2. 6	Heat exchanger between iron oxidation and 1st purification	-	+ 12.93
2. 7	Heat of evaporation in 1st purification	-	+ 5.06
2. 8	Heat exchanger between 1st and 2nd purification	-	+ 82.95
2. 9	Heat of evaporation in 2nd purification	-	+ 12.29
2.10	Heat of evaporation in cooling towers	-	+ 213.57
Net balance		0	+ 392.72
Net heat required or recovered		0	+ 392.72

* conductive and convective heat losses not included in these balances. 20 percent overall heat losses for hydrometallurgical section assumed in economic analyses for fuel requirements.

** excess heat dissipated by conduction and convection.

TABLE A2-17

ROASTING SECTION
INTEGRATED SULPHATION ROAST-DEAD ROAST MATERIAL BALANCE

	SOLIDS (TPD)																				GASES (TPD)												LIQUIDS (TPD)			
	S1	S2	S3	S4	S5	S6	S7	S8	S9	S10	S11	S12	S13	S14	S15	S16	S17	S18	S19	S20	G1	G2	G3	G4	G5	G6	G7	G8	G9	G10	G11	G12	L1	L2	L3	L4
FEEDS (SOLIDS)																																				
(Zn,Fe)S	197.41	197.41	197.41	197.61	-	197.41	-	-	-	-	254.12	254.12	254.12	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
PbS	26.29	16.19	16.19	16.19	-	16.19	-	-	-	-	3.72	3.72	3.72	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$CuFeS_2$	6.76	6.76	6.74	6.76	-	6.76	-	-	-	-	1.47	1.47	1.47	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
FeS_2	152.37	152.37	152.37	152.37	-	152.37	-	-	-	-	1.21	1.21	1.21	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SiO_2	2.69	2.69	2.69	2.69	-	2.69	-	-	-	-	0.00	0.00	0.00	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$CaCO_3$	1.36	1.36	1.36	1.36	-	1.36	-	-	-	-	0.48	0.48	0.48	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$MgSiO_4$	0.82	0.82	0.82	0.81	-	0.82	-	-	-	-	0.00	0.00	0.00	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ag	0.06	0.06	0.06	0.06	-	0.06	-	-	-	-	0.02	0.02	0.02	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CoS	0.09	0.09	0.09	0.09	-	0.09	-	-	-	-	0.06	0.06	0.06	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CdS	0.35	0.35	0.35	0.35	-	0.35	-	-	-	-	0.48	0.48	0.48	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MnS	0.23	0.23	0.23	0.23	-	0.23	-	-	-	-	0.13	0.13	0.13	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
AsFeS	4.05	4.03	4.05	4.05	-	4.05	-	-	-	-	0.82	0.82	0.82	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Al_2O_3	0.74	0.74	0.24	0.74	-	0.74	-	-	-	-	0.00	0.00	0.00	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Others	5.61	5.61	5.61	5.61	-	5.61	-	-	-	-	5.90	5.90	5.90	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H_2O (as moisture)	27.27	9.74	9.74	9.74	-	9.74	-	-	-	-	18.80	18.80	18.80	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-				
DROSS (SOLID)																																				
Zn	-	-	5.70	-	-	5.78	-	-	-	-	-	-	5.70	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CALCINES (SOLIDS)																																				
ZnO	-	-	-	-	0.00	0.00	0.00	0.00	0.00	0.00	-	-	-	26.29	96.39	42.06	18.52	175.25	175.25	-	-	0.00	0.00	0.00	0.00	-	148.97	52.58	10.52	0.00	-	-	-	-	-	-
$ZnSO_4$	-	-	-	-	21.48	21.48	17.47	4.81	200.72	256.18	-	-	-	0.31	1.15	0.50	0.13	2.09	258.27	-	-	126.21	25.49	21.48	0.00	-	1.78	0.63	0.13	0.00	-	-	-	-	-	-
$ZnO.2ZnSO_4$	-	-	-	-	3.75	3.75	3.05	0.78	35.09	44.78	-	-	-	0.00	0.00	0.00	0.00	0.00	44.78	-	-	39.54	4.45	3.75	0.00	-	0.00	0.00	0.00	0.00	-	-	-	-	-	-
$ZnO.Fe_2O_3$	-	-	-	-	4.70	4.70	3.82	0.88	43.92	56.05	-	-	-	6.92	25.37	21.07	2.77	46.13	102.18	-	-	49.58	5.58	4.78	0.00	-	39.21	13.84	2.77	0.00	-	-	-	-	-	-
(Zn,Fe)S (remnant)	-	-	-	-	0.38	0.38	0.31	0.87	3.60	4.59	-	-	-	0.44	1.60	0.78	0.17	2.91	7.58	-	-	4.05	0.45	0.38	0.00	-	2.47	0.87	0.17	0.00	-	-	-	-	-	-
PbO	-	-	-	-	0.80	0.00	0.00	0.00	0.00	0.00	-	-	-	0.52	1.90	0.83	0.21	3.45	3.45	-	-	0.00	0.00	0.00	0.00	-	2.94	2.04	0.22	0.00	-	-	-	-	-	-
$PbSO_4$	-	-	-	-	1.71	1.71	1.31	0.32	16.00	20.42	-	-	-	0.00	0.00	0.00	0.00	0.00	20.41	-	-	18.01	2.81	1.71	0.00	-	0.00	0.00	0.00	0.00	-	-	-	-	-	-
CuO	-	-	-	-	0.00	0.00	0.00	0.00	0.00	0.00	-	-	-	0.18	0.35	0.15	0.04	0.64	0.64	-	-	0.00	0.00	0.00	0.00	-	0.54	0.19	0.04	0.00	-	-	-	-	-	-
$CuSO_4$	-	-	-	-	0.66	0.66	0.54	0.12	6.20	7.91	-	-	-	0.00	0.00	0.00	0.00	0.00	7.91	-	-	6.98	0.78	0.66	0.00	-	0.00	0.00	0.00	0.00	-	-	-	-	-	-
Fe_2O_3	-	-	-	-	12.24	13.24	9.35	2.29	114.39	146.00	-	-	-	0.00	0.00	0.00	0.00	0.00	246.00	-	-	135.90	21.51	19.22	0.00	-	0.00	0.00	0.00	0.00	-	-	-	-	-	-
$MgSO_4$	-	-	-	-	0.06	0.06	0.05	0.01	0.57	0.73	-	-	-	0.00	0.00	0.00	0.00	0.00	0.73	-	-	0.64	0.07	0.06	0.00	-	0.00	0.00	0.00	0.00	-	-	-	-	-	-
$MnSO_4$	-	-	-	-	0.83	0.83	0.03	0.00	0.31	0.40	-	-	-	0.83	0.12	0.05	0.01	0.22	0.62	-	-	0.34	0.83	0.00	0.00	-	0.18	0.06	0.01	0.00	-	-	-	-	-	-
$CdSO_4$	-	-	-	-	0.04	0.04	0.03	0.01	0.36	0.46	-	-	-	0.09	0.35	0.15	0.04	0.63	1.09	-	-	0.41	0.05	0.04	0.00	-	0.54	0.19	0.04	0.00	-	-	-	-	-	-
$CoSO_4$	-	-	-	-	0.81	0.81	0.01	0.00	0.13	0.26	-	-	-	0.02	0.06	0.82	0.01	0.10	0.26	-	-	0.14	0.01	0.01	0.00	-	0.09	0.03	0.01	0.00	-	-	-	-	-	-
As_2O_3	-	-	-	-	0.27	0.27	0.27	0.04	2.06	2.63	-	-	-	0.08	0.29	0.13	0.03	0.53	3.16	-	-	2.37	0.31	0.83	0.00	-	0.45	0.16	0.03	0.00	-	-	-	-	-	-
SiO_2	-	-	-	-	0.28	0.28	0.23	0.85	2.62	3.35	-	-	-	0.00	0.00	0.00	0.00	0.00	3.35	-	-	2.95	0.33	0.03	0.00	-	0.00	0.00	0.00	0.00	-	-	-	-	-	-
$CaSO_4$	-	-	-	-	0.16	0.16	0.23	0.03	1.46	1.86	-	-	-	0.18	0.36	0.26	0.04	0.66	2.52	-	-	1.65	0.19	0.16	0.00	-	0.56	0.28	0.04	0.00	-	-	-	-	-	-
Ag	-	-	-	-	0.00	0.00	0.00	0.00	0.85	0.06	-	-	-	0.00	0.00	0.00	0.00	0.01	0.07	-	-	0.85	0.00	0.00	0.00	-	0.00	0.00	0.00	0.00	-	-	-	-	-	-
Others	-	-	-	-	0.06	0.06	0.05	0.01	0.60	0.77	-	-	-	0.12	0.45	0.19	0.05	0.81	1.58	3.91	-	3.42	2.12	2.81	2.75	-	1.84	1.41	1.22	2.17	3.92	-	-	-	-	-
GASES																																				
O_2	-	-	-	-	--	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	128.13	196.70	396.70	196.70	196.78	248.31	38.66	38.66	38.66	38.66	235.36	235.36	-	-	-	-
N_2	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	1409.24	1409.26	3409.26	1409.26	1409.26	816.65	816.65	816.65	816.65	816.65	2225.91	2225.91	-	-	-	-
SO_2	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	310.85	318.85	318.85	310.85	-	171.18	171.28	171.20	171.28	482.05	482.85	-	-	-	-
H_2O (vapour)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	212.31	212.32	232.32	212.32	-	18.80	18.00	18.00	18.80	231.12	64.87	-	-	-	-
AsO_3 (gaseous)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	3.85	3.85	3.05	3.85	-	0.42	0.42	0.42	0.42	3.44	-	-	-	-	-
Hg, Cl	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
NAL. SOLN. & LIQUIDS																																				
$FeSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	12.59	-	-	-
$Fe_2(SO_4)_3$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	142.74	-	-	-
$ZnSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	67.44	(-)	-	-
As (as AsO_4^{-3})	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	1.57	-	-	-	-	-	-	-	-	-	-	-	-	2.11	-	1.57	-
$CuSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	2.21	-	-	-
$MgSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	5.91	-	-	-
Na_2SO_4	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	1.07	-	-	-
$MnSO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	1.22	-	-	-
H_2SO_4	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	30.13	(-)	35.85	(-)
H_2O	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	(-)	-	-	-	-	-	-	-	-	-	-	-	-	202.58	(-)	166.25	(-)
Others	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	3.92	-
Total TPD (dry)(solids)	389.52	389.52	395.22	389.52	49.30	441.85	39.76	8.54	427.86	546.35	268.63	268.63	274.33	35.02	128.39	56.81	14.02	233.43	779.78	1.57	-	492.18	64.18	55.81	2.75	-	254.60	71.20	15.19	1.27	3.92	-	245.42	-	41.34	(-)
Total TPD (wet)(solids)	416.79	399.26	404.96	399.26	-	450.79	-	-	-	-	287.43	287.43	293.13	-	-	-	-	-	-	(-)	-	-	-	-	-	-	-	-	-	-	-	-	448.00	(-)	207.59	(-)
Total (m^3/day x 10^6)(gases)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	1.491	1.638	1.638	1.638	1.438	0.827	0.767	0.767	0.767	0.767	2.405	2.298	-	-	-	-

TABLE A2-18

LEACHING AND COPPER EXTRACTION CIRCUIT: INTEGRATED DEAD ROAST-SULPHATION ROAST MATERIAL BALANCE

	SOLIDS (TPD)			LIQUIDS (TPD)									PULPS (TPD)				GASES (TPD)		REAGENTS (TPD)		
	S1	S2	S3	L1	L2	L3	L4	L5	L6	L7	L8	L9	P1	P2	P3	P4	G1	G2	R1	R2	R3
$ZnSO_4$	256.18	-	4.27	160.90	113.50	-	-	-	-	-	67.46	934.23	-	-	-	-	-	-	-	-	-
$Fe_2(SO_4)_3$	-	-	6.19	-	-	-	-	-	-	-	142.73	-	-	-	-	-	-	-	-	-	-
$FeSO_4$	-	-	-	-	-	-	-	-	-	-	12.59	-	-	-	-	-	-	-	-	-	-
Total Fe	-	-	-	-	-	-	-	-	-	-	-	<20 mgpl	-	-	-	-	-	-	-	-	-
$CoSO_4$	0.26	-	-	-	-	-	-	-	-	-	-	0.26	-	-	-	-	-	-	-	-	-
$CdSO_4$	1.09	-	-	-	-	-	-	-	-	-	-	1.09	-	-	-	-	-	-	-	-	-
$CuSO_4$	7.91	0.93	0.15	-	-	-	-	-	-	-	2.21	6.98	-	-	-	-	-	-	-	-	-
$PbSO_4$	20.42	20.42	25.11	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$MnSO_4$	0.62	-	-	31.68	22.33	-	-	-	-	-	-	62.51	-	-	-	-	-	-	-	-	-
$MgSO_4$	0.73	-	-	52.52	39.11	-	-	-	-	-	-	92.37	-	-	-	-	-	-	-	-	-
$CaSO_4$	2.52	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$CaSO_4 \cdot 2H_2O$	-	3.19	3.19	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$ZnO.2ZnSO_4$	44.78	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Fe_2O_3	146.00	146.00	146.00	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$ZnO.Fe_2O_3$	102.18	102.18	5.11	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SiO_2	3.35	3.35	3.35	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
As_2O_3	3.16	3.16	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MnO_2	-	-	-	-	-	-	-	-	-	-	-	-	3.74	-	-	-	-	-	-	-	-
ZnO	176.30	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
PbO	3.45	3.45	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CuO	0.64	0.64	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
(Zn,Fe)S	7.50	7.50	2.92	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$Fe(OH)_3$	-	2.39	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
As	-	-	0.45	-	-	-	-	-	-	-	2.11	<1 mgpl	-	-	-	-	-	-	-	-	-
S^o	-	-	1.51	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ag(kg/day)	730	730	730	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Gangue(e.g.Al_2O_3)	1.58	1.58	1.58	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H_2SO_4	-	-	-	223.37	-	-	-	-	-	-	10.13	-	-	-	-	-	-	-	-	-	173.70
H_2O	-	132.47	79.67	1165.06	892.28	366.91	-	132.47	-	74.63	202.55	2498.97	-	-	-	-	12.00	8.50	-	47.00	-
Percol 351(kg/day)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	49.00	-
$KMnO_4$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	1.45	-	-
Suspended Solids(gpl)	-	-	-	-	-	-	0.10	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Specific Gravity (g/cc)	-	5.04	4.90	1.25	1.16	-	1.39	-	-	-	1.74	1.39	-	-	-	-	-	-	-	-	-
Pulp Density(% Sol.)	-	-	-	-	-	-	-	-	-	-	-	-	-	13.1	22.6	27.8	-	-	-	-	-
pH	-	-	-	-	-	-	-	-	-	-	-	4.0-5.2	-	-	-	-	-	-	-	-	-
Temperature (°C)	-	-	-	40	30	25	-	25	-	-	95	60	-	60	60	95	-	-	-	-	-
VOLUME(m^3/day)	-	-	-	1302.91	918.93	-	1861.80	-	731.58	-	253.20	2593.37	-	2616.85	785.05	373.61	-	-	-	-	-
TOTAL WEIGHT	798.74	427.33	279.51	1633.53	-	-	-	-	-	74.63	439.78	3596.41	-	-	-	-	-	-	-	-	-

TABLE A2-19

PURIFICATION AND ELECTROWINNING CIRCUIT
INTEGRATED DEAD ROAST-SULPHATION ROAST MATERIAL BALANCE

	SOLIDS (TPD)			LIQUIDS (TPD)															
	S1	S2	S3	L1	L2	L3	L4	L5	L6	L7	L8	L9	L10	L11	L12	L13	L14	L15	L16
Zn	8.87	11.99	-	378.23	-	13.31	-	-	-	-	-	-	-	-	-	405.83	-	119.90	0.70
Co	0.10	<0.10	-	0.10	-	-	-	-	-	-	-	<0.18mgpl	-	-	-	<0.2mgpl	-	-	-
Cd	0.59	-	-	0.59	-	-	-	-	-	-	-	<1.0mgpl	-	-	-	<1.0mgpl	-	-	-
Cu	2.78	-	-	2.78	-	-	-	-	-	-	-	<0.02mgpl	-	-	-	<0.2mgpl	-	-	-
Mn	-	-	-	22.73	-	-	-	-	-	-	-	-	-	-	-	23.98	-	21.20	0.73
Mg	-	-	-	18.66	-	-	-	-	-	-	-	-	-	-	-	19.81	-	19.52	0.57
Sb	<0.02	<0.02	-	-	-	-	-	-	-	-	-	-	-	-	-	<0.02mgpl	-	-	-
Zn°	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Co°	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Cd°	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Cu°	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Zn as dross	-	-	5.70	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Zn as dust	-	-	20.32	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Zn as slab	-	-	259.20	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Zn as cathode	-	-	285.22	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H_2SO_4	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	411.13	16.80
H_2O	-	-	-	2465.24	-	72.13	-	-	-	-	-	-	-	-	-	2607.02	39.20	2144.37	62.59
MnO_2	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sb_2O_3(kg/day)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Limestone	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Percol 156(kg/day)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Animal glue	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Na silicate	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sr carbonate	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Specific gravity (g/cc)	-	-	-	1.39	-	1.40	-	-	-	-	-	-	-	-	-	1.39	-	1.25	-
Pulp density (% Sol.)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
pH	-	-	-	4.0-5.2	-	-	-	-	-	-	-	-	-	-	-	≈5.0	-	-	-
Temperature (°C)	-	-	-	60	-	95	-	-	-	-	65	95	-	-	-	-	-	40	-
VOLUME (m^3/day)	-	-	-	2593.37	<1.00	75.85	0.00	0.00	0.00	0.00	-	2666.30	60.00	<1.00	5.00	2705.50	-	2398.09	70.00
WEIGHT OF SOLIDS	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
TOTAL WEIGHT	12.36	12.11	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-

TABLE A2-19 (Cont'd)

PURIFICATION AND ELECTROWINNING CIRCUIT
INTEGRATED DEAD ROAST-SULPHATION ROAST MATERIAL BALANCE

	PULPS (TPD)								GASES (TPD)				REAGENTS (TPD)					
	P1	P2	P3	P4	P5	P6	P7	P8	G1	G2	G3	G4	R1	R2	R3	R4	R5	R6
Zn	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Co	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Cd	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Cu	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Mn	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Mg	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	0.04
Sb	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Zn°	8.87	8.87	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Co°	0.10	0.10	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Cd°	0.59	0.59	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Cu°	2.78	2.78	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Zn as dross	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Zn as dust	-	-	-	-	-	-	-	-	-	-	-	-	0.00	-	17.42	2.00	-	-
Zn as slab	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Zn as cathode	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H_2SO_4	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H_2O	-	-	-	-	-	-	-	-	8.70	22.30	86.20	353.00	-	4.50	<1.50	0.00	-	-
MnO_2	-	-	-	-	-	-	-	3.74	-	-	-	-	-	-	-	-	-	-
Sb_2O_3 (kg/day)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	<27.00	0.00	-	-
Limestone	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	17.13
Percol 156 (kg/day)	-	-	-	-	-	-	-	-	-	-	-	-	-	4.50	-	-	-	-
Animal glue	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	0.05	-
Na silicate	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	0.13	-
Sr carbonate	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	N.A.	-
Specific gravity (g/cc)	1.40	1.55	-	-	-	-	1.55	-	-	-	-	-	-	-	-	-	-	-
Pulp density (% Sol.)	0.3	10.0	0.8	19.2	0.3	0.3	9.8	-	-	-	-	-	-	-	-	-	-	-
pH	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Temperature (°C)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
VOLUME (m^3/day)	2741.80	80.00	2765.50	60.00	2705.50	2705.50	80.00	-	-	-	-	-	-	-	-	-	-	-
WEIGHT OF SOLIDS	12.36	12.36	30.10	20.00	10.10	12.11	12.11	-	-	-	-	-	-	-	-	-	-	-
TOTAL WEIGHT	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-

TABLE A2-20

CADMIUM PLANT
INTEGRATED DEAD ROAST-SULPHATION ROAST MATERIAL BALANCE

	SOLIDS (TPD)				LIQUIDS (TPD)				PULPS (TPD)			REAGENTS (TPD)				
	S1	S2	S3	S4	L1	L2	L3	L4	P1	P2	P3	R1	R2	R3	R4	R5
Zn	8.87	0.56	-	-	4.04	12.91	-	13.31	-	-	-	-	-	-	-	-
Cu	2.78	2.80	-	-	-	<1mgpl	-	<1mgpl	-	-	-	-	-	-	-	-
Cd	0.59	0.20	-	-	-	0.39	-	<50mgpl	-	-	-	-	-	-	-	-
Sb	0.10	0.10	-	-	-	<50mgpl	-	<50mgpl	-	-	-	-	-	-	-	-
As	<0.02	<0.02	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Product Cd°	-	<0.04	-	0.39	-	-	-	-	-	-	-	-	-	-	-	-
H_2SO_4	-	-	-	-	13.83	-	-	-	-	-	-	<1.00	-	-	-	-
H_2O	-	-	-	-	72.13	72.13	-	72.13	-	-	-	-	-	-	-	-
Zn Dust	-	-	-	-	-	-	-	-	-	-	-	-	0.56	-	-	0.34
$CuSO_4 \cdot 5H_2O$	-	-	-	-	-	-	-	-	-	-	-	-	-	0.04	-	-
As_2O_3	-	-	-	-	-	-	-	-	-	-	-	-	-	0.04	-	-
$SrCO_3$	-	-	-	-	-	-	-	-	-	-	-	-	-	-	N.A.	-
Specific gravity (g/cc)	-	-	-	-	-	1.40	-	-	-	-	-	-	-	-	-	-
Pulp density (% Sol.)	-	-	-	-	-	-	-	-	3.4	N.A.	0.4	-	-	-	-	-
VOLUME (m^3/day)	-	-	-	-	80.67	74.85	<1.00	75.85	74.85	74.85	75.95	-	-	-	-	-
WEIGHT OF SOLIDS	12.36	3.72	N.A.	-	-	-	-	-	-	-	-	-	-	-	-	-

TABLE A2-21

LEAD/SILVER RECOVERY PLANT: INTEGRATED SR/DR MATERIAL BALANCE

	SOLIDS (TPD)							LIQUIDS (TPD)							
	S1	S2	S3	S4	S5	S6	S7	L1	L2	L3	L4	L5	L6	L7	L8
$ZnFe_2O_4$	5.11	5.11	5.11	–	–	–	–	–	–	–	–	–	–	–	–
(Zn.Fe)S	2.92	2.92	2.92	–	–	–	–	–	–	–	–	–	–	–	–
Fe_2O_3	146.00	146.00	146.00	–	–	–	–	–	–	–	–	–	–	–	–
$PbSO_4$	25.11	1.26	1.26	–	–	–	–	–	–	–	–	–	–	–	–
$CaSO_4 \cdot 2H_2O$	3.16	3.16	3.16	–	0.67	–	–	–	–	–	–	–	–	–	–
SiO_2	3.67	3.67	3.67	–	–	–	–	–	–	–	–	–	–	–	–
As	0.47	0.47	0.47	–	–	–	–	–	–	–	–	–	–	–	–
S°	1.51	1.51	1.51	–	–	–	–	–	–	–	–	–	–	–	–
Ag (kg/day)	73	≈ 9	≈ 5	–	≈ 67	≈ 67	–	≈ 64	≈ 3	≈ 67	–	–	–	–	–
Gangue	1.30	1.30	1.30	–	–	–	–	–	–	–	–	–	–	–	–
$ZnSO_4$	4.27	–	–	–	–	–	–	–	–	–	–	–	–	–	–
$CuSO_4$	0.15	–	–	–	–	–	–	–	–	–	–	–	–	–	–
$Fe_2(SO_4)_3$	6.19	–	–	–	–	–	–	–	–	–	–	–	–	–	–
$PbO.H_2O$	–	–	–	–	18.42	–	–	–	–	–	–	–	–	–	–
$ZnO.H_2O$	–	–	–	–	2.56	–	–	–	–	–	–	–	–	–	–
$CuO.H_2O$	–	–	–	–	0.09	–	–	–	–	–	–	–	–	–	–
$Fe(OH)_3$	–	–	–	–	3.21	–	–	–	–	–	–	–	–	–	–
PbO	–	–	–	–	–	17.10	–	–	–	–	–	–	–	–	–
ZnO	–	–	–	–	–	2.09	–	–	–	–	–	–	–	–	–
CuO	–	–	–	–	–	0.07	–	–	–	–	–	–	–	–	–
$CaSO_4$	–	–	–	–	–	0.53	–	–	–	–	–	–	–	–	–
Fe_2O_3	–	–	–	–	–	2.40	–	–	–	–	–	–	–	–	–
CaO	–	–	–	8.30	–	–	–	–	–	–	–	–	–	–	–
H_2O(Free)	74.66	39.35	29.64	–	2.18	–	–	471.69	49.43	521.12	78.61	436.50	17.68	39.72	436.50
H_2O(Non Free)	5.04	–	–	–	–	–	–	–	–	–	–	–	–	–	–
$PbCl_2$	–	1.68	0.63	–	–	–	–	20.19	1.05	21.24	–	–	–	–	–
NaCl	–	12.06	10.31	–	–	–	21.06	144.54	17.19	161.73	24.59	136.58	1.01	15.44	157.64
$CaCl_2$	–	1.34	0.50	–	–	–	8.05	16.02	0.84	16.86	4.55	25.24	0.19	–	33.29
$ZnCl_2$	–	0.28	0.10	–	–	–	–	3.33	0.18	3.51	–	–	–	–	–
$FeCl_3$	–	0.39	0.15	–	–	–	–	4.63	0.24	4.87	–	–	–	–	–
$CuCl_2$	–	0.01	<0.01	–	–	–	–	0.12	<0.01	0.13	–	–	–	–	–
Na_2SO_4	–	0.04	0.01	–	–	–	–	0.52	0.03	0.55	–	–	–	–	–
TOTAL TONS/DAY	279.63	220.56	206.75	8.30	27.20	22.26	29.11	661.10	68.97	730.08	107.75	598.32	18.88	55.16	627.43
MOISTURE (%)	28.5	25.0	20.0	0.0	8.0	0.0	0.0	–	–	–	–	–	–	–	–

PRODUCT = 22.26 MTPD; 71.31% Pb, 81.29 oz/sh.T.

TABLE A2-22

SPENT ELECTOLYTE NEUTRALIZATION
INTEGRATED DEAD ROAST-SULPHATION ROAST MATERIAL BALANCE

	SOLIDS (TPD)	LIQUIDS (TPD)			REAGENTS (TPD)
	S1	L1	L2	L3	R1
Zn	1.72	47.67	-	45.95	-
Mg	-	7.76	-	8.18	0.42
Mn	-	8.42	-	8.42	-
$CaSO_4.2H_2O$	286.78	-	-	-	-
H_2SO_4	-	163.45	-	-	-
H_2O	208.91	852.53	278.66	892.28	-
Limestone	-	-	-	-	166.80
pH	-	-	-	≈5.0	-
Temperature(°C)	-	40	-	30	-
VOLUME(m^3/day)	-	953.40	-	918.93	-
TOTAL WEIGHT	497.41	-	-	-	-

TABLE A2-23

INTEGRATED SULPHATION ROAST DEAD ROAST PLANT SULPHUR BALANCE IN ROASTING SECTION

	Sulphation Roast			Dead Roast		
	S (TPD)	SO_2 (TPD)	H_2SO_4 (TPD)	S (TPD)	SO_2 (TPD)	H_2SO_4 (TPD)
Feed to roasters	116.86	-	-	85.96	-	-
Calcines:						
from feed	24.73	-	-	0.18	-	-
from HAL solution	15.61	-	-	-	-	-
Off gases (final)	132.42	264.49	-	85.78	171.34	-
Bleed from scrubber$^{(2)}$	11.73	23.42$^{(1)}$	35.85	-	-	-
H_2SO_4 production$^{(3)}$	132.42	264.49	392.73	85.78	171.34	254.41
Losses	3.10	6.20	-	2.01	4.02	-

Gases Composition (%)

	SR		DR	Total$^{(4)}$
	after roaster	after scrubber		
O_2	8.41	9.88	3.16	7.24
N_2	68.83	79.53	86.93	82.44
SO_2	6.64	7.09	8.02	7.46
H_2O	16.13	3.50	1.89	2.87

(1) As SO_3

(2) Bleed contains 215.6 gpl H_2SO_4

(3) 97% efficiency in acid plant

(4) Combined off gases from SR + DR reactors

TABLE A2-24

INTEGRATED PLANT - SULPHATION ROAST-DEAD ROAST THERMAL BALANCE

	Heat (Kcal/day) x 10^6			
	generated or heat content		consumed	
	SR	DR	SR	DR
<u>1.0 Roasting Section</u>				
<u>1.1 Feed Dryer</u>				
Preheat feed (dry base)	-	-	+ 6.30	-
Vaporize water (8 to 2.5% SR) (7% DR)	-	-	+ 16.65	-
Net balance	-	-	+ 22.95	-
<u>1.2 Roasters</u>	SR		SR	
Preheat fluidizing air	-		+269.54	
Preheat feed (dry)	-		+ 35.12	
Vaporize moisture of feed	-		+ 8.83	
Reheat recirculated dust	-		+ 7.44	
Preheat solids in HAL solution:				
$FeSO_4$	-		+ 1.09	
$Fe_2(SO_4)_3$	-		+ 14.06	
$ZnSO_4$	-		+ 8.43	
H_2SO_4	-		+ 1.25	
Decompose HAL solution:				
$FeSO_4$	-		+ 4.92	
$Fe_2(SO_4)_3$	-		+ 77.79	
H_2SO_4	-		+ 5.14	
Vaporize water from HAL solution:	-		+183.75	
Heats of reaction:				
(Zn,Fe)S	-377.20		-	
FeS_2	-250.55		-	
$CuFeS_2$	- 11.38		-	
PbS	- 10.72		-	
Heat losses through shell	-		+ 32.49	
Net balance	-649.85		+649.85	

TABLE A2-24

INTEGRATED PLANT - SULPHATION ROAST-DEAD ROAST THERMAL BALANCE (cont'd)

	Heat (Kcal/day) x 10^6			
	generated or heat content		consumed	
	SR	DR	SR	DR
1.3 Waste heat boilers				
Gases to boilers (650°C SR, 920°C DR)	-313.39	-289.95	-	-
Gases off boilers (350°C)	-149.72	-138.52	-	-
Steam generation (350°C)	-	-	+163.67	+151.43
Equiv TPD steam	279.00		258.14	
Net balance (at exit)				
1.4 Electrostatic Precipitator	SR	DR	SR	DR
Gases to precipitator (350°C)	-149.72	-138.52	-	-
Gases off precipitator (300°C)	-128.33	-118.73	-	-
Heat losses through shell	-	-	+ 21.39	+ 19.79
Net balance (at exit)	-128.33	-118.73	+ 21.39	+ 19.79
1.5 Water scrubber	SR		DR	
Gases to scrubber (300°C)	-128.33		-	
Gases off scrubber (35°C)	- 14.97		-	
Gases cooling (300 to 35°C)	-		(+113.36)	
Water condensation (to 35°C)	-		(+120.44)	
Net balance (at exit)	-		(+233.80) (*)	
1.6 Overall balance	SR	DR	SR	DR
Net heat recovered or required:				
Feed dryer	-	-	+ 22.95	-
Waste heat boilers	-163.67	-151.43	-	-
Scrubber	-	-	(+233.80) (*)	-

(*) Heat dissipated in cooling towers.

TABLE A2-24

INTEGRATED PLANT - SULPHATION ROAST-DEAD ROAST THERMAL BALANCE (cont'd)

	generated or heat content	consumed
2.0 Hydrometallurgical Section*		
2.1 Neutral Leach		
Preheat calcine	-	+ 4.61
Preheat water	-	+ 25.36
Preheat neutral solution	-	+ 37.40
Preheat spent electrolyte	-	+ 39.17
Neutral solution (30°C)	- 5.57	-
Spent electrolyte (40°C)	- 18.94	-
Heats of dissolution and reaction	- 88.86	-
Heat of evaporation	-	+ 6.83
Net balance	-113.37	+ 113.37
Net heat required or recovered	0	0
2.2 Solution Cooling for 1st Stage		
Purification (71.1-65°C) **	- 13.65	-
2.3 Hot Acid Leach		
Preheat neutral leach residue	-	+ 12.65
Preheat acid	-	+ 4.12
Heat of reaction of sphalerite	- .36	-
Heat of reaction of zinc ferrite	- 27.56	-
Heat of solution of acid	- 30.08	-
Heat of evaporation	-	+ 4.68
Net balance **	- 58.00	+ 21.45
Net heat required or recovered	0	0
2.4 Heat of evaporation in 1st purification	-	+ 5.02
2.5 Heat exchanger between 1st and 2nd purification	-	+ 76.41

TABLE A2-24

INTEGRATED PLANT - SULPHATION ROAST-DEAD ROAST THERMAL BALANCE (Cont'd)

	generated or heat content	consumed
2.6 Heat of evaporation in 2nd purification	-	+ 12.19
2.7 Heat of evaporation in cooling towers	-	+ 213.57
Net balance	-	+ 307.19
Net heat required or recovered	0	+ 307.19

* conductive and convective heat losses not included in these balances. 20 percent overall heat losses for hydrometallurgical section assumed in economic analyses for fuel requirements.

** excess heat dissipated by conduction and convection.

TABLE A2-25

DUST COLLECTION BALANCE IN ROASTING SECTION

	Bulk Concentrate	Tailings Refloat Product	Integrated Plant	
			SR	DR
No. roasters	2	2	1	1
Roaster capacity, each (TPD)	420	450	400	280
Roaster diameter, each (TPD)	11.2	11.8	11.1	7.2
Total calcine produced (TPD)	1185.92	1191.63	546.35	233.43
Net calcine loss (TPD)	5.96	5.99	2.75	1.17
Net calcine loss (%)	0.5	0.5	0.5	0.5
Dust carry over (T/m^2/day)	10.89(1)	8.97(1)	8.97(1)	4.85(2)

	Bulk Concentrate		Tailings Refloat Product		Integrated Plant			
					SR		DR	
	% Eff.	TPD	% Eff.	TPD	% Eff.	TPD	% Eff.	TPD
Cyclones	90	960.59	90	882.69	90	427.86	24	56.01
Waste heat boilers	2	21.35	2	19.62	2	8.54	55	128.39
Electrostatic precip.	8	85.39	8	78.46	8	39.76	6	14.02
Total(3)	100	1067.32	100	980.77	100	476.16	100	198.42

(1) 685 °C, 75 cm/sec
(2) 930 °C, 95 cm/sec
(3) relative collection efficiency

APPENDIX A3

OVERALL THERMAL BALANCE FOR THE FLUIDIZED BED SYSTEM-TAILINGS REFLOAT PRODUCT

APPENDIX A3

OVERALL THERMAL BALANCE FOR THE FLUIDIZED BED SYSTEM-TAILINGS REFLOAT PRODUCT

A3.1 OPERATIONS CONDITIONS

Feed rate = 37.50 TPH (dry base)
Temperature = 685°C
Air factor = 1.80

A3.2 FEED AND CALCINES COMPOSITION

321 Feed Composition

900.23 TPD (dry base)
(Zn,Fe)S = 456.23 TPD = 4.7421×10^6 (moles/day)
$CuFeS_2$ = 15.60 " = 8.35×10^4 "
PbS = 37.43 " = 1.561×10^5 "
FeS_2 = 352.16 " = 2.9350×10^6 "

322 Calcine Composition

1191.63 TPD:
$ZnSO_4$ = 565.09 TPD = 3.5010×10^6 (moles/day)
$ZnO.2ZnSO_4$ = 106.06 " = 2.624×10^5 "
$ZnO.Fe_2O_3$ = 129.50 " = 5.372×10^5 "
(Zn,Fe)S = 10.63 " = 1.105×10^5 "
$PbSO_4$ = 47.20 " = 1.559×10^5 "
$CuSO_4$ = 15.36 " = 0.962×10^5 "
Fe_2O_3 = 274.60 " = 1.7195×10^6 "

A3.3 AIR REQUIREMENT *

Oxygen required to form desired composition according to:

$$(0.8712\ ZnS + 0.1288\ FeS) + 1.968O_2 \rightarrow$$
$$0.8712\ ZnSO_4 + 0.0644\ Fe_2O_3 + 0.1288\ SO_2$$

$$CuFeS_2 + 3.75O_2 \rightarrow CuSO_4 + 0.5\ Fe_2O_3 + SO_2$$

---

* Assuming that marmatitic sphalerite has a Zn:Fe ratio of 0.8712:0.1288

$PbS + 2O_2 \rightarrow PbSO_4$

$FeS_2 + 2.75O_2 \rightarrow 0.5\ Fe_2O_3 + 2SO_2$

$$\dot{n}O_2 = 0.0173\ (3.76/183.51) + 0.5068\ (1.968/96.207) + 0.0416\ (2.0/239.25) + 0.3912\ (2.75/119.97)$$
$$= 2.003 \times 10^{-2}\ (\text{moles } O_2/\text{gr concentrate})$$

$$q = 2.130 \times 10^{3}\ (m^{3}\ \text{air/Ton conc.})\ (\text{st. cond.})$$

For an air factor of Q = 1.8 (80% excess air into the fluidized bed),

$$q = 3.834 \times 10^{3}\ (m^{3}\ \text{air/Ton conc.})$$

or

$$Q = 3.834 \times 10^{3} \times 9.0023 \times 10^{2}$$
$$= 3.451 \times 10^{6}\ (m^{3}/\text{day})\ (\text{st. cond.})$$

A3.4 HEAT GENERATED AND CONSUMED

341 Heat Generated by Reactions

The values of Table A3-1 have been used to calculate the heat of reactions at 685°C (958°K).

TABLE A3-1

STANDARD HEATS OF FORMATION AND HEAT CAPACITIES

Compound	$\Delta H°_{298°K}$ (K cal/mole)	C_p (cal/mole-°K)
ZnS	- 48.2	$12.16 + 1.24 \times 10^{-3}\ T$
$ZnSO_4$	-234.9	$21.90 + 18.2 \times 10^{-3}\ T$
$ZnO.2ZnSO_4$	-551.3	75.5
$ZnO.Fe_2O_3$	-279.5	$34.3 + 13.3 \times 10^{-3}\ T$
FeS	- 22.8	$12.20 + 2.38 \times 10^{-3}\ T$
FeS_2	- 42.4	$17.88 + 1.32 \times 10^{-3}\ T$
Fe_2O_3	-196.3	$23.49 + 16.6 \times 10^{-3}\ T$
$CuFeS_2$	- 76.2	$22.89 + 40.33 \times 10^{-3}\ T$

TABLE A3-1 (Cont'd)

Compound	$\Delta H^\circ_{298^\circ K}$ (K cal/mole)	C_p (cal/mole-°K)
$CuSO_4$	-184.0	$18.77 + 17.20 \times 10^{-3}$ T
PbS	- 22.5	$10.66 + 3.92 \times 10^{-3}$ T
$PbSO_4$	-219.5	$10.96 + 31.0 \times 10^{-3}$ T
SO_2 (g)	- 71.0	$10.38 + 2.54 \times 10^{-3}$ T
O_2	-	$7.16 + 1.0 \times 10^{-3}$ T
H_2O (g)	- 68.3	$7.30 + 2.46 \times 10^{-3}$ T

.1 Sphalerite

To form $ZnSO_4$:

$(0.8712\ ZnS + 0.1288\ FeS) + 1.968\ O_2 \rightarrow$

$0.8712\ ZnSO_4 + 0.0644\ Fe_2O_3 + 0.1288\ SO_2$ ---------- (1-a)

Heat of reaction at 958°K calculated according:

$$\Delta H^\circ_{298^\circ K} = -(0.8712 \times 234.9 + 0.0644 \times 196.3 + 0.1288 \times 71) + (0.8712 \times 48.2 + 0.1288 \times 22.8)$$

$$\Delta H^\circ_{298^\circ K} = -180.63 \text{ (K cal/mole (Zn,Fe)S)}$$

(assuming no heat of solution between ZnS and FeS in the marmatitic sphalerite)

$$\Delta C_p = (19.08 + 15.86 \times 10^{-3} T) + (1.51 + 1.07 \times 10^{-3} T) + (1.34 + 0.33 \times 10^{-3} T) - (10.59 + 1.08 \times 10^{-3} T) - (1.57 + 0.31 \times 10^{-3} T) - (14.09 + 1.97 \times 10^{-3} T)$$

$$\Delta C_p = -4.32 + 13.9 \times 10^{-3} T \text{ (cal/mole (Zn,Fe)S)}$$

$$\Delta H^\circ_T = \int(-4.32 + 13.9 \times 10^{-3} T)dT$$
$$= -4.32T + 6.95 \times 10^{-3} T^2 + \Delta H_o$$

$$\Delta H_o = -180{,}630 + 4.32 \times 298 - 6.95 \times 10^{-3} \times \overline{298}^2$$
$$= -179{,}960 \text{ cal/mole (Zn,Fe)S}$$

$\Delta H^\circ_T = -179,960 - 4.32T + 6.95 \times 10^{-3}\ T^2$

Therefore, at 958°K:

$\Delta H^\circ_{958^\circ K} = -177.72$ (K cal/mole (Zn,Fe)S)

or

$\Delta H^\circ_{958^\circ K} = -203.99$ (Kcal/mole $ZnSO_4$)

To form $ZnO.ZnSO_4$:
$(0.8712\ ZnS + 0.1288\ FeS) + 1.532\ O_2 \rightarrow$
$0.2904\ (ZnO.2ZnSO_4) + 0.0644\ Fe_2O_3 + 0.1288\ SO_2$ ---------- (1-b)

$\Delta H^\circ_{298^\circ K} = -136.96$ (K cal/mole (zn,Fe)S)
$\Delta C_p = 1.66 - 1.52 \times 10^{-3}\ T$ (cal/mole °K)
$\Delta H_o = -137,352$ (cal)

$\Delta H^\circ_T = -137,352 + 1.66T - 7.62 \times 10^{-3}\ T^2$ (cal/mole (Zn,FeS))

$\Delta H^\circ_{958^\circ K} = -136.46$ (K cal/mole (Zn,Fe)S)

or

$\Delta H^\circ_{958^\circ K} = -469.90$ (K cal/mole $ZnO.2ZnSO_4$)

To form $ZnO.Fe_2O_3$:
$ZnO.2ZnSO_4 + Fe_2O_3 \rightarrow ZnO.Fe_2O_3 + 2ZnSO_4$ ------------------- (1-c)

$\Delta H^\circ_{298^\circ K} = -1.71$ (K cal/mole $ZnO.2ZnSO_4$)
$\Delta C_p = -20.9 + 33.1 \times 10^{-3}\ T$ (cal/mole-°K)
$\Delta H_o = +3048$ (cal)

$\Delta H^\circ_T = +3048 - 20.9T + 16.55 \times 10^{-3}\ T^2$ (cal/mole $ZnO.2ZnSO_4$)
$\Delta H^\circ_{958^\circ K} = -1.79$ (K cal/mole $ZnO.2ZnSO_4$ or $ZnO.Fe_2O_3$)

.2 Chalcopyrite

$CuFeS_2 + 3.75\ O_2 \rightarrow CuSO_4 + 0.5\ Fe_2O_3 + SO_2$ ---------- (2)

$\Delta H°_{960°K}$ = -309.21 (K cal/mole $CuFeS_2$ or $CuSO_4$)
(Incra Tables)

.3 Galena

$PbS + 2O_2 \rightarrow PbSO_4$ ---------------------------------- (3)

$\Delta H°_{298°K}$ = -197.0 (K cal/mole PbS)
ΔC_p = $-14.02 + 25.08 \times 10^{-3}$ T (cal/mole °K)
ΔH_o = -156,390 (cal)

$\Delta H°_T$ = $-156{,}390 - 14.02T + 12.5 \times 10^{-3}\ T^2$ (cal/mole PbS)

$\Delta H°_{958°K}$ = -158.35 (K cal/mole PbS or $PbSO_4$)

.4 Pyrite

$FeS_2 + 2.75O_2 \rightarrow 0.5\ Fe_2O_3 + 2SO_2$ -------------------- (4)

$\Delta H°_{298°K}$ = -197.90 (K cal/mole FeS_2)
ΔC_p = $-5.06 + 9.31 \times 10^{-3}$ T (cal/mole FeS_2)
ΔH_o = -196,706 (cal)

$\Delta H°_T$ = $-196{,}706 - 5.06T + 4.66 \times 10^{-3}\ T^2$ (cal/mole FeS_2)

$\Delta H°_{958°K}$ = -197.29 (K cal)

.5 Total Heat Generated

.51 From Sphalerite

Total $ZnSO_4$ in calcines = 3.5010×10^6 (moles/day)
Total $ZnSO_4$ produced by reaction (1-c)= 5.372×10^5 (moles/day)
Net $ZnSO_4$ produced by reaction (1-a)= 2.427×10^6 (moles/day)

$$\Delta H_1 = -203.99 \times 2.427 \times 10^6 = -4.9508 \times 10^8 \text{ (K cal/day)}$$

According to reaction (1-b) to form the total $ZnO.2ZnSO_4$ present in the calcines plus the $ZnO.2ZnSO_4$ to form the $ZnO.\ Fe_2O_3$ present in the calcines:

$ZnO.2ZnSO_4$ in calcines = 2.6238×10^5 (moles/day)
$ZnO.2ZnSO_4$ to form $ZnO.Fe_2O_3$ = 5.3719×10^5 (moles/day)
Total $ZnO.2ZnSO_4$ formed = 7.9957×10^5 (moles/day)

$$\Delta H_2 = -469.90 \times 7.9957 \times 10^5 = -3.7572 \times 10^8 \text{ (K cal/day)}$$

According to reaction (1-c) to form the total $ZnO.Fe_2O_3$ present in the calcines:

$$\Delta H_3 = -1.79 \times 5.372 \times 10^5 = -9.62 \times 10^5 \text{ (K cal/day)}$$

.52 Chalcopyrite

According to raction (2), for the total $CuFeS_2$ present in the feed:

$$\Delta H_4 = -309.21 \times 8.35 \times 10^4 = -2.582 \times 10^7 \text{ (K cal/day)}$$

.53 Galena

According to reaction (3), for the total PbS present in the feed:

$\Delta H_5 = -158.35 \times 1.56 \times 10^5 = -2.470 \times 10^7$ (K cal/day)

.54 Pyrite

According to reaction (4), for the total FeS_2 present in the feed:

$\Delta H_6 = -197.29 \times 2.935 \times 10^6 = -5.7905 \times 10^8$ (K cal/day)

(the total Fe_2O_3 present in the calcine is formed by oxidation of the FeS_2, $CuFeS_2$ and remnant FeS from the sphalerite).

.55 Total Heat Generated

$$\Delta H_R = \sum_{i=1}^{6} \Delta H_i = -1.5014 \times 10^9 \text{ (K cal/day)}$$

342 Heat Consumed in the System

.1 Heat Losses

The reactors are insulated to achieve 5% (as maximum) of heat losses (from the total heat generated by the reactions). Therefore,

$\Delta H_7 = +7.507 \times 10^7$ (K cal/day)

.2 Preheat the Fluidizing Air

$Q = 3.451 \times 10^6$ (m^3 air/day) (air blasted at 60°C after compression)
$= 3.234 \times 10^7$ (moles O_2/day)
$= 1.2171 \times 10^8$ (moles N_2/day)

Oxygen:
$H_T - H_{298°K} = -2313 + 7.16T + 0.50 \times 10^{-3} T^2$ at 958°K
$H_{958} - H_{333°K} = +4605$ (cal/mole O_2)

Total to heat $O_2 = 1.489 \times 10^8$ (K cal/day)

Nitrogen:

$H_T - H_{298°K} = -2031 + 6.66T + 0.51 \times 10^{-3} T^2$ at 958°K

$H_{958} - H_{333°K} = +4395$ (cal/mole N_2)

Total to heat $N_2 = 5.345 \times 10^8$ (K cal/mole)

Total heat required to preheat the air:

$\Delta H_8 = +6.8346 \times 10^8$ (K cal/day)

.3 Preheat the Feed

Entalphies calculated according to:

ZnS:

$$H_{958} - H_{298°K} = \int_{298}^{958} (12.16T + 0.62 \times 10^{-3} T^2)\, dT$$

$= +8.54$ (K cal/mole)

Total heat $= +8.54 \times 4.1313 \times 10^6 = +3.527 \times 10^7$ (K cal/day)

FeS:

$H_{958} - H_{298°K} = +9.04$ (K cal/mole)

Total heat $= +9.04 \times 6.1078 \times 10^5 = +5.521 \times 10^6$ (K cal/day)

$CuFeS_2$:

$H_{958} - H_{298°K} = +33.1$ (K cal/mole)

Total heat $= +33.1 \times 8.35 \times 10^4 = 2.764 \times 10^6$ (K cal/day)

PbS:

$H_{958} - H_{298°K} = +8.66$ (K cal/mole)

Total heat $= +8.66 \times 1.561 \times 10^5 = +1.352 \times 10^6$ (K cal/day)

FeS_2:

$H_{958} - H_{298°K} = +12.35$ (K cal/mole)

Total Heat $= +12.35 \times 2.9350 \times 10^6 = +3.6247 \times 10^7$ (K cal/day)

Total heat necessary to preheat the feed:

ΔH_9 = +8.116 x 10^7 (K cal/day)

.4 Reheat Recirculated Dust

Assuming, for simplification(*), that dust contains only $ZnSO_4$, Fe_2O_3, $PbSO_4$ and $CuSO_4$. Dust return to reactor at 600°C.

The dust elutriation flux from the reactors is:

K* = 1.038 x 10^{-4} (T/M^2 x sec) (685°C, u_o' = 75 cm/sec)
k* = 980.77 TPD dust recirculated

$ZnSO_4$:
$H_{958} - H_{873°K}$ = +3.28 (K cal/mole)
Total heat = +3.28 x 5.026 x 10^6 = +1.649 x 10^7 (K cal/day)

Fe_2O_3:
$H_{958} - H_{873°K}$ = +0.85 (K cal/mole)
Total heat = +0.85 x 1.7195 x 10^6 = +1.462 x 10^6 (K cal/day)

$PbSO_4$:
$H_{958} - H_{873°K}$ = +3.27 (K cal/mole)
Total heat = +3.27 x 1.559 x 10^5 = +5.098 x 10^5 (K cal/day)

$CuSO_4$:
$H_{958} - H_{873°K}$ = +3.23 (K cal/mole)
Total heat = +3.23 x 0.962 x 10^5 = +3.107 x 10^5 (K cal/day)

Total heat required to reheat the dust:

ΔH_{10} = +1.877 x 10^7 (K cal/day)

(*) The total heat required to reheat the dust is very small compared with the total heat consumed in the process.

.5 Vaporize and Decompose HAL Solution

Values for the heat of formation and heat capacities are given in Table A3-2.

TABLE A3-2

STANDARD HEAT OF FORMATION AND HEAT CAPACITIES OF HAL SOLUTION

Compound	$\Delta H^\circ_{298^\circ K}$ (K cal/mole)	C_p (cal/mole-°K)
Fe_2O_3	-196.3	$23.49 + 16.6 \times 10^{-3}\ T$
$FeSO_4$	-221.3	22.0
$Fe_2(SO_4)_3$	-617.0	66.2
$CuSO_4$	-184.0	$18.77 + 17.20 \times 10^{-3}\ T$
$ZnSO_4$	-234.9	$21.90 + 18.2 \times 10^{-3}\ T$
SO_2 (g)	- 71.0	$10.38 + 2.54 \times 10^{-3}\ T$
H_2SO_4	-194.6	33.2
O_2	-	$7.16 + 1.0 \times 10^{-3}\ T$
H_2O (g)	- 68.3	$7.30 + 2.46 \times 10^{-3}\ T$

HAL solution composition (main constituents):

$ZnSO_4$ = 112.78 TPD = 6.986×10^5 (moles/day)
$FeSO_4$ = 18.74 " = 1.234×10^5 "
$Fe_2(SO_4)_3$ = 177.71 " = 4.444×10^5 "
H_2SO_4 = 21.06 " = 2.148×10^5 "
H_2O = 472.76 " = 2.626×10^7 "

.51 Preheat Dissolved Solids and H_2SO_4 in HAL Solution

HAL solution is recirculated back into the roaster at 90°C

$$ZnSO_4 = 21.9\ (958 - 363) + 9.1 \times 10^{-3}\ (\overline{958}^2 - \overline{363}^2)$$
$$= 20.18 \text{ (K cal/mole)}$$

$FeSO_4$ = 22.0 (958 - 363) = 13.09 (K cal/mole)

$Fe_2(SO_4)_3$ = 66.2 (958 - 363) = 39.39 (K cal/mole)

H_2SO_4 = 33.2 (958 - 363) = 12.05 (K cal/mole)

Therefore,

$ZnSO_4$ = 20.18 x 6.986 x 10^5 = 1.410 x 10^7 (K cal/day)
$FeSO_4$ = 13.09 x 1.234 x 10^5 = 1.615 x 10^6 "
$Fe_2(SO_4)_3$ = 39.39 x 4.444x 10^5 = 1.750 x 10^7 "
H_2SO_4 = 12.05 x 2.148 x 10^5 = 2.588 x 10^6 "

Total to preheat dissolved solids in HAL solution:

$\Delta H_{11} = +3.581 \times 10^7$ (K cal/day)

.52 Decompose HAL Solution (no water)

$ZnSO_4$ remain as such.

$FeSO_4$:
$FeSO_4 \underset{\Delta}{\rightarrow} 0.5\ Fe_2O_3 + SO_2 + 0.25\ O_2$

$\Delta H°_{298°K}$ = +52.15 (K cal)
ΔC_p = 3.71 + 11.34 x 10^{-3} T (cal/mole °K)
ΔH_o = +50,541 (cal)

$\Delta H°_T$ = 50,541 + 3.71T + 5.67 x 10^{-7} (K cal)
$\Delta H°_{958°K}$ = +59.30 (K cal)

$Fe_2(SO_4)_3$:
$Fe_2(SO_4)_3 \underset{\Delta}{\rightarrow} Fe_2O_3 + 3SO_2 + 1.5O_2$

$\Delta H°_{298°K}$ = +207.7 (K cal)
ΔC_p = -0.83 + 25.7 x 10^{-3} T (cal/mole °K)
ΔH_o = +206,801 (cal)

$\Delta H°_T = +206{,}801 - 0.83T + 12.9 \times 10^{-3}\ T$ (K cal)
$\Delta H°_{958°K} = +217.85$ (K cal)

H_2SO_4:
$H_2SO_4 \underset{\Delta}{\rightarrow} H_2O + SO_2 + 0.5O_2$

$\Delta H°_{298°K} = +55.3$ (K cal)
$\Delta C_p = -12.07 + 5.6 \times 10^{-3}\ T$ (cal/mole °K)
$\Delta H_o = +58{,}648$ (cal)

$\Delta H°_T = +58{,}648 - 12.07T + 2.8 \times 10^{-3}\ T^2$ (K cal)
$\Delta H°_{958°K} = +49.66$ (K cal)

Therefore heat to decompose each component is:

$FeSO_4 = 59.30 \times 1.234 \times 10^5 = +7.318 \times 10^6$ (K cal/day)
$Fe_2(SO_4)_3 = 217.85 \times 4.444 \times 10^5 = 9.6810 \times 10^7$ "
$H_2SO_4 = 49.66 \times 2.148 \times 10^5 = +1.067 \times 10^7$ "

Total to decompose solids and H_2SO_4 in HAL solution:

$\Delta H_{12} = +1.1480 \times 10^8$ (K cal/day)

A3.5 OVERALL HEAT BALANCE

.51 Heat Generated (ΔH_R)

By chemical reactions:

$\Delta H_R = -1.5014 \times 10^9$ (K cal/day)

.52 Heat Consumed (ΔH_C)

By heat losses through reactor shell:

$\Delta H_7 = +7.507 \times 10^7$ (K cal/day)

To preheat the fluidizing air:

$\Delta H_8 = +6.8346 \times 10^8$ (K cal/day)

To preheat the feed:

$\Delta H_9 = +8.116 \times 10^7$ (K cal/day)

To reheat recirculated dust:

$\Delta H_{10} = +1.877 \times 10^7$ (K cal/day)

To preheat HAL solution:

$\Delta H_{11} = +3.581 \times 10^7$ (Kcal/day)

To decompose HAL solution:

$\Delta H_{12} = +1.1480 \times 10^8$ (K cal/day)

.53 Excess Heat to Vaporize H_2O

$\Delta H_E = \Delta H_R - \Delta H_C = -4.923 \times 10^8$ (K cal/day)

Water to vaporize:

$H_2O = (-4.923 \times 10^8/-16.326) \times 18/10^6$

$H_2O = 542.8$ TPD

(total H_2O in HAL solution from tailings refloat products is 472.76).

Therefore, the reactor has an extra capacity to consume 70.1 TPD equivalent water.

.54 Spent Additional to Consume

Assuming feed contains 2.5% moisture, heat to vaporize this water is:

ΔH moisture = $(22.51 \times 10^6/18) \times 16.326$

= $+ 2.04 \times 10^7$ (K cal/day)

(The total net heat surplus allows to use a feed with up to 7.8% moisture).

Net surplus after vaporizing 2.5% moisture:

ΔH surplus = $+4.349 \times 10^7$ (K cal/day)

Spent electrolyte of 60 gpl Zn, 165 gpl H_2SO_4 consumes 9.84×10^5 (K cal/Tm spent).

Total spent additional = $4.349 \times 10^7/9.84 \times 10^5$

= 44.20 TPD

The total consumption of heat by the HAL solution and spent electrolyte represents the net amount of heat to be eliminated from the fluidized bed alone, not considering heat recovered from the hot gases above the fluidized bed level. In practice, since the spray gun for the HAL solution and spent are located above the bed level, some heat is recovered from the hot gases. This will increase the overall capacity of the reactor for consuming HAL solution and spent, as is discussed in detail in Appendix A1.

A3.6 SOURCES OF THERMODYNAMICAL DATA

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APPENDIX A4

DETERMINATION OF MINERALOGY OF ROASTER CALCINES AND LEACH RESIDUES

APPENDIX A4

DETERMINATION OF MINERALOGY OF ROASTER CALCINES AND LEACH RESIDUES

A4.1 QUANTITATIVE DETERMINATION

The roaster calcines were routinely leached in both water and dilute acid solutions. Acid leaching in 25 gpl sulphuric acid solution removes zinc and copper sulphates ($ZnSO_4$ and $CuSO_4$) and oxysulphates ($ZnO.2ZnSO_4$ and $CuO.CuSO_4$). Zinc in the leach residue is present only as remnant sphalerite (Zn,Fe)S and ferrite ($ZnO.Fe_2O_3$) both of which are relatively insoluble in dilute acid. The relative proportion of each compound was determined from chemical analyses (Zn,Pb,S) where available. Zinc sulphate completely dissociates in water at 25°C whereas zinc oxysulphate is only partially soluble (42.2% from x-ray examination of synthetic $ZnO.2ZnSO_4$ leach ppts).

Hence the wt % of each zinc compound in calcines is determined as follows:

$$\text{wt\% ZnO.2ZnSO}_4 = \frac{\text{wt\% Zn(acid leach soln) - wt\% Zn(water leach soln)}}{0.485 \times 0.578}$$

$$\text{wt\% ZnSO}_4 = \frac{\text{wt\% Zn(acid leach soln) - wt\% Zn(ZnO.2ZnSO}_4)}{0.405}$$

$$\text{wt\%(Zn,Fe)S} = \frac{\text{wt\%S - wt\% S(PbSO}_4)}{.333}$$

$$\text{wt\% ZnO.Fe}_2\text{O}_3 = \frac{\text{wt\% Zn(acid leach residue)} - \text{wt\% Zn(sphalerite)}}{.2712}$$

where the factors 0.485, 0.405, 0.2712 and 0.592 are the stoichiometric %Zn in each phase and 0.333 is estimated %S in sphalerite.

.2 QUALITATIVE DETERMINATIONS

X-ray diffractometer scans at 1 °/min from 28.5° to 43.0° 2 Θ were routinely prepared from all roaster calcines to assess the relative amounts of various reaction products present.

.3 SEMI-QUANTITATIVE DETERMINATIONS

The 25 gpl acid leach residues from all runs were ground slightly to eliminate any agglomeration effects and the powders pressed into aluminum holders such that the powder presented a uniform, compact, flat surface in the plane of the x-ray beam. Diffractometer scans were run from 40.5° to 42.5° 2 Θ at 1°/min and the relative peak intensities of the most intense hkl reflection from ferrite, the 311 reflection were recorded.

All wt% ferrite determinations calculated from chemical analyses (Tables A1-4a-c and A1-10} can be found to conform to reasonable linear relationships of peak intensity to wt % (Figure A4-1). The method of least squares was used to fit the best curve ($y = .789x$) to the experimental data. The wt% ferrite could thus be determined using the 311 ferrite

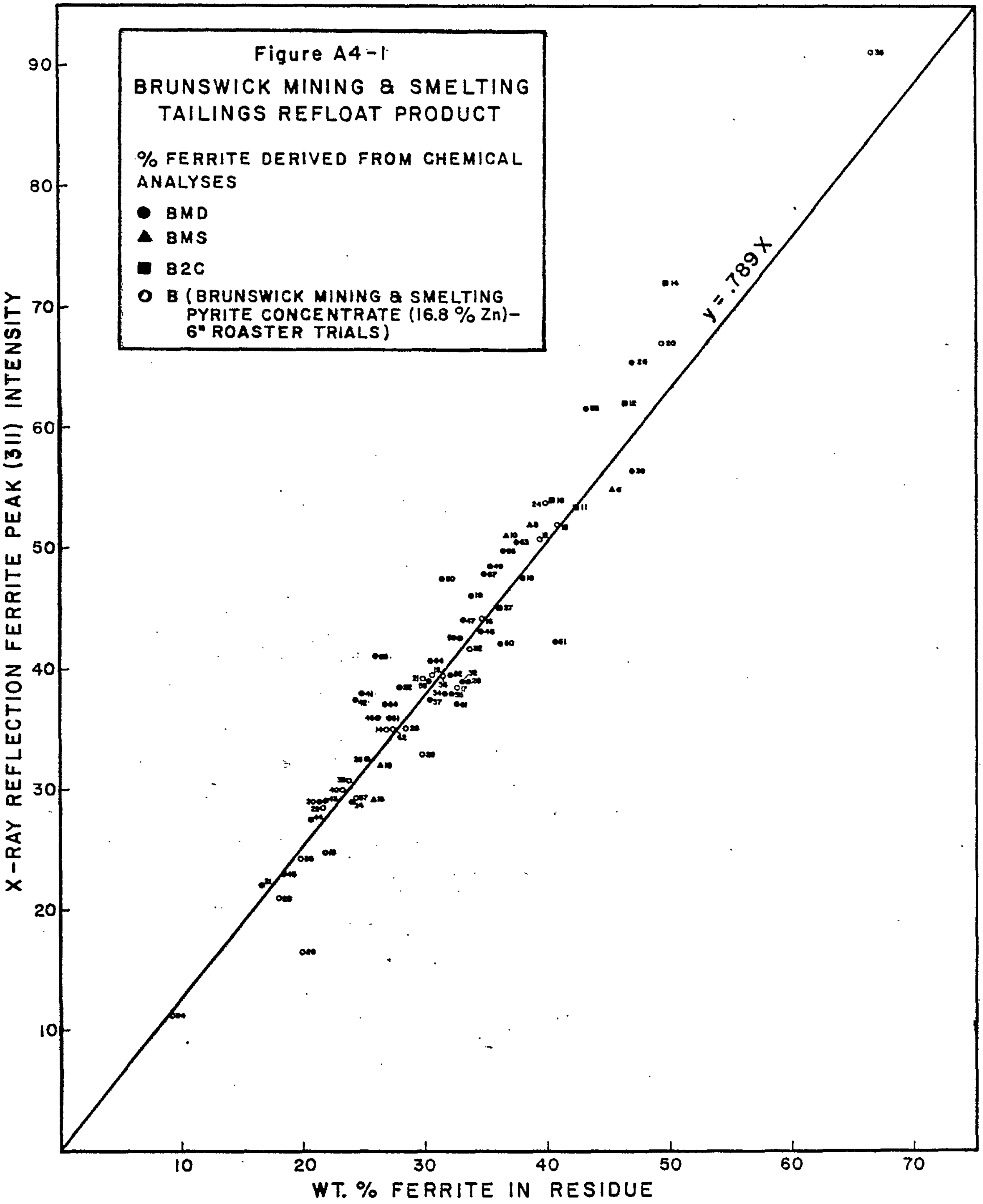


Figure A4-1
BRUNSWICK MINING & SMELTING TAILINGS REFLOAT PRODUCT

peak intensities for calcines where complete chemical analyses were not available and the wt% sphalerite calculated by difference from the total zinc analyses.

APPENDIX A5

ALTERNATIVES TO LIMESTONE NEUTRALIZATION OF SPENT ZINC ELECTROLYTE

APPENDIX A5

ALTERNATIVES TO LIMESTONE NEUTRALIZATION OF SPENT ZINC ELECTROLYTE

A5.1 LEAD SLAG LEACH-SAND BED DECOMPOSITION OF FILTRATE

An alternative to limestone as the neutralizing agent for excess spent electrolyte to maintain sulphate balance exists in lead smelter slag, such as is produced at the Belledune Smelter of Brunswick Mining and Smelting. Approximately 140,000 tons of lead slag of average chemical analysis as shown in Table A5-1 is added each year to an already substantial stockpile of slag at Belledune. Limited microscope and electron microprobe studies indicate that 15% of the zinc is intimately associated with wustite either as zinc oxide solid solution or zinc ferrite inclusions. The wustite grains, which account for 15% of the mass of the slag, are included in a complex amorphous Ca-Zn-Fe silicate glass which contains 85% of the zinc.

TABLE A5-1
AVERAGE CHEMICAL ANALYSIS OF LEAD SMELTER SLAGS
(wt%; Ag in oz/ton)

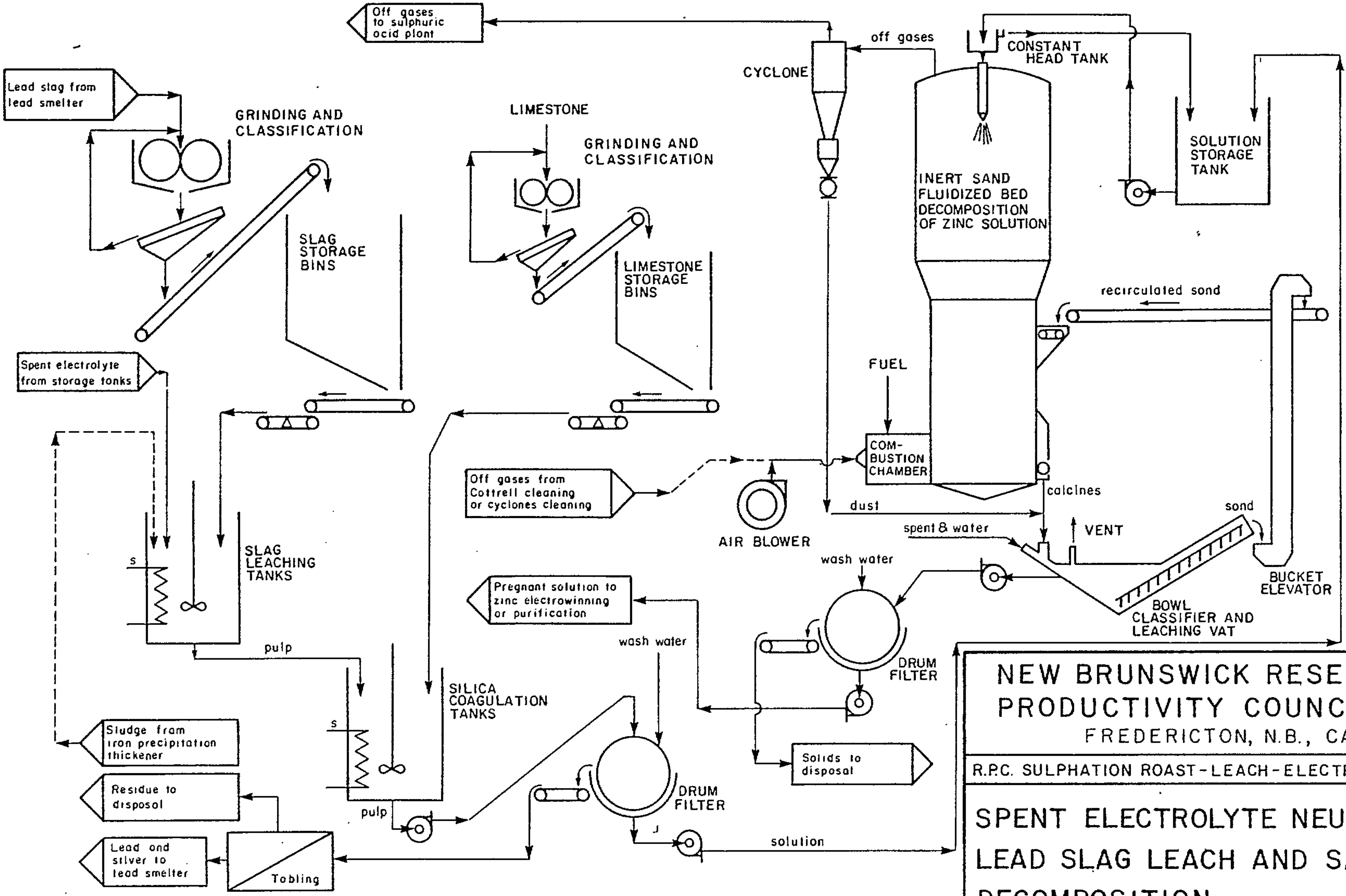
Zn	Pb	SiO_2	FeO	Al_2O_3	S	CaO	Ag
10.5	2.05	21.7	38.9	4.3	2.6	13.4	0.5

The stockpiled slag which, despite appreciable levels of zinc, is not as suitable for fuming as fresh slag due to the energy cost in reheating this material. Subsequently, the zinc in this material is lost and may never be recovered by the lead smelter.

The advantages in using this material as a neutralization agent for the sulphation roast plant spent zinc electrolyte are threefold:

1. approximately 70-80 per cent of the limestone requirement for the plant would be eliminated.
2. a significant fraction of the zinc, lead, and silver in the slag would be recovered. The throughput of zinc in the plant would be correspondingly increased.
3. the overall efficiency of the sulphation roast plant would be improved in that the MnO_2-$Fe(OH)_3$-gypsum sludge from the iron oxidation stage could be added to the spent electrolyte prior to neutralization with the lead slag. The present general plant flowsheet requires this sludge to be recycled to the neutral leach. Handling of the sludge in the slag neutralization circuit would considerably increase the settling rate in the neutral leach thickener (approximately 50%) as well as increase the filtration rate by a factor of at least four. This would effectively simplify the filtering operation in the plant and reduce capital costs significantly. An additional advantage would be the considerable reduction of gypsum content in the hot acid leach residue, which would result in a reduction in filter cake moisture content, ease in washing and generally less bulk to carryover to the brine leach.

A proposed flowsheet of a spent electrolyte neutralization circuit with lead slag is shown in Drawing A5-1. The slag after grinding to -100 mesh is contacted with the spent electrolyte at 50°C for 2-3 hours in order to decompose the zinc, calcium, and iron silicates and consume most of the sulphuric acid present in the spent. The pulp is then treated to coagulate the silica gel formed, as in the RTZ Process (1) developed recently in Australia. The coagulated pulp is then filtered and washed, and the residue further processed by gravity concentration to recover the lead and silver present as small metallic shots.



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R.P.C. SULPHATION ROAST-LEACH-ELECTROWINNING PROCESS

SPENT ELECTROLYTE NEUTRALIZATION-LEAD SLAG LEACH AND SAND BED DECOMPOSITION

APPENDIX 5.1

DATE: 14 APRIL, 1978

The solution, containing the zinc from the spent and most of the zinc leached from the slag, is further processed by thermal decomposition in an inert bed-fluidized bed reactor. The bed is composed of a silica sand-type material, fluidized with air or the hot gases from the roasters.

The reactor which operates at 650-700°C, decomposes the iron sulphate and sulphuric acid residual present in the solution. Zinc is recovered as sulphate in the form of small pellets or coating on the sand.

The bed overflow is quenched and leached directly in water or water and spent, and the sand classified for continuous recycle to the reactor. The classifier overflow is then filtered and the solution is directed either to the leaching section or to the purification circuit, if the iron levels are low enough, while the residue from the filter is discarded.

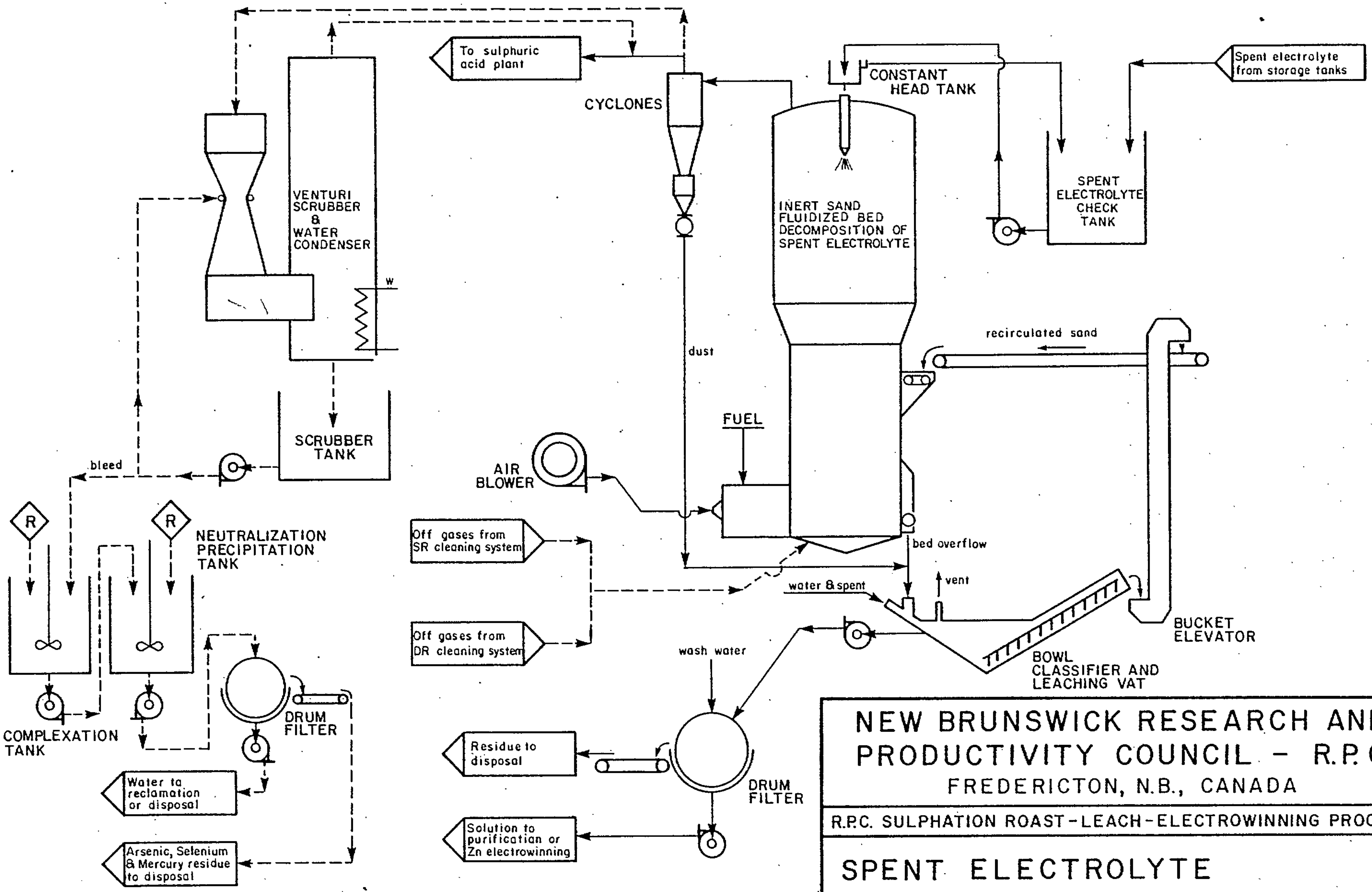
Since in most cases, the thermal energy recovered from the hot gases from the roasters is not enough to vaporize and decompose all of the solution from the slag leach circuit, additional fuel must be added to the sand bed reactor in the form of fuel oil, natural gas, or pulverized coal.

An alternative to the silica coagulation and filtering may exist if conditions can be found to minimize the polymerization of silicic acid in the slag leach. According to RTZ (1), proper control of pH, temperature, and retention time produce a leach pulp from a silicate ore from which the leach residue may be separated by thickening prior to coagulation of colloidal silica. This residue would then be either filtered or counter-current washed prior to treatment for lead and silver recovery. The decant solution from the leach, containing colloidal silica would be treated in the sand bed decomposing the silicic acid to silica which remains inert in the sand leach and reports to the leach residue.

Several preliminary leach trials have been carried out to determine the neutralization capacity and zinc extraction from the slag. The neutralization capacity is approximately 2 m^3 spent electrolyte per ton of slag, against 5 m^3 spent per ton of limestone. Also, 38% of the zinc was recovered from the slag without attempting to recover the zinc associated with colloidal and polymeric silica. These are very encouraging results and it appears from the mineralogy of the slag, that very high zinc extractions may be achieved using an optimized system.

A5.2 SPENT ELECTROLYTE DECOMPOSITION IN AN INERT FLUIDIZED SAND BED

As an alternative to neutralization, spent electrolyte from the zinc electrowinning cells can be decomposed in an inert fluidized bed reactor consisting of silica-type sand particles fluidized with hot gases from the sulphation and/or dead roast operations. The bed not only acts as a trap for zinc sulphate in the spent but also acts as a medium for cleaning fine particles from the roaster off-gases. If necessary, bed temperature may be maintained at a suitable sulphation temperature by burning oil, gas or pulverized coal. The schematic flowsheet is shown in Drawing A5-2.



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R.P.C. SULPHATION ROAST-LEACH-ELECTROWINNING PROCESS

SPENT ELECTROLYTE DECOMPOSITION IN A FLUIDIZED SAND BED

APPENDIX 5·2

DATE: 14 APRIL, 1978

The bed overflow, containing the sand and small pellets of zinc sulphate, including small particles of roaster off-gas solid carryover, can be leached directly in water or water plus spent electrolyte in a bowl-type classifier which continuously recycles the sand back into the reactor. The H_2SO_4 in the spent electrolyte decomposes in the reactor to increase the strength of the SO_2 from the sulphation and/or dead roaster off-gases. The leach solution is sent directly to the neutral leach.

If gas or fuel oil combustion is used to preheat the air for bed fluidization, instead of roaster off-gas, then a clean zinc sulphate product is produced which may be sent directly to electrowinning. The reactor temperature may also be decreased to 250-350°C, sufficient to decompose sulphuric acid and vaporize water, since no iron is present to resulphate.

A5.3 REFERENCES

1. Matthew, I. G. and Elsner, D.; The Hydrometallurgical Treatment of Zinc Silicate Ores, March Met. Trans. B, 8B, pp. 73-78 (1977) and Matthew, I. G. and Elsnew, D.; The Processing of Zinc Silicate Ores - A Review, March Met. Trans. B 8B, pp. 85-91. (1977)

APPENDIX A6

CALIBRATIONS

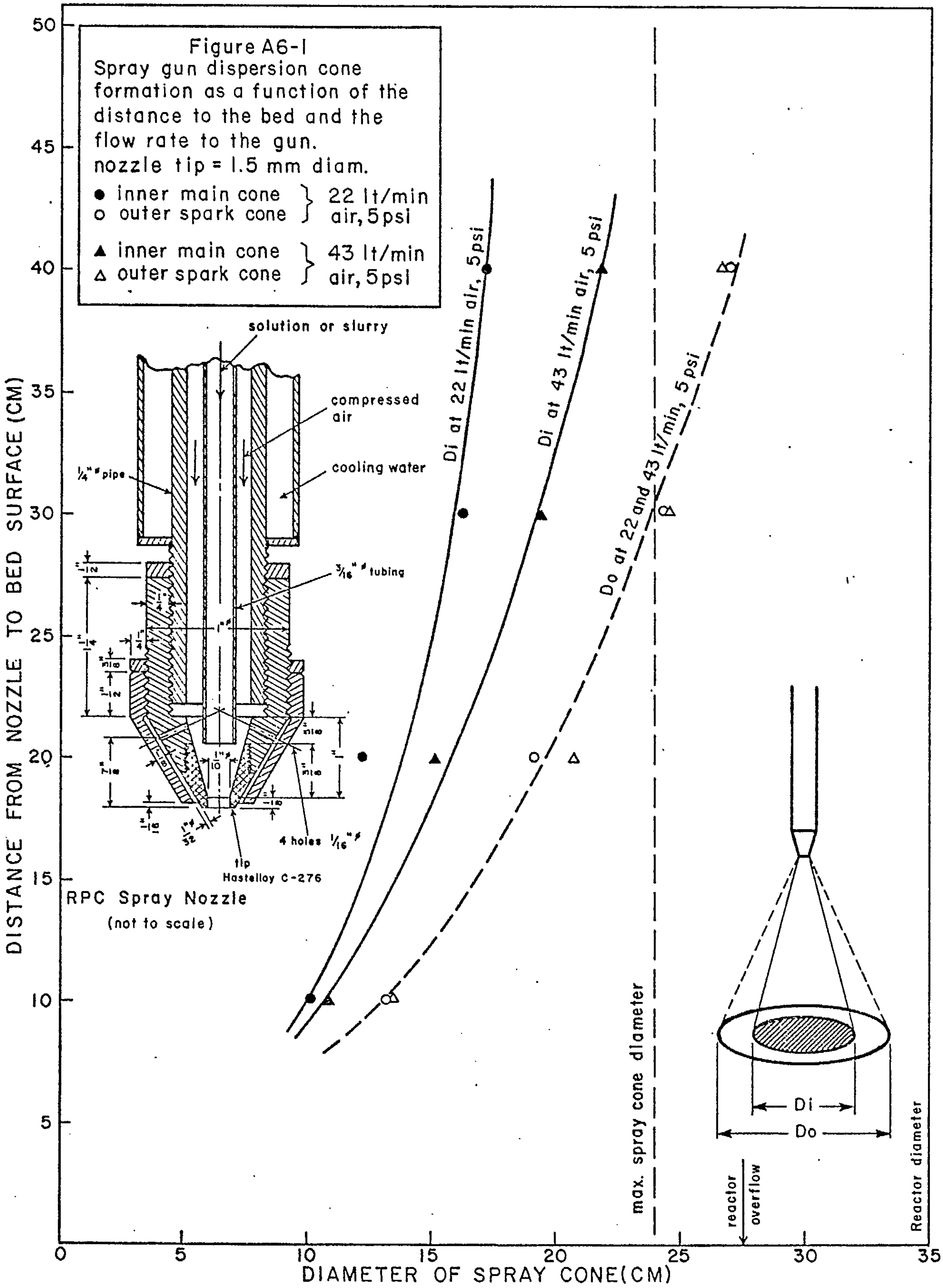

Figure A6-1
Spray gun dispersion cone formation as a function of the distance to the bed and the flow rate to the gun.
nozzle tip = 1.5 mm diam.

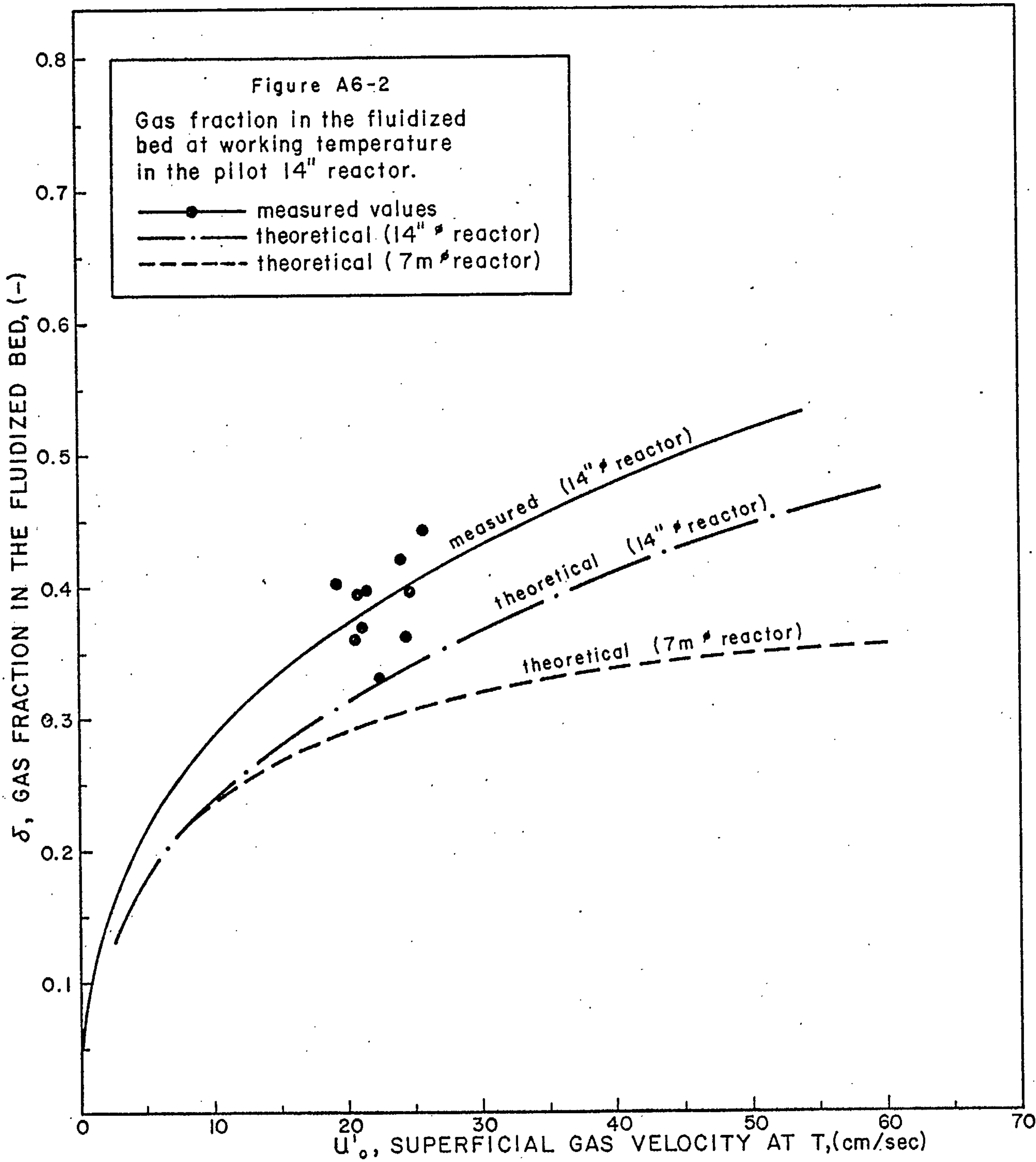

Figure A6-2

Gas fraction in the fluidized bed at working temperature in the pilot 14" reactor.

APPENDIX A7

PROCESS ECONOMICS

LIST OF TABLES IN APPENDIX A7

Table	Title	Page
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APPENDIX A7

PROCESS ECONOMICS

A7.1 CAPITAL COSTS

711 Roasting Section and Acid Plant

The direct capital costs determined for this section are presented in Table A7-1. Costs for the roasting section were determined as a function of hearth area from information supplied by Lurgi Canada Ltd. and are plotted in Figure A7-1. The required hearth area for scale-up purposes was computed on the basis data obtained from the mini-pilot trials. The hearth area to capacity relationship is shown in Figure 5 (see Section 6.1).

Waste heat boiler costs were supplied by Foster Wheeler Ltd., based on the projected gas flows and temperatures.

Additional handling, storage and feed preparation includes such items as the additional drying and feed preparation required by the sulphation roaster, feed unloading facilities, as well as storage and blending. Inclusion of this cost is a conservative measure (see Section 6.1 for further discussion).

Acid plant costs are based on a double contact type unit for zinc dead roasters. The increase in capital cost for the sulphation roast acid plants over the dead roast case reflects the lower off-gas SO_2 content of the former. Plant cost versus SO_2 content follows an exponential relationship for a constant acid production. Further discussion of acid plant costs is given in Section 6.1.

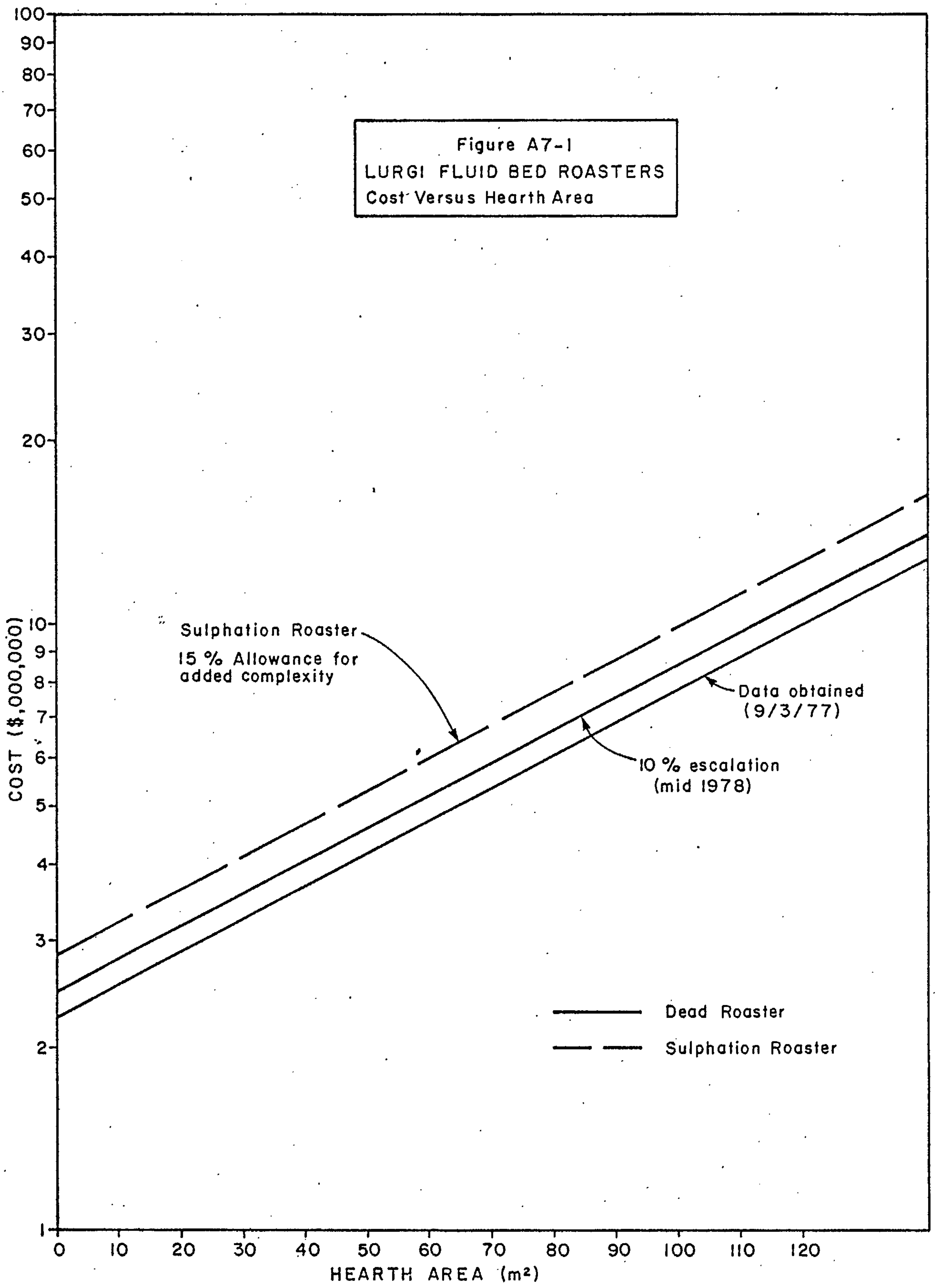

Figure A7-1
LURGI FLUID BED ROASTERS
Cost Versus Hearth Area

TABLE A7-1

CAPITAL COSTS
ROASTING SECTION AND ACID PLANT
($,000)

	DR	SR (Bulk)	SR (Tailings Refloat)	SR/DR
[†]Roasters (2)	19,400	20,000	21,000	12,850
Waste Heat Boilers (2)		6,610	6,900	9,430
Additional Handling, Storage and Feed Prep.	1,364	2,570	2,890	2,610
ROASTING SECTION	20,764	29,180	30,790	24,890
ACID PLANT	11,520	17,370	18,760	15,570

[†] Includes Per Roaster: main blower, roaster shell and refractory, basic instrumentation, hot cyclones, evaporative gas cooler, electrostatic precipitator[*] wet scrubber, connecting ducts and flues, feed storage bin, feed system, rotary calcine cooler, calcine surge bin, cold calcine conveying system. Also included are foundations, electrical systems, process piping, buildings and site development.

[*] Although allowance is for two electrostatic precipitators, only one will be required for the SR plants.

712 Leach Purification-Cadmium Recovery Plant

The direct capital costs determined for this section are presented in Table A7-2.

Neutral leach costs are higher for the sulphation roast plants than for a dead roast plant because of the need for filtration of the neutral leach thickener underflow. It was assumed that continuous vacuum filters can be used for this service.

The cost of the strong acid leach section of the sulphation plants are less than that for a dead roast plant because of the small volume of neutral leach residue that must be leached in the sulphation plant (see mass balances Appendix A2). Jarosite precipitation is eliminated from the sulphation roast flowsheet because the hot acid leach solution is recycled to the roaster. Lower building costs for this section reflect elimination of the jarosite precipitation equipment.

The solution purification and cadmium recovery plants are judged to be the same size and cost for dead roast and sulphation roast plants.

713 Electrolyses, Melting and Casting

The capital costs for this section were assumed to be identical for the sulphation roast and dead roast plant.

714 Neutralization

This section is necessary to maintain the sulphate balance of the plant. The design and costs were based on

TABLE A7-2

CAPITAL COSTS
LEACH-PURIFICATION-CADMIUM-RECOVERY SECTION
(mid 1978 $,000)

	DR	SR (Bulk)	SR (Tailings Refloat)	SR/DR
1. Neutral leach & Settling	3,012	4,599	4,599	4,288
2. Strong Acid Leach	2,846	2,362	2,362	1,708
3. Jarosite Precipitation	3,936	-	-	-
4. Purification-Cadmium Recovery	6,273	6,273	6,273	6,273
5. Building	5,182	4,146	4,146	4,146
	21,249	17,380	17,380	16,415

phosphoric acid fertilizer plant practice and bench scale testing (see Appendix A1.3). Equipment costs were determined from standard sources, and the price of a Bird-Prayon filter, with countercurrent washing, a unit widely used in the fertilizer industry, was obtained from the Bird Machine Company of Canada Ltd. Filter size was based on the tests described in Appendix A1.3. The usual factors, including buildings, were applied to complete the estimate.

716 Copper SX-Electrowinning Plant

Costs for a basic turnkey plant were obtained from Holmes and Narver. In addition to the turnkey plant, a precoat pressure filter was included for polishing the SX feed solution plus a 70 ft diameter thickener for settling the iron oxidation-neutralization products as well as pumps and reagent feeders. Factors were used to bring the equipment costs to the turnkey plant status.

717 Lead/Silver Recovery Plant

The flowsheet shown in Drawing No. 6 was used as a guide for costing this section. It was assumed that wood stave tanks and rubber covered ship propeller type mixers would be used for leaching. Rubber lined slurry pumps were assumed to be required for all fluid transfer operations. The three filters were assumed to be rubber or plastic covered rotary drum units with a filtrate rate of 2 US gpm/ft^2 of active filtration area, based on some very preliminary bench scale results. This is the largest source of possible error in the estimate. Purchased equipment costs were taken from published data and a factored estimate of total fixed capital was developed in the usual manner. A comparison with an updated estimate for a similar plant for treatment of the same product developed by Battelle Memorial Institute for St. Joseph Lead in 1956, agrees to within less than 5%.

The capital cost of this section might be reduced if a horizontal belt filter with countercurrent washing could be substituted for the brine leach filter, repulp equipment and repulp filter.

718 Other Capital Costs

Other capital costs, auxiliary facilities, services, indirect costs and working capital based on conventional plant costs.

A7.2 OPERATING COSTS

Labour, power, reagent and fuel costs are largely self-explanatory and are shown in Tables A7-3 through A7-6. The values were based on data for conventional zinc plants and were adjusted where necessary for changes in standard practice. Other expenses were determined using factors.

TABLE A7-3

LABOUR REQUIREMENTS
(Jobs)

	DR	SR (Bulk)	SR (Tailings Refloat)	SR/DR
Operating				
Roasting and Sulphuric Acid	38	45	45	45
Leach Purification and Cadmium Recovery	64	64	64	64
Electrolysis*	55	55	55	55
Melting, Casting and Shipping	36	36	36	36
Spent Elect. Neutr.		8	8	8
Copper-SX-Electrowin.		11	11	-
Lead/Silver Recovery	4	12	12	12
Maintenance	105	121	121	121
Staff and Administration	96	97	97	97
Total Jobs	398	449	449	438
Cost ($,000)/yr. Mid 1978	7496	8457	8457	8249

* assumes automatic stripping

TABLE A7-4

ELECTRICAL ENERGY REQUIREMENTS
kwhr/Tonne Zinc Metal

	DR	SR (Bulk)	SR (Tailings Refloat)	SR/DR
Concentrate Handling	5.84	9.00	9.60	7.02
Roasting and Calcine Delivery	53.93	82.57	88.07	64.39
Acid Plant	26.23	30.45	39.49	37.13
Leaching, Purification and Neutralization	82.12	82.12	84.00	82.12
Cadmium Recovery	6.00	8.56	8.56	6.33
Cooling Water	78.92	78.92	78.92	78.92
Utilities (Zinc Plant)	72.75	72.75	72.75	72.75
Zinc Electrolysis	3747.82	3747.82	3747.82	3747.82
Zinc Melting and Casting	123.67	123.67	123.67	123.67
Miscellaneous	100.00	100.00	100.00	100.00
Copper Plant	-	53.07	50.12	-
Lead/Silver Plant	-	28.00	11.13	6.59
Total kwhr/Tonne	4297	4417	4414	4327
Cost ($,000/yr) @ $0.01866/kwhr	$7274	$7477	$7472	$7325

TABLE A7-5

REAGENT COSTS
($,000/YR)

	DR	SR (Bulk)	SR Tailings Refloat)	SR/DR
ZINC PLANT				
Potassium Permanganate	318.5	-	-	558.0
Strontium Carbonate	276.0	276.0	276.0	276.0
Soda Ash or Equiv. Jarosite ppte. agent	200.0	-	-	-
Manganese Dioxide	90.9	2.5	2.5	-
Limestone				
Scrubber	27.0	280.7	330.1	156.0
Iron Oxidation	-	48.4	48.4	-
Bleed Neutral.	51.7	94.6	94.6	66.0
Spent Neutral.	-	1197.9	1205.6	643.5
Hydrate of Lime	159.8	279.8	273.1	159.8
Flocculents	80.0	80.0	80.0	80.0
Leveling Agents	110.0	110.0	110.0	110.0
Copper Sulphate				
Arsenous Trioxide				
Ammonium Chloride				
Antimony Trioxide				
Sodium Hydroxide				
COPPER SX-ELECTROWINNING PLANT				
Organic Reagent	-	72.9	68.8	-
Precoat and Body Mix	-	27.7	26.2	-
LEAD/SILVER RECOVERY PLANT				
Lime (CaO)	-	612.2	291.2	200.4
Sodium Chloride	-	226.8	249.8	127.8
Calcium Chloride	-	453.6	431.2	281.8
TOTAL	1313.2	3763.1	3487.5	2659.3

Relevant Unit Costs (where available)

Potassium Permanganate	$ 0.50/lb
Strontium Carbonate	$ 0.24/lb
Soda Ash	$ 0.04/lb
Manganese Dioxide	$ 72.0 /MT
Limestone	$ 11.0 /MT
Lime	$ 69.0 /MT
NaCl	$ 10.0 /MT
$CaCl_2$	$100.0 /MT

TABLE A7-6

FUEL REQUIREMENTS

	DR		SR (Bulk)		SR (Tailings Refloat)		SR/DR	
	Kcal/day (x10^6)	$/year ($000)	Kcal/day (x10^6)	$/year ($000)	Kcal/day (x10^6)	$/year ($000)	Kcal/day (x10^6)	$/year ($000)
1. Leach-Purification Electrowinning	564.7	$1,961	471.0	$1,635	471.0	$1,635	369.1	$1,282
2. Waste Heat Boiler Recovery	-423.5	-1,471	-352.8	-1,225	-378.3	-1,313	-315.1	-1,094
Net Requirement for Hydromet Section	141.2	490	118.2	410	92.7	322	54.0	188
3. Feed Dryer	-	-	49.3	171	53.1	184	23.0	80
4. Lead/Silver Product Dryer	-	-	15.7	55	6.5	23	4.2	15
5. Misc. (including propane)		577		627		627		627
TOTAL $,000		$1,067		$1,263		$1,152		$ 910

* includes 20% allowance for convective heat loss
Bunker C (180,000 BTU/Imp. gal) @ 45¢/gal
Propane (120,000 BTU/Imp. gal) @ 46¢/gal

A7.3 FEED COSTS

731 Concentrates

The schedule of smelter payments used is shown in Table A7-7 and was derived from published and confidential sources. As noted on the table, minor and precious metals, other than silver, were neglected in the payment calculations, as were penalties, because the effect of these on return on investment is generally negligible for these ores.

The zinc concentrates were divided into two catagories for the purpose of computing zinc concentrate payments. High grade concentrates included only the 56% zinc concentrate which was paid for using schedule a. on Table A7-7 (i.e. gross payment = 20 x 56. x 0.85 x 0.32 = $304.64/S.T.). Other concentrates are paid for using schedule b. on Table A7-7 (e.g. gross payment = 20 (52.- 8.) x 0.32 = $281.60/S.T.) which reflects a penalty charged against lower grade concentrates. The two schedules coincide at about 52.25% zinc.

Net concentrate payments to the mine were calculated on the assumption that the per ton return would be the same whether the concentrate was sold in Europe or in the Bathurst area. Therefore net payment to a smelter located in Bathurst = Gross payment - (treatment + transportation) + $6.00:where transportation is inland transport and overseas shipping. The $6.00 is the inland freight and handling charge which must be paid by the smelter if mine is to receive same net return as from European smelter.

Table A7-8 lists computed payment for concentrates at zinc prices of $0.24, $0.32 and $0.40/lb.

TABLE A7-7

SMELTER PAYMENTS

1. Zinc Concentrate

Payments

zinc

a. 56% zinc concentrate: 85% of the zinc price paid for the current U. S. producer price.

b. Other concentrates 52%, 50% and 48% Zn: deduct 8% from the zinc assay and pay for 100% of the remainder at the current U. S. producer price.

Treatment Charge

40% of the above zinc payment.

2. Lead Concentrate and Lead Oxide Product

Payments

lead

a. Concentrate: 90% of the lead paid for at the current U. S. producer price, less a deduction of 5¢/lb.

b. Oxide product: 92% of the lead paid for at the current U. S. producer price, less a deduction of 5¢/lb.

*copper — 50% of the copper paid for at the current U. S. producer price.

silver — 95% of the silver paid for at the Handy and Harman quotation, less a deduction of 15¢/troy oz.

Treatment Charge

$50.00 per dry short ton.

3. Copper Concentrate

Payments

copper — 100% of the contained copper paid for at the current U. S. producer price, less a deduction of 6¢/lb.

silver — 0.5 troy oz/ton is deducted per dry short ton and 90% of the remaining silver is paid for at the Handy and Harman quotation.

Treatment Charge

$48.00 per dry short ton plus a penalty of $2.00 per dry short ton for each 1% below 26% copper.

Penalties and credits other than those above were neglected.

Freight Charges: From mine to European smelter $24.00/dry s. ton.
From mine to Bathurst area facility $6.00/dry s. ton.

* Copper in the lead concentrate was neglected except in the Heath Steele bulk conventional models, Tables 17 and 18.

TABLE A7-8

NET PAYMENT TO MINE FOR CONCENTRATES

Zinc price ($/lb)	0.24	0.32	0.40
	Net Value $/S. Ton Concentrate		
DR* (52% Zinc)	108.72	150.96	193.20
DR* (56% Zinc)	119.09	164.78	210.48
SR (Bulk) (32% Zinc)			
30% Metal Value	82.14	97.50	112.86
35% Metal Value	95.83	113.75	131.67
40% Metal Value	109.52	130.00	150.48
SR (Tailings Refloat) (30% Zinc)			
30% Metal Value	59.47	73.87	88.27
40% Metal Value	79.30	98.50	117.70
SR (BMS Tailings Refloat)	40.90	55.42	69.93

* includes $6.00/S. Ton transportation credit for transportation within Bathurst area.

732 Bulk Concentrate

The payment schedule for bulk concentrates is shown in Table A7-8. Values were computed on the basis of current prices for lead, copper and silver ($0.32/lb, $0.62 and $5.40/troy oz respectively) and prices of $0.24, $0.32 and $0.40/lb for zinc at 30%, 35% and 40% of the contained metal values. For discussion of the per cent of full value required by the mines, see Section 6.5.

733 Brunswick Tailings Refloat-Zinc Middlings

Brunswick Mining and Smelting presently refloat tailings with three stages of cleaning to produce a 50% zinc concentrate with 30% recovery of zinc from the tailings. The tailings refloat product identified in this report is removed at the second cleaning stage. At this stage the flotation product, approximately 240 TPD, contains 50% of the zinc from the tailings as a concentrate containing 30% Zn, 3.6% Pb, 0.6% Cu and 4.4 oz/ton silver.

This tailings refloat product was used as the principal feed material in the mini-pilot plant trials and is the feed for the "tailings refloat" full plant scale-up in Section 5. In this case the product was assumed to be representative of a low lead middlings product, which might be obtained from zinc circuits. As such, reasonable payment for this concentrate was <u>assumed</u> to be 30% of full contained metal value.

In the integrated sulphation roast-dead roast general model described in Sections 5 and 6, which is the Brunswick Model of Section 7, and in the Heath Steele Model, feed to the sulphation roaster is a blend of a true tailings refloat product and a zinc middlings product. The latter might be recovered from a zinc flotation circuit as a consequence of upgrading final zinc concentrates at Brunswick. The composition of the middlings for purposes of this evaluation is assumed to be the same as the true tailings refloat product.

For the Brunswick and Heath Steele Models discussed in Section 7, payment for the tailings refloat product (i.e. 240 STPD second stage cleaner), which is presently being produced, is based on the value received for the upgraded third stage, 50% zinc concentrate, as calculated from the smelter schedule shown in Table A7-7. A freight deduction of $24.00/ ST is subtracted to determine the net value of the 50% zinc concentrate at the mine. A freight adjustment of $6.00/ST for the second-stage cleaning, tailings refloat product is credited to the mine to give the same annual net return to the mine as for the 50% zinc, third-stage cleaning, refloat product. Payment for the zinc middlings product in these models is at 30% of full value of contained metals.

734 Pressure Leach

Tables A7-9 through A7-11 summarize costs and revenue for the Sherritt Gordon pressure leach process. The data was obtained from Sherritt Gordon summary reports dated January 24, 1978 and March 22, 1978. For comparative purposes, the capital costs (Table A7-9) were escalated from second half 1977 to mid 1978 (assumed escalation 6%/yr.). Operating costs (Table A7-10) were also changed, where possible, to the same basis as that used in the roasting plant i.e. $18,843/yr average cost/job, $0.01866/ kwh (from $0.0143/kwh) and Bunker C $0.45 Imp. gal. (from $0.54/Imp. gal.).

TABLE A7-9

SHERRITT GORDON - PRESSURE LEACH
CAPITAL COSTS

		Bulk Concentrate $,000		Tailings Refloat $,000	
		Last Half 1977	Mid 1978*	Last Half 1977	Mid 1978*
1.	Capital	110,000	114,763	110,000	114,763
2.	Preproduction Expense				
	Pilot Plant	4,000	4,000†	4,000	4,000†
	Feasibility Study	300	313	300	313
	Training and Relocation	760	792	760	792
3.	Start-up Costs	880	918	880	918
4.	Working Capital	5,150	5,373**	5,500	5,737
	TOTAL	121,090	126,159	121,440	126,523

* Escalation = 4.3% estimate for 9 months
† Not escalated
** Working capital considered too low compared to dead roast plant

TABLE A7-10

SHERRITT GORDON - PRESSURE LEACH
OPERATING COSTS

	(Bulk Concentrate) (As Presented) Last Half 1977	(Bulk Concentrate) Relative to SR[2] Mid 1978	(Tailings Refloat) (As Presented) Last Half 1977	(Tailings Refloat) Relative to SR Mid 1978
1. Labour				
operating 239 jobs	3,200	6,347[1]	3,200	6,347[1]
maintenance 98 jobs	1,400		1,400	
2. Operating Supplies				
general	3,160	3,362[2]	3,160	3,362[2]
acid	2,240	2,240[3]	1,440	1,440[3]
3. Maintenance Materials	1,250	1,330[2]	1,250	1,330[2]
4. Utilities				
Fuel (5.4 Mgals BunkerC)	2,916	2,430[4]	2,866	2,388[4]
Electricity @ 4284 kwh/S.T.	6,126	7,994[5]	6,126	7,994[5]
Other	358	381[6]	358	381[6]
5. Taxes and Insurance	950	950	950	950
DIRECT OPERATING COSTS	21,600	25,034	20,750	24,192

[1] Av. cost/job (fringe benefits incl.) = $18,834

[2] Escalation 8.5%/yr

[3] Acid at $20/Ton assumes availability from Belledune lead smelter i.e., closure of fertilizer plant?

[4] Sherritt Gordon cost at $0.54/Imp gal reduced to $0.45/Imp gal

[5] Power cost mid 1978 = $0.01866/kwh

[6] presumed balance for propane for casting lines etc., escalated at 8.5%/yr

TABLE A7-11

SHERRITT GORDON-PRESSURE LEACH
ECONOMIC SUMMARY ($,000/YR)

	Bulk			Tailings Refloat	
Revenue Zinc Price ($/lb)					
0.24	67,000			54,500	
0.32	83,000			70,500	
0.40	99,000			86,500	
Direct Operating Cost	25,034*			24,192 (est)*	
Depreciation	7,918			7,918	
Cost of Concentrates	Percentage of Total Metal Value as Payment				
Zinc Price ($/lb)	30%	35%	40%	30%	40%
0.24	26,745	31,203	35,660	20,656	27,541
0.32	31,746	37,037	42,329	25,657	34,209
0.40	36,748	42,872	48,997	30,658	40,877
Profit (before tax) Zinc Price ($/lb)					
0.24	7,303	2,845	- 1,612	1,734	- 5,151
0.32	18,032	13,011	7,719	12,733	4,181
0.40	29,300	23,176	17,051	23,732	13,511
Total Capital Investment	126,159			126,523	
Return on Investment (before tax) Zinc Price ($/lb)					
0.24	5.79	2.26	- 1.28	1.37	- 4.07
0.32	14.51	10.31	6.12	10.06	3.30
0.40	23.22	18.37	13.51	18.76	10.68

* Sulphuric acid purchased @ $20/ST.