THE

HAROLD A. STEANE
MEMORIAL AWARD
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1. Summary
The Harold A. Steane memorial Award was established in May 1989 with major donations from Coast Technical Services Inc. and Placer Dome Inc. and with lesser donations from a few friends and colleagues of Harold. Subsequent donations have been made by former colleagues, the Steane family and Barrick Gold Corporation. The intent is to provide a perpetual award of $1500 to $3000 to a student at the University of British Columbia, registered in the UBC Department of Mining Engineering and entering either of the last two years of the Mineral Process Undergraduate Program or any year in the masters Program.

This booklet has been prepared by the trustees to give the recipients of the award a brief insight into the life of Harold Steane; where he came from, what he did and why he was so highly regarded in the mining fraternity.

Included in the booklet is a brief biography of Harold and a collection of some of the papers that Harold prepared and presented during his career. There were many more papers but these are the only ones that could be located by the trustees.

The trustees hope that the Harold A. Steane Award recipient will appreciate the preparation of this booklet and find it interesting.
2. Biography
Biography – Harold A. Steane

Harold A. Steane was born in Australia in the famous gold mining town of Ballarat on December 3, 1910. Harold completed high school in Ballarat and then attended the Ballarat School of Mines, graduating with a diploma in metallurgy in 1930.

In March 1930 Harold joined the staff of Mt. Isa Mines Limited at Mt. Isa, Queensland as Junior Metallurgist. In June 1933 he was awarded a scholarship to attend the University of Utah in Salt Lake City, where he earned a Master of Science in Metallurgy (ore dressing).

In August 1934 Harold joined the staff of Cyprus Mining Corporation on the Island of Cyprus as Assistant Mill Superintendent. He remained until July 1936 then returned to Mt. Isa Mines in December 1936 as Research Metallurgist. In December 1937 he returned to Cyprus Mines as Assistant Mill Superintendent but left when World War II forced the closure of the mine in June 1940.

At that time, Harold returned to Australia to work as Research Metallurgist and Assistant Mill Superintendent for Lake George Mines Ltd. at Captain’s Flat in New south Wales. He remained at Captain’s Flat until May 1946 when he returned to Cyprus Mines Ltd. as Assistant Mill Superintendent.

In February 1948 Harold moved to Uruwira Minerals Ltd. at Mpanda, Tanganyika, where he was employed as mill Superintendent until November 1951. Harold was then engaged as Superintendent of Mills for Mount Morgan Ltd., at Mount Morgan, Queensland, from January 1952 until December 1952 when he left with his family to take up residence in Canada and joined the staff of Canadian Exploration Company Limited, at Salmo, B.C., a subsidiary of the Vancouver-registered company, Placer Development Limited.

Canadian Exploration was at that time operating a tungsten mine and mill and a lead-zinc mine and mill on the same site. Harold moved his family to the mine site in the spring of 1953 and worked as Chief Metallurgist for the next seven years. It was during this period that he gained a strong reputation in the industry for treatment of tungsten ores as well as for the treatment of lead-zinc ores.

Placer Development acquired the Craigmont property at Merritt B.C., in 1958 and when a production decision was made in 1960 Harold moved to Vancouver to work with the design consultants on the 5,000 tpd copper concentrator. This was the first of many concentrator and plant designs on which Harold played a major guiding role.

With the move to Vancouver, Harold along with three or four other Placer engineers, formed the nucleus of what grew into Placer Development’s Project Development Group.
In the December 1960 issue of *Western Mining & Oil Review*, Harold was named Mining Man of the Month. This award was for his investigations preparatory to the mill design and flow sheet for Craigmont Mines, and recognition for his many technical papers. These included notably a feature article on tungsten in Canada and separate papers on the Salmo tungsten and lead-zinc mills of Canadian Exploration: all of which appeared in the Milling Volume of the Sixth Commonwealth Mining and Metallurgical Congress.

Harold was appointed General Mill Superintendent, a position he filled until his retirement in 1978. During the 20 year period of Placer’s most successful growth he was responsible for the metallurgical design of; Endako Mines – Fraser Lake B.C., and its subsequent expansion to 30,000 tpd, the Marcopper mine – a 15,000 tpd copper concentrator in the Philippines which was later expanded to 30,000 tpd, Cortez gold concentrator – a 1,500 tpd gold cyanide plant in Nevada, Gibraltar Mines concentrator – a 40,000 tpd copper-molybdenum plant in central British Columbia, near Williams Lake and the McDermitt mercury flotation and roasting plant in Nevada.

While deeply involved in the design of these plants, Harold also found the time to give general guidance and expertise to Placer’s operating concentrators and worked with Noranda’s metallurgists on the design of the Mattagami Lake lead-zinc concentrator in Northern Quebec. During this period he also established the Placer Development metallurgical and research laboratory in Vancouver which opened in 1970.

Harold found the time to pass much of his knowledge on to Placer’s metallurgists, and gave annual lectures to University of British Columbia and McGill university metallurgical students.

Harold reached retirement age in 1975 but was retained by Placer Development as a consultant until 1980. During this period he played a major role in the development of a process to remove arsenic and antimony from the Equity Silver ore, located near Houston B.C., and assisted in the design of the Mineral Real de Angeles concentrator – a 10,000 tpd silver-lead-zinc plant in Mexico. He also engaged in private consulting on tungsten and copper properties in places as far away as Turkey.

Family commitments played a major role in Harold’s life. On his second trip to Cyprus he met and married Isobel Innes, the daughter of a ship’s captain employed by Cyprus Mines. They raised four sons: Alan, an engineering physicist, was born in Australia; Brian, a mechanical engineer, was born in Cyprus; Robert, a metallurgist, was born in Tanganyika; and David, an electrical engineer, was born in Canada.
Harold was a true gentleman, and because of his sincere and willing manner coupled with his thorough and competent technical expertise he became a mentor to the many Metallurgists trained and employed at Placer’s operations and also to many outsiders who sought his advice. He would not give an answer to any engineering or scientific question without giving due thought to his answer. In the words of one who knew him well; “Harold was an incisive, tenacious, clear thinker possessing the remarkable capacity to reconcile analytical theory with the practical operating requirements and as a result was highly respected in the mining industry for his knowledge and expertise in the field of mineral dressing”.

He was a member of the Canadian Institute of Mining and Metallurgy, the Australian Institute of Mining and Metallurgy, and the American Institute of Mining Engineers.

Harold remained active and interested in the mining industry from the time of his full retirement in 1980 until he suffered a stroke in 1987 which confined him to bed until his death on April 3, 1989.

Harold’s wife Isobel continued to live in the family home in Vancouver and attended the awarding of the H.A.Steane Memorial Scholarship each year until her death on May 12, 1997. A member of the Steane family continues as a trustee on the board.

NB: Attached please find the employment history form that Harold filled out at the start of his employment with Placer Development. The salaries noted are in Pounds Sterling and Pounds Australian.
EDUCATION

Highest grade completed at school? Senior High School
Which school? Balket High School
What is the year you left? 1926
Have you passed your Junior Matriculation? Yes
Senior? Yes

If an undergraduate, state years required for degree:

Name of College or University attended:

University of Utah

Degrees obtained: A.A.

What did you specialize in? Geology

Graduated year: 1926

Name and location of Vocational, Technical or Correspondence Schools:

Length of Courses:

Certificate:

Year:

List any special Certificates, Licenses, etc., which you may have. B. C. Assayer Certificate, Land Surveyor, etc.

List any special qualifications not listed above:

Member Australian Inst. of Mining & Metallurgy, Member A.I.M.E.

What Languages other than English do you read?

Write?

Speak?

RECORD OF PREVIOUS EMPLOYMENT

Give a consecutive record of past employment, include any service in the Armed Forces.

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<th>Leaving Date</th>
<th>Name and address of Employer</th>
<th>Type of Business</th>
<th>Occupation</th>
<th>Your Salary</th>
<th>Reason for Leaving</th>
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<td>Tinner Metallurgist</td>
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<td>Assn Mill Sup't</td>
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<td>M. S. Taylor, Mill Superintendent</td>
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<td>Mount Isa Mines Ltd, Mount Isa, Qld</td>
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<td>$6.00 $6.00</td>
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<td>L. K. Jacobsen, Mill Superintendent</td>
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<td>Apr 1936</td>
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<td>Mill Sup't</td>
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<td>Voluntary, cations education</td>
<td>C. H. Richards, Gold Manager</td>
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<td>Mining</td>
<td>Superint. of Mills</td>
<td>$15.50 $16.75</td>
<td>Voluntary</td>
<td>G. Sheer, General Manager</td>
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EMPLOYMENT RECORD WITH THIS COMPANY

Please sign here: [Signature]

From | To | Occupation | Dept. Superintendent | SALARY | SUPERINTENDENTS REMARKS
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Type of Employment you consider yourself most suited for:

3. Collection of Work
DEVELOPING AND IMPLEMENTING
MINERAL PROCESSING FLOWSHEETS
FOR NEW MINING VENTURES

by

H. A. STEANE
GENERAL MILL SUPERINTENDENT
PLACER DEVELOPMENT LIMITED
VANCOUVER, B. C.

November 1974
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1.0 ABSTRACT

The capital cost of the milling plant is a large proportion of the total capital investment to bring a mine into production. A successful operating plant is essential to the profitability of the project.

The design of the milling plant involves the coordination of past experience from successful plants, test work on the characteristics of the particular ore to be treated, and consideration for innovations for improvement over past experience with the objective of improving the overall profit return on the venture.

This discourse covers the design and construction of a concentrating plant, commencing with the primary crusher and ending with concentrate ready for shipment.

It covers the period from the time a prospect becomes an exploration venture to the start of a producing mill. I will endeavour to cover all the important factors affecting ore treatment that have to be considered and resolved during that period, including economic considerations and the preparation of a feasibility study.

The INTRODUCTION reviews the intention in preparing the lecture and gives the basis for the information given.

The section on SAMPLING THE ORE DEPOSIT outlines the purposes to be considered in obtaining assay information affecting the process, and the sampling for metallurgical testing.

The section on METALLURGICAL TESTING covers the parameters that should be determined in the test work to provide the information necessary for flowsheet development. It includes a study of recovery relative to the degree of grinding required to give optimum economic return.

The section on FLOWSHEET DEVELOPMENT covers the factors influencing the development of the process flow scheme, selection of type and size of equipment, and arrangement for plant design.

The section on FEASIBILITY STUDY outlines the assembly of information with capital and operating cost to assess the profitability of the venture to enable management to decide whether to proceed to production.

The section on DESIGN AND CONSTRUCTION outlines the type of organization involved in the design and construction of the plant.
2.0 INTRODUCTION

This presentation will cover the functions to be considered in developing and implementing mineral processing flowsheets for a new mining venture. The first decision had to be the form it would take.

I do not propose to give a detailed step by step account of how to get a mill built and put into production, but to give my views on the important aspects of mineral processing as it affects design and operation of a flotation mill. These views might differ considerably from those of others, but they have been accepted by the management of Placer Development.

To let you know the basis for what I have to say, I have prepared SLIDE No. 1, which lists the mining properties that Placer has brought into production and for which I had responsibility for the metallurgical design.

I want to refer to Mattagami Lake Mines first. My association with that project was as a member of a milling committee of three, of which Lyall Ames of Noranda Mines was chairman and Mike Bennett of Quemont was the other member. My association and responsibility with this project was not as close as with the others.

For the other five plants, the supervision for design and construction was provided by Placer. The responsibility was a team effort by the Placer engineering group, with each team member covering his respective field of mechanical, electrical, metallurgical, plus a group providing the services of estimating, purchasing, and ore testing. My responsibility was the metallurgical design.

The actual detailed design work was done by the consulting firm of Wright Engineers, and the construction was carried out by Commonwealth Construction. Placer retains overall supervision, guides the progress, and gives approval at all stages throughout the design and construction.
3.0 SAMPLING THE ORE DEPOSIT

3.01 Objectives

A sampling program is the first step when investigating a property for production.

The two objectives in this sampling program are:
(a) To provide samples for assay and for geological study. These will be the basis for determining the tons and grade of ore in the deposit.
(b) To provide samples for metallurgical testing, to determine the treatment required and the results expected, and to determine and quantify any zonal difference that might have an influence on metallurgical design or production results.

3.02 Samples for Assay and Geological Study

In assaying the samples, it is important that assays are carried out on every sample for each constituent expected to influence the treatment or the economics.

Analyses should include all metals expected to contribute to, or detract from, the economic value of ore, and should include all constituents likely to have a detrimental influence on the metallurgical results.

For a sulphide ore, it is important to determine the extent of oxidation of the valuable mineral. My experience has been that often this has not been done, with the result that ore is blocked out showing total metal grades, but with only sketchy information of the metal in oxide minerals, for which recovery will be low.

Other possible interfering constituents could be pyrite, pyrrhotite, graphite or mercury. If these minerals are a problem, and the quantity varies to give a significant variation in the results, then we need to know the variation throughout the deposit. If their effect is constant, there would be no need to include them in the routine assaying. The metallurgical testing would determine and allow for their uniform detrimental effect.

If variations in mineral composition occur that affect metallurgical results, and if these variations can be determined by analytical methods, the determinations should be performed on a routine basis and possibly establish whether a zonal difference exists.
Portions of the same core samples will be examined for geological structure and mineral composition to assist in outlining the ore deposit. Mineralogical examination by a microscope should be made on selected samples to assist in metallurgical testing.

3.03 Samples for Metallurgical Testing

Samples should be submitted for metallurgical testing at an early stage of the exploration program. The value of a deposit cannot be determined by total metal assay alone. We need to test the metallurgical characteristics for recovery, concentrate grade, hardness of the ore, degree of grinding needed, and any other characteristics that would influence mill performance, capital or operating costs.

I would like to stress at this stage that there is no such thing as a representative sample of ore to be processed in a plant. There will be variations throughout the deposit and therefore a range of samples should be tested to determine the range in characteristics to be encountered. Zonal differences should be quantified if they will affect the economic results of the operation.

The mill never treats a representative sample. It treats the ore that is in front of the shovel from day to day. If it can be determined that there is a real necessity for blending, this would be important in the planning. Keep in mind that successful blending requires elaborate layout and is expensive. It should not be undertaken lightly.

The samples for testing may be a section split from the assay core samples, or crushed rejects from assay samples. The crushing should be kept in as coarse a size range as possible, certainly not pulverized. I like to ask for a sample of about 200 lbs. for bench testing though smaller samples down to 50 lbs. could be accepted for preliminary tests, if necessary.

Rotary drilling costs less than diamond core drilling and I have heard long discussions trying to justify rotary cuttings for metallurgical testing. We tried this on one occasion with discouraging results. The sample is likely to be wet and require air drying before it can be handled, encouraging oxidation. It will have an undue portion of fine pulverized material thus making it unreliable for work index determination.
Drilling with 2½" core bits will give 4.5 lb. per ft., and with 3" core bits will give 8 lb. per ft., providing adequate sample for test work.

Care must be taken in the operation of the drill to avoid oil contamination of the core. The use of drilling-mud should be avoided if at all possible.
4.0 METALLURGICAL TESTING AND REPORTING

4.01 Scope of Program

The metallurgical test program would continue over a period of twelve months or more, concurrent with the procurement of samples and accumulation of assays and tonnage data for the deposit.

Metallurgical reports should be issued frequently during this period so that the information can be used in computing economics of the ore reserves and formulating a flowsheet.

The metallurgical test program for Gibraltar commenced October 1969 and continued to August 1971, a period of twenty-three months. During that period, nineteen reports were issued.

These metallurgical reports must cover all the parameters that will influence flowsheet, equipment sizing, and plant design. It is important to determine all of these points from the earliest sample that is large enough for the work, and continue to repeat them on later samples as they are received. It is important to know what variations might be expected.

Following are the important parameters to be determined:

a) Flotation reagents and conditions. This will include retention times and circulating loads.

b) Recovery

c) Fineness of grind required for optimum recovery and the relationship of grind to recovery.

d) Hardness of ore. Work index.

e) Concentrate grade. Regrinding required and impurities affecting marketing.

f) Thickening and filtering rates for concentrate. Thickening rates for tailing if applicable.

At the conclusion of the test work, an estimate will be required to show a plant metallurgical balance with recovery and the concentrate grades expected at the degree of grinding chosen as optimum for the plant.

If there is a marked variation or trend in results from various parts of the orebody, this will need to be recorded.
4.02 Flotation Reagents and Conditions

First it must be shown that the ore is amenable to concentration by flotation, with reasonable recovery to a marketable concentrate.

For this preliminary investigation, the sample of ore should be ground to a size that is thought to be adequate or finer than adequate. Reagent types and quantities are chosen from experience and a flotation test is carried out using flotation times indicated by observation and experience to get a "feel" for the behaviour of the ore.

Having determined that flotation offers reasonable possibilities, more detailed testing is planned, varying each of the conditions, only one at a time, to arrive at the combination of conditions best suited to this ore.

It will be necessary to try several reagent types and to vary the quantities used to arrive at the best conditions for optimum recovery and concentrate grade. These points will have to be determined in conjunction with other variables of fineness of grinding, flotation times, and other conditions.

This testing procedure is not a straight line path from one variable to the next, because a change in one variable can affect some or all other variables. There will be considerable repetition, which may be minimized with experience.

The investigation is not intended to cover large varieties of reagents similar in performance. Based on the investigator's experience, reasonable reagent types are soon chosen and then varying quantities are tried to narrow down the range required.

It is understood that when the plant is in operation, a more detailed investigation of reagents can be made at the property by testing in the laboratory and trial in the plant.

In flotation testing it is important to record the time interval used for each step of conditioning and flotation, and aim to determine the minimum laboratory time required for each step for adequate results.

In mill design, it will be necessary to choose a time interval for each step of the process. The laboratory test work will be a guide to which the mill designer can apply a factor based on his experience.
It will be necessary to assess the plant circulating loads in cleaner tailing and middling products. These can be calculated from assays of the various products. Individual batch tests may provide adequate information for a simple ore, but for a complex ore, locked cycle tests may be necessary to minimize the proportion of valuable metal tied up in intermediate products.

4.03 Recovery

To assess an orebody, we must determine how much valuable metal or mineral will be recovered, the cost for recovering it, and the value of the product. We must avoid spending more for recovering an additional increment of metal than it is worth. The grade of the concentrate will influence its unit value, therefore, an optimum grade must be established relative to profit.

The recovery can be influenced by the grade of the concentrate produced and by the fineness of grinding for flotation feed. Additional grinding of intermediate flotation products may influence recovery, or grade of concentrate, or both.

The tests to establish a recovery-grade relationship will be much more difficult on a fine-grained lead-zinc or copper-zinc ore than on a simple low grade copper ore, or low grade molybdenite ore.

For these low grade ores it is sufficient to establish the relationship of rougher recovery relative to fineness of grinding. Regrinding rougher concentrate and establishing the cleaner concentrate grade and recovery relative to the fineness of regrinding would then follow the rougher tests.

4.04 Fineness of Grinding Required

The fineness of grind will affect the recovery obtained and the relationship must be determined. If the grinding is too coarse, some of the valuable mineral, either free or locked in middling grains, will not be floated.

Grinding generally is the greatest single operating cost. It is expensive and we do not want to grind any finer than is justified economically.

Here I would like to refer to one of my pet aversions, that of relying too strongly on a recovery number that states the recovery as the percent of the total metal in the head. I think a more informative number is the metal assay in the tailing that is being discarded. The pounds of metal being discarded is the only true measure of the dollars lost.
There will be an economic limit to the amount of money that can be spent in attempting to lower the tailing assay. For the same tailing assay, the percent recovery figure will be much higher when starting from a 1% head assay then when starting with a four-tenths of a percent head; yet the recovery from the lower grade head may be just as good an achievement, or perhaps even a better one.

For a copper ore, the tailing could be in the order of three to seven hundredths of a percent. This is the direct measure of the copper being lost in every ton of tailing.

The following example will illustrate the point:

For a head of 1%, tailing of 0.04%, recovery is 96%.
For a head of 0.4%, tailing of 0.04%, recovery is 90%.

In either case, the cost of lowering the tailing further might exceed the return. These tailing assays might even be lower than the economic limit.

To determine the effect of fineness of grind on recovery, several recovery tests should be run after varying degrees of grinding. A screen analysis of the tailing should be made with an assay of each size fraction. With this information it will be seen at what screen size the assay figure starts to increase markedly.

I like to use the Bond expression of the 80% passing size to define the degree of grinding that has been done. I think this is far more meaningful than to refer to the percent minus 200 mesh when this is much less than 80%. Tailing losses can only be reduced by finer grinding when they are associated with the coarse particles. We are really not interested in 200 mesh.

4.05 Grindability or Hardness of the Ore

The Bond work index is the commonest method of expressing a measure of the effort that has to be expended in grinding. Discrepancies might occur in measuring it or in applying it, but it is a reasonably close practical measurement, and the formula is simple to use.

The Bond work index is the kilowatt hours required to reduce a ton of rock from an infinite size to 80% passing 100 microns. The formula allows a calculation to be made for other size ranges. I have found the formula very useful.
Other investigators have done work and are continuing to try to develop other methods to determine the grinding effort required, but their formulas are difficult to use and I doubt whether the end result is any better.

4.06 Power Required. Increase for Finer Grinding

I would like to examine the effect of the degree of grinding on the capacity of a grinding plant.

Those of you that have worked with the Bond formula may know this already; but those that are not familiar with the calculation may be surprised at the amount of additional power and plant that is required for grinding one screen size finer.

On SLIDE No. 2, the Bond formula has been set out in equation 1 and the symbols defined.

After making one approximation, equations 2 to 7 proceed to show the power increase required to grind one screen size finer. The example given is from 65 mesh to 100 mesh, but the same relationship will hold for any two consecutive screen sizes in the square root of two screen series, within the range of flotation feed sizing.

Conclusion. The power required will increase by the fourth root of two, which is 1.19, for each screen size finer grind. That means a power and plant increase of 19% that must be provided in the design for one screen size finer.

4.07 Plant Throughput. Increase by Accepting Coarser Grind

This study is an extension and a corollary to the thoughts expressed in the previous slide.

SLIDE No. 3 has been prepared to show the increase in capacity that can be obtained from an existing plant by accepting a product that is one screen size coarser.

It assumes a plant treating 100 TPH, producing a product at 100 mesh. Starting with equation 7 from SLIDE No. 2, equation 8 is developed showing the increase in throughput by accepting a 65 mesh product, which is one screen size coarser.

A summary tabulation follows showing the increase in throughput successively for one, two, three and four consecutive screen sizes in the square root of two screen series.

Conclusion. The plant throughput will increase by the fourth root of 2, or 19% additional throughput for each screen size coarser. It will increase by 41%
4.08 Fineness of Grinding. Effect on Economic Recovery

Now let us return to the relationship of degree of grinding and economic recovery.

SLIDE No. 4 shows the results of some bench tests carried out on a sample of ore from a property not yet being brought into production.

A series of flotation tests were carried out over the 80% passing size range from 35 mesh to 270 mesh, with the recovery and tailing assay as tabulated.

The overall grinding cost of $0.04 per kilowatt-hour was a realistic cost in 1973. It would be closer to $0.05 now. Using the net smelter return of $0.50 per lb. copper in concentrate, the net smelter return per ton of ore has been calculated for each grinding size. The final column shows the net smelter return per ton of ore after deducting the cost of grinding.

This shows no economic benefit in grinding finer than 150 mesh, even though the recovery continues to increase with finer grinding.

SLIDE No. 5 shows a continuation of the information from SLIDE No. 4.

Column 2 shows the relative tons that could be ground in a plant, over a range of coarseness of grind, assigning the value of 100 for grinding to 270 mesh.

All other mine operating costs have been taken at $1.00 per ton, which was a realistic cost at a large open pit mine in 1973. The present figure would be closer to $2.00 per ton.

The column headed net smelter return per ton of ore, gross, has been taken from SLIDE No. 4.

The next column shows the net smelter return per ton of ore, net, after deducting grinding and all other operating costs.

The last column shows the relative profitability from the plant, considering the increased throughput that can be obtained at the coarser grinds.

This indicates that the greatest revenue is obtained at a 65 mesh grind even though, referring back to SLIDE No. 4, this gives a recovery of 71.7% and tailing assay of 0.113% Cu.
This poses a policy decision for management on the conservation of resources as to whether the lower recovery is acceptable, even though it is the most profitable.

4.09 Concentrate Grade and Regrinding Required

Most of the properties that Placer has brought into production recently are large low grade deposits.

The general treatment has been to float the rougher concentrate at as coarse a grind as possible consistent with recovery. The rougher concentrate is then ground finer in a regrind mill to improve the liberation of the valuable mineral prior to two or more cleaning stages.

Generally, regrinding to minus 325 mesh has been adequate to give final concentrate grade. This is determined by bench scale test work with sizing analysis and assays on each fraction.

Marcopper final concentrate is 70% minus 325 and Gibraltar concentrate is 70% to 80% minus 325. In both of these concentrators the regrind mill should be larger.

Endako is the exception in that much finer grinding on the concentrate was required to free quartz grains from molybdenite to produce marketable concentrate grade. The sizing of this concentrate is 80% minus 12 microns.

Testwork had not indicated that this degree of grinding would be required, but it was quite definite from the beginning of plant operation. Fortunately, we had provided flint pebbles in the regrind mills, so that we were able to provide an adequate increase in regrind capacity by replacing the pebbles with balls.

The pebble grinding was incorporated in the original design because of advice from Climax that they had found the use of steel balls in regrind mills resulted in lower recovery. However, at Endako we have not been able to substantiate this and have not reverted to pebbles for regrinding, though a larger mill was ordered and installed in case this should be necessary.

The testing required to determine the concentrate grade would be to grind the rougher concentrate then carry out cleaner flotation in one or more stages as may be indicated by the results obtained. Hopefully on these simple ores, the amount of valuable mineral remaining in the cleaner tailing will be small and therefore an estimate can be made of its effect in a plant, based on the experience from other plants.
For more complex ores, a locked cyclic test may be required to obtain a reliable estimate of plant expectations.

As a note of caution, some ores tend to oxidize rapidly and storage of rougher concentrate for even a few hours, sometimes can result in poor test results.

When sufficient concentrate has been produced that is believed to be reasonably representative of plant production, a complete qualitative spectrographic analysis should be made and followed by quantitative chemical analysis for all elements that could affect the further processing of the concentrate.

4.10 Thickening and Filtering

For plant design it will be necessary to choose sizes for a concentrate thickener, a concentrate filter, and also a tailing thickener if it is necessary to install one.

These tests would be deferred until the conditions for grinding and flotation are well established so that the samples for testing are close to the products that will be produced in the mill.

The variation in concentrate characteristics throughout the orebody seldom vary greatly. Thus the testing for thickening and filtering can usually be confined to one or two well chosen samples.

Thickening tests on tailing need to be carried out if a thickener will be incorporated in the mill. A thickener is not always incorporated.

It is important to carry out the tests on a tailing sample with the poorest settling characteristics, because a plant thickener must be able to handle this type of tailing for an indefinite period. Settling aids and pH conditions should be investigated so that an estimate can be made of the cost of settling aids against the increased cost of a larger thickener.

4.11 Pilot Plant Testing

The question arises as to whether testing should be carried out on a pilot plant scale. I consider that pilot plant testing is not necessary unless the metallurgy is difficult and would result in large circulating loads in the flotation section.
For simple ores, some pilot plant operation might be desirable as an overall confirmation of the results predicted from the bench tests, but it is not essential unless the economics of the operation are tight. It might be required for production of large concentrate samples for distribution to prospective buyers. The use of pilot plant testing should not be entered into lightly as such tests are generally both time consuming and costly.

Placer has a pilot plant with crushers and storage bins to accept samples of several tons. There is provision for grinding, classification, and single mineral flotation for 100 to 500 lbs. per hour depending on ore hardness and fineness of grind required. We are in the process of expanding this plant to provide for production of three separate concentrates.

Placer has not conducted exhaustive pilot plant tests on properties brought into production. The policy has been to pattern the plants after similar successful operations, and to provide sufficient flexibility or reserve for adjustments. To-date we have been successful in meeting requirements.

The main parameters of reagent conditions, fineness of grind and hardness of the ore are best determined by bench tests.

The pilot plant has been used only as an overall confirmation of the results predicted from bench tests, and also for the production of a large concentrate sample for distribution to prospective buyers.

For Gibraltar, it was used also to provide concentrate samples of sufficient size to allow bench scale tests for the separation of molybdenite.

As the revenue from molybdenite was not a crucial factor in the economics of Gibraltar, the molybdenite separation was not investigated extensively. A much larger pilot plant would have been required to incorporate a molybdenite separation section. The ore sample for this was not available.

The Gibraltar plant was designed incorporating a molybdenite separation section similar to other plants, and providing leeway for in-plant adjustments. In this way the final testing for the molybdenite separation was deferred for determination on a full plant scale.

The greatest amount of pilot plant work was done on Endako. Here the low molybdenite content of the ore, 0.2% MoS₂, made it necessary to treat a large sample of ore to get enough concentrate for regrinding and upgrading. Also to get enough final concentrate to provide samples of several pounds to give to prospective buyers.
Adequate sample was available for this purpose as a result of a large scale sampling program that was considered necessary to establish the grade of the ore reserves.

One adit was driven with raises off it. One shaft was sunk, with drifts off the shaft and raises off the drifts. This completed a sampling program, which produced about twelve thousand tons of ore.

A portable crushing plant was set up to receive the whole of the ore as it was mined and crush it to about 3/4 inch. The stream of crushed ore from the conveyor was sampled continuously by a chain bucket sampler to give the primary sample cut, which was kept as separate sample lots. These sample lots were further cut down with additional crushing between stages. Portions were kept for pilot plant testing.

The cost of obtaining a large sample for pilot plant operation is an important factor in design cost. In the case of Endako, the justification for the large sampling program was to have assurance that the ore grade was adequate for an economic operation.

Gibraltar was at the other extreme and only drill core was available until after plant design was well under way and pit stripping exposed some near surface ore.

While on the matter of pilot plant and cost of getting a sample, I want to mention an interesting example of a program for a lead-zinc property in Queensland that Placer was investigating.

The ore is fine grained sulphide requiring fine grinding in the order of 325 mesh. Even at that grind, the separation by flotation is poor, with high cleaner tailing assay and circulating loads. Bench tests leave too much of the lead and zinc in cleaner tail and middling products to get a clear assessment of overall results.

We conducted a pilot plant test on 25 tons of sample at the rate of 50 Kg per hour, about 1 ton per day. The sample was obtained by 8 in. diamond drill core at a depth of 900 ft. The overall exercise of drilling to obtain the sample, transportation, and pilot plant treatment charges gave a total cost around $500,000. That is $20,000 per ton.

4.12 Autogenous Grinding

The cost of getting sufficient sample can be a large expense when autogenous grinding is being considered.
The parameters for conventional grinding can be determined adequately by bench tests. However, I believe those that have installed autogenous grinding consider a pilot plant test to be essential to determine the behaviour of the ore in grinding. This sample would be several hundred to several thousand tons and should be broken in a similar manner to the way it would be mined. The cost of getting this sample must be considered a direct cost of investigating autogenous grinding.

There seems to be a lot still to be learned in predicting the behaviour of an ore for autogenous grinding, even after the pilot plant testing. It will be a still greater step forward when it becomes possible to predict autogenous plant requirements without having to treat a large tonnage sample in a pilot plant.

We, at Placer, have been interested in following the reports and information on autogenous grinding operations at other plants, but we have not been convinced that there would be a real economic advantage in any of our operating plants to date.

We are constructing a plant in Nevada to treat mercury ore, consisting mainly of clay, which includes some sand and occasional boulders up to eighteen inches. The deposit includes bands of mercury-bearing opalite that can be mined selectively and stockpiled separately.

Its clay content made it impossible to crush or screen the ore. Pulping the clay was also a problem. These conditions led to the decision to install an autogenous mill that could take the eighteen inch lumps of rock with the clay. The opalite will be fed separately to the mill in the amount required for pulping and grinding the clay portion.

This installation should give us an opportunity to observe some of the factors in autogenous grinding, though under unusual conditions.
5.0 FLOWSHEET DEVELOPMENT

5.01 General Requirement

A flowsheet is required as a basis for the feasibility study and should be as near final as possible. In its final form for plant design, the flowsheet must give a flow balance at every point, stating the flow rate for solid, water, and pulp, in units convenient for specifying equipment. It must also state the specific size for every piece of equipment; however, experience has shown that the sizing of minor equipment is not finalized until the design is well advanced.

The flowsheet development is based on results of the metallurgical testing and a knowledge of plants treating similar ores. It is natural to follow earlier successful installations and hopefully provide improvements, if possible, without creating additional problems. Thus there is a general similarity in Placer's plants.

The metallurgical test work may have shown a range of values for work index; the relationship of fineness of grind to recovery may leave some doubt as to the optimum fineness required; yet a choice must be made for the work index and fineness of grind to be adopted as mill design criteria.

The number of cleaning stages to be provided in flotation, and the retention time for the rougher and each cleaning stage must be chosen from test results and related to experience in similar plants.

The general skeleton outline of a flowsheet is fairly straightforward, with a few major concept decisions to be made at the outset. When plant layout and sizing of equipment is considered, the detailed flowsheet needs more careful decisions.

I shall omit consideration of autogenous grinding because we have not used it. We know that some recent autogenous plants have had delays in reaching full production. A loss in revenue from this delay could easily offset any possible gains.

The general items of the flowsheet will consist of the following.

Item 1. Material flow balance.

Item 2. Primary crusher. With or without a grizzly?
Item 3. Coarse-ore stockpile. With or without the fines removed? The purpose is to provide storage capacity and have the minimum amount of crushing and conveying equipment tied in with the mine shovel and truck haul, to minimize interruptions to the mining schedule and mining equipment. The storage capacity of the pile must be decided.

Item 4. Secondary crushing plant. The sizing required for the final product has to be decided.

Item 5. Fine-ore storage capacity to give a buffer between crushing and grinding must be decided.

Item 6. Grinding section. A decision is required whether it is to be a rod mill and ball mill combination or all ball mills, size and number of units.

Item 7. Regrind mill. Number and size?

Item 8. Flotation section. Rougher section retention time, number of circuits and number of banks per circuit, size and type of machine. Cleaner section, number of circuits, number of banks per circuit, size and type of machine.

Item 9. Concentrate thickener. Size?

Item 10. Concentrate filter. Number and size?

Item 11. Concentrate dryer. Is it needed?, and if so what type and size?

Item 12. Concentrate storage. Size?

Item 13. Disposal of tailing. Method?

The list covers the skeletal flowsheet fairly well. Determining the size and number of the units requires some careful study. Many of the items are interdependent, allowing for several possibilities, and can be influenced by plant layout.

The instructions that management would like to give would be something like this.

1. Build a plant for X thousand tons per day, to give optimum recovery and concentrate grade, at lowest capital cost for a plant adequate for the job, giving due consideration to the effect of design on the operating cost.

2. Provide for the possibility of expansion, but do not spend any money on it.
This latter instruction is one that must be modified. In the concentrator, reasonable consideration can be given to the provision for expansion without great increase in cost; however, in allowing for the possibility of expansion, there must be some cost increase, though it may not be readily apparent.

For the crushing plant it is not possible to provide for expansion without increasing the cost substantially. A crushing plant comprises several crushers with an integrated conveyor system. Therefore it is of no use to allow space for adding further crushers unless the conveyors are already oversize; or expenditure has been made to allow the conveyor to be replaced by a wider belt; or allowance has been made for speeding up an existing belt, already slower, wider, and with a drive of greater power rating than needed. We may be considering power in the region of 300 to 500 horsepower.

The alternative for the crusher expansion is to add additional crushers, either as an extension to the present building or in a separate building, but in either case its own additional conveying system will be required from the coarse ore stockpile right through to the fine-ore storage.

The following comments will examine the requirements in detail at each step in the flowsheet.

5.02 Material Flow Balance

This item must be compiled in detail soon after the general flow scheme and equipment type is settled, and before specifying the final size and choice of each item of equipment.

The material flow balance must be complete and detailed for solid, water, and pulp, to make sure that there is a balance between everything entering the system and everything leaving the system. It must show the rate for all streams and circulating loads within the system. This flow balance provides the reference information for sizing all major equipment and auxiliary equipment, conveyors, pumps, pipelines, etc.
5.03 Primary Crusher. With or Without a Grizzly?

At first thought the grizzly ahead of the gyratory seems quite logical; it would remove the undersize material that is too small to be crushed in passing through the crusher, and thus result in an increased throughput of the installation; however, there is some doubt as to whether there is a material increase in the throughput rate with a grizzly.

The addition of a grizzly makes a considerable increase in the capital cost. It requires a substantial increase in elevation of the dumping point above the crusher. It requires a concrete bypass and storage chute to get grizzly undersize to join the crusher discharge on the conveyor belt. The design is further complicated if provision is made for truck dumping at two sides of the crusheer.

As trucks have increased in size, the depth of the load has increased, so that on dumping the truck, this greater depth of ore hitting the grizzly has made it less and less effective in allowing the fines to pass through. The wear and maintenance on a grizzly is costly.

The decision has to be made at the outset; a grizzly cannot be added later. I do not recommend use of a grizzly ahead of a gyratory crusher.

Following is the situation at the Placer operations in the order in which they were built.

Craigmont - 1961 No grizzly One side dump
Endako - 1965 Grizzly One side dump
Marcopper - 1969 Grizzly One side dump
Gibraltar - 1972 No grizzly Two side dump

5.04 Coarse-Ore Stockpile

For the usual conical stockpile, discharging through drawpoints to one or more conveyors underneath, the live storage is about 20% of the total storage. This live storage should be made large enough to allow the remainder of the plant to operate during day to day interruptions in mine production, and also to allow mine operation to be independent of interruptions to the secondary crushing plant.

The dead storage should be adequate to allow the remainder of the plant to continue through downtime of the primary crusher for mantle or concave replacement, requiring several days. Bulldozers would be used to push the ore in dead storage to the drawpoints.
There should be no expectation of any ore blending in this stockpile. There may be segregation, which is not an advantage. If blending is desired a very large expenditure of capital is required.

For climates with freezing weather or for hard to handle sticky ores, it is desirable to screen out the fines ahead of the stockpile. This separation of the primary fines aggravates the regularity of treatment later in the grinding circuit, but has to be accepted for free flow of ore from the stockpile.

5.05 Secondary Crushing Section

This heading covers both secondary crushers and tertiary crushers.

The size of the product required will determine whether tertiary crushers are required and the number and size of each type. It is customary in large plants to have both secondary and tertiary crushers.

The size of product required from the crushing section must be considered in conjunction with the grinding section. A two-stage rod mill and ball mill combination can accept a coarser crushing product than a single stage ball mill.

The grinding sections in the Placer plants have been a rod mill and ball mill combination. We have tended to provide a greater proportion of rod mill capacity than most plants, and I consider this gives smoother operation and relieves the demand on the crushing plant. We also tend to accept a coarser crusher product.

For Gibraltar we have gone to the extreme by not providing tertiary crushers, though we have put the secondary crushers in closed circuit with 1½" slotted screens. That operation has been successful.

I recall an old statement, "It is cheaper to crush than to grind". This might have been true when ball mills were much smaller, with horsepower about the same as a crusher, but in these days of large grinding mills and no increase in crusher size, I do not think this statement is true.

For Gibraltar, the capital cost of the secondary crushing plant was $7 million for four 7 ft. crushers, with foundations, building, conveyors, and screens. This did not include the cost for any ore storage.

The Marcopper crushing plant built two years earlier cost $4 million for the same number of crushers.
Each of these crushers has a 350 HP motor with an average load of 300 HP, making 1,200 HP for the whole plant, with an operating time of 70% to 75% for seven days a week, which is normal for a crushing plant.

The Gibraltar grinding section cost $10 million. The grinding mills consume 13,000 HP and have an operating time of 95% to 97% for seven days a week.

The following table gives the comparison of capital and operating (1973) costs for Gibraltar crushing and grinding related to the power used directly for the crushers and the grinding mills.

<table>
<thead>
<tr>
<th></th>
<th>Crushing</th>
<th>Grinding</th>
</tr>
</thead>
<tbody>
<tr>
<td>Operating time</td>
<td>75%</td>
<td>95%</td>
</tr>
<tr>
<td>Capital $ per KW</td>
<td>7,900</td>
<td>1,000</td>
</tr>
<tr>
<td>Capital $ per KW</td>
<td>10,500</td>
<td>1,050</td>
</tr>
<tr>
<td>Operating %</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Operating KWH per ton</td>
<td>0.4</td>
<td>5.6</td>
</tr>
<tr>
<td>Operating cents per ton</td>
<td>7</td>
<td>26</td>
</tr>
<tr>
<td>Operating cents per KWH</td>
<td>17.5</td>
<td>4.6</td>
</tr>
</tbody>
</table>

This comparison is for seven days a week operation for both crushing and grinding. If crushing were limited to five days a week the relative capital cost for crushing increases still further, with no change in operating cost.

In that overlapping size area where it is possible to crush and also it is possible to grind, I conclude that it is cheaper to grind, and that crushing should be carried no further than is necessary to provide suitable feed for the grinding mills.

Further, it has been my observation that grinding capacity follows the prediction of the Bond formula as the fourth root of two, which is 1.19, for each screen size, so that the grinding section must be enlarged by 19% to grind one screen size finer. Crushing capacity appears to follow closer to the square root of two, which is 1.41, for one screen size, so that the final stage of crushing must be increased by 41% for one screen size finer and doubled for two screen sizes finer.

Support for this statement is shown in the following table compiled from information given in a Symons crusher catalogue.
Capacity Rating for Symons 7 Ft. S.H. Crusher

<table>
<thead>
<tr>
<th>Product Size</th>
<th>T.P.H. Symons Table</th>
<th>Ratio Symons Table</th>
<th>T.P.H. Using Ratio of $\sqrt{2}$</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 in.</td>
<td>550</td>
<td>1.31</td>
<td>596</td>
</tr>
<tr>
<td>3/4 in.</td>
<td>420</td>
<td>1.40</td>
<td>423</td>
</tr>
<tr>
<td>1/2 in.</td>
<td>300</td>
<td>1.25</td>
<td>300</td>
</tr>
<tr>
<td>3/8 in.</td>
<td>240</td>
<td>1.50</td>
<td>212</td>
</tr>
<tr>
<td>1/4 in.</td>
<td>160</td>
<td></td>
<td>150</td>
</tr>
</tbody>
</table>

Average Ratio 1.36

Note 1. The T.P.H. is net finished product.
2. The Ratio is calculated from T.P.H. given in Symons Table.
3. The last column has been calculated using the ratio of $\sqrt{2}$ from the base of 300 T.P.H. for 1/2 in. product.

The secondary crushing plant contains several crushers with screening after each stage to separate the finished product, and pass the oversize to the next stage. This involves considerable conveying equipment to take the products from one stage to the next.

Following is the distribution of the installed cost for the Gibraltar secondary crushing plant.

Building (includes all concrete & sub-floors) 28%
Crushers 12%
Screens 2%
Electrical (conveyor motors not included) 10%
Conveyors (includes galleries and electrical motors) 39%
Other (includes chutes) 9%

Most machine sizes these days are becoming larger, but this has not been true for crushers. The largest size in operation is 7 ft. with a capacity of 100 to 600 tons per hour in the final stage, depending on the product size. There are three 10 ft. size machines being manufactured and I am sure there will be great interest in these when they go into operation.
The sizes of screens have shown an increase in the last ten years. I believe Craigmont in 1960, installed the first 8 ft. wide screen in North America. Screens 8 ft. wide are now common with several manufacturers. These 8 ft. screens have simplified plant layout in that one 8 ft. machine is easier to incorporate in a plant design than two five or six ft. of equivalent capacity.

5.06 Fine-Ore Storage

All ore storage is expensive and therefore the size should be no more than necessary for a satisfactory buffer between crushing and grinding. In large plants with both crushing and grinding on a seven day week, a twenty-four hour buffer of live ore storage should be sufficient.

The least expensive storage seems to be a stockpile on a concrete slab. An 'A' frame roof structure provides rain and snow protection to the live portion of the pile and supports the tripper conveyor. Draw holes in the slab, with feeders and conveyors in tunnels below reclaim the ore. This form of storage permits a bulldozer to be used to clear away hung-up ore and 'rat holes'.

Opinions will differ on the type of feeder to have under a fine-ore storage system. They may be pipe feeders, slot feeders, or mechanical feeders. A gyrating conical bottom is also a contender, though still not quite proven.

5.07 Grinding Section

In all the Placer plants to date we have chosen a rod mill and ball mill combination. Except for Craigmont, one grinding section has consisted of one rod mill and one ball mill of the same size. This arrangement has proved satisfactory for the relatively coarse final product that has been accepted. It has been satisfactory for the design product size of 80% passing 100 mesh and has permitted increasing the feed rate and accepting a product size of 80% passing 48 mesh. The acceptance of this product size, two screen sizes coarser, has multiplied the feed rate by the square root of two, which gives an increase of 41%, and this is the explanation of the large increase in throughput at Gibraltar over the published design capacity, from 30,000 T.P.D. to 42,000 T.P.D.
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This size relationship between rod and ball mills at Gibraltar has also permitted the rod mill to handle the coarse rod mill feed with negligible tramp oversize in the rod mill discharge. The rod mill feed is the undersize from 1 1/4" slotted screens in the secondary crushing plant and has screen analysis of 30 to 35% plus 3/4".

For a plant that requires grinding finer than 80% passing 100 or 150 mesh, the ball mill capacity would need to be greater relative to the rod mill, either by one larger ball mill or two ball mills.

Following are the sizings of flotation feed for Placer mills.

<table>
<thead>
<tr>
<th></th>
<th>Plus 48</th>
<th>Plus 65</th>
<th>Plus 100</th>
</tr>
</thead>
<tbody>
<tr>
<td>Craigmont</td>
<td>13</td>
<td>25</td>
<td></td>
</tr>
<tr>
<td>Endako</td>
<td>15</td>
<td>27</td>
<td></td>
</tr>
<tr>
<td>Marcopper</td>
<td>20</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Gibraltar</td>
<td>18</td>
<td>30</td>
<td></td>
</tr>
</tbody>
</table>

The speed of our mills is higher than most installations. We have been increasing the speed of 13 1/2 ft. diameter rod mills in steps and observing the results. Following are the speeds tabulated.

- **Marcopper**
  - Original installation: 64% critical
  - Increased to: 68% critical
  - Planned increase to: 72% critical

- **Gibraltar**
  - Original installation: 67% critical
  - Increased to: 74% critical
  - Planned trial increase to: 78% critical

Power has increased proportional to the speed. Gibraltar has had twelve months operation with one mill at the higher speed. Indications are that the additional grinding accomplished is in line with the power increase.

All the ball mills are at 78% critical speed.

### 5.08 Regrind Mill

Middling grains that have floated in the rougher cells require further grinding to free the valuable mineral from gangue and thus permit an improvement in the grade of final concentrate. A regrind mill is required for this purpose.
There are differences of opinion as to where the regrind mill should go in the flowsheet. It may be used to grind the whole of the rougher concentrate, or the cleaner tailing, or some other intermediate product.

I prefer to send the whole of the rougher concentrate to a cyclone in closed circuit with a regrind mill. This will grind every middling grain that is coarser than the set classification size. It also will grind every free mineral grain to the same size.

The rougher concentrate could contain a significant proportion of coarse, free mineral grains that do not require further grinding. Finer grinding of these grains may increase losses in cleaning. It is also a consequence of regrinding that the finer the concentrate, the higher the moisture content of the filter cake.

If instead, the regrinding is done on a cleaner tail product, or on an intermediate product after removing some of the free mineral, the larger free mineral grains will have the opportunity of reaching the final concentrate at a coarser size, but also there is the danger that some of the middling grains might be equally as readily floatable and reach final concentrate without having a chance to be regrinded.

There is also the possibility that perhaps there should be two regrind units, one for rougher concentrate and one for scavenger concentrate. I have not had an opportunity to pursue this.

Estimating the size for the regrind mill in design stages is a real guessing game. There is a great leeway in estimating the tons of material to be regrinded and also the hardness of it. A larger mill would have been an advantage at both Marcopper and Gibraltar.

Following is the concentrate sizing from our mills.

- Craigmont: 77% minus 44 microns (325 mesh)
- Marcopper: 70% minus 44 microns
- Gibraltar: 70% to 80% minus 44 microns
- Endako: 80% minus 12 microns

5.09 Flotation Section

The usual flowsheet for this section includes one or more banks of rougher cells followed by one or more stages of cleaning. There is a choice of planning for discarding the first cleaner tailing, or of returning it to the head of the roughers. Our choice has been to discard the first cleaner tailing.
Deciding on the retention time to be allowed for flotation is always a weighty question and perhaps a good guess is better than a bad calculation. I have arbitrarily chosen 15 minutes as a reasonable time for the ores Placer has treated.

Flotation machines were set at a relatively small size for many years. The largest size was 40 cu.ft., for one manufacturer, 100 cu.ft. for another. The 40 cu.ft. machine was by far the more popular cell. This necessitated multiple banks of cells and distributors.

I don't like distributors. It is difficult to get one to split large flotation flows evenly. They are clumsy, heavy, and need considerable supporting structure.

Now larger and deeper flotation cells are available, up to 600 cu.ft. and 6'-6" deep. It is possible to pass 15,000 tons per day through one single bank of rougher cells, with elimination of distributors, a saving in floor space, and a reduction in the places requiring operator attention.

Slurry pumps are important items in a flotation section. They must be able to cope with the whole range of the variations in operating conditions encountered, or results will suffer and the basement floor will be flooded with froth and pulp.

In sizing pumps in the concentrator a great deal of reliance is placed on experience in similar operations. The figures on the material flow balance cannot be taken as absolute for some flows. It is known that some flotation froth streams can increase and diminish rather quickly when an operator turns an air valve.

For pumps handling froth, an arbitrary froth factor is applied to the gallons per minute of slurry to allow for the effect the froth with entrained air will have on the performance of the pump that will handle it. It is hard to predict a value for this factor, which might be as high as three or even higher, depending on the nature of the froth. It also is a very good guessing game.

For centrifugal pumps handling concentrate thickener underflow, a factor is required for the effect of viscosity on the power required. This factor is of the order of four. For Gibraltar the supplier recommended a 7.5 HP motor. We installed a 30 HP motor, which is well loaded.
5.10 Concentrate Handling

This section will consist of concentrate thickeners, filters, perhaps a dryer, storage and loading facilities.

For concentrate thickener area I have used 5 to 10 sq.ft. per ton per day. The bulk of the concentrate will thicken with much less area, but a slight murkiness of this valuable product will add up to an appreciable loss. It is better to be on the generous side.

For the filter area I have used 2 sq.ft. per ton per day. Generally two filters have been installed, allowing for one in operation and one standby for normal operation. During periods of above average mill head and increased mill throughput, both filters can be operated together for the time required.

Concentrate drying and concentrate storage must be planned according to the requirements at each location.

5.11 Tailing Disposal

We will end the flowsheet with tailing disposal.

Storage of tailing and reclamation of water must be provided to eliminate pollution of the streams and countryside.

Another reason for water reclamation is to reduce the requirement of new water, which is likely to cost more than reclaimed water, or might not be available in the quantity required.

I shall not cover the problems in the design of tailing storage. It is general to provide a dam or dams to enclose a settlement area where the solids will settle, and provide a method of returning the free water back to the system. The storage area is generally downstream from the concentrator and at a considerably lower elevation.

After the starter dam is made of suitable borrow material, the dam may be raised continually either by deposition of the sand portion from the tailing or by borrow material or mine waste rock.

For water reclamation there is the question of whether there should be a tailing thickener at the concentrator.

The advantages of a tailing thickener at the concentrator are as follows:
1. Immediate availability of reclaim water with the minimum pumping head for returning it to the system.
2. A smaller volume of thickened tailing for transportation to the tailing storage area, thus requiring a smaller diameter pipe.
3. A smaller volume of reclaimed water for return from the tailing storage area, thus requiring a smaller diameter pipe and a lower capacity pumping station.
4. Reduced water loss from seepage or from evaporation in hot climates, where the availability of new water is restricted or very costly.

The disadvantages of a tailing thickener are as follows:
1. The capital cost of the thickener installation and the operating cost, particularly if chemicals must be used as settling aids.
2. When sand from the tailing is used for raising the dams, additional water may be required to obtain adequate classification for producing sand.

An engineering study should be made to compare the economics of the alternatives for each particular situation, with and without a tailing thickener.
6.0 FEASIBILITY STUDY

To bring a property into production there is a large capital expenditure required for detailed design, purchase of equipment, and construction. Before a commitment can be made on this expenditure a feasibility study will be made.

The purpose of this study is two-fold. First to satisfy management that the operation will be successful and profitable. Second to convince those providing the finance, generally banks, of the same things before financing can be obtained.

The feasibility report for Gibraltar consisted of six volumes:

Volume 1 Geology Volume
   Drilling results (core footage, assays)
   Calculated geologic reserves, which includes all grade classifications, not necessarily to be mined.

Volume 2 Mining Volume
   Mining Method
   Mineable ore reserves.
   Production schedules, annual
   Mining equipment selection

Volume 3 Metallurgy Volume
   Metallurgical test results
   Predicted plant metallurgical results
   Reagent and grinding media consumption
   Flowsheet
   Milling equipment list

Volume 4 Capital Cost Volume
   Cost estimate for mine, mill and plant
   General arrangement drawing for whole plantsite showing location of various facilities from the mine to the tailing storage.
   General arrangement drawings of mill layout.

Volume 5 Cash Flow Volume
   Operating cost estimate
   Estimated cash flow for the life of the property, but close detailed for first ten years.
Volume 6  Marketing Volume

Projected metal price and sales.
Sales agreements, including net smelter return and freight.
7.0 DESIGN AND CONSTRUCTION

7.01 General

Following on from the feasibility study, a decision is made whether to proceed with financing arrangements and bring the property into production.

The decision to proceed is made first by management, then submitted as a recommendation to the Board of Directors for their approval.

The Gibraltar cost estimate for the whole property was $74 million and the actual cost incurred was $68 million.

Generally, financing is by a bank loan, and a realistic interest rate these days is 10% or higher. On a loan of $70 million, this would amount to $7 million a year, which is about $20 thousand a day, or $800 per hour.

All this is not an immediate charge, but, as time goes on, the charge will increase to reach this figure. With a continuing charge like that, time to get into operation is really important.

There is both the direct interest charge and also the loss in cash revenue that is not flowing. An operation such as Gibraltar, which could gross $50 million a year, would lose $130 thousand a day, or $6,000 per hour in cash revenue.

Every ton not treated today is not treated tomorrow; it is not treated until the end of the life of the property. At 10% interest, the present value of one dollar due twenty years from now is only 15 cents.

Therefore time now is of the essence.

SLIDE 6 shows the historical record of the major phases for Gibraltar.

For the Gibraltar project, it is of interest that, during August 1970, we were given a tentative go ahead pending the signing of a concentrate sales contract and being able to arrange finance.

As these two problems could not be resolved before mid-November, it was decided to commit $1.5 million for the period August to December to assure that the schedule of July 1972 start-up could be achieved.

It was realized that this money would be a direct loss if the financing and sales contract could not be arranged. As it turned out, this early expenditure was probably the most significant factor in being able to improve on the schedule and achieve a March 1972 start-up.
The tax free period was another incentive for an early start-up. This tax free period would end December 1973 regardless of start-up date. Every day earlier start-up was one more day of operation in the tax free period.

7.02 Organization, Responsibility, Supervision

It takes the utmost cooperation between the owner, the designer, and the construction contractor to accomplish the desired result within the time schedule and the budget.

In our experience this is done by forming a group, in effect a separate team devoted entirely to the project. Each member contributes his expertise to his area of responsibility in a cooperative manner with all other members of the team.

This team includes people from Placer, from the design consultant, and from the construction contractor, all working closely together with free discussion and interchange of ideas as each phase or problem is developed. There has not been a clash of personalities or conflict of interest, and no uncertainty as to the person whose responsibility it was to make each decision.

The Placer plants have all been designed with Wright Engineers as the Civil-Mechanical Design Consultant, M.A. Thomas and Associates as the Electrical Design Consultant, and Commonwealth Construction as the General Contractor.

Placer Development, through the Project Developments Department, keeps close liaison with the consultants and contractor to coordinate and direct the activities. Furthermore, and of paramount importance, the Placer Project group has complete authority to make all decisions with respect to design, equipment selection, and construction. Centralizing this authority eliminates time-consuming and confusing chains of command.

The general concept and layout has been developed by Placer Engineering and submitted in the feasibility report.

The design consultant then proceeds with detailed engineering design and layout. The initial concept may be modified or changed if there is an advantage. At all stages, Placer Engineering keeps close liaison with the design consultant. We see all drawings from the preliminary stage through revisions to the final approval issue.

Frequent meetings and discussions are held between the consultant's group and the Placer representatives involved in the particular design or construction problem, to reach mutual agreement on the best solution.
There are also frequent general meetings attended by all member of the team that might be affected by the outcome of the meeting, to discuss general progress and any particular aspects that need to be discussed, in order to provide full communications and to avoid creating a new problem through an inappropriate decision or lack of communication.

For the Gibraltar project design, there were a total of 151,000 man hours for 3,100 drawings. This included 27,000 man hours for 700 electrical drawings. The basis for charges for consultants work is an hourly rate for the man doing the work plus an agreed surcharge for overhead and profit.

We have enjoyed a good working relationship with Wright Engineers and M.A. Thomas. It is an advantage that our offices are within walking distance so that we can keep in close touch with all phases of the design as it develops.

For plant construction, we have used Commonwealth Construction as the general contractor. Placer has a project supervisor on the property throughout the construction, to whom the contractor is responsible.

There is also close liaison between the Project Supervisor in the field and Placer Engineering in Vancouver, so that problems that arise can be settled with the minimum delay.

The basis of our agreement with the contractor is a fixed fee and target unit cost arrangement. The project is divided into small sections, and unit target costs are negotiated for each, prior to the start of construction. On the overall completed project, any overrun is absorbed by the owner, any underrun is shared by both the owner and the contractor. This arrangement permits an early start on construction and dispenses with messy negotiations for changes or additions to the work.

The agreement is reached with the contractor in the early stages of design when there is very little detailed design available. Not uncommonly, we involve the contractor during the final stages of the feasibility study, to discuss the scope of the work, invite recommendations and establish a basis for preparation of target cost estimates.

7.03 Equipment Procurement

The type and sizing of most of the equipment has been specified on the flowsheet and the equipment list in the feasibility study.
For each piece of equipment, Wright Engineers prepares a draft specification, which Placer Engineering reviews and amends if necessary. The final specification is distributed by Placer Purchasing to suppliers for quotation.

Quotations received are analysed by Wright Engineers, who make a recommendation. The final decision is made by Placer Engineering on the basis of suitability for the job, cost, delivery time, and ability to provide ongoing service.

Placer Engineering conducts final negotiations with the supplier to examine or amend any of the specifications in the proposal and, where there is an option, to decide exactly what components or specifications are required, with a price adjustment if applicable. The delivery date may also be negotiated or amended at this time.

Placer Engineering prepares a requisition setting out clearly the equipment to be ordered, referring to specifications in the proposal or as amended, with agreed price and delivery date, and forwards this to Placer Purchasing.

Placer Purchasing places orders on the basis F.O.B. Factory. Placer Purchasing arranges the freight handling, which may include freight assembly from several manufacturers in an area to make up a carload or truckload shipment if necessary.

Suppliers are required to make reports at agreed intervals showing the status of manufacture and changes in expected delivery date if any, and give adequate reasons for the change.

Purchasing expedites and coordinates the delivery of every order to the mine site and issues a report every two weeks listing all orders outstanding, giving the current status and the expected delivery date.
<table>
<thead>
<tr>
<th>Mine</th>
<th>Location</th>
<th>Production Date</th>
<th>Achieved Capacity TPD</th>
<th>Ore Type</th>
</tr>
</thead>
<tbody>
<tr>
<td>Craigmont</td>
<td>B.C.</td>
<td>Sept. 1961</td>
<td>5,000</td>
<td>Copper</td>
</tr>
<tr>
<td>Endako</td>
<td>B.C.</td>
<td>May 1965</td>
<td>15,000</td>
<td>Molybdenite</td>
</tr>
<tr>
<td>Endako Expansion</td>
<td>B.C.</td>
<td>Dec. 1967</td>
<td>10,000</td>
<td>Molybdenite</td>
</tr>
<tr>
<td>Cortez</td>
<td>Nevada</td>
<td>Jan. 1969</td>
<td>1,700</td>
<td>Gold</td>
</tr>
<tr>
<td>Marcopper</td>
<td>Philippines</td>
<td>Oct. 1969</td>
<td>22,000</td>
<td>Copper</td>
</tr>
<tr>
<td>Gibraltar</td>
<td>B.C.</td>
<td>March 1972</td>
<td>43,000</td>
<td>Copper &amp; Molybdenite</td>
</tr>
</tbody>
</table>

**PLACER IN CONJUNCTION WITH NORANDA MINES**

| Mattagami       | Quebec   | Oct. 1963       | 4,000                 | Copper & Zinc |
APPLICATION OF BOND FORMULA

POWER INCREASE REQUIRED TO GRIND ONE SCREEN SIZE FINER

1. \[ W = 10W_i \left( \frac{1}{\sqrt{P}} - \frac{1}{\sqrt{F}} \right) \]

   Bond Formula

   All sizing is 80% Passing Size
   \( W_i = \text{KWH/Ton from infinite size to 100 micron} \)
   \( F = \text{Feed Size Micron} \)
   \( P = \text{Product Size Micron.} \quad P_{100} = \text{Micron size for 100 mesh product.} \)
   \( W = \text{KWH/Ton to Grind from F to P.} \quad W_{100} = \text{the power for 100 mesh product.} \)

   At \( F_{0.75} \) inch \( \frac{1}{\sqrt{F}} = 0.007 \)

   At \( P_{100} \) mesh \( \frac{1}{\sqrt{P}} = 0.082 \)

   \( \frac{1}{\sqrt{F}} \) is small compared to \( \frac{1}{\sqrt{P}} \) and may be neglected.

2. \[ W = \frac{K}{\sqrt{P}} \] approximated E.Q. 1.

3. \[ P_{65} = P_{100} \times (2)^{0.5} \] Tyler Screen Series

4. \[ W_{100} = \frac{K}{\sqrt{P_{100}}} \] Substitution in E.Q. 2.

5. \[ W_{65} = \frac{K}{\sqrt{P_{65}}} = \frac{K}{\sqrt{P_{100} \times (2)^{0.5}}} = \frac{K}{\sqrt{P_{100} \times (2)^{0.25}}} \] E.Q. 2 & 3

6. \[ \frac{W_{100}}{W_{65}} = \frac{K}{\sqrt{P_{100}}} \times \frac{\sqrt{P_{100}} \times (2)^{0.25}}{K} = (2)^{0.25} \] E.Q. 4 & 5

7. \[ \frac{W_{100}}{W_{65}} = (2)^{0.25} = 1.19 = 119\% \] E.Q. 6 cont.
SLIDE 3

BOND FORMULA CONTINUED

TONNAGE INCREASE OBTAINED BY ACCEPTING ONE SCREEN SIZE COARSER

Plant Treating 100 TPH at $P_{100}$

What will be treatment rate at $P_{65}$

$W_{100} = \text{KWH/Ton to 100 mesh}$

$W_{65} = \text{KWH/Ton to 65 mesh}$

7. \[
\frac{W_{100}}{W_{65}} = 1.19 = 119\% \quad \text{Slide 2}
\]

8. \[
100 \text{ TPH} \times \frac{W_{100}}{W_{65}} = 119 \text{ TPH} \quad \text{New Rate}
\]

SUMMARY

\begin{align*}
(2)^{0.25} &= 1.19 = 119\% \quad \text{One Screen Size Coarser} \\
(2)^{0.50} &= 1.41 = 141\% \quad \text{Two Screen Sizes Coarser} \\
(2)^{0.75} &= 1.68 = 168\% \quad \text{Three Screen Sizes Coarser} \\
(2)^{1} &= 2.00 = 200\% \quad \text{Four Screen Sizes Coarser}
\end{align*}
### SLIDE 4

**GRINDING INFLUENCE ON REVENUE**

Ore grade 0.4% Cu. Work index 10.0. Grinding from 0.75 in. All sizes given 80% passing. Grinding Cost $0.04 per KWH.

<table>
<thead>
<tr>
<th>Grind Mesh</th>
<th>KWH/Ton</th>
<th>Grinding $/Ton</th>
<th>Recovery %</th>
<th>Tail % Cu.</th>
<th>NSR/T Gross $</th>
<th>NSR/T Minus Grind</th>
</tr>
</thead>
<tbody>
<tr>
<td>35</td>
<td>4.2</td>
<td>0.168</td>
<td>56.0</td>
<td>0.176</td>
<td>2.24</td>
<td>2.07</td>
</tr>
<tr>
<td>48</td>
<td>5.1</td>
<td>0.204</td>
<td>63.8</td>
<td>0.145</td>
<td>2.55</td>
<td>2.35</td>
</tr>
<tr>
<td>65</td>
<td>6.2</td>
<td>0.248</td>
<td>71.7</td>
<td>0.113</td>
<td>2.86</td>
<td>2.61</td>
</tr>
<tr>
<td>100</td>
<td>7.5</td>
<td>0.300</td>
<td>75.5</td>
<td>0.098</td>
<td>3.02</td>
<td>2.72</td>
</tr>
<tr>
<td>150</td>
<td>9.1</td>
<td>0.364</td>
<td>79.5</td>
<td>0.082</td>
<td>3.18</td>
<td>2.82</td>
</tr>
<tr>
<td>200</td>
<td>10.9</td>
<td>0.436</td>
<td>81.5</td>
<td>0.074</td>
<td>3.26</td>
<td>2.82</td>
</tr>
<tr>
<td>270</td>
<td>13.0</td>
<td>0.520</td>
<td>84.0</td>
<td>0.064</td>
<td>3.36</td>
<td>2.84</td>
</tr>
</tbody>
</table>

NSR/T = Net smelter return per ton of ore, based on NSR $0.50 per lb. Cu.
### SLIDE 5

**GRINDING INFLUENCE ON REVENUE (CONTINUED)**

Metallurgy & Grinding Costs from Slide 4.

Other operating costs assumed fixed at $1.00 per ton.

**NSR/T Gross** = NSR $ per ton of ore from Slide 4.

**NSR/T Net** = **NSR/T Gross** minus grinding and other operating cost per ton of ore.

Relative profit = **NSR/T Net** x relative tons.

<table>
<thead>
<tr>
<th>Grind Mesh</th>
<th>Relative Tons</th>
<th>Grinding $/Ton</th>
<th>Other Operating $</th>
<th>NSR/T Gross $</th>
<th>NSR/T Net $</th>
<th>Relative Profit</th>
</tr>
</thead>
<tbody>
<tr>
<td>35</td>
<td>282</td>
<td>0.168</td>
<td>1.00</td>
<td>2.24</td>
<td>1.07</td>
<td>302</td>
</tr>
<tr>
<td>48</td>
<td>238</td>
<td>0.204</td>
<td>1.00</td>
<td>2.55</td>
<td>1.35</td>
<td>321</td>
</tr>
<tr>
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<td>1.00</td>
<td>2.86</td>
<td>1.64</td>
<td>328</td>
</tr>
<tr>
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<td>168</td>
<td>0.300</td>
<td>1.00</td>
<td>3.02</td>
<td>1.72</td>
<td>289</td>
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<tr>
<td>150</td>
<td>141</td>
<td>0.364</td>
<td>1.00</td>
<td>3.18</td>
<td>1.82</td>
<td>257</td>
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<tr>
<td>200</td>
<td>119</td>
<td>0.436</td>
<td>1.00</td>
<td>3.26</td>
<td>1.82</td>
<td>217</td>
</tr>
<tr>
<td>270</td>
<td>100</td>
<td>0.520</td>
<td>1.00</td>
<td>3.36</td>
<td>1.84</td>
<td>184</td>
</tr>
</tbody>
</table>
SLIDE 6

DESIGN CONSTRUCTION SCHEDULE

Gibraltar Mines Ltd. (N.P.L.)

Location: British Columbia, 40 miles north of Williams Lake.
Ore Reserve: 358 million tons, 0.373% Cu, 0.016% MoS₂.
Design Capacity: 30,000 T.P.D.

<table>
<thead>
<tr>
<th>Major Phases of Project</th>
<th>Starting Date</th>
<th>Month Elapsed Time</th>
</tr>
</thead>
<tbody>
<tr>
<td>Metallurgical Testing</td>
<td>October 1969</td>
<td>0</td>
</tr>
<tr>
<td>Engineering concept and design</td>
<td>August 1970</td>
<td>10</td>
</tr>
<tr>
<td>Clearing of plantsite and initial open pit area</td>
<td>October 1970</td>
<td>12</td>
</tr>
<tr>
<td>Installation of construction camp and stripping of open pit overburden</td>
<td>January 1971</td>
<td>15</td>
</tr>
<tr>
<td>Construction of crushing, milling and plant facilities</td>
<td>April 1971</td>
<td>18</td>
</tr>
<tr>
<td>Continuous operation, 30,000 T.P.D.</td>
<td>March 31, 1972</td>
<td>30</td>
</tr>
<tr>
<td>Plant throughput month average - 41,600 T.P.D.</td>
<td>June 30, 1972</td>
<td>33</td>
</tr>
</tbody>
</table>
SOME CRUSHING AND GRINDING CONSIDERATIONS
IN PLANT DESIGN

BY

H.A. STEANE
General Mill Superintendent
Placer Development Limited
Vancouver, B.C.

For Presentation to the
Canadian Mineral Processors
January 1976

Ottawa, Canada
Introduction

This presentation will cover some crushing and grinding considerations in plant design and operation. There is nothing basically new in the subject matter presented here; however, the viewpoint and interpretation may be novel and some of it controversial.

The presentation is based on experience gained in the metallurgical design of plants for Placer Development Limited from 1960 to the present. The plants are listed on SLIDE No. 1.

These plants have conventional crushing, followed by rod mill and ball mill grinding. I would like to think that autogenous grinding plants are the answer for eliminating secondary crushing plants and reducing the capital and operating costs; however, the autogenous plants seem to have some problems of their own and I have had no experience with them.

Fineness of Grinding Required

First let us consider the fineness of grinding required.

The fineness of grind will affect the recovery obtained, and the relationship must be determined. If the grinding is too coarse, some of the valuable mineral, either free or locked in middling grains, will not be floated. However, as grinding is probably the greatest single operating cost, we do not want to grind any finer than is justified economically.

Finer grinding should not be carried beyond the point where the net smelter return for the increment saved becomes less than the added operating cost. Even this fineness will be beyond the economic limit because of the additional capital cost for the equipment needed to achieve it. These factors will be discussed in more detail later.

The Bond work index is the commonest method of expressing a measure of the effort that has to be expended in grinding. It is the kilowatt-hours required to reduce a ton of rock from an infinite size to 80% passing 100 microns. The formula allows a calculation to be made for other size ranges.

Other investigators have done work and are continuing to try to develop other methods to determine the grinding effort required, but their formulas are difficult to use and I doubt whether the end result is any better.
Power Required. Increase for Finer Grinding

Let us examine the effect of the degree of grinding on the capacity of a grinding plant.

Those of you that have worked with the Bond formula may know this already; but those that are not familiar with the calculation may be surprised at the amount of additional power and plant that is required for grinding one screen-size finer.

On SLIDE No. 2, the Bond formula has been set out in equation 1 and the symbols defined.

After making one approximation, equations 2 to 7 proceed to show the power increase required to grind one screen-size finer. The example given is from 65 mesh to 100 mesh, but the same relationship will hold for any two consecutive screen-sizes in the square-root-of-two screen series, within the range of flotation feed sizing.

Conclusion. The power required will increase by the fourth root of two, which is 1.19, for each screen-size finer grind. That means a power and plant increase of 19% that must be provided in the design for one screen-size finer.

Plant Throughput. Increase by Accepting Coarser Grind

This study is an extension and a corollary to the thoughts expressed in the previous slide.

SLIDE No. 3 has been prepared to show the increase in capacity that can be obtained from an existing plant by accepting a product that is one screen-size coarser.

It assumes a plant treating 100 tons per hour, producing a product at 100 mesh. Starting with equation 7 from SLIDE No. 2, equation 8 is developed showing the increase in throughput by accepting a 65-mesh product, which is one screen-size coarser.

A summary tabulation follows, showing the increase in throughput successively for one, two, three, and four consecutive screen-sizes in the square-root-of-two screen series.

HAS/mt
December 18, 1975
This poses a policy decision for management on the conservation of resources as to whether the lower recovery is acceptable, even though it is the most profitable.

Secondary Crushing Section

Now let us consider the secondary crushing plant to feed the grinding plant.

The size of the product required will determine whether tertiary crushers are required and the number and size of each type. It is customary in large plants to have both secondary and tertiary crushers.

The size of product required from the crushing section must be considered in conjunction with the grinding section. A two-stage rod mill and ball mill combination can accept a coarser feed than a single-stage ball mill.

The grinding sections in the Placer plants have been a rod mill and ball mill combination. They have tended to provide a greater proportion of rod mill capacity than most plants, to give smoother operation and relieve the demand on the crushing plant. They also tend to accept a coarser crushe-product.

The Gibraltor plant has gone to the extreme by not providing tertiary crushers, though the secondary crushers have been put in closed circuit with 1½" slotted screens. That operation has been successful.

I recall an old statement, "It is cheaper to crush than to grind". This might have been true when ball mills were much smaller, with horsepower about the same as a crusher, but in these days of large grinding mills and no increase in crusher size, I do not think this statement is true.

For Gibraltor, the capital cost of the secondary crushing plant was $7 million for four 7 ft. crushers, with foundations, building, conveyors, and screens. This did not include the cost for any ore storage.

Each of these crushers has a 350 hp motor with an average load of 300 hp, making 1,200 hp for the whole plant, with an operating time of 70% to 75% for seven days a week, which is normal for a crushing plant.

The Gibraltor grinding section cost $10 million. The grinding mills consume 13,000 hp and have an operating time of 95% to 97% for seven days a week.

HAS/m
t
December 18, 1975
The following table gives the comparison of capital and operating (1973) costs for Gibraltar crushing and grinding related to the power used directly for the crushers and the grinding mills.

**Comparison of Gibraltar Crushing and Grinding Costs**

<table>
<thead>
<tr>
<th></th>
<th>Crushing</th>
<th>Grinding</th>
</tr>
</thead>
<tbody>
<tr>
<td>Operating time</td>
<td>75%</td>
<td>95%</td>
</tr>
<tr>
<td>Capital total $1,000</td>
<td>7,000</td>
<td>10,000</td>
</tr>
<tr>
<td>Horsepower consumption</td>
<td>1,200</td>
<td>13,000</td>
</tr>
<tr>
<td>Kilowatt consumption</td>
<td>895</td>
<td>9,698</td>
</tr>
<tr>
<td>Capital $ per kw</td>
<td>7,800</td>
<td>1,000</td>
</tr>
<tr>
<td>Capital $ per kw Operating%</td>
<td>10,400</td>
<td>1,050</td>
</tr>
<tr>
<td>Operating kw/hr per ton</td>
<td>0.4</td>
<td>5.6</td>
</tr>
<tr>
<td>Operating cents per ton</td>
<td>7</td>
<td>26</td>
</tr>
<tr>
<td>Operating cents per kw/hr</td>
<td>17.5</td>
<td>4.6</td>
</tr>
</tbody>
</table>

This comparison is for seven days a week operation for both crushing and grinding. If crushing were limited to five days a week the relative capital cost for crushing would increase still further, with no change in operating cost.

In that overlapping size area where it is possible to crush and also it is possible to grind, I conclude that it is cheaper to grind, and that crushing should be carried no further than is necessary to provide suitable feed for the grinding mills.

Further, it has been my observation that grinding capacity follows the prediction of the Bond formula as the fourth root of two, which is 1.19, for each screen-size, so that the grinding section must be enlarged by 19% to grind one screen-size finer. Crushing capacity appears to follow closer to the square root of two, which is 1.41, for one screen-size, so that the final stage of crushing must be increased by 41% for one screen-size finer and doubled for two screen-sizes finer.

Support for this statement is shown in the following table compiled from information given in a Symons crusher bulletin.

HAS/mt
December 19, 1975
Capacity Rating for Symons 7-Ft. S.H. Crusher

<table>
<thead>
<tr>
<th>Product Size</th>
<th>T.P.H. Symons Table</th>
<th>Ratio Symons Table</th>
<th>T.P.H. Using Ratio of $\sqrt{2}$</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 in.</td>
<td>550</td>
<td>1.31</td>
<td>596</td>
</tr>
<tr>
<td>3/4 in.</td>
<td>420</td>
<td>1.40</td>
<td>423</td>
</tr>
<tr>
<td>1/2 in.</td>
<td>300</td>
<td>1.25</td>
<td>300</td>
</tr>
<tr>
<td>3/8 in.</td>
<td>240</td>
<td>1.50</td>
<td>212</td>
</tr>
<tr>
<td>1/4 in.</td>
<td>160</td>
<td></td>
<td>150</td>
</tr>
</tbody>
</table>

Average Ratio 1.36

Note 1. The T.P.H. is the tons per hour of net finished product.
2. The Ratio is calculated from T.P.H. given in Symons Table.
3. The last column has been calculated using the ratio of $\sqrt{2}$ from the base of 300 T.P.H. for 1/2 in. product.

The secondary crushing plant contains several crushers with screening after each stage to separate the finished product, and pass the oversize to the next stage. This involves considerable conveying equipment to take the products from one stage to the next.

Following is the distribution of the installed cost for the Gibraltar secondary crushing plant.

Building (includes all concrete & sub-floors) 28%
Crushers 12%
Screens 2%
Electrical (conveyor motors not included) 10%
Conveyors (includes galleries and electrical motors) 39%
Other (includes chutes) 9%

Most machine sizes these days are becoming larger, but this has not been true for crushers. The largest size in operation is 7-ft. with a capacity of 100 to 600 tons per hour in the final stage, depending on the product size. There are three 10-ft. size machines being manufactured and I am sure there will be great interest in these when they go into operation.

HAS/mt
December 19, 1975
Grinding Section

Now let us return again to the grinding section. When accepting a coarser product from the crushing section, it is important that the rod mill be large enough relative to the ball mill so that the ball mill capacity becomes the limiting factor in grinding throughput and not the rod mill.

All the Placer plants to date have a rod mill and ball mill combination. Except for Craigmont, one grinding section has consisted of one rod mill and one ball mill of the same size. This arrangement has proved satisfactory for the relatively coarse flotation feed size that has been accepted. It has been satisfactory for the design product size of 80% passing 100 mesh and has permitted increasing the feed rate and accepting a product size of 80% passing 48 mesh. The acceptance of this product size, two screen-sizes coarser, has multiplied the feed rate by the square root of two, which gives an increase of 41%, and this is the explanation of the large increase in throughput at Gibraltar over the published design capacity, from 30,000 T.P.D. to 42,000 T.P.D.

This size relationship between rod and ball mills at Gibraltar has also permitted the rod mill to handle the coarse rod mill feed with negligible tramp oversize in the rod mill discharge. The rod mill feed is the undersize from 1½" slotted screens in the secondary crushing plant and has screen analysis of 30 to 35% plus 3/4".

For a plant that requires grinding finer than 80% passing 150 mesh, the ball mill capacity would need to be greater relative to the rod mill, either by one larger ball mill or two ball mills.

Following are the sizings of flotation feed for Placer mills.

<table>
<thead>
<tr>
<th></th>
<th>Plus 48</th>
<th>Plus 65</th>
<th>Plus 100</th>
</tr>
</thead>
<tbody>
<tr>
<td>Craigmont</td>
<td>13</td>
<td>25</td>
<td></td>
</tr>
<tr>
<td>Endako</td>
<td>15</td>
<td>27</td>
<td></td>
</tr>
<tr>
<td>Marcopper</td>
<td>20</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Gibraltar</td>
<td>18</td>
<td>30</td>
<td></td>
</tr>
</tbody>
</table>

HAS/mt
December 19, 1975
# SLIDE 1
## PLACER DESIGNED PROPERTIES

<table>
<thead>
<tr>
<th>Mine</th>
<th>Location</th>
<th>Production Date</th>
<th>Achieved Capacity TPD</th>
<th>Ore Type</th>
</tr>
</thead>
<tbody>
<tr>
<td>Craigmont</td>
<td>B.C.</td>
<td>Sept. 1961</td>
<td>5,000</td>
<td>Copper</td>
</tr>
<tr>
<td>Endako</td>
<td>B.C.</td>
<td>May 1965</td>
<td>15,000</td>
<td>Molybdenite</td>
</tr>
<tr>
<td>Endako Expansion</td>
<td>B.C.</td>
<td>Dec. 1967</td>
<td>10,000</td>
<td>Molybdenite</td>
</tr>
<tr>
<td>Cortez</td>
<td>Nevada</td>
<td>Jan. 1969</td>
<td>1,700</td>
<td>Gold</td>
</tr>
<tr>
<td>Marcopper</td>
<td>Philippines</td>
<td>Oct. 1969</td>
<td>22,000</td>
<td>Copper</td>
</tr>
<tr>
<td>Gibraltar</td>
<td>B.C.</td>
<td>March 1972</td>
<td>43,000</td>
<td>Copper &amp; Molybdenite</td>
</tr>
</tbody>
</table>

**PLACER IN CONJUNCTION WITH NORANDA MINES**

| Mattagami     | Quebec   | Oct. 1963       | 4,000                 | Copper & Zinc  |

HAS/mt
December 29, 1975
APPLICATION OF BOND FORMULA

POWER INCREASE REQUIRED TO GRIND ONE SCREEN-SIZE FENER

1. \[ W = 10W_i \left( \frac{1}{P} - \frac{1}{F} \right) \]
   Bond formula

   All sizing is 80% passing size

   \( W_i \) = kwhr/ton from infinite size to 100 micron

   \( F \) = Feed size micron

   \( P \) = Product size micron. \( P_{100} \) = Micron size for 100 mesh product.

   \( W \) = kwhr/ton to grind from \( F \) to \( P \). \( W_{100} \) is the power for 100 mesh product.

   At \( F_{0.75} \) inch \( \frac{1}{F} = 0.007 \)

   At \( P_{100} \) mesh \( \frac{1}{P} = 0.082 \)

   \( \frac{1}{F} \) is small compared to \( \frac{1}{P} \) and may be neglected.

2. \( W = \frac{K}{\sqrt{P}} \) approximated E.Q. 1.

3. \( P_{65} = P_{100} \times (2)^{0.5} \) Tyler screen series

4. \( W_{100} = \frac{K}{\sqrt{P_{100}}} \) Substitution in E.Q. 2.

5. \( W_{65} = \frac{K}{\sqrt{P_{65}}} = \frac{K}{\sqrt{P_{100}}} \times (2)^{0.5} = \frac{K}{\sqrt{P_{100}}} \times (2)^{0.25} \) E.Q. 2 & 3

6. \( \frac{W_{100}}{W_{65}} = \frac{K}{\sqrt{P_{100}}} \times (2)^{0.25} = (2)^{0.25} \) E.Q. 4 & 5

7. \( \frac{W_{100}}{W_{65}} = (2)^{0.25} = 1.19 = 119\% \) E.Q. 6 cont.

HAS/mt
December 29, 1975
SLIDE 3
BOND FORMULA CONTINUED

TONNAGE INCREASE OBTAINED BY ACCEPTING ONE SCREEN-SIZE COARSER

Plant treating 100 TPH at \( P_{100} \)

What will be treatment rate at \( P_{65} \)

\[ W_{100} = \text{kwhr/ton to 100 mesh} \]

\[ W_{65} = \text{kwhr/ton to 65 mesh} \]

7. \[ \frac{W_{100}}{W_{65}} = 1.19 = 119\% \] Slide 2

8. \[ 100 \text{ TPH} \times \frac{W_{100}}{W_{65}} = 119 \text{ TPH} \] New Rate

SUMMARY

\[
\begin{align*}
(2)^{0.25} & = 1.19 = 119\% \\
(2)^{0.50} & = 1.41 = 141\% \\
(2)^{0.75} & = 1.68 = 168\% \\
(2)^{1} & = 2.00 = 200\% 
\end{align*}
\]

One Screen-Size Coarser
Two Screen-Sizes Coarser
Three Screen-Sizes Coarser
Four Screen-Sizes Coarser

HAS/mt
December 29, 1975
SLIDE 4

GRINDING INFLUENCE ON REVENUE

Ore grade 0.4% Cu. Work index 10.0. Grinding from 0.75 in.
All sizes given 80% passing. Grinding cost $0.04 per kwhr.

NSR/T = Net smelter return per ton of ore, based on NSR $0.50 per lb. Cu.

<table>
<thead>
<tr>
<th>Grind Mesh</th>
<th>kwhr/Ton</th>
<th>Grinding $/Tonn</th>
<th>Recovery %</th>
<th>Tail % Cu.</th>
<th>NSR/T Gross $</th>
<th>NSR/T Minus Grind</th>
</tr>
</thead>
<tbody>
<tr>
<td>35</td>
<td>4.2</td>
<td>0.168</td>
<td>56.0</td>
<td>0.176</td>
<td>2.24</td>
<td>2.07</td>
</tr>
<tr>
<td>48</td>
<td>5.1</td>
<td>0.204</td>
<td>63.8</td>
<td>0.145</td>
<td>2.55</td>
<td>2.35</td>
</tr>
<tr>
<td>65</td>
<td>6.2</td>
<td>0.248</td>
<td>71.7</td>
<td>0.113</td>
<td>2.86</td>
<td>2.61</td>
</tr>
<tr>
<td>100</td>
<td>7.5</td>
<td>0.300</td>
<td>75.5</td>
<td>0.098</td>
<td>3.02</td>
<td>2.72</td>
</tr>
<tr>
<td>150</td>
<td>9.1</td>
<td>0.364</td>
<td>79.5</td>
<td>0.082</td>
<td>3.18</td>
<td>2.82</td>
</tr>
<tr>
<td>200</td>
<td>10.9</td>
<td>0.436</td>
<td>81.5</td>
<td>0.074</td>
<td>3.26</td>
<td>2.82</td>
</tr>
<tr>
<td>270</td>
<td>13.0</td>
<td>0.520</td>
<td>84.0</td>
<td>0.064</td>
<td>3.36</td>
<td>2.84</td>
</tr>
</tbody>
</table>

HAS/mt
December 29, 1975
Metallurgy & Grinding costs from Slide 4.

Other operating costs assumed fixed at $1.00 per ton.

NSR/T Gross = NSR $ per ton of ore from Slide 4.

NSR/T Net = NSR/T Gross minus grinding and other operating cost per ton of ore.

Relative profit = NSR/T Net x relative tons.

<table>
<thead>
<tr>
<th>Grind Mesh</th>
<th>Relative Tons</th>
<th>Grinding $/Ton</th>
<th>Other Operating $</th>
<th>NSR/T Gross $</th>
<th>NSR/T Net $</th>
<th>Relative Profit</th>
</tr>
</thead>
<tbody>
<tr>
<td>35</td>
<td>282</td>
<td>0.168</td>
<td>1.00</td>
<td>2.24</td>
<td>1.07</td>
<td>302</td>
</tr>
<tr>
<td>48</td>
<td>238</td>
<td>0.204</td>
<td>1.00</td>
<td>2.55</td>
<td>1.35</td>
<td>321</td>
</tr>
<tr>
<td>65</td>
<td>200</td>
<td>0.248</td>
<td>1.00</td>
<td>2.68</td>
<td>1.64</td>
<td>328</td>
</tr>
<tr>
<td>100</td>
<td>168</td>
<td>0.300</td>
<td>1.00</td>
<td>3.02</td>
<td>1.72</td>
<td>289</td>
</tr>
<tr>
<td>150</td>
<td>141</td>
<td>0.364</td>
<td>1.00</td>
<td>3.18</td>
<td>1.82</td>
<td>257</td>
</tr>
<tr>
<td>200</td>
<td>119</td>
<td>0.436</td>
<td>1.00</td>
<td>3.26</td>
<td>1.82</td>
<td>217</td>
</tr>
<tr>
<td>270</td>
<td>100</td>
<td>0.520</td>
<td>1.00</td>
<td>3.36</td>
<td>1.84</td>
<td>184</td>
</tr>
</tbody>
</table>
Canadian Exploration, Limited
Salmo, B.C.

Maps A and B, 22

By H. A. Steane, General Mill Supt.

Tonnage
1,700 tons per day. Five-day week operation.

Location
South from Salmo, 7 miles along Salmo-Nelson highway.

Ore
Galena, sphalerite, silver, pyrite, in dolomite limestone.

Metallurgy
1st September, 1955 to 31 May, 1956.

<table>
<thead>
<tr>
<th></th>
<th>Tons</th>
<th>Wt. %</th>
<th>Assays %</th>
<th>Distribution %</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td>Pb</td>
<td>Zn</td>
</tr>
<tr>
<td>Mill Feed</td>
<td>287,414</td>
<td>100.00</td>
<td>1.5</td>
<td>4.4</td>
</tr>
<tr>
<td>Pb con.</td>
<td>5,249</td>
<td>1.83</td>
<td>70.8</td>
<td>6.9</td>
</tr>
<tr>
<td>Zn con.</td>
<td>19,313</td>
<td>6.72</td>
<td>1.4</td>
<td>58.2</td>
</tr>
<tr>
<td>Tailing</td>
<td>262,852</td>
<td>91.45</td>
<td>0.14</td>
<td>0.35</td>
</tr>
</tbody>
</table>

Crushing

- Op'ing: +0.742" -0.525" +3M +10 +200 -200
- % Wt. 16.1 24.9 19.4 14.9 15.1 9.6
The Milling of Canadian Ores

Grinding

Primary, rod mill: 8-ft. x 12-ft., 18.0 r.p.m., 65.0 per cent critical, in open circuit. Grind 30 per cent, -200 mesh at 80 per cent solids. 3-in. rods at 0.2 lb. per ton. Classifier: 72-in. Wemco receives rod mill discharge, making finished overflow and sending sand to secondary mills.

Secondary, ball mill: 9-ft. x 48-in. conical, 22.8 r.p.m., 87.7 per cent critical in closed circuit with classifier No. 2; 2-in. balls at 0.45 lb. per ton. Ball mill: 6-ft. x 12-ft. cylindrical, 20.9 r.p.m., 65.0 per cent critical, in closed circuit with classifier No. 3; 2-in. balls at 0.45 lb. per ton. Classifier No. 2: 72-in. Wemco. Classifier No. 3: 45-in. Akins.

Combined overflows from three classifiers, 33 per cent solids.

Mesh  +65  +100  +150  +200  -200
% Wt.  5.0  10.0  15.0  10.0  60.0

Regrind, ball mill: 7-ft. x 36-in. conical mill, 25.0 r.p.m., 84.2 per cent critical in closed circuit with 12-in. Dorrecone. Feed to this unit is zinc rougher concentrate. Balls size 1-in., at 0.10 lb. per ton of concentrator feed. As a result of this regrind unit, the screening on the final zinc concentrate is 93 per cent -325 mesh.

Flotation

**Lead Circuit**

<table>
<thead>
<tr>
<th>Reagent</th>
<th>Lb./Ton</th>
<th>Added to</th>
</tr>
</thead>
<tbody>
<tr>
<td>NaCN</td>
<td>0.10</td>
<td>Ball mill</td>
</tr>
<tr>
<td>Aerofloat 25</td>
<td>0.16</td>
<td>Rod mill &amp; cells</td>
</tr>
<tr>
<td>Zinc sulphate</td>
<td>0.09</td>
<td>Final cleaner</td>
</tr>
<tr>
<td>Flotation contact time — 15 min.</td>
<td></td>
<td></td>
</tr>
<tr>
<td>pH — 8.5 (natural)</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**Zinc Circuit**

<table>
<thead>
<tr>
<th>Reagent</th>
<th>Lb./Ton</th>
<th>Added to</th>
</tr>
</thead>
<tbody>
<tr>
<td>Lime</td>
<td>0.35</td>
<td>Feed pump</td>
</tr>
<tr>
<td>CuSO₄ ( \cdot )H₂O</td>
<td>0.36</td>
<td>Conditioner</td>
</tr>
<tr>
<td>Z-11</td>
<td>0.07</td>
<td>Feed pump &amp; cells</td>
</tr>
<tr>
<td>Methyl Amyl Alcohol</td>
<td>0.15</td>
<td>Conditioner &amp; cells</td>
</tr>
<tr>
<td>NaCN</td>
<td>0.02</td>
<td>Con. regrind mill</td>
</tr>
<tr>
<td>Conditioning time — 2 min.</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Flotation contact time — 12 min.</td>
<td></td>
<td></td>
</tr>
<tr>
<td>pH — 9.8 to 10.2</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Power

West Kootenay, hydro-electric, 20 k.w.h. per ton.
Crew
 Crushing and conveying, two-shift operation for both lead-zinc and tungsten ores:

Milling:

| Operators | 8 |
| Repairs   | 1 |
| Staff     | 1 |
| **Total** | **10** |

| Operating | 9 |
| Tailing Dam | 3 |
| Repairs   | 5 |
| Staff     | 5 |
| **Total** | **22** |

Features

The crushing plant is underground at the mine, and is used for crushing both the lead-zinc ore and the tungsten ore.

The lead-zinc ore, after crushing, passes over a conveyor system 7,050 ft. long, consisting of seven belt sections, and four vertical underground ore-storage pockets. The point of delivery to the mill is 1,800 ft. below the mine portal.

The crusher operation is two shifts per day, one for lead-zinc ore, and one for tungsten ore.

Both lead and zinc concentrate is loaded onto trucks by a front-end loader, weighed and trucked to Salmo, where it is loaded on railroad cars.

Lead thickener overflow is returned to the lead cleaner tailing. Lead filtrate is returned to the lead final cleaner cell.

Zinc thickener overflow is returned to the zinc conditioner. Zinc filtrate is returned to the zinc final cleaner cell.

Automatic samplers are installed for final tailing and both final concentrates.

Tailing is impounded on a 28-acre lot, across the highway adjacent to the mill. The flow and distribution system is by gravity, through 10-in. wood-stave pipe, set level, and completely surrounding the area. Due to the rate of rise of the impounded tailing, it will be necessary to pump the tailing commencing in the spring of 1957.

The mill building is heated by eight 20 k.w. heaters during the week-end shut-down. Heating seldom is necessary when the mill is operating.

History

1900, prospecting done on the property.

1908, John Waldbeser was the first recorded owner of claims, lead-zinc.

1910, Iron Mountain Limited, a subsidiary of Pacific Coast Steel was organized with J. Waldbeser as manager.

1915, Iron Mountain Limited shipped 12,000 tons of lead ore.

1917-1919, Three short adits were driven at the present 4,100 level of the present Jersey Track area.

1919-1920, Concentrator, 25-ton per day, lead-zinc ore, constructed and placed in operation.

1924, Minor exploration and geological work done.

1926, Production and development suspended because of low prices for metals.

1934, Forest fire destroyed the mill and burned most of the timber in the area.

1936-1941, Further exploration on the property was done by Harold Lakes, and a detailed map was prepared showing all outcrops on the property from the Dodge claims on the north, to the Jersey claims on the south. Up to this time, interest had been confined to lead-zinc.

1942, Scheelite deposits were recognized and exploration continued under Harold Lakes. In September, 1942, Wartime Metals took over Iron Mountain Limited.

1943, Under Wartime Metals, the tungsten find of Harold Lakes was developed to production and milling in 14 months, and operated 6 weeks, to close in November, 1943. The 300-ton per day mill was on the site of the present lead-zinc mill.

1947, Canadian Exploration Limited took over the ownership and active mining and milling of tungsten ore continued until the end of 1948.

1949, Plans were completed for conversion from tungsten to lead-zinc mining and milling by March, 1949, at 400 tons per day, 8,000 tons per month. The ore was delivered to the mill by aerial tramline. The grinding circuit consisted of the 6-ft. x 12-ft. Hardinge rod mill and the 7-ft. x 36-in. Hardinge ball mill.
The Milling of Canadian Ores

1950, Milling capacity was increased to 500 tons per day on the addition of a bank of ten Denver No. 24 cells, completed March, 1950.

1951, Milling capacity was increased to 900 tons per day by November, 1951, with the installation of the 9-ft. x 48-in. Hardinge ball mill, and a 60-in. Akins classifier. Truck haul by road was used to supplement the aerial tramline.

1952, The decision was made to enlarge the lead-zinc mill and install the underground crushing plant and transportation system. This was completed by June, 1953 with milling at the rate of 1,700 tons per day. The truck haul and tramline were discontinued and all ore was delivered by the conveyor system.

1953-1956, No major changes or addition of equipment have been made during this period. One existing mill has been used to regrind the zinc rougher concentrate, together with Dorr cones installed for classification.

References
See bibliography.

Carnegie Mines of B.C., Limited
Sandon, B.C.

MAPS A AND B, 25

By B. W. Price, Mill Supt.

Tonnage
140 tons per day.

Ore
Galena, sphalerite, silver and pyrite in quartz and limestone.

Metallurgy, September, 1956

<table>
<thead>
<tr>
<th>Pb conc.</th>
<th>Zn conc.</th>
<th>Tailings</th>
</tr>
</thead>
<tbody>
<tr>
<td>59.1</td>
<td>8.8</td>
<td>0.25</td>
</tr>
</tbody>
</table>

Assays %

<table>
<thead>
<tr>
<th>Pb</th>
<th>Zn</th>
<th>Fe</th>
</tr>
</thead>
<tbody>
<tr>
<td>59.1</td>
<td>8.8</td>
<td>4.5</td>
</tr>
<tr>
<td>1.3</td>
<td>54.5</td>
<td>5.9</td>
</tr>
<tr>
<td>0.25</td>
<td>0.6</td>
<td></td>
</tr>
</tbody>
</table>

Crushing
The crushing plant is two-stage closed circuit producing -4-in. product. It operates 5 and 6 days per week with 2 men.

Flotation

Lead Circuit

<table>
<thead>
<tr>
<th>Reagent</th>
<th>Lb./Ton</th>
<th>Added to</th>
<th>pH</th>
</tr>
</thead>
<tbody>
<tr>
<td>NaCN</td>
<td>0.16</td>
<td>Ball mill</td>
<td></td>
</tr>
<tr>
<td>Z-3</td>
<td>0.01</td>
<td></td>
<td>7.2</td>
</tr>
<tr>
<td>Aerosol 31</td>
<td>0.12</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Zinc Circuit

<table>
<thead>
<tr>
<th>Reagent</th>
<th>Lb./Ton</th>
<th>Added to</th>
<th>pH</th>
</tr>
</thead>
<tbody>
<tr>
<td>Lime</td>
<td>1.87</td>
<td>Zn conditioner</td>
<td></td>
</tr>
<tr>
<td>CuSO4</td>
<td>1.25</td>
<td>&quot;</td>
<td>10.3</td>
</tr>
<tr>
<td>Z-3</td>
<td>0.05</td>
<td>&quot;</td>
<td></td>
</tr>
<tr>
<td>Cresylic</td>
<td>0.21</td>
<td>&quot;</td>
<td></td>
</tr>
</tbody>
</table>

Power
Carnegie hydro-electric.

Crew
Operating, 4 men; staff, 1 man.

History
The present milling plant was built in 1952 and is on the site of the 125-ton Silversmith concentrator which was erected in 1922. This plant used jigs and tables and the "oil flotation" process, using Callow pneumatic cells.

Features
Crushing on day shift. Both lead and zinc concentrates filtered direct from flotation machines. Hand sampling done throughout the mill. Final tailings to pond.

Cody-Reco Mines, Limited
Sandon, B.C.

MAPS A AND B, 25

By the Editorial Staff

Tonnage
150 to 200 tons per day.
Tungsten Milling in Canada

By H. A. Steane,
General Mill Supt.,
Canadian Exploration, Ltd.

In Canada the principal tungsten ore mineral is scheelite; wolframite-ferberite occurrences are minor. The occurrence of tungsten in Canada is widespread and more than 150 localities and properties are known, though the majority of occurrences are of mineralogical interest only. Numerous gold mines, particularly in Ontario and Quebec, have small to trace amounts of scheelite in gold-bearing veins, and often the enriched scheelite veins are low in gold. There are other occurrences of scheelite or wolframite, or both, associated with base-metal sulphides.

Several mines, perhaps twenty or more, have been able to produce small quantities of ore to yield a few pounds or tons of tungsten concentrate at a time of national emergency, but were not economic tungsten producers. The significant producers have
been Canadian Exploration Limited at Salmo, B.C.; the Red Rose mine near Hazelton, B.C.; and the Hollinger Consolidated Gold Mines Limited at Timmins, Ontario. These and minor producers will be mentioned later.

Of these mills, Canadian Exploration Limited has produced 78 per cent of the total Canadian production from the earliest record to the middle of 1956. Next in importance is the Red Rose mill, with 17 per cent of the total production, followed by the Hollinger mill. A new producer, Burnt Hill Tungsten Mines Limited in New Brunswick, commenced operation late in 1954.

Tungsten production in Canada has been spasmodic, in definite periods related to times of international conflict. The first period was 1912 to 1918; the second period, 1939 to 1943; the third period, 1947 to 1948; and the fourth period, 1951 to the present.

1912 to 1918

Scheelite was discovered on Moose river in Halifax county, Nova Scotia, in 1908. The property was brought into production by Scheelite Mines Limited, who erected a mill and made a shipment in 1912 of 14 tons of tungsten concentrate carrying 72 per cent WO₃. This was the first shipment from Nova Scotia, and the only important production of tungsten in Canada reported prior to 1918. The price at that time is unknown. The Engineering and Mining Journal, New York, records that the price for 1913 was higher than for 1912, and on January 24, 1914, ranged from $6.00 to $7.50 per unit of 20 lb. of tungsten trioxide.

In 1915, the British Government commandeered all supplies of tungsten concentrates within the Empire, at a fixed price equivalent to $12.00 per 20-lb.-unit of contained tungsten trioxide.

In 1916, a concentrating mill was erected at Burnt Hill, New Brunswick, by Acadia Tungsten Mines Company. This property was inspected in 1917 for the Munitions Resources Commission, Ottawa, and it was reported then that there was some tonnage of wolframite ore, but that the operators could not afford to produce concentrates at the official British price of 55 shillings per unit.

In 1917, a small test shipment was made from Halifax County, Nova Scotia, and another from Dublin Gulch, Mayo District, Yukon, amounting in all to 580 pounds, assaying 69.41 per cent WO₃, and netting $234.

In 1918, Canadian production amounted to 133 tons valued at $11,700, and containing 995 units WO₃. This production consisted of 11 tons of concentrate shipped to New York by the Acadia Tungsten Mines Limited, operating at Burnt Hill, New Brunswick, and a few small consignments to the Mines Branch Testing Plant, Ottawa, from Nova Scotia, Manitoba, and the Mayo district, Yukon. Scheelite was discovered near Falcon Lake, Eastern Manitoba, in March, 1918, and operations were carried on in the district during the year by the War Metals Production Company Limited. It was reported also that the Cariboo Chisholm Creek Mining Company Limited, Van Winkle, British Columbia, had been operating a deposit on Hardscrabble Creek in the Cariboo District. The price of scheelite on the New York market was around $26 per unit during January and February, 1918, but with the signing of the armistice, tungsten operations ceased; there were no quotations for November and December, 1918. Canadian production remained dormant until 1939.

1939 to 1943

Total Canadian production for this period was 74,100 units WO₃. Production was resumed in 1939, the first commercial producer being Columbia Tungsten Company Limited, renewing operation on the deposit on Hardscrabble Creek, near Wells, British Columbia, operating 1939 to 1940, with a gravity mill grinding in rolls to minus 8 mesh, and producing concentrate of over 70 per cent WO₃ and containing some 1,600 units.

Other producers followed, the largest producer being the Red Rose mine near Hazelton, operated by the Consolidated Mining and Smelting Company of Canada Limited. This mill of 75 tons per day capacity, treating ore of 1.7 per cent WO₃ operated from February, 1942 to November, 1943, and produced over 48,000 units WO₃, being 65 per cent of Canadian production for the 1939-43 period. This mill consisted of a rod-mill grinding through 10 mesh in closed circuit with a screen. A jig was installed within the grinding closed circuit to treat the minus 1/4-in. plus 10-mesh fraction in the rod-mill discharge, removing a high-grade scheelite concentrate, and returning the jig tailing to the rod-mill. The minus 10-mesh product passed over Deister tables, making a high grade concentrate, and tailing passed to a
scheelite flotation section, which made low-grade concentrate, 13 per cent WO₃, for chemical treatment. The gravity concentrate required leaching with hydrochloric acid to remove phosphorus and soluble copper, followed by further grinding and flotation to remove sulphides of copper and arsenic, leaving scheelite concentrate of marketable grade.

Next in importance was the tungsten mill at the Hollinger Consolidated Gold Mines Limited, Timmins, Ontario. This mill operated from March, 1942 to end 1943, and treated some 53,000 tons of specially selected scheelite-rich ore, averaging less than 0.5 per cent WO₃, from the Hollinger gold mine, together with over 1,000 tons of mostly low-grade custom ore from neighboring gold mines, and produced concentrate containing 16,800 units WO₃, being 22.4 per cent of Canadian production for the 1939-43 period. The mill was a gravity concentration plant, built mostly from used equipment that was lying idle, and it had a capacity of 120 tons per day. The ore was ground in a ball-mill in closed circuit with a 26 x 61-mesh screen, then passed to a flotation section for removal of sulphides prior to gravity concentration on 7 Deister sand-tables and 14 Deister slime-tables, with auxiliary classifying, sizing, and pumping equipment. The scheelite recovery ranged between 72 and 78 per cent, in concentrate assaying 75 per cent to 78 per cent WO₃, 0.15 per cent sulphur, and 0.03 per cent phosphorus.

The remainder of the production for this period came from several small producers, the chief being the following:

From Nova Scotia the production for the period amounted to 1,160 units. Kirkpatrick Tungsten Syndicate commenced production from a deposit at Goff, Halifax county, and shipped concentrate in 1940. Guysborough Mines Limited developed the Indian Path mine near Lunenburg and erected a 100-ton mill for production of scheelite concentrate in 1942. Old properties of Scheelite Mines, worked in the 1912-18 period were examined by the Geological Survey, who reported that the workings were caved in and full of water, but some ore was exposed.

From Northwest Territories, the production for the period amounted to 1,637 units. Slave Lake Gold Mines Limited, later International Tungsten Mines, on Outpost Islands in Slave Lake, commenced operation of a 50-ton mill in 1941, and produced a ferberite-wolframite concentrate from ore containing 0.6 per cent WO₃, also copper and gold, by ball-mill, jig, copper flotation, tables, with grinding to 70 per cent minus 200 mesh. Other small mills were erected at a tungsten deposit on Consolidation Lake, near Gilmour Lake, also at the Ruth property (C. M. & S.) about 10 miles east of Gilmour Lake, and at the Goodrock property on Gordon Lake.

From the Yukon, the production during this period amounted to 658 units. This production in mixed wolframite and scheelite concentrate, came chiefly from placer operations in the vicinity of Dublin Gulch and Haggart Creek, north of Mayo.

Other production amounting in all to some 4,000 units, came chiefly from a custom scheelite mill installed at the Quebec Provincial Mines School west of Val d’Or; a custom mill at the Bureau of Mines, Ottawa; a custom mill at Little Long Lac Gold Mines Limited, at Geraldton, Ontario; and Bralorne Mines Limited at Bralorne, B.C. The scheelite deposit at Salmo was discovered and a mill erected, which operated six weeks for negligible production.

Production at all mines ceased in December, 1943.

1947 to 1948

This third period had only one producer, Canadian Exploration Limited at Salmo, B.C., operating from June, 1947 until the end of 1948, with a 300-ton mill, and producing 104,000 units, mostly high-grade concentrate, but some low-grade concentrate that required chemical treatment. This production exceeded the whole previous Canadian production.

1951 to the Present

This fourth period commenced with a small production from the Yukon. This was followed by reopening of the mine of Canadian Exploration Limited at Salmo, B.C., the Red Rose mine near Hazelton, B.C., and the Hollinger mill in Ontario. Columbia Lead and Zinc Mines Limited, near Revelstoke, B.C., was a new producer, and some small production came from the Northwest Territories, and Burnt Hill Tungsten Mines Limited, New Brunswick.

The chief producer has been Canadian Exploration Limited, who built a new 250-ton mill, which commenced operation November, 1951, and has been expanded to the present 700 tons per day, and is still in operation. Production in this period
The Milling of Ores of Less Common Minerals

to mid-1956 has been 450,000 units $WO_3$. This mill has been described in detail elsewhere in this Milling Volume, and does not require further detail here.

The Red Rose mine, near Hazelton, B.C., was reopened by Western Tungsten Copper Mines Limited, in 1952 and closed in December, 1954. Production during this period was 76,482 units. The mill has been described earlier in this paper.

The mill of the Hollinger Gold Mines Limited, resumed operation in 1951 and ceased in 1953. The production during this period was 4,976 units.

The Columbia Lead and Zinc Mines Limited, operating the Regal Silver mine in the Selkirk mountain district, 20 miles east of Revelstoke, B.C., built a 50-ton mill, which was started in 1953 and operated first on tungsten ore, then changed to lead-zinc ore. The tungsten ore milled was about 2,800 tons, from which 10 tons of scheelite concentrate was produced.

There was a small production of 613 units $WO_3$ from Northwest Territories and the Yukon. The Consolidated Tungsten Mining Corporation of Canada, Limited, operating on Outpost Islands in Slave Lake, Yellowknife district, Northwest Territories, erected a mill of 50 tons daily capacity and commenced operation in September, 1951; however, operations proved unprofitable and were suspended August, 1952.

The Burnt Hill district in New Brunswick, a small producer in 1912-18 period, is active again, with operation by Burnt Hill Tungsten Mines Limited, which has erected a 150-ton mill. Operation commenced in August, 1954, treating wolframite ore, and concentrate containing 1,438 units $WO_3$ was shipped in 1955. Milling operations were suspended to install additional mill equipment, and it is expected that operation will be resumed in 1957. Sufficient ore averaging 1.25 per cent $WO_3$, has been indicated for 5-year operation at 150 tons per day.

Acknowledgement

The information for this paper has been obtained from several published articles, the Dominion Bureau of Statistics Annual Reports on the Miscellaneous Metal Mining Industry, the Annual Reports of the Minister of Mines, Province of British Columbia, the Financial Post Annual Survey of Mines, and private communications from individual mines. Much information was obtained from articles by McLaren (415), Eardley-Wilmot (61), Cummings (17), Blackshaw (81), and Wright (132).

Table of Canadian Tungsten Production

<table>
<thead>
<tr>
<th>Period</th>
<th>Unit $WO_3$</th>
</tr>
</thead>
<tbody>
<tr>
<td>1912 to 1918</td>
<td></td>
</tr>
<tr>
<td>Scheelite Mines Limited, Nova Scotia</td>
<td>980</td>
</tr>
<tr>
<td>Acadia Tungsten Mines Limited, Burnt Hill, N.B.</td>
<td>1,016</td>
</tr>
<tr>
<td>Miscellaneous</td>
<td></td>
</tr>
<tr>
<td>Total</td>
<td>1,996</td>
</tr>
<tr>
<td>1939 to 1943</td>
<td></td>
</tr>
<tr>
<td>Red Rose Mine, C.M. &amp; S., near Hazelton, B.C.</td>
<td>48,000</td>
</tr>
<tr>
<td>Hollinger Consolidated Gold Mines Limited, Timmins</td>
<td>16,800</td>
</tr>
<tr>
<td>Columbia Tungsten Company Limited, Hardscrape Creek</td>
<td>1,600</td>
</tr>
<tr>
<td>Northwest Territories</td>
<td>1,637</td>
</tr>
<tr>
<td>Nova Scotia</td>
<td>1,160</td>
</tr>
<tr>
<td>Yukon</td>
<td>658</td>
</tr>
<tr>
<td>Miscellaneous</td>
<td>4,245</td>
</tr>
<tr>
<td>Total</td>
<td>74,100</td>
</tr>
<tr>
<td>1947 to 1948</td>
<td></td>
</tr>
<tr>
<td>Canadian Exploration Limited, Salmo</td>
<td>104,000</td>
</tr>
<tr>
<td>1951 to 1956 (Production continuing)</td>
<td></td>
</tr>
<tr>
<td>Canadian Exploration Limited, Salmo</td>
<td>450,000</td>
</tr>
<tr>
<td>Red Rose Mine, Western Tungsten Copper Mines Ltd.</td>
<td>76,482</td>
</tr>
<tr>
<td>Hollinger Gold Mines Limited</td>
<td>4,976</td>
</tr>
<tr>
<td>Columbia Lead and Zinc Mines Limited (estimated)</td>
<td>700</td>
</tr>
<tr>
<td>Northwest Territories</td>
<td>613</td>
</tr>
<tr>
<td>Burnt Hill Tungsten Mines Limited</td>
<td>1,438</td>
</tr>
<tr>
<td>Total</td>
<td>534,209</td>
</tr>
<tr>
<td>Total Canadian Production to end 1956</td>
<td>714,305</td>
</tr>
</tbody>
</table>

It has been difficult to obtain an accurate record of some outputs, especially for the 1939-43 period. The figures given above and in this paper, even if not strictly correct, will show the relative importance of the producers.

References

5. The Beneficiation of Canadian Tungsten Ores, by D. C.
Tungsten Milling at
Canadian Exploration Limited
Salmo, B.C.

By H. A. Strange, General Mill Supt.

HISTORY

The history of the property now worked by Canadian Exploration Limited has been outlined in detail in a paper by Little from which the following summary has been made for the tungsten operation.

The property is located on Iron Mountain, which owes its name to the presence of large iron-stained outcrops, and was prospected for gold from 1895 through the turn of the century without discovery of commercial gold ore. The first recorded owner of claims that form the old Emerald lead-zinc mine was John Waldhouser, who commenced production of lead ore in 1908, and was the largest lead producer in the Nelson mining district in 1909. Iron Mountain Limited, a subsidiary of Pacific Coast Steel, of San Francisco, was organized in 1912 to operate the property, and a small but steady output was maintained until 1928, when operations ceased because of low market prices.

Exploration by the same company was resumed in 1936 under the direction of Harold Lakes, and interest being confined to lead and zinc. Scheelite was first discovered on the property in 1942 when exploration samples submitted to the B.C. Department of Mines in Victoria for examination for "strategic" minerals were reported to contain scheelite. Early samples indicated the WO₃ content to be of the order of 0.2 per cent. Exploration for scheelite continued, with higher grade ore being discovered.

The property was taken over from Iron Mountain Limited by the Wartime Metals Corporation in September, 1942. In fourteen months the property was developed and a 500-ton mill built, which operated six weeks, to close in November, 1945, when the war demand for tungsten ceased. The mill was on the site of the present lead-zinc mill.

Canadian Exploration Limited took over the property in 1947 and active mining and milling of tungsten ore was resumed in June of that year and continued until the end of 1948 when the tungsten operation ceased. The mill was converted to treat lead-zinc ore and has continued as a lead-zinc mill.
The milling of Canadian Ores

The present tungsten mill had its origin in 1951, when because of renewed demand for tungsten for the Korean conflict, an agreement was made between the Federal Government and Canadian Exploration Limited, for the sale of two blocks of ground to the government, and for this company to erect and equip a 250-ton mill and operate the mine and mill for the government. This mill commenced operation November, 1951. Further exploration by the company resulted in the discovery of another tungsten deposit outside the government-owned blocks. To treat ore from this deposit, the company added to the mill, a duplicate 250-ton section, which was completed in April, 1952. In September, 1952, the company purchased the government's interest in the mine and the mill, and operated the whole as one unit.

Except for the completion in 1953 of the central underground crushing and conveying plant to serve both mills, there have been no further additions to the tungsten mill, but by small alterations, and better understanding of the metallurgy, the milling rate has been increased steadily to the present 700 tons per day.

LOCATION

The Salmo property of Canadian Exploration Limited, is comprised of forty-eight Crown-granted mineral claims situated on Iron mountain between Sheep Creek on the north, and Lost Creek on the south, in the Nelson mining division. A good road branches off the main Nelson-Nelway highway at Sheep Creek valley, 4 miles south of Salmo, then climbs four miles to the main mine camp at elevation 4,000 ft. Present operations on these claims are three tungsten mines, the Dodge, Feeny, and Emerald, and one lead-zinc mine, the Jersey.

The tungsten mill, situated near the 3,800 level portal of the Emerald tungsten mine, treats the output from the three tungsten mines. The lead-zinc ore from the Jersey mine is milled in a separate mill situated along the main Nelson-Nelway highway, 7,000 ft. distant in a straight line from, and 1,800 ft. below the 3,800 Emerald mine portal.

CRUSHING

The crushing and conveying system has been described in detail in a paper by McCutcheon, and Walkey, from which the following summary for the tungsten operation has been taken.

The underground crushing plant is situated on the 3,800 level of the Emerald tungsten mine, 800 ft. from the portal. Crushing is two-stage by a jaw crusher followed by a screen and Allis-Chalmers Hydrocone in closed circuit with the screen.

There are two coarse-ore storage pockets, one of 5,000 tons capacity for the lead-zinc ore, the other of 3,000 tons capacity for the tungsten ore. The discharge from each pocket may be cut off completely by an up-cutting guillotine-type chute gate, thus allowing each ore to be crushed separately. Both pockets are arranged to discharge to a common Stephens Adamson manganese-steel pan-feeder, the rate of ore-flow being controlled by a set of five anchor chains, each weighing 100 pounds per foot, and operated by a 12-in. diameter air cylinder. The feeder delivers ore to the Allis-Chalmers, type A-1, 36-in. all-steel sectionalized jaw crusher, driven by a 150 h.p., 900 r.p.m. motor. The close setting is four inches.

The product from the jaw crusher is conveyed to a Tyler 6-ft. x 12-ft, two-deck type F-900 heavy-duty Ty-Rock screen, the bottom deck having 14-in. square openings, and the top deck having 2-in. square openings. The screen undersize is discharged to No. 1 conveyor, and carried by the conveyor system to the respective mill.

The oversize screen-product is conveyed to an Allis-Chalmers size 7-60 Hydrocone crusher equipped with a speed-set control, reset accumulator system. The cone has a close side setting of one-half inch and has a one-inch throw. It is driven by a 200 h.p., 900 r.p.m. motor. An even distribution of feed is maintained by hanging 3-in. chains three inches apart around the perimeter of the conveyor discharge chute. These chains extend six inches below the wobble-plate feeder of the cone. The Hydrocone product is conveyed with the jaw crusher product to the screen, and so operates in closed circuit with the screen.

A 15-ton single-hook electric crane is installed in the main chamber. It has a 26-ft. span, 48-ft. lift, and is controlled from a pendant that will reach to all floor levels.

Fresh air is drawn into the crushing chamber at the rate of 9,000 c.f.m., which is sufficient for three air changes per hour. Dust at the jaw and cone crushers, screen, and all transfer points, is collected and discharged to the portal by a 6,000 c.f.m. fan. This leaves 3,000 c.f.m. air to move along the 3,800 level, so preventing mine air from entering the chamber.
The ore to the tungsten mill is carried 2,920 ft.
by two conveyors from the crushing plant to the
lead-zinc and tungsten transfer point, then 450 ft. by
a third conveyor from the transfer point to the

tungsten mill.

A system of interlocking controls is installed
to facilitate operation, so that failure of one unit will
stop automatically all preceding conveyors or feeders.
To ensure that the ore being crushed is directed to
the appropriate mill, an interlock control is arranged
to control the operation of the coarse-ore chutes
according to the position of the lead-zinc and tungsten transfer chute.

The underground crushing plant operates one
shift per day on tungsten ore, at the rate of 150
tons per hour.

At the tungsten mill, the original crushing
plant of 40 tons per hour capacity still remains in
place, and is used to crush the small amount of ore
from the lower levels, about ten per cent of the
whole ore supply. On completion of the inclined
shaft being sunk, this ore will be delivered to the
underground crushing plant.

**TUNGSTEN MILLING**

The tungsten mill and the development of the
ore dressing practice have been described in detail in
two papers by McLeod and Kipp from which the
following has been taken.

The tungsten ore occurs in limestone and
skarn country rock. Some ore is heavily mineralized
with pyrrhotite, and contains also some pyrite,
quartz, and biotite; other ore is a quartz-rich type
with quartz replacement of limestone and small
amounts of pyrrhotite, pyrite, and sericite; a third
type is high in garnet.

The ore mineral is scheelite with trace
amounts of powellite. The principal gangue minerals
are calcite, pyrrhotite, pyrite, apatite, quartz, mica,
siderite, garnet, hornblende, pyroxene, and traces of
fluorite and molybdenite. The average mill feed
assay for the nine months to 31 May, 1956, has been
0.7 per cent WO₃ and 5 per cent sulphur.

Most of the scheelite is liberated in a grind
through 65 mesh, though some locked scheelite may
be found in particles down to 200 mesh. In a
coarsely ground product, some free scheelite occurs
in particles as coarse as 20 mesh. Scheelite is seen
to be more finely disseminated in the pyrrhotite than
in its other associations. The grain boundaries are
strongly knit, with the result that many scheelite
grains carry minute to fractional scabs and smears
of sulphide and carbonate minerals. In a similar
manner, particles of gangue minerals, both sulphide
and non-sulphide, carry scabs and smears of
scheelite.

The process is designed and operated to pro-
duce scheelite concentrate of marketable grade, and
following are the specification limits that concern us.

<table>
<thead>
<tr>
<th></th>
<th>Minimum</th>
<th>Maximum</th>
</tr>
</thead>
<tbody>
<tr>
<td>Tungsten trioxide (WO₃)</td>
<td>60.00%</td>
<td></td>
</tr>
<tr>
<td>Sulphur maximum</td>
<td></td>
<td>0.50%</td>
</tr>
<tr>
<td>Phosphorus maximum</td>
<td></td>
<td>0.05%</td>
</tr>
</tbody>
</table>

The high sulphide content of the ore is the
major factor in choosing the milling scheme, there
being ten to fifteen times as much sulphide mineral
as there is scheelite. Consideration was given to
removal of scheelite at coarser sizes than at present,
by jigs or coarse tables, but sulphide removal would
be a major problem, and such a scheme is not war-
ranted in this case. The best way to remove this
large amount of sulphide is by flotation, and this
necessitates grinding the whole of the ore to flotation
size before any concentration is attempted. With
the efficient scheelite flotation to recover fine
scheelite, it is not necessary to recover the scheelite
at the coarsest possible size of liberation.

The general treatment scheme in the mill is
as follows:

1. Grinding to 40 per cent -200 mesh.
2. Flotation of sulphide minerals for removal and discard.
3. Separation of remaining feed into sand and slime
fractions.
4. Hydraulic sizing of the sand fraction into five sizes.
5. Tabling of sized fractions to recover scheelite table
concentrate of marketable grade.
6. Thickening and flotation of the combined table tailing
and slime fraction, to recover scheelite flotation con-
centrates.
7. Leaching of scheelite flotation concentrate with hydro-
chloric acid to dissolve calcite and apatite, leaving
leached scheelite concentrate of marketable grade.
8. Drying and blending of table concentrate and leached
concentrate by batches to give a reasonably uniform
product to meet specifications.

**Grinding**

Two fine-ore bins receive the ore from the
underground crusher and the mill crusher. They
are made from 2-in. x 12-in. B.C. fir nailed on the
flat. The bins are 32 ft. high by 24 ft. diameter,
octagonal in shape, with flat bottoms. Each bin has
a theoretical capacity of 1,000 tons; the live capacity
is approximately 600 tons. Each bin is discharged
### Table I

**Size Analyses of Grinding Circuit September, 1956**

<table>
<thead>
<tr>
<th>Screen Fraction</th>
<th>Mill Feed % On</th>
<th>R.M. Disch. % On</th>
<th>Prim. CL. Sand. % On</th>
<th>B.M. Disch. % On</th>
<th>Sec. CL. Sand. % On</th>
<th>Prim. CL. O'Flow % On</th>
<th>Sec. CL. O'Flow % On</th>
<th>Combined Class O'Fows % On</th>
</tr>
</thead>
<tbody>
<tr>
<td>+.742&quot;</td>
<td>16.1</td>
<td>16.1</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
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</tr>
<tr>
<td>-.742&quot; + .525&quot;</td>
<td>24.9</td>
<td>41.0</td>
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<td></td>
<td></td>
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<tr>
<td>-.525&quot; - .375&quot;</td>
<td>10.6</td>
<td>31.6</td>
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</tr>
<tr>
<td>-.375&quot; - .253&quot;</td>
<td>8.8</td>
<td>60.4</td>
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<td>86.3</td>
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<td>88.8</td>
<td>9.4</td>
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<td>100.0</td>
<td>76.4</td>
<td>100.0</td>
<td>65.0</td>
<td>100.0</td>
<td>34.7</td>
<td>47.3</td>
</tr>
</tbody>
</table>

Calculated proportion of Mill Feed:

- 30% Sulphide
- 70% Tail

Weighted average of two classifier o'flows, 30:70 ratio (% solids): 41.5

### Table II

**Size Analyses of De-Sliming Circuit September, 1956**

<table>
<thead>
<tr>
<th>Mesh</th>
<th>Sulphide Flot'n. Tail % On</th>
<th>Table Tail % On</th>
<th>Slime Feed to Thickener % On</th>
<th>Secondary Dorclocie u'flow % On</th>
<th>Primary Dorclocie u'flow % On</th>
</tr>
</thead>
<tbody>
<tr>
<td>+.35</td>
<td>2.5</td>
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<td>3.1</td>
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<td>7.3</td>
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<tr>
<td>-.48</td>
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<td>14.7</td>
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<td>-.65</td>
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<td>-.100</td>
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<td>60.8</td>
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<td>16.1</td>
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<td>-.150</td>
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<td>71.6</td>
<td>10.0</td>
<td>11.1</td>
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<tr>
<td>-.200</td>
<td>49.9</td>
<td>100.0</td>
<td>100.0</td>
<td>35.4</td>
<td>29.8</td>
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<tr>
<td>Total</td>
<td>100.0</td>
<td>100.0</td>
<td>100.0</td>
<td>100.0</td>
<td>100.0</td>
</tr>
</tbody>
</table>

% Solids:
- 24.4 Sulphide
- 29.3 Tail

Mill feed rate - 660 tons per day. Feed containing some oxidized surface ore.
The Milling of Canadian Ores

origin to natural processes in the ore body rather than to overgrinding of the mill feed.

Sulphide Flotation

Sulphur occurs in the ore mainly as pyrrhotite; there is little pyrite. The head assays 4 per cent to 7 per cent sulphur. Most of the sulphide mineral fed to tables will report in the table concentrate, and therefore a method of removing the sulphide mineral must be provided to keep the sulphur content of the final table concentrate below 0.5 per cent. This outlet is provided by the sulphide flotation section treating the pulp from the grinding section, for removal of as much as possible of the sulphide mineral ahead of tabling. The table section can handle a small amount of sulphide mineral and still make a marketable scheelite concentrate, but a heavy load of sulphide on the tables could not be tolerated. The fine sulphide that does not float readily is depressed in the scheelite flotation.

The sulphide section consists of a bank of No. 24 Denver Sub-A flotation machines, fourteen rougher cells and four cleaner cells. These cells are fitted with conical-disc impellers and wear plates. The original drive was a 10 h.p. motor per pair of cells, impeller speed 275 r.p.m., peripheral speed 1,400 ft. per min. This drive has been changed to a 15 h.p. motor per pair of cells, impeller speed 320 r.p.m., peripheral speed 1,800 ft. per min. A great improvement was obtained when the speed was increased; the aeration was doubled, which increased the froth-making ability of the machine, and resulted in increased metallurgical efficiency and a saving of frothing and collecting agents.

The flotation is carried out at 40 per cent solids and at pH 8.4, the natural pH of the ore. A bulk rougher float is made and cleaned once. The cleaned sulphide concentrate is discharged to waste, the tailing is pumped to the desliming circuit. The reagents used per ton of ore are: pentasol xanthate, 0.20 lb.; AeroFloat 25, 0.133 lb.; copper sulphate, 0.03 lb.; methyl amyl alcohol, 0.02 lb.

Sulphuric acid up to 6 lb. per ton was used for several years in the sulphide flotation, resulting in improved floatability of pyrrhotite, but was suspected of lowering scheelite recovery in the subsequent scheelite flotation. The amount of calcium sulphate formed, increased the consumption of soda ash in the scheelite flotation, and at times made it impossible to obtain the desired pH of 10.2. Its use was discontinued early in 1955. The corrosive effect of the acid has required much replacement and repair to tanks and mechanisms in the sulphide flotation cells.

Soda ash has been tried in the sulphide flotation, with possible improvement in pyrrhotite flotation, but it interfered seriously with subsequent settling and thickening of the pulp for scheelite flotation and therefore its use could not be continued.

Desliming, Sizing, and Tabling

The purpose of tabling is to remove the granular scheelite in a concentrate that will meet specifications without further treatment. Efficient tabling requires a deslimed feed, hydraulically sized. Factors influencing the amount of table concentrate produced are: fineness of scheelite in the ore, fineness of grinding, and efficiency of the desliming circuit. Table recovery varies between 25 per cent and 45 per cent of the WO3 in the head.

The desliming and sizing circuit consists of three 12-in. x 20-deg. Dorrcrones, and a Dorrcro type E.X. five-compartment sizer, together with the necessary pumps.

The tailing from the sulphide flotation is pumped to two primary 12-in. Dorrcrones in parallel, the overflows by-pass the tables and flow directly to the flotation-feed thickeners, the underflows are cleaned further in one secondary 12-in. Dorrcrone, and finally in the sizer. The overflows from the secondary Dorrclone and the sizer can be closed circuited with the primary Dorrcrones, or can flow direct to the thickeners.

The Dorrcro Sizer is a Fahrenwald-type hydraulic classifier in which the sizing is effected by controlling the density in each pocket with a rising current of water, and regulating the rate of discharge of pulp. The rising water current and density are graded highest in the first pocket, producing the coarsest product, and progressively lower in each succeeding pocket to the finest discharge in the fifth pocket. The excess water overflows the top of the sizer, taking the last of the slime with it. The discharge from each pocket goes to a revolving distributor and is split to feed three or four tables.

The tables are No. 6 Deister Super Duty Diagonal Deck, with linoleum covers and hardwood riffles. They are pitched both longitudinally and transversely according to the size of the feed being tabled. On the coarse tables, the longitudinal pitch
is \( \frac{4}{3} \)-in. per ft., and the transverse \( \frac{5}{2} \)-in. per ft. The fine tables are pitched \( \frac{1}{4} \)-in. and \( \frac{3}{4} \)-in. per ft. The speed of the tables is 314 strokes per minute, and the length of the stroke varies from \( \frac{4}{3} \) to 1-in.

There are eighteen tables used for roughing, four for middling, and four for cleaning. The feed to the middling tables, and the feed to the cleaner tables, are sized hydraulically by smaller sizes. This roughing, middling, cleaning circuit is similar to the usual closed-circuit flotation. The rougher tables make rougher concentrate and middling, leaving tailing, which goes to scheelite flotation. The middling cut is dewatered in a 12-in. Dorrcleone and flows to middling tables, whose concentrate joins the rougher concentrate, and whose tailing is dewatered in another 12-in. Dorrcleone and returns to the grinding circuit where it joins the new feed and is re-ground. The rougher concentrate is pumped to a secondary sulphide-removal circuit, then to cleaner tables making a finished concentrate of 76 per cent \( \text{WO}_3 \), 0.2 per cent sulphur, and 0.02 per cent phosphorus, leaving a tailing, which is pumped to the middling tables.

In the early operation, much effort was made at high reagent cost, to try to float and remove all the sulphide that otherwise could report in the table rougher concentrate, but this never was successful. The present scheme accepts the fact that some sulphide will pass through to the table feed and will report in the rougher table concentrate. A secondary removal scheme, consisting of a Dings-Crockett wet magnetic separator and four small Denver flotation cells, has been installed to remove this sulphide from the rougher table concentrate prior to retabling on the cleaner tables. The sulphide reject removed by these units is returned to the main circuit, and eventually reaches either the main sulphide flotation concentrate or the final tailing.

**Thickening and Scheelite Flotation**

Two forty-foot thickeners receive the table tailing, the slime fraction of the ore. Dorrcleone overflows from the middling circuit, and miscellaneous wash water. The purpose is to thicken the pulp for scheelite flotation to follow. Part of the thickener overflow water is used in the desliming circuit, tabbling, and other places where possible. A settling agent, either Separan 2610 or Aerofloc 3000, 0.005 lb. per ton is used to help thickening. The use of these reagents does not affect the scheelite flotation.

The scheelite flotation section consists of four 5-ft. x 6-ft. Denver conditioners, and twenty No. 24 Denver Sub-A flotation cells, divided into sixteen rougher cells and four cleaner cells, arranged for three stages of cleaning. The cells are equipped with 15 h.p. dual drives instead of the original 10 h.p. drives, and the speed increased from 275 to 320 r.p.m. This speed increase has been worth while in improving the aeration and frothmaking ability of the machines.

The thickened pulp, about 45 per cent solids, is conditioned for sixteen minutes with conditioning agents, at \( \text{pH} \) 10.1 to 10.2, and a temperature of 16 to 20°C. The soda ash addition, all to No. 1 conditioner, is regulated to control the conversion of oleic acid to soap. The rate of addition is controlled by an automatic \( \text{pH} \) recording and controlling system, with the electrode assembly in No. 2 conditioner. The temperature is important. It is difficult to maintain froth on the machines at temperatures below 16°C., and excessive frothing occurs at temperatures above 20°C. A 100 h.p. boiler supplies steam to No. 1 conditioner for heating the pulp.

The dispersant reagents used in lb. per ton, and their function, are as follows: soda ash, 2.5, alkalinity control; sodium silicate, 0.25, depressant for quartz and silicates; quebracho, 0.80, depressant for calcite; sodium cyanide, 0.15, depressant for sulphides. Oleic acid is the collector and is stage fed to the cells at about 0.40 lb. per ton.

The scheelite flotation section receives feed of average 0.5 per cent \( \text{WO}_3 \), and produces concentrate of 30 per cent \( \text{WO}_3 \), 0.24 per cent sulphur, 3.5 per cent phosphorus, with varying amounts of calcite, silicates, and occasionally fluorite. The concentrate is filtered on an American-type disc filter to about 14 per cent moisture. The average flotation tailing is 0.10 per cent \( \text{WO}_3 \). A size analysis of the flotation tailing is given in Table IV.

**Leaching**

The flotation concentrate is brought up to shipment grade and phosphorus specification by leaching with hydrochloric acid to dissolve calcite and apatite. Silicates or fluorite if present, are unaffected and dilute the final concentrate.

The acid storage consists of three rubber-lined steel tanks, 9 ft. diam by 23 ft. 6 in., with total storage capacity of 171 tons of acid. One tank is located at the railroad at Salmo, and the other two at the mill, with a gravity flow in rubber hose line to
The Milling of Canadian Ores

Table III

Size Analyses of Dorrco Sizer Products September, 1956

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<tr>
<th>MESH</th>
<th>NO. 1 SPIGOT % On</th>
<th>NO. 1 SPIGOT Cum</th>
<th>NO. 2 SPIGOT % On</th>
<th>NO. 2 SPIGOT Cum</th>
<th>NO. 3 SPIGOT % On</th>
<th>NO. 3 SPIGOT Cum</th>
<th>NO. 4 SPIGOT % On</th>
<th>NO. 4 SPIGOT Cum</th>
<th>NO. 5 SPIGOT % On</th>
<th>NO. 5 SPIGOT Cum</th>
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<td>73.7</td>
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<td>64.2</td>
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<td>32.6</td>
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<td>33.8</td>
<td>100.0</td>
<td>67.4</td>
<td>100.0</td>
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</table>

TOTAL: 100.0

% SOLIDS: 45.0

% DISTRIBUTION OF SPIGOT PRODUCTS: 19.7

MILL FEED RATE — 660 Tons per day. Feed containing some oxidized surface ore.

Table IV

Elutriation of Tungsten Flotation Tailings
Week Composite 16 to 22 July, 1956

<table>
<thead>
<tr>
<th>Prog.</th>
<th>Weight %</th>
<th>Assay % WO₃</th>
<th>% Distribution of WO₃</th>
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<tr>
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<td>18.3</td>
</tr>
<tr>
<td>- 48</td>
<td>11.7</td>
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</tr>
<tr>
<td>- 65</td>
<td>13.5</td>
<td>40.9</td>
<td>16.7</td>
</tr>
<tr>
<td>- 100</td>
<td>13.2</td>
<td>50.6</td>
<td>10.0</td>
</tr>
<tr>
<td>- 150</td>
<td>10.8</td>
<td>100.0</td>
<td>37.9</td>
</tr>
</tbody>
</table>

TOTAL: 100.0

NOTE: 325 mesh aperture 42 microns. Elutriation sizes are for quartz.

Table V

Metallurgical Balance
1 September, 1955 — 31 May, 1956

<table>
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<tr>
<th></th>
<th>Weight Tons</th>
<th>% WO₃</th>
<th>WO₃ Units</th>
<th>% Dist. WO₃</th>
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<td>Mill Feed</td>
<td>150,125.00</td>
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<td>106,137.54</td>
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<td>Iron Concentrate</td>
<td>12,294.10</td>
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<td>1,475.30</td>
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<tr>
<td>Table Concentrate</td>
<td>470.77</td>
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<td>35319.53</td>
<td>33.28</td>
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<tr>
<td>WO₃ Flotation Feed</td>
<td>137,360.13</td>
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<td>69,342.66</td>
<td>65.33</td>
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<tr>
<td>Leached Concentrate</td>
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<td>WO₃ Flotation Tailing</td>
<td>135,361.78</td>
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<td>15,307.23</td>
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<tr>
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<td>148,831.46</td>
<td>0.12</td>
<td>18,059.21</td>
<td>17.01</td>
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</tbody>
</table>

Total Marketable Recovery: 82.99%.
of a steam coil covered by a 35-mesh screen, forming a tray on top of steel bins. One hundred infra-red lamps are suspended in a rack over each drier. The wet table concentrate is sampled and spread on the drier, and falls through the screen when dry. Occasional stirring will hasten the drying. It remains in the bin until required for blending.

The drier for the leached concentrate was manufactured by the Conveyor Company of Los Angeles. It is a paddle-type double-screw conveyor 26-ft. long, screw diameter 9-in., equipped with sixteen 4-kilowatt Syntron radiant heating panels, suspended over the screws. The drier is equipped with a feed hopper, into which the leached concentrate from the filter press is shovelled. The dried concentrate is discharged continuously to barrels on the floor below, where it is sampled and stored in batches until assays for impurities and grade are available.

A small flat-hearth roaster, complete with bag-type dust collecting system, is installed and may be used when necessary for an additional drier, or as a roaster to reduce the sulphur content of the concentrate.

The blending of the table and the leached concentrates is done in a 24-in. horizontal screw mixer, 18-ft. long, manufactured by the United Steel Corporation, and is driven by a 25 h.p. motor.

The blending is done to meet the buyer's specification. The concentrate is shipped in steel cans, content 150 lb. net weight. The canned concentrate is handled and stored on pallets by a Yale fork lift truck.

Assaying

The use of instruments for assaying at this property has been described in detail in a paper by Wilson, from which the following summary has been taken.

Much research and study has gone into the methods for assaying the various mill samples, and now there is close agreement between our results and those of other laboratories that have been working with us. The methods for analysis of concentrate samples were well established and agreement was close between experienced technicians here and elsewhere, but for low-grade products, agreement was far from close in the early operation.

Assays for WO₃ on low-grade daily control samples are done colourimetrically using a Beckman DU Spectrophotometer for reading the light absorption. The results with this method agree with the gravimetric method performed by a sufficiently skilled technician. Assays for WO₃ on concentrate daily control samples are done by a short gravimetric method, giving results within eight hours. The long gravimetric method is used for shipment samples and monthly composite samples.

One determination that has required much attention is the WO₃ assay on the sulphide concentrate. This sample is high in iron and sulphur, with 0.1 per cent WO₃. Until the trouble was understood and corrected, the results by the gravimetric method were invariably only a fraction of those given by the colourimetric method. The solution to the problem was the use of alpha benzoin oxime in addition to cinchonine for complete precipitation of WO₃ in the presence of high iron. With cinchonine
alone, it was found that precipitation was not complete, but that a constant weight of \( \text{WO}_3 \) remained in solution. For high-grade samples this was not important, but the lower the grade, the greater the proportion of \( \text{WO}_3 \) not precipitated.

Colourimetric methods are used also for phosphorus and molybdenum determinations. A Leco high-frequency induction-furnace is used for sulphur analysis by the combustion method. This has reduced the work considerably and gives reliable results.

Acknowledgement

The subject of this paper has been so well covered in sections by earlier authors of several papers, that the writer could not do better than attempt to make a composite paper combining the important points from each, much of it taken verbatim. Acknowledgement to these authors has been given at the appropriate part in the text. Acknowledgement is made also to Canadian Exploration Limited for permission to write and publish this paper.

References

See Bibliography.
Re-Grind Practice at Canadian Exploration Limited

By H. A. STEANE *

(Annual General Meeting, Vancouver, April, 1938)

(Transactions, Volume LXI, 1938, pp. 125-128)

INTRODUCTION

THIS PAPER describes the re-grind section in the 2,000-ton per day, lead-zinc flotation mill of Canadian Exploration Limited, near Salmo, B.C.

There are two re-grind units in operation, one in the lead circuit, and one in the zinc circuit. In each unit, the rougher concentrate is re-ground ahead of the cleaning circuit. These re-grind units have made valuable improvements in the concentrate grade, and with no loss in recovery by over-grinding galena or sphalerite. Since the initial installation of this re-grind section, the plant throughput has been increased gradually from 1,600 tons to 2,000 tons per twenty-four hours, and, as shown by the screen analysis of mill tailing, a coarser overall grind has been accepted, from 55 per cent to 45 per cent minus 200 mesh. There has been an increase in grinding capacity, but two mills have been taken from the main grinding circuit and used for the re-grind circuit.

ONE CHARACTERISTICS

The ore contains galena, sphalerite, and pyrite, in calcite-dolomite gangue. The galena is particularly fine-grained; the sphalerite is considerably coarser. Intergrowth of one sulphide mineral with another is rare, but grinding through 325 mesh is required to give adequate liberation of the sulphide minerals from the calcite-dolomite gangue.

PLANT PRIOR TO RE-GRIND INSTALLATION

The mill flow-sheet was conventional, with a lead flotation section followed by a zinc flotation section. In each section, a rougher concentrate and a scavenger concentrate was made; the rougher concentrate went to three-stage cleaning; the cleaner tailing and the scavenger concentrate were returned to the head of the rougher cells.

At 1,600 tons per twenty-four hours, the grinding for flotation feed was 55 per cent minus 200 mesh. At this grinding there were middling grains of calcite containing one or more particles of galena, sphalerite, and pyrite. These middling grains could be floated readily, and, depending on conditions, they would report either in the final concentrate, thus lowering the grade, or in the tailing, giving lower recovery.

There was the possibility of finer grinding on the whole of the ore to liberate the locked minerals. This would have required much greater grinding capacity, thus increasing both capital and operating costs. The fact that the middling grains could be floated readily has allowed similar results to be accomplished at lower cost. In fact, since installing the re-grind section, the trend has been to coarser initial grinding of flotation feed.

RE-GRIND FLOWSHEET No. 1

![Diagram of re-grind flowsheet]

D1, D2, D3, DorrClones size 12'; M1, Ball mill Harding, 7' x 36'.

Re-Grind Section

The middling re-grind section was not conceived and put into operation in its present form but grew step by step, by trial and observation, from June 1953 to January 1957.

A previous paper (1), written in December, 1955, gave information on the operation of DorrClones in the middling re-grind circuit. The paper dealt with DorrClone operation rather than flotation metallurgy.

For clarity in reporting the progress, the changes may be grouped under three definite stages, shown in the accompanying three flowsheets: Re-Grind Flow-sheet No. 1, grinding middling products only, from both lead and zinc sections; Re-Grind Flow-sheet No. 2, grinding zinc rougher concentrate before cleaning, in addition to the lead and zinc middling products as before; and Re-Grind Flow-sheet No. 3.

grinding lead rougher concentrate and zinc rougher concentrate before cleaning, and the zinc scavenger concentrate.

Re-Grind Flow-Sheet No. 1

The re-grind was started in June, 1953, with one DorrClone and one ball-mill treating the zinc scavenger concentrate and the zinc cleaner tailing. The DorrClone overflow was returned to the head of the zinc roughers; its underflow passed through the ball-mill in open circuit, then to the head of the lead roughers.

It may seem strange to return this zinc-section product to the lead section, but some galena was liberated in the re-grind, and, by using cyanide in the re-grind mill, the zinc did not float in the lead section.

Work continued on this re-grind circuit throughout 1954, with changes in reagent addition, and only minor changes, but still working on scavenger concentrate and cleaner tailing.

A second DorrClone was added, to be in closed circuit with the re-grind mill for finer grinding. A third DorrClone was added to treat the lead scavenger concentrate and the lead cleaner tailing; the sand product from this DorrClone joined the zinc middling products in the re-grind mill. This completed the circuit shown in Flow-sheet No. 1.

This re-grind unit gave a definite improvement in the grade of the zinc concentrate, raising it from 50 per cent to 55 per cent, but still there were middling grains in the final zinc concentrate. These middling grains being sphalerite in calcite and dolomite, increased the CaO and MgO content, which the smelter found undesirable.

The presence of middling grains in the final zinc concentrate was proved by microscopic observation, which showed composite grains in the plus 325 mesh fraction. These grains floated readily in the zinc rougher cells, and again in all three cleaning stages, and therefore never had a chance to be re-ground. The next step was to change the re-grind circuit so that the zinc rougher concentrate would be re-ground before going to the cleaner cells.

Re-Grind Flow-Sheet No. 2

In December, 1954, the re-grind section was changed so that the re-grind mill, in closed circuit with No. 2 DorrClone, treated the zinc rougher concentrate alone. The middling products were classified in No. 1 and No. 3 DorrClones as before, but the underflows were sent to the ball-mill in the main grinding circuit. The re-grind circuit is shown in Flow-sheet No. 2.

The cyanide addition to the re-grind mill was discontinued now that the re-ground rougher concentrate would remain in the zinc section and go direct to the cleaner cells. Note this deletion of cyanide. There were other effects, not expected, of which more will be told.

The first results were disappointing. After grinding the zinc rougher concentrate, the zinc flotation was sluggish in the cleaner cells, the grade of final zinc concentrate was not improved, and the loss in the tailing increased.

Several reagent additions were tried. These were copper sulphate, sodium aeroxol, and xanthate, singly or in combination, added to the cleaners; and copper sulphate and lime, singly and together added to the re-grind mill. These additions appeared to help a little, but not much. They have been discontinued.

Success was achieved on adding sodium cyanide to the re-grind mill, at the rate of 0.05 lb. per ton of original ore. The final zinc concentrate was raised above 58 per cent zinc.

This re-grind section now was giving good results on zinc flotation, but there was little if any improvement on lead flotation. Middling grains consisting of galena, sphalerite, and pyrite, together in calcite, occurred in the final lead concentrate, and in the intermediate products of rougher concentrate, scavenger concentrate, and cleaner tailing.

To liberate this locked galena, sphalerite, and pyrite, in calcite, it seemed logical to send the lead rougher concentrate to a re-grind unit similar to the zinc unit. We hesitated to do this because the galena in the ore is particularly fine grained, and additional grinding might have caused a loss of the already fine, free galena.

A screen analysis of the final lead concentrate at that time is given in Table II. From this Table it appeared that, by grinding through 325 mesh, the grade of the lead concentrate could be raised from 68 per cent lead to 75 per cent, with a substantial reduction in zinc and iron.

Re-Grind Flow-Sheet No. 3

In January, 1957, it was decided to make a trial of grinding the lead rougher concentrate. A ball-mill for this purpose was taken from the main grinding circuit, and some pump lines were re-routed to complete the circuit shown in Flow-Sheet No. 3, which is still in use.

This re-grind unit has been even more successful than the screening in Table II indicated, with the result that the lead concentrate average for six months has been 76.7 per cent Pb, 4.4 per cent Zn, and 1.1 per cent Fe.
in Table III. Operating conditions for the three DorrClones are shown in Table IV.

Moisture in Lead Concentrate Filter Cake

The only adverse result of the lead re-grind unit was the increase in moisture content of the lead concentrate filter cake. The moisture content increased from 6.5 per cent to 11 per cent as a result of the fineness of the grinding. At this moisture content the concentrate was sticky and sloppy, which added difficulties to the loading and trucking operation.

A front-end loader is used to pick up the concentrate from a pile and load it onto a truck. The wet concentrate would stick to the scoop of the loader, making filling and discharging of the scoop difficult.

The vibration from the truck traveling on the highway would cause the concentrate to flow, and it was necessary to have a perfect seal between the tail-board and the body to prevent loss of concentrate.

It was found that, by adding light fuel-oil to the pump feeding the filter, the moisture content of the filter cake fell to 9 per cent, and the cake discharging from the filter pelletized itself in rolling down the sides of the conical pile on the loading-floor, so that it looked like a pile of buckshot. This concentrate handled well in the scoop of the front-end loader, and in the truck. The fuel oil addition rate is 1 lb. per ton of dry solids.

Details of Equipment for the Re-Grind Section

The three DorrClones, size 12-in., were purchased for this section.

The zinc re-grind mill is a 7 ft. by 36 in. Hardinge conical ball-mill. The lead re-grind mill is a 6 ft. by 12 ft. cylindrical mill. These mills were part of the main grinding circuit. They are probably larger than required, but they were available and are serving the purpose adequately. No attempt has been made to determine the minimum size of ball-mill required.

Provision has been made to return part of the overflow from each DorrClone back to its feed pump so that a steady feed-pressure can be maintained. This feed-control is automatic on the lead rougher concentrate Dorr-Clone, and will be made automatic for the other Dorr-Clones.

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