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CHEMICAL AND METALLURGICAL USES FOR NATURAL GAS IN SOUTH AUSTRALIA

by

J.E.A. Gooden, R.G.P. McMahon, J.R. Tuffley and W.M. Walker

Investigated by: Chemical Metallurgy Section
Officer in Charge: J.E.A. Gooden

F.R. Hartley. Director

THE AUSTRALIAN MINERAL DEVELOPMENT LABORATORIES
Adelaide South Australia

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SUMMARY

Background

With the establishment of a pipeline from the north-east corner of South Australia to the capital, natural gas has become available within a quite small distance of a number of metalliferous ore deposits, rail heads, ports, and industrial centres. Amdel has, since its inception 12 years ago, worked on a number of projects, both for the Department of Mines of South Australia and commercial clients, which involve or could involve the use of natural gas. Such topics as the direct reduction or magnetising roasting of iron ores, lead-, copper- and zinc-smelting and refining, copper- and nickel-hydrometallurgy, ilmenite reduction leading to production of synthetic rutile, production of copper powder, and sulphur from gypsum have all been investigated.

The natural gas marketing company in South Australia is active in arranging contracts for sale of natural gas for industrial heating and their representatives have had several discussions with Amdel staff on metallurgical uses. Following a discussion in January 1970 the proposal was put to the Department of Mines that Amdel should make a thorough study of the potential chemical and metallurgical uses for natural gas in South Australia.

Objectives

The proposed programme consisted of a preliminary review of technology and selection of relevant overseas developments, followed by four and a half weeks' overseas investigation, concluding with a final appraisal of technology, study cost estimates and recommendations regarding establishment in South Australia of operations shown to be economically attractive. The Department of Mines, in accepting the proposal, specified three extractive metallurgical operations for study, viz:

1. the treatment of zinc silicate from South Australian deposits;
2. the production of copper powder from solutions formed by the acid or ammonia leaching of copper ores or concentrates; and
3. the production of iron powder and sponge-iron from South Australian iron ore deposits near the gas pipeline.

Summary of Work Done

Following the preliminary literature review and the overseas investigations, the technology of each of the above aspects was examined in detail.

The production of sponge-iron, steel strip, reinforcing rod and wire, and iron-powder, is considered and a preliminary economic evaluation for each of these products is presented.

A study of capital and operating costs for the Anaconda-Treadwell process for the production of copper powder or bars has been carried out and the results are compared with figures obtained in the overseas investigation.

Since there is no established technology for the production of zinc from zinc silicates using natural gas, a preliminary investigation of possible processes and types of equipment is presented.

Conclusions

It appears possible to mine iron ore in the Olary region of South Australia, rail it to Salisbury, there beneficiate it to produce iron ore concentrate pellets and reduce these by means of natural gas to produce 130,000 tons per year of iron in reduced pellets. It is assumed that 60,000 tons of this iron would be sold as sponge iron, the greater part overseas, and 70,000 tons melted to produce steel. Of this amount approximately 65,000 tons would be teemed into ingots for rolling to strip and reinforcing bar, and drawing into wire. This output appears to be equivalent to local South Australian demand. The remaining 5,000 tons approximately would be converted to iron powder for which there is an Australian demand for 3,000 tons per year, and it is assumed in the light of a preliminary appraisal of overseas markets carried out by International Technical Services Limited that 2,000 tons could be sold overseas. These subsequent steps appear economically attractive in this first feasibility study.

The flowsheet and materials balance provided by Treadwell Corporation were used to make independent estimates of the capital cost and heat consumption of the major items of plant equipment used in the Anaconda-Treadwell copper process. These independent estimates agreed closely with Treadwell Corporation's estimates and show that the costs of an Anaconda-Treadwell type custom refinery, at the 80,000 tons per annum scale, are favourable even assuming the worst circumstances considered. At the 10,000 tons per annum scale the advantages are not so clearly defined. If the capital investment must be raised by loans with payment of interest charges, then the custom refinery is clearly uneconomic. If no interest charges are applicable, however, then the custom refinery would show a saving of 0.3 to 3.3 cents per pound of copper when compared with shipping to Port Kembla for smelting and refining. The best saving of 3.3 cents per pound of copper on 10,000 tons per annum is equivalent to a 6.19% return on the required capital investment of $\$12.1 \times 10^6$.

1. INTRODUCTION

With the establishment of a pipeline from the north-east corner of South Australia to the capital, natural gas has become available within a quite small distance of a number of metalliferous ore deposits, railheads, ports, and industrial centres. Amdel has, since its inception 12 years ago, worked on a number of projects, both for the South Australian Department of Mines and commercial clients, which involve or could involve the use of natural gas. Such topics as the direct reduction of iron ores, magnetising roasting, lead, copper and zinc smelting and refining, copper and nickel hydrometallurgy, ilmenite reduction leading to production of synthetic rutile, production of copper powder, and sulphur from gypsum have all been investigated.

The marketing company is active in arranging contracts for sale of natural gas for industrial heating and their representatives have had several discussions with Amdel staff on metallurgical uses. Following a discussion in January 1970 the proposal was put to the Department of Mines that Amdel should make a thorough study of the potential chemical and metallurgical uses for natural gas in South Australia. The proposed programme consisted of a preliminary review of technology and selection of relevant overseas developments, followed by four and a half weeks' overseas investigation, concluding with a final review of technology, budget cost estimates and recommendations regarding establishment in South Australia of operations shown to be economically attractive. The Department of Mines in accepting the proposal specified three extractive metallurgical aspects for study, viz:

1. the treatment of zinc silicate from South Australian deposits;
2. the production of copper powder from solutions formed by the acid or ammonia leaching of copper ores or concentrates; and
3. the production of iron powder and sponge-iron from South Australian iron ore deposits near the gas pipeline.

It is assumed that this final report will probably be made available by the Department of Mines to commercial firms having a bona fide interest in applying the results. Such a firm might well require further detailed study of some aspect to be carried out at its own expense, but the present project will have served as a foundation.

1.1 Price of Natural Gas

The price of the gas is vital in any economic study, and will be taken for budget estimating purposes as 3.5 cents per therm at Peterborough, and 4.0 cents per therm in Adelaide, (1 therm = 10^5 Btu. The specification for calorific value of gas supplied from Gidgealpa is 950 - 1050 Btu per cubic foot).

2. PRELIMINARY EVALUATION (MID-1970) OF METALLURGICAL USES OF NATURAL GAS

2.1 Introduction

On the basis of data for South Australia and published information on overseas developments, a preliminary evaluation was made of the feasibility of using natural gas for the production of iron, copper, and zinc in this State. This evaluation, made in mid-1970, formed the basis of the planning of a series of plant visits overseas by J.E.A. Gooden in November 1970, (see Appendix A).

2.2 Production of Sponge-Iron

2.2.1 Definition

The term "sponge-iron" will be used to describe highly "pre-reduced" or "metallised" iron ore in the form of lumps or agglomerates. Conzinc Riotinto of Australia Limited (CRA) planned to build a plant in the Pilbara region to produce 1 million tons per annum (tpa) of such a product (which they termed "Himet") by SL/RN rotary kiln reduction of iron ore with coal. The product, for which a typical analysis is given below (Mackenzie, 1969) was quoted at approximately $38.5 per ton free on wharf, Port Dampier.

	%
Total iron	93.2
Metallic iron	90.2
Degree of reduction	96.8
C	0.43
S	0.027
P	0.045
SiO_2	3.2
Al_2O_3	1.3
CaO + MgO	<1.0

In Mexico the HyL process using natural gas has been operating at Monterrey since 1955 and new plants have been built at Vera Cruz and Puebla. Plants are understood to be mooted for southern USA and a number of other countries (Anon, 1969; Miller, 1970).

2.2.2 Uses

Two major uses for a sponge-iron product are envisaged in the Adelaide area:

1. as a feedstock for electric-furnace steelmaking in a mini-steel plant;
2. as a cupola feed in foundries.

A further application in South Australia for which the product may be exceptionally well suited is:

3. as a precipitant for copper from acid leach solutions.

Regarding (1), the suitability of sponge-iron for electric-furnace steelmaking has been established, and it can be used in combination with the conventional scrap, or as the sole charge. Large developments in this method of steelmaking on a large scale, particularly in Japan, are foreseen by CRA due particularly to falling power costs with the installation of nuclear power generators, and to the development of ultra-high power (UHP) electric furnaces with short fat arcs capable of more rapid melting (Innes, 1969; Innes and Melouney, 1969). This method is, however, attractive on a small scale and Miller (1970) lists no less than thirty mini-steel plants in the USA with steel outputs ranging from 50,000 to 300,000 short tons per year, based on electric-furnace melting (in most cases of scrap), in many cases continuous casting (down to 60,000 tons per year) and rolling. As mentioned in a letter to the Department of Mines (2 September 1970) two representatives from an Adelaide company consuming 12,000 tons per year of steel strip visited Amdel to discuss the possibility of building a mini-steel plant to supply their own needs and those of another company using approximately the same tonnage. They indicated that they at present buy mainly Japanese strip, as the quality of the corresponding BHP product is inadequate.

Regarding (2), it is believed that a sponge-iron product of suitable size and analysis would be attractive to foundries as cupola feed and might command a price between good scrap at approximately $38 per ton and virgin pig iron at approximately $54 per ton. This belief is partly based on the development of the Dwight-Lloyd McWane (DLM) process to produce large metallised pellets for foundry use and the proposed use of SL/RN and FIOR products for this purpose.

Regarding (3), it is likely that the high surface area of the sponge-iron would make it a uniquely good precipitant for copper from acid leach solutions, and it may be possible to produce, in a two-stage cementation process, a copper precipitate close to specification for milling to copper powder, and hence of much higher value than the usual "cement copper" (Groves, 1964). This outlet for sponge-iron is likely to be small in comparison with (1) and (2), however, and will not be further discussed here.

2.2.3 Production Processes

The most comprehensive reviews of direct reduction processes for iron ores are those by Wild (1961 and 1969). Considering only processes using methane (natural gas) as fuel, Wild (1969) lists nine new processes since his 1961 review, viz:

Purofer
Futakushi
FIOR
Phelps Dodge
Armco
IRSID
ICEM
Russian rotary kiln
Midrex

and the existence of a number of less-well-known processes is known to the present author (J.E.A. Gooden) from patent literature, etc. In such a field it seems wisest to select for detailed consideration only the processes for which commercial plants have been built and operated, or which have been proven at large pilot-plant scale resulting in construction or planning of commercial units. On this basis, only the following processes qualify:

a. Commercial Plants Believed to be Operating at Present

Process	Location	Output (approx.) tons/day
HyL	Monterrey, Mexico	200
	Monterrey, Mexico	500
	Vera Cruz, Mexico	500
	Puebla, Mexico	700
H-iron	USA	50 (?)
Phelps Dodge	Arizona	50
FIOR (Esso)	Dartmouth Nova Scotia	300 ?

Process	Location	Output (approx.) tons/day
Purofer	Oberhausen, Germany	500
Midrex	Portland, Oregon	1200

b. Commercial Plants under Construction or Planned

Process	Location	Output (approx.) tons/day
HyL	Arizona	?
(Nu Iron? HIB?)	Venezuela	3000
Futakushi	Philippines? Malaya?	300

Of the processes believed to be operating commercially the H-iron plant (if it has in fact been re-erected) and the Phelps Dodge plant are small special purpose plants, leaving only:

HyL
FIOR
Purofer
Midrex

for serious consideration. The HyL process reduces lump ore or pellets batchwise in reactors with reformed natural gas; the Purofer process uses a shaft furnace to reduce lump ore or pellets with reformed natural gas; the Midrex process also reduces pellets with reformed natural gas in a shaft furnace, and the FIOR process is a fluidised-bed reduction treating minus 4 to 8-mesh ore with steam-reformed natural gas.

2.2.4 Sources of Raw Material

Figure 1 gives a diagrammatic summary of sources of raw material for a sponge-iron plant in South Australia. It has been assumed that the operation would be independent of BHP though there would be nothing to prevent the undertaking buying Middleback Range or Western Australian ore if the price were right. It would be unwise, however, to rely heavily on BHP for ore while producing products expected to compete with BHP's pig iron or strip in the Adelaide area. Since an output of say 30,000 tpa of sponge iron (100 tons/day approximately) would appear from a general appraisal of plants to be a minimum, corresponding to say 60,000 tons of ore per year, and since at least

15 years' reserves would be necessary to build a plant it follows that iron ore deposits of less than 1 million tons cannot be considered, unless in the case of a number of nearby very easily mined smaller deposits.

<u>South Australian Ore Deposits</u>. Outside the Middleback Ranges, the only deposits believed, on the basis of reports by Whitten (1962 and 1963) to exceed 1 million tons are:

Razorback Ridge,	(100 million tons above plain).
Billeroo,	(5 deposits aggregating 1.5 million tons per 100 feet. Could possibly be mined in conjunction with Koolka, Plumbago and Old Toraminga Dam which aggregate 0.7 million tons per 100 feet).
Wilgena Hill,	(60 - 200 million tons).
Mount Christie,	(20 - 100 million tons).

The Razorback Ridge ore, while well situated, is unlikely to be treatable by any but chemical means (Gooden, 1963 and 1965), though the possibility of a high-temperature reduction with grain-growth cannot be completely discarded. It will be ignored in this section, but would technically be suitable feed for hydrometallurgical iron powder production (Section 2.3.2).

The Billeroo deposits only just aggregate sufficient ore for consideration. The ore is high in phosphorus but dry magnetic separation of a sample gave a concentrate assaying 61% Fe, 0.065% P_2O_5 (Bollen, 1962a), which would be acceptable. The ore could be transported by road to Olary and railed approximately 110 miles to Peterborough to the natural gas pipeline.

The deposits at Wilgena Hill and Mount Christie, while of acceptable size, appear too remote. However, an estimate of freight cost for 100,000 tons per year in bulk from Wilgena to Peterborough was obtained from the Commonwealth Railways. The ordinary classified rates for iron ore in bulk, excluding loading and unloading would be $5.20 for the 302 miles by Commonwealth Railways to Port Pirie, and $3.66 for the 72 miles by South Australian Government Railways to Peterborough, a total of $8.86 per ton. However, the spokesman for the Commonwealth Railways said they would be very interested in such a contract, and could probably reduce their price by $2.00 per ton. If the South Australian Government Railways did the same, the total would be approximately $5 per ton of ore (or concentrate). Samples of the Wilgena Hill ore, however, did not respond well even to flotation (which would be impracticable) (Hosking and Moskovits, 1960) and the Mount Christie material, unfortunately more remote, shows more promise of beneficiation. Sandman (1961) obtained a 62.3% Fe concentrate by dry magnetic separation and Bollen (1962b) obtained similar concentrates, the concentrate grade depending on the depth of the ore. The phosphorus content of a concentrate for which a complete analysis is shown was 0.10% P_2O_5 which could well be acceptable.

In summary, the most likely utilisation of South Australian iron ore would be the treatment of ore or concentrate from the Billeroo-Koolka-Plumbago-Old Toraminga Dam area with natural gas either at Peterborough or nearer to Adelaide.

South Australian Pyrite Calcine. Sulphuric acid production from Brukunga pyrite produces approximately 70,000 tons per annum of fine calcine assaying approximately 56% Fe which is understood to be disposed of cheaply at present. While direct magnetic separation is not very effective for upgrading, reduction-roasting to magnetite using gaseous reductant followed by magnetic separation is understood to yield a high-grade concentrate at high iron recovery. This appears an ideal feed for producing pellets for metallisation. The company involved has been actively engaged in studying outlets for their upgraded calcine and has indicated willingness to discuss results and possible commercial exploitation. (1972 Note - This operation has now ceased)

Local Scrap. The most likely use for local scrap in the enterprises considered would be as part of a composite sponge-scrap charge to an electric steelmaking furnace. The US Bureau of Mines (Prasky et al. 1970) has developed a process of heating light-gauge scrap with low-grade non-magnetic ore in a rotary kiln to yield a calcine which can be beneficiated magnetically to a high-grade magnetite concentrate. Enquiries will be made regarding local prices for such scrap. The process solves a waste disposal problem in the USA.

Imported Ore. Prices have not yet been obtained for Western Australian iron ore, but it is likely that coastal freight will be considerable.

2.2.5 Summary

On the basis of the above preliminary evaluation it was recommended that visits be made to the Midrex plant at Portland, Oregon, and the Purofer plant at Oberhausen, Germany, and to several mini-steel plants in the USA, particularly the Oregon Steel Mill supplied by the Midrex plant, and Soulé Steel, San Francisco; this last representing a smaller plant (60,000 tpa) having continuous casting. It was also recommended that discussions with the US

Bureau of Mines at Minneapolis, where a great amount of work on pre-reduction of iron ore has been carried out, and with Swindell Dressler (licensees for the HyL process) in Pittsburgh would also be fruitful.

2.3 Production of Iron Powder

2.3.1 Uses and Market

Finlayson and Morrell (1969) state that world consumption of ferrous metal powders has risen from less than 10,000 tpa to over 150,000 tpa in the last 20 years, and suggest an even more rapid rise in the next 5 years (500,000 tpa by 1973) if high tonnage applications are developed. The main applications are in sintered components (50% of powder used in the UK), welding and cutting, preformed forgings, and powder rolling to produce strip and foil.

Enquiries were made of a large Adelaide firm manufacturing certain automotive parts by powder-metallurgy techniques. Their spokesman estimated that they and three other (interstate) Australian undertakings probably used an aggregate of 500 tons of iron powder per year which is presently imported. The balance, and by far the greater part of the Australian consumption of 3,200 tpa (1968-1969) was, he said, used in welding rods. This Adelaide company has examined the possibility of producing their own iron powder, and have obtained samples of coal-reduced Yampi fines, rejected silica magnetically, and obtained excellent results in the use of the concentrate in their production line.

Regarding future Australian expansion of powder metallurgy techniques, he said this was bound to come. Regarding production of gears by this method he said it would not be suitable for gears requiring case-hardening, but for gears operating at high speed requiring good lubrication and not a high tooth load, powder-metallurgical techniques would be suitable. However, there were no presses in Australia at present capable of producing gears of greater than 3 inches outside diameter.

It is apparent that any iron-powder producing enterprise should plan to export.

It was announced (Anon, 1970) that the Titan Manufacturing Co. would produce iron powder in Australia, but the announcement describes the various processes for producing iron powder in fairly general terms. BHP representatives indicated in late 1970 that a final decision had not been made on a process, but it was most likely to be atomisation of liquid metal. Whether firm commitments have been made is not known.

2.3.2 Production Processes

World production of iron powder in 1970 was analysed by Finlayson and Morrell (1969) as follows:

Type of Process	% of World Production	Suitable Raw material
Reduction	28	High-grade iron ore or mill scale.
Atomisation	60	High-quality scrap or hot metal.
Electrolytic	2	Low-grade steel plate, cast iron, etc.
Hydrometallurgical	10	Any iron-bearing raw material, including scrap, ilmenite, pyrites and pickle liquor.

Reduction. The Hoganas (Sweden) or Sieurin method reduces high-grade magnetite with coke and limestone in saggers in a tunnel kiln. The Petrofina and Pyron processes (Canada) reduce iron oxide with hydrogen on a moving belt. Crushing, anealing, grinding, screening, blending are carried out on the reduced product. The International Nickel Company of Canada Limited use the carbonyl route.

Atomisation. The Mannesmann process atomises a stream of molten metal with a jet, the DPG process by impingement on a spinning disc. Numerous variations are extant and various post-treatments are used to give required powder properties. It is stated that many small-scale plants exist throughout the world for the production of iron, iron-alloy and non-ferrous metal and alloy powders by atomisation.

Electrolytic. Anodes of impure iron are used in solutions of iron sulphate or chloride.

Hydrometallurgical. The RCA-Peace River-Woodall Duckham process produces ferrous chloride from a wide range of feedstocks including pyrite, ilmenite, pickle-liquor, low-grade scrap and ore, reduces the ferrous chloride with hydrogen and grinds the sponge. A pilot-plant commenced production at the end of 1969 at Windsor, near Detroit.

2.3.3 Value of Product, and Cost of Production

Reduced or atomised powder (minus 100 mesh) sells at approximately

£Stg80 (Australian prices of $170 - $190 have been quoted). Electrolytic powder sells at £Stg224 per ton.

From their analysis, Finlayson and Morrell (1969) state that for plants of capacity less than 15,000 tpa atomisation is likely to provide the cheapest powder. Above 15,000 tpa the hydrometallurgical process will, they consider, win provided that a saleable by-product, such as synthetic rutile, is made.

Since electrolytic powder appears rather an expensive speciality, and the above authors consider gaseous reduction of ore to be the most expensive of the remaining routes, attention will be confined to:

a. the atomisation process, and
b. the hydrometallurgical process.

2.3.4 Sources of Raw Material in South Australia

Molten metal would be available for atomisation from a mini-steel plant as envisaged in Section 2.2.2. A number of small atomisation operations are understood to be operating around the world, and the route is very versatile.

Suitable feed for the hydrometallurgical process would be local ore, scrap, or pyrite, or imported ilmenite, the two latter producing sulphur and synthetic rutile, respectively, as valuable by-products.

The former process would use natural gas in annealing, and the latter process would use gas to produce hydrogen for reduction of the ferrous chloride.

2.3.5 Summary

On the basis of the preliminary evaluation presented above it was recommended that a small atomisation operation, and the Peace River pilot plant at Windsor, Ontario, be both visited to gain insight into pilot and labour requirements and costs. In addition, consumption trends should be discussed with an institute or research association authoritative in this field, for example Battelle Memorial Institute at Columbus, Ohio, or BISRA in England.

2.4 Production of Copper Powder

2.4.1 Uses and Market

Copper powder is used in powder metallurgy mixed with tin powder to make sintered bronze components. Brass powder is generally pre-alloyed. Copper powder is also used in sintered iron components as a cementing agent.

Evans, Romanchuk and Machiw (1961) give the annual US consumption of copper and copper base-metal alloy powder in 1959 as 40 million pounds (20,000 short tons), a per capita consumption of about 0.25 lb per annum for that year. Annual US consumption from 1960 to 1964 is shown in Figure 2.

The Australian imports of copper and copper-base alloy powders and flakes are shown in Figure 3. The 1967-1968 values were 4,228 cwt of copper powder and flake and 4,291 cwt of copper-alloy powder and flake, a total of 8,519 cwt or 950,000 lb, a per capita consumption of about 0.08 lb per annum. Linear projection of the values shown in Figure 2 suggests an Australian consumption of 570 tons valued at $1 million in 1973. However, the disparity between Australian and United States per capita consumption may indicate that the Australian consumption will increase much more rapidly than is shown by the linear projection.

It is not possible to separate the amounts of flake and powder imported, but it is believed from enquiries that powder would predominate, an assumption which is supported by the US data in Figure 2.

Statistics have been sought regarding copper-powder consumption in New Zealand, South Africa and South-east Asia. No values have been obtainable, however, either locally or from the Copper and Brass Information Centre in Sydney.

A copper powder operation should aim to export, for example to south-east Asia. Processes are believed to be well developed overseas for rolling copper powder to strip and the strip output might well exceed the output sold as powder. Considerable tonnages of copper strip are known to be used in South Australia. (Later information favours melting rather than direct-rolling of powder.)

2.4.2 Production Processes

A recent literature survey by the present author (J.E.A. Gooden) lists the following routes:

Electrolytic
Reduction of oxide
Reduction from solution by hydrogen
Comminution of metal
Atomisation of liquid metal
Evaporation of metal
Thermal decomposition of salts
Cementation from solution
Other reduction processes

While the first is believed to be currently the most important commercially, the relative importance of the other routes is not quantitatively known.

The atomisation route could be followed in a plant considered in Section 2.3 and alloy powders could be produced in this way, but South Australia's main interest is in hydrogen-reduction of acid or ammoniacal leach solutions.

Hydrogen - Reduction of Ammoniacal Copper Solutions. Sherritt Gordon, at Fort Saskatchewan in Canada, have done a great deal of pioneering in this field and demonstrated that a copper powder very suitable for powder metallurgy could be produced. They also produce a number of other metal and alloy powders by this route. The process is also used by Universal Metals (now Metal Powders Inc.) at Kansas City to produce 10 tons of copper per day.

Hydrogen - Reduction of Acid Copper Solutions. This more recent development has also been worked on by Sherritt Gordon, and is used by Arizona Chemcopper Company at Bagdad, Arizona in a plant designed to produce 25 tons of copper powder per day. (Personal communication from Mr J. Rosenbaum of the US Bureau of Mines suggests that the plant was initially under-designed, as due to various problems only a fraction of this output was achieved. Recent reports indicate that problems are being overcome and the output is rising slowly towards the design value.)

Hydrogen Reduction of Cuprous Cyanide. This new route is a feature of the Anaconda-Treadwell process which is believed to have considerable potential and relevance to South Australia. Briefly, it is a process for treating copper sulphide concentrates by slurrying them with concentrated sulphuric acid and coating the slurry onto pebbles which are baked in a rotary kiln. The products are copper sulphate, elemental sulphur and sulphur dioxide. The latter is reconverted to sulphuric acid in a conventional contact acid plant, the sulphur is purified for sale and the copper sulphate dissolved and the solution purified. While solvent extraction followed by electrowinning is a possible copper recovery method, the companies prefer to precipitate cuprous cyanide which is reduced with hydrogen to give pure copper powder. This acid attack is seen as promising because it eliminates air pollution due to SO_2, converting sulphide copper quantitatively to elemental sulphur, its most concentrated and saleable form. It is considered relevant to South Australia because it permits the marrying of sulphide and oxide ore treatment, producing a copper sulphate solution from the sulphide concentrate which can be combined with that from acid vat-leaching of oxide ore.

For this reason we have kept in touch with the Treadwell Corporation, who have supplied us with samples of the baked, coated pebbles from the kiln. The acid leach section of the pilot plant in Arizona has started and was expected to be operating smoothly by November 1970. It was understood that the cyanide section would be slightly delayed.

2.4.3 Source of Raw Materials in South Australia

The cost of transport to interstate or overseas smelters provides an incentive for local treatment of the State's numerous deposits of copper ore. While Burra ore is to be ammonia-leached, a number of other oxide deposits are amenable to acid-leaching. Attractively low prices are being quoted for sulphuric acid produced as a by-product of non-ferrous smelting. The Anaconda-Treadwell process gives the ability to treat sulphide copper concentrates and oxide ores or concentrates using a common copper recovery process. Scrap should also be taken into consideration.

2.4.4 Summary

Several South Australian companies are already interested in the possibility of entering hydrometallurgical extraction of copper.

It was recommended that visits be made to Sherritt-Gordon's hydrogen-reduction plant in Canada, Chemetals Corporation in New York, Arizona Chemcopper and the Anaconda-Treadwell pilot plant (both in Arizona) and possibly the plant at Kansas City which produces copper powder by hydrogen reduction from ammoniacal solution.

2.5 Production of Zinc from Silicate Ore

The use of natural gas to produce zinc from silicate ore is an interesting possibility, but no technology has been developed. More details should be sought of the chemical composition and mineralogy of the South Australian zinc silicate ore. It would be attractive to produce zinc metal directly from the ore or a concentrate by the use of natural gas, though the metal would almost certainly require fractional distillation for lead removal to achieve the high purity currently required.

Production of zinc oxide by allowing the zinc vapour to reoxidise, as in slag fuming, would not be so attractive. It would admittedly save freight and simplify leaching, but the zinc oxide product would require densification or agglomeration and would still only be a feed for acid leaching. Reduction and reoxidation would thus only be a concentration stage. Nevertheless, the economics of such a step will be briefly examined in later sections of this report. Work has been done at Amdel on the reduction of solid 17.5% Zn blast-furnace slag with hydrogen and carbon monoxide. To avoid melting, reaction temperatures were limited to below 950°C, at which temperature the rate is slow and the equilibrium constant small. Natural zinc silicate will have the advantage of a higher melting point. If the zinc silicate were melted a slag-fuming-type operation could be envisaged, but natural gas, despite certain Russian claims (Gooden, 1964) appears to be an inferior fuel for slag fuming. The use of a cyclone furnace should be considered.

Doerner (1936) made a thorough practical study of the reduction of zinc oxide sinter with natural gas at temperatures up to 1,050°C firstly in laboratory equipment, and then in a semicontinuous retort producing approximately $^1/_3$ lb per hour of metallic zinc. A number of problems were overcome, the reoxidation of zinc vapour by CO_2 and H_2O being prevented by catalysing the reaction of these latter with excess methane. A large pilot plant was designed, but as far as is known, not built.

3. RESULTS OF OVERSEAS VISITS AND DISCUSSIONS

3.1 Notes on Overseas Visits

Notes on each of the nineteen overseas visits made during November - early December 1970 are presented in Appendix A. Included in these notes are reprints of brochures and papers relevant to the subjects under study.

3.2 Summarised Review of Technology Based on Overseas Visits

3.2.1 Sponge Iron and Iron Powder

Figure 4 has been prepared to show in a generalised way the routes using natural gas which have been examined overseas, and which, with the exception of iron-ore leaching, are all in commercial operation now, or have been recently, or will be shortly. It must, of course, be borne in mind that the iron powders produced by the various routes differ in their properties and applications, but various post-treatments are capable of making a given type suitable for other applications.

A comparative assessment will be made in Section 4 of this report of the applicability of these processes in South Australia. Companies engaged in metallised pellet production, and atomised powder production indicated the possibility of their supplying capital if such operations were shown to be viable in South Australia.

The commercial utilisation of pyrite cinders was also inspected and discussed.

3.2.2 Copper Powder

Information has been obtained on three methods of producing copper powder from an ammoniacal copper solution:

1. Precipitation of metal powder by hydrogen produced by reforming natural gas. This operation on a commercial scale has been inspected.
2. Precipitation of black cupric oxide by boiling and subsequent solid-state reduction with gas.
3. A possibility mentioned by research staff was to precipitate basic copper carbonate and reduce this hydrothermally with hydrogen (produced by reforming natural gas).

The commercial plant in Arizona producing copper powder by hydrogen precipitation from acid solutions is understood to have encountered severe corrosion problems.

The Anaconda-Treadwell process being piloted in Arizona has been inspected and detailed cost estimates obtained. This process would use natural gas to produce copper briquettes for melting, or powder, or continuously-rolled strip. The proponent suggested that another company (General Electric) for which he consults might be interested in establishing a commercial copper-production operation in South Australia.

A paper assessment of these copper processes using natural gas is made in Section 5 of this report.

3.2.3 Zinc from Silicate Ores

Data have been obtained from laboratory work on the gaseous reduction of zinc silicate, and from pilot-scale gaseous reduction of zinc oxide sinters. Woodall-Duckham have developed in the laboratory an HCl leach process which they would like to try on samples of South Australian zinc silicate ores, but the process would not necessarily use much natural gas. This area is also assessed on paper, in Section 6 of this report, as far as is possible with the research data available.

4. ECONOMICS OF AN INTEGRATED OPERATION IN SOUTH AUSTRALIA TO PRODUCE SPONGE-IRON, STEEL AND IRON POWDER

4.1 Markets

4.1.1 Sponge Iron

It is very likely that a highly-metallised sponge-iron product could be sold in South Australia and interstate for foundry use, particularly if briquetted, and any excess exported to Japan. The market would probably have to be worked up and iron founders convinced of the value of this material. Hamersley Iron were active in advertising their "Hi-met" product which they were considering producing in Western Australia at the rate of 1.4 million tons per year from Pilbara ore and Blair Athol coal by the SL/RN process. The preliminary quotation of price in 1970 was $38.50 on wharf at Dampier.

The interstate and overseas markets should be examined as exports would permit larger-scale pre-reduction at lower unit cost, but may require purchase of ore or pellets from Whyalla, Tasmania or Western Australia to supplement local reserves (see Section 4.2.4). For the purpose of this report it will be assumed that 60,000 tons per annum of metallic iron are sold as sponge, either as-reduced or briquetted, in South Australia, interstate, or overseas.

4.1.2 Steel Strip

A South Australian tube manufacturer has stated his steel strip requirement as 12,000 tpa. He also states that another firm may use 12,000 tpa so that a market for (say) 25,000 tpa of strip will be assumed. Welded tube manufacture probably restricts the steel-making operation to the use of a major proportion of virgin iron and a minor proportion of purchased scrap, in order to keep the content of residual metals down. It will be assumed, therefore, that the proposed steel-making operation will use only sponge iron and "in-plant" scrap.

The above manufacturer has also indicated his interest in later making alloy steels.

4.1.3 Reinforcing Bar and Wire

Most of the mini-steel plants in the USA produce reinforcing bar and wire, which appear ideal products not requiring the massive rolling equipment necessary for large sections and plate.

The manager of a South Australian plant involved in this field estimated in 1971 the South Australian consumption of merchant reinforcing bar, of

$^{3}/_{8}$-inch diameter and above, including deformed bar of sizes 4 to 11, as 20,000 tons (±5%) and the consumption of drawn wire, including galvanised wire for fencing, as 20,000 tons (±10%).

On this basis a local market for 40,000 tons of bar will be assumed, covering bar used as such and bar subsequently drawn to wire. Such a bar-rolling and wire-drawing operation is carried out at Etiwanda, California (see Appendix A) as well as elsewhere.

4.1.4 Iron Powder

The Australian consumption of iron powder is at present approximately 3,000 tpa, of which by far the greater part is used in acid (rutile-coated) welding electrodes. Mr W. Speed of ETSA, Federal President of the Australian Welders Institute, said he believed the time was ripe to produce the iron powder in Australia and listed suppliers of the welding rods on the Australian market (in approximate order of importance), as:

Lincoln Electric Co.
EMF
Murex (subsidiary of EMF)
Weldex (Philips) (Made in Holland)
Australian Welding Industries.

This application is particularly interesting as Mannesmann in Germany (see Appendix A) produce a powder suitable for this application, simply by water-atomisation of electric-furnace-melted metal, drying and screening - without any reduction or heat-treatment after atomisation. The plant required would therefore be fairly simple. Glidden-Durkee have a similar process and AO Smith also, it is believed.

A survey of overseas markets has been made for this project by International Technical Services Limited (Appendix E). While there is competition from Europe, and markets for iron powder have scarcely developed in many of Australia's neighbours, an export market of 2,000 tpa will be assumed making a total market of 5,000 tpa.

4.1.5 Total Outlet for Sponge Iron

From the foregoing sections, the following will be assumed:

	Tons Fe pa
Sponge iron, as-reduced and briquetted	60,000
Steel strip	25,000
Reinforcing bars and wire	40,000
Iron powder for welding rods	5,000
	130,000

A production of 130,000 tpa of metallic iron will require the production and processing of 210,000 tons of iron ore concentrate assaying approximately 64% Fe.

4.2 Sources of Raw Material

These were considered in a preliminary way in Section 2 of this report.

4.2.1 Iron Ore from Olary

Appendix B lists a number of deposits within 30 to 40 miles of the Olary Railway Station. After brief discussion with Mr G. Whitten the assumptions set out in Appendix C have been made giving assumed reserves which can be summarised as follows:

	million tons	Grade % Fe
High-grade ore	1.8	66
Concentrates from lower grade ore*	2.0	61
Total	3.8	-

* 3.0 million tons at 48% Fe.

At a production rate of 160,000 tpa of concentrates the above reserves would last 23 years; at 210,000 tpa they would last 18 years.

A freight rate from Olary to Salisbury, which has been assumed as the site for a pelletising, pre-reduction and melting plant, was sought from the South Australian Railways. A senior officer advised that while the scheduled rate would be $8.14 per ton, a rate of $4.50 per ton could be assumed for 100,000 tpa of iron ore. Bogey exchange would take care of change of gauge and a siding exists at Olary. The rate would be subject to adjustment each 12 months, and amounts to approximately 2 cents per ton-mile.

4.2.2 Iron Ore from Mount Christie

The Mount Christie deposit is 17 miles from the Wynbring railway station which is 377 miles from Port Pirie. Assuming a final negotiated freight rate of 1 cent per ton-mile on Commonwealth Railways and 2 cents per ton-mile on South Australian Railways, the rail freight to Salisbury would be $6.00 per ton. Proved reserves approach 20 million tons, with possible reserves of 20 to 100 million tons. At the freight rate quoted it would be necessary to beneficiate the ore, grading about 38% Fe, at the mine. Beneficiation testwork on drill-core material has been carried out at Amdel.

It appears that the cost of concentrate from this source would be a little higher than from Olary, so that Mount Christie will be considered as a reserve which could be brought into production if a large increase in sponge-iron production were required.

4.2.3 Pyrite Calcine from Birkenhead

As mentioned in Section 2 of this report, sulphuric acid production from Brukunga pyrite produced approximately 70,000 tpa of fine calcine assaying approximately 56% Fe. It was shown to be possible to upgrade this calcine to better than 67% Fe, so that a supply of 50,000 tpa of concentrates assaying at least 65% Fe was assumed. However, this operation ceased in 1972 and the calculations have therefore been revised to omit this source of iron ore feed.

4.2.4 Purchased Ore or Pellets

Since the purpose of this project is to develop unused South Australian resources, purchase of ore or pellets will not be considered in the basic operation. If markets develop it may be possible later to expand production and hence decrease unit costs. In this case, due to limited reserves in the Olary province, iron ore or pellets may have to be purchased.

4.2.5 Local Scrap

"Overseas Trade" for the year 1968-1969 shows 123,578 tons of iron and steel scrap waste, valued at $3.15 million (ie. $25.1 per ton average) exported from South Australia to Japan. While local scrap is the feed for almost all of the "mini-steelplants" in the USA, it will not be considered for the plant under discussion.

4.2.6 Cost of Olary Ore, Delivered to Plant

In order to produce as steady a feed as possible to the reduction stage, it is assumed that both lower-grade Billeroo ore, and high-grade ore are mined and concentrated separately but simultaneously and the concentrates combined in equal proportions. The high-grade ore will probably require grinding and magnetic separation to remove phosphorus as well as for pelletising. On the basis of work by Bollen on surface samples only, it will be assumed that 48% Fe lower-grade Billeroo ore can be upgraded to a 61% concentrate representing 65% of the original weight. The high-grade ore apparently assays 63 to 67% so that production of a 67% concentrate from it will be assumed in 95% of ore weight. The required tonnage of 210,000 tpa of 64% Fe concentrate is assumed to be made up as follows:

	Conc Grade Fe %	Conc Weight tpa	Tons Ore*
Billeroo low-grade	61	115,000	177,000
High-grade ore	67	95,000	100,000
Total	63.7	210,000	277,000

* Tons of ore to be mined and transported.

It will be assumed that the annual tonnage of feed (277,000 tons) is mined by open-pit methods in an annual campaign of several months by a large contractor. He would move in on several of the deposits, mine ore and waste rock, load the ore into motor trucks and cart approximately 30 miles to stockpiles (low-grade and high-grade) at the Olary siding. From these stockpiles the broken ore would be loaded throughout the succeeding 12 months on to rail trucks for transport to the plant.

Budget costs assumed are as follows:

	Cost per ton of Ore Railed, $	
	Low-grade	High-grade
Mining ore, ($0.80 per ton)*	0.80	0.80
Mining waste rock, ($0.50 per ton)*	0.50	2.00
Transport of ore to stockpile*	0.80	0.80
Loading on to rail trucks*	0.20	0.20
Freight to Salisbury**	4.50	4.50
Cost of ore on rail trucks at Salisbury	6.80	8.30

* Budget estimates by W.P. Boyce, Operations Research/Computer Services Section, Amdel.

** See Section 4.2.1.

From the previous tabulation the annual cost of ore on rail trucks at Salisbury is $2,030,000 or $9.68 per ton of resultant concentrate.

4.3 Production of Concentrate

The beneficiation plant will treat per annum:

177,000 tons as-mined low-grade ore (48% Fe)
100,000 tons as-mined high-grade ore (65-67% Fe)

A first estimate flowsheet by L.J. Weir is given in Figure 6. It assumes that the ores mined will contain predominantly magnetite, but includes provision for gravity cleaning of a hematitic material. The following budget estimates have been made:

	$
Purchased equipment cost	520,000
Installation 200%	1,040,000
Fixed Capital Investment (FCI)	1,560,000

	$/ton Ore
Power	0.10
Steel	0.07
Water	0.05
Stores	0.10
Direct labour	0.40
Maintenance	0.20
Direct Operating Cost	0.92
Indirect Costs:	
Payroll overhead, laboratory, plant overhead	0.40
Fixed Operating Costs	
Depreciation (10% of FCI)	0.57
Interest on capital (7% of FCI)	0.40
Total Operating Cost	2.29

ie. $3.10 per ton of concentrate.

4.4 Pelletising

It is assumed that 210,000 tpa of fine concentrate, 64% Fe, is balled and then fired on a travelling grate. It is assumed, without actual experimental data for this material, that firing to 1,050°C will give the pellets sufficient strength for the reduction process. The capital cost of the balling and firing plant will be taken as $2 million.

The direct operating cost will be taken as $1 per ton of concentrate and indirect and fixed costs as $1.80 per ton of concentrate, giving a total cost of pelletising as $2.80 per ton.

4.5 Reduction

From overseas visits and enquiries, there are five possible processes for metallising iron ore pellets using solely a gaseous reductant:

a. Allis Chalmers Direct Reduction Process
   (rotary kiln, natural gas is reformed within the kiln burden).

b. Armco Process
   (shaft furnace, natural gas reformed externally).

c. HyL Process
   (batch reactors, external reforming).

d. Midrex Process
   (shaft furnace, external reforming).

e. Purofer Process
   (shaft furnace, external reforming).

Of these, little information has been released on the Armco or Purofer processes and only HyL and Midrex processes are operating on full commercial scale. Midland-Ross will not consider a Midrex plant smaller than 400,000 tpa of metallised pellets. The HyL process is well-proven and plants of the required size have been built, and it will therefore be used for the present budget costing. However, the Allis Chalmers process is very attractive in that the expense of external reformers is removed, so that the economics should be more favourable. Further information is awaited from the USA.

The capital cost of a HyL plant consisting of one reformer furnace, four reducing reactors and four heat accumulators, and including water recirculating system, transformation from 13.2 kV incoming electric power, control house and other directly-related facilities within the battery limits will be taken as $8 million.

The direct and indirect operating cost breakdown by Swindell-Dressler for a plant of similar size can be itemised for the plant under consideration as given in Table 1, giving a *direct plus indirect operating costs* of $11.50 per ton of iron in reduced pellets. Fixed charges will be taken as follows:

	$/ton Fe in Reduced Pellets
Fixed Operating Cost	
Depreciation	0.61
Interest on capital	0.43
Total Operating Cost	12.54

4.6 Melting and Casting

Swindell-Dressler recommend continuous casting only if the production rate is greater than 100,000 long tons per annum of metal. The assumed melting capacity is 74,000 tpa of iron in reduced pellets which at 95% yield gives 70,000 tons of molten metal, so that a single 20-ton electric arc furnace would be used teeming into ingots. The budget *Fixed Capital Investment* quoted by Swindell-Dressler is $4 million.

The above company gives an analysis of direct plus indirect operating costs, excluding electricity, for a 250,000 tpa electric furnace plant melting 60% HyL sponge and 40% scrap. Adjustment of their costs for South Australian conditions gives $17.2 per metric ton of iron. For the smaller plant presently under consideration, with teeming into ingots, a figure of $20 per ton of iron in reduced pellets will be assumed, to which will be added $6* for electricity (600 kWh per ton of iron). Fixed charges will be taken as follows:

	$/ton Fe in Reduced Pellets
Depreciation (10% of FCI)	0.54
Interest (7% of FCI)	0.38
Total Operating Cost	26.92

4.7 Rolling

The cost of rolling the ingots to strip and reinforcing bar, and drawing of wire will not be considered at this stage.

4.8 Production of Iron Powder

Woodall-Duckham Limited have supplied a capital cost estimate and quantities

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* The Tariff Officer of ETSA advised that while the regular rate for approximately 70 million kWh per annum would be approximately 1.1 ¢ per kWh, a rate lower than 1 ¢ may be negotiable, particularly if interruptable supply were agreed to by the user.

of items required for operation for the production of 5,000 tpa of iron powder using Glidden-Durkee technology, including induction-furnace melting of low-carbon scrap (see Appendix D). The heat-treatment step included would produce a product suitable for powder metallurgy, in addition to the as-atomised product for electrode coating. It will be assumed that the powder plant is located adjacent to the electric furnace steelplant, and that ladles of molten metal are carried from the latter to the former, so that no melting is required in the powder plant. However, it would be wise to include the induction furnace for holding of the molten metal when required, temperature and carbon adjustment, refining or alloying if required, etc. Minor adjustments have, therefore, been made to the data supplied.

The plant is assumed to produce 5,000 tpa of iron powder working two shifts per day for 250 days per annum. Woodall-Duckham suggest a Fixed Capital Investment, under Australian conditions, of $1.3 million.

Since molten metal will be transferred from the melt-shop, the power consumptions will be taken as only 150 kWh per ton in the direct operating cost analysis set out in Table 2, giving:

	$/ton Product
Direct Operating Cost -	25.78
To which must be added:	
Indirect Operating Cost - (Payroll overhead, laboratory, plant overhead)	12.00
Fixed Operating Cost - Depreciation (10% of FCI)	26.00
Interest on capital (7% of FCI)	18.20
Total Operating Cost -	81.98
	(say)$82.00

It should be noted that these capital and operating costs apply to the production of a powder metallurgy grade of powder by subsequent heat treatment after atomisation.

The cost of production of the molten metal will be taken as $5 less than the cost of producing billets (Table 3), viz. $62.30 per ton of metal. Hence total production cost for iron powder is $144.30 per ton of metal.

4.9 Summary of Costs of Products Compared with Market Values

The total costs of production of the various intermediate and final products are set out in Table 3. The costs of iron-ore concentrate and pellets produced in this small-volume operation from small and relatively remote deposits appear somewhat high by comparison with prices obtained by Australian mining companies exporting to Japan, but the pellet price is comparable with prices paid by USA iron and steelmakers. The marketable products are discussed below.

4.9.1 Reduced Pellets (Sponge Iron)

HyL sponge-iron produced from Alzada pellets is 87% metallised, but higher metallisations can be obtained with a little additional gas consumption. Hamersley Lurgi pellets used by Cavaghan and Harris are quoted as containing 86.4% metallic iron, 91.6% total iron, ie. 94% metallised. A detailed appraisal would require examination of gangue constituents, but an estimated total production cost of $37.50 per ton of iron in a South Australian product assumed to assay 90% total iron, ie. $34 per ton of product, appears to have a margin for profit when compared with $38.5 per ton of Himet on the wharf at Dampier. (This latter price was quoted before the recent labour cost increases.)

4.9.2 Steel Billets

The estimated total production cost of $67.70 per ton for steel billets compares with the following Australian (wholesale) prices cif Australian capital city ports supplied by BHP (20 May 1971):

	$ per Ton
Billets for re-roll	85 (basic price) plus surcharges eg. mild steel $+1
Merchant bar-rounds	103 (basic price) plus surcharges eg. $^1/_2$ inch $+4 $^3/_8$ inch $+8

There appears to be ample margin for profit. Distribution charges would be added to the above prices to give the retail prices from distributors which are understood to be about 20% higher.

4.9.3 Iron Powder

Australian prices of $170 and $190 per ton were quoted in 1969 for two of the more common grades of imported iron powder. There appears to be a small margin to cover sales costs and profit with an estimated total production cost of $144.30 per ton. It should be noted that this total production cost corresponds to manufacture of a heat-treated powder for powder-metallurgy, which probably commands a price higher than those quoted above.

4.9.4 Capital Investment

The estimated fixed capital investment for each stage of processing is also shown in Table 3. The total for the whole complex is estimated at $17 million.

4.10 Conclusions

It appears economically feasible to mine iron ore in the Olary region of South Australia, rail it to Salisbury and there beneficiate it, produce iron ore concentrate pellets and reduce these by means of natural gas to produce 130,000 tpa of iron in reduced pellets. It is assumed that 60,000 tons of this iron would be sold as sponge iron, the greater part overseas, and 70,000 tons melted to produce steel. Of this amount approximately 65,000 tons would be teemed into ingots for rolling to strip and re-inforcing bar, and drawing into wire. This output appears to be equivalent to local demand. The remaining 5,000 tons approximately would be converted to iron powder for which there is an Australian demand for 3,000 tpa, and it is assumed in the light of a preliminary appraisal of overseas markets (Appendix E) that 2,000 tons could be sold overseas. These subsequent steps appear economically attractive in this first feasibility study.

4.11 Recommendations

It is important that plans be made to prove the iron ore reserves and confirm the amenability to concentration of ore at depth. Diamond drilling (to the depth assumed to be mined) should be carried out at several of the larger deposits (both high and lower-grade) and the core supplied to Amdel for benefication testing.

5. THE ANACONDA-TREADWELL COPPER PROCESS

5.1 Introduction

During visits to Arizona and New York late in 1970, Mr Gooden obtained flowsheet and cost information for a recently developed hydrometallurgical process for pure copper production. The information obtained related to a plant capacity of 80,000 short tons of copper per annum.

The aim of this Section of the Report is to determine the economic feasibility of this process for South Australian conditions. Estimates for an Australian plant of 80,000 tpa copper capacity will be made, therefore and compared with Anaconda-Treadwell estimates for 80,000 short tons per annum of copper. From the former, estimated costs will be derived for a plant of a capacity of 10,000 tpa of copper. This was chosen as a realistic order of magnitude for South Australia since it corresponded to the sum of the outputs from Kanmantoo (7,000 tpa Cu) and from Mount Gunson (3,000 tpa Cu).

5.2 Process Flowsheet

The flowsheet adopted is shown in Figure 7. The process is briefly described below.

The copper concentrate is attacked with 90 to 98% sulphuric acid to form copper sulphate and other metal sulphates, releasing the sulphide sulphur as elemental sulphur and the sulphuric acid being reduced to water and sulphur dioxide:

$$CuFeS_2 + 4H_2SO_4 = CuSO_4 + FeSO_4 + S_2 + 2SO_2 + 4H_2O$$

This attack is carried out partially in a pug mill and is completed in an externally-heated kiln from which air is excluded. The temperature is held at 250 to 280^{o}C. Pebbles preheated in a kiln are also fed, the copper and other metal sulphates forming a coating on the pebbles. The water, sulphur dioxide and elemental sulphur are largely vaporised from the kiln. The coated pebbles are cooled in a second kiln by a countercurrent stream of sulphur dioxide, which also volatilises the remaining sulphur. The pebbles are then countercurrently washed on two screens, dried and preheated before recycling to the kiln.

The gases from the kiln and sulphur vaporiser pass to two water scrubbers in series where elemental sulphur is recovered as a slurry. This sulphur assays 97 to 98% S and can be purified by melting and filtration. The sulphur dioxide from the scrubbers is oxidised for conversion to sulphuric acid for re-use.

The metal-sulphate-containing solution (60 g/l Cu) from the pebble washers is thickened and filtered to remove waste insoluble residue, and then purified by sparging in sulphur dioxide to precipitate silver and other metals which are removed by filtration.

The purified solution then passes to a reactor where the copper is precipitated as cuprous cyanide by the reaction with hydrogen cyanide (as an aqueous solution containing 25% HCN) and sulphur dioxide, sulphuric acid being regenerated:

$$2CuSO_4 + SO_2 + 2HCN + 2H_2O = 2CuCN + 3H_2SO_4$$

The precipitated cuprous cyanide is removed by continuous pressure filtration, washed with copper sulphate solution and then twice with water to eliminate free cyanide, dried in a multi-hearth dryer to 3% moisture, and then briquetted in compacting, pocketed rolls.

These briquettes are reduced in a kiln by the passage of hydrogen from a gas reforming plant thus regenerating hydrogen cyanide gas for recycling:

$$CuCN + {}^{1}/_{2}H_2 = Cu + HCN$$

The gas is water-scrubbed to form the 25% HCN solution required for the precipitation stage.

The copper-barren solution, after the cuprous cyanide precipitation, contains 18% H_2SO_4 and ferrous and other metal sulphates. This solution is concentrated in a triple effect evaporator to nearly 40% H_2SO_4. Ferrous sulphate heptahydrate is crystallised from this solution by refrigeration to $0^{o}C$ and removed by centrifuging, the 40% H_2SO_4 solution then being further concentrated to 80% H_2SO_4 by submerged combustion. This acid is then blended with oleum to give 90 to 98% acid for re-use.

The ferrous sulphate heptahydrate crystals are melted and spray dried at $260^{o}C$ to give the sesquihydrate, which is roasted to give magnetite and sulphuric acid for recycle.

The raw materials required for the process are oleum, to replace sulphuric acid losses; hydrogen cyanide, to replace losses; and natural gas, both for heating purposes and for reforming to hydrogen and also for the production of inert gas for purging and blanketing for safety reasons.

The products from the plant are copper, magnetite and elemental sulphur.

5.3 Approximate Sizing and Costs of Major Items of Equipment

The main factors which have been used in determining the approximate equipment sizing for cost purposes are discussed in the following sections.

5.3.1 Overall Plant Capacity

Data have been obtained from the Treadwell Corporation for the costs of an 80,000 short tons per annum plant. It was suggested that the minimum economic size of plant would probably be of the order of 3,500 to 5,000 tpa. A custom treatment plant in South Australia would require a capacity of about 10,000 tpa (Section 5.1). It was decided, therefore, to check out the data obtained for the 80,000 tpa plant and to estimate the probable costs for a 10,000 tpa plant.

5.3.2 Mass Balances

The mass balances throughout the process are shown on the flowsheet in Figure 7. The data are presented as long tons per hour for an 80,000 tpa copper plant, assuming a working year of 8,000 hours. The values shown in parenthesis refer to a 10,000 tpa copper plant.

With effective recovery of SO_2 from the ferrous sulphate and other streams, then the only make-up sulphuric acid required will be to replace stack and filter losses. If other metal sulphides are also present in the concentrate, then the SO_2 from these sources should be sufficient to replace losses. In fact, Anaconda-Treadwell derived a nett gain in sulphuric acid of 3 tons per day. For this exercise it has been assumed that no make-up will be required.

Hydrogen cyanide will be required to make up losses from filtration and gas scrubbing. Anaconda-Treadwell assume a make-up of 2% of the circulating load. This would appear to be adequate and the same value has been assumed for this exercise.

5.3.3 Heat Balance

Process heating, using both natural gas combustion and steam, will be required. Cooling, both by cooling water and by refrigeration, will also be required. The following estimates were prepared for the major users for comparison with the Anaconda-Treadwell data. The values refer to an 80,000 tpa plant, whilst the values in parenthesis refer to a 10,000 tpa plant:

1. Leaching System

Heat required to raise to 280°C:

$$\begin{array}{lll} \text{concentrate} - & 33.3 \times 2240 \times 260 \times 0.24 \times 1.8 & = 8.4 \times 10^6 \text{ Btu/hr} \\ \text{acid} - & 76.1 \times 2240 \times 260 \times 0.402 \times 1.8 & = \underline{32.0 \times 10^6 \text{ Btu/hr}} \\ & & 40.4 \times 10^6 \text{ Btu/hr} \end{array}$$

Heat recovered from pebbles preheated to 400°C:

$110 \times 2240 \times (400 - 280) \times 0.24 \times 1.8 = 12.8 \times 10^6$ Btu/hr

Hence heat to be supplied $= 27.6 \times 10^6$ Btu/hr

$(3.45 \times 10^6$ But/hr)

Note: The overall heat of reaction, assuming preheated reactants and volatilisation of steam and sulphur is not expected to be great, and has been neglected.

Heat to dry and preheat recycled pebbles, assuming 5% moisture:

water - $110 \times 2240 \times 0.05 \times 1000 = 12.3 \times 10^6$ Btu/hr

pebbles - $110 \times 2240 \times 380 \times 0.24 \times 1.8 = 40.4 \times 10^6$ Btu/hr

52.7×10^6 Btu/hr

$(6.6 \times 10^6$ Btu/hr)

Hence total heat to be supplied: 80.3×10^6 Btu/hr

$(10.05 \times 10^6$ Btu/hr)

Assuming 50% thermal efficiency, then heat to be supplied as natural gas: 160.6×10^6 Btu/hr

$(20.1 \times 10^6$ Btu/hr)

ie. 3860×10^6 Btu/day, cf. 4000×10^6 Btu/day from Anaconda-Treadwell.

2. Sulphuric Acid Evaporation

From the flowsheet we find that an 18.5% solution containing 37.9 tons per hour of H_2SO_4 is to be evaporated by triple effect evaporation to give a 40% solution, ie. 110 tons per hour (tph) of water is to be removed.

Assuming 2.5 lb of water removed per pound of steam for triple effect evaporation, gives a steam requirement of 2.46×10^6 lb per day, cf the Anaconda-Treadwell estimate of 2.23×10^6 lb per day. After crystallisation of the process of ferrous sulphate, the H_2SO_4 is further concentrated by submerged combustion to 80%. The removal of the water of crystallisation associated with the $FeSO_4$ leads to a 50% H_2SO_4 leaving the crystalliser. Hence to give an 80% concentration requires the removal of a further 28.5 tph of water, ie. a heat load of 90×10^6 Btu/hr. Assuming a

75% heating efficiency leads to a requirement of 2900 x 10^6 Btu per day, cf Anaconda-Treadwell estimate of 6000 x 10^6 Btu per day.

3. Ferrous Sulphate Crystallisation

Considering a two-stage crystalliser, with the feed solution being cooled in the first stage by heat exchange from the ferrous sulphate-free liquor leaving the second stage, and with the second stage being cooled directly by refrigeration, we find:

Total feed,	119 tph at 60°C
Total liquid effluent,	75 tph at 50°C
Total solid effluent,	44 tph
Feed temperature to second stage,	29°C
Feed to second stage (assuming half of solid removed from first),	95 tph acid and 12 tph (dry) solids

Heat to be removed by refrigeration is:

$$\begin{aligned} 95 \times 29 \times 1.8 \times 0.68 \times 2240 &= 7.55 \times 10^6 \text{ Btu/hr} \\ 12 \times 29 \times 1.8 \times 0.25 \times 2240 &= \underline{0.35 \times 10^6 \text{ Btu/hr}} \\ & 7.9 \times 10^6 \text{ Btu/hr} \end{aligned}$$

ie. 660 tons of refrigeration. Using ammonia as refrigerant, the power required is 700 hp.

Assuming 2 x 50 hp centrifuges, then total power required is 800 hp. Hence daily requirement is 16,000 kWh (assuming a power factor of 0.9), cf. Anaconda-Treadwell estimate 17,200 kWh.

4. Ferrous Sulphate Drying and Roasting

The roasting is exothermic and so the only heat required is for spray drying of the heptahydrate. The heat requirement for drying heptahydrate to anhydrous ferrous sulphate is 93 kcal/g mole, ie. 59.4 x 10^6 Btu per hour. The daily requirement is, therefore, 1425 x 10^6 Btu per day, and assuming 50% efficiency, natural gas is required to provide 2850 x 10^6 Btu per day, cf. Anaconda-Treadwell estimate of 2690 x 10^6 Btu per day.

5. Overall Balance

In general, the above calculations serve to prove the validity of the data supplied by Anaconda-Treadwell, the major users agreeing fairly well. The power requirements for pumping, etc. from Anaconda-Treadwell have been accepted without question. The steam requirements have been translated by Anaconda-Treadwell to a natural gas usage assuming a thermal efficiency of 75%. The expansion of high-pressure steam to 20 psi heating steam has also been credited as electrical power. The overall requirements given by Anaconda-Treadwell, therefore, appear accurate. For an 80,000 tons per day plant these are:

Power - 119,000 kWh/day
Natural gas - 17,700 x 10^6 Btu/day
Water - 2,600 x 10^3 gal/day

For a 10,000 tpd plant, utilities will be required in approximately direct proportion to throughput, ie. -

Power - 14,900 kWh/day
Natural gas - 2,210 x 10^6 Btu/day
Water - 324 x 10^3 gal/day

5.3.4 Cost of Sulphate Leach Equipment

Assuming that a half-hour residence time is required in the pugging operation, the required equipment is 4 x 10,000-gallon vessels each with 200-hp agitators.

Assuming 1 hour residence times in the kilns, then two kilns each 13-ft diameter by 400-ft long are required.

Estimated approximate purchase costs for these main items of equipment are as follows:

Pugging,	vessels, 4 x 10,000 gal	\$0.04 x 10^6
	agitators 4 x 200 hp	\$0.2 x 10^6
Kilns,	2 off each 13-ft diameter x 400 ft	\$2.2 xx 10^6
	Total	\$2.44 x 10^6

cf. \$US2.54 x 10^6 Anaconda-Treadwell estimate.

5.3.5 Cost of Solubles Separation from Pebbles

Assuming drag classifier type washers and assuming 10 tons per day (tpd) per foot width, then classifiers 18-ft wide are required. Assuming a half-hour residence time in the rotary drier for pebble drying and preheating, an 11-ft diameter by 250-ft long drier is required.

Approximate purchase costs are as follows:

Washers,	\$4,000/ft wide, 2 off	$\$0.14 \times 10^6$
Drier,	11-ft diameter x 250 ft	$\$0.50 \times 10^6$
		$\$0.64 \times 10^6$

cf. \$US0.43 x 10^6 Anaconda-Treadwell estimate.

5.3.6 Cost of Gas Scrubbing

A total of 30.2 tpd SO_2 requires scrubbing before acid manufacture, ie. approximately 10,000 scfm. From Semrau (1971), the cost of gas cleaning this amount of SO_2 by wet scrubbing, direct-contact gas cooling and wet electrostatic precipitation is \$0.45 x 10^6, cf. \$US0.71 x 10^6 Anaconda-Treadwell estimate.

5.3.7 Sulphur Dioxide Handling

The cost of a compressor to handle 10,000 scfm is \$0.1 x 10^6, cf. \$US0.28 x 10^6 Anaconda-Treadwell estimate.

5.3.8 Solution Purification

Assuming 50 sq ft per ton per day solids for the thickener, then the required thickener area is 3,200 sq ft.

Assuming a filtration rate of 50 lb per hour per square foot for the underflow filter, then an area of 200 sq ft is required.

For sulphur dioxide treatment of the clarified liquor, assuming a half-hour residence time, a 16,000-gallon vessel with 30 hp agitator is required.

Assuming a flowrate through the final clarifying pressure filter of 1 gallon per minute per square foot, we find a required filtration area of 800 sq ft.

The approximate purchase costs are as follows:

Thickener, 3200 sq ft	\$0.12 x 10^6
Rotary filter, 200 sq ft	\$0.06 x 10^6
Vessel, 16,000 gallons	\$0.02 x 10^6
30 hp agitator	\$0.01 x 10^6
Pressure filter, 800 sq ft	\$0.04 x 10^6
	\$0.25 x 10^6

cf. \$US0.31 x 10^6 Anaconda-Treadwell estimate.

5.3.9 Cyanide Precipitation

The total flow into the precipitation reactor is 40,000 gallons per hour and the SO_2 gas flowrate is 64,000 s cu ft per hour. Allowing a 2-hour residence time for liquid, plus a 5-minute residence time for gas, we find the total volume of reactor required is 120,000 gallons. The approximate purchase cost is estimated as:

12 x 10,000-gallon vessels	\$0.12 x 10^6
12 x 200 hp agitators	\$0.60 x 10^6
	\$0.72 x 10^6

cf. \$US0.81 x 10^6 Anaconda-Treadwell estimate.

5.3.10 Cyanide Filtration

Assuming 50 lb per hour per square foot as the filtration rate, then four filters each 700 sq ft are required, ie. one for initial separation, one for a sulphate solution wash, and two for water washing.

The estimated cost for each is \$0.06 x 10^6, ie. total \$0.24 x 10^6, cf. \$US0.47 x 10^6 Anaconda-Treadwell estimate.

5.3.11 Briquetting and Reduction

Assuming 50% moisture in the filter cake feed to the dryer, then the required evaporation is 32,000 lb per hour. An approximate cost for a double shell dryer of this capacity from Perry (1950) is \$0.08 x 10^6.

Assuming the purchase cost of briquetting equipment to be twice that of an equivalent capacity disc pelletising plant, then the approximate cost is \$0.07 x 10^6. Assuming a 4-hour residence time in a rotary reducing kiln to provide complete recovery of HCN, then a kiln of about 11-ft diameter by 250-ft long will be required. A conventional kiln of these dimensions would cost approximately \$0.5 x 10^6. Considering extra expense due to external heating and due to necessity for air exclusion, a cost of \$0.75 x 10^6 has been assumed.

For HCN scrubbing, a 10-ft diameter by 15-ft high packed column has been estimated at a cost of \$0.06 x 10^6. The total purchased equipment cost for briquetting and reduction is therefore estimated at \$0.92 x 10^6, cf. \$US1.06 x 10^6 Anaconda-Treadwell estimate.

5.3.12 Melting and Casting

No estimate has been made for this operation. The Anaconda-Treadwell estimate is \$US1.02 x 10^6.

5.3.13 Acid and Iron Recovery

The triple effect evaporator system has a total evaporation of 110 tph water. Assuming a heat transfer coefficient of 200 Btu per square foot per hour degree F, and a temperature difference of 50°F, an area of 8,000 sq ft is required in each effect. The estimated cost of each effect is \$0.3 x 10^6.

The total required throughput of the crystallisers is 1,050 tpd (say 2 x 500 tpd units). The estimated cost of each unit is \$0.25 x 10^6. Two 70-in. diameter centrifuges will be required for crystal separation, at an estimated cost of \$0.13 x 10^6, each. The spray dryer for the heptahydrate drying requires a capacity of 20 tph evaporation. From Perry (1950), a cost of \$10 per pound of water evaporated per hour seems reasonable, ie. an estimated cost of purchased equipment of \$0.4 x 10^6.

The dried sesquihydrate crystals will be roasted in a 10-ft diameter fluid-bed roaster. The estimated cost of the roaster is \$0.15 x 10^6. The cost of gas cleaning has already been estimated in Section 3.6. A disc pelletising plant for the magnetite residue would cost approximately \$0.06 x 10^6.

A cost of \$0.05 x 10^6 has been assumed for the submerged combustion acid concentration.

The estimated total cost of an acid plant to produce 38.2 tph (1,000 tpd) is \$1.6 x 10^6. Assuming the purchase cost of the major items to be one-third of this installed cost, we find the purchase cost to be \$0.53 x 10^6.

Hence the total estimated purchase cost of the acid and iron recovery section is \$2.85 x 10^6, cf. the Anaconda-Treadwell estimate of \$US3.00 x 10^6.

5.3.14 Tailings Treatment

The cost of tailings treatment will depend greatly on location. The Anaconda-Treadwell estimate of \$US0.13 x 10^6 has been assumed.

5.3.15 Utilities

The cost of the complete installation of packaged-unit utility facilities was estimated as follows:

1. Inert Gas

   A plant to produce 25,000 scfd of inert gas (for purging and other safety reasons) by the stoichiometric combustion of natural gas was estimated to cost \$0.08 x 10^6.

2. Hydrogen

   A steam-natural gas catalytic reformer to produce 3.9 tpd of H_2 (10^6 scfd) as a 70% hydrogen mixture was estimated to cost \$0.6 x 10^6.

3. Cooling Water

   A cooling tower to cool 36 x 10^6 gallons per day of water through 20°F, would cost \$0.5 x 10^6.

4. A steam-raising plant to generate 2.34 x 10^6 lb per day of steam at 600 psi has a total cost of \$0.6 x 10^6.

5. A steam turbine to expand 2.34 x 10^6 lb per day steam from 600 psi to 20 psi and to generate 100,000 kWh per day of electricity has a total cost of \$0.2 x 10^6.

6. A 660-ton refrigeration facility has a total cost of \$0.3 x 10^6.

Hence the cost of the complete installation of packaged unit utility facilities is estimated as \$2.28 x 10^6. Assuming a purchase cost of the major items to be one-third of this installed cost, we find the purchase cost of the utility facilities to be \$0.76 x 10^6, cf. the Anaconda-Treadwell estimate of \$US0.78 x 10^6.

5.4 Plant Total Capital Costs

It can be seen that the approximate costs estimated above for an 80,000 tpa plant are very similar to the costs estimated by Anaconda-Treadwell.

The total purchase cost for the major items of equipment is estimated to be \$10.52 x 10^6, and assuming an installation factor of 4.0 (the average installation factor, excluding working capital, quoted by Buchanan and Sinclair, 1964) we find the total fixed capital required to be \$42.1 x 10^6.

The equivalent fixed capital estimate by Anaconda-Treadwell for an

80,000 tpa plant is:

Total purchased equipment	\$US11.52 x 10^6	\$A 9.6 x 10^6
Installation, etc.	\$US30.2 x 10^6	\$A27.5 x 10^6
Total	\$US41.7 x 10^6	\$A38.0 x 10^6

Including working capital, the total capital required for an 80,000 tpa copper plant is:

$$4.64 \times 10.52 \times 10^6 = \$A49 \times 10^6$$

The capital cost for a 10,000 tpa plant is estimated from:

$$\frac{42.1 \times 10^6}{8^{0.67}} = \$10.4 \times 10^6$$

or \$12.1 x 10^6 including working capital.

5.5 Manufacturing Cost Data

The manufacturing costs are detailed in Table 4 for an 80,000 tpa plant and in Table 5 for a 10,000 tpa plant.

For a production of 80,000 tpa, the manufacturing costs, excluding the cost of the concentrate feed, amount to \$10.78 x 10^6 pa with a further \$2.94 x 10^6 pa required for interest charges, ie. a cost of 6.2 cents per pound of copper (7.9 cents per pound if interest charges are added).

For a production of 10,000 tpa, the manufacturing costs, excluding the cost of the concentrate feed, is \$2.27 x 10^6 pa with a further \$0.73 x 10^6 pa required for interest charges, ie. a cost of 9.9 cents per pound of copper (13.2 cents per pound if interest charges are added).

(Note: 1. No allowance for profit has been made, and
2. No credit has been assumed for the production of sulphur and magnetite).

5.6 Discussion

Very good agreement was achieved between this estimate of capital cost and that of Anaconda-Treadwell. Close agreement was also achieved in the heat balances over the major items of equipment.

The treatment costs can be compared with refining costs quoted by the E.R.S. Company of Australia Limited, for example, who charge a fixed cost per ton of material treated, plus an amount which varies with the price of copper.

The charges and deductions for a 30% copper concentrate (assuming no silver or gold values) amount to 9.9 cents per pound of copper with a recovery of 96%. This cost includes profit by E.R.S. Company of Australia Limited, and so the actual refining cost is somewhat less. If one allows for the shipping cost of the concentrate from Port Adelaide to Port Kembla at $13 per ton, then the total cost becomes 12.3 cents per pound of copper.

To assess the advantages of the installation of a custom refinery, using the Anaconda-Treadwell process, in South Australia, one must assess the probable accuracy of the foregoing capital and operating cost estimates.

It should be noted that the major cost centre in the Anaconda-Treadwell process, both for capital and operating costs, is in the acid recovery section. Recovery of acid from ferrous sulphate is also the centre of greatest uncertainty. A number of such plants for recovery of acid from pickle liquors and in titanium dioxide manufacture have been built and subsequently shut down as being uneconomic. The acid and iron recovery section for the adopted flowsheet accounts for 27% of the capital expenditure, 25% of the labour, 30% of the power and 5% of the natural gas supply, to recover and recycle some 300,000 tpa of H_2SO_4 (80,000 tpa Cu plant). From this it can be calculated that the cost of recovery amounts to $14.3 per ton of H_2SO_4 for the 80,000 tpa plant, and $24 per ton of H_2SO_4 for the 10,000 tpa plant.

In both cases, therefore, the cost of recovering the acid is apparently less than the cost of purchasing acid.

The total estimated operating costs are about 60% dependent on capital cost assessment and 40% on assessment of raw materials, utilities and labour. The probable tolerances on these estimates are ±10% on the assessment of raw materials, utilities and labour (about 50% of these are dependent on natural gas), and plus 30% minus 10% on the assessment of capital charges. Hence, the overall tolerance on the estimated operating cost is about plus 20% minus 10%.

The refining costs therefore become:

10,000 tpa plant - 9 to 12 cents per lb Cu
or 12 to 15.8 cents per lb Cu if interest charges are added.

80,000 tpa plant - 5.7 to 7.5 cents per lb Cu
or 7.2 to 9.5 cents per lb Cu if interest charges are added.

It can be seen, therefore, that at the 80,000 tpa scale, the costs of an Anaconda-Treadwell type custom refinery are favourable even assuming the worst circumstances. At the 10,000 tpa scale, the advantages are not so clearly defined. If the capital investment must be raised by loans with payment of interest charges, then the custom refinery is clearly uneconomic. If no interest charges are applicable, however, then the custom refinery would show a saving of 0.3 to 3.3 cents per pound of copper when compared with shipping to Port Kembla for smelter refining. The best saving of 3.3 cents per pound of copper on 10,000 tpa is equivalent to a 6.1% pa return on the required capital investment of \$12.1 x 10^6.

6. PRODUCTION OF ZINC FROM SILICATE ORES

6.1 Introduction

Exploration by the Electrolytic Zinc Company of Australia Limited has established the existence of a large deposit of high-grade zinc silicate ore near Beltana in the northern Flinders Ranges. The zinc silicate mineral present is willemite, Zn_2SiO_4. In their 1969 Annual Report, reserves were quoted at 827,000 tons containing 34.5% zinc and 2.8% lead, and 237,000 tons containing 14.3% zinc and 10.6% lead. This ore is currently being treated in their acid leaching process at Risdon, Tasmania. The Director of Mines directed that as one aspect of Project 1/1/129, the treatment of zinc silicate ores using natural gas should be examined.

6.2 Pyrometallurgical Processes Available for Zinc Extraction

Zinc extraction processes fall broadly into three types; electrolytic, electrothermic, and thermal reduction with carbon. Commercial pyrometallurgical reduction processes are carried out in horizontal retorts, vertical retorts, and lead-zinc blast furnaces. Some of the process equipment used in these processes may be applicable to a process for reducing zinc silicate ores with natural gas.

6.2.1 Horizontal Retort Process

The Horizontal Retort Process (Mathewson, 1959) treats oxidised zinc ores, usually sintered or nodulised, in batch retorts. The ore is mixed with a carbonaceous reductant and charged to a refractory retort. Each retort is fitted with its own condenser, and as many as five hundred retorts may be contained in one furnace. A carbon blanket at the mouth of each retort is used to reduce CO_2 to CO and prevent reoxidation of zinc during cooling.

The equipment and batchwise nature of the process are basically unsuitable for use in a natural gas reduction process.

6.2.2 Vertical Retort Process

In the Vertical Retort Process the charge of oxidised zinc ore and coal moves continuously by gravity through the retort. The charge is agglomerated to ensure free movement of the solids, permit heat transfer to the centre of the charge, and provide sufficient porosity for the escape of gases. The charge is premixed, briquetted and passed through a coking furnace before being charged to the retort.

The reduction reaction is carried out at 1,100 to 1,200°C to ensure that only a minimum of carbon dioxide is formed. A little zinc is re-oxidised and tends to form accretions inside the top of the retort. Zinc vapour is condensed rapidly in a zinc-splash condenser which avoids the formation of blue powder. The zinc product can then be refined by liquation and/or continuous fractional distillation to produce high-grade zinc.

The equipment used in the Vertical Retort Process is potentially suitable for conversion to a natural gas reduction process. Some difficulties are envisaged in the conversion, and these will be discussed in a later section of this report.

6.2.3 Blast Furnace Process

The lead-zinc blast furnace is fed with sinter and coke, and is run in a similar manner to the lead blast furnace. Molten lead and a molten slag are tapped from the bottom of the furnace, while zinc vapour is distilled from the top of the furnace. Zinc is condensed in a lead-splash condenser and the two metals are separated by liquation. Zinc produced by this method is of Prime Western grade containing about 1.5% lead. High-grade zinc can be produced by distillation.

The recirculating stream of lead contains in excess of 2% zinc. High-grade zinc can be produced from this stream by passing it through a continuous vacuum de-zincing unit. This vacuum de-zinced (VDZ) metal has a purity of 99.9% zinc, with impurity levels of 0.03% lead, 0.02% cadmium and 0.003% iron.

Treatment of even a high-grade zinc silicate ore or concentrate in an Imperial Smelting furnace would give a larger volume of slag (due to silica and the fluxes required by the silica) and hence a higher treatment cost than obtains at Cockle Creek where the silica content is approximately 2%. However, the condenser and vacuum de-zincing unit would be suitable in any process which produces zinc vapour.

6.2.4 Natural Gas Reduction

The basic units required in a process for reducing zinc silicate ore with natural gas are a reduction furnace or retort and a condenser. It is envisaged that either a zinc-splash or lead-splash condenser would be used; both of these being proven commercial units. Refining of the zinc product and production of high-grade VDZ metal are also proven processes.

Therefore, the only major item of equipment which requires investigation is the reduction furnace or retort. The vertical retort may provide the basis for a suitable unit and would probably require an agglomerated feed in the form of pellets or briquettes.

6.3 Thermochemical Considerations

6.3.1 Reduction of Zinc Oxide

Reduction of zinc oxide with natural gas was investigated extensively by the US Bureau of Mines (Doerner, 1936; Van Zante and Gorski, 1950). It was found that the primary reduction reaction was:

$$ZnO + CH_4 = \underset{(gas)}{Zn} + CO + 2H_2 \qquad \ldots (1)$$

This was accompanied by two secondary reactions:

$$ZnO + CO = \underset{(gas)}{Zn} + CO_2 \qquad \ldots (2)$$

$$ZnO + H_2 = \underset{(gas)}{Zn} + H_2O \qquad \ldots (3)$$

The amounts of CO_2 and H_2O formed were appreciable. Equilibrium gas partial pressures for the overall reduction (Reactions 1, 2 and 3) are listed in Table 6 (Doerner, 1936).

Zinc vapour was re-oxidised to ZnO by the reverse of Reactions 2 and 3 during cooling and condensation. This re-oxidation was prevented by reacting excess CH_4 with CO_2 and H_2O to form CO and H_2 by Reactions 4 and 5:

$$CH_4 + CO_2 = 2CO + 2H_2 \qquad \ldots (4)$$

$$CH_4 + H_2O = CO + 3H_2 \qquad \ldots (5)$$

Both of these reactions required a catalyst.

6.3.2 Reduction of Zinc Silicate

If a parallel is drawn with zinc oxide reduction, it can be assumed that the primary reaction for zinc silicate reduction is as shown in Reaction 6:

$$Zn_2SiO_4 + 2CH_4 = \underset{\text{(gas)}}{2Zn} + 2CO + 4H_2 + SiO_2 \qquad \ldots (6)$$

Secondary reactions would be:

$$Zn_2SiO_4 + 2CO = \underset{\text{(gas)}}{2Zn} + 2CO_2 + SiO_2 \qquad \ldots (7)$$

$$Zn_2SiO_4 + 2H_2 = \underset{\text{(gas)}}{2Zn} + 2H_2O + SiO_2 \qquad \ldots (8)$$

Values of the free energy of reaction ($\Delta G°_T$), log Kp, Kp, and equilibrium partial pressures of the gaseous components for Reaction 6 are listed in Table 7. Values of $\Delta G°_T$, log Kp and Kp for Reactions 7 and 8 are listed in Table 8.

If Reactions 6, 7 and 8 are considered together, the gas phase will contain six constituents - Zn, CH_4, CO, CO_2, H_2 and H_2O. The equilibrium partial pressures for the six gases can be calculated from the equilibrium constants and from the stoichiometry of the reactions. Partial pressures are listed in Table 9.

It can be seen that even at 1500°C the partial pressures of CO_2 and H_2O are very small, indicating that Reactions 7 and 8 do not proceed to any significant extent. As a result there should be negligible re-oxidation of zinc vapour.

It is possible, therefore, that reduction of zinc silicate with natural gas could be more easily accomplished than reduction of zinc oxide with natural gas. However, the presence of calcium and magnesium carbonates in the ore could alter conditions radically. Emission of large quantities of CO_2 from carbonate decomposition would create suitable conditions for re-oxidation of zinc vapour. This problem could possibly be overcome by prior upgrading of the ore to remove carbonates. Upgrading could be achieved by a physical separation process (eg. heavy media) or by calcining. The latter would provide a preheated feed to the reduction furnace.

Reaction 6 is endothermic and requires 107 kcal per gram mole of zinc at 1,100°C. Internal heating would not be possible because of the large quantity of carbon dioxide that would be produced. Hence the reduction unit must be externally heated and this almost certainly limits the reduction unit to a vertical retort.

6.4 Composition of Beltana Ore

In the 1969 Annual Report for Electrolytic Zinc Industries Limited, ore reserves at Beltana were quoted at 827,000 tons containing 34.5% zinc and 2.8% lead, and 237,000 tons containing 14.8% zinc and 10.6% lead. The compositions of various samples which have been forwarded to Amdel suggest that the ore contains approximately 5% Fe, 4% Ca, 4% Mg. The calcium and magnesium are both present as carbonates. However, heavy media separation produced an upgraded product containing approximately 57% Zn, 3% Pb, 4% Fe, 0.4% Ca and 0.4% Mg. The carbon dioxide content of the ore would then be slightly in excess of 1% and should not prove troublesome.

6.5 Vertical Retort for Zinc Silicate Reduction

The vertical retort requires an agglomerated feed in the form of briquettes, sinter or pellets. The feed should also contain a minimum of carbonates. The optimum method of achieving these two requirements would need investigation.

During the reduction stage, lead compounds would be reduced to metallic lead. If the lead completely volatilised, it should be no problem in a lead-splash condenser but could be troublesome if it formed a liquid pool in the base of the retort, or if it solidified causing blockages. It is possible that a separate outlet point could be provided for molten lead.

The US Bureau of Mines did considerable research into the reduction of zinc oxide by natural gas in the early 1930s and mid-1940s. They used a metal retort to facilitate heat transfer through the retort wall. Considerable difficulty was experienced in finding a suitable alloy. Hence, the question of refractory versus metallic retort would need to be considered.

By reference to Table 6, it can be seen that considerable CO_2 and H_2O are formed during zinc oxide reduction. Oxidation of zinc was prevented by reacting excess CH_4 with these gases according to Reactions 4 and 5. The reactions needed a catalyst which was provided by soaking alundum grains in a solution of nickel nitrate and reducing in a current of hydrogen at 600° to 700°C.

However, the catalyst bed often became poisoned with deposited impurities and proved troublesome.

The need for a catalyst in the zinc silicate reduction process is uncertain. If testwork shows that a catalyst is required then consideration as to the type of catalyst and arrangement of the catalyst bed would need further consideration.

The process envisaged may be summarised as follows:

1. Upgrade the ore by heavy-media separation to produce a feed which is high in zinc and low in carbonates.
2. Briquette, pelletise or sinter the feed.
3. Reduce with natural gas in a vertical retort made of thermally-conducting refractory or metal.
4. If necessary, provide a catalyst for Reactions 4 and 5.
5. Condense the zinc vapour in a lead-splash condenser.
6. Produce VDZ metal from the recirculating lead.

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APPENDIX A

DETAILED NOTES ON OVERSEAS VISITS AND DISCUSSIONS

Notes on the nineteen visits made by J.E.A. Gooden in November 1970 are set out in the following pages, and tables, figures, and extracts from reprints are given at the end of each set of notes. Only the most relevant sections of reprints are included.

		Pages
Visit 1.	Anaconda-Treadwell Pilot Plant, Arizona Production of Copper Sulphate & Copper Metal.	1 - 5
Visit 2.	Soule Steel Company, California Production of steel reinforcing bars & fence posts from steel scrap.	6 - 9
Visit 3.	Ameron Steel Producing Division, California Production of steel reinforcing bars etc from scrap steel.	10 -13
Visit 4.	Sherritt Gordon Mines Ltd., Alberta Production of Copper Powder.	14 -39
Visit 5.	U.S. Bureau of Mines, Minneapolis Information on Iron Ore pre-reduction.	40 -41
Visit 6.	Metal Powders Inc., Missouri Production of Copper powder from Scrap & Cement Copper.	42 -57
Visit 7.	Midrex Division, Midland-Ross Corporation, Ohio The Building & Operation of Metallised Iron Pellet Plants.	58 -82
Visit 8.	Peace River Mining & Smelting Co., Ontario Production of Iron Powder from scrap.	83 -99
Visit 9.	Swindell-Dressler Company, Pennsylvania The Hyl Direct Reduction Process in Steel Making.	100 -119
Visit 10.	Jersey Enterprise Inc., New York The Fior Process to Produce Briquettes for Steel Furnace.	120 -123
Visit 11.	Hydrocarbon Research Inc., New York Production of H-Iron Powder.	124 -138
Visit 12.	Treadwell Corporation, New York Production of Copper.	139 -144
Visit 13.	British Iron & Steel Research Association, London Discussion on Iron Powder Production.	145 -146
Visit 14.	Woodall-Duckham Ltd., Crawley, England Production of Iron Powder.	147 -154
Visit 15.	British Iron & Steel Research Association, Sheffield, England Production of Alloy Powder, Stainless Steel Strip and Iron Powder.	155 -156
Visit 16.	British Iron & Steel Research Association, Grangetown, England Direct Reduction Processes for Iron Ore.	157 -160
Visit 17.	Mannesmann Pulvermetall Germany Production of Iron Alloy Powders.	161 -164
Visit 18.	Montecatini-Instituto, Italy Treatment of Pyrite cinders.	165 -166
Visit 19.	Montecatini Edison Treatment of Pyrite cinders.	167 -185

APPENDIX A

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See previous page.

VISIT 1: ANACONDA-TREADWELL PILOT PLANT

Situated at rear of
ANACONDA RESEARCH CENTRE,
TUCSON, ARIZONA

DATE: Monday 2 November 1970

MET: Mr E.P. Cadwell, Vice-President, Treadwell Corporation

1. INTRODUCTION

Mr. Cadwell explained that he had come to live in Tucson to supervise construction and operation of the pilot plant, which will have a throughput of 6 tons chalcopyrite cons (2 tons copper) per 24 hours and will have cost $US 5 million. He is a mining and milling engineer who has worked in a number of copper operations in North and South America and retired from Cyanamid Corp. some years ago to join the Treadwell Corp., who now retain him to supervise the piloting of the joint Anaconda-Treadwell process.

The process consists of two stages - the first is production of a copper sulphate solution from the sulphide concentrate, and the second is the production of copper metal from the solution via a cyanide route. It is this latter stage which Mr Cadwell conceived as a result of observations when working for Cyanamid Corporation. The first stage is operating continuously on a 10-day campaign, 4-day shut-down basis, but the units for the second stage (cyanide process) are still being constructed. Mr Cadwell said he could talk freely as all I could see was covered by patents already issued or in process.

2. CHEMISTRY OF THE PROCESS

Amdel had already received a flowsheet from Mr Cadwell, but the acid-attack reaction is:

$$CuFeS_2 + 4H_2SO_4 = CuSO_4 + FeSO_4 + S_2 + 2SO_2 + 4H_2O \qquad \ldots (1)$$

The reaction occurs with chalcopyrite and other copper minerals above 180°C, and below 250°C there is no attack on pyrite or molybdenite. At higher temperatures some reversion occurs

$$CuSO_4 + S_2 = CuS + 2SO_2 \quad \text{... (2)}$$
(covellite)

(This is not necessarily disastrous as the covellite can probably be floated from the product, as can pyrite and molybdenite).

Sulphur and sulphur dioxide are volatilised and the sulphates dissolved in water. Precious metals remain in the insoluble residue and the pregnant solution is purified of selenium, tellurium and silver by sparging in sulphur dioxide to precipitate them. Hydrogen cyanide and further SO_2 are then added whereupon cuprous cyanide precipitates rapidly and completely as creamy white, rapidly settling flocs, insoluble in acid.

$$2CuSO_4 + SO_2 + 2HCN + 2H_2O = 2CuCN + 3H_2SO_4 \quad \text{... (3)}$$

(Copper metal can alternatively be used as reductant in place of SO_2). This precipitation is very selective as nickel and zinc cyanides are soluble in acid. The sulphuric acid formed in 3 is concentrated for re-use.

The cuprous cyanide after washing is reduced with hydrogen at 325°C.

$$CuCN + \tfrac{1}{2}H_2 = Cu + HCN \quad \text{... (4)}$$

It can be seen that, provided sulphur dioxide is recovered from ferrous sulphate and re-converted to sulphuric acid along with SO_2 not sparged into the solution, the only sulphuric acid required will be make-up for minor losses, and all sulphide sulphur reports as elemental sulphur. Similarly the only HCN required is that corresponding to minor losses.

3. PILOT PLANT ENGINEERING

All unit processes except the hydrogen-reduction of cuprous cyanide (and the present method of sulphur scrubbing) are carried out inside an enclosed approximately 90 x 60 x 30 foot high building. Equipment is all new and well laid-out, though as Mr Cadwell pointed out it was only a pilot plant - the next stage from laboratory testing - and numerous changes were expected and required.

Sulphide concentrates are blended with conc. sulphuric acid (98%) and fed into a Pfaudler kettle where approximately 50% of reaction occurs.

Minus ½ in andesite porphyry pebbles are preheated to 400°C in a shaft preheater and fed with the slurry into a 4 ft diameter x 45 ft (approximate) mildsteel rotating kiln, with seals to prevent access of air, and externally heated by approximately 16 natural gas burners. The temperature is held at 250-280°C. Dry, coated pebbles, from which approximately 65% of sulphur has been vaporised in the kiln pass through another shaft similar to the pebble heater, in which a countercurrent flow of cool SO_2 volatilises the remaining sulphur and cools the pebbles. The pebbles pass over two washing screens on which they are washed, first with recycled solution, then with process water, and are then returned with less than 1% breakdown to the pebble preheater. Pyrex pebbles will probably be tried.

The gases from the kiln and sulphur vaporiser pass to two water-scrubbers in series where sulphur is removed as a yellow-green powder slurry. The sulphur assays 97-98% S and can be purified by melting and filtration. The sulphur dioxide is at present neutralised with ammonia but would commercially be converted to sulphuric acid for re-use.

The copper sulphate solution (60 g/l) is filtered to remove insoluble residue, then purified by sparging in sulphur dioxide to precipitate selenium, tellurium and silver. The solution then passes to a tank where an aqueous solution containing 25% HCN and some sulphur dioxide is added. The precipitated cuprous cyanide is removed in a Huttenwerk Sonthofen "FEST" continuous pressure filter, washed with copper sulphate solution to eliminate free cyanide, then twice with water, dried in a Wyssmont multihearth dryer to 3% moisture, and compacted into "apricot-stone" briquettes in Komarek-Greaves pocketed rolls. These briquettes pass out of the building to the reduction shaft, fitted with sophisticated controls and purges to prevent ingress of air, where they are reduced to metallic copper with hydrogen at 325°C. Further details of this stage were given later by Mr E.S. Roberts in New York. The reduced briquettes of the original size and shape but of very low density consist of 100 Å crystallites of copper. The gases are scrubbed with water and the 25% HCN solution resulting is piped inside the building to precipitate cuprous cyanide.

The copper briquettes can be sintered to give metallic pellets suitable for melting. Alternatively the cuprous cyanide can be pressed (or possibly extruded, though it lacks lubricity) into sections which can be reduced, rolled and drawn into wire without melting. Such wire has been produced assaying 11 ppm total impurities with a conductivity of 102.5. Lack of melting means that impurities only reduce conductivity in proportion to their volume percent.

Sulphur recovery is 98-99%. The copper-barren solution after cuprous cyanide precipitation is evaporated from 18% H_2SO_4 to nearly 40% H_2SO_4 in a multiple-effect evaporator, ferrous and zinc sulphates crystallized out, then further concentrated to 80% H_2SO_4 by submerged combustion. This acid is blended with oleum to give 98% acid for re-use. Mr Cadwell states that this recovery process gives cheaper acid than they could synthesize.

4. ECONOMICS

Mr Cadwell states that while an 80,000 tons per annum (tpa) conventional copper smelter and refinery, including gas recovery, would cost $US65 million under Arizona conditions, a comparable plant using the new process would cost $US41 million. While the cost of processing copper concentrates to wirebar conventionally is 7.75 cents per lb, the new process would cost 3.8 cents per lb assuming no credits for sulphur, iron etc. (One ton of sulphur is produced per ton of copper from chalcopyrite concentrates). Cement copper could be dissolved for 0.5 cents per lb and converted to metal for a further 1.6 cents per lb. He considers the minimum economic size of plant would probably be 10-15 tons copper per day (3,500-5,000 tpa). More detailed cost data were given later by Mr E. Roberts in New York.

Mr Cadwell states that HCN can be made from natural gas for 0.8 cents per lb by the Andrussow process, which yields a 26% solution in water.

5. FURTHER ACTION

Mr Cadwell is chief consultant to General Electric Corp. which is the third largest consumer of copper metal, and already manufactures in Australia. They would like to make 50,000 tpa of copper, and after a description of local production of copper ore Mr Cadwell is very interested in the possibility of establishing a custom treatment plant in South Australia with local participation.

Mr Cadwell also arranged for me to see Mr E. Roberts at Treadwell Corporation head office in New York, and to see a laboratory demonstration of the cuprous cyanide precipitation stage.

VISIT 2: SOULÉ STEEL COMPANY

Situated at
2201 East Carson Street,
CARSON CITY, (Nth Long Beach)

DATE: Tuesday 3 November 1970

MET: Mr Earl Neeb, Steelplant Manager
Mr Guy Haskins, Plant Engineer

1. INTRODUCTION

Soulé Steel produce 80,000 tons per annum (tpa) of 0.30%C re-inforcing bars ("re-bars") and 10,000 tpa fence posts using scrap steel at present costing about $US.35 per long ("gross") ton. Rebars are produced to A432 (60,000 lb yield). The re-bar is used in Soulé reinforcing fabrication plants on the same site and elsewhere. Mr Haskins said that a paper describing the plant was presented by Williamson at the Spring meeting of the 1968 or 1969 AISE Convention.

2. PLANT DESCRIPTION

A sketch of the plant is shown in Figure A2:1

The scrap seen on the heap consisted of plate scrap, bumpers, wheels, toys etc. They try to avoid auto scrap because of tramp elements (Cu, Ni, Cr, S) and also try to avoid machine-shop turnings because of bulk. They specify 5 ft maximum length. Scrap is loaded from the heap into old rail trucks and taken into the furnace bay for loading by overhead 7½ ton electromagnet crane into bottom-opening charging buckets. 3-5 charges lifted by the 10 ton charging crane fill a furnace.

The two furnaces, 25 ton capacity, 3-electrode, size 15 KK Whiting Corporation units are of 11 ft shell diameter and draw 6 MW at 240 V. A transformer station steps down the 66,000 V supply to 12,000 V which feeds the furnace primaries.

Two baghouses, stated to cost approximately $50,000 each (approx. $3 per square ft of bag area) handle 60,000 cfm from each furnace hood and 200,000 cfm from the building roof.

Furnace bottoms last four years, sidewalls are replaced (metal-cased "Interpace" bricks) in 12-16 hours once per month, and the arch is replaced twice per month using approximately 1/3 of amount of refractory required for the whole shell. Hot patching is done between heats. The plant operates around the clock 7 days per week with eight holidays per year and a mechanical maintenance shut-down of one shift per week.

A furnace heat cycle is approximately 2 hours consisting of charging, melting, refining, slagging, pouring, inspection and patching. The bottom-pour ladles into which the steel is poured hold more than 20 tons and last 200 heats before relining. Samples are taken during the heat for carbon analysis as well as estimation by grinding or breaking.

Each full ladle is lifted by the 40 ton ladle crane to its position above the 2 - nozzle tundish of the "Concast" continuous casting machine which casts the steel at 3050°F (100°F higher than for individual ingots) in an 80 degree arc of 14 ft radius. The red-hot continuous billet 5 by 5 inches then passes through a 3-roll straightener before being cut by a flying shear into four 12 ft lengths. A ladle takes 40 minutes to cast followed by 20 minutes before casting a ladle from the other furnace. The ingot cross-section (5 by 5 in.) was chosen such that a 12 ft by 6 inches long billet (weighing 1,100 lb) would yield an 80 ft length of 18-bar The cooled ingots are discharged and stacked in the yard in stacks identified by heat number. One stack of approximately fifty 4 ft ingots (10 tons) is carried by the forklift. Two stacks constitute one heat.

Billets are brought back by forklift to the natural-gas-fired reheat furnace into which they are pushed on water-cooled skids. The soaked billets then pass through the rolling mill which consists of two stages. The first (roughing) stage of diamond box-pass design consists of six stands and yields a 2 inch round which is transferred laterally to go through the finishing mill which consists of up to fourteen stands. One 4 ft billet yields 780 ft of $^3/_8$ inch bar which is cut into three 260 ft lengths which cool on the hot-bed before being cut into lengths and wired into bundles.

Small sections are produced in 2½ ton bundles, large in 5 ton bundles which are loaded by travelling crane into warehouse bins or on to semi-trailers.

3. LABOUR

The total plant staff (all shift-plus day-workers) (excluding fabrication) consists of approximately 150. A furnace shift crew consists of 1 melter and first and second helpers, and a ladleman and helper and 2 general helpers.

4. ANALYTICAL CONTROL

The laboratory is sited alongside the furnaces and has LECO Carbon and Sulphur Analysers and a Baird Atomic Spectromet. A preliminary carbon assay made during a melt requires 5-8 minutes. Average figures copied from an assay sheet were:

	C	0.30 %
	Mn	0.4 %
	P	0.15 %
	S	0.04 %
	Si	0.1 %
	Mo	0.03 %
	Cr	0.2 %
	Ni	0.15 %
	Cu	0.25 %
Final	C	0.3 %

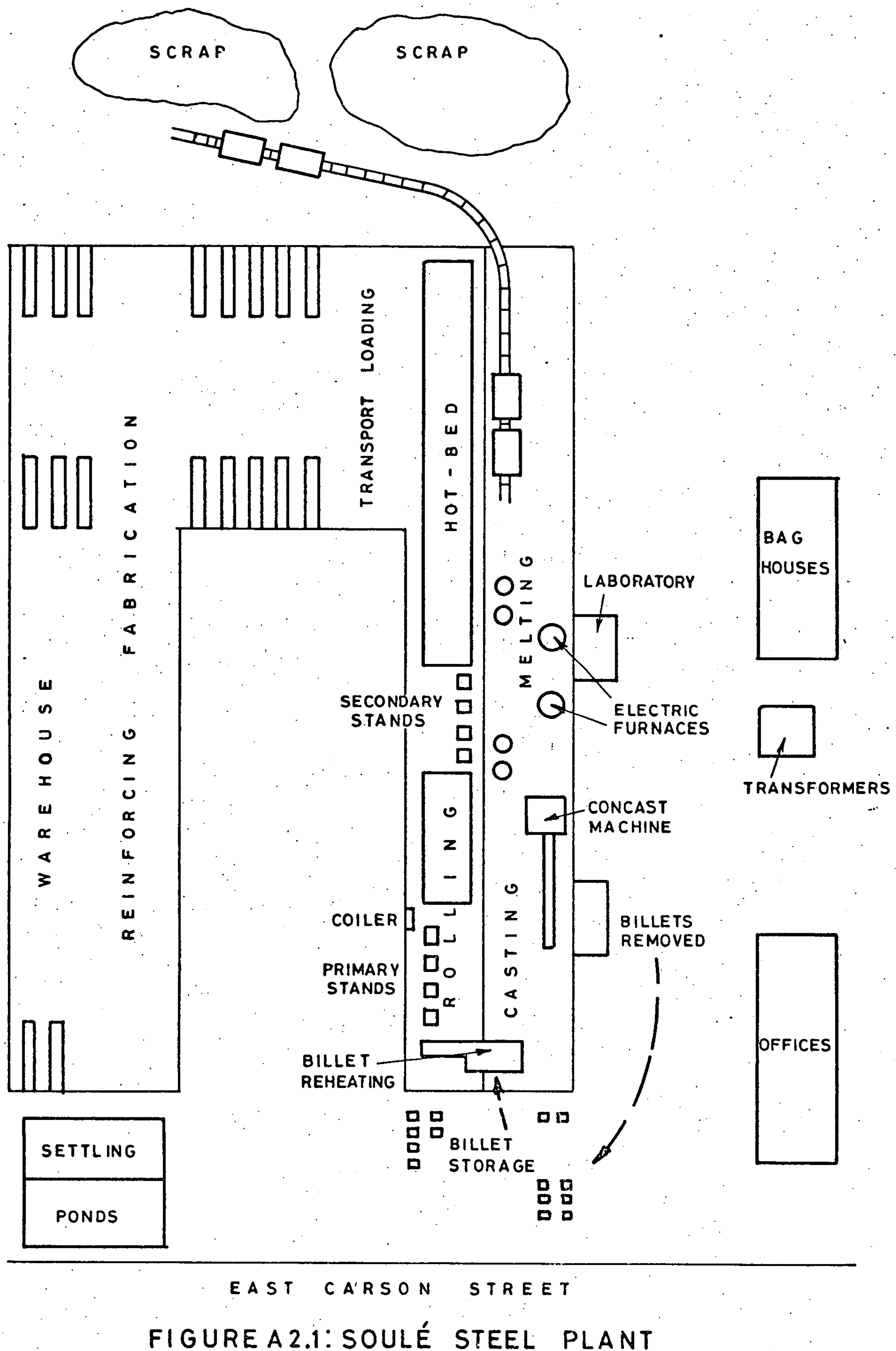


FIGURE A2.1: SOULÉ STEEL PLANT
(approx. 1 inch = 100 ft)

VISIT 3: AMERON STEEL PRODUCING DIVISION
(formerly Etiwanda Steel Co)

Situated at
12459 Arrow Highway,
ETIWANDA, California 91739

DATE: Tuesday 3 November 1970

MET: Mr Richard Pasma,
Manager - Industrial Engineering

1. INTRODUCTION

This plant produces rebars, plain rounds, and coiled reinforcing material for concrete pipes. The area is relatively poor for scrap, being some 50 miles east of Los Angeles in a grape-growing area. Current scrap prices are shown in Table A3.1, and the company uses No. 1 heavy melt grade including auto bodies etc which inevitably contains tramp elements. They have two 10 ton Whiting furnaces from which they tap 13 ton heats, and are adding a 15 ton Lectromelt furnace which will produce 20 ton heats. Capacity will be almost 100,000 tons per year. It is understood that the company does not attempt to produce to merchantable quality dimensional tolerances. Rod is produced for a nearby wire mill, but they do not produce prestressing wire. Carbon content varies with the intended final size of bar and hence cold work - large bars are made from up to 0.60%C, wire from steel containing 0.08%C.

2. DESCRIPTION OF PLANT

The plant layout is sketched in Figure A3.1.

Product storage is outside due to the dry climate, otherwise the general arrangement of a melting aisle and rolling aisle side-by-side resembles that at Soulè Steel. The furnaces draw 6.5 MW and melt cycle is 2 hours. Furnace relining is done after 300 heats. The continuous caster was built by United Engineering and Foundry Co., Pittsburgh but the ladle turntable was engineered and patented by Etiwanda. They cast a ladle in 20 minutes then there is a wait of 40 minutes for a ladle from the other furnace. 4½ x 4½ in billets are produced and stacked in the open by overhead crane which also returns them to the reheat furnace in which, as it is rather small they have found it necessary to burn oil as well as gas to obtain sufficient heat. The chain-drive mill was designed and engineered by Etiwanda, and Mr Pasma suggested that blueprints would be available for sale if the project was proceeded with in South Australia. Roughing and finishing stages of rolling are in-line, with intermediate cropping of the leading end of the roughed bar, as at Soulè Steel. Four stands reduce the billet to 2¼ inches diameter round and a further six stands were being used to yield 7-bar ($^{7}/_{8}$ inch diameter). The mill produces from 7/32 rod to 11-bar ($1^{3}/_{8}$inch diameter). The hot bed for cooling is 182 ft long. It appears that most of the rebars are coiled and coils butt-welded together to give 2,000-3,000 lb coils handled by forklift. Tramp elements have to be watched as they affect welding and would be very important if making welded tube in South Australian Copper for example needs to be less than 0.35% but other elements also interfere. Baghouses from Wheelabrator or Pangborn are used and Mr Pasma said that one \$180,000 unit would be required per furnace, and regulations were tightening.

3. LABOUR

Including supervision, clerical and maintenance 275 people are employed for melting and rolling (not including wire mill). A shift crew in the melt shop including continuous casting (which saves considerable labour over individual ingots) is about 20, with 12 in the rolling mill, including inspectors and laboratory staff.

4. OTHER COSTS

Ferrosilicon and manganosilicon additions are made, but ferromanganese only if the heat is low in manganese. Carbon content can be raised in the furnace by dipping the electrodes, which cost 30 cents per lb or in the ladle using carbon raisers.

5. VALUE OF PRODUCT

"Iron Age" gives current prices for rebars - in the Los Angeles area the base price is \$6.25 per hundredweight, with surcharges on (a) diameter (\$0.50 per cwt. for $^{3}/_{4}$ inch, \$1.40 per cwt for 3-bar $^{3}/_{8}$ inch) (b) length (most clients buy 20, 40, 60 ft which incur no surcharge) (c) quantity (no surcharge for lots above 20 tons).

TABLE A3:1

Ferrous Scrap Prices, Los Angeles, effective October 26 1970 (ex "Iron Age"; Prices obtained in trade based on representative tonnages per gross ton (2240 lb) delivered to customer)

	$US
No.1 Heavy melting	41.00
No.2 Heavy melting	37.00
No.1 Dealer Bundles	35.00
No.2 Dealer Bundles	31.00
Machine Shop Turnings	14.00
Shovelling Turnings	18.00
Cast Iron Borings	14.00
Elec. furnace 1 ft and under (foundry)	52.00
No.1 Cupola Cast	45.00

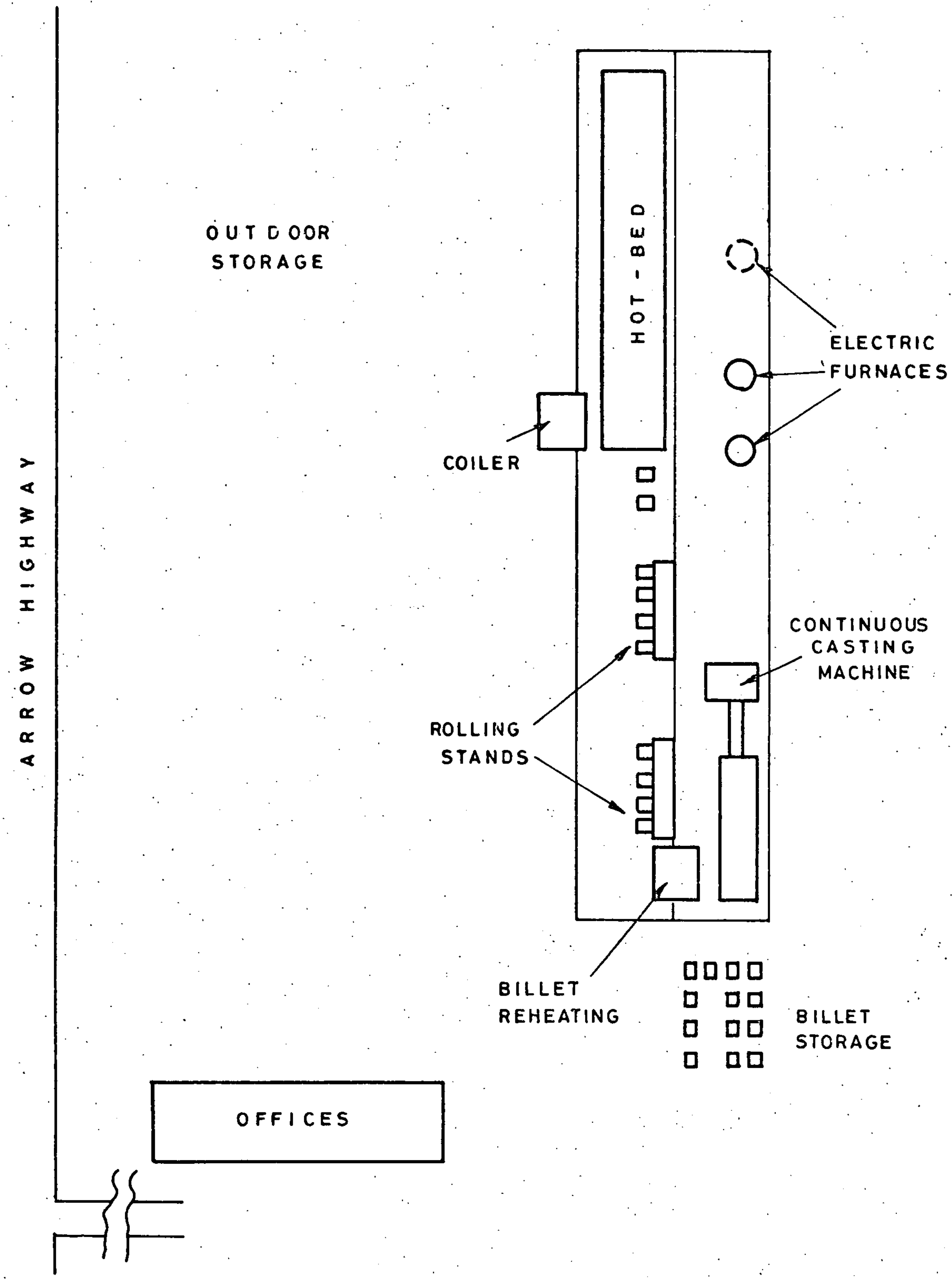


FIGURE A3.1: AMERON PLANT, ETIWANDA
(approx. 1 inch = 100 ft)

VISIT 4: SHERRITT GORDON MINES LTD

Situated at
FORT SASKATCHEWAN
ALBERTA, CANADA
(17 miles from Edmonton)

DATE: Friday 6 November 1970

MET: Dr David J.I. Evans, Director, Res. & Dev. Division
Mr Herb Veltman, Assistant Director, Res. & Dev. Division
Mr Wasil Kunda, Head, Chemical Metallurgy Res. Department
and a number of other staff members.

1. COPPER POWDER PRODUCTION

While Sherritt Gordon supplied know-how for the Bagdad plant in which copper is hydrogen-precipitated from acid solution, and for the Kansas City plant using ammonia-ammonium carbonate solution, and did pilot the use of ammonia-ammonium sulphate solution for Marinduque Iron Mines, they do not currently produce copper powder. Messrs Veltman and Kunda supplied preprints of two recent papers relating to the ammine carbonate system and I spent several hours talking to research staff who had worked on copper precipitation, and then examining methods of metal powder characterisation.

1.1 Hydrogen Precipitation

Dr Evans said that hydrogen precipitation could produce any type of copper powder required, and showed numerous samples ranging in coarseness and apparent density. Precipitation from acid solution is more difficult as the solution becomes increasingly acidic and corrosion is a problem. Stainless steel can be used for ammoniacal solutions but titanium is required for acid solutions.

The Samin concentration of 20-40 g/l copper was considered too dilute, and 120-140 g/l is preferable. Mr Kunda has some idea on how to increase the concentration, or alternatively, if the solution is sufficiently pure, precipitate copper oxide as currently planned for Burra and reduce with hydrogen. Conditions for hydrogen precipitation are 350°-375°F, 500 psi partial pressure of hydrogen. Sherritt Gordon have tried hydrothermal reduction of copper oxide without success, but one of the research staff said that the green basic carbonate precipitated below 200°C can be reduced hydrothermally with hydrogen to copper. Typical concentrations for hydrogen precipitation are NH_3 100-130, CO_2 80-90 gpl. EMA-11 polyacrylate gives quicker reduction and "seed" copper powder can be used. Re-dissolution tends to occur as soon as the hydrogen atmosphere is replaced by air.

1.2 Physical Properties of Powder

The properties determined for nickel and cobalt powders (and hence presumably for copper powder) are Flow Number (time in seconds for 50 g powder to flow through a Hall Flowmeter made by Metal Disintegrating Co. New Jersey), Apparent Density, and Tapped Density (measured by weighing a receptacle placed below the Hall Flowmeter) and the Fisher No. (measured by means of a Fisher Sub-Sieve Sizer). Sizing is determined on 1 g samples by Buckbee-Mears (St. Paul, Minnesota) micromesh sieves down to 5 microns. These are cleaned ultrasonically using a unit supplied by Dynasonics Corp (Pipestone, Minnesota). Sherritt used Dow Chlorthene NU costing about $4 per gallon, at 70-80°C, and state that this fluid is much cheaper than that supplied by Dynasonics.

2. VISIT TO NICKEL-COBALT PLANT

Mr Veltman supplied an Annual Report, a company brochure "Enriching the Earth's Riches", an illustrated flowsheet, and an "Outline of Plant Operations". Extracts from the latter two publications are appended. Mr Bauke Weizenbach then showed me over both the nickel plant (50 tons per day) and the cobalt plant (1 ton per day).

2.1 Nickel Plant

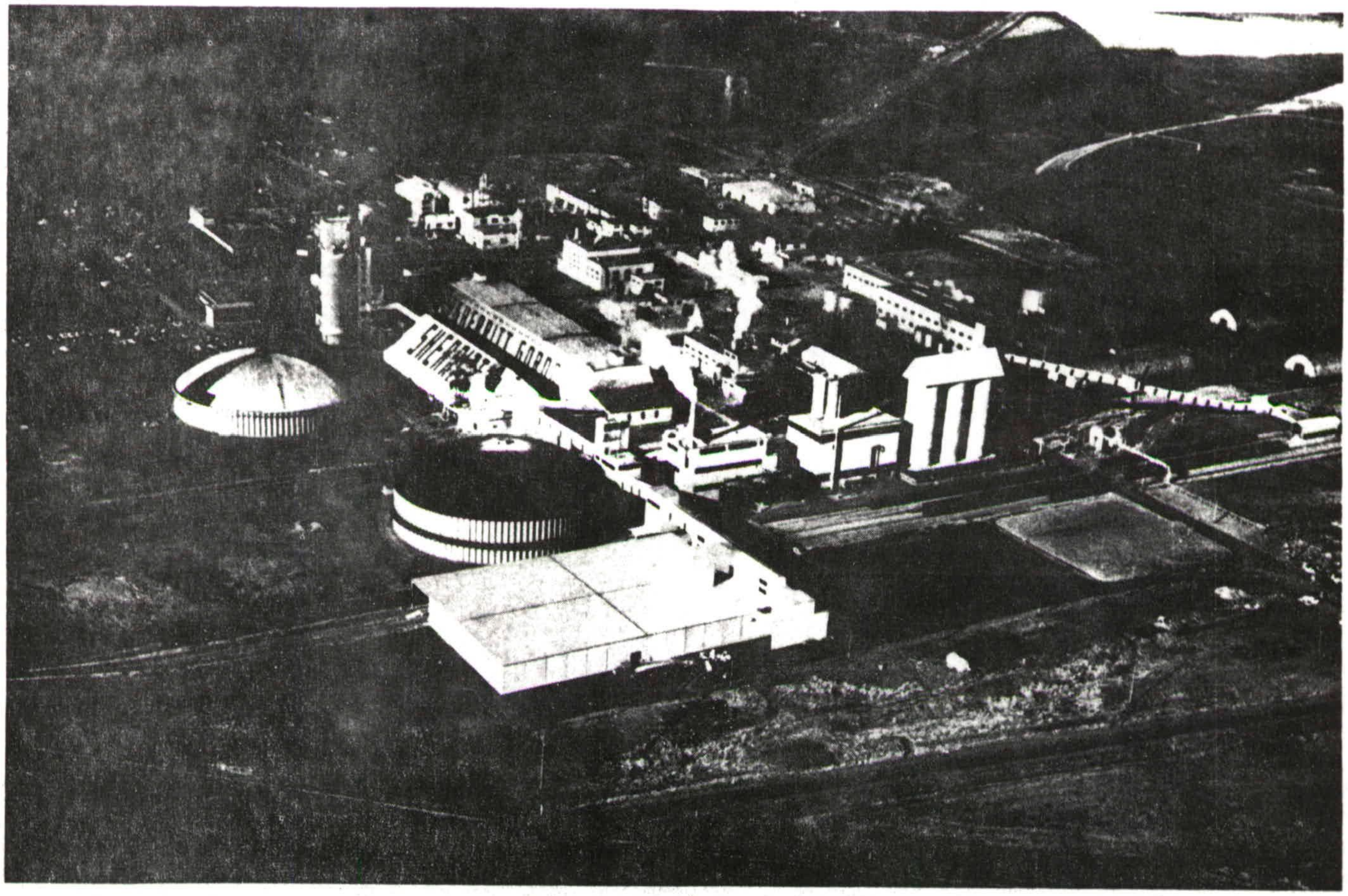
Nickel sulphide concentrate from Sherritt's own operations and from Kambalda is leached in one pair of Primary Leach autoclaves in parallel, followed by three pairs of Final Leach autoclaves. The two stages are countercurrent, the products being final tailing containing 0.8% Ni and nickel pregnant solution. Subsequent operations are described in the printed literature appended.

2.2 Cobalt Plant

This was formerly the pilot-plant operated on the Marinduque copper-zinc project. Details of the process, which is carried out in essentially pilot-scale manner, are given in the printed literature appended.

3. FURTHER APPRAISAL

If Sherritt-Gordon were asked to further appraise the production of copper-powder in South Australia they would carry out a "desk-study" (involving no laboratory work) for $Can5,000, corresponding to an engineer's salary for the time required, plus 100% overhead charges. They would send us a questionnaire to obtain details of conditions to be assumed.



OUTLINE OF PLANT OPERATIONS

SHERRITT GORDON MINES LIMITED

Fort Saskatchewan, Alberta, Canada

SHERRITT GORDON MINES LIMITED
Fort Saskatchewan, Alberta

COMPANY BACKGROUND

Sherritt Gordon has a history going back to 1927, as a copper-zinc mine at Sherridon, Manitoba. The mine commenced production in April 1931. Low copper prices forced the shut-down of the mine in June 1932. Production started up again in August 1937 and continued until September 1951 with the exhaustion of the ore-body. During this period a total of 8,500,000 tons of ore were mined.

Sherritt was and still is very active in prospecting, and in September 1941 located a likely prospect at Lynn Lake, Manitoba. To prove up this prospect took considerable time and was slowed down because of World War II. By the end of 1943 there was definite evidence of nickel-copper mineralization. No further work was done until April 1945. The ground was staked and drilling commenced in July. The finding of the high grade "EL" ore-body in January 1947 was the turning point in the development of the mine.

Much work was required to prove up sufficient ore to make a mine: geophysical surveying, prospecting, drilling, and development. While this was going on work was proceeding, to develop a process for treatment of the ore, in planning facilities and procedures for the mine and a refinery, and in arranging financing for the overall complex. Test work on the Lynn Lake ore started in 1946, a pilot plant was operated at Sherridon in 1947, in 1948 a pilot plant was set up at Lynn Lake and operated in 1949, 1950 and 1952. In 1947 Professor Frank A. Forward, then Head, Department of Mining and Metallurgy at the University of British Columbia, was retained to work on Lynn Lake concentrate to find a method of recovering metals at a lower operating and capital cost than the conventional smelting and refining method. Professor Forward came up with an ammonia leaching process which looked promising. From this Sherritt research men duplicated the results in the laboratory at Sherridon. Following this a pilot plant was built in Ottawa. The results of the work at this pilot plant from 1949 to 1952 provided the design data and trained operating personnel for the commercial plant built at Fort Saskatchewan.

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It was just under nine years from the discovery of the first ore-body at Lynn Lake on September 22, 1945 until the first shipment of nickel left Fort Saskatchewan on August 7, 1954.

FORT SASKATCHEWAN

The Town of Fort Saskatchewan is located 17 miles northeast of Edmonton and 800 miles from the mine at Lynn Lake. Prior to the coming of Sherritt, Fort Saskatchewan was a small rural town with a population of 1,200. It is now a thriving rural-industrial town of over 4,000 people.

The selection of Fort Saskatchewan as the plant site was influenced by several important factors. First was the availability of an abundant supply of natural gas, at a reasonable price, which could be used as fuel and principally as a raw material for making ammonia. The site was on the Canadian National Railway which was important because the C.N.R. had agreed to build the $15,000,000 connecting link between Sherridon and Lynn Lake. Also from the North Saskatchewan River there was plenty of water for cooling and process use. The general locality was attractive and within easy reach of a large centre of population. It was in a good farming area and enroute to the Northwestern United States, and the Pacific Coast, making it ideally located for marketing of the Company's large tonnage of chemical fertilizer, and for purchase of custom material for treatment.

PRODUCTION DEPARTMENT - METALS

The process developed by Sherritt for the Fort Saskatchewan plant differs from conventional smelting and electrolytic refining in that no melting or plating operations are involved. It also differs from all other hydrometallurgical processes in three ways: by the reagents used, by the sequence of processing steps, and by conditions of operation. More important, the nickel produced is in the highly desirable form of pure metallic powder, instead of an oxide or ferro-nickel. The powder is ideally suited for many applications, and is pressed easily into briquettes or strip.

Sherritt's hydrometallurgical refining process is basically simple: dissolve the nickel, copper and cobalt contained in the concentrate in a

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solution of ammonium sulphate and ammonia; then recover the metals from the solution, one at a time.

PROCESS - METALS PLANT

1) LEACHING - The first stage in the refining process is the leaching of the nickel, copper and cobalt from the concentrate by means of an ammonia solution. The leaching operation is basically a continuous, counter-current process performed in autoclaves - totally enclosed agitator-equipped pressurized vessels. Optimum operating pressures and temperatures vary from 100 to 110 psig and from 170° to 180° F. A counter-current leach system is used to provide maximum metal recovery.

In the system, fresh incoming concentrate is first treated by leach liquor which has been produced in the second stage of the leach process. This intermediate liquor takes up the most easily leached portion of the concentrate, yielding a full-strength solution of dissolved metals. The partially leached concentrate is then treated in a second set of autoclaves with fresh leach liquor high in ammonia.

The solution containing the dissolved metals is then separated from the leach residue (iron oxide, insolubles, etc.) by means of thickeners. One phase of this liquid-solid separation occurs between the leach stages, and the other after the final leach treatment. Following the final leach treatment, the residue is carefully washed by repulping and filtering in order to remove all soluble nickel before it is pumped to the residue pond.

A certain amount of sulphur is also leached with the metals. Most of the sulphur forms ammonium sulphate, while a small portion becomes unsaturated sulphur compounds such as ammonium thiosulphate, ammonium thionate, and others. These unsaturated compounds play a very important part later on in the process.

2) COPPER SEPARATION - The pregnant solution containing copper, nickel, cobalt, ammonium sulphate, ammonium sulphamate, and the unsaturated sulphur compounds, together with some free ammonia, is then pumped to the first stage of the copper separation circuit. Here the solution is heated to the boiling point in an enclosed five-stage boiler unit. The boiling sequence produces

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two results: first, most of the uncombined ammonia in the solution is boiled off and returned to the leach circuit to be used again; and secondly, a reaction between the unsaturated sulphur compounds and the copper in solution precipitates the copper in the form of a black sludge of copper sulphide.

Most of the copper sulphide is then removed by passing the solution through a filter press. The traces of copper remaining after this step are stripped from the solution by passing hydrogen sulphide through it, thereby bringing the copper content down to practically zero.

The copper sulphide produced in the copper boil is suitable for shipping directly to a custom smelter for recovery of copper. However, the copper sulphide obtained by using hydrogen sulphide contains considerable amounts of nickel which is returned to the leach circuit for re-dissolving. This portion represents only about 10% of the total copper processed.

3) NICKEL RECOVERY - The solution leaving the copper separation circuit contains nickel, cobalt, ammonium sulphate, ammonium sulphamate, and a very small amount of unsaturated sulphur compounds. Before the nickel and cobalt can be recovered, it is necessary to decompose the ammonium sulphamate and the unsaturated compounds contained in the solution. This stage is known as "oxydrolysis" - the oxidation of the unsaturated sulphur and the hydrolysis of sulphamate to sulphate. Heating under pressure in the presence of air converts the sulphamate and the unsaturated compounds into more ammonium sulphate.

The solution is now ready for the recovery of pure nickel. To this point the process has been continuous, but from this point on the recovery of nickel is on a batch basis.

The solution is first fed into an autoclave containing a small amount of fine nickel powder. When the autoclave is filled, the agitators bring the nickel powder into suspension. Hydrogen is then passed into the vessel up to a total pressure of about 500 psi. This precipitates the nickel in solution on to the nickel powder particles originally in the autoclave in the form of pure nickel metal. The process continues until almost all the

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nickel has been precipitated. Very little cobalt will precipitate by proper control of chemical conditions.

After most of the nickel has been removed, the agitators are stopped, the nickel powder is allowed to settle, the depleted solution drawn off, and fresh solution added. This operation is repeated some 50 times until the nickel particles become so heavy that it is difficult to keep them in proper suspension.

When this occurs, the solution is drawn off with the agitators running, thereby removing the precipitated nickel with the solution.

The entire contents of the autoclave from the completed cycle are then discharged to cone-bottomed flash tanks where the mother liquor overflows to a storage tank and the nickel metal settles to the bottom of the cone. The nickel, now in the form of a slurry, is washed, dried, and packaged as powder; or pressed into briquettes, sintered, and packaged for shipment.

A completely new cycle of reduction is then begun with a nucleation step which produces the very fine nickel seed particles.

4) COBALT RECOVERY - The solution leaving the nickel reduction circuit contains a very small amount of nickel, cobalt, and a high concentration of ammonium sulphate. The nickel and cobalt are removed from the solution together by treatment with hydrogen sulphide, which yields a mixed precipitate of nickel and cobalt sulphides. The sulphide is filtered off and processed elsewhere in the plant for the recovery of pure cobalt metal. The recovery technique is similar to that used in the reduction of nickel.

5) AMMONIUM SULPHATE RECOVERY - The remaining solution contains only ammonium sulphate which is recovered by evaporation. This sulphate, in the form of crystals, is sold as nitrogen fertilizer. The entire 21% nitrogen content of the crystals is available in a water soluble form. The product also contains 24% sulphur, another important plant food.

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6) ROLLING MILL - Sherritt Research and Development Division at Fort Saskatchewan developed a process for converting nickel and cobalt powder into strip. A Rolling Mill has been established at Fort Saskatchewan using this powder metallurgy process.

The main product of the mill is not strip but coinage blanks which are punched from pure nickel strip. At the present time we are supplying nickel blanks to the Royal Canadian Mint for Canadian 5 cent, 10 cent and 25 cent coins. Sherritt has also supplied coinage blanks to South Africa, the Netherlands and Brazil. The recent increase in the price of silver has greatly stimulated the demand for pure nickel coinage blanks. Several countries have already converted from silver to nickel coinage, others are planning to do so.

In 1966 Sherritt put into operation a minting press to produce finished coins and medallions. The first production from the Sherritt Mint was a pure nickel medallion in three sizes which portrayed the entry into Alberta of one of the early fur traders, Anthony Henday. A second medallion featuring Henry Kelsey has also been minted by Sherritt. Many custom medallions were struck during the 1967 Centennial Year. The Sherritt Mint is now busy with orders for pure nickel coins from several foreign countries.

The operation of the Rolling Mill has provided a sizeable market for nickel and cobalt powder and has also provided a means of profitably upgrading our products.

7) GRINDINGS PLANT - In addition to treating the nickel concentrate from the mine at Lynn Lake, Sherritt purchases or acts as a custom refinery for nickel cobalt bearing material from others. Most of these other feeds are handled in the normal operation. Some materials require special treatment - to meet this requirement, in August 1965, the Grindings Plant was put into operation.

One of the main materials treated is "grindings", hence the name Grindings Plant. Grindings are a by-product generated in the production of high temperature alloy components for the jet propulsion industry. These

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grindings are purchased or treated on toll basis and from them substantial amounts of nickel and cobalt are recovered.

The development of the process used in the Grindings Plant is another example of the work done by the Research and Development Division and the Engineering Department. It has resulted in a substantial increase in cobalt production.

8) SPECIAL POWDERS - The Sherritt refining process produces nickel and cobalt powders of very high purity. Using techniques developed by the Research and Development Division, various grades of these powders are being produced for specific applications in industry.

Sherritt's refining process was modified to allow production of COMPOSITE POWDERS. The Company's patented process deposits nickel and/or cobalt on core materials of finely divided metal and non-metal particles. Composition of these particles can be controlled to produce a wide range of materials having unique properties. Such composite powders are applied by metal spraying or sintering techniques to solve critical materials problems in aerospace, appliance and other industrial markets.

Sherritt metal production of nickel, cobalt and copper, from Fort Saskatchewan is at the rate of 50 tons per day.

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PRODUCTION DEPARTMENT - FERTILIZERS

Sherritt entered the fertilizer manufacture as a planned result of the metals plant process. An anhydrous ammonia plant was built to produce ammonia which was required as the leaching agent for the recovery of metals. The plant produced more ammonia than was required by the metals plant. This surplus ammonia was available for sale as a fertilizer. Anhydrous ammonia is the highest nitrogen content fertilizer material available - 82% N - and may be applied directly or as an aqua solution. In addition to producing ammonia, hydrogen and carbon dioxide are also produced for use in various parts of the plant.

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To meet the requirements for the metals plant and fertilizer plant, the ammonia plant was increased in size from 75 tons per day to 150 tons per day and then to 250 tons per day. These expansions allowed more ammonia to be sold as a fertilizer, to use more for increasing the manufacture of ammonium sulphate, and to enter the manufacture of urea.

The next step in the expansion of the fertilizer plant was the building of a urea plant. This was started in 1961, with the first shipments of urea being made in late 1962. Urea is the highest nitrogen content dry fertilizer available, 46% nitrogen. The urea produced by Sherritt is a prilled form containing less than .25% biuret. Biuret is an impurity which is harmful to plants when the urea is used as a foliar application. Sherritt is the only producer in the world of this high quality urea. It is also made in a feed grade, as a supplement to cattle feed, and in a technical grade for various industrial uses. Urea is made by combining anhydrous ammonia and carbon dioxide, both of which Sherritt had available in good supply.

With the increasing demand for fertilizers in Canada, United States of America, and other parts of the world, it was evident that Sherritt was in an excellent position to expand its fertilizer production. In 1964 work was started on increasing the capacity to produce ammonia and urea and to add phosphate fertilizers to the Sherritt line. Phosphate fertilizers are produced from ammonia, sulphuric acid and phosphate rock - all of which Sherritt had or could secure at a reasonable price.

The first Sherritt phosphate fertilizers were shipped in August 1965.

The expansion into phosphate fertilizers was a logical move for Sherritt, as throughout Western Canada the great need is for phosphate. This is also true for a large part of the American market area which Sherritt can serve economically. By being in a position to serve these markets Sherritt is in a better position to market its straight nitrogen fertilizers by making up mixed orders. The more fertilizer that can be marketed close to home means less that has to be marketed in more distant markets, with a subsequent reduction in freight charges. Also in the manufacture of phosphate fertilizer, the nitrogen content is supplied by Sherritt ammonium sulphate, urea, and

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anhydrous ammonia. Market surveys have indicated an increasing demand throughout Western Canada for fertilizers, particularly phosphate fertilizers.

The sulphuric acid is purchased from a neighboring plant, Inland Chemicals, who expanded their plant from 100 tons per day to 600 tons per day, to meet Sherritt's needs. This is delivered to Sherritt by pipeline connecting the two plants. The phosphate rock is brought in from Florida, by boat to Vancouver where bulk handling facilities have been set up for transferring the rock from the boat to rail cars. From Vancouver the rock is brought in by the Canadian National Railways. About 150,000 tons of the phosphate rock per year is required.

The 1964-65 expansion to the fertilizer plant with the expanded utilities and other services cost approximately 25 million dollars. The size of this expansion can best be realized by comparing it to the original plant which at the end of 1954 was 24 million dollars. By the end of 1966 the cost of the plant will be in excess of 60 million dollars: a plant investment of $65,000 - $70,000 per employee.

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AMMONIA PRODUCTION

The Ammonia process consists of high pressure reforming and purification of the gases to produce a synthesis gas that can be converted to anhydrous ammonia.

This is done by passage of a steam-gas mixture over heated catalyst contained in tubes; the catalyst being heated indirectly from outside the tubes by burning natural gas with preheated air in the primary reformer furnace, thereby transforming natural gas and steam into hydrogen, carbon monoxide and carbon dioxide. Air is then added to the reformed gas in such an amount that the nitrogen in the final gas will be in the correct proportion for the production of ammonia.

The reformed gas then passes through two beds of shift catalyst, where essentially all of the carbon monoxide is converted to carbon dioxide. The gas is then contacted with monethanolamine for removal of carbon dioxide. The final traces of carbon oxides and oxygen are removed by methanation catalyst.

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The purified mixture of hydrogen and nitrogen at a pressure of approximately 5,200 psi then reacts in the presence of a catalyst to produce gaseous ammonia. The gaseous ammonia is liquified by cooling and sent to the weigh and storage area.

The ammonia is stored as a liquid in Hortonspheres and in storage tanks for distribution to the Metals and Fertilizers plants and to fertilizer markets.

Hydrogen is produced separately in the ammonia plant also by catalytic steam reforming of natural gas. The purified hydrogen gas is used mainly for the precipitation of nickel and cobalt in metals recovery operations.

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AMMONIUM SULPHATE PRODUCTION

The ammonium sulphate plant was constructed to process an ammonium sulphate solution which is produced as a by-product in the metals recovery process.

The ammonium sulphate plant consists of two similar units of triple effect evaporators. In these the solution is concentrated, supersaturated and crystallized. The crystal slurry, which is drawn off by pump, is then centrifuged and dried. The dried crystals are then screened, with the product being treated with an anti-caking agent and stored for distribution.

The major use of ammonium sulphate is as a fertilizer. It is an extremely good fertilizer for use in sulphur deficient areas, because it contains 24% sulphur. As a fertilizer it may be applied directly as 21-0-0 or be used as a blending component with urea and ammonium phosphate fertilizers.

Ammonium sulphate fines are used as a feed source in the manufacture of 16-20-0 and they may also be used in the metals leach process.

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UREA PRODUCTION

Urea was the first organic compound to be produced synthetically. In its initial stages of development considerable engineering ingenuity was required to overcome such process difficulties as corrosion, methods of recycle and recovery of gases.

The major use of Urea is as a fertilizer, but other uses are as animal feed supplements and in the manufacturing of plastics and adhesives. The dual use of Urea as a fertilizer and animal feed supplement has made it an exceedingly popular product in agriculture.

The advantages of using Urea as a fertilizer are its high nitrogen analysis (46%), the fact that it is non-explosive, is very soluble, is resistant to leaching and is non-corrosive to fertilizer applicators.

The main features of Sherritt Gordon's Urea plant are the total recycle of unreacted CO_2 and NH_3, use of oxygen in air as a corrosion inhibitor and the extremely high purity of product. A fertilizer with a biuret contamination of less than 0.25% is produced.

Biuret is a constituent that is formed during the manufacturing process when too long a residence time at elevated temperatures is spent in a vessel. It has proved to be harmful to certain plants and crops (it acts as a weed killer), but has been found to be a protein builder in cattle.

The Urea production process is essentially the reaction of liquid ammonia, gaseous carbon dioxide and recycle ammonium carbamate at elevated temperatures and pressure inside a reactor. The reaction takes place at about 2,800 psig and is de-pressurized in two stages in order to recover unreacted gases. The flashed off ammonia, water vapor and carbon dioxide are collected in a wash column, where they react to form ammonium carbamate. This is then recycled back to the reactor as a liquid.

The solution remaining (approximately 75% Urea in water) is stored in a large tank from where it enters the crystallization section. Subsequent centrifuging and drying will produce a dry crystal that is remelted and then

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prilled. The prilled Urea enters a coating drum where an anti-caking agent is added prior to storage.

BASIC REACTIONS

1. $2\ NH_3 + CO_2 \longrightarrow NH_4CO_2NH_2$

Ammonia   Carbon Dioxide   Ammonium Carbamate

2. $NH_4CO_2NH_2 \longrightarrow NH_2CONH_2 + H_2O$

Ammonium Carbamate   Urea   Water

3. $2\ NH_2CONH_2 \longrightarrow NH_2CONHCONH_2 + NH_3$

Urea   Biuret   Ammonia

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PHOSPHORIC ACID PRODUCTION

The phosphate rock used by Sherritt Gordon is mined in Florida and shipped through the Panama Canal to Vancouver, B. C. where it is then loaded in closed hopper cars and sent by rail to the plant site at Fort Saskatchewan. Here it is unloaded at a rate of 8 tons per minute into three rock silos, each having a storage capacity of 6,000 tons.

Normal Florida phosphate rock is considered a fluorapatite and contains many impurities. Its main component is Bone Phosphate of Lime ($Ca_3(PO_4)_2$) which is the main inorganic constituent of bones.

The key to successful phosphoric acid plant operation is the attack system where the objectives are to obtain maximum conversion to phosphoric acid and to produce an easily filterable and washable gypsum crystal. To achieve these objectives, the phosphate rock is ground by a ball mill into a powdered form. It is then added with a small excess of sulphuric acid into the attack tank, where conditions are controlled to obtain maximum conversion and maximum gypsum crystal growth.

PRINCIPLE REACTION

$$Ca_3(PO_4)_2 + 3H_2SO_4 + 6H_2O \longrightarrow 3\ CaSO_4 \cdot 2H_2O + 2H_3PO_4$$

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After a retention time of approximately 8 hours in the attack tank, the slurry is pumped to a pan filter where the 31% P_2O_5 acid is removed first and stored. The remaining acid is separated from the gypsum cake by a three stage counter-current wash stream. After the final wash the gypsum cake is discharged and pumped out to the gypsum pond.

The stored 31% P_2O_5 acid can be used in the granulation plant for specific formulations or it can be concentrated to 40% P_2O_5 acid by evaporation and stored for use in the granulation plant.

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GRANULATION PLANT

The granulation plant is capable of producing four types of granular fertilizer. They are 11-48-0, 23-23-0, 11-55-0 and 16-20-0.

The basic granulation process consists of producing slurry of correct chemical and solids composition in the saturator and bringing it into contact with dry recycled granules inside the blunger. Here, through vigorous mixing, each granule is enlarged with a coating of slurry. Subsequent drying and screening will result in a low moisture fertilizer granule. All undersized material will be recycled to the blunger for further coating and product obtained is transferred out to bulk storage from where it is shipped to customers.

BASIC SATURATOR REACTIONS

1. $H_3PO_4 + NH_3 \longrightarrow (NH_4)H_2PO_4$

   Phosphoric Acid — Mono-ammonium phosphate

2. $H_3PO_4 + 2NH_3 \longrightarrow (NH_4)_2HPO_4$

   Phosphoric Acid — Diammonium Phosphate

3. $H_2SO_4 + NH_3 \longrightarrow (NH_4)_2SO_4$

   Sulphuric Acid — Ammonium Sulphate

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RESEARCH AND DEVELOPMENT DIVISION

The Sherritt nickel and cobalt refining processes are the direct result of Sherritt's own research and development effort which expanded the scope of the company from mining to metal refining and manufacturing. The success of Sherritt's unique hydrometallurgical processes firmly established "research" in its organization. Today about 80 professional and technical people are employed in the new Research Centre of the Research and Development Division located in Fort Saskatchewan.

The work carried out by our research staff concentrates essentially on four areas: hydrometallurgical process development, new product development, technical assistance to the production plants and agronomic research.

Research studies in the field of pressure hydrometallurgy resulted in the development of numerous processes for the recovery of not only nickel and cobalt, but also copper, zinc, lead and other metals as well as sulphur and iron from a wide range of metal-bearing concentrates, mattes, industrial residues and secondary materials. While some of these processes have already been applied commercially, others are under intensive development for either Sherritt's own use or for adaptation by other companies under suitable licensing agreements. Of considerable future interest are the processes developed for the recovery of nickel and cobalt from laterite ores. Processes for the recovery of non-ferrous metals and elemental sulphur from sulphide concentrates also offer interesting new metallurgical possibilities.

The new product development work led to the production of pure nickel and cobalt strip by the direct rolling of metal powders. This, in turn, provided the means of further vertical integration into the production of coinage blanks and, finally, to the minting of coins for a number of foreign governments. Extensive investigation of the mechanics of hydrogen reduction of nickel and cobalt from solution brought about the development of a variety of monocomponent and composite powders presently in commercial production at Fort Saskatchewan. The combined knowledge of pressure hydrometallurgy and powder metallurgy built up over the years has formed the basis for Sherritt's development of advanced new materials such as thoria dispersion strengthened metals and alloys for high temperature application

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in future jet propulsion systems.

Research and production groups pool their efforts in a continuous program of technical assistance to Sherritt's production plants which is carried out with the object of introducing new processes, raising the efficiency of existing processes and production procedures and maintaining the highest quality of metal and fertilizer products.

With the expansion of Sherritt's fertilizer operations a long range agronomic research program has now been initiated to study the effectiveness of Sherritt fertilizers throughout our principal marketing area, the Canadian prairie provinces. This enables Sherritt to make better recommendations and give better service to its customers.

The company's research plans for the future are designed to stimulate further growth by diversification.

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SERVICES AND UTILITIES

Sherritt, to a very large extent, supplies the services and utilities it requires to operate the plant.

The power house and water treatment plants supply all the requirements for steam, air, heat and water and almost 33% of the electric power. The daily water circulating load through the plant is about 75 million gallons. About 2,600,000 gallons of water per day are pumped from the North Saskatchewan River to make up for evaporation and other losses, equivalent to 9 tons per minute. The water treatment plant treats, purifies, cools and recycles the water.

Fertilizer shipping is very seasonable, up to 70% of shipments are made during a three month period. This necessitates large storage facilities. Sherritt's original storage building for ammonium sulphate has a capacity of 35,000 tons, urea storage 12,000 tons, and ammonia storage 2,000 tons. With the expanded plant, urea storage was increased by 35,000 tons, ammonia by 2,000 tons and for ammonium phosphates a 50,000 ton bulk storage building and

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a 6,000 ton bagged storage building was required, in addition three 6,000 ton silos for the storage of phosphate rock were required.

The new 50,000 ton bulk storage building for phosphates and the 35,000 ton bulk storage building for urea are circular buildings, or "agrodomes". They are two of the lightest steel dome structures in the world, for their size and applied load, having a 225' clear span roof. They provide a most efficient and economical solution for storage, and are very striking in appearance.

The storage and shipping facilities will permit the loading of over 100 trucks per day, in addition to shipments by rail.

The new track scale installed, at the time of the phosphate expansion, is one of the largest in North America, being designed to accommodate the jumbo rail car which was just coming into service. The size of the scale will allow these cars to be weighed as a whole instead of weighing in two parts.

Other facilities include maintenance shops, analytical laboratories, stores, purchasing, administrative and personnel offices.

About 850 to 900 persons are employed at the Sherritt Plant; operations are continuous, 24 hours a day, seven days per week.

Production from the plant, on a daily basis is about:

450 tons	Anhydrous Ammonia (90T available for sale)
540 tons	Ammonium Sulphate (370T available for sale)
6 tons	Hydrogen
300 tons	Urea
400 tons	Ammonium Phosphates
45 tons	Nickel
2,300 lbs.	Cobalt
11,500 lbs.	Copper (as Copper Sulphide)

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In addition to handling products, a considerable quantity of raw materials must be handled daily, including:

Water	2,600,000 gals.
Natural Gas	24,000,000 cu. ft.
Sulphuric Acid	600 tons
Phosphate Rock	520 tons
Lynn Lake Concentrate	300 tons
Other Metal Feed	25 tons

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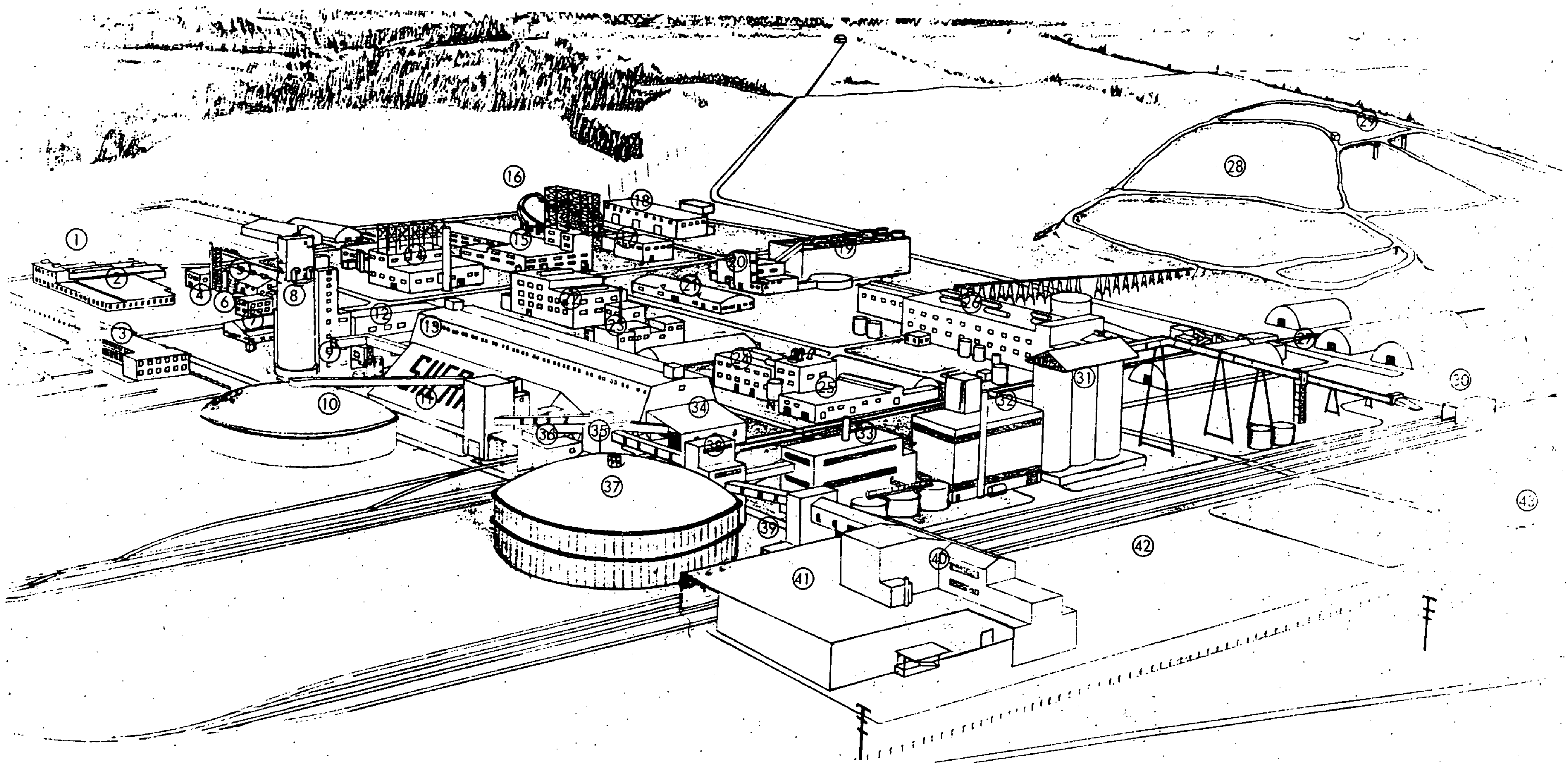
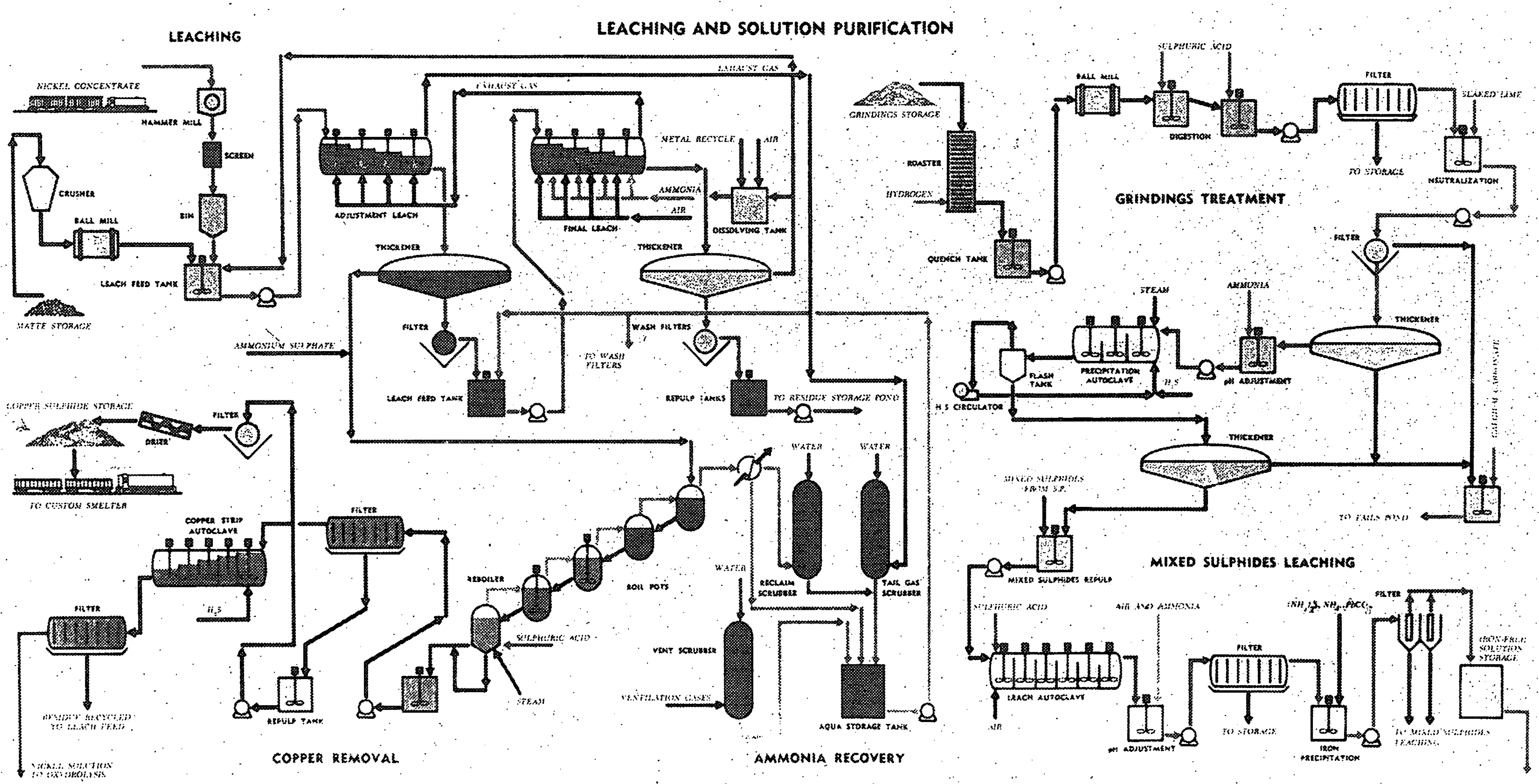


1. ROLLING MILL
2. RESEARCH AND DEVELOPMENT DIVISION
3. ADMINISTRATION BUILDING
4. PILOT PLANT
5. COBALT RECOVERY PLANT
6. OPERATIONS LABORATORIES
7. PERSONNEL AND MEDICAL BUILDING
8. UREA PRILL TOWER
9. UREA PRODUCTION PLANT
10. FERTILIZER UREA STORAGE BUILDING
11. FEED UREA STORAGE BUILDING
12. AMMONIUM SULPHATE PRODUCTION PLANT
13. AMMONIUM SULPHATE STORAGE BUILDING
14. GAS REFORM PLANT
15. AMMONIA PRODUCTION PLANT
16. AMMONIA STORAGE SPHERES
17. AMMONIA WEIGH
18. MAINTENANCE SHOPS
19. WATER COOLING TOWER
20. WATER TREATMENT PLANT
21. PURCHASING AND STORES
22. POWER HOUSE
23. SULPHIDE PRECIPITATION PLANT
24. METALS RECOVERY PLANT
25. METAL HANDLING BUILDING
26. LEACH PLANT
27. CONCENTRATE STORAGE BUILDINGS
28. LEACH RESIDUE POND
29. GYPSUM POND
30. PHOSPHATE ROCK UNLOADING BUILDING
31. PHOSPHATE ROCK STORAGE SILOS
32. PHOSPHORIC ACID PRODUCTION PLANT
33. AMMONIUM PHOSPHATE GRANULATION PLANT
34. AMMONIUM SULPHATE BAGGING PLANT
35. BULK WEIGH BINS
36. FEED UREA BAGGING PLANT
37. AMMONIUM PHOSPHATE STORAGE BUILDING
38. FERTILIZER BLENDING PLANT
39. TRACK SCALE
40. FERTILIZER BAGGING PLANT
41. BAGGED FERTILIZER STORAGE BUILDING
42. TRUCK PARK
43. TRAFFIC BUILDING


LEACHING AND SOLUTION PURIFICATION
LEACHING
NICKEL CONCENTRATE
HAMMER MILL
SCREEN
BIN
CRUSHER
BALL MILL
LEACH FEED TANK
MATTE STORAGE
EXHAUST GAS
EXHAUST GAS
ADJUSTMENT LEACH
FINAL LEACH
METAL RECYCLE
AIR
AMMONIA
AIR
DISSOLVING TANK
THICKENER
THICKENER
FILTER
LEACH FEED TANK
AMMONIUM SULPHATE
TO WASH FILTERS
WASH FILTERS
REPULP TANKS
TO RESIDUE STORAGE POND
COPPER SULPHIDE STORAGE
DRIER
FILTER
TO CUSTOM SMELTER
COPPER STRIP AUTOCLAVE
FILTER
FILTER
H_2S
RESIDUE RECYCLED TO LEACH FEED
NICKEL SOLUTION TO OXYDROLYSIS
REPULP TANK
COPPER REMOVAL
REBOILER
BOIL POTS
SULPHURIC ACID
STEAM
WATER
WATER
WATER
RECLAIM SCRUBBER
TAIL GAS SCRUBBER
VENT SCRUBBER
VENTILATION GASES
AQUA STORAGE TANK
AMMONIA RECOVERY
GRINDINGS STORAGE
ROASTER
HYDROGEN
QUENCH TANK
BALL MILL
SULPHURIC ACID
DIGESTION
FILTER
TO STORAGE
SLAKED LIME
NEUTRALIZATION
GRINDINGS TREATMENT
FILTER
STEAM
AMMONIA
THICKENER
FLASH TANK
PRECIPITATION AUTOCLAVE
H_2S
pH ADJUSTMENT
H_2S CIRCULATOR
CALCIUM CARBONATE
THICKENER
MIXED SULPHIDES FROM S.P.
TO TAILS POND
MIXED SULPHIDES REPULP
MIXED SULPHIDES LEACHING
SULPHURIC ACID
AIR AND AMMONIA
$(NH_4)_2S$, NH_4HCO_3
FILTER
LEACH AUTOCLAVE
AIR
pH ADJUSTMENT
FILTER
TO STORAGE
IRON PRECIPITATION
TO MIXED SULPHIDES LEACHING
IRON-FREE SOLUTION STORAGE

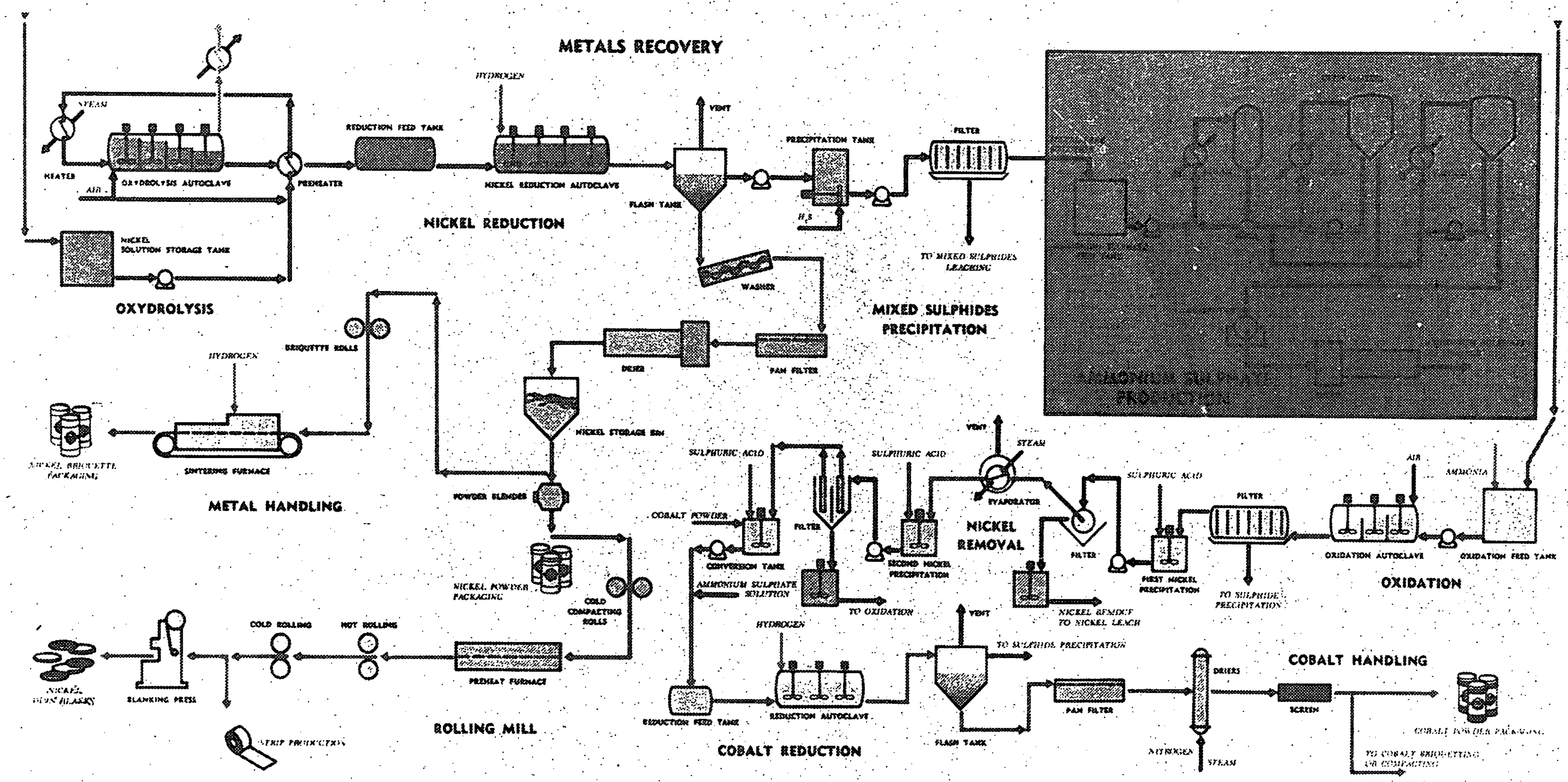

METALS RECOVERY
STEAM
HEATER
AIR
OXYDROLYSIS AUTOCLAVE
PREHEATER
NICKEL SOLUTION STORAGE TANK
OXYDROLYSIS
REDUCTION FEED TANK
HYDROGEN
NICKEL REDUCTION AUTOCLAVE
NICKEL REDUCTION
VENT
FLASH TANK
PRECIPITATION TANK
H_2S
FILTER
TO MIXED SULPHIDES LEACHING
WASHER
MIXED SULPHIDES PRECIPITATION
PAN FILTER
DRIER
AMMONIUM SULPHATE PRODUCTION
BRIQUETTE ROLLS
HYDROGEN
SINTERING FURNACE
NICKEL BRIQUETTE PACKAGING
METAL HANDLING
NICKEL STORAGE BIN
POWDER BLENDER
NICKEL POWDER PACKAGING
COLD COMPACTING ROLLS
PREHEAT FURNACE
HOT ROLLING
COLD ROLLING
BLANKING PRESS
NICKEL COIN BLANKS
STRIP PRODUCTION
ROLLING MILL
SULPHURIC ACID
COBALT POWDER
CONVERSION TANK
AMMONIUM SULPHATE SOLUTION
FILTER
TO OXIDATION
SULPHURIC ACID
SECOND NICKEL PRECIPITATION
VENT
STEAM
EVAPORATOR
NICKEL REMOVAL
FILTER
NICKEL RESIDUE TO NICKEL LEACH
SULPHURIC ACID
FIRST NICKEL PRECIPITATION
FILTER
TO SULPHIDE PRECIPITATION
AIR
AMMONIA
OXIDATION AUTOCLAVE
OXIDATION FEED TANK
OXIDATION
HYDROGEN
REDUCTION FEED TANK
REDUCTION AUTOCLAVE
COBALT REDUCTION
VENT
TO SULPHIDE PRECIPITATION
FLASH TANK
PAN FILTER
DRIERS
NITROGEN
STEAM
COBALT HANDLING
SCREEN
COBALT POWDER PACKAGING
TO COBALT BRIQUETTING OR COMPACTING

FIGURE A

UNIVERSAL MINERALS AND METALS INC.: SIMPLIFIED FLOW DIAGRAM FOR PRODUCTION OF COPPER POWDER FROM SCRAP: AMMONIUM CARBONATE PROCESS

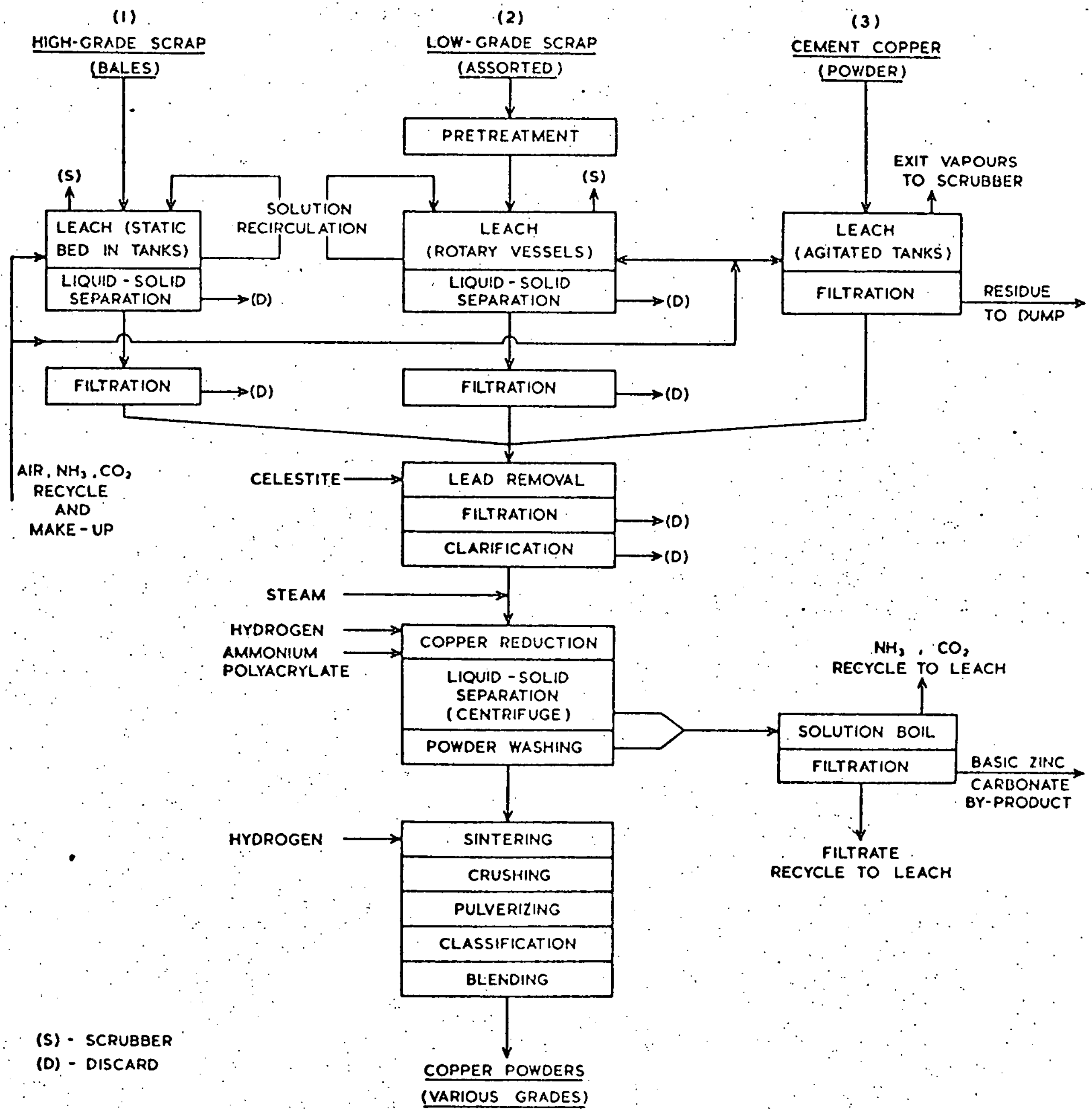
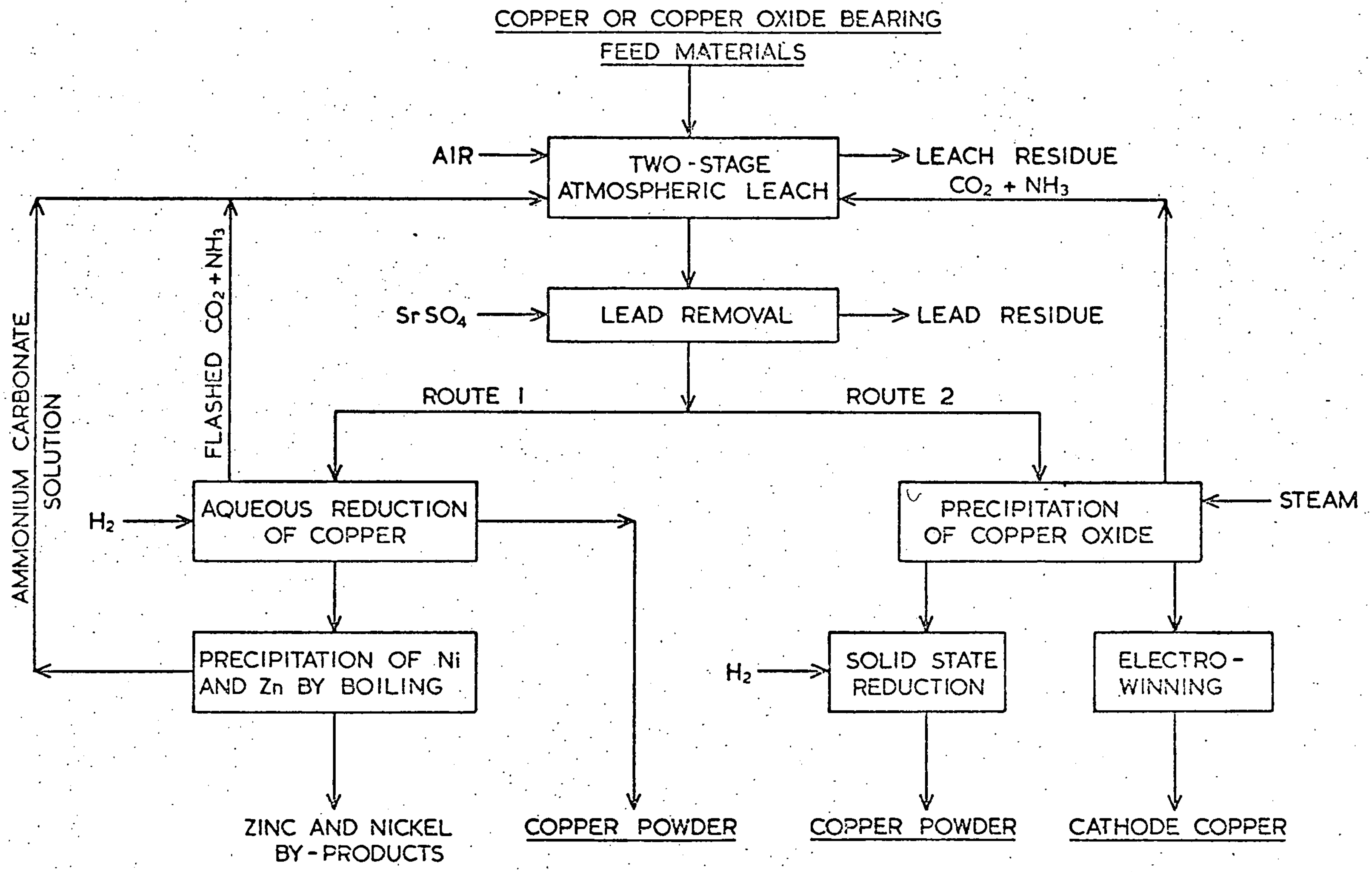


FIG. 12 FLOW SHEET FOR RECOVERY OF COPPER FROM COPPER OR COPPER OXIDE BEARING MATERIALS IN AMMONIUM CARBONATE SYSTEM.



VISIT 5: US BUREAU OF MINES

Situated at
Twin Cities Metallurgy Research Centre,
MINNEAPOLIS

DATE: Monday 9 November 1970

MET: Mr Morris M. Fine, Supervisory Metallurgist,
Mr Charles Praske
Dr John Nigro

Mr Norwood B. Melcher, Research Director was out of town but Mr Fine has worked and published extensively in the field of iron ore pre-reduction and the morning was spent in discussing the current status of developments. Their most recent review (Fine, Melcher and others 1970) is US Bureau of Mines Bulletin 651. Other useful references cited by Mr Fine were:

1. Jeffery, W.C., 1970 Chas. Ed Hoyt Memorial Lecture.
   AFS Transactions p 101 (1970) (Use of DLM product in foundries)
2. Fogleman, Gloven and Jensen
   BF and Steel Plant p 733 (October 1970)
   (Use of Fior product in an electric furnace - refractory life adversely affected)
3. McManus G.J. - Iron Age, August 17, 1970
   Gun Goes Off for Direct Reduction
4. Anon, Iron Age, May 21, 1970.

He said that an H-Iron plant treating 300,000 tons had been estimated to cost $US8.6 million and that it had been estimated that ore could be reduced for $US8.87 per ton. It had also been stated that Midland-Ross would sell reduced ore at $US30 per metric ton at the mouth of the Mississippi. He also referred to activities by Armco Steel near Houston, Texas. Mr Fine also said that North Star Steel Co., PO Box 3189, St. Paul, Minnesota. 55101 (Phone 612-735-2110) had a very nice mini-steel plant and were planning to use metalised pellets and also produce alloy steels. They have continuous casting using a self-levelling tundish and produce rebars, angles and custom bars. Dr J.C. Gay is Raw Material and Process Control Superintendent.

Mr Fine has just recently however handed over responsibility for iron ore research to Mr Charles Praske and assumed responsibility for copper. He supplied copies of US Bureau of Mines Bulletin R.I. 7314 and 7339 and a copy of a paper by himself and R.B. Schluter presented at the AIME Annual Meeting at Denver, February 1970 - the latter two dealing with recent work on treatment of a chalcocite concentrate.

Most of the afternoon was spent in discussions with Mr Charles Praske and Dr John Nigro, and in a tour of the facilities, mainly the pilot plant for metallisation of iron ore pellets, and conversion of a mixture of light automobile scrap and non-magnetic taconite into a magnetite concentrate. The plant is illustrated in Bulletin 651. An interesting unit for rapid magnetic determination of pellet metallisation, designated a "Ramsey Converter, RC-10" was built by the Ramsey Eng. Co. St Paul, Minn.

VISIT 6: METAL POWDERS INC.

Situated at
115 East 13th Avenue,
NORTH KANSAS CITY MISSOURI

DATE: Tuesday 10 November 1970

MET: Mr Robert Swafford, Vice President and General Manager of the Kansas City operation.

1. INTRODUCTION

Mr Swafford explained that Metal Powders Inc. is a subsidiary of McAlester Fuel Corp. which owns a mine and produces cement copper in Arizona. He supplied an illustrated brochure, extracts from which are appended, describing the operations and giving product specifications. He was able to spend 2½ hours describing their operations and showing me through the plant. Despite the necessity of my signing a non-disclosure agreement he was able to speak freely on all aspects of the process on which I enquired. He said the US consumption of copper powder, including that in pre-blended bronze powder was approximately 300,000 tons per annum. He said that products suitable for all applications except high conductivity electrical, could be made by hydrogen precipitation, and such powders, due to their irregular shape, gave highest green strengths. Another company had shown that his company's moulding-grade powder could be converted into flake by ball-milling. He suggested that friction-grade powder might even prove a substitute for flake. Two general grades of powder are produced - fine, light"friction-grade" and denser "moulding grade" (see Tables reproduced on pages A-54 to A-57).

2. PROCESS CONDITIONS

Metal Powders' main feedstock is cement copper produced by the parent company, but they also leach scrap (No. 1 and No. 2) and such special items as copper-clad iron wire, off-cuts of printed circuit board etc. The copper must be available to the lixiviant. Brass can be used but the pregnant solution should not exceed 10-15 g/l Zn as zinc salts deposit in the ammonia still which treats a bleed-off stream. Leaching is done at slight negative pressure as ammonia-pollution of the atmosphere is viewed seriously (probably without cause) by the plant's neighbours. Air is sparged in as required to oxidise copper from the elemental form, and the cuprous/cupric ratio is an important aspect which affects subsequent hydrogen precipitation. Generally the plant solutions are approximately 1:1 CuI:CuII but reagent requirements are affected. From memory Mr Swafford quoted the following:

$$1.03 \text{ g/l } NH_3 \text{ per g/l CuII}$$
$$0.67 \text{ g/l } NH_3 \text{ per g/l CuI}$$

These are theoretical requirements and in practice excess reagents have to be present to combine with zinc and other consumers. Generally the pregnant solution assays around 1 lb/gal copper, i.e. 120 g/l Cu, 120-140 g/l NH_3, 80 g/l CO_2. Ammonia is purchased but carbon dioxide enters the solution in the hydrogen used for reduction which is produced by reforming natural gas.

2.1 Hydrogen Precipitation

Mr Swafford said that there is a lot of "art" in producing the desired powder, but suggested that 75% of the control of powder form was in the precipitation stage and 25% of it in the sintering and milling post-treatment. Friction-grade powder is produced in a single precipitation and moulding-grade powder by multiple densification.

Hydrogen is produced by reforming desulphurised natural gas or refinery by-products. Assuming 4 vols. of hydrogen per vol. of methane:

$$CH_4 + 2H_2O = CO_2 + 4H_2$$

Mr Swafford estimated that they use 2-3 cu ft of hydrogen per pound of copper for precipitation and protection in the sintering furnace.

3. DESCRIPTION OF PLANT

Mild steel is generally used except in contact with ammonia vapour where stainless steel (Type 316 or possibly lower-grade) is required.

3.1 Leaching

Scrap is leached in three 8 ft diameter by 16 ft percolation tanks, cement in five 15 ft diameter by 7 ft 8,000 gallon tanks agitated by a 50 inch turbine blade at 50 rpm, and very light printed circuit board in a 60,000 gallon tank holding 200,000 lb. Scrap is loaded by clam-shell bucket and cement by front-end loader, and hatches are water-sealed. Vented ammonia is scrubbed and the solution used for washing etc. Scrap residue is similarly unloaded, but the 10-12% residue from cement is pumped out and removed on the filter, an 8 ft diameter by 10 ft Eimco vacuum unit operated with about 5 inches of precoat (5-600 lb per day) and fitted with a spray wash. The filtered solution goes to storage, whence it is drawn to the hydrogen-reduction autoclaves via a preheater.

3.2 Hydrogen Precipitation

Hydrogen is produced by a Chemical Construction Co. reformer, stored in a 5,000 c ft gasometer, and compressed from 6 ounces to 1,000 psi by a compresser with a capacity of 180 cfm at 1,000 psi. A 600 hp boiler (and a newer, slightly larger unit) supply process steam for heating solution and distilling ammonia and carbon dioxide from a bleed stream of lean solution. Distillation of this bleed stream is done to maintain the water balance and keep the impurity level down, and is accomplished with live steam in a 2 ft diameter mild steel column packed with Berl saddles, steam passing up the column countercurrent to the solution. The condensers are of stainless steel. Ammonia-water is recycled and ammonia and carbon dioxide gases piped to the leaching tanks.

The two reduction autoclaves, horizontal stainless steel-lined vessels approximately 5 ft diameter by 15 ft are fitted with three agitators. Solution is pumped in by a reciprocating positive displacement pump, made in stainless steel. The autoclaves are operated at approximately 300 deg. F, 1,000 psi total pressure (approximately 400 psi hydrogen). 25 minutes is required for charging, and 50-60 minutes for reduction. The slurry is allowed to settle for 5-15 minutes before discharging it through heat exchangers, which cool it to 100°F, into a stainless steel-lined holding tank. Discharge requires 20-30 minutes. A reduction rate of 39 lb per min is usual, but 60 lb per min has been attained.

The slurry is centrifuged in a suspended basket centrifuge and the solids washed before being sintered in a 20 ft long hydrogen-swept gas-fired furnace through which the precipitated copper passes as a 2 inch cake on a 3.5 ft wide stainless steel belt.

Furnace temperature is 1,300°F and throughput 1,700 lb per hr. The sintered product is broken up by saw-tooth crushers and milled and sized by a Mikro pulveriser hammermill and air classifier and Sweco screen. Fines are collected by cyclone and Wheelabrator baghouse. Run-of-mill product is drummed in black drums, and blending to specification is carried out by two men on day shift only in trailer-loads of 40,000 lb constituting 3-4 batches in a Patterson blender. The final product is then packed in blue 600 lb drums.

3.3 Quality Control

The physical testing laboratory, in addition to the usual equipment for determining sizing and flowrate, had presses and equipment for determining green strength of compacts and furnaces in which compacts could be sintered to determine "growth".

The analytical laboratory, staffed by a chemist and two assistants, had facilities for determining copper in feed electrolytically and assaying the final product. Shift operators make routine determinations on plant solutions for CO_2 (add phosphoric acid, CO_2 displaces salt solution), NH_3 (Kjeldahl and back titration - 15 minutes), total Cu (add acetic acid, ammonia, use colorimeter).

3.4 Labour

The shift crew consists of five men, of which one is the leading hand:

Leaching and Filtration:	1 operator
Reduction and Distillation:	1 operator
Centrifuging:	1 operator
Drying and Sintering:	1 operator
Reforming:	(fraction of 1 operator's time)

4. FURTHER ACTION

Mr Swafford would be happy to supply samples to bona fide enquirers in Australia, and to supply commercial amounts in Australia if tariffs etc. did not price their product out of the market.

METAL POWDERS, INC.
A VERY SPECIAL CAPABILITY FOR A VERY SPECIAL COPPER POWDER

The Metal Powders, Inc., plant in North Kansas City, Missouri, was the first to produce copper powder commercially by hydrogen reduction of a complex of cuprous and cupric ammonium carbonates in solution.

The process is called hydrorefining. It is unique. The copper powder produced by this process is also unique in its physical characteristics. These characteristics, in turn, offer significant advantages to the producers and users of powdered metal parts and components. Details of these advantages, and the process that makes them possible, are covered on the following pages.

However, there is one distinctive feature of the process which should be mentioned here: The hydrorefining process has the capability of extracting copper powder economically from oxide ore concentrates and almost any copper-bearing scrap in concentrations as low as 5% by weight.

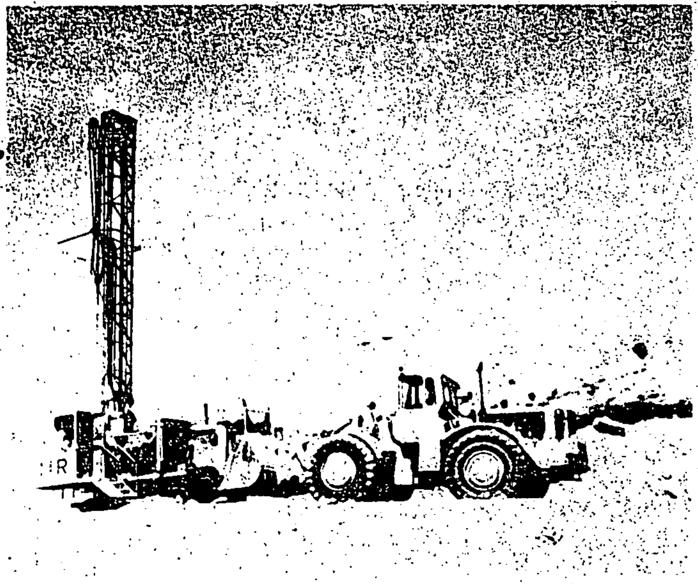
Since the process can be used economically with ore concentrates, the price of finished powder need no longer be tied to the volatile and unpredictable cost of copper scrap.

Given a source of raw material in the ground, independent of the secondary copper market, price stability is assured and certain economies can be passed on to copper powder customers. Metal Powders, Inc. has such a source.

The company is a subsidiary of McAlester Fuel Company, an organizaztion which specializes in the extracton of natural mineral resources. McAlester's Zonia mine in Kirkland, Arizona, provides the copper ore for the Metal Powders refining plant in North Kansas City.

This plant, in the heart of the midwest industrial complex, also has access to copper scrap which can be blended, when economically warranted, with the ore concentrates in the refining process.

From this plant comes a wide range of high purity, tailor-made copper powders created to meet the individual needs of our customers.



The Zonia copper mine has a minimum of fifteen years proven reserves, with thousands of acres showing a high degree of mineralization still to be drilled out.

THE IMPORTANCE OF BEING CLUSTERED

Copper powders produced electrolytically, have a branched, or dendritic, structure. Powders produced by atomization are nugget-like. But copper powders produced by the Metal Powders, Inc. hydrorefining process have a grape-like cluster, or botryoidal, structure.

Our research, and our customers' on-the-job experience, have demonstrated that the clustered structure with its irregular surface area has several important effects on physical properties.

Most important, MP powders have higher green strength. This translates directly into lower production losses between the die and the sintering furnace.

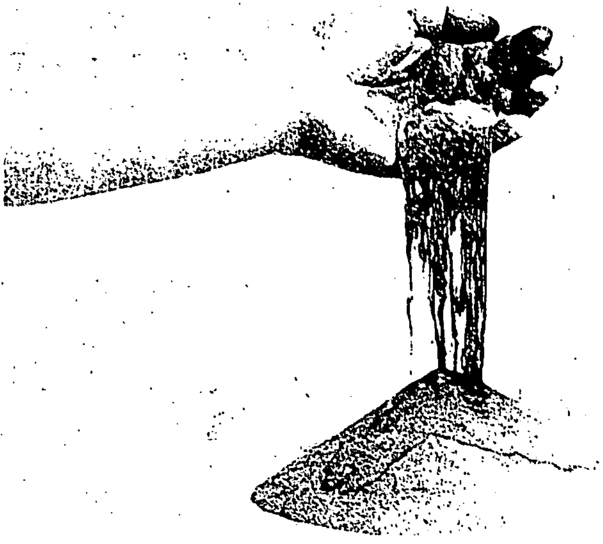
Second, the clustered configuration provides higher flowability. This means shorter cycle times and increased press production.

Third, hydrorefining results in highly uniform powders. Since it is a chemical process, it is susceptible to strict quality control and makes possible the production of powders made to exacting specifications. The obvious result is higher quality sintered parts.

It has also been found that hydrorefined copper powder sinters at lower temperatures, and in a shorter period of time, than powders produced electrolytically or by atomization.

MP powders have exceptional stability and resistance to oxidation over long periods of storage and at temperatures considerably above ambient.

These properties add up to considerable savings to fabricators in terms of powder utilization, power consumption, and speed of fabrication.



47 OTHER WAYS TO UTILIZE COPPER POWDERS

MP copper powders are produced in molding and friction grades. Both grades are available in a wide range of screen sizes, apparent density, green strength, crystal growth and flowability. These and other characteristics can be varied to meet your requirements by altering the hydrorefining process parameters.

The range of applications can be still further extended [illegible] many ways: By the addition of other powders or materials, or by the use of copper powders as alloying agents with other metals.

Probably the widest use of molding grade copper powders is in the manufacture of oilless bearings. Friction grade powders are used primarily in clutch plates and brake linings.

However, imaginative engineers have discovered dozens of other applications. The list below includes 47 of them. Perhaps, their ideas will stimulate some of your own.

Abrasive wheels

Agricultural applications: Farm machinery, food enrichment, fungicides, lawn and garden equipment, soil conditioning.

Aerospace: Brake lining, counterweights.

Automotive: Bushings, brake bands and linings, engine [illegible]d transmission applications.

Building and Construction: Conductive and non-spark flooring, decorative plastics and linoleum, pipe joint compounds.

Catalysts

Coatings: Anti-fouling paints, conductive paints and plastics, corrosion-resistant paints, decorative paints, spray coatings, vacuum metallizing.

Electrical and Electronic: Brushes, contacts, crystal supports, printed relays.

Electrical Discharge and Electromechanical Machining: Electrodes, preforms.

Heat treatment: Infiltration

Industrial (general): Bearings and bushings, friction material.

Joining: Brazing, resistance welding electrodes.

Lubricants: Anti-galling pipe joint compound, copper lubricants, plastic-filled metals.

Magnets

Nuclear Engineering: Gamma ray shielding.

Ordnance: Armor-piercing cores, incendiary bombs, flares, torpedos, tracers, etc., projectiles rotating bands.

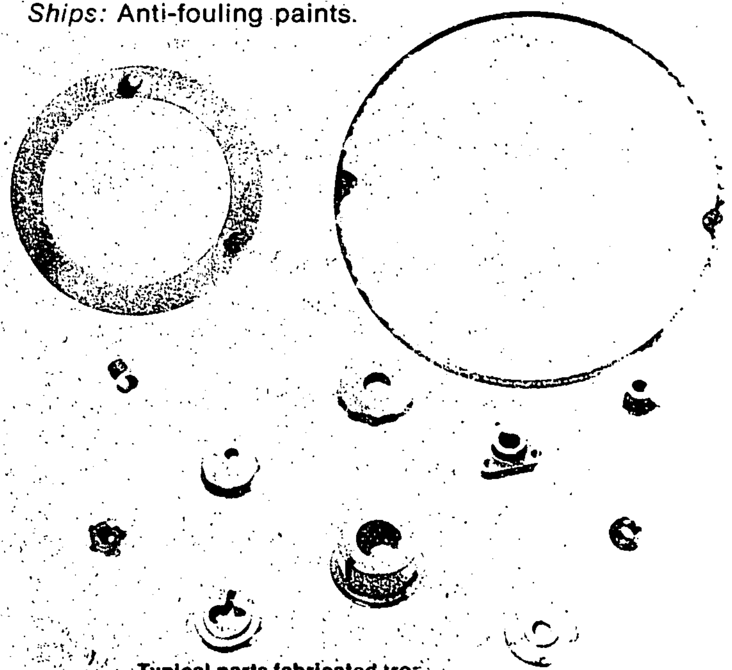
Personal Products: Cordless electric toothbrush and razor, fingernail lacquer.

Radio and Television: Permanent magnets, printed circuits.

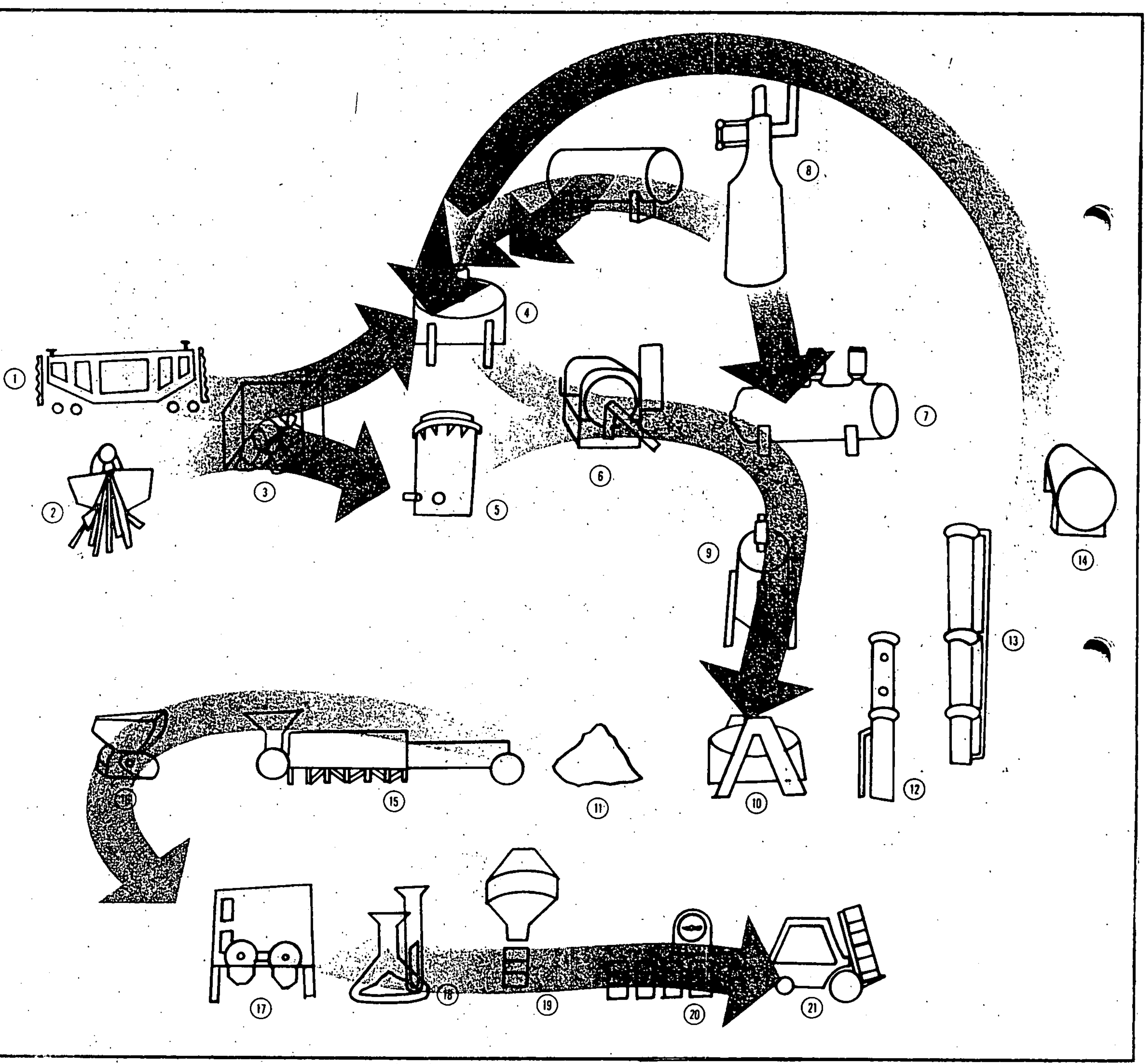
Railroads: Brake linings, friction strips on pantograph.

Self-lubricating Parts: Oil-filled, plastic-filled.

Ships: Anti-fouling paints.



Typical parts fabricated from copper powder.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THE SECRET OF OUR SUCCESS

The hydrorefining process at the MP plant in North Kansas City, Missouri, currently produces 300-350 tons of high purity copper powder per month. Because of increased demand for the special characteristics of hydrorefined copper powder, plans have been completed to expand the plant to handle 750 tons per month.

① & ② Copper cement from the Zonia mine and copper scrap arrive in gondola cars or by motor freight.

③ Raw materials are assayed and stored in separate bins.

④ & ⑤ Copper cement is charged by weight to one of five cement leach tanks. Copper scrap is charged to one of three scrap towers. Raw materials are solvated in a solution of ammonium carbonate and sparged air to produce rich copper-bearing liquor.

⑥ The rich liquor is purified to remove contaminants such as tin, lead, sulfate, phosphate, etc. Then it is filtered through a large rotary vacuum precoat filter. The rich liquor is then stored for future use or fed directly to the autoclaves.

⑦ In the two high-pressure autoclaves the copper powder is reduced by hydrogen at 1000 lbs. gauge, at 350°F.

⑧ Hydrogen for the autoclaves is produced from natural gas or propane in the reform plant.

⑨ When the reaction is complete, the charge is decanted overhead.

The ecologically-minded reader should note that the refinery does not produce noxious odors, loud noises or movement of heavy equipment, and that all reagents, except stripping chemicals, are reused.

For the process-minded reader, a schematic of the plant is shown at left. This is the way it works:

⑩ & ⑪ The charge is then run to the centrifuges where refined copper powder is harvested as wet cake.

⑫ & ⑬ Ammonia and carbon dioxide are recovered in stripping columns and absorbers.

⑭ Lean liquor is stored in tanks before recycling to the leachers.

⑮ The wet cake is fed to a hydrogen oven where it is dried and reduced to bright copper cake.

⑯ At the exit end of the drier, a sawtooth crusher breaks up the cake.

⑰ The powder is screened and classified into various grades.

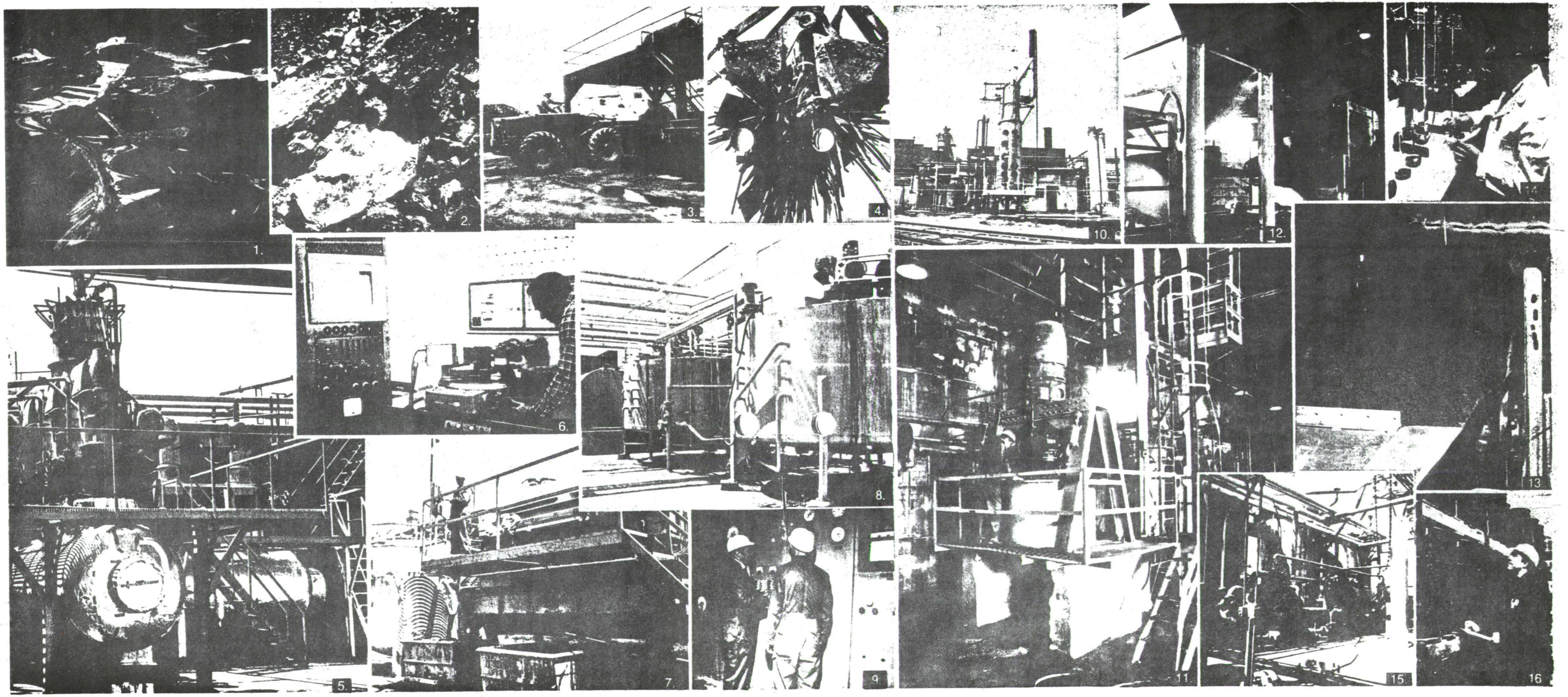
⑱ & ⑲ Finished powder is checked by quality control and then blended to the specifications of various customers.

⑳ & (21) Powder is weighed and loaded into 600-pound vapor-proof drums. These are banded four to a pallet and shipped by motor freight.

FOR THE PICTORIALLY-MINDED

Shown below are some of the facilities and equipment of Metal Powders, Inc. used in various stages of the production of copper powder from ore concentrates.

1. Aerial view, Zonia mine
2. Copper oxide ore
3. Loading copper cement at the Zonia mine
4. Unloading scrap at the refinery
5. Autoclaves
6. X-ray spectrophotometer
7. Rotary precoat filter
8. Cement leach tanks
9. Autoclave panelboard
10. Reform plant
11. Centrifuge
12. Hydrogen reduction oven
13. Exit end of drier
14. Quality control laboratory
15. Scrap leach tanks
16. Warehousing



CREATIVE TECHNICAL SERVICE BEGINS IN THE LABORATORY . . .

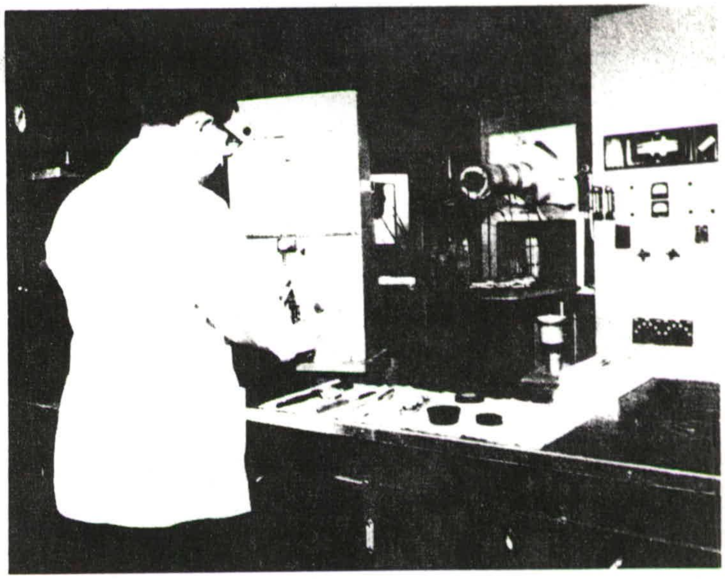
Metal Powders, Inc. maintains a well-equipped quality control laboratory to insure that each lot of powder meets your most exacting specification.

In addition, MP is well equipped to handle both physical and chemical laboratory analyses on behalf of its customers for use in the development of powders for specialized applications.

The capabilities of MP's own experienced personnel are further extended by our continuing contract services with two highly respected independent research organizations.

Midwest Research Institute in Kansas City, Missouri, provides highly sophisticated physical and chemical analyses. SKC P/M Engineering is utilized for physical and mechanical testing and the development of new applications in the use of MP copper powders.

This combination of in-house and independent research assures Metal Powders, Inc. customers of accurate information and the most advanced thinking the use of copper powders across a broad spectrum of applications.



. . . AND CONTINUES INTO YOUR PLANT

MP technical service starts with consultation on the use or development of copper powders to fit your requirements . . . and doesn't end until your job is coming smoothly out of the sintering furnace.

Metal Powders, Inc. supplies free samples for testing and evaluation. All powders produced by MP are specifically tailored to our customers' requirements and manufactured to customers' specifications. Chemical and Physical Specifications Sheets, certified by Metal Powders, Inc., are forwarded with preshipment samples to assure strict quality control.

And, of course, MP engineers and technical personnel are always available for consultation at any stage of the process from powder development through final product production.

BASIC FACTS YOU SHOULD KNOW ABOUT MP POWDERS

MP COPPER POWDERS

MP Grades	CHEMICAL ANALYSIS					PHYSICAL PROPERTIES					
						Screen Analysis					
	Copper Content % Min.	Hydrogen Loss % Max.	Acid Insol % Max.	Apparent Density	Flow (Sec./50 g)	100 +	− 150 +	− 200 +	− 270 +	− 325 +	−
Granules	99.5	0.30	0.02	3.80	25	− 40 + 100 40-45	35-40	10-15	2 Max.	–	–
MP 170-90	99.0	0.75	0.05	1.50-2.20	0	0.2 Max.	0.5 Max.	5 Max.	5 Max.	5 Max.	90 Min.
MP 250-80	99.0	0.75	0.05	2.40-2.70	0	0.2 Max.	0.5 Max.	0.5 Max.	2 Max.	5 Max.	95 Min.
MP-250-60 LG	99.50	0.30	0.02	2.55-2.75	30	0.2 Max.	1-5	5-10	10-15	5-10	60-70
MP-250-60 HG	99.50	0.30	0.02		30	0.2 Max.	1-5	5-10	5-10	5-10	65-75
MP 280-60 MG	99.50	0.30	0.02	2.75-2.95	30	0.2 Max.	1-5	5-10	10-15	5-10	55-65
MP 280-60 HG	99.50	0.30	0.02		30	0.2 Max.	1-5	5-10	10-15	5-10	60-70
MP 300-40 LG	99.50	0.30	0.02	3.00-3.20	30	0.2 Max.	1-5	10-20	10-20	5-10	35-55
MP 300-40 HG	99.50	0.30	0.02	3.25-3.40	30	0.2 Max.	1-5	10-20	10-20	5-10	40-50

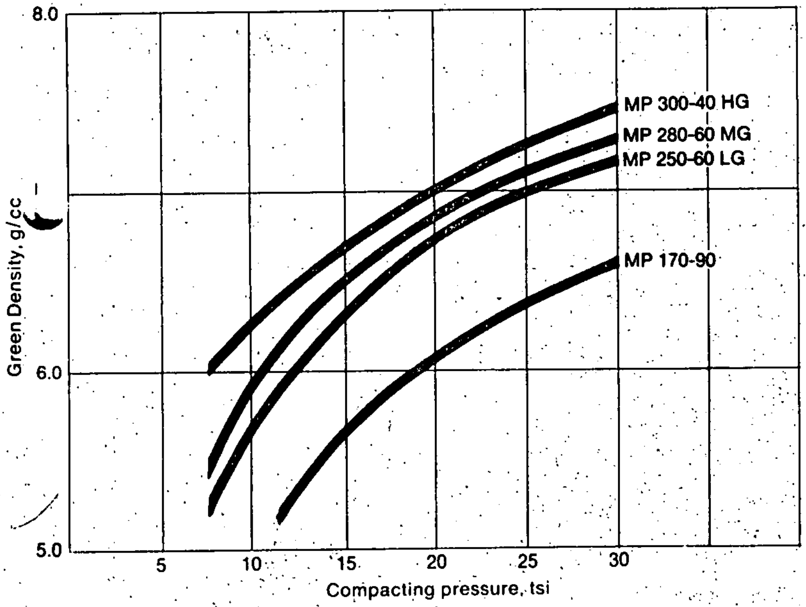
MECHANICAL PROPERTIES OF MP COPPER POWDER

MP	Compacting Pressure	Sintering Time	Density, g/cc		Hardness	Tensile Strength	Elongation	Dimensional change
Grade	tsi	hrs.	Green	Sintered	Rockwell	psi	%	%
MP 170-90	20	1/2	6.44	7.53	H-79	30,600	21.0	− 9.4
	20	1		7.54	H-78	32,500	31.0	− 10.0
	30	1/2	6.80	7.71	H-74	31,000	21.9	− 7.0
	30	1		7.60	H-74	30,600	27.1	− 7.2
MP 250-60	20	1/2	6.82	7.45	H-67	24,900	14.8	− 5.2
	20	1		7.57	H-71	28,500	23.0	− 6.2
	30	1/2	7.34	7.67	H-69	26,400	17.5	− 3.1
	30	1		7.77	H-71	28,700	23.7	− 3.9
MP 280-60	20	1/2	6.84	7.49	H-63	20,600	11.2	− 4.9
	20	1		7.59	H-66	25,500	18.4	− 5.9
	30	1/2	7.40	7.75	H-68	21,400	11.5	− 3.5
	30	1		7.85	H-69	25,100	16.1	− 4.4
MP 300-40	20	1/2	6.87	7.13	H-41	14,500	7.6	− 1.9
	20	1		7.31	H-49	17,900	10.1	− 3.3
	30	1/2	6.55	7.40	H-50	16,500	8.8	− 1.3
	30	1		7.56	H-55	21,400	15.1	− 2.4

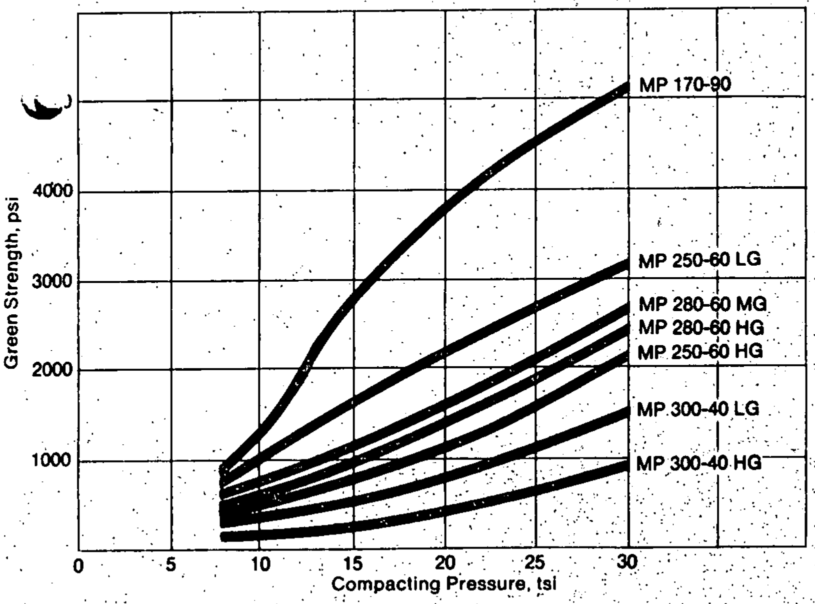
TEST CONDITIONS:
LUBRICANT 3/4% ZINC STEARATE
SINTERING TEMP. 1832°F (1000°C
SINTERING ATMOSPHERE CRACKED NH_3

MP COPPER POWDERS

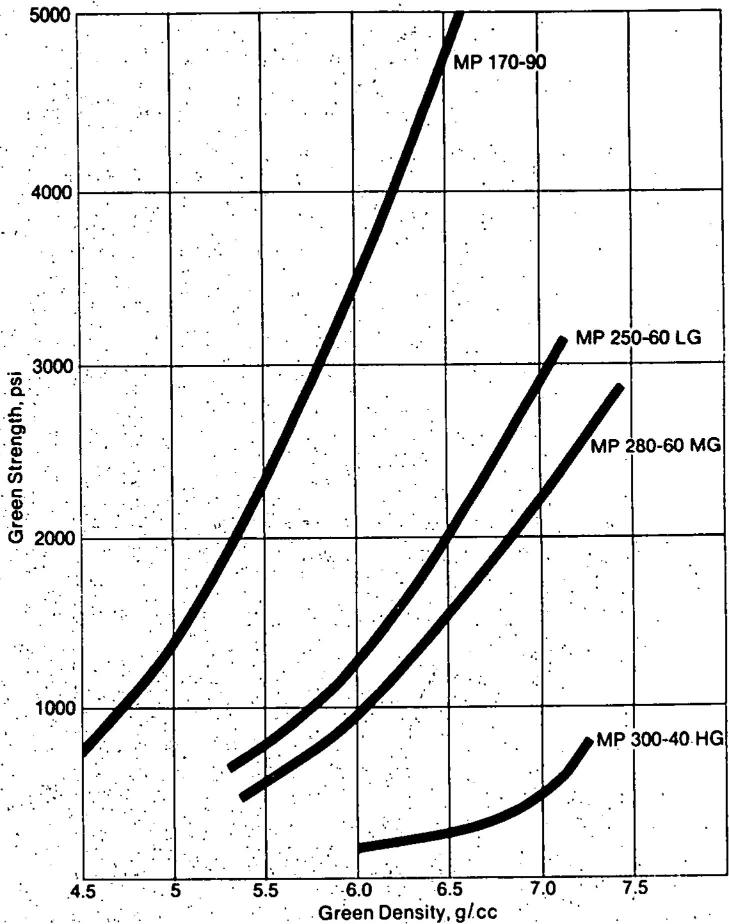
GREEN DENSITY vs. COMPACTING PRESSURE
¾ % ZINC STEARATE



GREEN STRENGTH vs. COMPACTING PRESSURE
¾ % ZINC STEARATE



GREEN DENSITY vs. GREEN STRENGTH
¾ % ZINC STEARATE



90/10 BRONZE

COMPOSITION: 90% COPPER
5% TIN—COARSE
5% TIN—FINE
1% GRAPHITE
¾ % ZINC STEARATE

SINTERING TEMPERATURE: 1525° F
SINTERING ATMOSPHERE: NH_3

TENSILE PROPERTIES OF 90/10 BRONZE

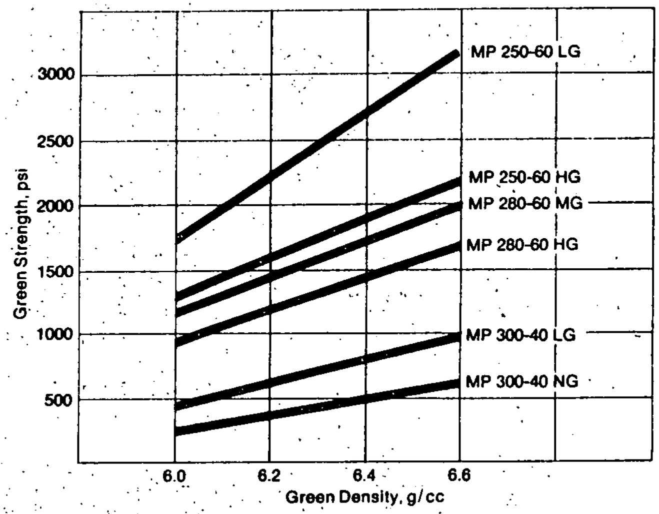
COPPER POWDERS	Density, g/cc		Hardness	Tensile Strength	Elongation	Dimensional Change, Length
	Green	Sintered	Rockwell	psi	% 1″	%
MP 280-60	6.07	6.02	H-39	14,300	6.7	– 0.57
Electrolytic Cu	6.00	5.99	H-28	12,200	6.5	– 0.54
Reduced Cu	5.96	5.90	H-30	13,300	6.7	– 0.57

90/10 BRONZE BUSHINGS

	Green Density, g/cc	6.0		6.6	
	Sintering Time, Min.	10	15	10	15
MP 280-60 M.G.	Sintered Density, g/cc	5.81	5.84	6.12	6.20
	K Factor, psi	23,200	23,600	29,500	29,500
	Change in Diameter in./in.	– 0.001	0	+ 0.011	+ 0.008
MP 280-60 H.G.	Sintered Density, g/cc	5.48	5.80	5.98	5.94
	K Factor, psi	18,900	20,700	25,500	25,800
	Change in Diameter in./in.	+ 0.015	+ 0.003	+ 0.020	+ 0.021

BUSHING SIZE: 1″ OD x ¾″ ID x 1″ L

90/10 PREBLENDED BRONZE, GREEN DENSITY vs. GREEN STRENGTH



MP COPPER POWDER ADDITION TO IRON POWDER

	Copper Powder	Density, g/cc		Hardness	Tensile Strength	Elongation	Dimensional Change
		Green	Sintered	Rockwell	psi	% 1"	% Length
Composition 97% Iron 3% Copper 1% Graphite 3/4% Zinc Stearate *Conditions* Compacting Pressure, 30 tsi Preburned 1200°F. 1/2 hr. Sintering Temp., 2050° F. Sintering Time, 1 1/2 hrs. Sintering Atmosphere, Cracked NH_3		REDUCED IRON					
	MP 280-60	6.49	6.35	B-49	45,800	2.7	+ 0.45
	Electrolytic Cu	6.43	6.33	B-45	44,700	2.6	+ 0.48
	Reduced Cu	6.44	6.32	B-47	45,200	2.2	+ 0.48
		ATOMIZED IRON					
	MP 280-60	6.71	6.64	B-81	72,300	1.8	0
	Electrolytic Cu	6.65	6.60	B-78	73,400	1.9	0
	Reduced Cu	6.69	6.63	B-75	69,700	1.8	0

VISIT 7: MIDREX DIVISION, MIDLAND-ROSS CORPORATION

Situated at
900 North Westwood
TOLEDO OHIO

DATE: Wednesday 11 November 1970

MET: Mr Norman J. Klatt, Sales Manager

1. INTRODUCTION

Mr Klatt said that Midland-Ross Corporation policy was to build and operate metallised pellet plants throughout the world and to supply metallised pellets at a stable price on a guaranteed basis. Midland-Ross Corporation would invest their own money and own the plant, and would normally require a 10 year contract for sale of product. An exception to the above policy has been published in the press (page A-62) where it is stated that Midland-Ross Corporation will participate in a joint venture with Chugai Ro Kogyo Kaisha Limited to build a 400,000 tons per annum plant in Japan. Midland-Ross Corporation would be very interested in investing in Australia.

2. CURRENT ACTIVITY

The metallised pellet plant at Portland, Oregon was stated to be operating (though competitors have since suggested that continuous smooth operation has not yet been achieved). The plant at Georgetown, S. Carolina will be run-in in January 1971 and pellets supplied to the steel mill in April 1971. The plant in Hamburg, Germany will be run-in in April 1971 and operating in June-July 1971, this latter plant involving a new steel-mill as well as pre-reduced pellet plant. A plant is projected for Japan as mentioned above, and plants are under consideration for New Orleans (800,000 tons per annum - to supply mini-steel mills and ironmakers) and Montreal (Dominion Steel and Sidbeck?)

3. LOGISTICS

The most important siting considerations are an ample supply of natural gas and an assured market for the product. A plant to produce 400,000 long tons per annum of metallised pellets, i.e. 400,000 short tons per annum of steel, will use 650,000 cf per hour of gas. The budget price in Adelaide (35-40 cents per thousand cft) was considered reasonable, the estimated cost at New Orleans being 30-35 cents. Other requirements for a 400,000 tons per annum plant are:

30-50 acres land

700 gpm water (100 gpm potable, 600 gpm cooling for closed-circuit operation

100 kw electricity per ton

The ore to be used must be low in Cu, P, S and gangue, as the gangue content is increased by approximately 50% by removal of oxygen. The Portland plant was designed to use Marcona ore but the plants in Georgetown and Hamburg will use Swedish oxide pellets. A magnetite ore or concentrate could be processed in a shaft kiln as at Portland, but a hematite would require a grate which would increase the capital cost of the plant. A deep-water port is very desirable, although the depth available at New Orleans will only be 40 ft (Savage River pellets have been tried and proved satisfactory, and the TiO_2 content can be fluxed.)

4. COSTS

4.1 Capital

Mr Klatt confirmed that the Portland plant, consisting of both pre-reduction plant and steel plant, cost $US35 million. The pre-reduction plant including pelletisation cost $US12 million.

4.2 Treatment

Two alternatives are possible, Midland-Ross Corporation purchasing either iron ore concentrate, or fired oxide pellets. In the first case an ore cost of $15 plus a conversion cost of $21 gives a total cost of $36 per long ton of ore treated. In the second a pellet cost of $18 plus a conversion cost of $18 gives a total cost of $36 per long ton of oxide pellets. These conversion costs are conservatively high. 1.38 tons concentrate give 1 ton pellets and 550,000 tons oxide pellets yield 400,000 tons reduced pellets. If the plant capacity were 1.2 million tons per annum the conversion costs would be reduced by $2 per ton. Based on Midland-Ross Corporation experience in installing the shaft kilns at Savage River, Australian costs are a little lower than US.

Regarding the use of lump ore, Mr Klatt said they had made some trials, but this was a grey area. Asked if Midland-Ross Corporation had tried reducing green or dried balls without prior induration he said that strong pellets were required to avoid attrition and fines production. Oxide pellets normally have a strength of 500 lb and metallised pellets 100-150.

5. PRODUCTION ASSESSMENT

Analysis of the Portland product was given by Midland-Ross Corporation in correspondence (Table A-7.1). A sample of Portland metallised pellets was provided. These were stated to have been produced 8 weeks previously and to assay 2%C. The New Orleans plant is programmed to produce a portion of its output at 1.5%C for steel mill use, and Copperweld and Babcock & Wilcox are potential customers. The balance of the output will contain 3.0% + C for foundry use. The pellets can be briquetted and a sample briquette (old, produced in lab. and probably containing 1% P) was obtained. General Motors produced 5 inch diameter slugs described in a paper by Hafner & Clow (pages A-76 to A-82). Steelmaking tests at Portland are described in a Midland-Ross Corporation report dated December 1967 (pages A-71 to A-75).

6. FURTHER ACTION

If Midland Ross Corporation are asked to carry out a detailed feasibility study they require $100,000 to be deposited. The study normally costs $30,000-$50,000, of which the balance is either refunded or credited to the client as product. If Midland Ross Corporation decide not to proceed the whole $100,000 is refunded. Two such studies have been made to date, one being for the Canadian government.

The concept of Midland Ross Corporation supplying capital as well as engineering is attractive and this process will be further considered in detail.

TABLE A7.1: SPECIFICATION OF MIDREX METALLISED PELLETS

	Indicated %		Typical	
	Chemical Analysis			
Metallisation	96		95	min.
Total Fe	93		91.0	min.
Mn	.08		0.1	max.
Cu	0.017		0.07	max.
P	0.007		0.05	max.
S	0.002		0.02	max.
Pb, Sn, V, Mo, Zn, Cr, W, As, Sn, Co, Ni	0.05	max.	0.1	max.
Carbon	1.0		.7%	min.
Gangue (total - excluding carbon)	5.0	max.	8	max.
SiO_2	1.93	approx.	2.90	max.
Al_2O_3	.80	approx.	.85	max.
MgO	.46	approx.	0.67	max.
CaO	.47	approx.	0.55	max.
TiO_2	.01	approx.	0.07	max.
	Physical Characteristics			
Size	Nominal $^3/_8$" to $^1/_2$" 1% max. minus 28 mesh		3% max. minus 28 mesh	
Strength per pellet	250 lb min.		150 lb min	
Bulk Density	120 lb/cu ft min.		110 lb/cu ft min	
Tumble Index (ASTM)	28 mesh 90		85 max.	

Note: Percent Metallization = $\frac{\text{Metallic Iron}}{\text{Total Iron}}$

MIDLAND-ROSS FORMS JOINT VENTURE IN JAPAN

Chugai Ro Kogyo Kaisha Will Be Partner in Building and Running Pelletizing Facility for Steel

By a Wall Street Journal *Staff Reporter*

CLEVELAND - Midland-Ross Corp. said it will join Chugai Ro Kogyo Kaisha Ltd of Osaka, Japan, in a venture to build and operate a multimillion-dollar plant in Japan to produce metalized pellets used in electric-furnace steelmaking.

The facility, which is subject to approval of the Japanese government, will have an initial capacity of at least 400,000 tons of metalized pellets a year. Construction is expected to begin about the middle of next year and production is scheduled about 18 months later.

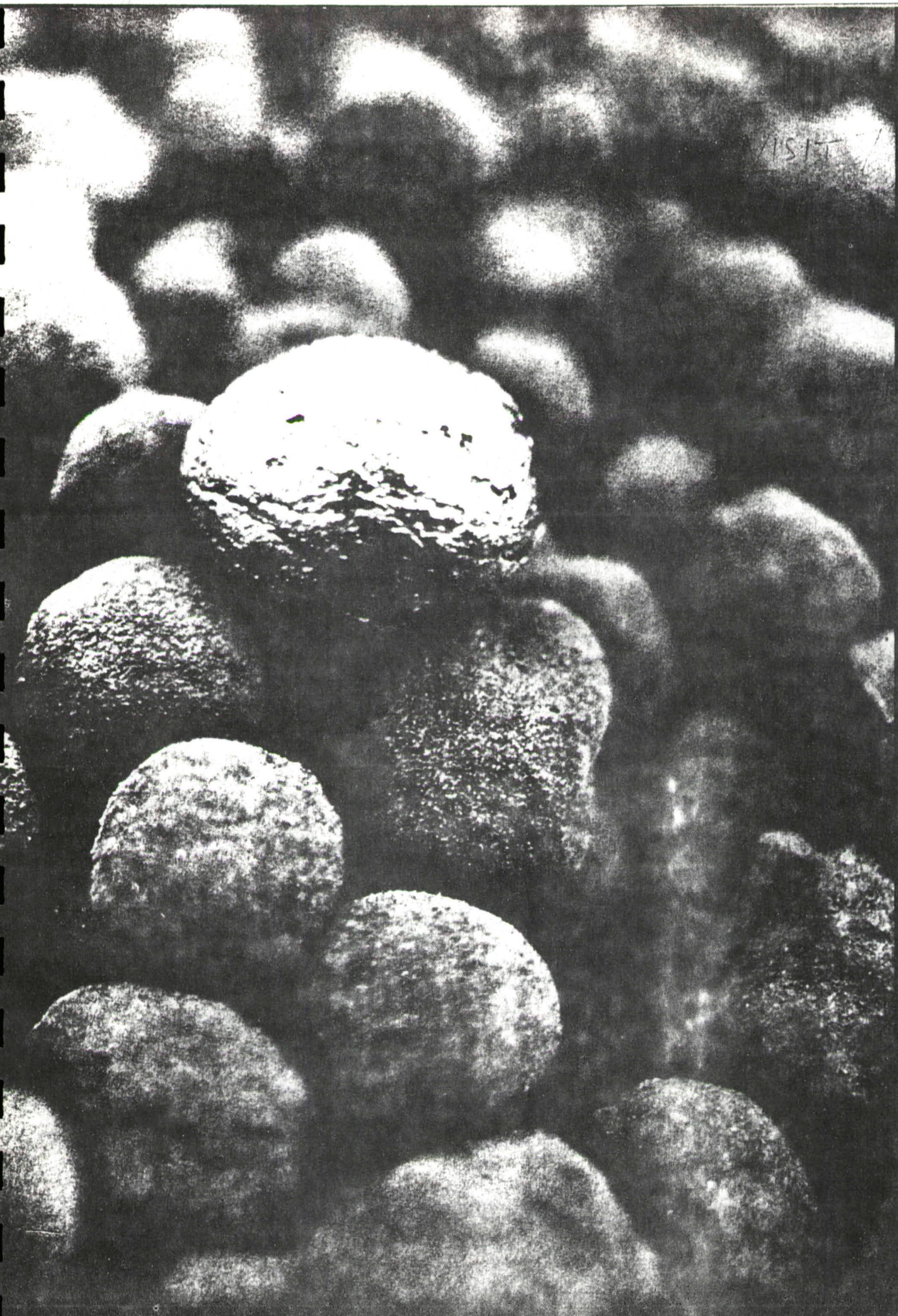
Production of the plant will be marketed under long-term contracts through trading companies.

The announcement said Midland-Ross will receive an initial fee for use of its metalizing process, will be paid a royalty on the output of the plant and will own 50% of the joint venture. The agreement also provides that future metalizing installations in certain areas of the Far East will be constructed by Chugai Ro for the joint venture.

Midland-Ross operates a metalized pellet plant at Portland, Ore. Other of its plants are under construction in Georgetown, S.C., and Hamburg, West Germany.

The Japanese plant will increase the worldwide capacity for Midland-Ross metalized pellets to at least 1.6 million tons a year, the company said. By the end of 1971, the company expects to have additional metalizing plants under construction that will increase the combined capacity of the facilities when completed to about 2.5 million tons annually.

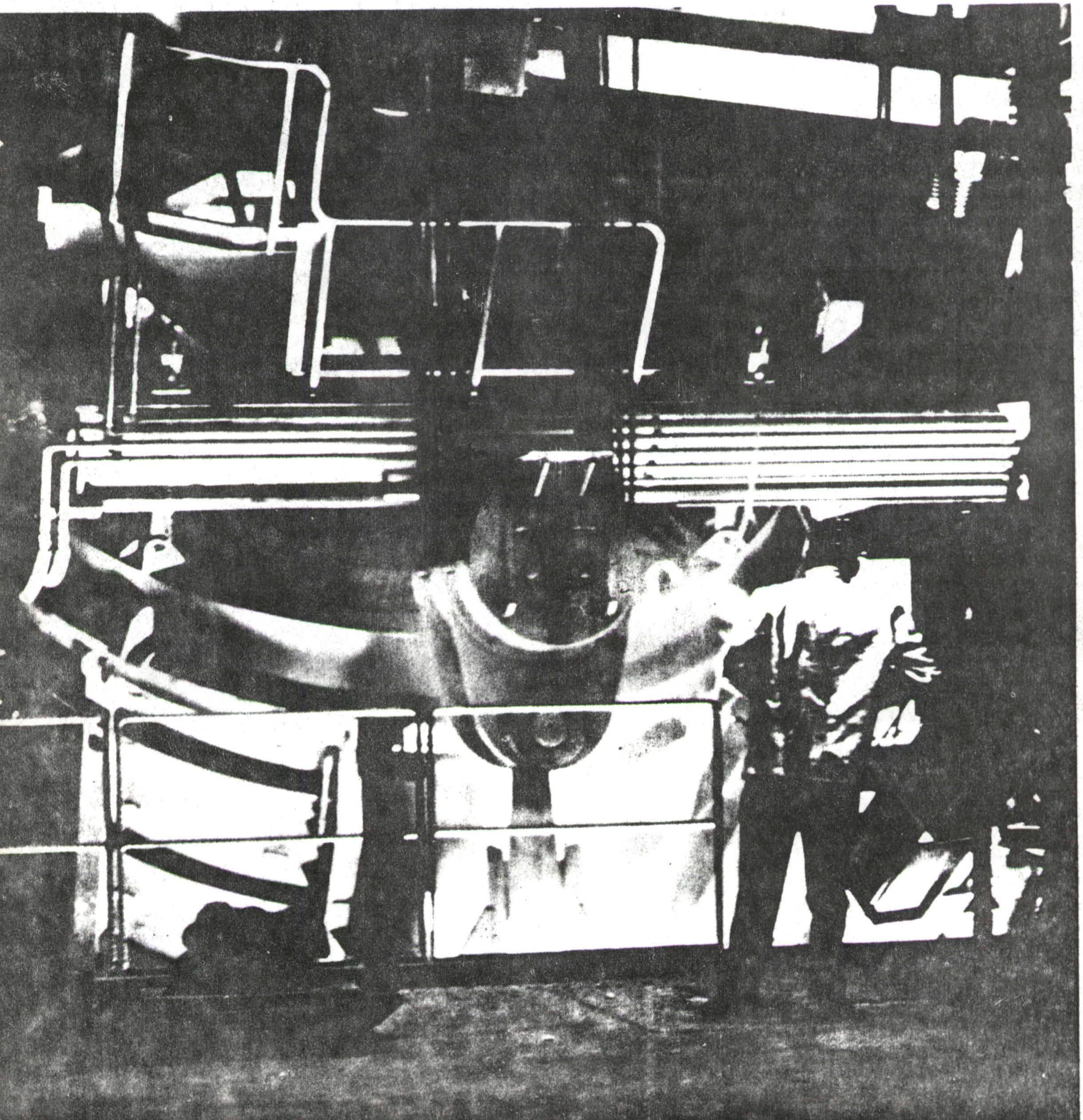
M·R PELLET a new dimension in steelmaking



VISIT 1

MR

The first commercial-scale production of 95 percent metallized pellets
for use in electric steel furnaces . . .
a development of Midrex Division, Midland-Ross Corporation



a new dimension in steel manufacture

The highly metallized pellet has burst onto the scene at the very moment the mini steel plant has begun to make its move. Small steel facilities of 500,000-ton capacity a year and less are increasing steadily in number as sources of specialty steels for regional customers. A significant factor in proliferating this growth will be the highly metallized pellet, providing higher melt rates, increased productivity and a more precisely-controlled end product. Through continuous feed these pellets increase steel furnace output 35 percent or more as compared with conventional scrap feed. By using M•R Pellet a steel producer with three electric arc furnaces can achieve additional output equal to a fourth furnace. Unlike scrap, M•R Pellets are not subject to fluctuations in supply or price. At the same time the producer using M•R Pellets develops a high quality end product suited to plate, rebar, wire and a host of special shapes and sizes.

a new dimension in steel markets

M•R Pellet is expected to become the feed of choice for smaller steelmakers with electric arc furnaces. Conservative estimates —based on scrap as the feed— indicate that electric furnace steelmaking will grow from a current 14 percent of raw steel output in the U.S. to 17 percent by 1975. Midland-Ross, however, foresees the electric furnace share of total production reaching at least 20 percent by 1975—based on the now-commercialized pellet of high metallization. This heightened activity in electric furnace steelmaking is expected to evolve some distance from conventional steelmaking centers —closer to ocean transportation and closer to customers who will benefit from a convenient source of steel.

a new dimension for steel-poor nations and areas

M•R Pellet permits a steel plant of limited size to be constructed virtually anywhere. An underdeveloped country which cannot afford nearly $200 million for a conventional large steelmaking facility may welcome a mini steel plant and pelletizing complex which represents an investment of approximately $35 million. Also, a smaller steelmaking facility is all that is needed to serve markets of limited size. The smaller installation, therefore, permits underdeveloped nations to manufacture steel at a cost and quantity which will stimulate industrial progress, not overwhelm it.

a new dimension for Midland-Ross Corporation

The first commercial-scale installation for highly metallized iron pellets is another forward step in the growth of Midland-Ross. Long a converter of steel into products for varied industries and a leading supplier of components to steel and automotive producers, Midland-Ross for the first time finds itself on the ground floor of steel production. Through its Midrex Division, the corporation plans to own or partially own and operate all M•R Pellet plants. Growth rate of such installations will be accelerated by the growth rate of new electric furnaces. One facet of this picture —particularly gratifying to Midland-Ross—is that the M•R Pellet/electric furnace installation provides a clean operation virtually free of air and water pollution.

Cover. *A single M•R Pellet, polished to highlight its sponge metallic surface, dominates a pile of oxide pellets.*

Left. *Electric furnace at Oregon Steel Mills produces steel directly from M•R Pellets.*

long discussed, now a commercial reality

Oregon Steel Mills, a division of The Gilmore Steel Corporation, has constructed a modern integrated steel mill in the Pacific Northwest. The site is in Portland, Oregon, on the Willamette River. Oregon Steel Mills felt that a metallized ore would provide more efficient production than scrap metal in a small facility equipped with electric furnaces.

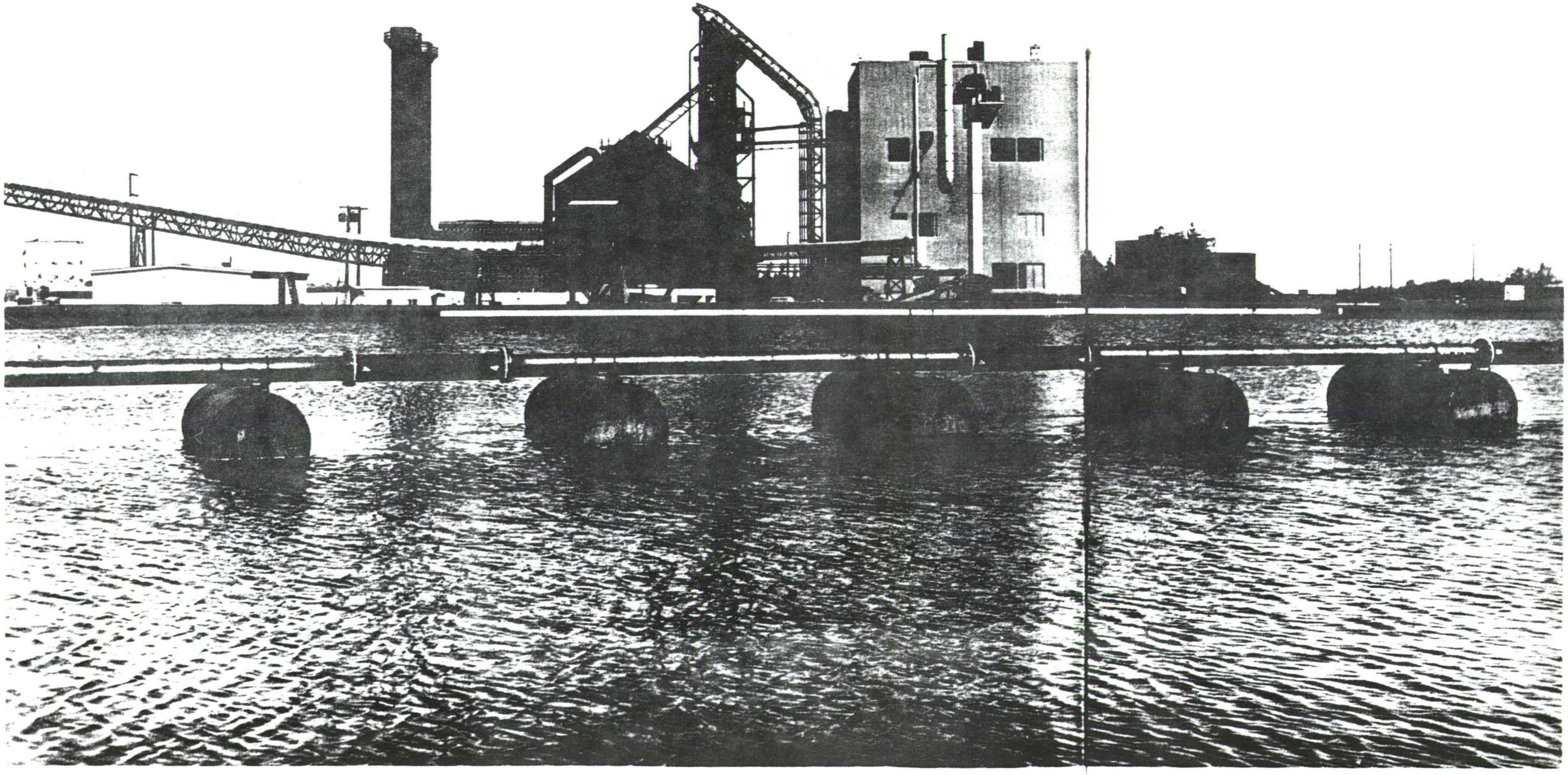
All conditions favored adoption of the Midland-Ross M•R Pellet as the feed. Power and natural gas were readily available. Labor and service facilities were in good supply. Ocean-going ships could navigate the river channels from the Pacific Ocean to a dock alongside the plant site.

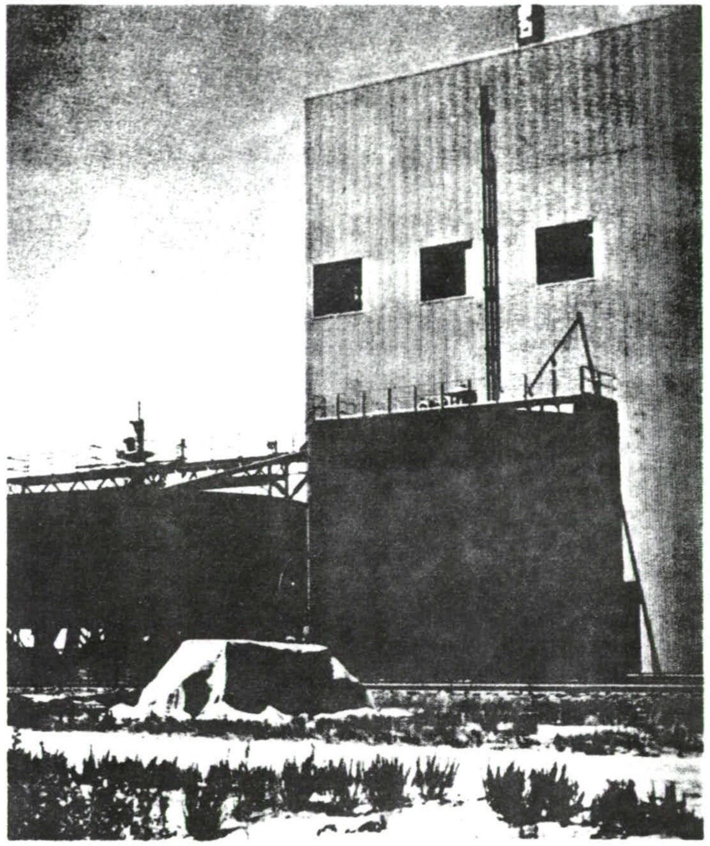
Total cost of mill and pellet plant came to $35 million. Midland-Ross constructed, owns and operates an M•R Pellet facility to supply 400,000 long tons a year. The plant converts high grade magnetite iron ore concentrate (69-70 percent Fe with low gangue) to M•R Pellets (95 percent metallization).

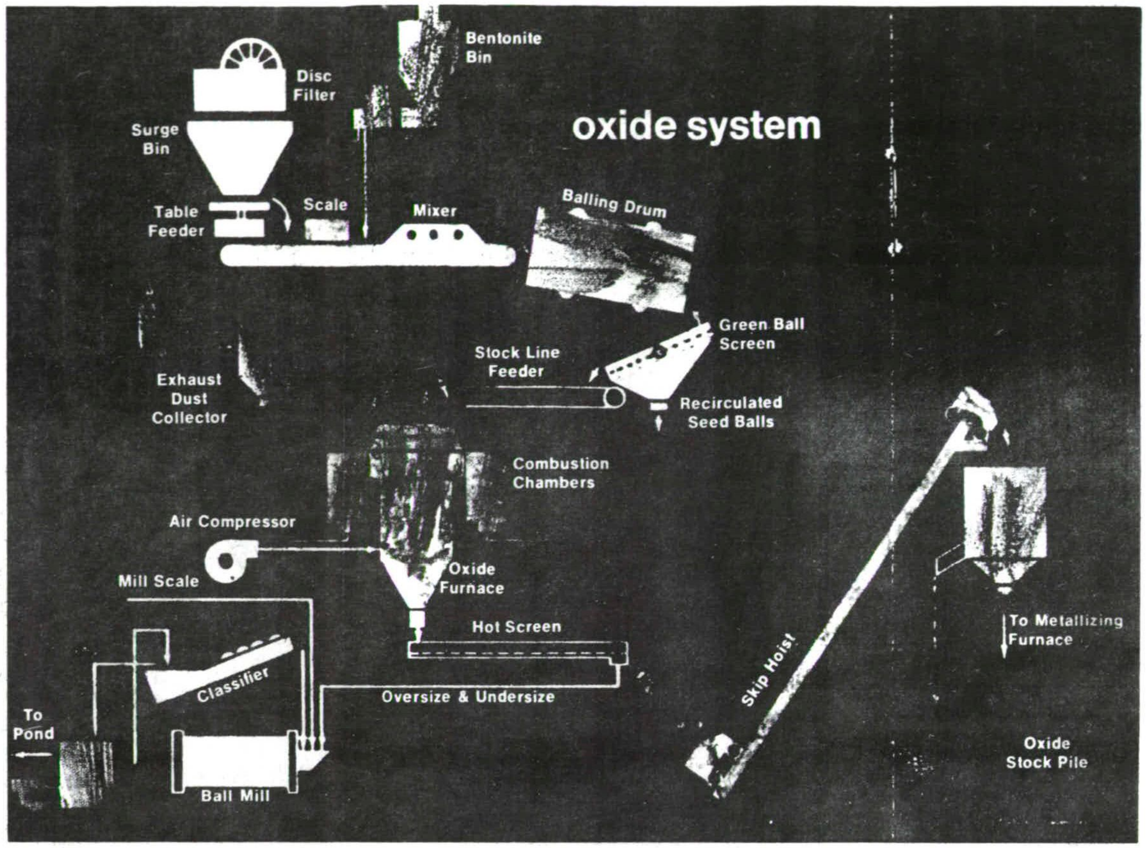
An onsite conveyor runs M•R Pellet continuously to the Oregon Steel Mills plant. Oregon Steel runs metal from the electric furnace into a slab caster which delivers plate up to 100 inches wide and $\frac{3}{16}$ through 3 inches thick.

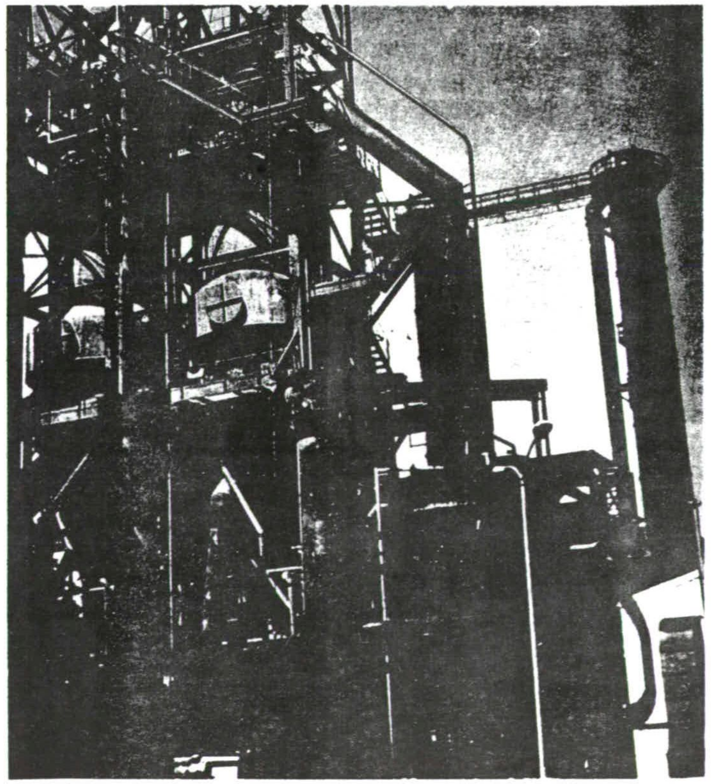
Result is the world's first totally integrated plate mill served by an electric arc furnace—from the iron ore directly to the plate.

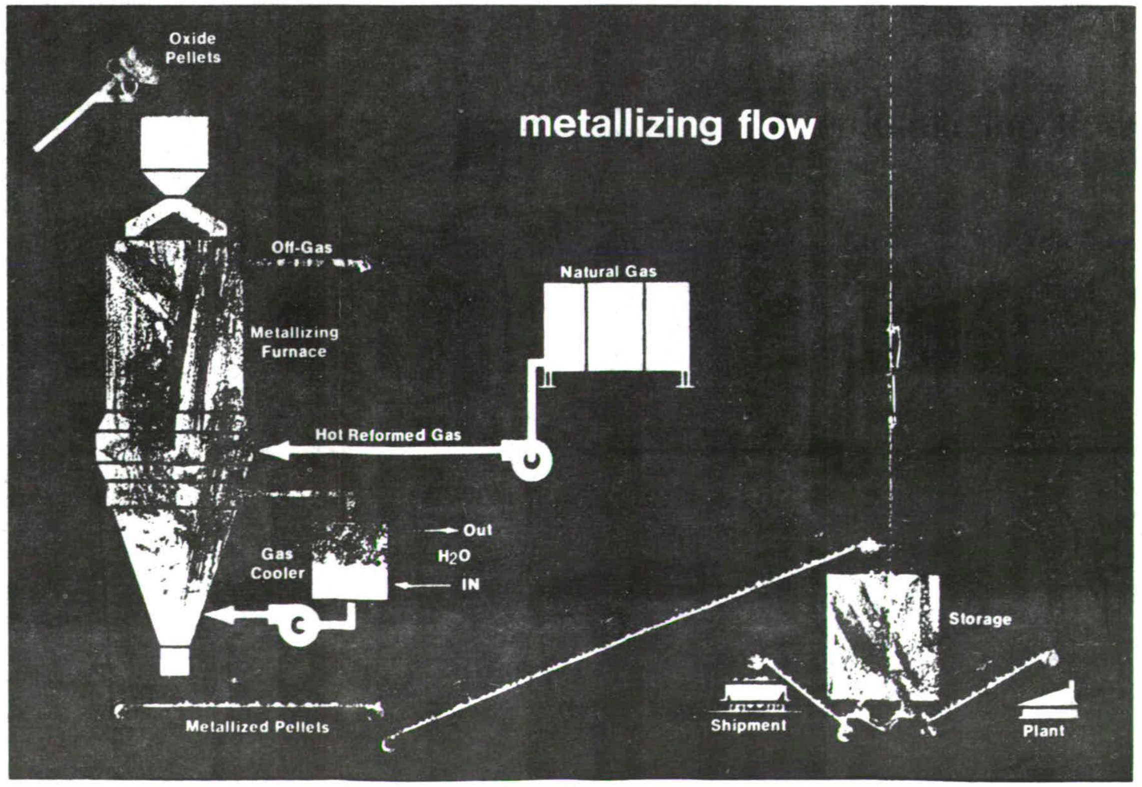
The Midrex pellet plant and slurry pond at Portland.






Bentonite Bin
Disc Filter
Surge Bin
oxide system
Table Feeder
Scale
Mixer
Balling Drum
Green Ball Screen
Stock Line Feeder
Exhaust Dust Collector
Recirculated Seed Balls
Combustion Chambers
Air Compressor
Oxide Furnace
Mill Scale
Hot Screen
To Metallizing Furnace
Classifier
Oversize & Undersize
Skip Hoist
To Pond
Ball Mill
Oxide Stock Pile




Oxide Pellets
metallizing flow
Off-Gas
Natural Gas
Metallizing Furnace
Hot Reformed Gas
Out
H2O
IN
Gas Cooler
Storage
Shipment
Plant
Metallized Pellets

M•R Pellet—a development of Midland-Ross Corporation

The process starts with any high grade iron ore. This installation uses an iron oxide slurry, an unusual feature in its own right. The Midland-Ross M•R Pellet system then requires two steps to produce 95 percent metallized pellets.

step I
an oxide pellet

This begins with a slurry fed to a disc filter which dewaters to a controlled moisture at 64 dry long tons an hour. The filter cake is mixed with bentonite to form nominal ½-inch pellets in a balling drum. This then is the feed to a Midrex vertical shaft furnace which, at 2,400°F., produces hardened iron oxide pellets. The furnace then cools the product following which special handling equipment transfers the oxide pellets to the metallizing operation.

Excess oxide pellets are stockpiled in the yard. For reclaiming they pass over the furnace discharge product screen ahead of the skip hoist serving the metallizing furnace. The oxide pellet operation is an independent system.

step II
reduction to high metal content

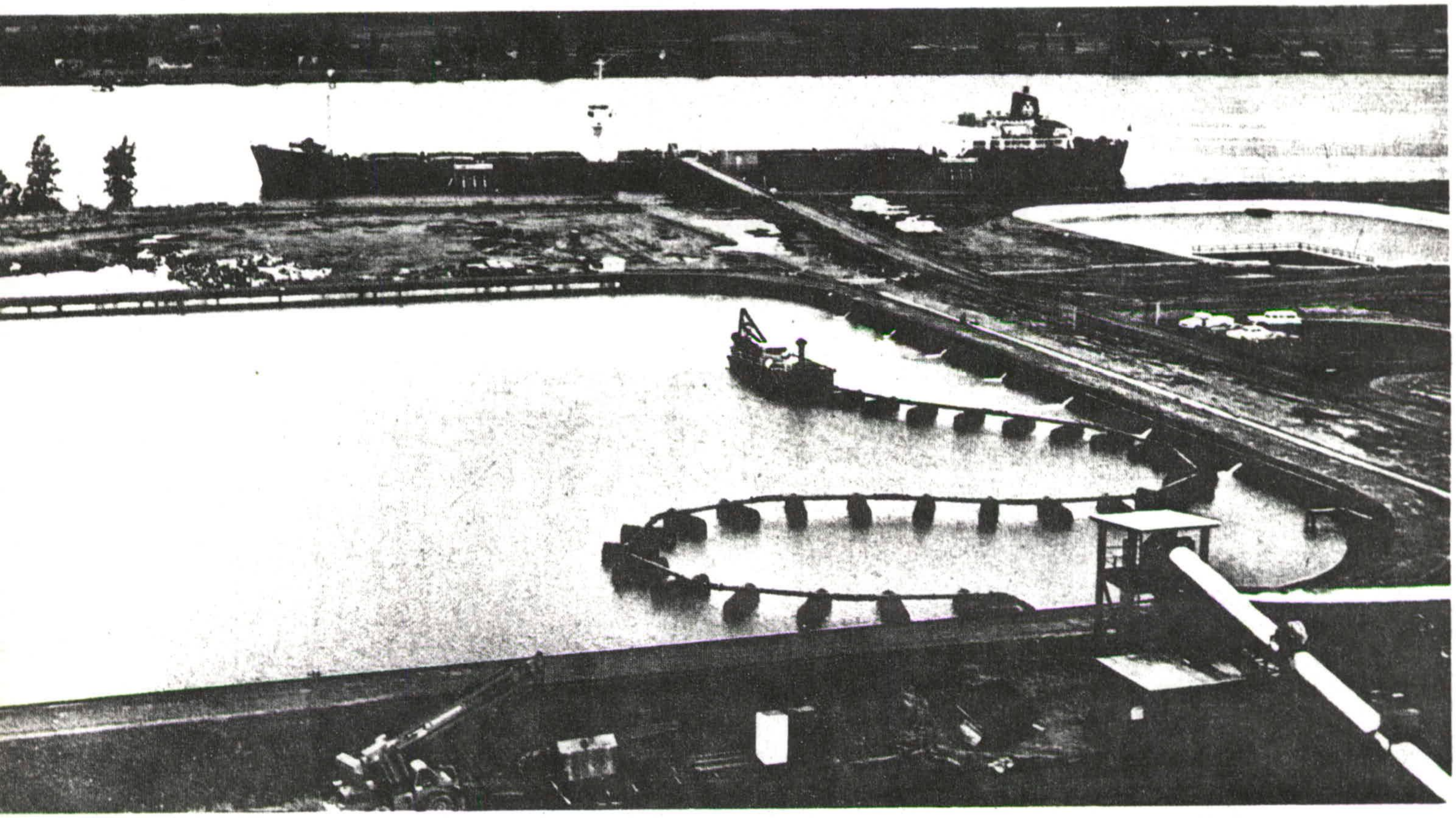
The metallizing operation receives hot oxide pellets directly from process or cooled recirculated pellets from the stockpile. This section, however, is independent of the foregoing oxide production. That is, the versatile metallizing plant can use screened lump, oxide pellets from other sources or input from yet other starting points.

This is an area of significant input from Midrex Division which was able to develop a reformed gas thirsty for the oxygen bonded to the oxidized pellets—at most economical costs under highly efficient conditions. Overall heat consumption in metallizing approximates 13 million B.T.U. for each long ton of metallized product.

Metallizing operations involve two subsystems: a natural gas reformer and the metallizing furnace served by the gas. The gas reformer, developed by Midrex Division, converts various gases to a blend of hydrogen, carbon dioxide, carbon monoxide, water vapor and nitrogen. High pressure blowers jet this hot reformed gas as a counterflow through the bed of pellets descending through the furnace shaft. In the lower section of the furnace the pellets cool to 120°F., well below the temperature of reoxidation.

Final product passes over a scale for inventory tabulation en route to a 5,000-ton holding silo and to Oregon Steel's electric arc furnace. Delivered price of finished M•R Pellet is competitive with scrap prices in this and most other instances.

Midland-Ross designed the M•R Pellet process in modules at fixed capacities so that efficient duplication is available should expansion needs so dictate. The site will permit at least two more modules, requiring no more than increased use of the hydraulic dredge in the slurry pond.



hydraulics applied to ore handling

Economical transportation of the iron oxide to the Oregon Steel Mills site is achieved by an inventive slurry system developed by Marcona Corporation of San Francisco. The system provides the benefits of liquid handling and low cost unloading.

Marcona ships quality ore from its own mine in Peru as a 70 percent minus 325 mesh concentrate. Marcona pumps aboard ship (an ocean-going vessel of 40,000 dry long ton capacity) a slurry at 75 percent solids by weight. As the slurry settles in the ship hold, Marcona decants surplus water, providing a shippable material with less than 8 percent moisture.

Destination is a 10-million-gallon slurry pond on Midrex property adjacent to the Oregon Steel Mills site. Unloading the cargo requires slurry pumps aided by water jets. This system delivers to the pond a slurry at better than 1,000 dry long tons an hour.

An 8-inch suction dredge distributes incoming slurry over the pond to avoid segregation. The dredge tows a flexible pipe from the onboard pumps.

The same dredge, complete with cutter bar assembly at the suction head, removes settled solids from the bottom of the pond. This supplies a 50-ft.-diameter thickener tank with a slurry of 33 percent solids by weight. The tank system increases density of the slurry to about 70 percent solids. This then is pumped to a 40-ft.-diameter storage tank. Solids remain in suspension because of a powerful turbo blade agitator in the holding tank. From here pipes carry the concentrate at regulated rates into the initial stages of M•R Pellet production.

Recirculated water returns to the slurry pond for cooling and clarification before discharge to a settling area for natural clarification. This properly reconditions the water before it returns to the river.

Reground fines (screened from the oxide product), dust collection slurries and a small quantity of furnace fume from the steel mill itself are all returned to the pond by a recirculation system. This eliminates pollution.

M•R Pellet—a new process for a changing world

The M•R Pellet process is expected to alter the shape of the steel industry throughout the world. Emerging nations and steel-hungry areas are beginning to embrace a steelmaking capability which requires modest capital investment. The use of M•R Pellets will substantially improve the return on investment for electric furnace steelmaking. And this is highly significant because, even now, electric furnace steelmaking with conventional feeds is recording the best returns on investment in the steel industry. Least impact from the new process will be felt by the major steel producers who already have huge investments in conventional facilities.

interest worldwide

The technology of the small plant is a reality today with nearly 50 facilities in the U.S. and others scattered about the world. Small steel plants may be found in Canada, Mexico, Japan and the European Common Market. Other areas expressing interest include Northern Ireland, Venezuela, Mozambique, Tanzania and Rhodesia, to pinpoint a few. Limited markets and limited finances make these countries ideal locations for electric furnace steel production. For example, a basic oxygen furnace complex requires output on the order of 2 million tons a year to pay off. An installation with blast furnace, BOF, coke ovens and sinter plant might cost $200 million. The same tonnage from a series of electric furnaces would require an investment of only $70 million.

location consideration

The small steel plant with highly metallized pellet feed ideally should be convenient to water transportation. Any country today can shop the world markets for ore, carried in increasingly sizeable ships. Currently 25 percent of the world iron ore travels by sea and this is expected to jump to 40 percent within the ensuing decade.

increased output

The electric steel furnace achieves at least 35 percent more output with M•R Pellet feed than it does with scrap in a given time period. Some major reasons:
The furnace feed hopper is continually replenished with pellets. The sponge-like M•R Pellets melt down at unmatched speed.

reduced post-processing

The M•R Pellet minimizes the time needed to refine hot metal for either rebar or plate. Scrap feed requires considerable refining to achieve the quality which M•R Pellet produces right from the pour.

another major advantage

The M•R Pellet/electric furnace complex operates virtually free of pollution. Midrex reports no gaseous effluent from the pellet preparation process. At the furnace, the clean charging characteristics of M•R Pellet help by generating considerably less fume than scrap does. As to liquid waste, the slurry pond—360 x 400 ft. and 15 ft. deep—serves as a cooling, clarification and settling basin. Much of the water is reused time and again. The excess, however, drains back into the river pure as when drawn.

Midland-Ross — a new source of iron for steelmaking

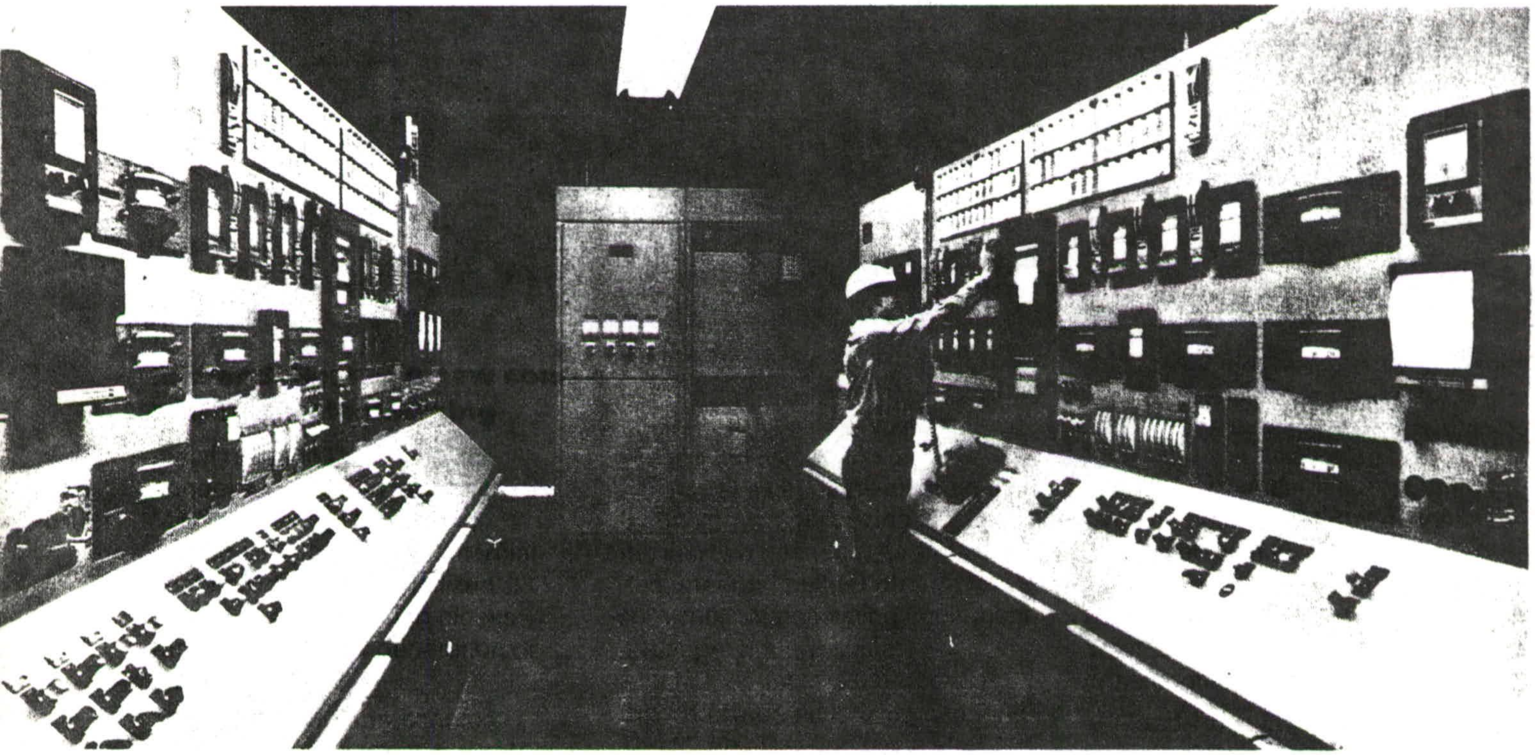
Initial reaction to M•R Pellet augurs a significant future for Midland-Ross in the steel industry. As the first commercial-scale producer of metallized pellets, the corporation closes ranks with the new generation of steelmakers.

The Midrex Division enables Midland-Ross to retain control over the M•R Pellet process. This guarantees that the quality designed into the process will be evident in each pellet. In preparing for its role in assisting the steelmaker, Midland-Ross has developed a capability with the interfaces at both ends of the M•R Pellet installation—raw material into the system and the finished pellet into the furnace. The pellet plant fits smoothly into the total steelmaking facility.

Such involvement for Midland-Ross is not so sudden as it seems. It has been developing a technology in pelletizing for the last two decades. As a major designer and builder of conventional pelletizing equipment, it has provided installations which account for 20 percent of the worldwide pelletizing capacity. The Midrex Division has a staff of wide plant experience—specialists in electrical, mechanical and civil engineering supported by teams of construction and purchasing authorities. Therefore, Midland-Ross is qualified to design and build complete systems for processing metals or portions of systems as the need may be.

Now Midland-Ross has expanded its role in steelmaking—through its M•R Pellet process and the Midrex Division. The need was there and will take years to begin to satisfy. In anticipation of the overseas market, Midland-Ross has joined with Korf Industrie Und Handel GmbH to produce and sell metallized pellets and specialized steel products for European consumption.

Midland-Ross is justifiably excited over the future of the mini steel industry and its own role as a significant participant. The corporation provided an important contribution to basic steelmaking with its conventional pelletizing system. Now, with M•R Pellet as a major product, Midland-Ross expects this business will represent an increasingly substantial portion of corporate sales in ensuing years.



SURFACE COMBUSTION DIVISION MIDLAND-ROSS CORPORATION

Inter-Office Correspondence

To Mr C.W. Sanzenbacher - Prod. Engr.
From J.B. Allyn - Product Engineering
Date December 6, 1967
Subject M-R Pellet Melting Test at Oregon Steel Mills

CC: Mr D. Beggs - Engineering
Dr L.W. Smith - Engineering
Mr T.A. Blanchard - R & D
Mr H.G. McVeigh - Minerals Proc
Mr W.A. Ahrendt - R & D
Mr D.C. Meissner - R & D
Mr J.E. Bonestell - MP Admin.
Mr R.E. Kranz - Minerals Proc.

On December 1, at the Oregon Steel plant, an electric arc furnace steelmaking test was performed. The purpose was to test the use of M-R pellets to find any difficulties that might occur in the future. The furnace used was of 22 tons capacity, used hollow graphite electrodes, had a power input of 10,000 KW, and was made by Lectromelt. The initial charge consisted of 7 tons of sized angle scrap, 5 tons of M-R pellets, and 500 lbs. lime. After melt-down, another 10 tons of pellets were charged continuously through the furnace roof at a rate of 700 to 850 lbs. per minute. The average pellet composition was 95.6% total iron, 96.3% metallization, and 1.4% carbon.

SUMMARY OF TEST RESULTS:

The test was successful. Initial melt-down was accomplished in about 35 minutes. Pellets were continuously fed for another 40 minutes. The pellets had few chips or fines, fed smoothly and well, and were absorbed easily during melting. The power-on-to-tap time, less delays, was 1 hour 16 minutes, a new record for the furnace. Despite the 1.4% average pellet carbon level, only a 3 minute oxygen blow was needed to bring the melt to a correct carbon level. An 11 minute blow was originally anticipated. Slag weight was low, the yield was 98.5%, and the product had very low sulphur and phosphorus. The product was in all respects acceptable for further processing.

Based on this test, M-R pellets appear suitable for continuous-charge steelmaking, provided that the carbon level is lower, perhaps between 0.5% and 1.0%.

TEST DESCRIPTION:

Tests were performed using the above described furnace, plus a continuous feed system designed jointly by Oregon Steel and Steel Company of Canada. This system allows the accurate weighing, feeding, and distribution of pellets from a storage bin to three feed pipes. Each of the pipes then directs pellet flow to a hole in the roof of the furnace. See attached sketch. The pellets then free fall through the hole, into the slag slayer, and slowly sink through into the melt. The tests were performed by regular Oregon melt shop personnel, with Oregon R & D people acting as advisors and observers.

The tests were well planned. Thursday night, a test run was made using S-L pellets and several malfunctions were spotted. The calibration of the pellet weighing system was checked and furnace operation at 10,000 kW was tried with no ill effects (normal operation is at 7,500 kW). Also on Thursday, a meeting was held to outline test procedure. A copy of the procedure is attached.

On Friday morning, at 9:30, melting of the initial charge commenced. Several delays occurred, all of a minor nature. Furnace power was set at 10,000 kW at the start, with 150 volts electrode potential. Typical erratic scrap melting operation was observed during melt-down, with violent arcing, power shifts, and much noise and smoke. Such operation tripped the power supply several times so the limit was set somewhat higher. No further difficulty occurred. A temperature and carbon test was made just before melting was complete. The temperature was about 2980°F, with 0.31%C. About 0.50%C was expected.

After melt-down was completed, pellet feeding was started at 700 lbs per minute. Feeding was uniform with few chips or fines. A few coal black pellets were noted. Eight minutes later, the bath temperature was up to 3100°F whereas about 2950°F was desired. The pellet feed rate was increased to 850 lbs per minute so as to lower the temperature. Eight minutes later the temperature was down to 2970°F so the feed rate was cut back to 750 lbs per minute. It was interesting to note that at 850 lbs per minute the pellets were piling up inside the furnace. Apparently, for the conditions, the upper feed limit had been passed. The test continued at 750 lbs per minute, with power fluctuations from 9700 to 10300 kW occurring during continuous feeding.

Further carbon and temperature determinations were made during the duration of the test. Carbon rose to 0.375% whereas 0.66% had been expected. Carbon boil was evident during continuous charging, which is beneficial.

At the end of the continuous charging period, an oxygen blow was used to lower the carbon level. Oxygen was blown at a rate of 30,000 CFH, for 4 minutes. Much smoke and flame occurred during this time, which is typical.

A slag raking operation then ensued, with the slag being of proper viscosity. Further lime was added, along with some ferrosilicon and ferromanganese. The lime was not really necessary but the operating crew was following standard procedure. The slag was then raked off. Slag viscosity appeared correct. Later tests showed it to be slightly basic because of the extra lime. The basicity ratio was 1.05. Included FeO was 8%. Total slag was 3,300 lbs but Oregon anticipates that only 2,500 lbs will occur if an updated practice is used.

The heat was then poured and later weighed. Ingot weight was 22.2 tons which was expected. Electrode loss was 350 lbs total, or 15.7 lbs per ton tapped. The yield, on the basis of metal tapped to iron charged, was 98.5%. Total power used was 513 kW per ton, but if delays and extra operations are deducted, a 480 kW per ton figure results. The ingots were acceptable in every respect.

DISCUSSION OF RESULTS

Test results were expected except for the unaccountably low carbon level that resulted. So far no one has explained this. Theoretically, 1% carbon pellets will reduce all the FeO present and will give a 0.3% carbon heat. The 1.4% carbon pellets used should have given 0.66% carbon at the end of the heat but only 0.375% occurred. Further tests are needed to determine if this was just a fluke or a real effect.

The tests proved that a 10,000 kW input is feasible for a continuously-charged 22 ton furnace. However, further increases may be limited by the ability of the slag to accept a higher rate of pellet flow.

It is interesting to compare the test results with those using the Stelco-Lurgi pellets. Oregon has tested S-L pellets many times using the continuous charging technique. M-R pellets differed from S-L pellets in the following ways:

1. Less slag - A typical S-L slag weight is about 4,000 lbs for a 22 ton heat, whereas 2,500 lbs, is expected for M-R pellets. This means a savings in power of about 30 kWh per ton.
2. Lower sulfur and phosphorus - The test heat had 0.012% S and 0.015% P Lurgi pellets typically give 0.02 - 0.03% S and 0.01 - 0.03% P.
3. A more uniformly sized feed - Lurgi pellets vary from ¼" to over 1" in diameter.

4. A higher carbon level - Lurgi pellets typically have 0.2% C, which requires carbon additions during a heat. M-R pellets in the 0.5 - 1% C range will provide enough carbon for the heat and will give a continuous strong carbon boiling action for good purification.

5. Both M-R and Lurgi pellets have about the same metallization, about 95 - 96%. As a consequence, they both have from 2 - 3% FeO, which is desirable to promote a carbon boil during the heat.

A comparison of the tested M-R pellets operation with normal scrap melting practice is also interesting.

M-R Pellets	Scrap
Charge to tap time - 1 hour 16 minutes	1 hour 45 minutes to 2 hours
kWh per ton output - 480	480 - 520
Furnace output - 17.5 tons per hour	11 to 12.5 tons per hour
Electrode consumption - 15.7 lbs per ton	12 lbs per ton

It is expected that electrode consumption will be about the same for pellets or scrap operation. Several pellet tests are needed to establish a valid consumption figure. 70% S-L pellet operation normally has a consumption of 12 lbs per ton for hollow electrodes.

A representative from Stelco was present during the tests, Dr Gordon Forward. He was invited by Oregon because of the past help they have received from Stelco in acquiring pellet melting technology. Dr Forward did not have any particular comments to make on the test results.

CONCLUSION AND COMMENTS:

On the basis of one test only, M-R pellets in the 0.5 - 1.0% carbon range appear to be suitable for continuous charge steelmaking.

Oregon is dealing with Lectromelt for a new 65 ton arc furnace for the new plant. Oregon is designing the continuous feed mechanism. They expect to get at least 30 tons per hour production from this furnace. For typical scrap operation, only 16 to 20 tons per hour could be expected. The use of reduced pellets as a continuous feed for arc furnaces can certainly be called a breakthrough in steelmaking.

J.B. Allyn

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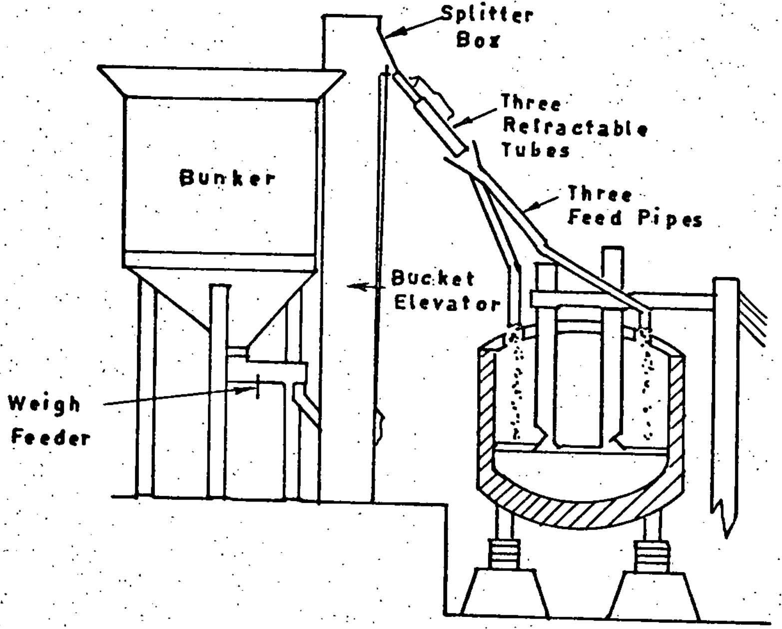

FIGURE A – 2: THE CONTINUOUS CHARGING APPARATUS

Melting Prereduced Iron-Ore Pellets in the Cupola

by R. H. Hafner, Jr., Rsch. Met. and S. C. Clow, Tech. Director
Clow Corp., Coshocton, O.

ABSTRACT

An investigation was made to determine whether prereduced pellets could be successfully melted in a cupola and to study their influence on the quality of the resulting gray and ductile iron. Heats conducted in a 60-in. experimental cupola, in which steel scrap had been replaced by increasing amounts of prereduced pellets, demonstrated that charges containing up to 70% prereduced material could be melted in an acid cupola. At this level the average blast pressure increased from 14 oz to 19 oz with a momentary peak of 24 oz. Additional coke was required to slag off the 6-8% siliceous gangue in the pellets. Carbon and silicon values were not appreciably affected provided the ferrosilicon and coke charges were adjusted at the beginning and end of each heat to compensate for the segregation of the pellets within the cupola. The percentage of tramp elements except titanium appeared to be almost complete and no increase in the iron oxide content of the slag could be found. Two full scale heats in a 108-in. production cupola confirmed the results obtained in the experimental cupola. Using up to 40% prereduced pellets resulted in an increase in the average blast pressure from 22 to 25 oz. Although the economics of using pellets has not been established, realistic estimates place their future cost somewhere between the cost of foundry steel and pig iron.

Introduction

To keep abreast of changing foundry technology and to prepare for the challenges of the future, today's progressive foundryman must be constantly seeking new and improved methods and materials. One of the new materials appearing on the horizon, certain to be of future interest to many producers of quality ductile, malleable and gray iron castings, are prereduced iron-ore pellets.

Prereduced pellets result from the direct reduction of iron-ore pellets to metallic iron pellets without an intermediate melting step.[1] Their development was motivated by a desire to find a means of increasing blast furnace melting rates and to provide potential melting stock for the electric steel furnace. The pellets (Fig. 1) are $\frac{3}{8}$ to $\frac{5}{8}$ in. diameter; however, a small percentage will fall between $\frac{3}{8}$ and $\frac{3}{16}$ in. This is not the most desirable size to charge into a cupola but it is the optimum size for efficient production of high-grade, reduced pellets by the SL/RN process used to produce the pellets for this investigation.[2] The pellets are magnetic and can be handled conveniently by a crane and magnet. Their size and shape also make them ideal for mechanical handling by an automated storage and charging system. The pellets are not friable and can withstand the roughest treatment without creation of fines. Because the reduced pellets can be reoxidized, they must be shipped and stored under cover.

Those considerations that make prereduced pellets attractive as cupola melting stock are:

1. A source of virgin metal with a low "tramp" element content.
2. The chemical composition is known and is consistent.
3. The availability and price should be stable once commercial production becomes established. Known reserves of iron ore will insure an adequate supply of ore of known composition for an extended period.
4. A lower cupola-charge cost might be possible depending on price differentials between pellets and competing charge materials.

Attracted by the potential of prereduced pellets as melting stock for producing ductile iron, the authors' company conducted a research program in 1966 to determine the effect of prereduced iron-ore pellets on cupola operations and to study the influence of pellets on the quality of the resulting gray and ductile iron. Exploratory gray and ductile iron heats utilizing pellets were conducted in an experimental cupola capable of melting nine tons per hour. After determining that it was practical to melt pellets in the small cupola, two ductile iron heats were melted in a production cupola that normally melts 28 tons per hour.

Three types of prereduced material produced in a pilot plant were obtained for evaluation.

1. "A" pellets produced from virgin ore.

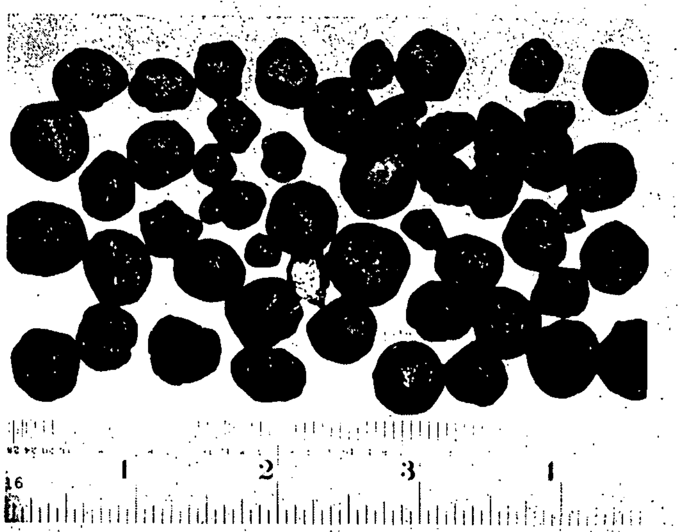

Fig. 1. Prereduced iron ore pellets.

2. "B" pellets made from the tailings of a prior extraction process.

3. Fines consisting of improperly formed or broken "B" pellets largely resulting from reduction disintegration and obtained when passing the reduced product over a 3/16 in. mesh screen. A screen analysis of the fines was as follows:

31.6% whole pellets
56.8% broken pellets—6 to 20 mesh
11.6% fines under 20 mesh

Because a small percentage of fines is unavoidably produced during the manufacture of pellets, they were of interest as a possible source of low-cost material that could be injected into the cupola with a borings injector.

The chemical analyses of the pellets used in the investigation are listed in Table 1. The information in the table can be grouped into four areas of major importance.

1. The metallic iron content is dependent on the efficiency of the reduction process and the amount of gangue present. A good grade of pellets should contain a minimum of 92% metallic iron.

2. The Fe^{++} represents unreduced iron oxide and should be as low as possible. Since a cupola is not a smelter, the iron oxide will normally be lost in the slag.

3. The metallic oxides comprise the gangue portion of the pellets. Since beneficiation of ore does not provide 100% separation, the pellets contained 6 to 8% gangue. At a basicity of 0.35 to 0.57, the gangue is self-fluxing and will combine with the regular cupola slag during melting in an acid cupola.

4. The residual elements with the exception of carbon and sulfur are controlled by selection of the ore and will be very low.

TABLE 1. Chemical Analyses of Prereduced Pellets.

	"A" Pellets	"B" Pellets	"B" Pellet Fines
Total Iron	94.10%	93.40%	89.30%
Metallic Iron	92.20	92.20	87.40
Fe ++	0.80	0.4	0.80
SiO_2	2.64	3.38	4.38
Al_2O_3	0.83	1.05	1.91
CaO	0.30	0.70	1.60
MgO	1.68	0.86	1.58
TiO_2	0.05 to 0.10	0.22	----
S	0.014	0.012	0.059
C	0.068	0.250	0.395
Ni	---	0.16	---
P	0.023	0.018	---
Mn	N.D.	0.007	---
Cr	N.D.	0.014	---
V	N.D.	N.D.	---
Mo	N.D..	N.D.	---
Cu	N.D.	N.D.	---
Al	0.006	0.008	---
Mg	N.D.	N.D.	---
Sn	0.022	0.012	---
Ti	Tr.	N.D.	---
Pb	N.D.	N.D.	---
As	0.012	0.003	---

Description of Experimental Cupola Heats

Three experimental heats designed to study the effects of adding pellets to a cupola charge were conducted in a 60-in. experimental cupola lined down to 40-in. with an acid refractory. The cupola was equipped with a 1000 F (538 C) hot blast and the necessary instrumentation for the measurement and control of the cupola operation. The cupola was tapped continuously into a 5-ton forehearth. Slag was removed by a rear siphon slag spout.

A 1½-ton transfer ladle placed on a platform scale in front of the forehearth permitted the metal poured from the forehearth to be weighed each time the transfer ladle was filled. Spout temperatures were measured at 15-minute intervals and monitored continuously with a recording radiation pyrometer. Samples for spectrographic analysis were poured from the spout each 15 minutes and from the forehearth each time the transfer ladle was filled. The three heats were primarily gray iron, although 5 ladles of ductile iron were made by the open-ladle treatment during the heat of May 20th. Each of the heats were of 7 to 8 hours duration during which period the hot metal was used for production of centrifugally cast pipe.

A borings injector was installed in the area of maximum lining burnout 24 in. above the tuyeres to determine the possibility of injecting pellets or pellet-fines directly into the melting zone.

Details of the cupola operations are listed in Table 2. The cupola charges listed in Table 3 were devised to determine the effects of varying the percentage of pellets in the charge on the operation of the cupola. During the first heat on April 30th, the pellets were weighed on a platform scale, then dumped into the cone-bottom charging bucket on top of the balance of the charge. During the second and third heats, the pellets were stored in an overhead hopper normally

TABLE 2. Cupola Operating Data.

	Experimental Heats			Production Heats	
	April 30	May 13	May 20	Sept. 28	Oct. 5
Shell Diameter	60 in.	60 in.	60 in.	108 in.	108 in.
Lining Diameter	40 in.	40 in.	40 in.	72 in.	72 in.
Hot Blast Temperature	1000F	1000F	1000F	760F	700F
Blast Rate – Lb Min.	275 & 325	300	300	740 to 840	740 to 840
Nominal Metal Charge-Lb	1500	1500	1500	5000	5000
Total Metallics Charged-Tons	64.1	73.1	58.9	234.2	234.4
Melting Time – Hours	7.0	8.42	7.25	8.92	9.17
Melting Rate – Tons/hour	9.16	8.68	8.12	26.2	25.6
Average Spout Temperature	2760F	2840F	2840F	2880F	2890F
Average Spout Analysis					
Carbon	3.26%	3.63%	3.69%	3.36%	3.43%
Silicon	1.50	1.92	1.54	1.41	1.15
Sulfur	0.074	0.073	0.073	0.104	0.098

used for limestone and fed into the coke-stone weigh-hopper by a vibrating feeder. After weighing, the pellets were discharged into the charging bucket on top of the balance of the charge.

The pellet-fines were injected into the cupola by the borings injector during the second heat on May 13th. Feed rate was determined by an automatic timer set to inject fines at a rate estimated to be 10% of the total metal charged. Difficulty was experienced with plugging of the discharge end of the injection tube due to the slow melting rate of the pellets and to molten slag and iron droplets flowing down the cupola lining and dropping inside the bottom lip of the tube. Use of the injector was abandoned after 4½ hours due to stoppage of the injection tube.

During the third heat, pellet fines were weighed on a platform scale and then added to the balance of the charge in the charging bucket.

Results of Experimental Heats

Blast Pressure

Although it was expected that increasing amounts of pellets would ultimately result in prohibitive windbox pressures due to the decreasing permeability of the charge, the actual increase was not excessive. The ranges of pressures obtained at different percentages of pellets are listed in Table 4. An increase of pellets up to 50% of the charge resulted in an average pressure of 15 oz with a momentary peak of 17 oz. This pressure is within the operating range of most cupola blower installations and would not be considered a significant pressure increase.

The injection of 12% pellet-fines directly into the melting zone with the borings injector produced no significant increase in windbox pressure. The addition of 15% fines together with 45% pellets on May 20th,

TABLE 3. Cupola Charges.

Date of Heat	Number of Charges	Coke Ratio	Ductile Returns	Steel Scrap	Foundry Pig	Silvery Pig	Ferro-Silicon	Pellets	Pellet Fines
					Experimental Heats				
Apr. '30	14	8.4/1	–0–	1,000	500	100	–0–	–0–	–0–
	20	7.7/1	–0–	900	450	112	–0–	160	–0–
	20	6.6/1	–0–	800	400	125	–0–	320	–0–
	19	6.5/1	–0–	700	350	137	–0–	480	–0–
	7	6.6/1	–0–	600	300	150	–0–	640	–0–
May 13	11	6.3/1	750	750	–0–	75	–0–	–0–	–0–
	20	6.4/1	750	600	–0–	75	–0–	–0–	205
	20	5.7/1	750	300	–0–	75	–0–	160	175
	15	5.8/1	750	150	–0–	75	–0–	380	135
	11	5.1/1	750	–0–	–0–	75	–0–	480	–0–
	3	6.3/1	750	–0–	–0–	75	–0–	800	–0–
May 20	20	6.0/1	750	750	–0–	12	–0–	–0–	–0–
	30	6.0/1	750	–0–	–0–	12	–0–	550	250
	16	6.3/1	600	–0–	–0–	98	–0–	710	250
	6	6.0/1	450	–0–	–0–	–0–	36	870	250
					Production Heats				
Sept. 28	21	7.0/1	750	4,250	–0–	–0–	90	–0–	–0–
	2	5.9/1	750	4,250	–0–	–0–	120	–0–	–0–
	24	6.6/1	750	3,250	–0–	–0–	90	1,090	–0–
	2	6.0/1	750	3,250	–0–	–0–	105	1,090	–0–
	31	6.4/1	750	2,750	–0–	–0–	92	1,635	–0–
	12	6.4/1	750	2,750	–0–	–0–	84	1,635	–0–
Oct. 5	24	7.0/1	750	4,250	–0–	–0–	90	–0–	–0–
	2	5.4/1	750	4,250	–0–	–0–	135	–0–	–0–
	19	6.4/1	750	2,750	–0–	–0–	92	1,635	–0–
	2	5.9/1	750	2,750	–0–	–0–	107	1,635	–0–
	45	6.2/1	750	2,250	–0–	–0–	96	2,180	–0–

TABLE 4. Effect of Pellets on Wind Box Pressure Experimental Cupola.

Date	Charge	Blast Rate Lb/Min.	Pressure – Oz Low	Ave.	High
April 30	Base Mix	275	5	6	7
	10% Pellets	275	6	7	9
	20% Pellets	325	7	11	14
	30% Pellets	325	9	11	13
	40% Pellets	275	5	7	10
May 13	Base Mix	275	10	12	13
	12.5% Fines	300	7	11	15
	12.5% Fines + 10% Pellets	300	11	13	16
	9.9% Fines + 22% Pellets	300	11	13	18
	35% Pellets	300	13	13	14
	50% Pellets	300	14	15	17
May 20	Base Mix	300	10	14	17
	15% Fines + 35% Pellets	300	15	19	23
	15% Fines + 45% Pellets	300	15	19	24
	15% Fines + 55% Pellets	300	10	15	20

increased the average pressure from 14 oz to 19 oz with a transient peak of 24 oz. This is a significant increase in pressure but one that could be handled by a blower of sufficient reserve capacity. It is not recommended that pellet-fines be used as top charge, cupola material because of their effect on blast pressure. A further disadvantage is that a large amount of fines are discharged with the effluent stack gas.

Metal-Coke Ratio

Because the pellets and pellet-fines contain gangue, additional coke is required to melt this slag-forming constituent. The first heat on April 30th began with a metal-to-coke ratio of 8.4/1. The average spout temperature of 2760 F (1516 C) was considerably below the desired tapping temperature of 2850 F (1566 C) required for satisfactory carbon pickup. The ratio in the subsequent heats was reduced to 6/1 on May 13th and 6.3/1 on May 20th. The average tapping temperature during the latter two heats was 2840 F (1560 C). No attempt was made to determine the optimum coke ratio but past experience with the cupola indicated that a ratio of 8/1 should result in an average metal temperature of 2850 F (1566 C). It appeared that the melting of 50 to 60% pellets required an additional 30 to 35% coke.

Metal Yield

The weight of metal tapped exceeded the weight of metal charged. After making material balances to account for carbon pickup and other changes in the chemical analysis, the calculated yields were slightly in excess of 100%.

April 30	102.7%
May 13	101.1%
May 20	100.4%

The FeO and MnO lost in the slag were not determined as the slag was not weighed. Correction for these losses in the material balances would undoubtedly reduce the yields to less than 100%. It does appear that the metallics in the pellets were substantially recovered.

Cupola Slag

The gangue was self-fluxing and after combining with the normal cupola slag, flowed freely from the cupola. A comparison of the gangue with cupola slag from a pellet-free reference heat is as follows:

	Reference Slag	"A" Pellets	"B" Pellets	Fines	Normal Range
SiO_2	46%	48.5%	56.5%	46.4%	40/50
Al_2O_3	15	15.2	17.5	20.2	10/20
CaO	34	36.4	26.0	33.6	25/38
MgO					
Basicity	0.56	0.57	0.35	0.51	0.5, 0.9

A chemical analysis of slag samples taken at ½ hour intervals indicated that the slag compositions fell within the range of normal acid slags. Increasing the amounts of pellets in the charge did not appreciably alter the slag composition; however, the slag volume was increased.

Because the pellets and fines contained from 1.2% to 1.9% Fe^{++} as unreduced iron oxide, an appreciable increase in the FeO content of the slag was expected. It was found, however, that the pellets produced no measurable increase in FeO.

% Pellets	Number of Samples	% FeO
0	4	2.07
10	5	2.07
20	2	1.42
30	5	1.92
40	3	5.94°
50	7	2.02
60	2	1.93
70	1	1.91

°Low bed and spout temperature at end of heat.

Metal Composition

The silicon analysis of each ladle of iron taken from the forehearth during the heat of May 20th is plotted in Fig. 2. It should be noted that a significant, temporary drop in silicon occurred each time the percentage of prereduced material was increased. This occurrence was definite and could be detected in all three heats each time a change was made in the percentage of pellets charged. The silicon drop was especially pronounced when the charge was changed from a base mix to 50% prereduced material. It is speculated that the pellets, because of their size and shape, dropped through the open charge and melted several charges in advance of the charge to which they were originally added. This temporarily diluted the metal in the well with respect to silicon and upset the metal-to-coke ratio. Eventually a balance was restored as the scrap and ferrosilicon portion of the charge "caught up" and the silicon was returned to a normal level. The charges remaining to be melted during the last 20 to 30 minutes of a heat became increasingly depleted in pellets. As a result, these charges contained increasing proportions of ferrosilicon and scrap and thus became increasingly higher in silicon. The silicon content of the last ladle on May 20th was in excess of 3.60% which was the calibration limit of the spectrograph. To obtain a uniform silicon level, it appears necessary to adjust ferro alloy additions at the beginning and end of a heat to compensate for the melting of the pellets in advance of their respective charges.

Carburization and the resulting carbon level was related directly to spout temperature. No significant variations in carbon were found to occur with a change in pellet charges unless that change was accompanied by a drop in metal temperature.

Description of Production Heats

Following the successful melting of charges containing up to 70% prereduced material in the 60 in. experimental cupola, two 8-hour ductile heats were run in a 108-in. cupola to evaluate the effects of replacing 20, 30 and 40% of low phos steel scrap with pellets. Details of the cupola preparation and operation are listed in Table 2. The cupola charges are listed in Table 3. To establish a benchmark, the normal ductile iron charge consisting of 85% steel and 15% ductile returns was charged into the cupola at the beginning of each heat. As soon as melting conditions became stabilized (approximately 90 minutes after tap-out) substitution of pellets for part of the steel charge was begun.

Previous experience with pellets in the experimental cupola indicated that the pellets drop through the steel scrap and melt several charges ahead of the charge of which they were originally a part. To offset the temporary drop in temperature and dilution of silicon when this occurs, extra coke and ferrosilicon equal to the amount required by one charge containing pellets were added to the two base-mix charges preceding the first charge containing pellets. A similar procedure was followed each time the percentage of pellets in the charge was changed during the heat. The amount of coke charged with the base mix was increased when charging pellets to provide for the calcination, melting and superheating of the gangue in the pellets.

Spout temperatures were monitored continuously with a recording radiation pyrometer and measured at 15 minute intervals with an optical pyrometer. A spectrographic sample was poured from the forehearth each 15 minutes.

TABLE 5

Effect of Pellets on Wind Box Pressure Production Cupola.

CHARGE	Maximum Pressure at Blast Rate of:		
	740#/ Minute	790#/ Minute	840#/ Minute
BASE MIX	17 oz	19 oz	22 oz
20% PELLETS	20 oz	22 oz	27 oz
30% PELLETS	22 oz	26 oz	30 oz
40% PELLETS	20 oz	---	25 oz

Results of Production Heats

Blast Pressure

Although the stepup to the larger cupola resulted in an increase in melting rate from 9 to 26 ton per hour the effects of using pellets in the larger cupola almost duplicated the results obtained with the experimental cupola. Unlike the experimental heats, it was not possible to operate the production cupola at a constant blast rate due to varying production demands for hot metal. As a result, variations in windbox pressure resulted from both a change in the amount of pellets charged and changes in blast rate.

Table 5 summarizes the influence of pellets on the maximum pressures obtained at the three blast rates used during the two heats. The pressures obtained at blast rates of 740 and 790 lb/min occurred during reasonably stable melting periods. The high values obtained at the 840 lb/min blast rate with 20% and 30% pellets occurred during the transition periods from base to pellet mixes. Of most significance was the maximum pressure of 25 oz obtained with 40% pellets at an 840 lb/min blast rate. Cupola operations were stable throughout this two-hour period during which time the average pressure was 23 oz. This was an increase of only 3 oz above the base mix.

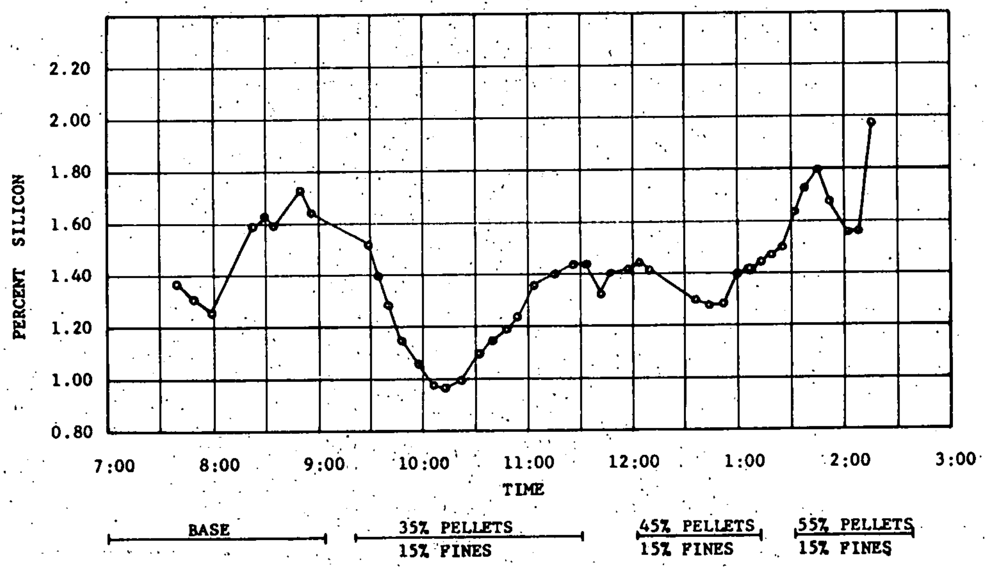

Fig. 2. Variation in silicon resulting from segregation of pellets in the cupola.

TABLE 6. Slag Analyses Production Cupola.

TIME	SiO_2	Al_2O_3	CaO	MgO	FeO	MnO	Basicity	Charge
September 28								
8:30	47.68	16.84	29.60	3.90	2.98	1.16	0.52	Base Mix
9:30					3.36			20% Pellets
10:30	49.08	15.32	20.40	11.48	2.56	1.46	0.50	20% Pellets
1:00					3.18			30% Pellets
2:00	55.44	15.00	14.16	12.42	3.04	1.19	0.38	30% Pellets
October 5								
8:00	52.54	13.68	27.60	4.16	2.30	0.54	0.48	Base Mix
9:30					1.47			30% Pellets
10:30	48.46	13.40	25.00	3.78	2.68	2.79	0.56	30% Pellets
11:30					1.72			40% Pellets
2:00	51.88	20.04	16.20	5.12	2.10	0.38	0.30	40% Pellets

Metal Temperatures

Spout temperatures were 2880 ± 40 F (1582 ± 22 C) during the production heats. This was below the spout temperatures of 2910 to 2925 F (1599 to 1607 C) usually experienced with a ductile base, pellet-free charge in the production cupolas.

Metal-Coke Ratio

Metal-coke ratios are listed in Table 3. As the percentage of pellets was increased, the coke charge was increased as follows:

% Pellets	Coke Ratio	% Increase
0	7.0/1	–
10	6.6/1	5.9
20	6.4/1	9.4
30	6.2/1	12.9

The average metal temperatures were about 30 F (17 C) lower than experienced with pellet-free heats, therefore the above coke charges may have to be increased if a spout temperature of 2910-2925 F (1599-1607 C) is to be attained.

Cupola Slag

Since the pellets used in the production heats contained 8% gangue, the addition of 40% pellets introduced 160 lb of gangue per charge that had to be slagged off. As the composition of the gangue made it self-fluxing, it was not necessary to add additional limestone. The slag volume, however, increased 140%. This presented no problem in cupola operation but it did result in increased labor to dispose of the larger volume. The large volume of acid slag resulting from the gangue makes it impractical to use prereduced pellets of this composition in a basic cupola.

The chemical analyses of the slags listed in Table 6 are within the range of typical acid cupola slags. Although calculations based on the 0.8% Fe^{++} in the pellets indicated that up to 8% FeO could be expected in the slag, the averages of the actual FeO values were approximately equal to the FeO formed in the absence of pellets. This confirmed observations made during the experimental heats wherein charges containing 70% prereduced material failed to produce an increase in the FeO content of the slag.

Metal Composition

The effect of pellets on residual tramp elements is shown in Table 7. The "base iron" listed in the table is a summary of 20 eight-hour pellet-free heats made with low-phos steel scrap and returns. Due to the dilution by the high-purity pellets, the percentage of all tramp elements except titanium decreased when pellets were charged.

Storage of Pellets

Thirty-five tons of prereduced pellets remaining after completion of the test heats were stored outdoors on the ground from Sept. 1966 to March 1967. For protection against the weather the pellets were covered with a sheet of plastic. At the end of the six months period, the surface of the pellets still exhibited their silvery-gray luster. The only pellets showing

TABLE 7.

Effect Of Prereduced Pellets On Residual Elements.

	Production Heats			
Element	Charge	Low	Average	High
Aluminum	Base	0.009	0.018	0.023
	20%	0.002	0.008	0.011
	30%	0.002	0.011	0.018
	40%	0.002	0.011	0.014
Arsenic	Base	0.002	0.004	0.007
	20%	0.001	0.001	0.002
	30%	N.D.	- - - -	0.002
	40%	N.D.	- - - -	0.005
Chromium	Base	0.028	0.042	0.060
	20%	0.012	0.019	0.022
	30%	0.018	0.025	0.031
	40%	0.018	0.025	0.036
Copper	Base	0.075	0.087	0.094
	20%	0.052	0.073	0.082
	30%	0.055	0.064	0.073
	40%	0.055	0.062	0.071
Manganese	Base	0.20	0.26	0.29
	20%	0.16	0.21	0.24
	30%	0.15	0.17	0.22
	40%	0.11	0.12	0.14
Molybdenum	Base	N.D.	0.002	0.007
	20%	N.D.	N.D.	N.D.
	30%	N.D.	- - - -	0.002
	40%	N.D.	N.D.	0.004
Phosphorus	Base	0.030	0.035	0.044
	20%	0.019	0.023	0.022
	30%	0.020	0.026	0.031
	40%	0.020	0.024	0.030
Tin	Base	0.008	0.011	0.012
	20%	0.001	0.005	0.006
	30%	0.004	0.006	0.007
	40%	0.002	0.006	0.008
Titanium	Base	0.002	0.003	0.006
	20%	0.001	0.007	0.014
	30%	0.002	0.009	0.030
	40%	0.001	0.002	0.003

signs of rust were found on the edges of the pile where they had not been covered. When fractured, a bright metallic luster was found throughout the interior of the pellet.

The remaining pellets were charged into the production cupola as 30% of a ductile base mix for a four-hour period. As a possible measure of oxidation during storage, the FeO content of the slag was determined during and after the pellet melting period. Surprisingly, the FeO was lower during the period the pellets were melted.

Time	% Pellets	% FeO
10:00 AM	30	0.67
1:15 PM	0	2.30
3:15 PM	0	1.13

Prereduced pellets appear to be stable enough to resist oxidation if they can be transported and stored in a manner that avoids direct exposure to rain and snow.

Economics

Although it was demonstrated that prereduced pellets can be melted successfully in an acid cupola, the economics of doing so has not been resolved. Because the production of prereduced pellets has been confined to pilot plants, a firm pellet price based on the actual operating experiences of a large commercial venture has not been established.

Several estimates have been published[3,4,5] indicating costs ranging from $30.00 to $48.00 per gross ton of metallic iron FOB lower lake ports. The most realistic estimates indicate a minimum cost of $40.00 for the better grades of pellets containing 92% or more metallic iron. Assuming a $40.00 base pellet price to which is added the cost of freight, extra coke and ferrosilicon, the cost of using pellets in Coshocton, Ohio would be approximately $49.00 G.T. At this price it would not be economical to substitute pellets for low-phos plate and structural steel scrap at today's scrap prices (Sept. 1967). Prereduced pellets will become attractive, however, if high-quality steel scrap becomes scarce or more costly than pellets.

No tests were made to evaluate prereduced pellets as a substitute for pig iron. However, it must be recognized that the chemical purity of the pellets also means freedom from the needed elements found in pig iron, i.e. carbon and silicon and for some applications manganese and even phosphorus. The cost of adding these elements to the pellet charge must be added to the price of pellets before making a comparison with pig iron costs. There are some reduction processes under development that produce a product containing up to 9% carbon, although the high carbon content is accomplished by using up to 40% FeO which makes the material unsuitable for cupola melting stock. Based on metallurgical behavior, prereduced pellets of the composition used in this investigation are more properly considered a replacement for steel scrap than as a substitute for pig iron.

Availability of Prereduced Pellets

Pickands Mather & Co., Steel Company of Canada and Lurgichemie made the prereduced pellets available to the Clow Corp. for this specific research program. Production has been confined to pilot-plant operations where limited quantities have been produced to prove reduction processes and to supply reduced material for specific research projects associated with the use of pellets in the blast furnace, electric arc furnace and cupola. There is, nevertheless, a snowballing interest throughout the steel industry accompanied by much activity in the research and development areas[6] and a number of firms are seriously considering construction of full scale reduction plants. Since the use of prereduced pellets in the cupola is not a current practice and because the economics of doing so is uncertain, it is probable that their availability to the cupola operator will follow the development of their use for the blast furnace and/or electric furnace. There is certain to be abundant supply of oxide pellets available for reduction. North American capacity to produce oxide pellets is now 52,600,000 tons annually and 21,250,000 tons of additional capacity are under construction. All of this production, however, will not be available or even suitable for reduction.

The technology of reducing oxide pellets by an external reductant such as the SL/RN process has been developed successfully and the ability of the cupola to melt prereduced pellets has been demonstrated. Just when prereduced pellets will become available to the foundryman, however, is largely a question of the economics of producing and utilizing this material by the iron and steel industry.

Conclusions

1. Charges containing up to 70% prereduced pellets can be successfully melted in a cupola without creating excessive windbox pressures or adversely effecting general cupola operations.

2. When substituted for steel scrap in cupola charges, prereduced pellets, because of their high purity, reduce the amount of undesirable tramp elements found in the resulting castings.

3. Estimates place the cost of pellets somewhere between the cost of foundry steel scrap and pig iron. However, in the absence of a firm pellet price, the economics of using pellets can not be accurately determined.

Acknowledgement

The authors wish to express their gratitude to William H. Dailey, Jr. and Woodrow W. Holden of Pickands Mather and Co.; J. George Sabikan and Frank J. Pearce of The Steel Company of Canada, Ltd.; and the staff of Clow-Coshocton for their cooperation and encouragement throughout the course of this investigation.

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VISIT 8: PEACE RIVER MINING AND SMELTING LTD

Situated at
Concession 2 Road,
AMHERSTBURG, ONTARIO CANADA

DATE: Thursday 12 November 1970

MET: Dr Tyson Rigg, Chief Chemist

1. INTRODUCTION

Dr Gravenor has very recently left the company to take up a position at the University of Windsor, and the new General Manager is Mr Ken G. Whiteley. Dr Rigg supplied a number of papers regarding the iron powder process and related topics which give a flowsheet for the process. The new plant is designed to produce 50,000 tons per annum of iron powder from scrap, and while certain units have been operated, the plant is not yet running. The three papers appended give technical and cost data. A further paper was seen later "Peace River Iron Powder Plant in Ontario" Steel Times June 1969.

2. DISCUSSION

2.1 Feedstocks

In answer to questions regarding iron ore types Dr Rigg said that hematites and some magnetites, especially those containing apatite and titania, dissolved readily. Pure pyrite does not readily dissolve but pyrrhotite does. Peace River have had some discussions with Murphyores regarding ilmenite treatment by the Peace River process, and this is economically attractive because of the value of the synthetic rutile. Dr Rigg has been examining vanadiferous ilmenites and titanomagnetites, including one from Australia which dissolves readily, iron and vanadium both going into solution. Red mud is a less attractive feed because of filtration problems and loss of hydrogen chloride by combination with sodium and calcium. Chlorides which will not hydrolyse or reduce are to be avoided. Dr Rigg considered that with sufficiently cheap ore a plant of half the present capacity could be viable. The present plant treats wet, oily turnings.

2.2 Product

Dr Rigg said that the big possibility for powder exploitation was in preformed shapes for hot forging. When it was realised that in forging from bar stock 35-40% of the metal was wasted, in comparison with only 5% when using a preformed compact, the prices were competitive. In addition the tools last three times as long when a preformed compact is used. Alloys can be produced by mixing powders. The Peace River powder consists essentially of single crystals and contains none of the inclusions present in directly - reduced iron powders. He showed high-tensile parts produced by this means.

The product will sell for 8-10 cents per lb.

3. DESCRIPTION OF PLANT

Dr Rigg showed me over the plant, commencing with elevators conveying superficially rusted steel turnings over the fence from an adjacent scrap heap into storage hoppers which feed the dissolving tanks. The batch hydrogen-reduction vessels, four in all, were massive vessels and were undergoing modification. A continuous reduction process is being sought. Reduction is carried out by hydrogen at a temperature well above the melting point of the ferrous chloride briquettes, which do not melt because the reaction is endothermic.

4. FURTHER ACTION

Peace River Mining and Smelting Ltd would be prepared to test ore samples submitted and schedules have been prepared giving prices for the supply of information. A copy could not be found while I was there, but was subsequently posted to Adelaide (pages A-85 & A-86). It appears that my visit constituted the greater part of a Level 1 disclosure.

Pricing Schedule	DISCLOSURE FEE PEACE RIVER IRON POWDER PROCESS	SCHEDULE NO. PP-2 DATE OF ISSUE: Mar. 16, 1970 SUPERSEDES NO. PP-1

PEACE RIVER IRON POWDER PROCESS

Interested parties can arrange to obtain the details of the Peace River Iron Powder Process by signing a secrecy agreement and payment of a disclosure fee. The size of the fee depends upon the depth of the disclosure as described below.

Level	Depth of Disclosure	Fee
I	Plant tour and two hour discussion meeting.	$ 1,000 non-redeemable.
II	Economic and technical disclosure of process details.	$15,000 non-redeemable.
III	Preliminary feasibility study for a specific plant location and feedstock.	$50,000 minimum plus $200 per day for all Peace River staff time in excess of 90 mandays plus all travel expenses-non-redeemable.

The depths of the disclosure levels can be described as follows:

Level I

A tour of the Peace River Iron Powder Plant at Amherstburg, Ontario, plus a two hour discussion of the general economics and technology of the Peace River Process with senior research, engineering and management personnel.

Continued over

PEACE RIVER MINING & SMELTING LTD.
AMHERSTBURG - ONTARIO - CANADA
(519) 736-2125 TELEX 024-77737

Level II

A detailed written disclosure of the Peace River Iron Powder Process with technical reasons for the choice of the process steps, an outline flow diagram with process conditions and mass balance (excluding details of the Aman reactor), materials of construction with reasons for choice, description of the control system, estimated utilities consumptions, manning schedules, plant efficiences sufficient to allow operating costs to be evaluated, typical North American capital equipment costs for main process items, land and site service requirement, raw materials requirements, and descriptions of capabilities of the products produced by Peace River at Amherstburg. Discussion of the written disclosure.

Level III

A preliminary feasibility study prepared by Peace River technical personnel with respect to a specific plant location and feedstock as selected by the interested party.

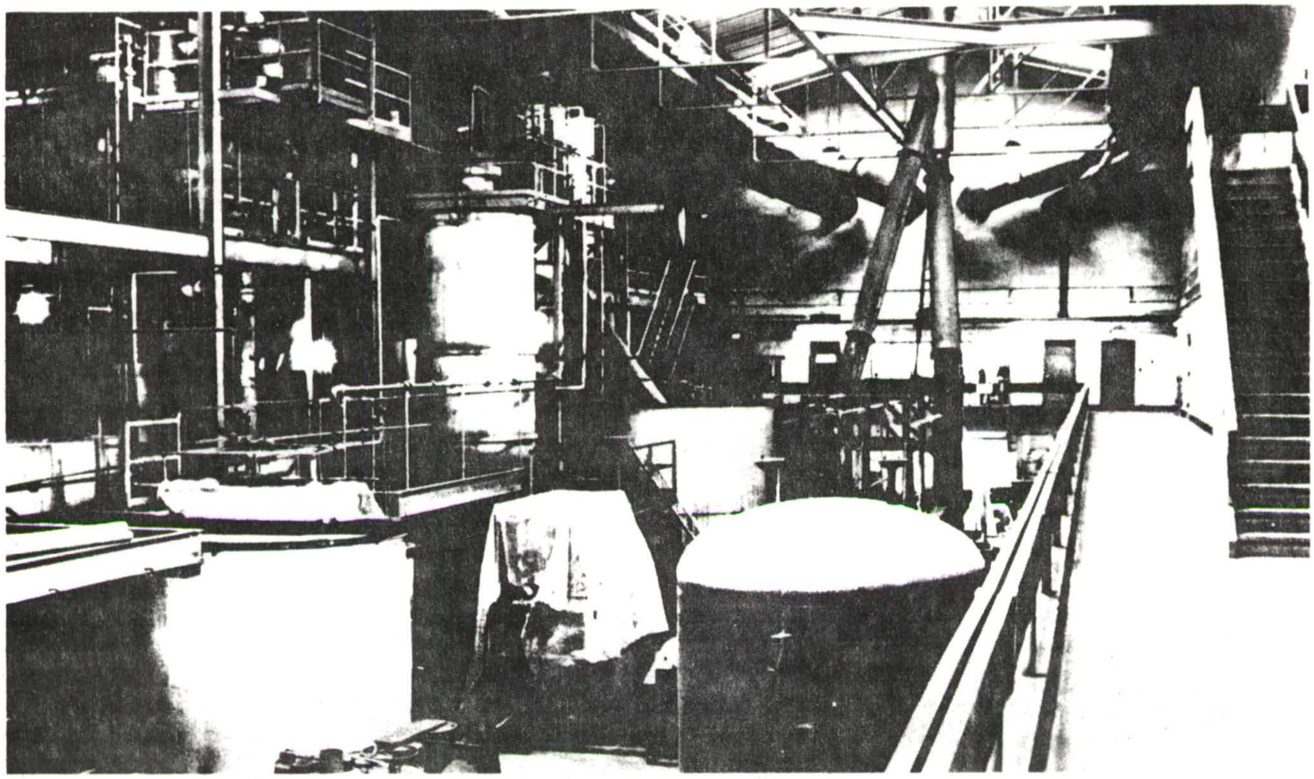
FOOTNOTES

❸ The fees shown above for Level I and Level II Disclosure are the net fees to be realized by Peace River, payable in Canadian funds in advance of receipt of the disclosures. The details of the fees for Level III. Disclosure will be defined in a written contract between the parties concerned.

This disclosure offering is subject to withdrawal or change without notice and the normal limitations of prior agreements between Peace River Mining & Smelting Ltd. and other interested parties applies.

Schedule No. PP-2
Date of Issue Mar. 16, 1970.

❸ Changed since previous issue



PILOT PLANT near Edmonton was constructed to test the ferrous chloride reduction stage on a large scale basis.

Reduction of Hot Ferrous Chloride Is Key to Iron Powder Production

Initially developed to exploit a low-grade iron deposit, this process starts by leaching scrap iron with hydrochloric acid. The resultant ferrous chloride is reduced to sponge iron in a batch-type reactor. Product yield is high, and the iron obtained is of high purity.

CHARLES LAW, *World News, Toronto*

A plant that uses a novel process for making iron powder is scheduled to go onstream next year in Canada. Engineered and built for Peace River Mining & Smelting, Ltd. (Edmonton, Alta.) by its wholly owned subsidiary Great Lakes Forgings, Ltd., the $14.5-million, 150-ton/day installation will use the hydrogen reduction of ferrous chloride to obtain molding-grade powders for the automotive industry.

Iron powders are presently made by reducing specially selected oxides or atomizing molten metal. But Peace River Mining, working with the Research Council of Alberta, and Woodall-Duckham Ltd. of England developed the hydrochloric acid leach process to exploit a low-grade oolitic iron deposit it holds in the Clear Hills region of northwestern Alberta.

In the near future, however, scrap rather than Clear Hills ore will be used as raw material for the new plant; this is because Peace River Mining believes that a large-volume demand for pressed parts must first be established before exploitation of its remote ore property is warranted. Thus, the decision to locate the commercial unit in the Windsor-Detroit area is based on the proximity of established powder markets and an abundant supply of scrap.

Now under construction in the town of Anderdon, the new plant will be in operation next August. One interesting aspect of the venture is the prospect that the high-grade powder product can be rolled and

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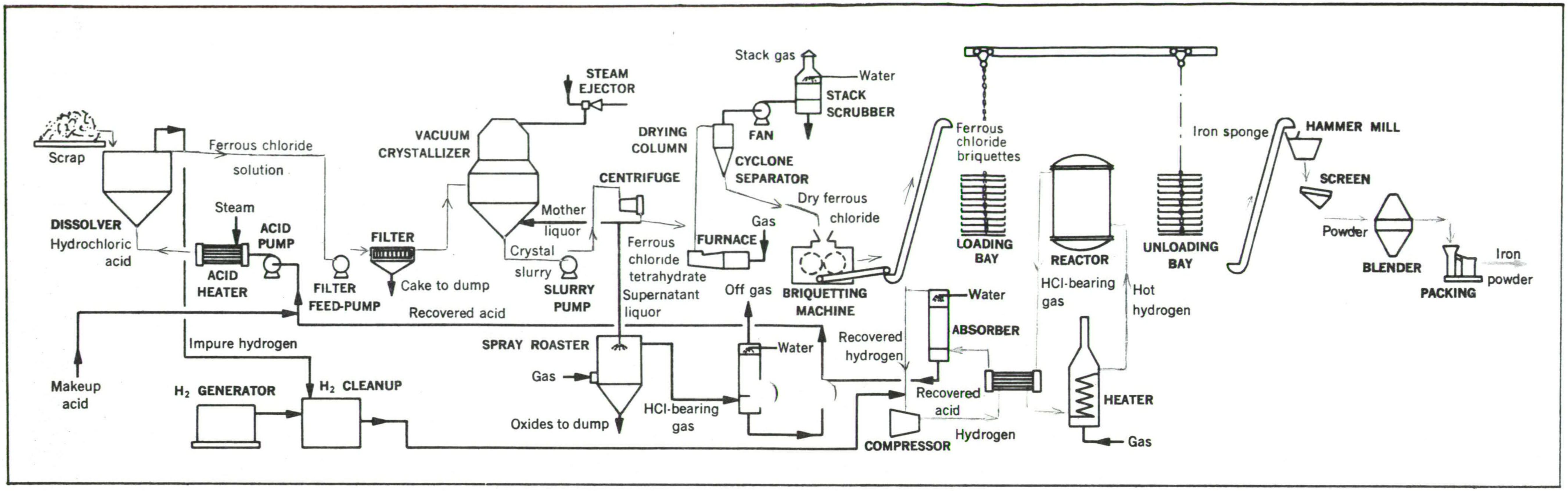


sintered to a steel strip attractive to a wide variety of steel fabricators. (No significant demand now exists for sintered strip—primarily because there has been little powder available at costs sufficiently low to spark a development program.)

▶ **Development Work**—In a joint effort, Peace River Mining and the Research Council of Alberta first tried a sulfuric acid leach on the iron deposits; this was abandoned because of high acid-loss and the difficulty of obtaining a high-purity, sulfur-free iron.

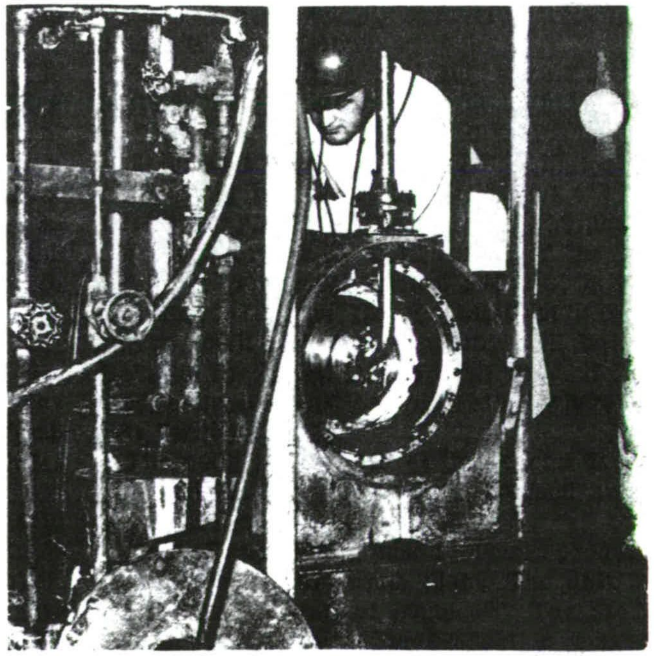
The chemistry of the chosen hydrochloric acid route is simple enough, although considerable engineering development was required before it became workable on a large scale. Process variables, especially those that regulate ferrous chloride reduction, were determined in a $1-million, 5-ton/day pilot plant that has been operating near Edmonton since early 1966.

Dissolution of the iron is the first step, the following reaction taking place at 194 F. and atmospheric pressure:

$$Fe^{++} + 2HCl \rightarrow FeCl_2 + 2H^+$$

Subsequent reduction of the ferrous chloride is what poses major design problems. Essentially this is a high-temperature (1,112 to 1,472 F.) step in which the above reaction is reversed by combining the briquetted ferrous chloride crystals with hydrogen at atmospheric pressure. The heat needed for reacting is about 37,000 kcal./mole, but an additional 23,000 kcal./mole are provided to preheat the feed and evaporate residual water.

The reduction step takes place in batch fashion inside fixed-bed reactors. An attempt was made to develop a continuous reactor, but this idea was dropped because of difficulties in handling hot ferrous chloride. In the chosen units, bed depth is about 2 ft.; the charge is loaded and unloaded in a stack of trays. Peace River Mining says over-all process yield is higher than 93% in terms of iron input. Final product is 99.9% Fe.


CENTRIFUGE separates ferrous chloride from mother liquor.

▶ **Operating With Scrap**—At the new plant, scrap is fed into a dissolver and reacted with hot acid containing 20 to 25% by weight HCl. Iron dissolves in the acid, and the ferrous chloride solution containing 3 to 5% residual acid is pumped to a filter that removes insoluble impurities.

Crystallization of ferrous chloride tetrahydrate is done in a vacuum crystallizer, and separation from the mother liquor is carried out in a subsequent centrifuge. The mother liquor is returned to the vacuum crystallizer, but the supernatant liquor from the centrifuge, which contains soluble impurities such as metal chlorides, is sent to a spray roaster. Here, the impurities are hydrolyzed to oxides, and the hydrogen chloride released is recovered for use in the dissolving step.

Because these vessels handle ferrous chloride solutions, they must use rubber-lined steel, whereas the piping is made of glass-fiber-reinforced plastic. Pump internals are made of either silicon iron or graphite; the centrifuge has an all-titanium front end.

To improve both the strength of the recovered acid and the quality of briquettes, the tetrahydrate is dried to the dihydrate stage in a dispersion dryer. A roll press then shapes the hot, dry material into pillowlike briquettes that improve gas-solids contacting in the reduction stage.

▶ **Reduction Step**—The $FeCl_2$ briquettes are charged into trays and loaded into the reactor; meanwhile, hot hydrogen is injected through the bottom of this unit. As the reaction proceeds at temperatures up to 1,472 F., HCl gas issues from the top of the reactor. After heating hydrogen from the compressor, the gas is sent to an absorber. Here, hydrogen is obtained from the top, recompressed, and returned to the reduction step; liquid HCl recovered from the bottom is sent to the dissolver.

Peace River Mining engineers say no unusual precautions are necessary to protect individual units against high temperatures and corrosion. However, certain reactor sections are built of high-alloy steels to guard against temperatures above 1,472 F. Carbon steel is adequate for other pieces of equipment.

Since hydrogen is a byproduct of the dissolving step, it is recovered, purified and used in the reduction stage. Because the raw material is scrap iron, only a small hydrogen generator will be needed to make up for losses. On the other hand, if iron ore is used, hydrogen must be produced elsewhere in a separate facility.

The material that leaves the reactor is a mass of sponge iron. This is ground, screened and blended to give a specification product.

Peace River Mining hasn't released production cost figures for its Anderdon plant. The unit, however, will employ a staff of 80 to 100. The $14.5-million capital cost includes preproduction expenses and working capital.

POWDER METALLURGY / AND IRON POWDER PRODUCTION

A Hydrometallurgical Process to Produce Iron Powder from Scrap Iron

C. P. GRAVENOR, General Manager,
T. RIGG, Chief Chemist,
J. N. STONE, Chief Engineer,
Peace River Mining & Smelting Ltd.,
Amherstburg, Ont.

ABSTRACT

The process described has been developed to utilize low-grade scrap — turnings, borings, tin cans and other types of ferrous scrap which currently has been of limited or no use in conventional steelmaking processes. If scrap of this type can be converted into high-purity iron by chemical means, then there is a potential saving of several million tons of metal annually in North America.

The process is chemically quite simple and consists of dissolving ferrous scrap in hydrochloric acid, followed by evaporation and crystallization of the resultant solution to yield ferrous chloride crystals. These crystals are dried, briquetted and converted to iron sponge by reduction in hot hydrogen. The HCl produced from the reduction step is absorbed in water and the hydrochloric acid so produced is returned to the dissolver circuit. The hydrogen produced from the initial dissolution of the scrap is used to reduce the ferrous chloride. Purity is maintained by bleeding off impure chloride solutions from the evaporator-crystallizer circuit, and converting these chlorides to oxides and HCl in a high-temperature spray roaster. The HCl produced in the roaster is absorbed in water and the hydrochloric acid returned to the dissolver.


C. P. GRAVENOR


T. RIGG


J. N. STONE

C. P. GRAVENOR, vice-president and general manager of Peace River Mining and Smelting Ltd., is a University of Toronto graduate with a B.A. in Honours Science (geology, 1949), an M.S. in 1950 from the University of Wisconsin and a Ph.D. in geology from Indiana University in 1952. From 1952 to 1956 he taught geology at the University of Alberta, and from 1956 to 1965 he was chief of earth sciences and assistant director of research, Research Council of Alberta. In 1965, he left the Research Council to join Peace River Mining and Smelting Ltd. Since 1960, he has been involved in developing hydrometallurgical techniques to extract metals from low-grade raw materials and industrial waste products.

TYSON RIGG graduated (B.Sc. 1949, M.Sc. 1951, and Ph.D. 1953) from the University of Newcastle-on-Tyne, England, specialising in inorganic chemistry. After a year of post-doctoral studies, he joined Marchon Products Ltd., later taking up an appointment with the U.K. Atomic Energy Authority. In 1960, he came to Canada to work at the Research Council of Alberta, transferring in 1966 to his present position as chief chemist with Peace River Mining and Smelting Ltd.

J. N. STONE, chief engineer, Peace River Mining and Smelting Ltd., is a graduate in chemical engineering from the University of London, England (1957). After several years experience in an organic chemical manufacturing company and a contracting company specializing in corrosion-resistant chemical plant, he joined Peace River in 1965, and has since been actively involved in the design, start-up and evaluation of both pilot and commercial iron powder plants.

PAPER PRESENTED: at the 71st Annual General Meeting of the Institute, Montreal, April, 1969.

KEYWORDS: Powder metallurgy, Iron powders, Hydrometallurgy, Scrap iron, Hydrochloric acid, Alloys, Dissolution, Centrifuges, Briquetting, Aman reactor, Ferrous chloride, Manganese alloys.

INTRODUCTION

In recent years, there has been increasing concern over the build-up of indestructible wastes in North America. Tailings piles, and slag heaps from mining and refining operations, red mud from alumina plants and increasing amounts of iron and steel scrap in garbage dumps and in the form of abandoned automobiles are creating unsightly blights on the countryside. Not only are these waste materials a problem in terms of land usage and appearance, but they also represent a loss of potential raw material to the metallurgical industry.

It has been pointed out by Rosenbaum (1966) in an address to the Engineers' Joint Council Meeting that, until recently, automobile scrap has been considered as a valuable source of salvage and scrap. However, the over-all composition of automobiles is changing with the increasing use of stainless, aluminized and galvanized steel, aluminum, zinc and copper. This means that the automobiles which will appear in the scrap yards in the 70's will have a more complex composition and hence will be more difficult to use in conventional steelmaking processes. In addition, there are large tonnages of steel in the form of tin cans and other refuse which are being used as land fill. This represents a loss of several million tons of metal annually in North America and hence is a wasteful practice.

It has long been recognized that if hydrometallurgical processes could be applied economically to steelmaking a close control over quality could be obtained and consequently low-grade scrap could be used to produce high-purity products. At present, there is a modest amount of low-grade scrap used in the steel industry, but materials such as machine-shop turnings are not considered as good feed stocks. Consequently, the value placed on scrap iron is proportional to the physical nature and purity of the scrap. In industrial areas, the current value of high-purity scrap is in the order of \$20.00 to \$40.00 per ton. In the same areas, impure scrap has a value of \$6.00 to \$20.00 per ton, even though the total intrinsic value of the impure scrap may be higher than high-purity scrap. It is evident that if wet chemical processes can be developed to handle the lower-cost material, such processes would aid considerably in the control of impurities build-up in recycled scrap, and would provide a means of decreasing the mounting supplies of metal waste while at the same time producing high-purity metals to the benefit of the economy.

HISTORY OF PROCESS DEVELOPMENT

A research program was set up in 1960 at the Research Council of Alberta to develop a process to produce iron powder from the low-grade Peace River iron deposits. The results of the early laboratory and small-scale pilot-plant work were published in 1964 (Gravenor *et al.*, 1964).

At the conclusion of the small-scale pilot-plant investigations, it was recognized that the process would have to be tested in a much larger pilot plant using commercially available equipment before a full-scale commercial plant could be built. Consequently, in 1964, funds were raised by Peace River Mining & Smelting Ltd. to start the design and construction of a 5-tpd plant. Design was started early in 1965 and in the late fall of 1965 erection of equipment commenced on space leased from the Research Council of Alberta. The erection was completed in the spring of 1966 and testing started in March. By the fall of 1966, all equipment had been commissioned and the first iron sponge produced. Experimental work on the scrap process has now been completed and the plant is being used for other developmental purposes.

In order to simplify the initial start-up of the pilot-plant, it was decided to use scrap metal as a source of iron and later to convert to Peace River iron ore as the feed. Various types of scrap have now been tested, including heavy scrap, turnings and tin cans. Experiments on the dissolution of Peace River ore were started in the spring of 1967, and this program is being continued.

After running the plant on scrap for several months, it was decided to explore the possibility of using scrap as the feed-stock in the first commercial development. There are many reasons for this decision, but primarily a scrap plant can be economical on a smaller scale than an ore plant and hence would be an excellent interim development allowing time to improve the technology prior to the construction of a large plant based upon Peace River ore.

SCRAP PROCESS

Essentially any type of ferrous scrap can be used as feed to the process. It is economically advantageous to use low-cost scrap such as machine-shop turnings and mixed turnings and borings, tin cans and automobile scrap. In general, this type of scrap also offers the advantage of having a large surface area which leads to rapid dissolution in hydrochloric acid and, in addition, usually contains appreciable amounts of other metals, some of which may be recovered as by-products.

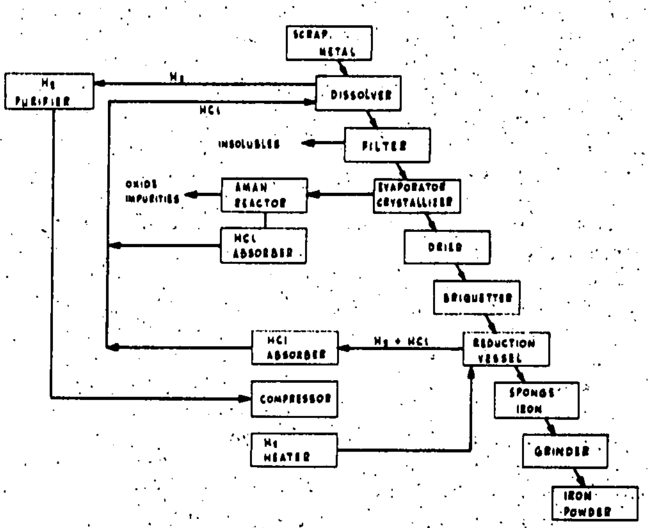


FIGURE 1.—The Scrap Process.

Briefly, the process consists of the dissolution of scrap iron in hydrochloric acid, followed by the evaporation of the solution to yield crystals of ferrous chloride. The ferrous chloride crystals are dried and briquetted, and the briquettes are reduced in hydrogen to produce iron sponge and hydrochloric acid; the acid is recycled to the dissolution stage. An outline of the process steps is shown schematically in *Figure 1*, and the following is a description of each stage of the process.

DISSOLUTION

As there is little information available on the deliberate dissolution of large quantities of steel scrap in hydrochloric acid, a laboratory and pilot-plant program was undertaken to provide the necessary data for up-scaling the dissolver system to commercial size. The results of this lab and pilot program have been published recently by Rigg (1968).

Iron dissolves readily in hydrochloric acid and the rate of dissolution increases markedly with an increase in surface area and temperature. Although the reaction is exothermic:

$$Fe + 2HCl\ (aq) \rightleftharpoons H_2 + FeCl_2\ (aq)\quad \Delta H = -20.96\ \text{K.cal.},$$

in practice it has been found that it is necessary to preheat the acid to about 95°C. After this initial preheat and using machine-shop turnings as the feed, it has been found that the reaction is self-sustaining.

The hydrochloric acid fed to the dissolver also reacts with carbides and sulphides in the steel turnings to release hydrocarbons and H_2S. Essentially all of the sulphur in the initial feed comes off as H_2S, except for a small part which reacts with certain metallics (i.e. copper) to form insoluble sulphides which are trapped in the filter residue. Gas chromatographic analysis of the dissolver off-gas shows it to contain 0.13 per cent CO_2 and 0.11 per cent hydrocarbons (C_1 to C_7). As the hydrogen from the dissolver is to be used to reduce ferrous chloride, it is first passed through an absorber to remove traces of HCl and then sent to a MEA scrubber to remove H_2S before being placed in storage for use in the reduction stage.

Pilot-plant results indicate that a typical solution coming from the dissolver contains about 31 per cent ferrous chloride and 3 per cent free acid.

FILTRATION

The ferrous chloride liquor from the dissolver contains a mixture of insoluble materials which must be removed before evaporation and crystallization. Using clean turnings as the feed, the dried insoluble residue amounts to about 0.1 per cent of the weight of the feed. For this purpose, a pressure-edge filter was chosen for high solid capacity and ease of manufacture with acid-resistant materials. The composition of the insoluble residue varies considerably, depending on the nature of the scrap. An analysis of the insoluble residue is given in Table I (metals determined spectrographically) and represents an average of several analyses.

Table I — Analysis of Insoluble Residue

Constituent	%	Constituent	%
C	3.51	Fe	matrix
S	0.02	Pb	trace
Al	1.0	Mg	0.4
Sb	0.4	Mn	0.5
As	0.25	Mo	0.3
Ba	trace	Nb	N.D.
Be	N.D.*	Ni	0.7
Bi	N.D.	Si	matrix
B	N.D.	Ag	trace
Cd	N.D.	Sr	trace
Ca	0.3	Sn	0.003
Cr	0.8	Ta	N.D.
Co	0.01	Ti	0.3
Cu	0.1	W	0.1
Ga	N.D.	V	0.03
		Zn	trace

*N.D. — non-detectable

It is evident that the combination of dissolving and filtering provides a major purification step in the treatment of scrap. The bulk of the carbon, sulphur and silica is removed in these stages, as well as a substantial amount of aluminum and other metals presumably present as refractory inclusions in the scrap.

EVAPORATION AND CRYSTALLIZATION

The ferrous chloride liquor from the filter is fed to an evaporator-crystallizer, where the concentration of ferrous chloride in the solution is increased to approximately 48 per cent. For the first few months of pilot-plant operation, heat was supplied to the evaporator-crystallizer circuit by submerged combustion in which combustion gases from a natural gas burner are passed down a titanium dip tube into a tank of circulating liquor. As the gases pass through the liquor, rapid heat transfer takes place and the hot concentrated solution is then pumped directly to a vacuum crystallizer. Alternatively, and in the system now in use, heat may be supplied from a graphite-tube heat exchanger in which steam is used to supply heat to the circulating liquor. Both systems operate satisfactorily, but there are some process and heat economy advantages in the use of tube heat exchangers.

The crystallizer is maintained under a vacuum (50 mm Hg absolute) by a steam ejector system and, as the hot liquor (75-80°C) is sprayed into the vacuum chamber, concentration is effected by the evaporation of water. Concentration and cooling of the liquor causes crystallization of ferrous chloride tetrahydrate and the crystal slurry passes down a barometric leg into a settler-classifier. A spectrographic analysis of unwashed crystals is given in Table II. The elements are reported as a percentage of the iron content of the ferrous chloride.

CENTRIFUGE

The crystal slurry from the base of the settler-classifier is pumped to a horizontal pusher-type centrifuge. The centrifuge basket is made of titanium wire screen, with the slots parallel to the horizontal axis. The basket rotates about the horizontal axis and the crystal slurry is fed at one end onto a plate normal to the basket axis, which rotates at the same speed as the basket. As the cake forms on the basket, the plate oscillates along the basket axis and so gradually pushes the crystals along the basket until they are discharged at the open end into a catcher ring. As the crystals are pushed along toward the open end, they may be washed to remove surface impurities present in the mother liquor adhering to the outside of the crystals. The crystals, as discharged, contain 3.5 per cent free moisture.

Table II — Analysis of Unwashed Ferrous Chloride Crystals

(reported as per cent of the iron content—

Constituent	%	Constituent	%
Al	0.004	Pb	N.D.
Sb	N.D.*	Mg	0.001
As	N.D.	Mn	0.18
Ba	trace	Mo	0.007
Be	N.D.	Nb	N.D.
Bi	N.D.	Ni	0.029
B	N.D.	Si	0.07
Cd	N.D.	Ag	trace
Ca	trace	Sr	N.D.
Cr	0.07	Ta	N.D.
Co	trace	Sn	N.D.
Cu	0.011	Ti	0.04
Ga	trace	W	N.D.
Au	N.D.	V	trace
		Zn	N.D.

*N.D. — non-detectable

Table III — Screen Analysis of Ferrous Chloride Tetrahydrate Crystals

Tyler Screen Size	% Wt.
+ 10	2.5
+ 20	48.1
+ 40	42.9
+ 60	5.0
+ 80	1.5

The efficiency of the centrifuge washing operation has been determined by the addition of a small quantity of phosphoric acid to the evaporator-crystallizer circuit and by measuring the amount of phosphoric acid in the mother liquor and in the crystal cake with and without washing. By this technique, it has been found that a washing efficiency of 80 per cent or better is readily attainable. Washing efficiency is defined as the percentage of adhering mother liquor washed out of the filter cake.

The crystals of ferrous chloride tetrahydrate are allowed to grow in the crystal settler, so that they can be centrifuged without blinding the screen and present a small surface area which facilitates washing. Table III shows a typical screen analysis of the ferrous chloride crystals as delivered from the centrifuge.

FLASH DRYER

The ferrous chloride tetrahydrate crystals delivered from the centrifuge are not free-flowing, are difficult to briquette, would add too much water to the final reduction reactor gas system and have a low iron-to-volume ratio, which would add considerably to the size of the reduction reactor. For these reasons, it is necessary to dry the crystals to the dihydrate state prior to briquetting. This drying is accomplished in a Raymond flash dryer wherein the wet crystals are entrained in the hot off-gases from a natural-gas burner and the dry crystals removed in a cyclone. During drying, a small amount of ferric oxide is formed and the hydrogen chloride produced is removed from the stack gas by means of an absorber. A typical analysis of the dried product is $FeCl_2.1.7\ H_2O$, and 2 per cent of the iron is present as ferric oxide. Table IV gives a typical screen analysis of dried product from the flash dryer. If the ferrous chloride hydrate is dried to an extent corresponding to the composition $FeCl_2.1.0\ H_2O$, the oxide content increases to about 3 per cent and then increases rapidly with further dehydration.

Table IV — Screen Analysis of Dried Ferrous Chloride

Tyler Screen Size	% Wt.
+ 10	0.6
+ 20	11.8
+ 40	21.1
+ 60	15.4
+ 80	7.9
+ 100	4.9
+ 150	4.4
+ 200	9.8
+ 250	1.6
+ 325	8.4
− 325	12.6

BRIQUETTING

A considerable amount of early experimental work was done at the Research Council of Alberta on the reduction of ferrous chloride powder in both fluidized beds and batches. The fluidized bed proved difficult because of particle density changes, agglomeration and consequent defluidization, and the presence of excessive amounts of chloride in the product. Fixed beds of powder can be reduced, but the time of reduction is unacceptable due to the difficulty in contacting large volumes of hydrogen with the solid unless very high gas pressure drops are acceptable. Consequently, in order to obtain good gas-solid contact with the minimum pressure drop across a bed, it was decided to briquette the ferrous chloride. Dried ferrous chloride can be briquetted, without a binder, in a Komarek-Greaves opposed-roll press. The briquettes produced in the pilot plant are pillow-shaped and are ⅝ in. in thickness by 1½ ins. in length by ⅞ in. wide. They are stored in hoppers prior to charging to the reduction equipment. If the briquettes are exposed to humid conditions for prolonged periods, they will take up moisture, expand and disintegrate. Hence, if these conditions are anticipated, the briquettes should be stored under a dry inert gas blanket.

AMAN HYDROLYSIS

In *Figure 1*, a bleed-off of mother liquor from the evaporator-crystallizer circuit is shown. This bleed-off is required to maintain a low impurity level in the mother liquor and, consequently to prevent breakthrough of impurities to the ferrous chloride tetrahydrate crystals. Using machine-shop turnings or other mild steel scrap, it has been found necessary to remove from 3 to 7 per cent of the iron at this stage in order to maintain a high level of purity in the ferrous chloride crystals. As this represents the main loss of iron in the over-all circuit, it is estimated that it will require about 1.1 tons of feed to produce 1 ton of powder. The additional loss is primarily due to oxidation of the iron scrap prior to treatment.

The Aman reactor was developed by the Woodall-Duckham Company in England, and is essentially a high-temperature spray roaster. The waste chloride solution (containing primarily ferrous chloride with smaller amount of other metallic chlorides) is sprayed into the top of a brick-lined tower and the droplets fall through an up-flowing spiral stream of hot gas generated by a natural-gas burner near the base of the tower. As the droplets fall through the hot gas at temperatures of 600°C to 700°C, the chloride solutions dry and the chlorides are hydrolyzed to oxides. The ferrous chloride undergoes oxidative hydrolysis according to the following reaction:

$$2FeCl_2 + 2H_2O + \tfrac{1}{2}O_2 \rightleftharpoons Fe_2O_3 + 4HCl$$

Displacement reactions may also take place during the drying of the waste chloride solutions, e.g.:

$$3CaCl_2 + 2H_3PO_4 \longrightarrow Ca_3(PO_4)_2 + 6HCl$$

The combustion gases and HCl are taken off from the top of the tower and passed through an absorber to remove the HCl as hydrochloric acid, which is returned to the dissolver circuit. The oxides are removed from the base of the tower and go to solid waste disposal.

If the feed material is rich in alloying elements (i.e. nickel, copper or molybdenum), then it could be economically feasible to remove these constituents either from the waste liquor before it enters the Aman reactor or from the oxide product of the Aman.

REDUCTION OF FERROUS CHLORIDE

Pilot-plant results on the reduction of ferrous chloride briquettes with hydrogen were presented in 1968 by Stone and Rigg.

Prior to the pilot-plant work, a considerable amount of research was carried out on a smaller scale in an attempt to develop various types of reduction equipment. This work, which was done by the Research Council of Alberta and the Woodall-Duckham Company of England, included investigations on the use of fluidized beds, travelling belt reactors and static-bed reactors. Although the fluidized bed is, *a priori*, an attractive technique, three major problems were encountered. First, the particles in the bed tended to agglomerate, which made it difficult to maintain fluidization; second, a considerable amount of ferrous chloride volatilized; and third, there was an excessive amount of ferrous chloride in the product. The latter problem is perhaps the most serious, as it is imperative that all of the ferrous chloride be essentially reduced to iron sponge. By-passing of gas proved to be a problem in the use of travelling belts and, although static beds of ferrous chloride crystals can be reduced, in practice this is not feasible because of the high pressure drop created by the large volume of gas which must be contacted with the solid (12,000 scf of hydrogen per cu.ft of solid).

To overcome the pressure drop problem in fixed beds, it was decided to briquette the ferrous chloride in order to provide sufficiently large pore spaces for gas passage; the uniform size of the briquettes cut down the problem of channelling, which is one of the most difficult problems to overcome in any fixed-bed reactor. After considerable experimentation with large fixed beds, it has been determined that beds of up to 3 feet in thickness can be reduced without encountering serious gas channelling problems.

A considerable amount of equilibrium data on the reaction:

$$FeCl_2 + H_2 \longrightarrow Fe + 2HCl$$

has been published and has been summarized by Rigg (1964). The reaction starts below 400°C, but does not reach appreciable percentages of HCl in the off-gas until temperatures of 600-700°C are attained. For example, at 500°C the mole per cent HCl at equilibrium is about 3 per cent and at the melting temperature of ferrous chloride (673°C) the per cent of HCl is about 16 per cent. It is therefore evident that the temperature should be raised to the maximum allowable in order to keep the amount of recycling of the gas to a minimum.

In early experiments, most of the heat was supplied by radiant heating; when higher temperatures were reached (i.e. above the M.P. of ferrous chloride), the chloride melted, which resulted in severe reactor design problems. It was therefore decided to put all the heat into the system by preheating the hydrogen to about 800°C — well above the melting point of ferrous chloride. Under these conditions, the ferrous chloride does not melt, due to the fact that the reaction is strongly endothermic (approximately 34 K.cal./mole iron at 600°C). At the reaction front in the briquette, this heat requirement acts as a heat barrier, so that the temperature of the centre of the briquette does not exceed the melting point of ferrous chloride.

The reaction is usually topochemical and it has been reasoned that iron transfer occurs *via* the vapour-phase migration of ferrous chloride. In this way, reasonably large cubic crystals of Fe (α) are formed. The size of the crystals may be increased, and ultimately iron "whiskers" are formed, by decreasing the hydrogen flow and introducing heat from a source other than the hydrogen. The general rate equation for the topochemical reaction is as follows (Rigg, 1966):

$$\frac{f}{t} = \frac{75.8}{\rho_{Fe}\, d} \times \frac{P_e - P}{P_e} \times e^{-15,500/RT}$$

where f is the fractional penetration of the reaction front (as a fraction of the radius of the particle), ρ_{Fe} is the density of iron (initially present as chloride) in gm/cm^3, t is the reaction time in minutes, d is the particle diameter in cm, R is the universal gas constant, T is the reaction temperature (°K), P is the partial pressure of hydrogen chloride in the reaction zone and P_e is the equilibrium partial pressure of hydrogen chloride. When f is plotted against time, it can be shown that f is a linear function of reaction time and that the reaction front advances at a constant rate (Rigg, 1966). Hence, the geometry of the briquette is important, as the minimum dimension will determine the length of time for complete reduction.

In practice, where large beds of briquettes are reduced, it has been found that the rate of reaction is roughly one-third the theoretical rate (Stone and Rigg, 1968). This slower rate is attributed to the fact that there is some breakage of the briquettes in handling and hence some of the pore space is blocked by powder and small pieces of briquettes, which in turn gives rise to channelling.

The pilot-plant reactor is 3 ft, 6 inches in diameter and has a total bed depth of 10 feet and a capacity of 8,000 lbs of ferrous chloride dihydrate briquettes. As mentioned earlier in this paper, the maximum bed depth which can be used without encountering severe channelling is about 3 feet. Consequently, the reactor was loaded with beds separated by stainless steel grids. Generally, three to four beds were employed, using combinations of 2- and 3-foot beds. At first, the reactor was installed with a downward flow of hydrogen; however, it was found that as the reaction proceeded the pressure drop started to build up and the flow of hydrogen had to be cut back. This build-up of pressure was caused by the squashing of the bed ~~of~~ briquettes, which are very weak when they are fully reduced and at high temperatures (i.e., 800°C — the temperature of the incoming hydrogen).

After several tests with downward-flowing hydrogen, the reactor was converted to up-flow and, although the pressure drop did not increase appreciably during the reaction, it was found that there was a long "tail" on the reaction. By examining partly reduced beds, it was found that the long reaction time was due to channelling up the side walls of the reactor. Once the material on the side was reduced, there was increased pore space at the sides of the reactor due to shrinkage of the briquettes away from the side walls, and hence it took a long time to reduce the centre of the bed, thus accounting for the "tail" on the reaction. This problem was overcome by increasing the flow of hydrogen and driving the hydrogen directly through the beds, so that reaction took place as a moving front rather than from the side walls into the centre of the beds. Pilot-plant results indicate that it requires about 40 moles of hydrogen per mole of iron for complete reduction.

The HCl - hydrogen gas mixture from the reactor is first passed through a heat exchanger to recover some of the heat and cool the gas before the next stage. The gas exits from the heat exchanger at about 250°C and next passes to a desuperheater, where it is contacted with a spray of concentrated acid in order to cool the gas to about 60°C. From the desuperheater, the cooled gas is passed to an adiabatic absorber, where the HCl is removed from the hydrogen stream. This acid, which has a concentration of about 25-30 per cent, is returned to the dissolver circuit to dissolve new scrap. After the HCl has been removed, the hydrogen is compressed and from there passes through the heat exchanger previously mentioned and then to the main hydrogen heater, where it is heated to 800°C and then passed up through the reactor. More than 200 batches have been processed in the reactor at the pilot plant.

POWDER PREPARATION

The briquettes are found to be lightly sintered together when taken from the reactor and the beds are first chopped into lumps about fist size and then ground to powder in a hammermill. From the hammermill, the powder is first passed over a screen to remove most of the coarse (plus-100-mesh) material and to an elutriator where some of the minus-325-mesh material is removed. The fines from the elutriator are recycled to the briquetter circuit, where they are blended with ferrous chloride dihydrate prior to briquetting. This step is an important part of the process, as straight ferrous chloride reduced in hot hydrogen and then hammermilled produces a powder which consists of 80 to 90 per cent -325-mesh material. However, by seeding the ferrous chloride with recycle powder it has been found possible to grow aggregates of powder particles within the sponge. After grinding, this produces a coarser material suitable for a moulding-grade powder. Consequently, by varying the amount and size of the recycle powder and varying the screen openings of the hammermill, it has been found possible to produce powders differing in apparent density from 1.8 to 3.3. In commercial practice, three powders will be made. The first is a very fine powder for special purposes, and the second and third powders will have apparent densities of 2.4 and 2.8, respectively.

Table V gives a typical chemical analysis of the powder produced in the pilot plant; Table VI gives the screen analysis and other physical characteristics of 2.8 A.D. powder.

Powders with a density of about 2.40 show similar chemical and screen analyses, but have a somewhat lower green density (6.83 at 40 tsi) and higher green strength (4,500 psi at 40 tsi). Powders of higher chemical purity can be obtained by giving the product powders a wash in very dilute acid. This is possible, because the impurities present are largely external to the iron particles. Some powder may be given this treatment in the commercial plant if ultra-high purity is required by consumers.

Table V — Chemical Analysis of Iron Powder

Constituent	%Wt.	Constituent	%Wt.
Aluminum	0.003	Zirconium	N.D.
Arsenic	N.D.	Carbon	0.024
Bismuth	N.D.	Phosphorus	0.001
Boron	N.D.	Sulphur	0.001
Cobalt	0.002	Manganese	0.20
Copper	0.003	Silicon	0.011
Chromium	0.003	Iron	balance
Lead	N.D.	Oxygen (as hydrogen loss)	0.2*
Magnesium	0.020	Insolubles	0.023
Molybdenum	0.006		
Nickel	0.060		
Tin	N.D.		
Titanium	N.D.		
Vanadium	N.D.		

*This is primarily "skin" oxygen on the exterior of particles.

Table VI — Physical Characteristics of Powder

Apparent Density	2.8	
Screen Analysis	**Mesh**	**% Wt.**
	+ 100	2.5
	− 100 + 150	13.4
	− 150 + 200	21.4
	− 200 + 250	11.7
	− 250 + 325	22.5
	− 325	28.1
Subsieve Sizes (−325 fraction)	+ 30 μ	16.3
	+ 20 μ	7.8
	+ 10 μ	3.4
	− 10 μ	0.6

Flow Rate — 29 seconds —
Green density . . . (at 40 tsi — 1% Zn Stearate) — 6.85
Green transverse rupture strength . .
(at 40 tsi — 1% Zn Stearate) — 1,800 psi

ALLOYS

Alloys of nickel, cobalt, molybdenum and iron have been produced in the pilot-plant reactor by co-reducing salts of these elements mixed with iron chloride. The conditions required for producing alloys of this type (maraging steel) are the same as those for reducing ferrous chloride. The alloy powders produced have been compacted, sintered and hot-rolled. The characteristics of the fully dense strip are similar to those reported for commercial maraging steels of similar compositions.

Although nickel, cobalt and molybdenum can be reduced, it is not possible to reduce manganese compounds under the same conditions as ferrous chloride and a special technique has been developed to produce a manganese alloy. This involves mixing manganese dioxide with the ferrous chloride dihydrate prior to briquetting. During the low-temperature phase of reduction, the manganese dioxide reacts to form manganese chloride, which volatizes and migrates through the briquette. Eventually, as the ferrous chloride reduction nears completion, there is insufficient HCl to suppress hydrolysis of the manganese chloride by the moist hydrogen. Under these conditions, the manganese chloride hydrolyses to MnO and HCl, and hence a very fine-grained MnO is dispersed throughout the iron. In the manufacture of powder parts, the finely dispersed MnO is reduced by carbon, resulting in a homogeneous alloy.

The manganese alloy produced in this manner has an advantage in that the powder has good compressibility and green strength, but it has a disadvantage in that it requires a slightly longer sintering time to achieve homogeneity. The carbon can be added as graphite to reduce the manganese oxide or it can be added by changing the composition of the endo gas in the sintering furnace. Attempts have been made to add the manganese dioxide directly to the iron powder, but this technique has not been successful as it does not give the fine dispersion of manganese oxide required for complete reduction and rapid homogenization.

COMMERCIAL PLANT

A commercial plant to produce 50,000 tons per year of iron powder is now under construction near Amherstburg, Ontario. The feed will be steel turnings from the automotive industry. The plant will be completed toward the end of 1969 and powders will be available early in 1970.

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From scrap . . . to iron powder

By early fall 1969, the Peace River Process for making iron powder should be on stream. The initial plant will have a capacity of 50,000 tons of molding grade iron powder per year.

The process is unique. It all started as a project to produce iron economically from the large, low grade ore deposits in the Peace River area of Alberta, Canada. From the hydrometallurgical process envisioned, a high purity iron powder would result. This product, in turn, would be roll compacted, sintered, and hot rolled to form sheet and strip. But to prove out the process as quickly and economically as possible, the investigators decided to start with steel scrap instead of the iron ore. (There really was no difference except speed and development costs.)

Instead, Peace River Mining & Smelting Ltd. working with the Research Council of Alberta and Woodall-Duckham Ltd. of England, found that they had a low cost process for making very high purity iron powder from scrap material that is, presently, almost a drug on the market; chips, turnings, and light gauge trimmings. (They are continuing to work on the problem of producing iron powder from the low grade ore.)

How does the process work? Basically the process is quite simple. Dissolve the scrap in hydrochloric acid, crystallize out ferrous chloride, and dry it to form the dihydrate. This material is then reduced to iron by reacting it with hydrogen. The end product is an iron powder varying from 99.4 to 99.8 percent iron depending on the grade required. Hydrochloric acid is formed as a byproduct of the reduction step and is recycled back to the dissolver. The process is shown in simplified form in the flow chart in figure 1.

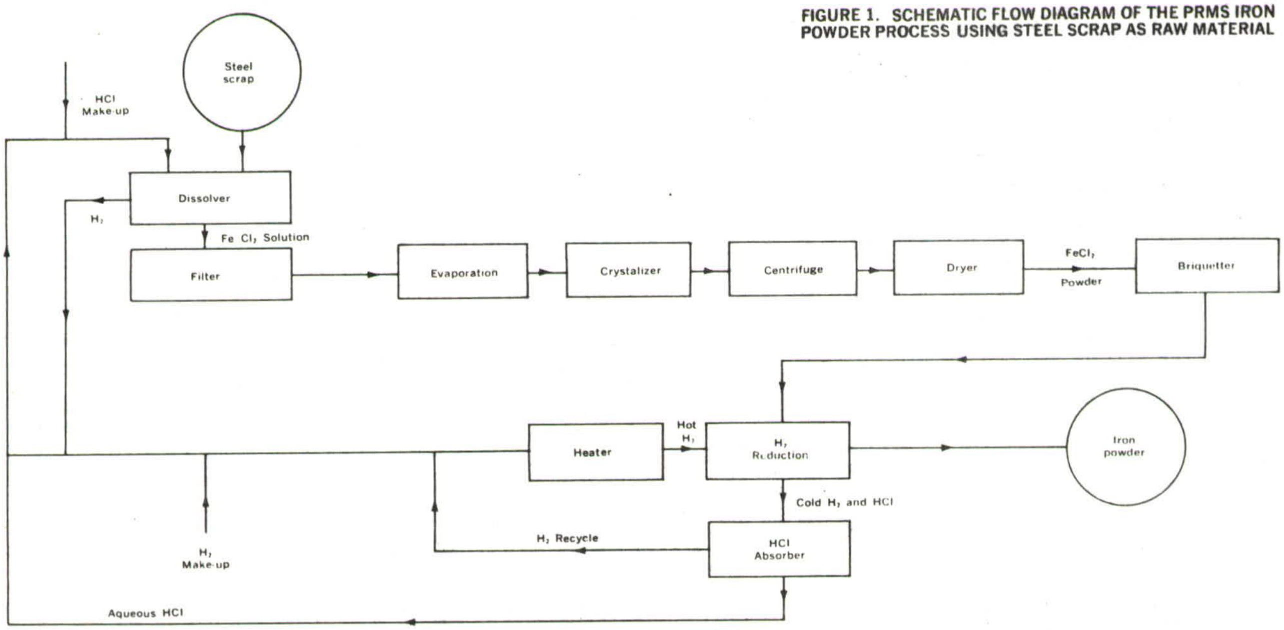
In theory the process is a completely closed cycle with only iron needed as a raw material since hydrogen is evolved when the iron is dissolved in acid. Practically, of course, small amounts of both hydrogen and hydrochloric acid will be needed for make-up to compensate for less than 100 percent reaction efficiencies and for normal losses due to leakage. The key, of course, is low cost scrap and that's why the initial plant will be located in the Windsor, Ontario area, just across the river from a huge scrap generator —the U. S. automobile industry.

Properties of the powder. Two grades of iron powder will be produced at the start. These are Peace River Powders 240 and 280, they differ slightly in chemistry and screen analysis, but the major difference is in the apparent density of the powders. Table I lists the properties of both powders as they are being produced now in the 5 ton per day pilot plant. When the main plant is in full operation it is confidently expected that an even better powder will be made. Specifically, the hydrogen loss will be lowered to about 0.1 percent and the carbon content lowered or perhaps eliminated.

For the design engineer. This new iron powder will have two important assets for the design engineer: A high degree of compactability so that parts can be molded to high density at low compacting pressures, or conversely much larger parts can be molded on the presently installed low tonnage forming presses. The second advantage will be price. Present indications are that the powder will sell for about 10 cents (Canadian) per pound, or roughly 9 cents per pound in U. S. money. At this cost, many parts which are

Dr. C. P. Gravenor, General Manager and Director of Research (left), D. H. W. Allan, Manager of Marketing, and G. R. Heffernan, President of Peace River Mining & Smelting, Ltd. (right) discuss a section of the 12 foot long model of the iron powder plant under construction at Amherstburg, Ontario, Canada.

FIGURE 1. SCHEMATIC FLOW DIAGRAM OF THE PRMS IRON POWDER PROCESS USING STEEL SCRAP AS RAW MATERIAL



Reprinted from Precision Metal, Oct., 1968.

mechanically feasible now, but priced too high, can be made from iron powder.

By using one percent of zinc-stearate lubricant and sintering for 30 minutes at 2050°F in an hydrogen atmosphere, usable strengths can be developed in the 240 powder even at quite low final densities as shown in figure 2.

The 280 powder shows about the same properties at the same densities, but when moderate amounts of copper are added there is an impressive increase in mechanical properties as shown by the two curves in figure 3.

The designer should note that the mechanical properties which have been given are for material essentially free of carbon. While data are not yet available on carbon additions, there is every reason to expect that significant increases in strength levels will result from small carbon additions and proper heat treatment.

For the parts producer. Other than the properties already given some data are available that are of importance to the parts fabricator. Figure 4 shows the increase in green density as a function of forming pressure and figure 5 shows green strength as a function of forming pressure.

Both powders shrink when sintered, but the amount of shrinkage is less than 0.2 percent at low densities and less than 0.05 percent at higher densities as shown in figure 6.

Equally important to the parts producer is the much lower ejection pressure required for this new powder as shown in figure 7. Based on the ejection pressure data it would seem that the amount of lubricant mixed with the iron powder might be greatly reduced.

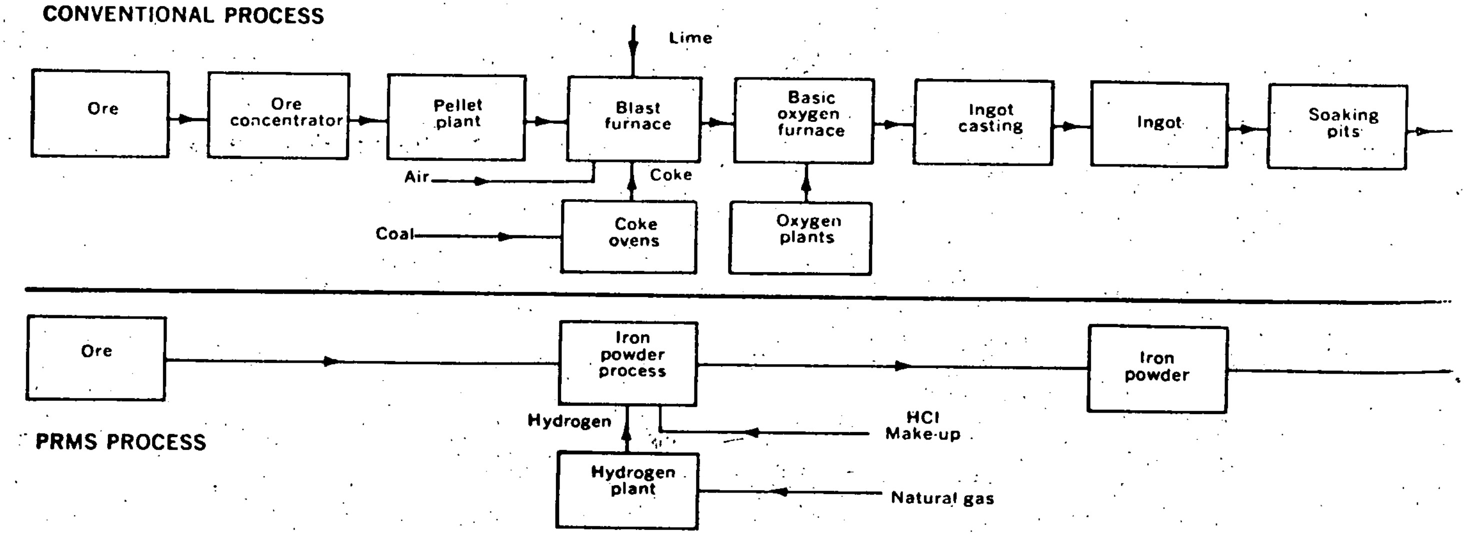
Markets. Three important markets seem likely targets for this new powder. The primary one, of course, is for parts which are larger than any now being made at moderate to high densities.

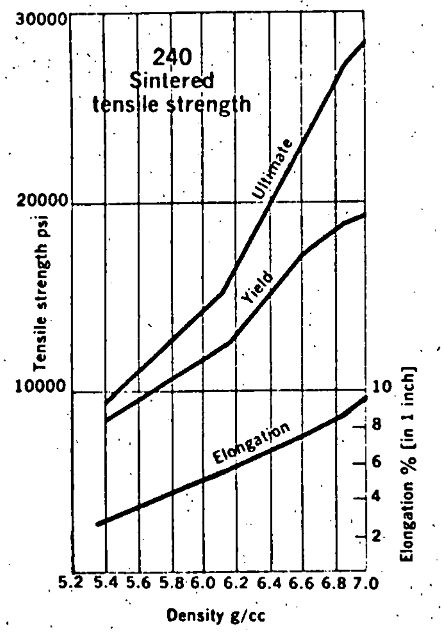
A developing market, particularly in the automotive industries is the use of powder metal preforms for closed impression die forging. This market will, for Peace River at least, probably have to wait for the development of pre-alloyed powders. However, some progress has been made using alloy mixtures with adequate diffusion occuring during sintering.

The third important market is the production of hot rolled strip (hot band) from metal powders. In this case considerable progress has already been made and pilot costs indicate that the process is quite feasible due to the low powder costs and reduced capital investment involved. Figure 8 gives comparative flow charts and illustrates the much simpler process possible through powder metallurgy compared to the conventional method. PM

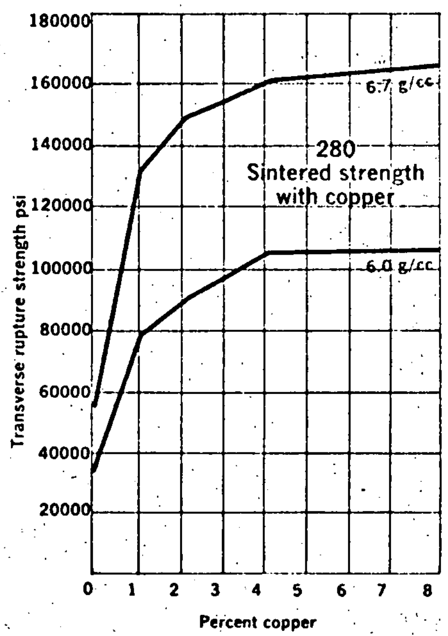
TABLE 1 — PROPERTIES OF PEACE RIVER IRON POWDERS

	240	280
Screen Analysis (weight percent)		
+ 100	1.5	2.5
- 100 +150	15.8	13.4
- 150 +200	23.1	21.4
- 200 +250	12.5	11.7
- 250 +325	18.0	22.5
- 325	28.6	28.1
Subsieve sizes (percent of -325 fraction)		
+ 30 microns	14.6	16.3
+ 20	8.6	7.8
+ 10	3.1	3.4
- 10	2.0	0.6
Apparent Density (g/cc)	2.41	2.79
Flow Rate (for 50 grams)	33.0 seconds	29 seconds
Chemical Analysis (weight percent)		
Manganese	0.273	0.19
Carbon	0.012	0.02
Sulphur	0.005	0.007
Phosphorous	0.0055	N. A.
Silica	0.006	0.006
Hydrogen loss	0.22	0.25

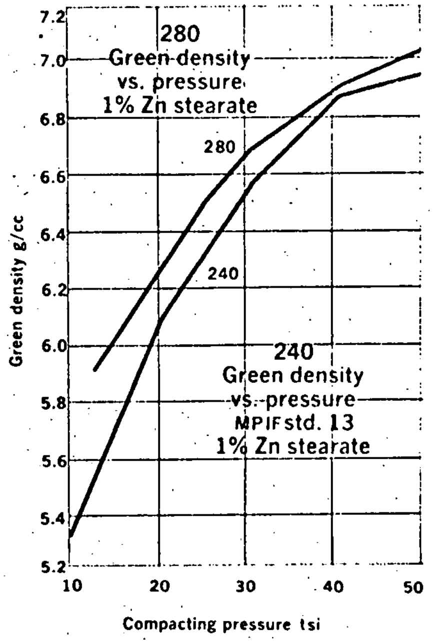




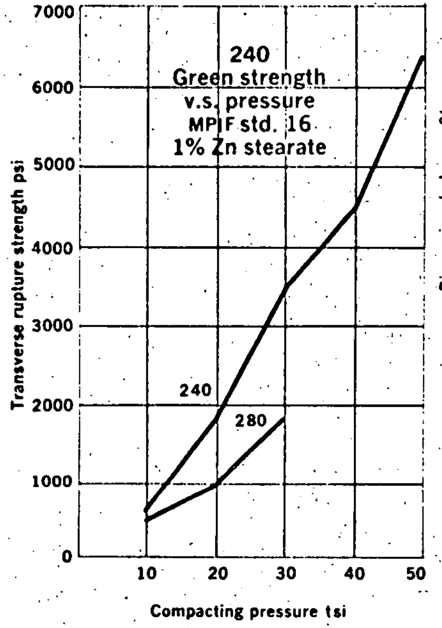
2. The mechanical properties of the pressed and sintered iron powder seem low, but this is a high purity iron, practically free of carbon. The high ductility even at low densities is notable.



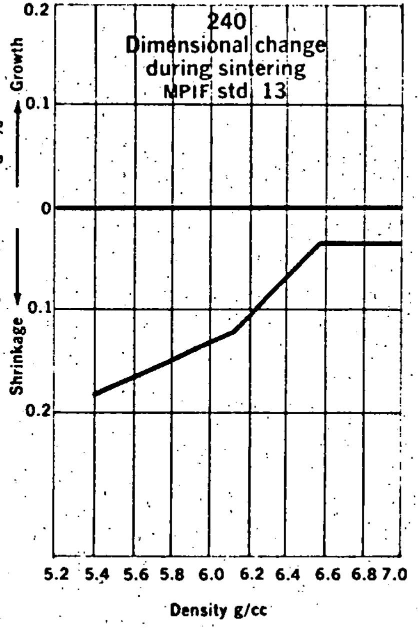
3. With copper added to the iron powder very high mechanical properties are secured at relatively low densities.



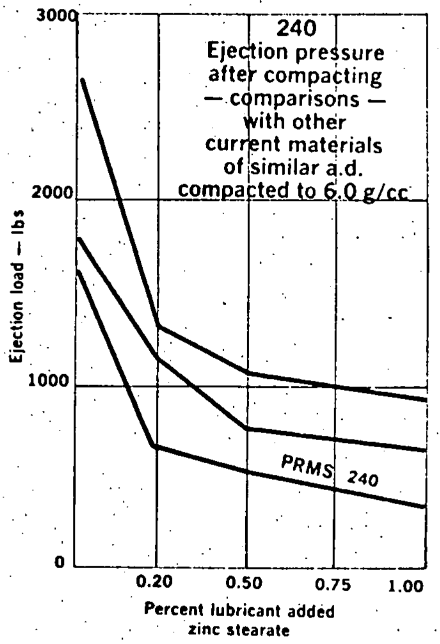
4. Good compressibility combined with high apparent density allows formation of high density parts at relatively low forming pressures.



5. Even at low compacting pressures the green strength of the compacts is good. Handling of the green compacts should present no problem.



6. Moderate shrinkage, especially at the higher densities will allow accurate dimensional characteristics.



7. The amount of lubricant needed for easy ejection may be drastically lowered with these new powders. This comparison is with two competitive iron powders widely used by the P/M parts industry.

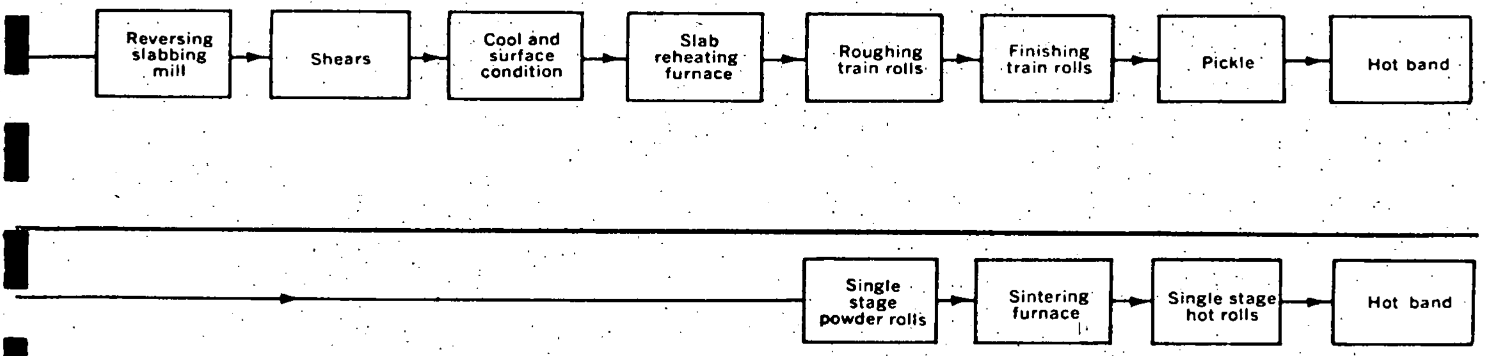


FIGURE 8 FLOW CHART OF INTEGRATED PLANT TO PRODUCE HOT ROLLED STEEL STRIP (HOT BAND)

Peace River Iron Powder Plant in Ontario

By our AMERICAN CORRESPONDENT

Peace River Mining & Smelting is currently building a high-purity iron powder production plant at Amherstburg, Ontario, having an annual capacity of 50,000tons. The installation employs the Peace River process, which is based on dissolving iron-bearing material in hydrochloric acid to produce a ferrous chloride solution. The end product is sponge iron varying from 99.4 to 99.5% purity. This sponge iron is crushed and blended to produce powders with specific characteristics.

PEACE RIVER MINING & SMELTING (PRMS), Edmonton, Alberta, owns extensive ore deposits in the Peace River area of Alberta, which are situated close to an inexpensive source of natural gas. The firm, in conjunction with the Research Council of Alberta and Woodall-Duckham, of Crawley, England, and assisted by grants from the Canadian Defence Research Board, has developed a hydrometallurgical process for producing iron powder from both this ore and other iron-bearing materials, including steel scrap. The use of iron powder for the production of structural parts, particularly by the automotive industry, is increasing rapidly. With a view to enlarging the potential markets for its powder, PRMS has carried out research and development on a semi-commercial scale on the rolling of iron powder directly into steel strip.

The company is now constructing an iron powder plant at Amherstburg, near Windsor, Ontario, with an initial capacity of 50,000tons/year to supply iron powder to the automotive parts industry in the Detroit-Windsor area and to other consumers in the United States and Canada. This plant will use low-grade steel scrap as its primary raw material and will enable PRMS to penetrate the rapidly growing North American iron powder market.

The distribution service facilities include two railway systems whose tracks cross at the corner of the plant property, well maintained secondary roads with ready access by both bridge and tunnel to major highway systems and to "Ontario's Main Street"—Highway 401. Docking facilities at Amherstburg and Windsor are available for the loading of ships using the St Lawrence Seaway system. The major automotive industry plants, which represent the larger individual consumers of iron powder parts, are located in the Detroit area across the river from Amherstburg.

Since the decision was made to proceed with the construction of the Amherstburg plant, there have been announcements by several major companies in the US and Canada that they are also proceeding with the construction of iron powder plants using a variety of different production techniques. While this rapid expansion of capacity may result in a temporary oversupply of iron powder in 1970, Peace River believes that its chemically-produced powders have a quality advantage and that it will be able to meet whatever competition develops.

PRMS PROCESS

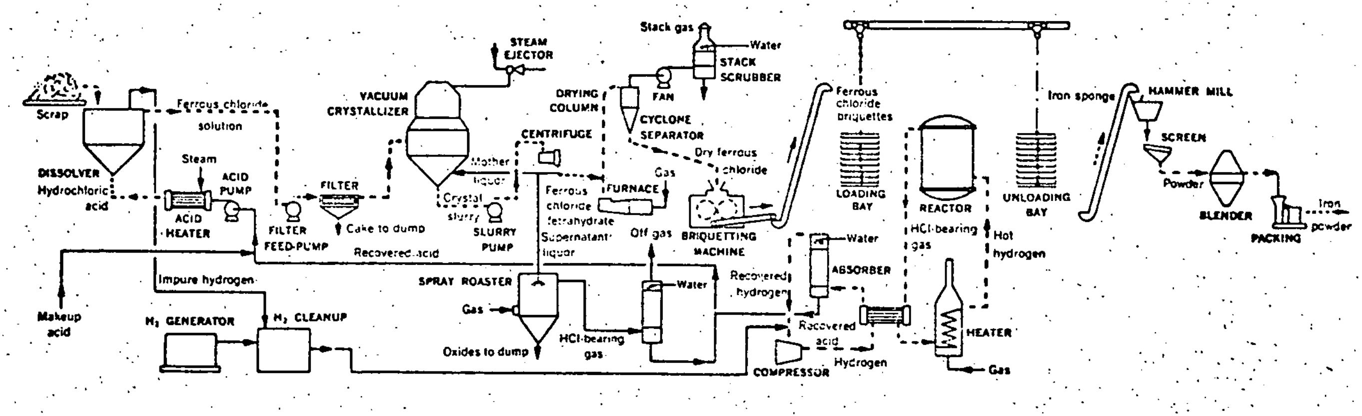
The process variables, especially those that regulate ferrous chloride reduction, were determined in a 5ton/day pilot plant that has been operating near Edmonton, Alberta, since early 1966.

The dissolution of the iron is the first step, (Fig1) the following reaction taking place at 194°F and atmospheric pressure:—

$$Fe^{++} + 2HCl \longrightarrow FeCl_2 + 2H^+$$

The subsequent reduction of the ferrous chloride is the

1 *Flow sheet of Peace River iron powder process.*



step causing major design problems. Essentially this is a high-temperature (1112 to 1472°F) step in which the above reaction is reversed by combining the briquetted ferrous chloride crystals with hydrogen at atmospheric pressure. The heat needed for reacting is about 37,000kcal/mole, but an additional 23,000kcal/mole are provided to preheat the feed and to evaporate the residual water.

The reduction step takes place in batch fashion inside fixed-bed reactors. An attempt was made to develop a continuous reactor, but this idea was dropped because of difficulties in handling hot ferrous chloride. In the chosen units, the bed depth is about 2ft; the charge is loaded and unloaded in a stack of trays. The overall process yield is higher than 93% in terms of iron input, and the final product is 99·4 to 99·95% Fe.

At the new plant, scrap will be fed into a dissolver and reacted with hot acid containing 20 to 25% by weight HCl. Iron dissolves in the acid, and the ferrous chloride solution containing 3 to 5% residual acid is pumped to a filter that removes insoluble impurities.

The crystallisation of ferrous chloride tetrahydrate is done in a vacuum crystalliser, and separation from the mother liquor is carried out in a subsequent centrifuge. The mother liquor is returned to the vacuum crystalliser, but the supernatant liquor from the centrifuge, which contains soluble impurities, such as metal chlorides, is sent to a spray roaster. Here, the impurities are hydrolyzed to oxides; and the hydrogen chloride released is recovered for use in the dissolving step.

Because these vessels handle ferrous chloride solutions, they must use rubber-lined steel, whereas the piping is made of glass-fibre reinforced plastic. The pump internal parts are made of either silicon iron or graphite and the centrifuge has an all-titanium front end.

To improve both the strength of the recovered acid and the quality of briquettes, the tetrahydrate is dried to the dihydrate stage in a dispersion dryer. A roll press then shapes the hot, dry material into pillowlike briquettes designed to improve gas-solids contact in the reduction stage.

$FeCl_2$ briquettes are charged into trays and loaded into the reactor; meanwhile, hot hydrogen is injected through the bottom of this unit. As the reaction proceeds at temperatures up to 1472°F, HCl gas issues from the top of the reactor. After heating hydrogen from the compressor the gas is sent to an absorber. Here, hydrogen is obtained from the top, recompressed, and returned to the reduction step; liquid HCl recovered from the bottom is sent to the dissolver.

Since hydrogen is a by-product of the dissolving step, it is recovered, purified, and used in the reduction stage. Because the raw material is scrap iron, only a small hydrogen generator will be needed to make up for losses. On the other hand, if iron ore is used, hydrogen must be produced elsewhere in a separate facility. The material that leaves the reactor is a mass of sponge iron. This is ground, screened and blended to give a specified product.

POWDER PROPERTIES

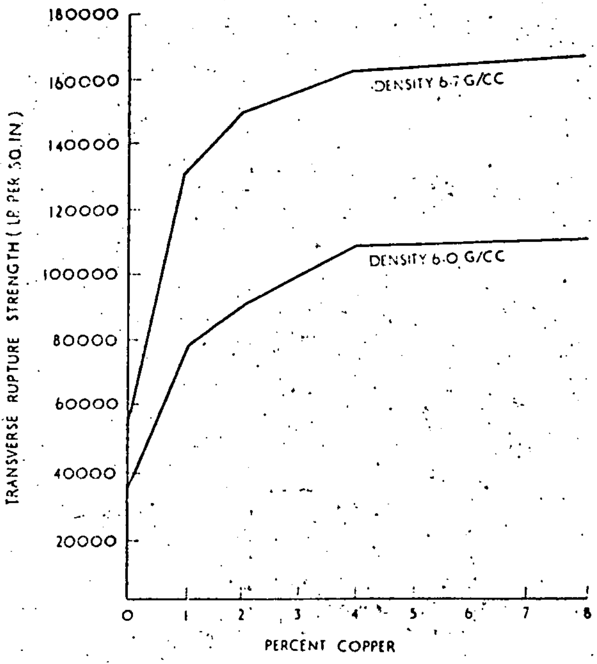
Two grades of iron powder will be produced at the start. These are Peace River Powders 240 and 280; they differ slightly in chemistry and screen analysis, but the major difference is in the apparent density of the powders. Table 1 below lists the properties of both powders as they are being produced now in the 5ton/day pilot plant. When the main plant is in full operation it is confidently expected that an even better powder will be made. Specifically, the hydrogen loss will be lowered to about 0·1% and the carbon content lowered or perhaps eliminated.

Table 1.—Properties of Peace River Iron Powders.

	240	280
Screen analysis (weight %):		
+100	1·5	2·5
−100 +150	15·8	13·4
−150 +200	23·1	21·4
−200 +250	12·5	11·7
−250 +325	18·0	22·5
−325	28·6	28·1
Subsieve sizes (% of -325 fraction):		
*30 microns	14·6	16·3
+ 20	8·6	7·8
+ 10	3·1	3·4
− 10	2·0	0·6
Apparent density (g/cc)	2·41	2·79
Flow rate (seconds for 50g)	33·0	29
Chemical analysis (weight %):		
Manganese	0·273	0·19
Carbon	0·012	0·02
Sulphur	0·005	0·007
Phosphorous	0·0055	N.A.
Silica	0·006	0·006
Hydrogen loss	0·22	0·25

This new iron powder will have two important assets for the design engineer: a high degree of compactability so that parts can be moulded to high density at low compacting pressures, or conversely much larger parts can be moulded on the presently installed low-tonnage forming presses. The second advantage will be price. Present indications are that the powder will sell for about 10cents (Canadian)/lb. By using 1% zinc stearate lubricant and sintering for 30min at 2050°F in a hydrogen atmosphere usable strengths can be developed in the 240 powder even at quite low final densities. The 280 powder shows about the same properties at the same densities, but when moderate amounts of copper are added there is an impressive increase in mechanical properties as shown by the two curves in **Fig2.** It should be noted that the mechanical properties which have been given are for material essentially free of carbon.

2 *Transverse rupture strength of No280 powders with copper additions.*



While data are not yet available on carbon additions, there is every reason to expect that significant increases in strength levels will result from small carbon additions and proper heat treatment.

Both powders shrink when sintered, but the amount of shrinkage is less than 0·2% at low densities and less than 0·05% at higher densities. Equally important to the parts producer is the much lower ejection pressure required for this new powder. Based on the ejection pressure data available it would seem that the amount of lubricant mixed with the iron powder might be greatly reduced.

POWDER MARKETS

Three important markets seem likely targets for this new powder. The primary one, of course, is for parts which are larger than any now being made at moderate to high densities. A developing market, particularly in the automotive industries, is the use of powder metal preforms for closed impression die forging. This market will, probably have to wait for the development of pre-alloyed powders. However, some progress has been made using alloy mixtures with adequate diffusion occuring during sintering. The third important market is the production of hot rolled strip from metal powders. In this case considerable progress has already been made and pilot costs indicate that the process is quite feasible due to the low powder costs and reduced capital investment involved.

PRODUCTION OF STEEL STRIP

PRMS has installed a semi-commercial mill at its pilot plant and successfully produced, from its iron powder, high-quality steel strip 6in wide and 0·040in thick. The powder rolling process consists in feeding the iron powder into a hopper that rests on two rolls placed side by side, and as the rolls rotate, the powder falls into the nip of the rolls and emerges below as a green strip compact. It is then fed continuously into a furnace having a reducing atmosphere where it is heated and minor oxide impurities eliminated. The hot strip then passes continuously into a second rolling mill where it is brought to full density in one rolling pass. The product of this simple process is equivalent to, or better than, hot-rolled strip made by conventional methods, can be made very thin, and does not require the usual acid pickling operation to remove scale before further processing.

The equipment for powder rolling is much less expensive than the requirements for the rolling of cast ingot or slab in the currently conventional processes. The high purity level of the iron powder and the decreased amount of rolling both contribute to the elimination of the undesirable directional characteristics found in strip made by present practices. Cold rolling of the hot-rolled strip made from iron powder requires essentially the same facilities as used in conventional practice.

The following are the approximate average yields based on iron content at each stage of the operation. In the "conventional process", ore to pellets 90%, pellets to blast furnace iron 93%, blast furnace iron to steel ingots 92%, ingot to hot rolled strip 70%. From ore to hot rolled strip the cumulative yield is of the order of 54%. In the "PRMS process", ore to iron powder 80%, iron powder to hot-rolled strip 95%, giving a cumulative yield of about 76%. The results of a comparative study have shown that the "PRMS process" requires significantly less capital cost per annual ton of production and the direct operating costs are marginally smaller than for the "conventional process". This indicates that the "PRMS process" for producing hot-rolled strip can be competitive with the "conventional process" at a much reduced production rate.

Davy-Ashmore Foundry Companies Reorganised

TO ACHIEVE higher efficiency and profitability through specialised production, Davy-Ashmore has planned a general reorganisation for the four foundry companies in the group. The programme will take one and a half years to complete.

It is understood there will be some redundancy in the case of the two companies operating on the north-east coast, but to what degree is not yet known since normal losses of personnel during the long period of reorganisation, coupled with retraining and interdepartmental transfers, will help to mitigate this.

The largest roll manufacturing unit in the foundries group, Armstrong Whitworth (Metal Industries), at Gateshead and Jarrow, will be futher enlarged by transferring the roll-making business of Davy and United Roll Foundry from Billingham to Gateshead and Jarrow.

In order to allow roll manufacture to proceed unhindered by other activities in Close Works, Gateshead, and the Jarrow foundry, the production of rock crushing machinery will be completely segregated from roll making within the Close works and operated as a separate and fully autonomous business in respect of management and production facilities. All other activities within the Armstrong Whitworth company will be discontinued.

Davy and United Roll Foundry, Billingham, Teesside, will concentrate on steel castings, in which it has established a reputation, while the general steel casting business carried on at the Jarrow works of Armstrong Whitworth will be transferred to the Billingham foundry. This is intended to result in a substantial increase in work load, which will provide a reliable foundation for further growth.

John M Moorwood and the Brightside Roll Foundry, both in Sheffield, will continue specialising in rolls, but with closer management and technical co-operation.

Top level management changes are also envisaged in the reorganisation. Mr F L Turnbull, managing director of Armstrong Whitworth, reached retirement age last year but consented to continue until the reorganisation scheme had been formulated.

Mr Turnbull has now retired from the post of managing director, but remains for a period as a non-executive director.

Mr H H Utley, managing director of Davy and United Roll Foundry, has been appointed managing director of Armstrong Whitworth (Metal Industries), responsible for the management of both the north-east companies and for that part of the reorganisation which involves them.

Mr J O Taudevin, chairman and managing director of John M Moorwood, has become managing director of Brightside Roll Foundry responsible for the management of both the Sheffield roll-making companies.

Continuous Casters Galore

WITH THE RECENT commissioning of a two-strand continuous casting machine, built by Officine Meccaniche Danieli & C, SpA, of Buttrio, Italy, at the works Ferriere Acciaierie Sarde, SpA, at Cagliari, Italy, the number of Danieli continuous casters in operation is now 21 with a total of 52 strands. This latest installation produces billets of 90mm sq (3½in sq) and 140mm sq (5½in sq). At the present time Danieli has a further seven continuous casters for a total of 19 casting strands under construction.

VISIT 9: SWINDELL-DRESSLER COMPANY

Situated at
441 Smithfield Street,
PITTSBURGH Pa.

DATE: Friday 13 November 1970

MET: Mr Ken A. Jackson - Manager, Process Engineering Department

1. INTRODUCTION

Mr Litton had advised that he would be out of the country all the week, but Mr Jackson filled in. He said if I required further information after returning to Australia it could be obtained either direct or via Mr Hutchison in Melbourne who had recently been to Pittsburgh. He supplied some papers and a brochure regarding the HyL process.

2. ECONOMICS

Table 3 (page A-119) in the paper by Gearhart and Jackson gives data for a modern plant to treat either lump ore or pellets. The original HyL plant had a capacity of 200 tons per day. The paper in EMJ giving a cost breakdown does not include ore-yard labour. The shift crew consists of 6 to 7 men so that at 9,000 paid man-hours per man per shift (which includes night differential) the labour per annum is 54,000 to 63,000 man-hours. In reply to a question on degree of instrumentation Mr Jackson thought that the crew could be reduced by several men so that 49,000 man-hours per annum could be assumed. A capital cost of $US8.5 million can be taken for the plant for which data is given on page A-119.

3. TECHNICAL ASPECTS

If lump ore is used, the ore must be hematite, >60% Fe, minus 2 in. plus $^1/_2$ in. A modified Gakushin test using some hydrogen is used to assess reducibility. The metallised pellets produced would probably be satisfactory as cupola feed as little breakdown occurs with Alzada pellets during reduction, but he would have to confirm this with Hylsa. Lump ore does disintegrate somewhat on reduction. With lump ore there has been a variation in degree of reduction between top and bottom of the reactor but with pellets metallisation has been over 90%, at the top and 84-85% at the bottom and subsequent handling effects blending. The pellets flow readily out of the reactor, whereas lump ore had to be extracted, and are bucket-charged to the electric furnaces. The pellets should be indurated just sufficiently for handling as over-induration glazes the surface and hinders reduction. Alzada pellets produced on a Lurgi grate are used at Monterrey and Puebla. Specifications call for a compressive strength of 300-500 lb and 200 lb might be satisfactory. Cold-bonding processes are a possibility, but a shaft furnace is the cheapest, on a small scale, of the high-temperature processes. HyL aim for 1.5-2.5% carbon in their product, but not for high metallisation (85%).

Refining of the steel is carried out in the electric furnace. El Encino lump ore produces sponge containing approximately 0.4% P which constitutes 60% of the electric furnace charge, the remaining 40% being scrap. When reduced pellets of lower phosphorus content were used to constitute 85% of the electric furnace charge there was a tremendous reduction in heat time and the metal was right on specification without refining.

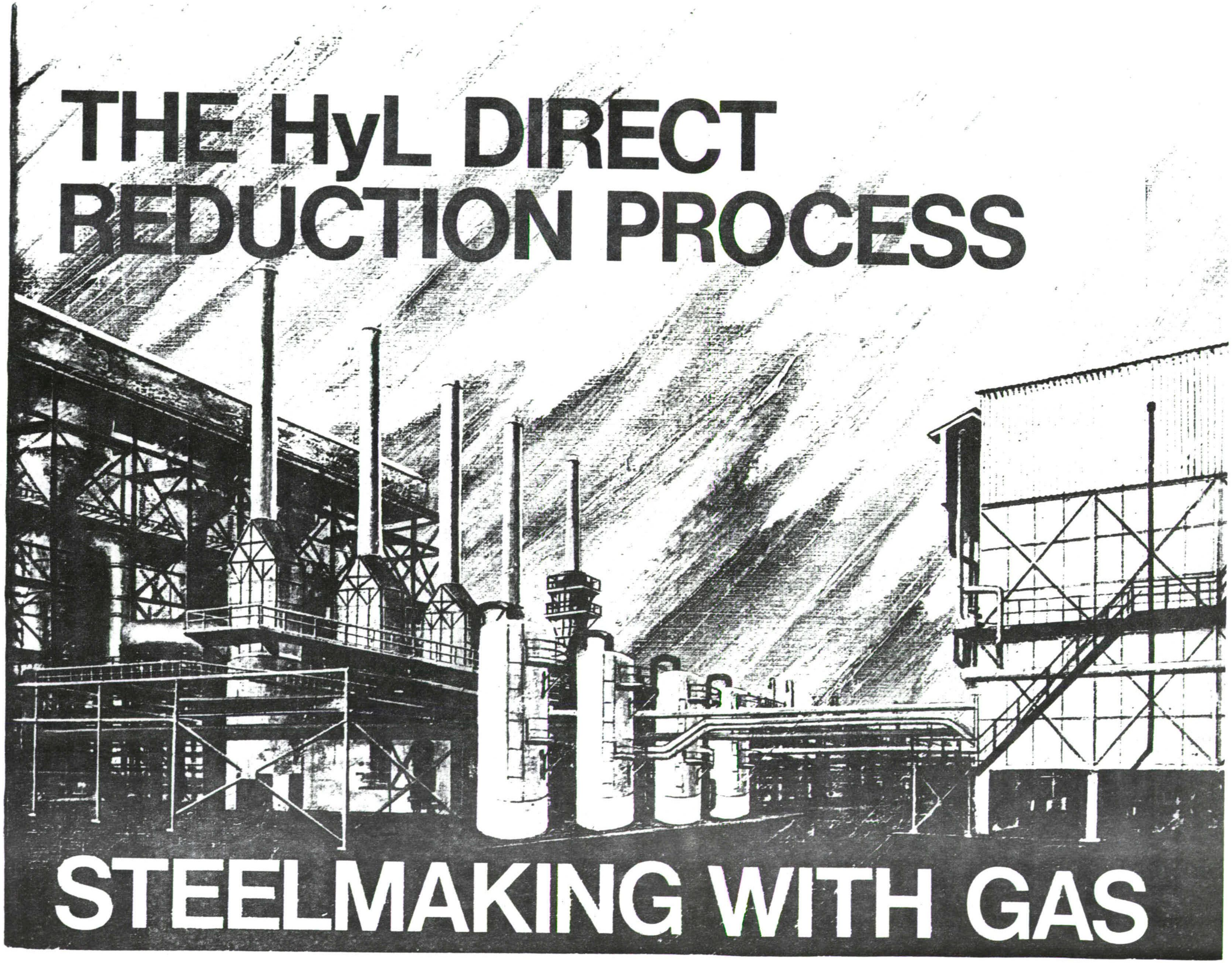
Some tests have been made of the use of HyL sponge for copper cementation and indications are that it should be crushed.

A plant at Salvador, Brazil to use LVRD Itabira pellets is scheduled to be complete in 2 years' time.

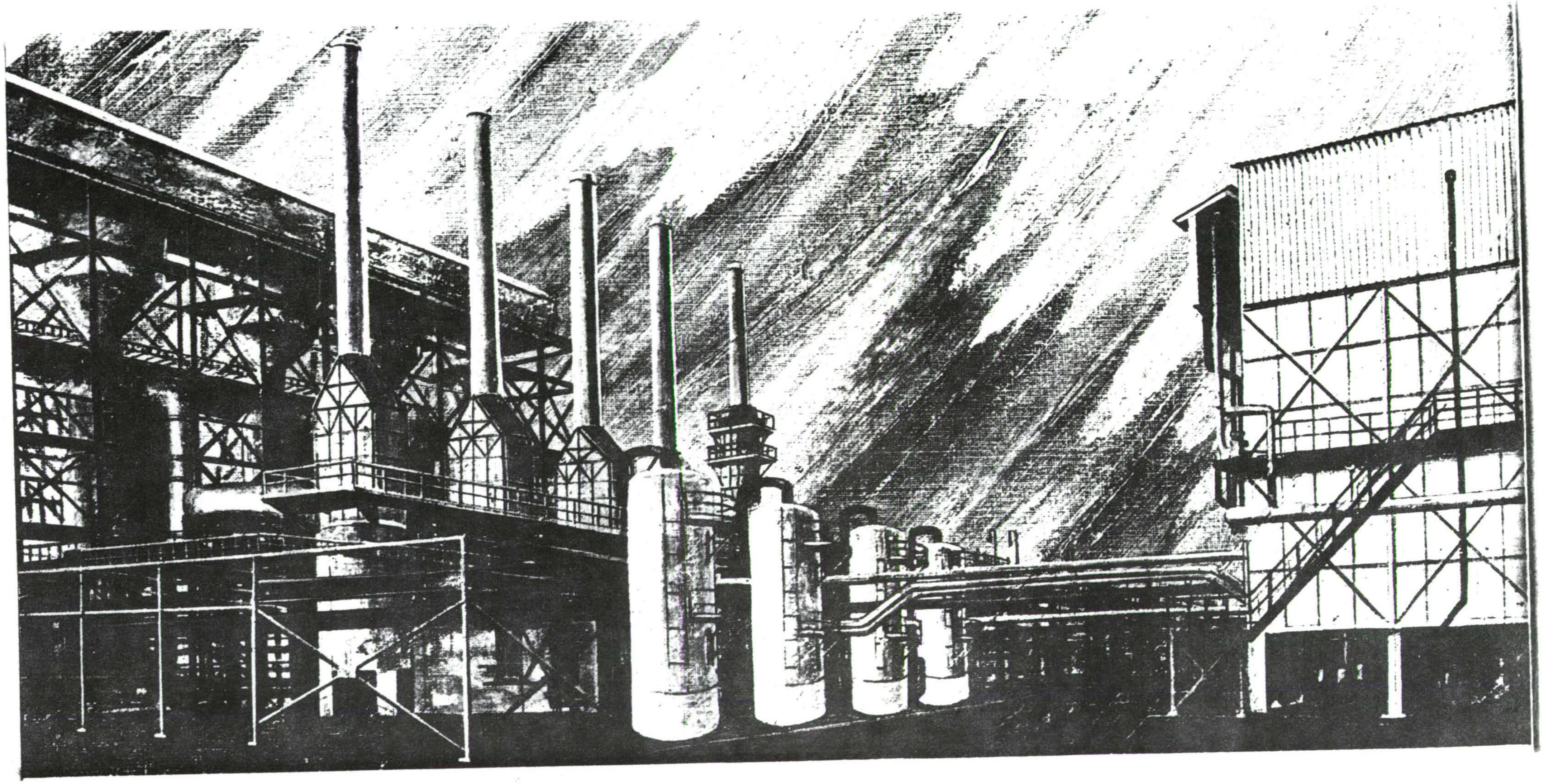
4. FURTHER ACTION

Preliminary bench-scale or "bag" tests costing $2,000-$3,000 can be carried out on a 50lb sample of ore to see whether an ore is definitely unsuitable for the HyL process, or not. A pilot-plant run at Monterrey would cost $30,000-$60,000 excluding freight and require 50 metric tons of usable material. An estimate of production capacity for a given ore can then be made. Melt-shop tests would have to be arranged with Hylsa.

THE HyL DIRECT REDUCTION PROCESS



STEELMAKING WITH GAS



THE HyL DIRECT REDUCTION PROCESS

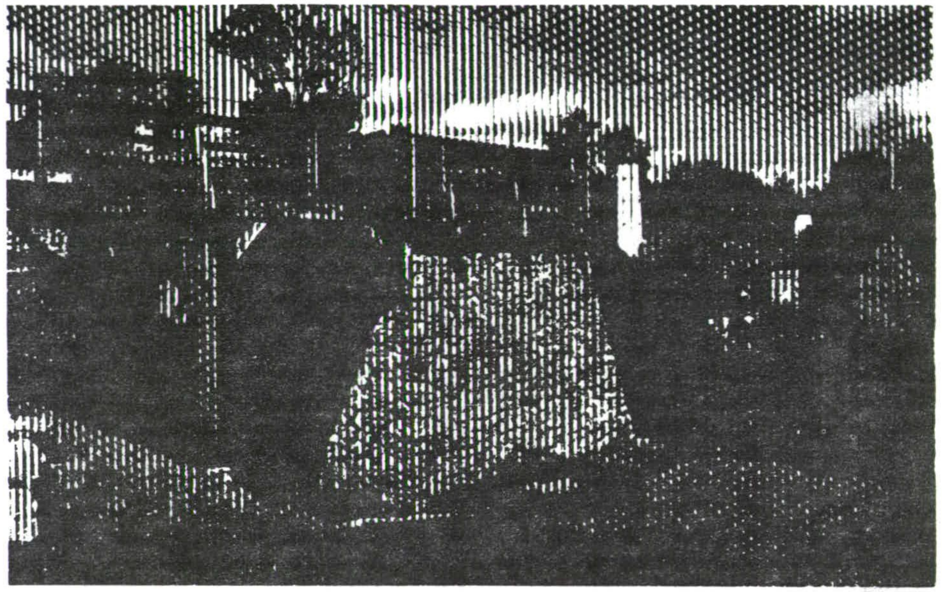
The evolution of the reduction process of iron ore began around 1000 B.C. in Asia Minor, when it was discovered that workable iron could be produced by burning wood charcoal with high grade ore. For the three hundred years which followed, iron was used only moderately for the fashioning of some war implements until the Assyrians recognized its full value and converted their metallurgy from bronze to iron, ushering in the Iron Age.

It took some 2500 years to result in the first major innovation when, in the 18th century, British ironmaster Abraham Darby discovered the use of metallurgical coke instead of wood as a reducing agent. In both Europe and Asia, forests surrounding iron-producing cities had become so depleted, that these centers faced an iron famine. Darby's discovery provided the basis of the modern blast furnace technology which, in the United States, swung over to the use of coke during the first half of the 19th century.

The use of coke and the blast furnace led to the incredible expansion in the use of steel during the industrial revolution; but this technology became its own limiting factor. Blast furnace reduction of iron ore depended on the availability of metallurgical coke, required major capital investments to construct the furnaces and coke ovens, and demanded a high market to make the operation economical.

The electric smelter appeared in Europe in the early 20th century partly in answer to the limitations of the blast furnace technique. It could use lower grades of reducing media, but it required vast quantities of electric power, keeping its operating expenses high. The low shaft furnace which, by virtue of its reduced height, does not need high grade metallurgical coke to support the charge, has been considered as an alternative, but it too has not proven sufficiently economical or efficient for wide-scale use.

In the past few decades, new techniques for reducing iron ore have been developed, particularly "direct reduction" processes. Of the many hundreds of direct reduction schemes proposed using a variety of fuels and reducing media, one has proven commercially successful—HyL direct gaseous reduction.



Natural gas is the basis for the HyL process. It is reformed by mixing with steam and catalytically treated in a furnace to produce hydrogen and carbon monoxide for use in ore reduction; the residual gas, after reducing the ore, is utilized as fuel for the gas reforming furnace.

Where gas is not available, other reformable hydrocarbons such as naphtha can be used as the source of hydrogen and carbon monoxide.

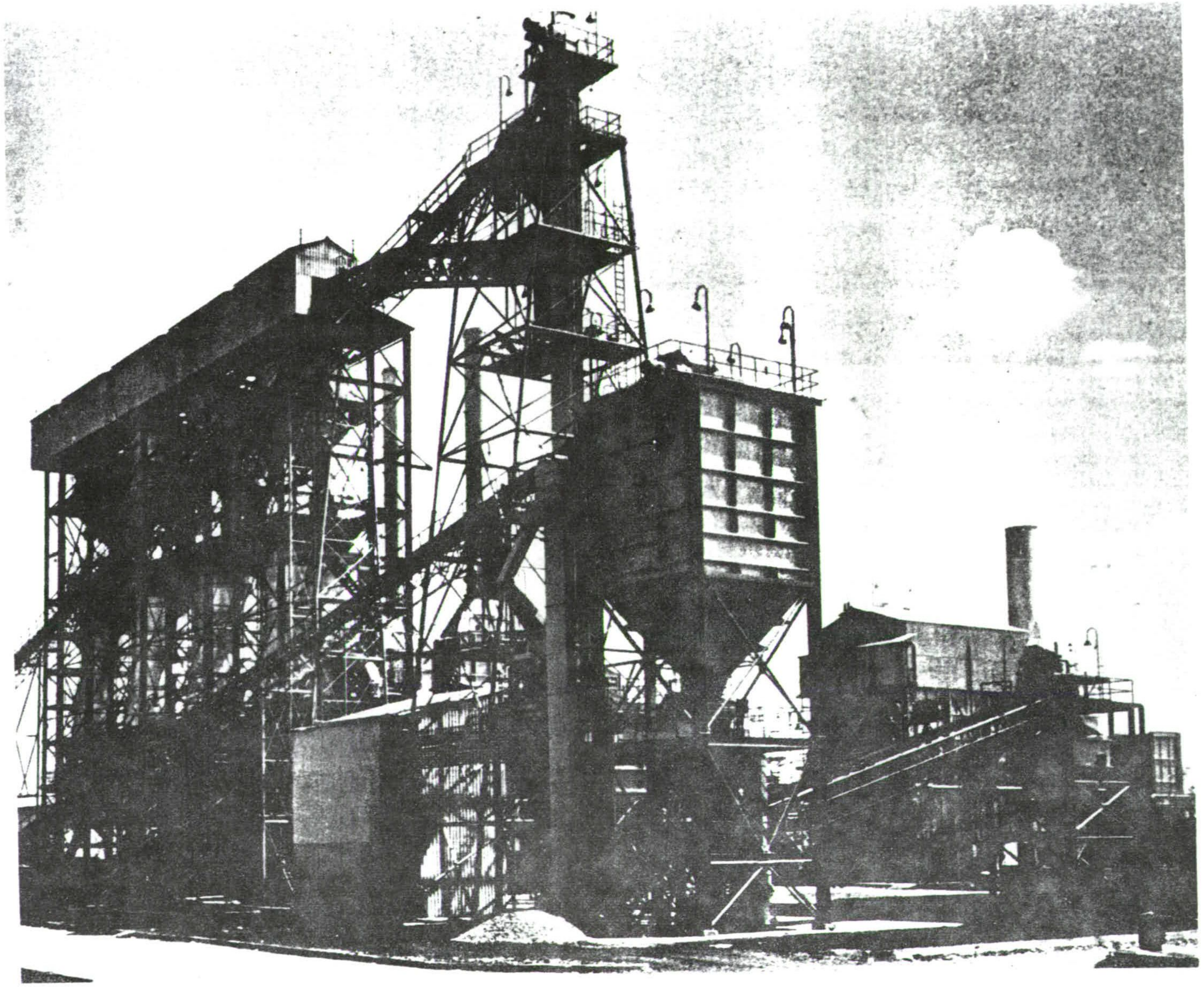
The HyL process frees the user from the dependence upon solid fuels and from the often expensive use of electric energy to reduce iron ore. In addition, the HyL process permits economical operation of steel plants with a wide range of output capacities and products.

To investment interests in nations with growing economies where a definite demand for steel exists, the HyL process offers a means to the economical formation of new steelmaking industries. With a natural gas direct reduction process available, a nation undergoing increased industrialization can take a new look at the problem of how to create a steel-based economy.

The 500 Ton/ Day Plant

A commercial 500-ton/day direct reduction plant has the general appearance of a modern chemical plant, rather than the traditional steel mill. Essentially, the plant is divided into two main sections; the gas-reforming and the ore-reducing systems.

Highly automated, the entire plant operates with a crew of only six men per shift. All valving is operated from a central control room employing a graphic control board system.



Gas-Reforming System

Natural gas piped into the 500-ton-per-day HyL plant is first desulfurized in an activated carbon desulfurizer. It is next mixed with steam in a molar steam/gas ratio of 2.1 to 1.0 and the mixture is heated to 430°C prior to entering the gas-fired vertical radiant tubes in the principal portion of the reforming furnace. A nickel catalyst and reformation temperature of about 850°C (1562°F) speed the reaction of gas with steam to produce reducing gases as follows: $CH_4 + H_2O = CO + 3H_2$. Composition of the reformed gas is approximately 74% H_2, 13% CO, 8% CO_2 and 5% CH_4.

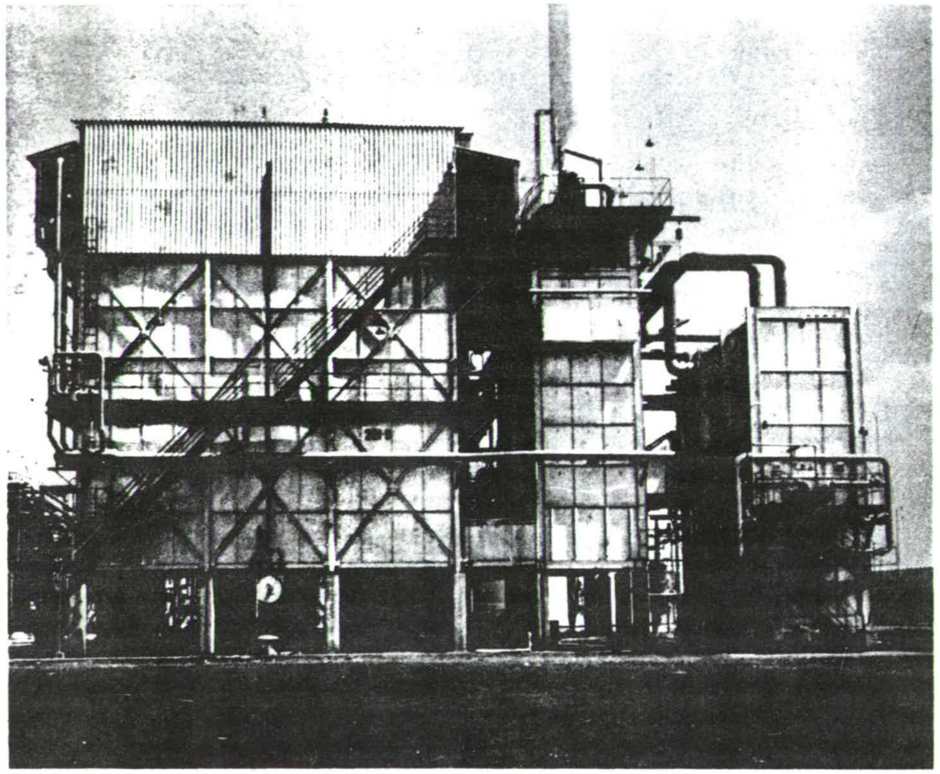
As the hot gas leaves the reforming furnace, it passes through a waste heat boiler or heat exchanger where the excess heat is recovered by converting water into steam for recycling. Most of the steam is used in the reforming operation to produce reducing gas; the remainder is utilized elsewhere in the plant to provide power for turbine driven pumps and other equipment.

A coil for superheating the steam is located in the preheater portion of the reforming furnace. This arrangement provides a recuperative heating system; steam is superheated and simultaneously the mixture of desulfurized gas is cooled.

The entire plant is self-sufficient and requires only natural gas and water from external sources.

Leaving the heat exchanger at a temperature of about 230°C (471°F) the gases are further cooled in the reformer quench tower. From the cooling tower, the gas is piped into the ore reduction reactor systems.

Though the 500-ton-per-day plants in Mexico use natural gas, the HyL process is flexible, and plants can be designed to handle other hydrocarbons, if available resources and other economic considerations so dictate.



Ore-Reducing System

Ore reduction occurs in four reactor vessels, which are coupled to gas preheaters and quench towers. Each reactor system is interconnected by piping. The piping is manifolded to allow the bypass of any reactor system for the purpose of maintenance.

Conveyors move the ore from storage bins to charge hoppers. A complete reducing cycle for each reactor requires 12 hours and the stages for conversion are as follows:

(1) Preheating raw ore with partially spent reducing gas (secondary gas)—3 hours.
(2) Reducing the preheated ore to sponge (exposure to primary gas)—3 hours.
(3) Cooling the sponge with fresh, cool reducing gas—3 hours.
(4) Unloading the cooled sponge iron and reloading with raw ore—3 hours.

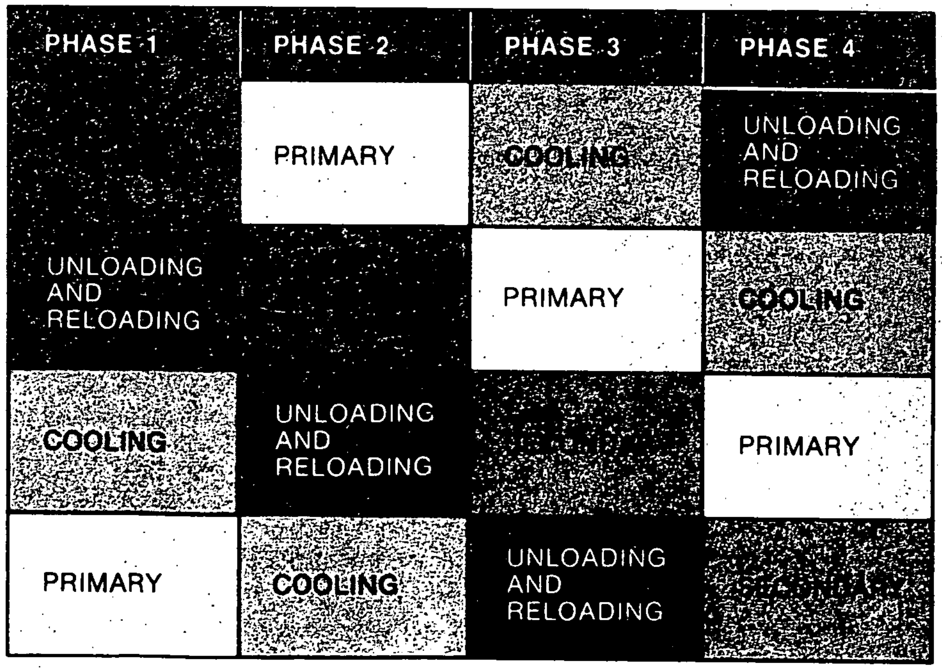
To accomplish the above cycle on an efficient, continuous basis, the flow of reducing gas through the reactors or reduction system is in accord with this pattern.

Heat is introduced to the reactor through a special patented procedure which was developed by the engineering staff of Hojalata y Lamina, S.A. After the reducing gas leaves the quench tower at each reactor, as previously described, it is heated in a conventional tubular gas heating furnace to a temperature of 1300°F to 1500°F. At the same time a stream of air is also heated to approximately the same temperature in another furnace of a design similar to that used for preheating the gas. A carefully controlled quantity of this preheated air is then injected into the reducing gas stream, causing the combustion of a portion of the reducing gas. The resulting heat of combustion raises the temperature of the gas mixture from 1800°F to 2250°F. This hot gas then enters a reactor containing hot, partially reduced ore which has just completed the secondary stage. In this primary stage the bulk of the ore reduction takes place and a considerable quantity of water vapor and carbon dioxide appear in the exit gases. As in the case of the gases leaving the cooling stage, the gases from the primary stage are also cooled in a quench tower to bring about the removal of water by condensation. The gas is again heated to a high temperature and flows into a reactor which has just been charged with fresh cold ore. During the ensuing secondary stage, the ore is heated and partially reduced by contact with the hot gas. The gases which emerge from the secondary stage still contain appreciable quantities of hydrogen and carbon monoxide and are used as fuel to supply heat to the various furnaces and heaters in other parts of the plant. The calorific power of this final gas is not quite sufficient to supply all of the heat needed in the process and it is therefore necessary to introduce some additional natural gas into the fuel gas system.

The reactors are switched from one stage to another of the 12-hour cycle by motor operated valves remotely activated from a panel in the control room.

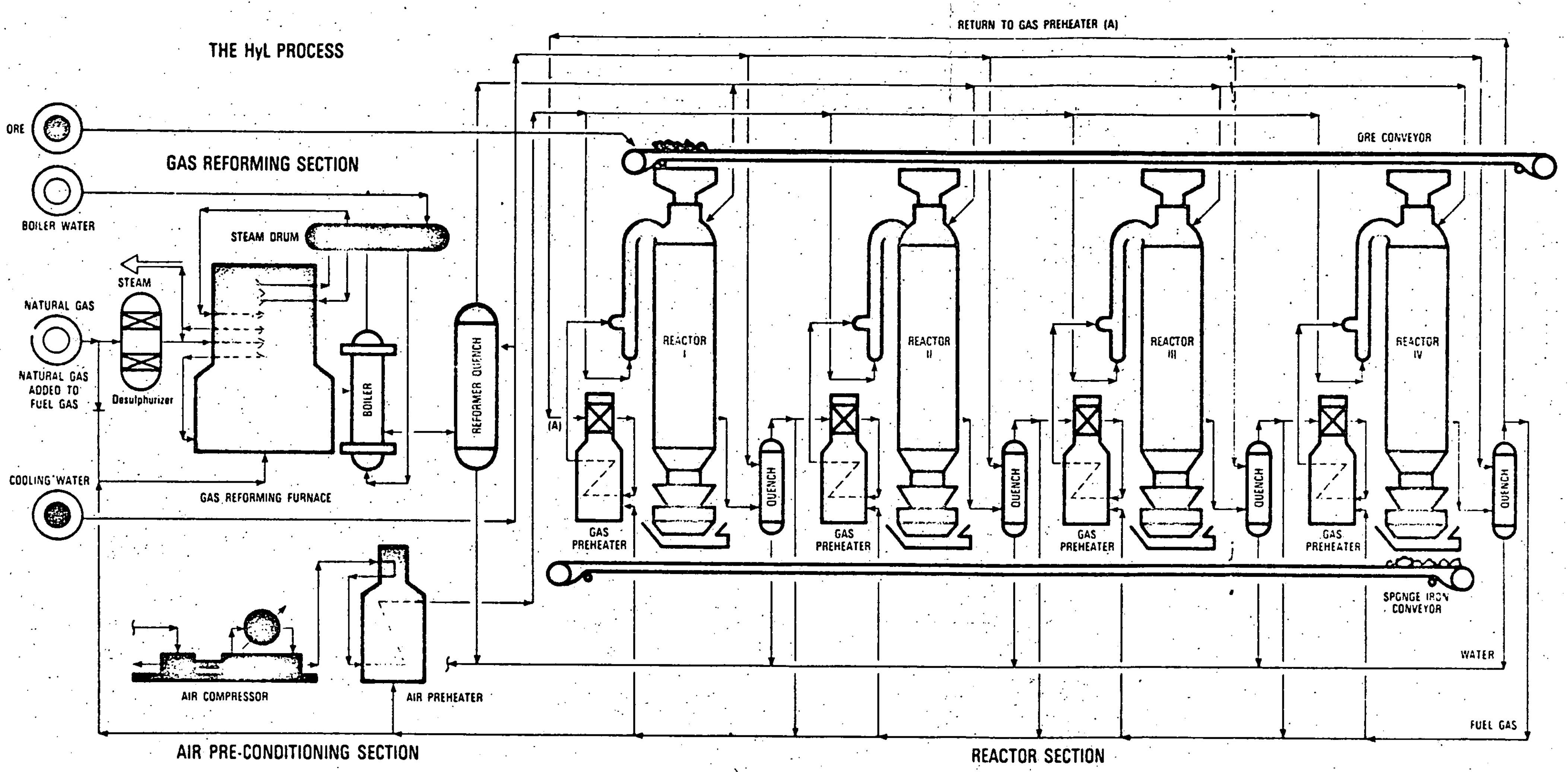
The reduction of ore with reducing gases in the temperature range of 1600 to 1900°F is accomplished by the following reactions:

$Fe_2O_3 + 3H_2 = 2Fe + 3H_2O$
$Fe_2O_3 + 3CO = 2Fe + 3CO_2$
$3Fe + 2CO = Fe_3C + CO_2$
$3Fe + CH_4 = Fe_3C + 2H_2$



PHASE 1	PHASE 2	PHASE 3	PHASE 4
[illegible]	PRIMARY	COOLING	UNLOADING AND RELOADING
UNLOADING AND RELOADING	[illegible]	PRIMARY	COOLING
COOLING	UNLOADING AND RELOADING	[illegible]	PRIMARY
PRIMARY	COOLING	UNLOADING AND RELOADING	SECONDARY

THE HyL PROCESS



Production

Millions of tons of HyL sponge have been converted to quality steel by melting in electric arc furnaces. These furnaces vary in size from 6 to 25 tons of metal per charge. Most heats contain from 50 to 75% sponge iron, with the balance made up of scrap.

The melt-down with sponge iron is characterized by its generally lower electric and thermal conductivity, as compared to all-scrap melting. A longer melting time is required, and additional energy is needed for melting the greater amount of slag constituents. Refractory life is often shorter due to acid slag conditions.

With the presence of sponge iron, however, there are fewer voids in the charge, and so less radiation escapes to the refractory lining of the furnace; moreover, the electrodes are completely surrounded by the charge, and this causes increased heat absorption by the charge.

Unlike the HyL process, the blast furnace and electric arc smelting processes are both based on the use of solid fuels to produce molten pig iron. Yet where blast furnaces are concerned, the need for large amounts of capital, the blast furnace's metallurgical coke requirements, and the large output required to make the blast furnace economical, have deterred many nations from developing their own steelmaking industries. Similarly, wide use of electric smelting has also been limited because of its heavy demand for electricity and its relatively high cost of operation in most countries.

These conditions, plus the ability to control the slag in melting, tend to offset the effects of lower conductivity and larger slag volume when melting sponge iron.

Any comparative data, however, must be based on the entire steelmaking process in order to realize sponge iron's true significance. While it costs slightly more to convert sponge iron to steel than it does to convert scrap, the over-all cost of steel per ton is lower if the steelmaking facility incorporates an HyL ore reduction unit with its inherent savings in over-all capital cost and the cost of energy.

The table on page 14 illustrates the performance of electric furnaces when melting mixtures of sponge iron and light scrap. These data permit calculation of the cost of electric furnace operation in any plant, using unit costs for the particular location.

Costs for casting sponge iron heats are the same as those for scrap heats, and the techniques used are the same in every case. The calculation of costs for steelmaking, considering the overall operation at the Tamsa installation, are given on pages 16 to 19.

Quality of Sponge Iron Produced

The quality of the sponge iron which can be produced by the HyL process is entirely contingent upon the grade and characteristics of the ore or concentrate charged into the reactors. The ores used at Monterrey and Tamsa during normal production range from 55 to 65% in iron content and are low in sulfur and phosphorus. The reduction process removes 90% or more of the oxygen. Up to 85% of the sulfur in low-sulfur ores also is removed. The reduced product averages 85% metallization. Gangue constituents not removed by preliminary processing are removed as slag during subsequent melting and refining operations.

Producing Steel with HyL Sponge Iron

HyL sponge iron is highly suited to use wherever cold melting stock is normally employed: the basic oxygen furnace, the open hearth, or the electric arc furnace. Experience to date has primarily been with electric arc furnace melting where HyL sponge is used with scrap for the production of rimming and low carbon steels. Electric furnace operation is flexible, producing either carbon or alloy steels.

Electric Furnace Data

(Basis: 17 ft. Furnace; Low Carbon Steel Product)

	Case 1	Case 2
Furnace Charge		
Scrap—%	40	15
Sponge Iron—%	60	85
Total Charge—%	100	100
Ingot Yield—%	91.4*	93*
Tap-to-Tap Time —Minutes	304	320
Lining Life—Heats	120	115
Roof Life—Heats	43	43
Consumption per Metric Ton of Ingot Produced		
Electric Power—KWH	680	790
Electrodes—KG	8.0	9.7
Lime—KG	82	100
Dolomite—KG	10.5	11.5
Magnesite—KG	5.2	6.6
Refractories—KG	10	10.4
Ferro Alloys—KG	5.8	5.8
Labor and Supervision—Man Hours	1.9	2.0

*Ingot yield experienced by Hojalata y Lámina, S. A. High yield in Case 2 is based on use of high grade home scrap only.

Advantages

A high grade ore or concentrate, when reduced in HyL Process Reactors, will produce a sponge iron which is lower and more stable in cost than scrap iron in many localities, and more dependable in supply. Also composition of the sponge iron will be more uniform and tramp metal content will be known and under control. The quality of the steel produced from sponge iron, whether by the basic oxygen, open-hearth or electric arc furnace will be benefitted by the controlled low residual content compared to steel produced from scrap iron.

Two fundamental advantages gained by using sponge iron are lower steel production costs derived from increased yields and improved surface finish of steel products.

Improved surface finish is especially important in the manufacture of galvanized, tinned and enameled products, both from the standpoint of product yield and quality.

Other factors which favor steel produced from HyL sponge compared to similar steels made from scrap are improved grain size control; improved cold workability pertaining to cold rolling and deep drawing operations; improved response to process annealing, allowing for shorter annealing cycles; and good weldability, as evidenced by the quality of electric-resistance welded pipe product where rejections based on weld defects have been markedly fewer.

Applicability of the HyL Process

Swindell-Dressler Company, the exclusive world-wide sales agent for HyL, makes feasibility-operability studies to determine the applicability of the HyL process to specific regional conditions. These studies use actual operating experience as a basis for relating costs and consider the following aspects:

(1) Anticipated market or required plant size;
(2) Type and quantity of products that will be produced;
(3) Location of the market;
(4) Existing steelmaking facilities, if any, that could be integrated into a new facility;
(5) Prospective plant location;
(6) Source, chemical analysis and cost of natural gas;
(7) Source and reserves of the iron ore and its grade;
(8) Evaluation of available ore;
(9) Preparation of sponge iron for sale as melting stock.

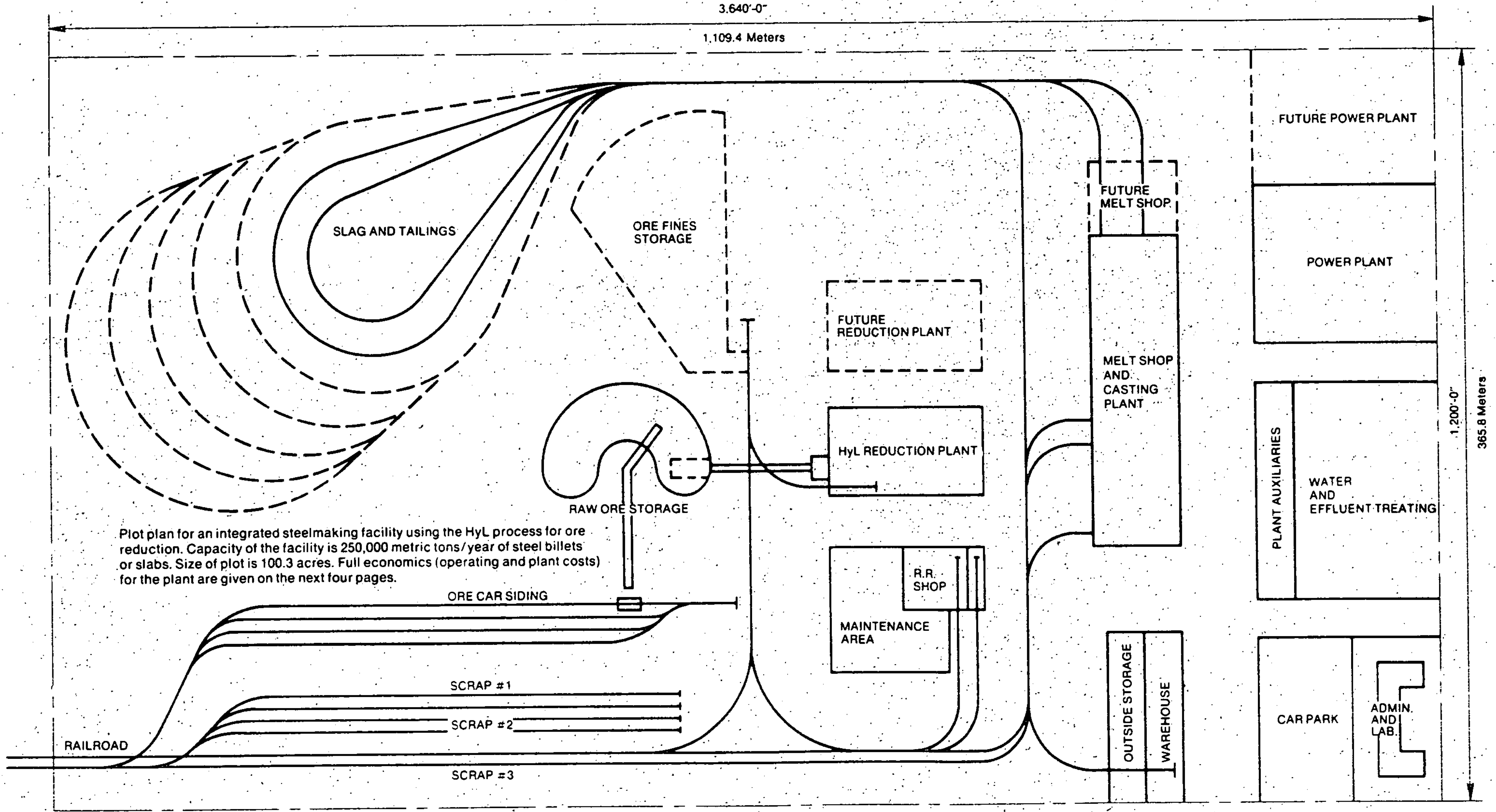
If the preliminary survey indicates that the HyL process would be feasible for a particular steelmaker Swindell-Dressler then requests a sample of available ore for a verification test-run in pilot plant facilities.

Construction of Facilities

Facilities using the HyL process for reduction of iron ore are designed and built by Swindell-Dressler Company. Installations can range from small plants devoted entirely to HyL reduction to complete, integrated steelmaking facilities involving ore beneficiation, HyL installations, melt and casting shops, rolling mills and finishing facilities.

HyL plants can be designed to be operated with simple or complex control techniques, as determined by existing local conditions. The ultimate goal is to provide a plant of economical design that can be operated by the user's own staff.

Hojalata y Lamina, S. A., of Monterrey, offers to assist in the training of personnel for operation of new HyL plants. Instruction manuals and other aids for plant operation are provided by the Swindell-Dressler Company.



Plot plan for an integrated steelmaking facility using the HyL process for ore reduction. Capacity of the facility is 250,000 metric tons/year of steel billets or slabs. Size of plot is 100.3 acres. Full economics (operating and plant costs) for the plant are given on the next four pages.

PLOT PLAN—AREA 100.3 ACRES OR 40.6 HECTARES

Congrès international sur la production et l'utilisation des minerais réduits
Evian (France), 29 - 30 et 31 Mai 1967

ELECTRIC ARC FURNACE STEELMAKING WITH H y L SPONGE IRON

J. CELADA S.
Director of Engineering
Hojalata y Lámina, S. A. Monterrey, México

Hojalata y Lámina has up to the present produced in its electric arc furnaces 3.2 million tons of steel using sponge iron as the major source of metallics. Melt shop practice is described and data are give on yield, power and materials consumed. A discussión is made of the effect on the melt of such sponge iron variables as metallization, carbon content, phosphorus and granulometry. There is growing worldwide appreciation of sponge iron due to its intrinsic advantages as low residuals and ease with which it can be handled.

TABLE I

AVERAGE SPONGE IRON COMPOSITION DURING FIRST QUARTER 1967

(Percent by weight, dry basis)

Fe	47.42	
Iron in $Fe_{0.95}O$	13.88	
Iron in Fe_3C	25.09	
TOTAL IRON		86.39
C	0.09	
Carbon in Fe_3C	1.80	
TOTAL CARBON		1.89
PHOSPHORUS		0.417
SULPHUR		0.023
SiO_2	6.6	
Al_2O_3	0.2	
CaO	0.3	
GANGUE		7.1
OXYGEN in $Fe_{0.95}O$		4.18
		100.00

Percent Metallization $\left(\frac{\text{non oxide iron}}{\text{total iron}}\right)$ 83.93 %

Enero de 1967

TABLE II

SPECIFICATIONS FOR STEELS CURRENTLY PRODUCED FROM SPONGE IRON HEATS

Spec.	C	Mn	P	S	Cu	Ni	Si	Cr	Sn	basic property or application
SAE 1006	.07	.25/.35	.015	.030	.08	.06	-	.06	.01	Extra deep drawing
" "	.08	.25/.40	.020	.030	.10	.08	-	-	-	Deep drawing and coating
" "	.08	.25/.40	.020	.035	.20	.08	-	-	-	Light drawing
" 1008	.10	.25/.50	.035	.045	.40	.12	-	-	-	Light drawing
" 1010	.09/.14	.35/.55	.040	.040	.20	.10	-	-	-	Pipe manufacturing Std.
" 1012	.11/.15	.30/.45	.035	.045	.35	.12	-	-	-	Light structural
" "	.11/.15	.30/.45	.035	.045	.20/.45	.12	-	-	-	Light structural
HyL 1541	.09/.14	.45/.65	.045/.075	.035/.065	.20	.10	-	-	-	Machined fittings
SAE 1020	.18/.23	.30/.60	.020	.030	.20	.10	-	-	-	Auto chasis
HyL 1530	.10/.14	.70/.90	.020	.040	.25	.10	.10	-	-	L.P.G. portable cylinders
SAE 1018	.16/.21	.65/.85	.035	.040	.30	.12	.10	-	-	Storage tanks
" 1027	.20/.25	1.20/1.60	.020	.045	.35	.12	.10	-	-	High tensile
HyL 1536	.15/.20	.85/1.15	.040	.050	.30/.60	.50/.70	.10	-	-	Heavy structural trailer & truck chassis
" 1527	.08	.25/.35	.030	.040	.12	.06	.60/.70	-	-	Motor's laminations
SAE 1045	.43/.50	.60/.90	.040	.050	.40	.15	.10/.20	-	-	Laminated springs & shovels
" 950	.06/.12	.30/.45	.090/.110	.035	.35/.50	.45/.55	.25/.35	.80/1.00	-	High tensile and atmospheric corrosion resistance

Notes: 1.- The specifications listed are produced in the forms of structural plate, hot rolled sheet and strip and cold rolled sheet and strip and include applications of extra deep draw and coating finish.

2.- Low carbon rimmed steel = 68% of steel produced
Medium carbon rimmed steel - 23% of steel produced
Killed and semi-killed steel - 9% of steel produced

TABLE III

SPONGE IRON SCREEN ANALYSIS MADE WITH ENCINO ORE

Mesh number	Retained	Accumulated
1/2	23.39	23.39
1/4	24.59	47.98
6	14.95	62.93
12	9.59	72.52
20	6.14	78.66
40	4.69	83.35
70	5.41	88.76
140	5.36	94.12
-140	5.88	100.00

PRODUCTION OF METALLIZED PELLETS

BY THE HyL PROCESS

BY: H. E. GEARHART, PROJECT MANAGER AND
K. A. JACKSON, MANAGER PROCESS ENGINEERING DEPARTMENT
SWINDELL-DRESSLER COMPANY
DIVISION OF PULLMAN INCORPORATED
PITTSBURGH, PENNSYLVANIA

ASSOCIATION OF IRON AND STEEL ENGINEERS
1970 ANNUAL MEETING

TABLE 1

COMMERCIAL HYL PLANTS

COMPANY	LOCATION	RATED CAPACITY	START OPERATION
HYLSA	MONTERREY, MEXICO	200 MT/DAY	1957
HYLSA	MONTERREY, MEXICO	500 MT/DAY	1960
TAMSA	VERACRUZ, MEXICO	500 MT/DAY	1967
HYLSA DE MEXICO	PUEBLA, MEXICO	500 MT/DAY	1969

RATED CAPACITY IS BASED ON OPERATION WITH EL ENCINO LUMP ORE

TABLE 2

SPONGE IRON PRODUCTION

ANNUAL METRIC TONS - TOTAL IRON

1960	114,000
1961	175,000
1962	166,000
1963	170,000
1964	203,000
1965	213,000
1966	265,000
1967	317,000
1968	363,000
1969	385,000
TOTAL	2,370,000

TABLE 3

	LUMP ORE OPERATION 1969	ALZADA PELLETS 6 MONTHS - 1970
RATED DAILY PRODUCTION METRIC TONS FE	500	680
NATURAL GAS SCFT METRIC TON FE	25,000	20,300
WATER GALLONS - METRIC TON FE	1,136	1,060
ELECTRIC POWER KWH METRIC TON FE	10	8
AVERAGE % METALLIZATION	82.3	87.4
AVERAGE % REDUCTION	88.2	91.5

VISIT 10: JERSEY ENTERPRISES INCORPORATED

Situated at
30 Rockefeller Plaza
NEW YORK. NY 10020

DATE: Monday 16 November 1970

MET: Mr F.P. Barrow Jr., Executive Vice-President and Director

1. HISTORY OF FIOR PROCESS

Mr Barrow said that the 300 tons per day semi-commercial plant in Nova Scotia was operated in a technically successful manner, correlation obtained with the Baton Rouge pilot plant, design parameters estimated, and a considerable tonnage of 4 by 2 by 1 inch pillow briquettes produced. The plant was then shut down in October 1969 as the process proved to be not economic at that small scale. Market development was accelerated and a commercial undertaking on the Gulf Coast of Texas to produce 1.5 million metric tons per annum of slightly thicker briquettes was planned for completion in the third quarter of 1972. The briquettes, 92% metallised, would contain over 90% metallic iron. Response to the product was quite good at a quoted price of $38.50 cif Japan and Europe and it was tested in Japan, USA, Europe in larger quantities than the product of any other developing process. In tests with joint participation of Esso and the steelmaker the product was universally acceptable, but in some tests in which Esso did not participate a learning period was required before the steelmakers were able to use the product's unique properties. Lukens Steel Company near Philadelphia carried out some tests which were reported in the first quarter of 1970 (Smailer and Price, 1970 - summary page A-123).

Esso found on re-estimation that the natural gas reserves on the Gulf Coast were largely committed and that the price would be substantially higher than previously estimated. Ocean freight had escalated and iron ore suppliers increased the prices quoted. The US Government also removed the investment tax credit. As a result the return was no longer sufficiently high to attract Esso into a new field, and design work was stopped early in 1970.

2. PRESENT STATUS

Mr Barrow pointed out that FIOR is a technical asset and that ESSO were considering sites for its exploitation outside the USA, and might consider a joint venture. The process requires hematite fines - not finely ground as for pelletising but minus $^1/_4$ inch. Four percent gangue in the ore would probably be the limit as it would constitute 6% in the product, which can be used for electric furnace steelmaking or cupola feed (though increasing the cupola slag volume). The FIOR Briquettes, of which he showed a sample, are produced hot and do not weather faster than scrap and can be shipped in open cars. While the Nova Scotia plant has been dismantled the pilot plant remains at Baton Rouge.

OPERATING BENEFITS OF CONTINUOUSLY CHARGING PREREDUCED ORE BRIQUETTES TO A 150-TON ELECTRIC FURNACE

By

R. M. Smailer and J. A. Price

For Presentation At

American Iron and Steel Institute Technical Committee On
Electric Furnace Steelmaking, February 6, 1970

And

American Institute of Metallurgical Engineering NOH And
BOSC Conference, April 7, 1970

1/16/70

I. SUMMARY

The basic objective of these trials was to develop a continuous addition practice with prereduced iron ore briquettes which would result in substantially improved productivity and reduced operating costs, while maintaining satisfactory product quality. The benefits obtained in the experimental briquette runs fulfilled this objective in all respects. Overall, 17,000 tons of steel were produced employing continuous addition practices in two complete furnace refractory campaigns. The results of 108 experimental heats compared with 200 conventionally charged all-scrap runs are summarized as follows:

(1) Productivity increased 18%
(2) Tap-to-tap time shortened 39 minutes
(3) Furnace refractory life and costs were unchanged
(4) Electrode consumption was lowered 22%
(5) Oxygen consumption decreased 40%
(6) Energy requirement increased 2.6%
(7) Metallic yield was unaffected
(8) Product quality was unchanged

The continuous addition of FIOR™* prereduced iron ore to a molten scrap bath resulted in a productivity 18% more than that obtained with standard charging practice. This value is based on tap-to-tap time including all shop delays. For this and other reasons discussed later, 18% is considered to be a conservative number. The increase in productivity resulted from a reduction in tap-to-tap time of 39 minutes. Most of this was due to the ability to compress the melting and refining operations into a single procedure during continuous addition. The furnace refractory life on a reline-to-reline basis during the briquette runs was equivalent to standard practice.

In addition, the above listing gives some of the important secondary effects arising from briquette usage. Electrode consumption was 22% lower than for conventionally charged heats. Oxygen usage decreased 40%, primarily due to the residual iron oxide content of the prereduced briquettes. The total energy requirement increased 2.6% for the FIOR briquette heats over conventional runs. Also, metallic yield was unaffected.

Other benefits of continuous operation with prereduced briquettes were realized in steel chemistry. Final residual metal contents such as copper, nickel and chromium were controlled to similar levels for briquette and all-scrap runs by accounting for the dilution effect of the briquettes. This effect was anticipated and permitted a larger percentage of lower grade scrap to be charged in the briquette heats. Slag practice allowed phosphorus and sulfur to be controlled as required. Rolling performance and metallurgical quality of steels made with briquettes were equal to standard production.

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*Trademark applied for by the FIOR Division of Jersey Enterprises, Inc., an affiliate of Standard Oil Company (N.J.).

VISIT 11: HYDROCARBON RESEARCH INCORPORATED

Situated at
115 Broadway,
NEW YORK NY 10006

DATE: Monday 16 November 1970

MET: Dr Edwin T. Layng - Executive Vice President
Mr H.O. Johansson - Manager, Contract Sales
Mr William Volk - in charge of Research

1. HISTORY OF H-IRON PROCESS

Dr Layng said that initial development began as a joint venture with Bethlehem Steel in 1953 and culminated in a 120 tons per day semi-commercial plant by Bethlehem Steel at Vernon, California, near Los Angeles. This plant operated successfully but was shut down due mainly to air pollution problems. Another plant of 50 tons per day capacity was designed for the Alan Wood Steel Co. at Conshohocken to produce iron powder for moulding and welding-rod coating. It was later sold to the Glidden Corporation of Hammond,Indiana and has operated for a total of 13 years but has now been shut down.

2. PRESENT STATUS

A brochure supplied is reproduced on page A-126

Dr Layng said that there appears to be a revival of interest in direct reduction and in the H-iron process. His company was concentrating on foreign sales and was interested in treatment of ilmenite, their original work having been co-sponsored by National Lead.

He said that a single H-Iron reactor could treat 200,000 short tons per annum of feed and that the complete facility including briquetting and passivating of product would cost $US8-$US9 million. It might be possible to introduce a magnetic concentration step after reduction.

Feed specification would be nominally 20 mesh, i.e. 100% minus 12 mesh, not more than 35% minus 325 mesh (a normal 20 mesh concentrate would contain 5-6% minus 325 mesh).

If the product were to be sold as powder it could be screened and the coarse fraction used for welding-rod coating and the fine for powder metallurgy. 95% of the oxygen present is removed.

3. COSTS

The capital cost of a plant to treat 200,000 short tons per annum feed was estimated to be $US8-$US9 million.

Hydrocarbon Research representatives would only outline operating costs as follows:

	$ per ton of product
Fuel	?
Powder	?
Water	?
Steam	?
Chemicals	?
Labour	2.7
Supervision	?
Maintenance	(4% annually of capital cost)
Total	12

They assumed that an investment of $8 million was amortised over 15 years, and used a 12% rate of return. They assumed a cost of 47 cents per thousand cubic ft for hydrogen and said that magnetite required 24,000 cu ft hydrogen per ton of product.

4. FURTHER ACTION

Dr Layng offered to supply a written cost breakdown for a specific case. They would carry out a preliminary evaluation test on a single ore for $12,000, made up of a $4,000 standard fee, six runs at $1,000 each on 300 lb samples and $2,000 for briquetting tests. They have a unit which will take a 5-7 ton batch.

DIRECT REDUCTION

BY

H-IRON®

DIRECT REDUCTION
BY
H-IRON®

TABLE OF CONTENTS

SECTION I. Introduction
SECTION II. Process Background
SECTION III. Process Description
SECTION IV. Utilization of H-Iron Powder
SECTION V. H-IRON Economics

DIRECT REDUCTION
BY
H-IRON®

TABLES AND FIGURES

TABLE 1. Electric Furnace Tests on H-IRON Powder
TABLE 2. H-IRON Investment and Operating Requirements
TABLE 3. H-IRON Processing Costs.

FIGURE 1. H-IRON Flow Diagram
FIGURE 2. Cost of Producing H-IRON Powder

I. INTRODUCTION

The H-Iron process, a joint development of HRI and Bethlehem Steel, is a method for making melting stock by direct reduction of iron ore with hydrogen. Reduction is carried out using finely divided ore in fluidized beds at a temperature below 1,000°F, and a pressure above 200 psig. The low reduction temperature chosen avoids stickiness of the iron powder, and elevated pressures are employed to simplify drying the hydrogen and to increase the rate of reduction.

This process represents an economical method for the manufacture of feed to a steel making furnace. Electrical furnace performance employing H-Iron has been demonstrated to be as good as the performance obtained with scrap. Product uniformity is excellent, and the problems of the tramp elements encountered in scrap are almost entirely eliminated.

HRI offers the H-Iron process for license, and is prepared to undertake the several technical functions required to bring a project to successful commercial realization. The services offered include:

Pilot plant evaluation of ore samples,
Technical and economic studies,
Plant design, engineering and construction.

II. PROCESS BACKGROUND

Pilot plant development of the H-Iron process was conducted at the Trenton research laboratory of Hydrocarbon Research, Inc., in collaboration with Bethlehem Steel. Early work was directed towards a study of reduction kinetics on various ores to provide assurance that satisfactory reduction rates could be achieved for commercial application. A 10 tpd pilot plant was then built at Trenton, in which more than 1,000 tons of product has been made for evaluation in steel making. Two possible applications include charge to the open hearth, as a substitute for the customary purchased scrap and ore charge, and a 95 percent reduced product for electric furnace charge.

Two commercial scale H-Iron plants were built. One plant, of 50 tpd capacity, operated by The Glidden Company, produces reduced iron powder for sale in the powder-metallurgy market. Mill scale is the charge to this operation. A demonstration plant of 110 tpd design capacity was built at the Vernon works of Bethlehem Steel Company in Los Angeles. This plant operated during 1961 and 1962, and its product was used as a melting stock for electric steel making furnaces.

During the final phase of the program, 120 tpd of 95 percent reduced powder was produced. Reduction of 95 percent or better was readily achieved and the day to day uniformity of reduction was excellent.

III. PROCESS DESCRIPTION

Iron ore fines, of the proper particle size, and hydrogen, are supplied to the battery limits of the H-Iron process. The H-Iron process flow is shown in the flow diagram.

Ore is sent by bucket elevator to storage bins, and flows by gravity to a charge hopper. After the hopper is loaded, it is valved-off from the ore supply and purged of air by the use of nitrogen. The hopper is then pressured with a slip stream from the hydrogen reduction circuit. When the pressure in the hopper is higher than the reducer pressure, ore is delivered through a pipe to the top bed of the reducer by merely opening a valve in the line. The powder, practically at its settled density, flows much like water through a pipe. This is known as dense phase transport.

Reduction is achieved in three fluidized beds in series. Fresh ore is partially reduced in the top bed. It is then transferred to the middle bed, where it is further reduced, and lastly, to the bottom bed, where final reduction takes place. The produce of this operation is subsequently transferred by dense phase transport to storage hoppers.

Dry hot hydrogen enters the base of the reducer and passes through the three fluid beds in series. The hydrogen reacts with iron oxide forming metallic iron and steam. This reaction is endothermic, and the heat required is supplied by the hot hydrogen stream.

Effluent gas from the reducer is cooled by exchange with the dry hydrogen being recycled to the reducer. This gas then passes through a water scrubbing tower where most of the water formed in the reducer is condensed from the gas. This water scrubber also serves to remove dust carried over from the reduction vessel.

A portion of the recycle hydrogen stream is vented to prevent accumulation of inert gases. The major portion of the dry hydrogen is recycled through a compressor and is then joined by a make-up hydrogen stream. The combined streams are preheated in the effluent/feed exchanger, further heated in a furnace, and returned to the reducer. By recycling the reducing gas (hydrogen), a considerable savings are achieved, compared to once through processes.

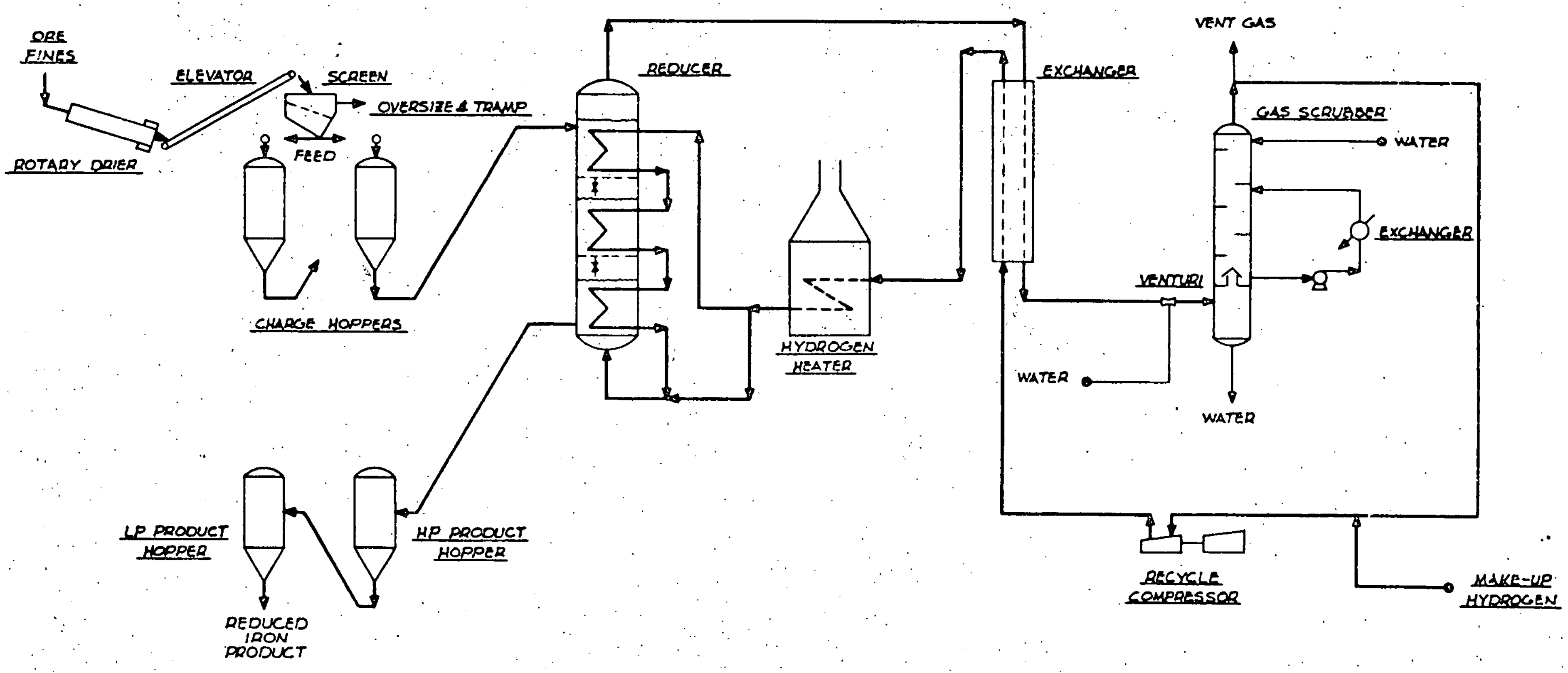


FIG. 1

PRIVATE DESIGN

NOT TO BE COPIED: RETURNABLE ON DEMAND; CIRCUMSTANCES AND FACT OF TRANSMITTAL, CONFIDENTIAL COMMUNICATION: DISCLOSURE OR USE PERMITTED ONLY ON WRITTEN AUTHORIZATION

HYDROCARBON RESEARCH, INC.

HYDROCARBON RESEARCH, INC
NEW YORK, N.Y.

JOB NO

TYPICAL H-IRON PLANT
FOR DIRECT REDUCTION

J. HUANG
10-10-1969
9907D-TL-D-1001

IV. UTILIZATION OF H-IRON PRODUCT

Briquetting

At Vernon, two methods of charging H-Iron to the electric furnace were used. Initially, reduced powder was briquetted and passivated in a furnace at 1,600°F. Briquettes were then charged to the electric furnace. Data obtained with briquettes of 95 percent reduced H-Iron product in an electric furnace indicated that yields, heat time, power consumption and fluxes were essentially the same as for scrap charges. Lower levels of reduction resulted in increased heat time and power consumption, and decreased yield. One of the outstanding advantages of the H-Iron process is its ability to produce 95 percent reduced iron uniformly and consistently, thereby achieving economical electric furnace operation.

Pyrophoric Powder

In another part of the program at Vernon, powder, not briquetted or passivated, was charged directly into an electric arc furnace. Up to 40 tons of 95 percent reduced powder (40 percent of the charge) was melted smoothly with normal yields (93 percent); while heat time, power consumption (504 kwh per ton) and fluxes were essentially the same as obtained with all scrap heats. In those tests no attempt was made to exclude air from the furnace, yet the extent of oxidation of the powder was negligible.

Experimental melting tests on active reduced H-Iron have been successfully conducted in a one-ton arc furnace at the U. S. Bureau of Mines, Albany, Ore., Metallurgical Research Center. Results of these tests demonstrated that active H-Iron powder with high gangue and high percentages of other metal oxides can be melted satisfactorily and produce specification steel.

In this program 27 heats were made, 24 on H-Iron and three with all scrap iron or pig for comparison. The melts were of four general types; charges of H-Iron only, charges of scrap only, charges of H-Iron and 25 percent scrap, and charges of H-Iron plus a synthetic slag mixture to simulate the refining of H-Iron from high slag ores. One typical melt from each group is shown in Table I.

The melt-down time for all scrap heats was longer than for the all H-Iron or the H-Iron and scrap heats; power consumption was essentially the same in both cases. Power consumption varied depending on the degree of preheat of the furnace (the tests were conducted during the day shift only

IV. UTILIZATION OF H-IRON PRODUCT

Pyrophoric Powder (continued..)

and normally two heats were made per day). Likewise, the yield figures varied depending upon the amount of material left in the furnace after pouring.

Hot Charges

If product from the reducer in a commercial plant is charged directly to the electric furnace, the sensible heat of the powder at 800°F represents a saving of almost 100 kwh per ton in the electric steel making step. In the production heats at Vernon charging cold H-Iron powder, the power requirement was 504 kwh per net ton and the yield was 93 percent. These values are in good agreement with the values shown in Table I.

In these experiments, as in the production heats at Vernon, no attempt was made to exclude air from the furnace and no measurable re-oxidation of the H-Iron product was observed. In some of the tests the H-Iron product was charged at the beginning of the heat; while in others, the powder was fed continuously to a molten bath. Both types of operation were entirely successful. The heats in which slag constituents were added demonstrated that active powder with a high gangue content can be successfully melted and refined.

DIRECT REDUCTION BY H-IRON

TABLE 1

ELECTRIC FURNACE TESTS ON H-IRON POWDER[1]

RUN NO.	266	271	250	261
RUN TYPE	H-Iron only	Scrap only	H-Iron & Scrap	H-Iron & Slag[2]
CHARGE, POUNDS				
H-Iron	966		679	586
Scrap		1,012	260	107
Slag Mix				129
TIME				
Melt down	0:37	1:02	0:52	0:45
Refining	0:48	0:43	0:25	0:18
Total	1:53	2:22	1:52	1:25
POWER, KWH				
Melt down	300	330	320	250
Refining	210	200	70	70
Total	530	590	460	380
PRODUCT, POUNDS				
Iron	863	920	857	590
Slag	231	342	198	214
IRON RECOVERY, %	98.5	90.8	98.0	92.3
POWER CONSUMPTION, KWH/NT Product[3]	547	486	528	547

[1] Tests conducted at the Albany Metallurgical Research Center, U.S. Bureau of Mines on melting H-Iron powder in an Electric Arc Furnace.

[2] Slag constituents corresponding to a titaniferous ore were added to the H-Iron.

[3] Values were corrected for heat leak.

V. H-IRON ECONOMICS

Investment and operating requirement data are presented in Table 2 for a plant to produce 300,000 MT/Year of reduced powder. In Table 3, these data are used to develop the processing cost - \$8.87/MT. This cost is exclusive of:

Hydrogen cost,
Ore cost,
Crushing and grinding - if required,
Briquetting and passivation - if required,
Materials handling and offsites.

Figure 2 develops the total cost of producing active reduced powder from fines, as a function of hydrogen and ore costs. Should ore grinding or product passivation be required, these would add the following approximate costs to that derived from Figure 2:

Ore crushing and grinding, 64% Fe	\$ 1.75/Ton of product
Briquetting and passivation	\$ 1.82/Ton of product.

Materials handling and offsite costs will vary considerably with the specific factors of a given location.

DIRECT REDUCTION BY H-IRON

TABLE 2

300,000 METRIC TON PER YEAR H-IRON PLANT INVESTMENT AND OPERATING REQUIREMENTS*

RAW MATERIALS	
Iron Ore fines, 64% Fe	470,000 MT/Year
Hydrogen	24,000,000 SCFD
PLANT INVESTMENT	$8,600,000
OPERATING REQUIREMENTS	
Fuel	167 MMBTU/Hr.
Power	1,250 KWH
Make-up water	440 GPM
Chemicals	$ 27/Day
Labor	8 Men/Shift

---

*These are battery limits requirements for the plant shown in Figure 1.

DIRECT REDUCTION BY H-IRON

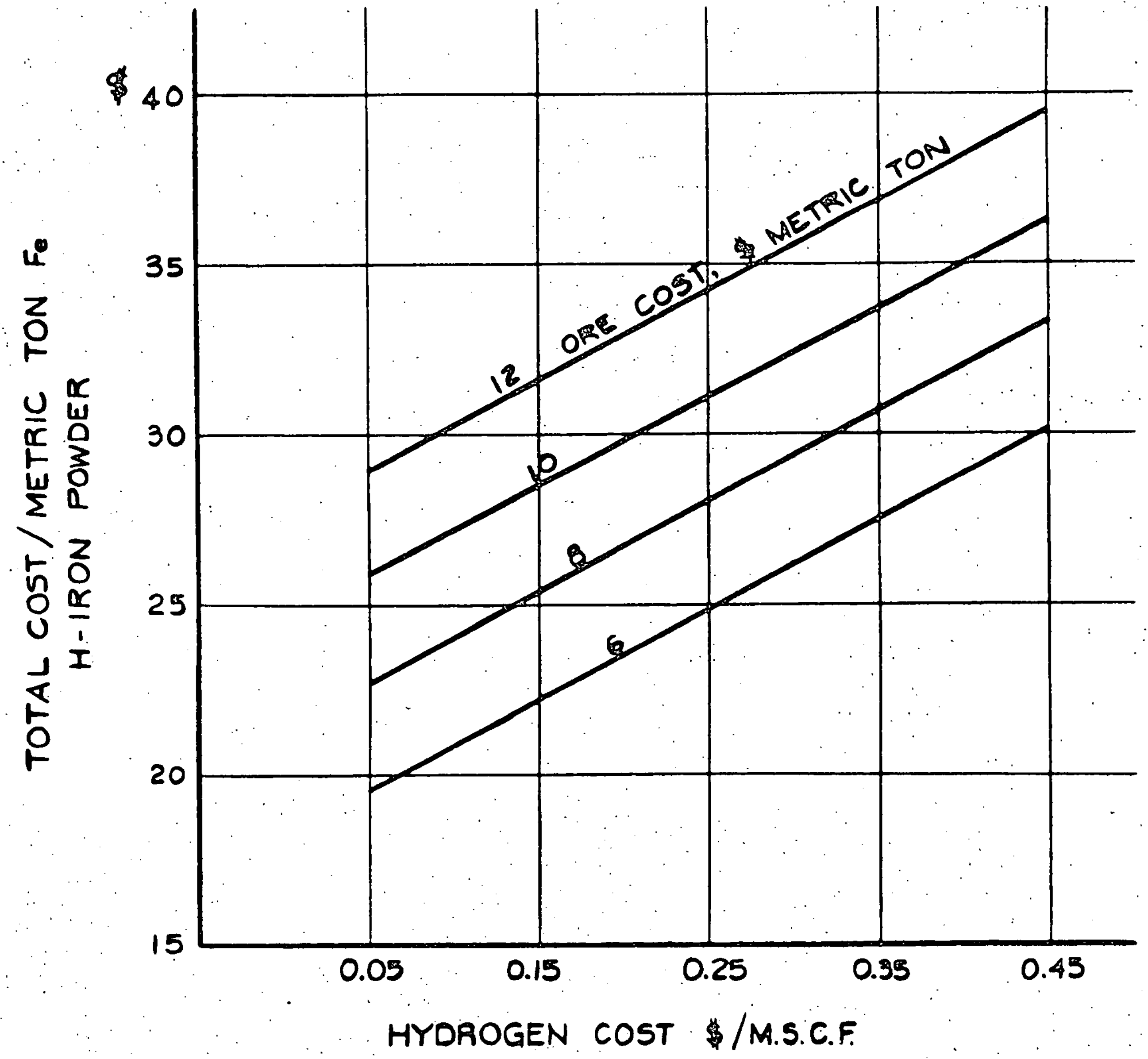
TABLE 3

300,000 METRIC TON PER YEAR H-IRON PLANT UNIT PROCESSING COST

		Cost/Metric Ton Fe Product
Fuel , 4.4 MM BTU/Ton	,@ $0.30/MMBTU	$ 1.32
Power , 33 KWH/Ton	,@ $0.007/KWH	0.23
Water , 0.7M Gal/ton	,@ $0.025/M Gal	0.02
Chemicals		0.03
Labor , 0.28 Man-Hrs/ton	,@ $3.50/Man-Hr.	0.98
Supervision	,@ 25% labor cost	0.25
Maintainance & Insurance	,@ 5% Investment	1.43
Plant Services & General Overhead	,@ 25% labor & Supervision	0.31
Total Unit Costs		$ 4.57
Capital Charges	,@ 15% of Investment	$ 4.30
Total Processing Cost		$ 8.87

FIGURE 2
TOTAL COST INCLUDING ORE FOR 300,000 METRIC TON/YEAR OF H IRON POWDER

BASIS: 64% Fe IN ORE



VISIT 12: TREADWELL CORPORATION, NEW YORK

Situated at
1700 Broadway,
NEW YORK NY 10019

DATE: Tuesday 17 November 1970

MET: Mr E.S. Roberts, Vice President

1. FURTHER TECHNICAL DETAILS RELATING TO ANACONDA-TREADWELL PROCESS

Mr Roberts, who used to be with Chemetals Corporation, said that the pebble heater and desulphuriser were not yet proven at large size. He said that he had inspected at Union Carbide an Orbital Furnace supplied by an English manufacturer through Dravo Corporation, and thought this might be relevant. He also referred to cracking of the fibreglass-resin sulphur scrubber, and the precipitation of hydrated silica by cyanide. I said that I did not want to see a laboratory demonstration of the copper cyanide precipitation but would like as complete information as possible on estimated process costs. He had a member of his staff give me details of the laboratory cyanide precipitation technique.

He also described the precautions taken in the design of the hydrogen reduction reactor to prevent ingress of air. There is a charge hopper above, and one below, the reduction shaft, and above and below each hopper are, from top to bottom, a slide valve and two ball valves. A multi-loop control system ensures that these are operated in proper sequence.

Ferrous sulphate is crystallised as heptahydrate by refrigeration to 0°C and the crystals melted and spray-dried at 500°F to give the sesquihydrate.

2. CAPITAL COST DATA

For a plant to produce 80,000 short tons of copper per annum, using copper powder as reductant as would be done if treating an oxide ore, Mr Roberts supplied the following data:

2.1 <u>Major Purchased Equipment</u>	<u>$US million</u>
Sulphate Leach Operation	2.54
Solubles separation from pebbles	0.43
Gas scrubbing	0.71
Sulphur Dioxide handling	0.28
Solution Purification	0.31

Major Purchased Equipment (Cont.)	$US million
Cyanide Precipitation	0.81
Filtration of Cuprous Cyanide	0.47
Briquetting and Reduction	1.04
Melting and Casting	1.02
Acid and Iron Recovery	3.00
Tailings Treatment	0.13
Utilities	0.78
Total Purchased Equipment	11.52

2.2 Other Items, including:

Concentrate supply
Piping
Electrical
Installation
Process Structure
Instrumentation
Painting
Industrial - Engineering
Drafting
Purchasing
Travelling
Field - Tools rental etc.
Taxes and Insurance
Start-up

Total	30.2
Total Fixed Capital Investment	$US41.7 million

(This estimate was made in October 1969 and should be adjusted if necessary. It does not include any working capital).

3. OPERATING COST DATA

For the same assumed plant Mr Roberts supplied the attached material flowsheet (Drg 2505) and Table A12.1 pages 142 & 143. He said that HCN could be produced by the Andrussow process at a molar cost three times that of ammonia. He supplied the following breakdown of labour:

	Strong Acid	Copper Recovery	Melting and Casting	Acid and Iron Recovery	Tailings	Utilities and Services
Foremen	0	1	2	0	0	1
Operators	2	2	14	2	1	1
Helpers	2	2	8	2	1	1
Shifts/day	3	3	1	3	3	3
Shifts/week	21	21	5	21	21	21
Shifts/year	1092	1092	260	1092	1092	1092

and estimated the following:

Operating Supplies and Maintenance: 2½% per year of cap. cost
Maintenance Labour: 2½% per year of cap. cost
Technical Supervision: 20% of Labour cost
Plant Overhead: 60% of Labour cost

(342 operating days per year assumed, 365 days for supplies and labour)

TABLE A12.1: UTILITIES AND PRODUCTS
Copper 240 ST/D
(Table supplied by Treadwell Corp.)

	Power Req'd KWH/D	Natural Gas MM-BTU/D	600 lb Steam M-LB/D	250 lb Steam M-LB/D	L.P. Steam (20psi) M-LB/D	Treated Water [a] M-LB/D	Cooling Water 20°F Rise M-GAL/D	Raw Water M-GAL/D	SO_2 Gas ST/D	H_2 Gas ST/D	N_2[b] Gas SCF/D	HCN Make up ST/D	Burned Lime LB/D	H_2SO_4 100% Basis ST/D	$FeSO_4$ 7 H_2O ST/D	Fe_3O_4 ST/D	Sulphur ST/D
Steam	(99500)[c]		1720	511	(2230)												
Water Supply																	
Cooling Tower	23600						(36000)	1080									
Water Treatment	697					(9020)		1200									
Steam Supply	3810	3130	(2340)			234											
H_2 and N_2 Supply																	
Gas Reformer	572	715	115				1290			(3.9)							
N_2 Plant	72	9	1				22				(24800)						
Yard and Offsites	2000					33		9									
Strong Acid																	
Bake Reaction	26700	4000					14400							1850			
Solubles Separation	3850																
Gas Scrubbing	12200				(480)	480	5040										(224)
SO_2 Handling	4590					6250	181		(646)								
Acid Cyanide																	
Purification	4870						8600		82								
Precipitation	24000				480				38			2.0					
Filtration	23000					1200											
Briquetting & Reduction	8830	240	509			4800	323			3.9	24000						
Melt and Cast	10200	800				480	1600										
Acid & Iron																	
Multiple Effect Concentration	15200			24	2230	(4460)											
$FeSO_4$ Crystallization & Centrifuging	17200						1610								(1060)		
$FeSO_4$ Drying & Roasting	6720	2690						216	(242)						1060	(293)	
Submerged Combustion Concentration		6000												(672)			
Contact Acid Reconversion	27100	148		(535)					768					(1180)			

Continued

TABLE A12.1: CONTINUED

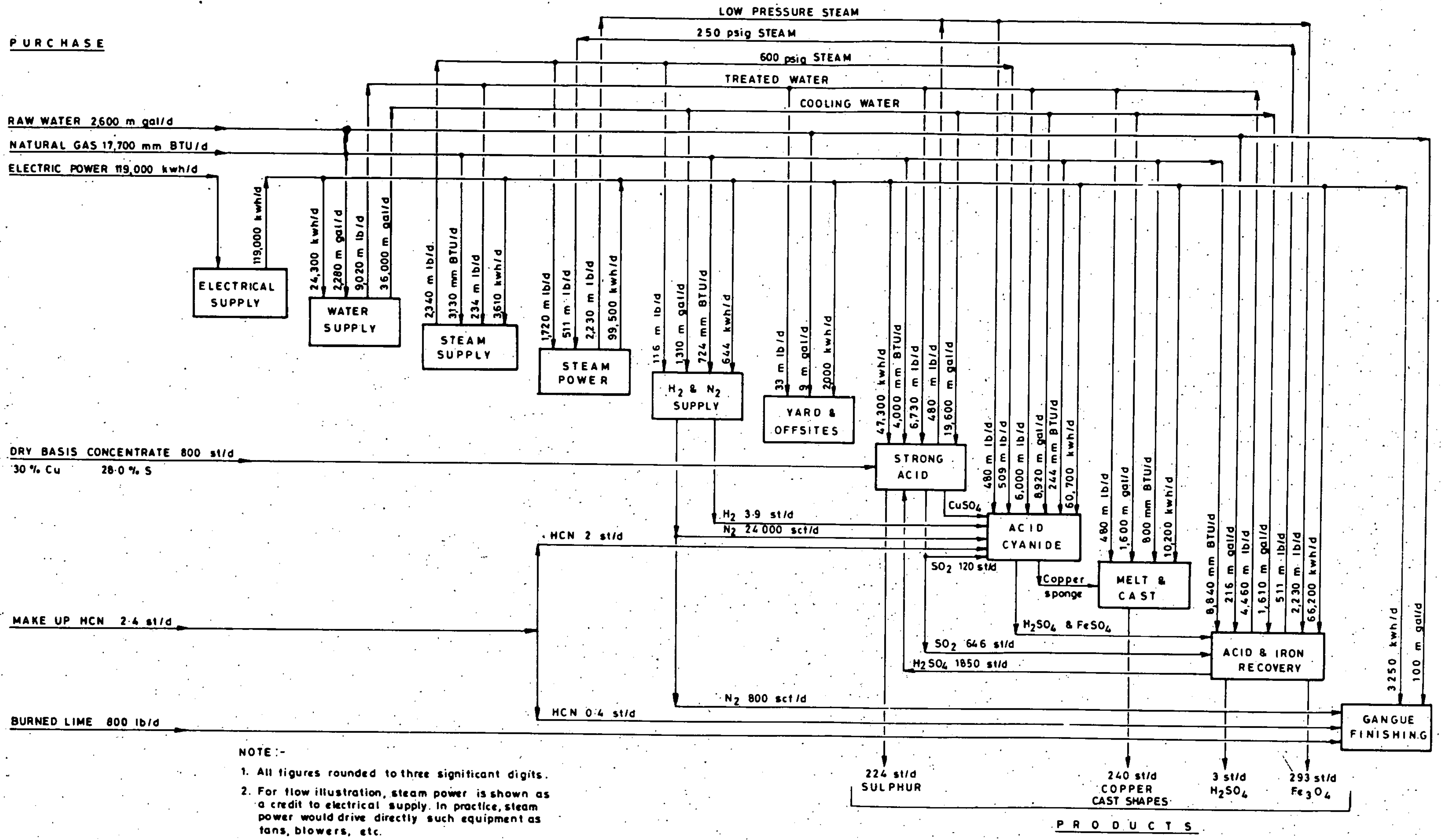
	Power Req'd KWH/D	Natural Gas MM-BTU/D	600 lb Steam M-LB/D	250 lb Steam M-LB/D	L.P. Steam (20psi) M-LB/D	Treated Water a M-LB/D	Cooling Water 20°F Rise M-GAL/D	Raw Water M-GAL/D	SO_2 Gas ST/D	H_2 Gas ST/D	N_2b Gas SCF/D	HCN Make up ST/D	Burned Lime LB/D	H_2SO_4 100% Basis ST/D	$FeSO_4$ $7H_2O$ ST/D	Fe_3O_4 ST/D	Sulphur ST/D
Gangue Finishing																	
Tailings Treatment	50																
Sulphides Reflotation	3000							50									
Precious Metals Cyanidation	200							50			800	0.4	800				
Credits	99500		2340	535	2710	13500	36000		888	3.9				1850	1060	293	224
Debits	219000	17700	2340	535	2710	13500	36000	2600	888	3.9	24800	2.4	800	1850	1060		
Surplus														3		293	224
Purchase	119000	17700						2600			24800	2.4	800				

Notes: a. By ion exchange - for boiler and washing cuprous cyanide.

b. For purging.

c. Bracketed figures are credits.

NB: L.P. steam is credited as kw of electrical power.



NOTE :-

1. All figures rounded to three significant digits.
2. For flow illustration, steam power is shown as a credit to electrical supply. In practice, steam power would drive directly such equipment as fans, blowers, etc.
3 Turbine efficiency has been taken as 60 % for power calculation.

TREADWELL CORPORATION
NEW YORK, N.Y. EASTON, P.A.
DRAWING Nº 2505

VISIT 13: BRITISH IRON & STEEL RESEARCH ASSOCIATION
Process Physics Section

Situated at
Battersea Park Road,
LONDON

DATE: Thursday 19 November 1970

MET: Dr A.A. Rush

Dr Rush explained that he had been endeavouring to produce powder, suitable for compaction into thin mild steel strip, by gas atomisation of molten steel. However the powder he produced was spherical and not fine enough for producing the thin strip desired at British Iron & Steel Research Association's Sketty Hill laboratories near Swansea, which required a minus 53 micron powder. The methyl cellulose binder ran through the powder he produced. He did not consider it economic to increase the velocity of the nitrogen jet above twice the speed of sound to produce finer particles as the energy requirement became excessive. He said that a suitable powder, finer and irregular, is now being produced using a water jet of sonic velocity, at Port Talbot, Wales, the model work having been done in London.

At British Iron & Steel Research Association's Sheffield laboratories alloy powders are being produced by water atomisation, and compacted into strip, and a visit to these laboratories was arranged. Dr Rush considered that production of alloy powders was more profitable and said that the economics of mild steel powder production were well known.

He said that research and development on the use of iron powders to make preformed shapes for forging is being carried out by Stanton and Stavely, now part of the Tubes Division of the British Steel Corporation. Their address is PO Box 72, near Nottingham NH 10 5AA, Phone Ilkeston (06072)-2121. Dr D. Cratchley is in charge of this work, deputy Mr T.B. Reeve.

Relevant references cited by Dr Rush were:

1. Michalke M & Scholz W. Chemie Ingenieur Technik <u>41</u>, 1-2 pp 16-22 (Jan. 17 1969) "Production of alloyed and non-alloyed iron powder by atomisation".
2. Tamura K. & Wanikawa S. Japan Powder Met. <u>15</u> 302-7 (Oct. 1968) "On the atomising variables in the production of metal powder by liquid atomisation"

3. Anon Metal Bull. Friday March 21, 1969. "Iron Powder in Perspective"

VISIT 14: WOODALL-DUCKHAM LIMITED

Situated at
Crawley, Sussex,
ENGLAND

DATE: Monday 23 November 1970

MET: Dr Peter Finlayson (Manager Metallurgical Industries Development)
Mr Frank Singleton (Chief Chemist)
Mr Jim Marshall (Senior Extractive Metallurgist)

1. INTRODUCTION

Dr Finlayson outlined how Nordac, now taken over by Woodall Duckham with expertise in HCl technology had been the successful tenderer to develop the process devised by the Research Council of Alberta for Mr Jerry Heffernan, Managing Director of Peace River Mining and Smelting and several other concerns. Dr Finlayson carried out a survey for a client, (BTP) 12 months ago to assess the market for iron powder, possibly produced as a byproduct of conversion of ilmenite to synthetic rutile. It was believed that a reduction of say two pence per pound in the price of iron powder might greatly increase its usage. The conclusions of that survey were that until the use of iron powder in preformed shapes for forgings is developed, there is no incentive to build a new plant.

Japanese demand for iron powder is currently 20,000 tons per annum but is rapidly increasing. Kawasaki produce 13,000 tons per annum and Kobe expect to open a plant to produce 15,000 tons per annum in 1971.

2. IRON POWDER PRODUCTION

2.1 Hydrometallurgical

Woodall-Duckham have exclusive world rights to engineer plants using the Peace River process. They are not active in the field at the moment as Peace River Mining and Smelting Ltd chose to engineer their Amherstburg plant themselves. If Nairne pyrite were used as feed to a Peace River process, despite its natural pyrrhotite content, it would almost certainly have to be heated to convert pyrite to artificial pyrrhotite to confer HCl-solubility.

2.2 Atomisation

Woodall-Duckham have technical know-how in this field and could build a plant, Dr Finlayson said.

3. ZINC EXTRACTION FROM SILICATE ORE

Some 10 years ago Woodall-Duckham developed in the laboratory an HCl-extraction process for extraction of zinc from a Brazilian silicate ore, followed by electrowinning. The process appeared attractive and Woodall-Duckham would welcome a sample of S.A. zinc silicate ore from the lease-holding company for a free preliminary appraisal.

4. OTHER TOPICS

The following were also discussed briefly:

1. Attack of HCl on apatite and clays.
2. Dilute, cold acid-leach of malachite-azurite ore to avoid attack upon dolomite.
3. HCl leach of uranium ores.

The Aman NCl-recovery process was discussed in detail with slides, literature (pages A-149 to-A-154) and comparison with the Lurgi process.

reprinted from The Engineer · December 1969

Acid-free effluent clears the way to bigger profits

Woodall-Duckham
WORLDWIDE PROCESS ENGINEERING AND CONSTRUCTION

Acid-free effluent clears the way to bigger profits

Steel picklers are being forced to control their effluent. Graham L. Hodgetts shows how a reappraisal of techniques can boost profitability

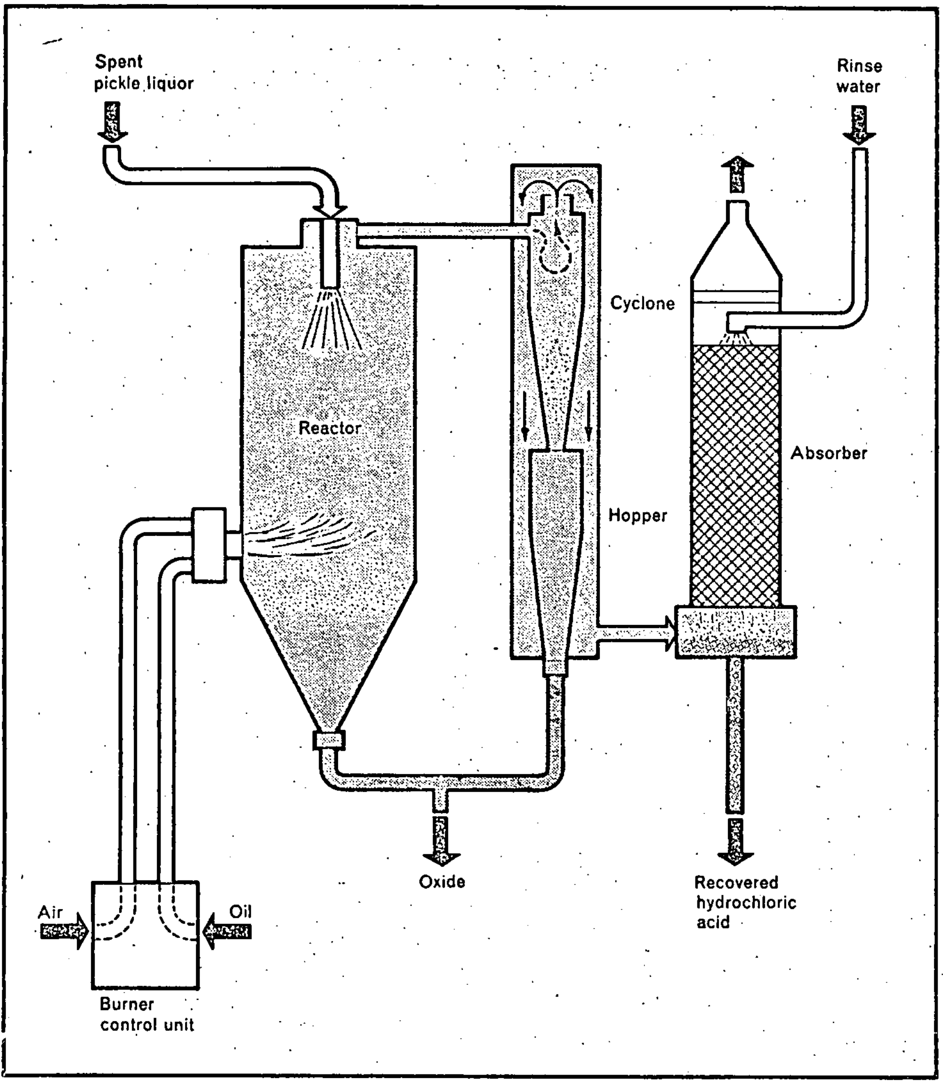
Britain's steel picklers today are under significant pressure from local authorities and river boards to control the effluent from their installations. In some cases the big stick has compelled operators to examine their techniques analytically.

Often such self-examination not only reduces the offending contaminant to an acceptable level but also improves other aspects of operation.

Particular instances are more economical usage of acid so less is consumed for every ton of steel pickled; indirect heating of the bath eliminating the condensation water associated with direct steam injection and reducing the liquor volume; and more effective use of rinse water.

In an attempt to capitalize on the present dilemma of the acid pickler, dry de-scaling techniques are now being offered. But these have made no significant inroads into the industry, either in Britain or abroad, despite the hovering effluent inspectors.

Fig. 1—Plant for recovery of HCl from spent liquors



Fresh approach. In recognizing that acid de-scaling is a thoroughly well proved and accepted technique, a fresh approach to effluent optimization has been made by the Woodall-Duckham Group.

Two of its member companies, Woodall-Duckham Ltd and Ames Crosta Mills, have combined their knowhow to provide industry, for the first time, with a complete package for tackling and solving pickle effluent problems — fig. 2.

Following the general swing to hydrochloric acid — HCl — for pickling flat products and wire, W-D/ACM has concentrated on the treatment of effluents from this medium.

Recovery. The achievements by W-D in the recovery of HCl from spent liquors are well known to the steel industry. This process not only recovers the free HCl in the liquor but it completely regenerates by hydrolysis in a spray roaster the iron chlorides back to HCl. This eliminates the spent liquor effluent — fig. 1. It overcomes the main problem of the medium-to-large pickler which, in most cases, currently has to rely on the acid supplier to haul away the resulting spent pickle liquor.

Thickener. Ames Crosta Mills, which is prominent in the field of effluent treatment, has taken a long hard look at the problems associated with the removal of gelatinous hydroxides from neutralized rinse waters. By nature ferrous hydroxide flocs are very slow to settle and extremely difficult to remove by filtering.

There are some processes available that oxidize the ferrous hydroxide to the ferric form. This does go some way to improving settlement and filtration, but these processes are expensive and susceptible to changes in the iron content of the rinse water.

Having rejected settling and filtration ACM now employs a novel technique. Neutralization is carried out normally but the resulting liquor is fed to the Komline flotation thickener — fig. 3.

Here, rather than attempt to settle the flocs, their buoyancy is increased by the action of minute air bubbles. These become attached to the flocs, causing them to rise to the surface. A portion of the clear water is recycled at a controlled rate and mixed by the action of a water ejector with air in the air-mix pressure tank.

The buoyant condition of the sludge blanket — normally 8 in to

24 in thick — causes the upper surface to rise above liquid level. Drainage of the flat then takes place before it is swept into an adjacent hopper by a variable-speed skimming mechanism. Scroll conveyors in the base of the unit remove any dense particles which reach the bottom of the tank.

With this unit the clarified water usually contains less than 30 ppm solids, and the resulting hydroxide sludge has a sufficiently high solids content — typically 7% by weight or more — to reduce the problems of external transport.

Sources. Before applying either or both these processes it is essential that both the pickler and the rinse sections are examined and modified so that effluent produced is the minimum without detriment to the pickled product.

In certain cases of continuous strip pickle lines operating at low line speeds — up to 100 ft/min — the rinse section can be arranged to achieve efficient rinsing with a sufficiently low volume of primary rinse water to correspond with the water feed required at the HCl absorber on the recovery plant.

In this way the effluent problem is reduced so that only nominal dosing of the secondary rinse is necessary, if at all, to maintain the required pH. In these systems the iron content of the secondary rinse water is usually low enough to be acceptable.

A bonus from using such techniques is that the rinsed strip is very much cleaner and does not rust or stain so readily as strip with a relatively high proportion of surface residual chlorides.

The counter-flow rinse principle of such a rinse section is shown in fig. 4. Two counter-current tanks are shown here, but more tanks, usually, are used.

Of particular importance are the high-pressure fine water jets directed on the strip before it leaves the pickle bath. Air can be used for this, but operators with under-hood fume extraction systems consider water jets play a double role in reducing dragout *and* maintaining the water balance on the pickle line.

Wire problem. It is not that easy, however, for high-speed strip lines where dragout is notoriously difficult to restrain within tolerable limits. Neither is it quite so straightforward for wire picklers, though a lot can be done to improve on the infamous 'hose in the corner' technique.

An example of a batch rinse system is shown in fig. 5. Naturally, some existing installations will not lend themselves readily to the approach suggested but in many cases reduction in rinse water volume can be achieved with nominal modifications.

It can be said that processes are available which, when applied efficiently, allow the acid de-scaling operator to observe the ecological standards of society despite the medium he employs and upon which he relies for his livelihood.

Graham L. Hodgetts is executive, industrial sales, Woodall-Duckham Ltd.

Fig. 2—Package effluent disposal plant

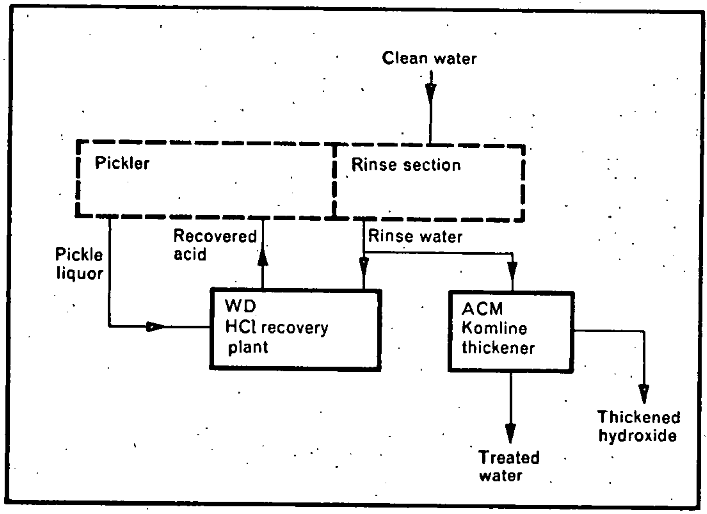


Fig. 3—ACM/Komline flotation thickener

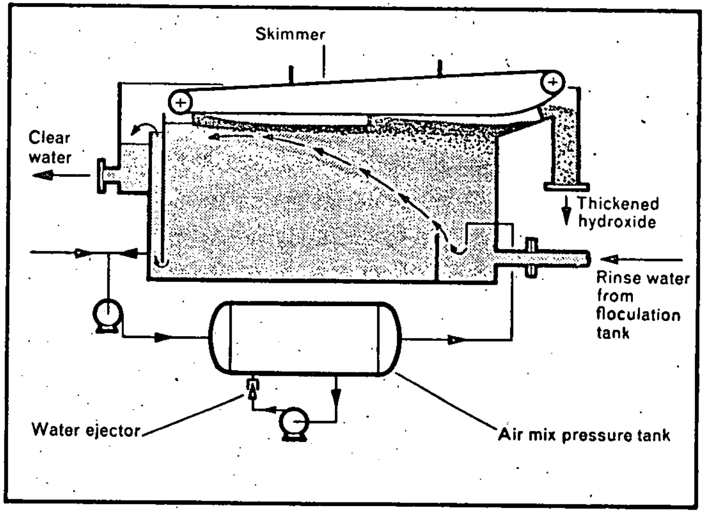


Fig. 4—Counter-current rinse line

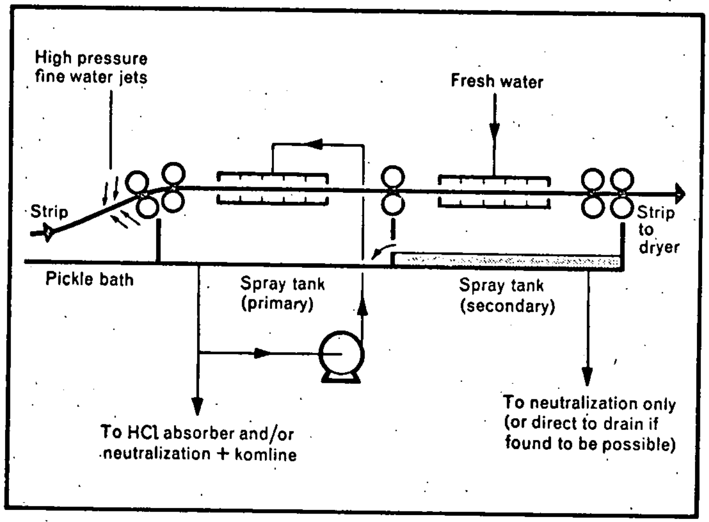
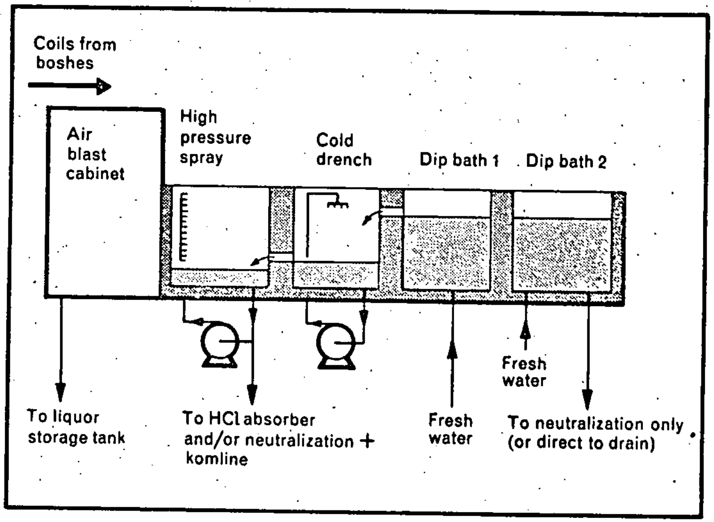
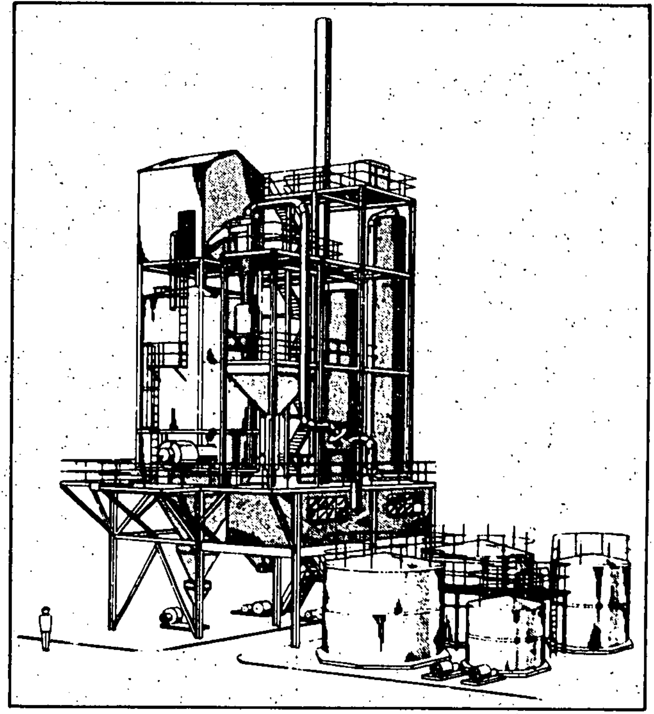


Fig. 5—Batch rinse system for high-speed wire lines



hydrolysis of metal chlorides



Woodall-Duckham

WORLDWIDE PROCESS ENGINEERING AND CONSTRUCTION

hydrolysis of metal chlorides

the Woodall-Duckham regeneration process

During the last 15 to 20 years the number of processes using chlorine or chloride technology has greatly increased in the inorganic chemical field. These fall into four groups, based on the use of:

1. Solid chlorides; 2. Gaseous chlorine; 3. Gaseous hydrochloric acid; 4. Aqueous hydrochloric acid.

The last group has expanded most of all, and for this the process described below is largely responsible.

Many chemically sound processes have in the past not been feasible due either to the poor economics or the problem of disposal of chloride effluents. The recycling of hydrochloric acid at low cost made possible by the Woodall-Duckham regeneration process makes such processes viable. Examples are:— 1. Descaling of steel; 2. Production of high purity oxides such as those of magnesium, cobalt, nickel, etc.; 3. The upgrading and beneficiation of ores, etc., such as ilmenite, bentonite, manganite, sands, etc.; 4. Production of high purity metal powders, e.g. iron, cobalt, etc.

The Woodall-Duckham regeneration process operates either as a process in its own right, or as the essential closing link in a larger process where it permits the recycling of hydrochloric acid back to the acid consuming stage.

Existing processes

It has been known for some time that a number of metal chlorides will hydrolyse in the presence of steam at high temperatures to give oxides and HCl gas.

However, the severe corrosive conditions involved in such reactions prevented their commercial exploitation until more resistant materials of construction became available. This has only happened in the last decade, and previously in most cases waste chloride liquors were treated as effluents and discharged to drain. Even if steps were taken to neutralise any free hydrochloric acid in such an effluent the resulting liquor was still obnoxious due to the high percentage of dissolved solids, present as chlorides.

In an early hydrolysis process developed in Germany, the liquor was first concentrated and the ferrous chloride precipitated as $FeCl_2 . 4H_2O$. The crystals were then roasted in a multiple hearth furnace followed by fluid bed calcination. This process was difficult to operate and was soon abandoned.

The Regeneration Process

The heart of the process is the spray roaster in which the metal chloride is hydrolysed to hydrochloric acid and the corresponding oxide according to the following general equations:

$$(1)\ MCl_2 + H_2O \rightarrow 2HCl + MO$$

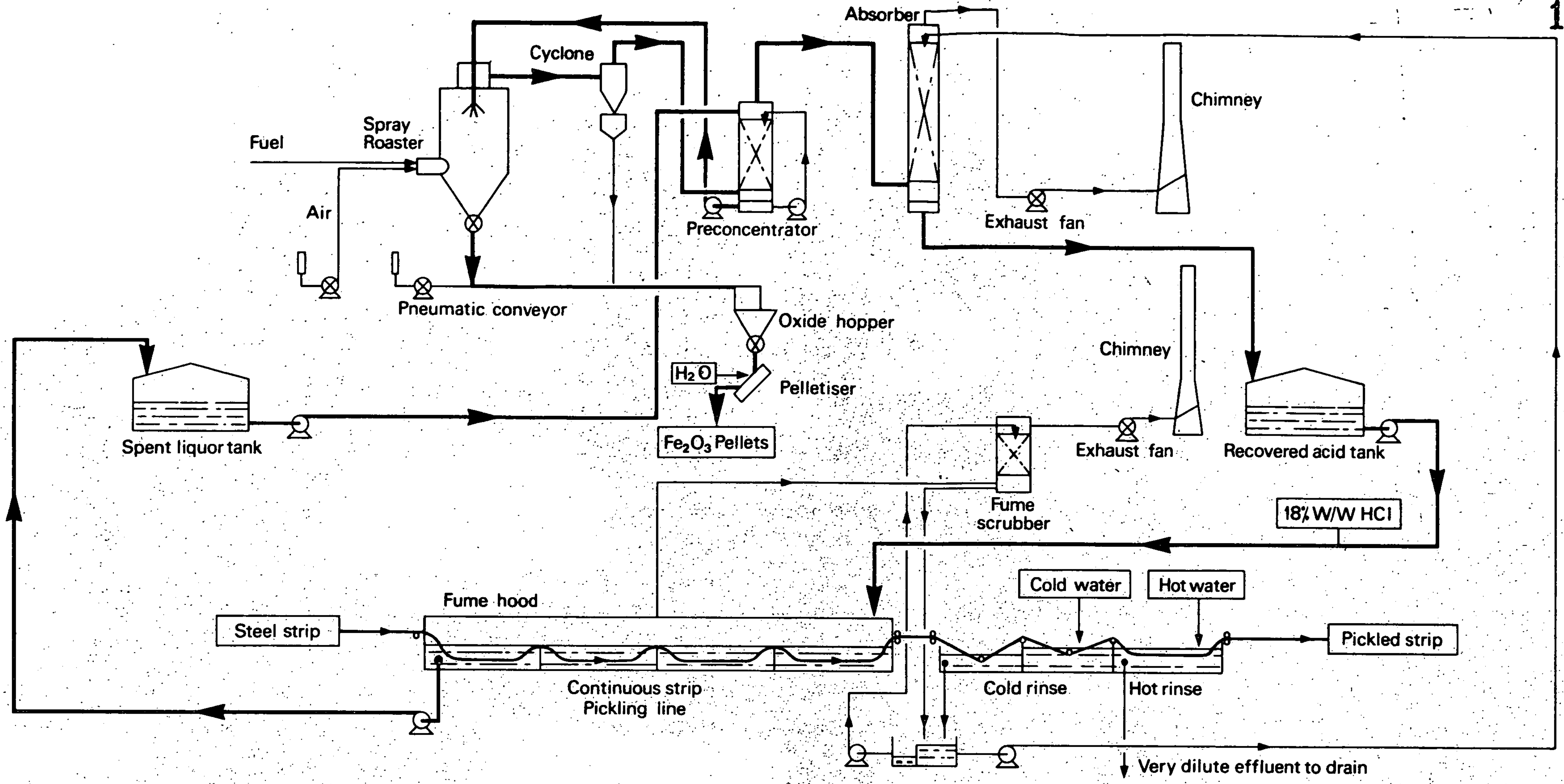
$$(2)\ 2MCl_2 + 2H_2O + \tfrac{1}{2}O_2 \rightarrow M_2O_3 + 4HCl$$

$$(3)\ 2MCl_3 + 3H_2O \rightarrow 6HCl + M_2O_3$$

where M represents the metal atom.

The main steps of the process are evaporation and hydrolysis in the spray roaster, removal of fine oxide carryover in a cyclone system and absorption of the hydrogen chloride in water, either in an adiabatic absorber or in a cooled absorber system. In many cases a pre-concentrator is used, while in some cases a submerged combustion or steam evaporator is also included in the plant.

The heat required for the reaction and for evaporating water, etc., is supplied by the combustion products of a gaseous or oil fuel which are introduced into the roaster near the bottom. The chloride solution is sprayed into the top, counter-current to the gas flow, using spray nozzles and pumps specially developed for the process. The drops as they pass downwards are first dried, then hydrolysed to the oxide and hydrogen chloride. The oxide is removed continuously from the bottom of the roaster. The mixture of combustion products, water vapour and hydrogen chloride gas leave the top together with an amount of finer particles of oxide. It then passes through a cyclone where most of the oxide is removed and then proceeds to the absorption system where the hydrogen chloride is absorbed in water to produce aqueous acid of the desired concentration.



A considerable degree of heat economy is achieved in many cases by installing a pre-concentrator upstream of the absorber. Like the absorber this is a simple mild steel column lined with rubber and acid resisting bricks and packed with Intalox saddles. The incoming feed liquor is circulated around it continuously. Hot exit gases from the roaster enter the base and are cooled by contact with the liquor. The heat transferred to the liquor results in the evaporation of 25-30% of the total water so that less heat is then required in the roaster.

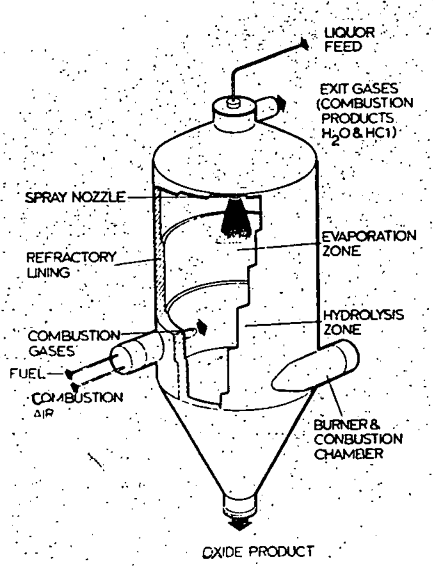
Mild steel is unattacked by dry hydrogen chloride gas below about 500°C. The roaster shell, cyclone, and ducting to the pre-concentrator are in contact with combustion gases and hydrogen chloride gas at temperatures less than this and can therefore be safely fabricated in mild steel. The roaster is additionally lined with refractory bricks or concrete to maintain the shell at 200-300°C.

The use of titanium for key items such as spray nozzles, pumps, fan impellers etc. has virtually eliminated the corrosion problems of handling chloride liquors and the associated wet gas streams containing hydrogen chloride, which have prevented the earlier development of a satisfactory process. In order to ensure that there is no leakage of HCl from the plant during the changing of a spray boom etc., the system is operated under a slight vacuum and this is achieved by a titanium suction fan installed after the absorber. The system can be operated with any normal oil or gaseous fuel although in extreme cases it may be necessary to use a distillate fuel or alternatively a high purity gaseous one when an extremely pure oxide is required.

Not all chlorides can be hydrolysed in the spray roaster. The alkali chlorides require theoretically higher temperatures than are practicable, but would in any case volatilise before this and leave the roaster in the gas stream. Calcium chloride will undergo a partial degree of hydrolysis only.

For one particular mixture of chlorides containing calcium chloride a fluid bed was installed at the bottom of the roaster in order that the reaction time could be extended and complete recovery of the chloride could be achieved. This was however, a most exceptional circumstance and in the event was shown not to be necessary.

With chlorides which hydrolyse easily, e.g. $FeCl_2$, the efficiency of regeneration of HCl is very high, being 98-99% at least. The losses are represented by the traces of HCl in the exit gases from the absorber and the residual chloride in the oxide. With reasonable maintenance and care in operation this can be reduced to 0·1% Cl^-.



The oxide is generally discharged from the roaster via a rotary valve. It is transferred from here to a hopper either by screw conveyor or pneumatic conveyor. If required in the powder form it is then bagged. In some cases it is more convenient for storage or transport to pelletise the oxide in a tilting pan pelletiser. With some applications, e.g. upgrading of silica sand, the product is a mixture of several different oxides and is of no value. It is then disposed of as a slurry in say a disused section of a quarry, by discharging from the roaster straight into an irrigated trough.

Operating parameters

The reaction in the roaster can be divided into two broad overlapping bands. In the upper one water vapour is evaporated from the drop along with any free hydrochloric acid. In the lower one the now anhydrous chloride and in some cases possibly oxychloride, still in contact with water vapour, is hydrolysed to the oxide and hydrogen chloride gas. For any given chloride the hot zone temperature at the bottom of the roaster is fixed and is related to the temperature at which hydrolysis takes place. For most applications it is in the range 600-850°C. Starting therefore with a fixed temperature at the base, a temperature pattern is set up in the roaster, the final temperature at the top and the temperature gradient being determined by a number of parameters such as concentration of feed, size of roaster, through-put, etc.

For maximum thermal efficiency in such a system it is necessary to operate with the lowest possible gas exit temperature. Counter-current operation of the roaster permits close control of the exit gas temperature, this usually being in the range 300-350°C. The addition of a direct contact pre-concentrator to the circuit reduces the effective gas exit temperature even further to below 100°C. This reduces the cost of operation considerably and results in regeneration of HCl more economically than by other processes.

Throughput and efficiency of hydrolysis are greatly influenced by the contact pattern between the falling drops of chloride solution and the rising stream of combustion products. To obtain maximum contact time a spiral motion is imparted to the combustion products entering the roaster by mounting the burners tangentially. Careful placing of the spray nozzles is required to ensure that oxide is not deposited on the walls. No exact science exists for the precise determination of the gas and spray patterns but investigations on a large number of commercial plants have given reliable design parameters. Operation over a wide range of these parameters is very stable and the roaster can operate at reduced throughput without significant loss of efficiency.

In the evaporation the liquid droplets are converted into hollow spherical particles of solid chloride. When these particles reach the hotter zone of the roaster and are hydrolysed the resultant product is a hollow spherical particle of oxide. Its ultimate particle size is usually extremely small, and this property is often of particular value in the subsequent use of the oxide. Thus, iron oxide produced in this type of plant is a suitable raw material for ferrite and pigment production.

Most of the heat input to the roaster is used for evaporating water. The heat of reaction and sensible heat in the product oxide are generally small by comparison. The concentration of chloride in the original feed is therefore of considerable importance. Any given sized roaster will have a maximum heat input. Thus the maximum output of oxide or hydrogen chloride is directly related to the feed concentration. This type of roaster is not the best tool in which to carry out evaporation either from consideration of efficiency or capital cost. It is therefore in many cases advantageous to preconcentrate the liquor before feeding to the roaster. In most cases superheat in the exit gases will preconcentrate the feed. Others require an evaporation step.

Feed concentration is also of great importance with regard to the final concentration of hydrochloric acid. Since HCl and water form an azeotrope at a concentration of approximately 21% w/w HCl, if the mixture of water vapour and hydrogen chloride in the gases entering the absorber is less than 21% w/w HCl, the maximum concentration achievable in an adiabatic absorber will be less than 21% w/w. Normally the acid produced is about 18/20% HCl. If higher concentrations of HCl are required then the HCl content of the gases must exceed 21% w/w HCl. To achieve this, higher chloride concentrations are necessary in the feed to the reactor and thus in most cases some form of preconcentration of the feed will be required. For product hydrochloric acid concentrations higher than 21% the adiabatic absorber is replaced by a cooled absorber system. Careful consideration therefore has to be given to each particular case to determine the best feed concentrations required to give lowest capital cost, maximum efficiency and the desired product acid strength.

Experience has shown that in spite of the very low wetting rate used in the adiabatic absorption column there is no risk of escape of HCl into the atmosphere with the stack exit gases. The Alkali Act lays down a maximum limit of 0·2 grains HCl/scf non-condensibles, and figures much lower are normally achieved without the need for a tail scrubber. Operation is so effective that it is possible to use dilute acid (up to say 2% HCl), such as would result from wash waters, as feed to the absorber and still not be in danger of exceeding the limit.

The success of the process can be measured by the increase in plant capacities which has occurred over the last seven years. From a pilot plant of 6 ft diameter the first commercial roaster of 15 ft diameter was commissioned in 1962, since when many plants have been operated with diameters of 10, 15, 20, 25 and 30 ft.

Although scale-up has been based on very conservative principles, the increase of plant capacity (which is approximately proportional to roaster cross-sectional area) is an indication of the successful operation of the process.

Applications

As an example of applications of the process we have chosen steel descaling.

In the rolling of steel to strip and other sections, a layer of scale forms on the outside, which must be removed before further finishing operations can be carried out. This scale consists of oxides of iron, being generally FeO adjacent to the metal, Fe_3O_4 in the intermediate layers, and Fe_2O_3 in the surface layers.

Traditionally this scale has been removed by mechanical or chemical means, or a combination of the two.

By far the most common treatment has been to immerse the steel in a bath of sulphuric acid, in which the bulk of the scale is dissolved. Disposal of the resulting spent acid has presented serious problems, and although various regeneration processes have been developed the overall economics have proved unfavourable.

If hydrochloric acid is used for scale removal, then the acid is easily regenerated from the resulting ferrous chloride liquors by the Woodall-Duckham regeneration process. Apart from the hydrochloric acid which is recycled to the descaling line, iron oxide is produced as a dry free-running powder.

Spent pickle liquor can vary considerably in composition according to the type of pickling line, contact time, type of steel, type of line heating, i.e. live steam or indirect heat exchangers, and other factors. Some lines operate with a spent liquor containing virtually no free acid, e.g. 25% w/w $FeCl_2$, 0·5% w/w HCl, while others can have a composition of 14% w/w $FeCl_2$, 8% w/w HCl. The Woodall-Duckham regeneration process can handle any composition of spent liquor found in practice. The presence of free HCl presents no difficulties. This is evaporated off with the water in the upper part of the roaster. The $FeCl_2$ reacts according to the equation

$$2FeCl_2 + H_2O + \tfrac{1}{2}O_2 \rightarrow Fe_2O_3 + 4HCl$$

The iron oxide is very pure and can be pelletised for return to the blast furnaces, or used as a valuable feed material for pigments or ferrites.

Disposal of rinse water resulting from washing the pickled steel is frequently a problem in a steel mill and can often necessitate a neutralisation plant.

Rinse water contains up to 1-2% HCl and is too dilute to justify handling in a regeneration plant. It is, however, perfectly acceptable as feed to the absorption column. Some modern large-scale steel strip pickling lines are designed to limit the production of rinse water to be no greater than the absorber requirements and to use it first in the pickle line fume scrubber. No neutralisation plant is then required.

Since the costs of regeneration are generally much lower than the cost of new acid purchase, the overall descaling process becomes more economic. At the same time use of the process solves a serious effluent disposal problem.

When the iron oxide is sold as a by-product then significant profits can be shown on the regeneration stage.

In addition to the straightforward benefits outlined above, there are very significant secondary benefits from the use of hydrochloric acid. Because the acid is more selective in dissolving iron oxide than metal, less pitting of the steel occurs, and a better surface finish results. Also hydrochloric acid dissolves iron oxide more rapidly than sulphuric acid, so that scale removal is faster. This permits much higher throughputs and the elimination of the expensive scale-breaking equipment used on sulphuric acid pickling installations.

Economic data for operation of a pickle liquor regeneration plant

Capacity of plant	2,900 l/h of liquor containing 25·0% w/w $FeCl_2$ and 0·5% w/w HCl
Recovered acid make	2,700 l/h of 18% w/w HCl
Wt. of Fe_2O_3	570 kg/h
Heat input	$2{\cdot}20 \times 10^6$ Kcal/h
Power	150 kW
Rinse water (to absorber)	2,600 l/h
Process water	200 l/h
33% HCl make-up*	10-20 l/h
HCl in exit gas	0·1 gm HCl/Nm^3 non-condensibles
Cl^- in Fe_2O_3	0·5% or less

*Excluding losses on the pickle line, e.g. drag-out, fume loss etc.

VISIT 15: BISRA MECHANICAL WORKING DIVISION

Situated at
Hoyle Street,
Sheffield ENGLAND

DATE: Tuesday 24 November 1970

MET: Mr Geo Sturgeon

1. ALLOY POWDER PRODUCTION

Stainless steels of the composition of 304, 316, 430 have been water-atomised in the laboratory and can be dried and used for compaction into strip without any post-treatment. The powder used averages 20% minus 100 plus 200, 30% minus 200 plus 300, 30% minus 300 plus 400 and 20% minus 400 mesh. Molten metal is poured at say 50°C above melting point and the ratio of water volume to metal governs powder fineness. 90% conversion into usable powder is obtained and a large proportion of particles of the powder have the desired irregular shape. Alloy powders are presently made by BSA (1,000-1,500 tons per annum) and sold at 8 shillings per lb and by Hoganaes in USA (soon to be produced in Sweden), costing 10 shillings per lb in USA. These prices appear very inflated and Mr Sturgeon believes that a powder suitable for compacting into strip, if not suitable without further blending etc for sale for powder-metallurgical uses, can be made quite cheaply. The whole process is being tried at present and budget estimates have yet to be made, but Mr Sturgeon thinks alloy-powder production could be viable at 1 ton per hour.

2. STAINLESS STEEL STRIP PRODUCTION

BISRA have been producing 4 inch wide strip from 0.08 inches down to foil. Mr Sturgeon thought that plate production from powder could not compete due to compaction speed limitation. Air has to be expelled in compaction and tends to fluidise the powder. No binder is used. Hacksaw blades may be a viable product and Mr Stugeon showed stainless strip in various stages of production from powder and finally rolled as cladding on to aluminium window T-section. The process is nearing possible commercialisation and he could not show me the operation. We could write later to find whether the British Steel Corporation would release the know-how. The only published paper appears in Powder Metallurgy 13 (26) 1970.

3. IRON POWDER GENERALLY

Mr Sturgeon said that mild-steel powder was not yet being produced at Port Talbot, and that his division was probably doing more than Stanton and Stavely on preformed shapes for forging. He agreed that adoption of this process would take time, but said that the Ford Motor Company in Europe were keen to use it and that they had tested powder-forged connecting rods and gears. He thought that cold isostatic compression would be used for larger components such as whole crankshafts, which had been produced by this method. Each part would have to be assessed to determine whether powder forging was the most economic method.

VISIT 16: BISRA IRONMAKING DIVISION

Situated at
Grangetown,
MIDDLESBROUGH. ENGLAND

DATE: Wednesday 25 November 1970

MET: Mr J.M. Ridgion - Head of Tees-side Laboratories
Dr R. Wild - Deputy Head of Tees-side Laboratories

1. GENERAL

Dr Wild is an authority on direct reduction processes, and had visited North America to see and discuss direct reduction processes about six weeks previously. We discussed each in turn and he gave his assessment, after which they gave me a tour of the laboratories.

They considered that the budget price I quoted for gas was not really cheap and should be capable of reduction.

2. DIRECT REDUCTION PROCESSES

2.1 Using Natural Gas As Sole Reductant

Dr Wild considers that if a plant was to be built immediately, the only process 100% sure of successful operation would be HyL. The process is somewhat inefficient thermally but recent use of pellets at Pueblo has improved this as well as output and uniformity of reduction. Gas consumption would now be 20,000 cu ft per metric ton of iron in reduced pellets and in a 1 million ton per annum plant might be reduced to 18,500 cu ft. The process is engineered by Swindell-Dressler who have enormous experience with gas reformers and the Mexican operation claim to have not replaced a reformer tube in 2 years. Their first plant, still operating, is 200 tpd.

The Mildland-Ross plant at Portland (400,000 tons per annum) consists of two shaft furnaces, that at Hamburg will use only one furnace of double capacity, but this is as large as can be built at present. In the Portland plant only the reduction section was built by Midland-Ross, and they consider that the electric furnace installation does not use their product to full advantage. At Georgetown also their plant will be only an adjunct to the steelplant, but the Hamburg plant which will use four Concast strands is designed as an integrated operation. The electric furnaces will use 80%

of reduced Swedish pellets, 10% in-plant scrap and 10% purchased scrap. The Portland plant is having to use purchased pellets of higher sulphur content than anticipated. The sulphur content is not high enough to cause trouble in steelmaking but is being removed in the reduction stage and over-loading the desulphuriser. One furnace has now been modified so that sulphur is re-absorbed by the cooling pellets, and the design output and fuel efficiency is stated to be attained. Gas requirement of the process is stated to be 13 million BTU/gross ton of product containing 91% Fe.

The proprietors of the Purofer process claim to be able to use green pellets, so that their process should be the cheapest. Mr Voice of BISRA saw the plant and said it had been made to work, but the efficiency is not known. von Bogdandy of HOAG claims that Midland-Ross have infringed his patents which contributes to Midland-Ross's unwillingness to admit visitors. Both are shaft-furnace processes and both can recycle top-gas. Purofer gas reforming is discontinuous using a reformer and a recuperator but probably insensitive to sulphur.

The Armco process is very similar and the reformer only differs in engineering - the method of getting gas in and out, pellet loading, etc. The process will probably go ahead if Armco can obtain sufficient capital.

Summarising these processes, it can be seen that Armco, Purofer and Midland-Ross on the one hand are essentially similar continuous shaft furnace processes, and on the other hand HyL is a batch process.

Mr Wild considered that the new process being developed by Allis-Chalmers should also be watched. This company wish to exploit their grate-kiln expertise, but have rejected their Acar process in favour of the present development which consists of a conventional drying-firing grate, preheating pellets to 1,100°C, followed by a rotary kiln which can admit either air or gas through rings of six peripheral ports spaced along the length of the kiln. Valves are arranged so that while the ports are submerged, gas is admitted and immediately cracked in the bed of hot iron ore. While the ports are not submerged, air is admitted. In this way one obtains the advantage of a rotary kiln with solid reductant in that a strongly reducing atmosphere can be maintained in the bed while combustion can be carried out in the less-reducing gas phase. The need for gas reforming equipment is also eliminated. The process is a little less efficient thermally, using 15.5 to 16 million BTU per ton of product, but can also operate with 10 to 20% solid reductant in the pellet and possibly with green pellets, reducing the fuel requirement (gas plus solid) to 13 to 14 million BTU per ton. Allis-Chalmers are very keen to apply their process, possibly as a joint venture with a steelmaker, but in any case to make steel by 1973.

The fluid bed processes comprise Fior and US Steel. The latter's process resembles their earlier Nu-Iron process and the product is known as HIB. The Venezuela plant, contrary to some reports, is only designed to make a 70% reduced product. Their Monroeville laboratories are testing the feasibility of making a 90% reduced product, but say that this will require more than a "once-through" process. The process is complicated and has a long throughput time and involves considerable reformer chemistry.

2.2 Using Solid Reductant

The SL-RN process seems to be on the way up. There are reasons for the New Zealand problems as an 8% TiO_2 concentrate, green pellets, and lignite with 25% moisture are being used. Stelco-Lurgi are putting considerable effort into solving these problems. The South Korean operation is working at specified output producing 70% metallised feed for electric pig-iron furnaces. The Sudbury, Ontario plant to produce 90% Fe, 1.5% Ni pellets is now being commissioned - it is a beautiful and highly integrated plant to treat pyrrhotite.

The Midland-Ross Heat Fast process is claimed to be technically satisfactory but is actually a complete engineering failure due to the metal stresses in the large annular firing arrangement. The plant built in the USA to produce oxide pellets initially then reduced pellets was converted to an Allis Chalmers grate-kiln plant.

McKee are not active in the field of direct reduction. The DLM process is operating at 400,000 tons per annum in Alabama to produce foundry iron for cast-iron pipe. There have been rumours of trouble with the electric arc furnace.

3. BISRA RESEARCH

The pilot plant area is well set up and equipped to study crushing and screening problems at quite high throughputs.

The Continuous Ironmaking Process (CIP) being investigated in the Teesside Laboratories resembles the Dored process, but BISRA are endeavouring to overcome refractory attack by the high FeO slag by rotating the 5 ft diameter by 10 ft vessel at high speed so that the slag is contained within the centrifuged molten metal. This creates numerous engineering and safety problems and such research is obviously expensive.

I met Mr R.A. Oliver with whom we had correspondence regarding reducibility testing. BISRA are engaged in standardising the VDE test for the Standards Association and mentioned that the constitution of the Standards Association sub-committees was changing in favour of technical people. They are also working on a reducibility test (Burghardt) using a constant load of 9 psi on the burden.

The laboratories are also studying sintering under high suction (100 inches W.G.)

VISIT 17: MANNESMANN PULVERMETALL GMBH

Situated at
Mönchengladbach.

DATE: Friday 27 November 1970

MET: Hein K.B. Schmidt - Sales Manager
Dipl Phys Joseph Hewing - Metallurgist

The $2^1/_2$ hour visit was in the form of an introductory discussion with Mr Schmidt, who spoke English and was well-informed regarding the processes, followed by a tour of the plant with him, and finally a discussion at which Mr Hewing was present to answer more technical questions.

1. INTRODUCTION

Mannesmann produce 25,000 tons per annum of iron alloy powders for all purposes by atomisation of liquid metal, with or without subsequent heat-treatment. Leaflets describing the products for powder metallurgy have been supplied.

Mr Schmidt said that his company sold little powder in Australia and that the Australian market was satisfied by Hoganas, Domtar and QMP. His company could not compete pricewise after paying all the freight costs, duties etc. He said that the cheapest grade of 40 mesh iron powder for welding rod coating would sell in large quantities in Germany for about DM 600 ($A150) per ton. When I estimated a price of around $A180 (DM 720) in Australia he said that DM300 ($A75) might be consumed in these on-costs. Since they could not economically supply from Germany, Mannesmann might therefore be interested in participating in a joint venture to manufacture in Australia.

Mr Schmidt said that some years ago Mannesmann had calculated a minimum viable output of iron powder by atomisation as 400 tons per month (5,000 tons per annum). The Australian consumption of 3,000 tons per annum is not far short of this. Now with increasing costs and reduced prices the minimum plant-size in Europe would be nearer 800 tons per month. Australia could not expect to gain much of the Japanese market as, in addition to the present Kawasaki plant, Kobe have been operating a pilot plant and will next year have a commercial plant using A.O. Smith's process. This is a water-atomisation process using an elliptical arrangement of jets which is understood to produce more coarse oversize than does Mannesmann's circular jet arrangement.

If forced to do so by rising costs, Mannesmann will transfer atomisation to the site of their Hooking(?) steel plant several miles away and use hot metal from the oxygen converter.

2. PLANT OPERATIONS

The works comprise (a) the older RZ operation consisting of cupola melting to give a relatively high-carbon metal which is air-atomised and quenched in water. In the subsequent heat-treatment the excess carbon reacts with the oxide layer formed in atomisation to give, without external reductant, a reduced iron powder of form suitable for powder metallurgy. The second, newer operation (b) is the WP plant, in which scrap is melted in an electric furnace and the melt atomised with water. It was intended to thus produce without post-treatment a more spherical powder for welding-rod coating. However, research by Mannesmann showed that the water-atomised powder, if reduced by reducing gas, gave a powder superior in many respects to the RZ product for powder metallurgy.

It suits Mannesmann to keep both plants operating, but if a new plant were now to be built Mannesmann would use water atomisation. The water-atomised powder is suitable without further treatment for acid rutile-coated welding rods, but basic lime-coated rods require a lower oxygen content which can be most cheaply obtained by use of the RZ product.

2.1 RZ Plant

High-purity scrap, consisting of offcuts from Mannesmann's tube plant is fed with a mixture of coke and graphite (to reduce the overall sulphur content of the fuel) and limestone to two cupolas operated in parallel. Slag is tapped but no other refining of the metal is possible, and the 5%C metal is poured from a ladle into the atomiser. Jets of air break up the falling stream and the droplets are quenched in water in the bottom of the chamber. The water-level and hence the time exposed to air governs the oxidation of the particles. The iron particles are removed from the water by cycloning, filtered on a continuous vacuum filter and dried. The product is analysed in batches, blended to obtain the correct C:O ratio and charged into 15 inch by 15 inch trays which pass through seven or eight electrically-heated furnaces in parallel, each furnace being 12 to 15 ft long with a cooling zone of similar length. No external reductant is used and the CO produced burns gently at the ends of the furnace. These Elino furnaces are capable of 1,300°C but were operating at 1,050°C at the time. The cake of powder, now silvery-grey can be just broken by hand, and is fed into a breaker followed by a toothed-disc disintegrator. Oversize amounting to less than 1% of production is removed by Siebtechnik vibrating screens and sold to the chemical industry.

This was frankly admitted to be a rather archaic and labour-intensive operation, but cupola-melting was stated to be cheaper than use of an electric furnace.

2.2 WP Plant

Scrap, of high grade, but not of such quality and uniformity as used in the RZ plant, is charged to a 10 ton electric furnace which is operated on an off-peak contract from 1300-0800 hours daily. Cr and Mn were stated to be removable, but copper and nickel cannot be, and Cu must be less than 0.1% in the scrap. The furnace is tapped into a ladle which is transferred to a position over one of three tundishes beneath each of which are water jets arranged circularly to give a conical envelope of high-pressure water. The particles fall into water and are recovered and dried as in the RZ plant. For acid-welding-rod coating the dried product is simply screened as in the RZ plant. For powder-metallurgical purposes the dried product is fed to Elino furnaces similar to those in the RZ plant, but with cracked ammonia supplied for reduction. (Mr Hewing said later in reply to my question that reformed natural gas would be satisfactory but would result in a carbon content of about 0.06% C, similar to the RZ product, instead of 0.002% currently obtained in the WP product). The small electrically-heated Elino furnaces are the bottleneck at present, and employees were busy on a much larger single unit (nicknamed the "jumbo-jet") which will be gas-fired and operating next year. The reduced product is treated in a manner similar to that in the RZ plant.

2.3 Powder Metallurgy Applications

Mannesmann do not consume any of their own powder, but have a powder metallurgy shop equipped with a full range of presses, sintering furnace etc. for assistance to clients. They design dies and supply presses which must exert a pressure of about 6 tons per sq. cm. A large range of products were exhibited and tool cost was quoted as averaging $5,000 so that powder metallurgy is only applicable to large runs. Powder-metallurgy parts aggregate to 3 kg in an average car (4 kg in a VW).

3. CONCLUSIONS

For a WP plant of 500-800 tpm (6,000-10,000 tons per annum) Mr Schmidt quoted a capital investment of DM12 million (without reduction) or DM 15 million ($A3.8 million) (with reduction) under German conditions. This includes the buildings (but not land) the expensive heavy foundations, etc. Water is recirculated at a high rate.

Mr Schmidt said that Mannesmann would be happy to discuss with a potential S.A. steelmaker a co-operative iron-powder operation.

VISIT 18: MONTECATINI INSTITUTO RICERCHE G. DONEGANI

Situated at
Via Del Lavoro 4,
NOVARA.   ITALY

DATE: Monday 30 November 1970

MET: Dr G. Sironi

The object of discussion with Dr Sironi was to assess the feasibility of applying fluid-bed magnetising roasting, upgrading and pelletisation, (possibly with reduction to metallic iron at some stage) to South Australian pyrite cinders.

Dr Sironi outlined Montecatini practice at Follonica, which I was to see the following day and supplied a number of reprints. 1,600 tons per day of calcine is produced assaying 50% Fe and yielding after dry magnetic separation using Sala-Mortsell machines 1,100 tons per day concentrates of 65-66% Fe and thence 1,100 tons per day pellets. All magnetic separation is now done dry. Magnetising reduction is carried out in a 7 ft diameter reactor at 750°C and a space velocity at temperature of 5 ft per sec, using direct injection of Bunker C fuel oil. Montecatini have a capacity of 80 tons per day per square metre of grate area, but space rate, size, capacity, and disposition of oil injection guns will depend on the particle size of the pyrite cinder being considered. Dr Sironi considered that a temperature of 900 or (probably) 950°C would be required if natural gas were to be directly injected due to the stability of methane. A less thermally-efficient process would result. Fuel consumption by Montecatini is 24 Kg oil per metric ton of calcine, and a residence time of 20-25 minutes is ample. However one calcine tested by Montecatini simply did not reduce under these conditions.

Montecatini have considered fluid-bed reduction to metal but found a number of difficulties in the course of experimental work in a 4 inch diameter reactor. Dr Sironi considers the process to be expensive in capital and heat, and while it might be justified for making high-priced iron powder, production of a scrap substitute is not economic. The project has therefore been shelved.

On the subject of scale, Dr Sironi considered that Montecatini were pelletising at the minimum economic throughput - approximately 1,000 tons per day or approximately 300,000 tons per annum. He would not consider pelletising

70,000 tons per annum. I pointed out the possibility of avoiding firing by charging green pellets to a reducing furnace, or the possibility of using a cold-bonding process. He said that he had tried charging his synthetic magnetite pellets green to a shaft, but they exploded. If a hematite were available to blend with the synthetic magnetite to reduce the exothermicity of the re-oxidation and hence cracking of the pellets (and also to reduce the impurity content) it would also help the economics. Montecatini themselves blend 10 to 20% of hematite with their magnetite at present to reduce the exothermicity of firing. The "cold-bonding" process using lime developed by Kilstehdt at Stockholm and Volin at Michigan, and the Grangesberg process using cement (which also introduces acidic oxides) were agreed by Dr Sironi to permit economic operation at lower throughput, but unlike the firing process, do not eliminate any sulphur which is a problem with pyrite calcines. Montecatini find that reduction eliminates 50% of sulphur at the 1% level, but less at the 0.1% level.

Another important consideration is the non-ferrous metals content. Steelmakers generally require Cu <0.04, Zn <0.06, Pb <0.06, As <0.03, S <0.02%. Montecatini operate a pilot plant at Follonica in which they are investigating at a scale of 50 tons per day, the removal of non-ferrous metals by fluid-bed treatment with chlorine. They have worked on numerous cinders - one at present assays: Fe 55, Cu 0.85, Zn 2.8, Pb 1.3, As 0.4%. A previous cinder assayed Cu 0.4, Zn 0.6, As 0.1%. A difference from the LDK process is that the latter does not remove As, and also that strong pellets can be made <u>after</u> removal of non-ferrous metals. However to the present S has only been reduced to 0.05% and a level of 0.02% is required.

On the whole Dr Sironi stressed the need for caution in any such venture to treat pyrite calcines, and said that the unit cost at the low throughput, possible penalties for impurities, and behaviour of the pellets would all have to be carefully assessed.

He said that Montecatini's pilot plant was suitable for reduction of ilmenite to the ferrous state and that they had a technically feasible route (similar to the Sinha-Waugh process) to produce synthetic rutile, using the know-how they had obtained on direct fuel injection. If they could find a partner producing ilmenite they would go ahead with this.

VISIT 19: MONTECATINI EDISON S.P.A.

Situated at
FOLLONICA ITALY

DATE: Tuesday 1 December 1970

MET: Dr Bruno Viviani of Montecatini) who are in charge of the pilot plant
Mr R. Moro of Dorr-Oliver, Milan) for non-ferrous metals recovery. The latter interpreted for the whole of my visit.
Dr Angelo Lorenzo - Manager of the Sulphuric Acid and Iron Ore Pellet plant.

1. INTRODUCTION

Montecatini mine pyrite 10 miles away and truck it to the acid plant, which has been described in various papers, (pages A-169 to A-178). I confined my attention to the stages of processing the calcine from removal of hot calcine from the pyrite roaster to production of green balls of iron ore concentrate.

Montecatini have also on the same site a "pilot-plant" for testing their chlorination process for removal of non-ferrous metals from pyrite cinders. They have tested Russian cinders, are now testing Rio Tinto cinders and expect to test Bulgarian and Portuguese cinders. This is a joint project between Montecatini and Dorr-Oliver. Mr Moro also spoke of the likelihood of using the plant to investigate production of synthetic rutile from ilmenite, when the current programme is complete.

2. CONVERSION OF PYRITE CINDER TO IRON ORE PELLETS

Some changes have taken place since earlier published descriptions, notably that all reduction is accomplished by direct fuel injection, and all magnetic separation is done dry. There are four roasting and reduction lines in parallel, the reduced calcine from all four being combined for magnetic separation. The reduction section is sketched in Figure A19-1 page A-179

The tuyeres are hexagonal cast alloy, protruding 3 to 4 inches above the grate with holes drilled in the flat faces.

3. PILOT PLANT FOR NON-FERROUS METAL REMOVAL

This plant, which treats 50 to 60 metric tons of calcine per day stands approximately 45ft by 22ft by seven levels of 8ft high (1 x 12 + 7 x 8 ≈ 70ft) exclusive of the rotary dryer and was stated by Mr Moro to have cost approximately \$1 million. It consists of one fluid-bed reducing reactor 5 ft diameter (6 ft diameter freeboard section) operated by direct fuel injection at approximately 700°C, followed by two chlorination reactors 4 ft diameter (6 ft diameter free board section) of which only one has been found to be necessary. The reducer eliminates some arsenic but its main purpose is to supply heat to the chlorination stage by means of the exothermic oxidation of magnetite, since it is preferable not to burn fuel in that stage. Chlorination is carried out with a chlorine-air mixture at 900 to 1,000°C A brochure is reproduced on pages A-180 to A-185.

Incoming cinders are dried, screened at 0.5 mm and the oversize ground. The chloride-bearing gases after dust removal are scrubbed with recycled solution which is simply filtered at present. The purified calcine goes to the commercial plant. An absorber had been included to remove excess chlorine from the tail-gases but it has been found that 0.8% S in the reduced calcine gives sufficient SO_2 to reduce all excess Cl_2 to HCl which is absorbed in the scrubber along with the SO_3

Rio Tinto calcine was quoted as containing Zn 2.8, Cu 0.8, Pb 1.3, As 0.7% and current final assays are Zn 0.08 to 0.1% similarly Pb, Cu 0.02 to 0.03% similarly As. Little work has been done on Ag and Au but Mr Moro thinks 60 to 70% could be recovered. No work has been done on Co or Ni.

4. FURTHER ACTION

Montecatini and Dorr-Oliver would be ready to reply in detail to any further enquiries.

FLUOSOLIDS REDUCTION OF HEMATITE BY MONTECATINI

By U. P. Colombo, T. D. Heath, G. Sironi, and G. Tomasicchio

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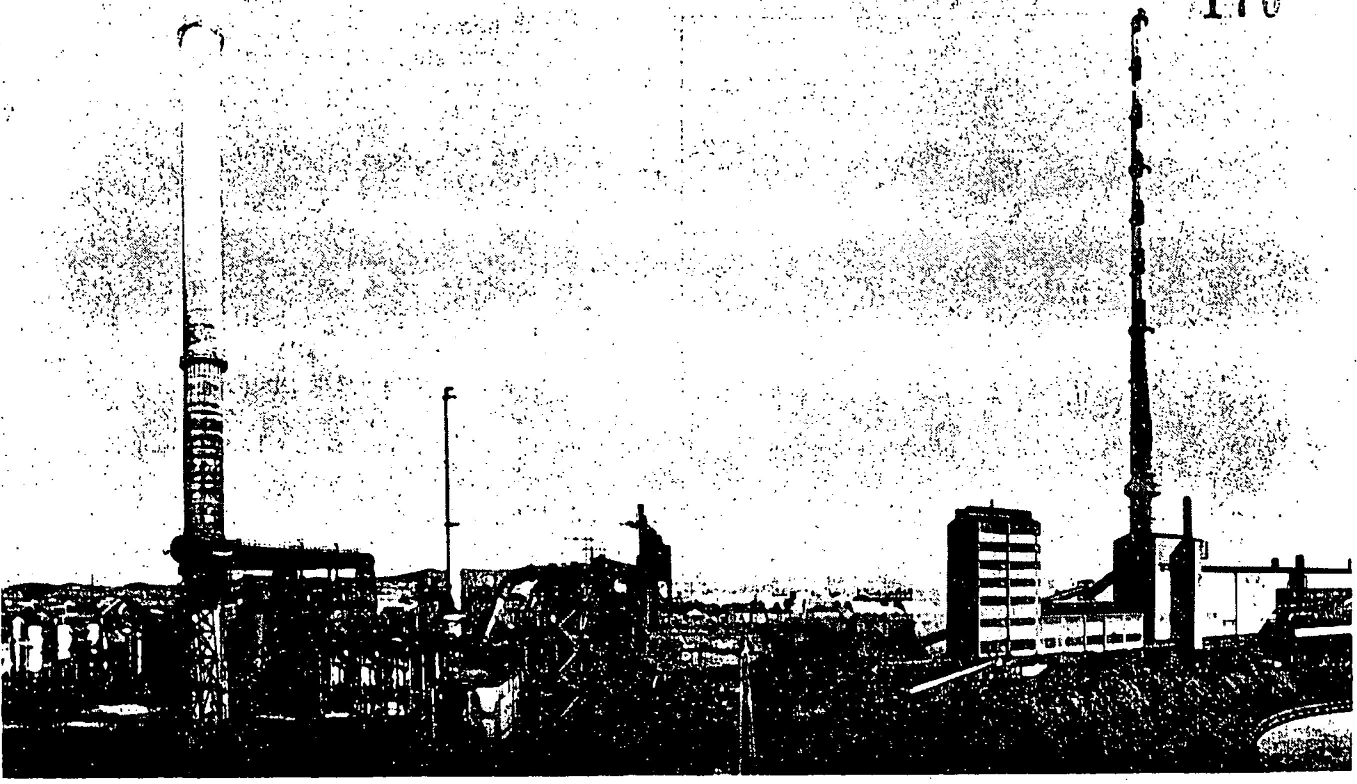

Montecatini's Follonica plant.

FLUOSOLIDS REDUCTION OF HEMATITE BY MONTECATINI

A major new plant in Italy is using a process that gets the last "squeal" out of pyrite. Key to the process is fluid-bed reduction by injection of fuel oil.

by U. P. Colombo, T. D. Heath, G. Sironi, and G. Tomasicchio

Roasting of iron pyrites for sulfur dioxide, used for the production of sulfuric acid, also yields hematitic calcines. These have so far been considered a by-product of limited commercial value.

While development of the fluidized bed roasting technique has permitted the construction of large-capacity units and the recovery of considerable quantities of the heat produced in the oxidation of the pyrites, it has further aggravated the problem of utilizing the calcines for metallurgical purposes. This is largely because the pyrite decrepitates to a considerable extent in the course of the fluidized bed roasting process, so that the resulting calcines are extremely fine.

Increasing competition from native sulfur and by-product sulfur recovered from hydrocarbons has placed the pyrites in a critical position on the world market. This situation is particularly serious in Italy, which is the world's fourth largest producer of pyrites (after Japan, the USSR, and Spain).

The Montecatini Co. which in 1963 extracted from its mines 1,375,000 net tons of pyrites, corresponding to about 90% of the whole Italian production, has met this situation by the complete utilization of pyrite. They now regard it as an iron ore as well as a sulfur ore. The final objective has been the production of high-grade iron oxide pellets eminently suitable for blast furnace feed.

In order to put the new process on an economic basis, Montecatini has concentrated the treatment of pyrites in a large plant built at Follonica, on the Tyrrhenian coast, in the neighborhood of the pyrite mines.

►►

U. P. COLOMBO and G. SIRONI are with the G. Donegani Research Institute, Montecatini Soc. Gen., Novara, Italy. T. D. HEATH is with Dorr-Oliver Inc., Stamford, Conn. G. TOMASICCHIO is associated with Dorr-Oliver, SpA Milan, Italy. This article was presented at the 94th AIME Annual Meeting.

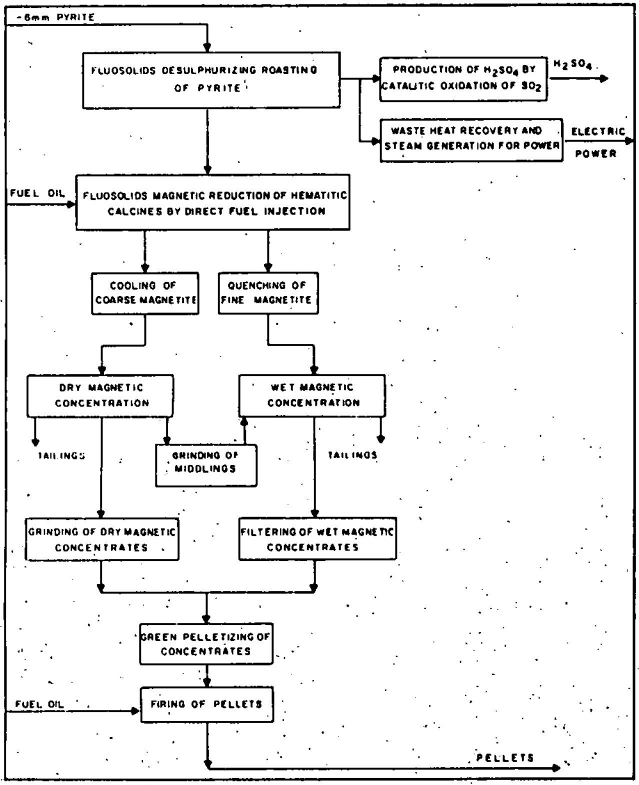

Fig. 1—Flowsheet of the process.

THE MONTECATINI PROCESS

Fig. 1 shows the flow-sheet of the Montecatini Process[1], as applied in the Follonica plant. It consists of the following steps:

1.—Dead roasting of 2200 metric tpd run-of-mine pyrites containing 35-39% S and 35-39% Fe in four parallel trains of FluoSolids* reactors (Fig. 2). resulting in a production of 2200 metric tpd of sulfuric acid (100% basis), 570,000 kw-hr of electricity (from waste heat), and 1500 metric tons of hot hematitic calcines.

2.—Reduction of the hot hematitic calcines to magnetite in four parallel trains of FluoSolids reactors (Fig. 3).

3.—Up-grading of the magnetite by combined wet and dry magnetic concentration, resulting in the production of 1100 metric tpd of concentrates with about 67% total Fe and 12% moisture.

4.—Green pelletizing in a rotary drum and firing on a travelling grate, yielding 1100 metric tpd of high purity iron oxide pellets with 66-67% Fe.

In the case of pyrites this process combines for the first time in one factory unit all four steps listed above and offers the following advantages:

1.—Cheaper, non-selective ore mining methods are employed for the production of the pyrite needed in the plant.

2.—Crude ore is not submitted to any dressing operation, but is fed to the roasting plant as run-of-mine pyrite crushed to ¼ in.

3.—Hot hematitic calcines, as discharged from the roasting step, are immediately utilized in the successive reduction step, without appreciable heat losses.

4.—Feeds to the magnetic separation and pelletizing steps need only a minor amount of grinding due to the considerable degree of decrepitation of the pyrites in the course of the roasting step.

5.—Magnetic concentration by both dry and wet methods produces a combined concentrate with almost the right amount of water needed for the green balling step.

6.—Thermal balance of the pelletizing step is substantially improved by the exothermic heat of oxidation of the magnetite to hematite. Furthermore, a better desulfurization of the product is achieved because of the resulting homogeneous temperature profile throughout each pellet.

MAGNETIC REDUCTION WITH THE FLUOSOLIDS SYSTEM

Magnetic reduction of the hot calcines prior to magnetic concentration can be considered, on account of its novelty, as the most significant step of the whole Montecatini process. It will be described in detail.

From the various techniques proposed for the reduction operation in the literature,[2,3,4] fluidization was selected not only for its well known inherent advantages, but also because of the nature and particle size distribution of the pyrite calcines obtained from FluoSolids roasting. Furthermore, only a fluidized-bed technique permitted the utilization of a new method of reduction involving direct use of oil as a reductant in the fluid bed. The method is called direct fuel injection and will be referred to in this paper as DFI.

Laboratory and Pilot Work

This cooperative Montecatini-Dorr-Oliver process was started in 1960, with extensive laboratory and pilot work on the reduction step of the process. The previous experience of Dorr-Oliver was primarily related to the reduction of natural hematites, and the need was felt for a thorough comparative investigation on the artificial hematite from pyrite.

As a first step, a thermogravimetric study on the reducibility of natural hematites and pyrite calcines was performed. High-iron-content samples were tested in order to eliminate side effects due to different gangue composition. A Stanton mass-flow thermogravimetric balance was used for this investigation.

Typical reduction curves at 650°C (1202°F) are shown in Fig. 4. Pyrite calcines exhibit a higher rate of reduction than natural hematites. This was observed throughout the whole temperature range under study (300° to 750°C).

The reoxidation of reduced samples was also investigated with the same thermobalance, and the results obtained at 300°C (572°F) are shoyn in Fig. 5.

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* FluoSolids is a registered trademark of Dorr-Oliver Inc.

Table I— Specific Surface Areas of Natural and Artificial Hematites, Compared with those of their Magnetic Reduction and Re-Oxidation Products (Fraction 100-325 Mesh)

	Specific surface area, m² per g		
	Unreduced Product	Reduced Product	Re-oxidized Product
Pyrite calcine	1.65	1.81	1.80
Hematite from Italy	0.20	1.46	0.93
Hematite from Brazil	0.21	1.54	1.12

The different samples are characterized by more similar behaviors in the reoxidation kinetics than in the previous reduction.

The specific surface areas of unreduced, reduced and reoxidized samples were measured by a modified Brunauer-Emmett-Teller[6] method, using a Perkin-Elmer-Shell Sorptometer. The results are shown in Table I.

Correlation of specific surface area data with reduction and reoxidation kinetics of the different samples led to the conclusion that the unusually high reactivity of hematitic pyrite calcines was due, essentially, to their high specific surface area arising from the structural changes during pyrite roasting. Magnetites obtained by reduction of natural hematites have specific surface area values comparable to those of reduced pyrite calcines, and this explains their similar behavior in the reoxidation kinetics. The mechanisms of the reoxidation of artificial magnetites to yield magnetic γ-Fe_2O_3, or non-magnetic α-Fe_2O_3 (hematite), were also investigated, and the conditions required to avoid any reoxidation to non-magnetic hematite ascertained. Full details of this investigation have been published elsewhere[6,7,8]. It has also been confirmed that the presence of water molecules and of hydroxyl ions in the lattice of the magnetite stabilizes its spinel structure, which disfavors the formation of α Fe_2O_3 in a possible reoxidation[9]. The findings of such studies led to the introduction of a steam-fluidized magnetite cooling step in the process.

Table II—Continuous Tests in a 4-in. Diam FluoSolids Reactor: Hematitic and Reduced Pyrite Calcines (Reduction with Separately Generated Gases)

Particle size distribution: cumulative % by wt

	Wt Distribution	Mesh (Tyler)								
		+4	8	14	35	65	100	200	325	−325
Composite unreduced calcines	100	0.1	13.3	27.2	48.3	61.8	70.1	82.1	87.3	12.7
Reduced calcines (bed)	92.8		12.1	22.2	46.0	58.9	80.0	92.0	96.1	4.9
Reduced calcines (cyclone)	7.2					0.5	2.1	7.5	17.1	82.9

Chemical analyses: % by wt

	Fe Total	Fe Ferrous	S Total	S Sulfate	S Sulfide	CaO	MgO	SiO_2	Al_2O_3	CO_2
Composite unreduced calcines	48.14	1.32	3.28	2.85	0.43	7.76	0.67	12.82	2.07	0.62
Reduced calcines (bed)	48.25	16.01	3.40	2.98	0.42	8.30	0.71	13.83	2.21	0.61
Reduced calcines (cyclone)	61.52	20.40	2.60	2.05	0.55	3.18	0.39	3.90	0.95	0.87

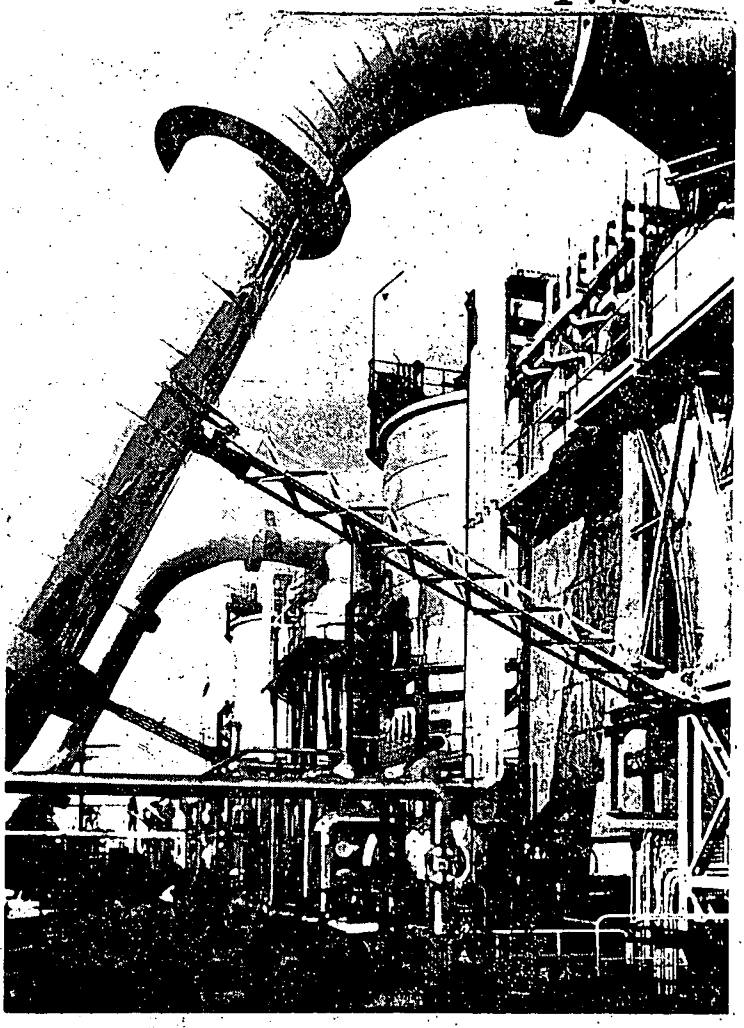


Fig. 2—The pyrite roasting plant.

Comparative reducing tests were made in a 4-in. diameter testing reactor on Montecatini pyrite calcines and on natural hematites. In one series of tests, reduction was obtained by the use of gaseous reductants; in another series, reduction was obtained by DFI.

Tests With Gaseous Reductants

From continuous single-stage tests using a fluidizing gas containing about 33% H_2 + CO and 67% N_2, it was found that the Montecatini pyrite calcines were amenable to fluidization and that they could be easily and homogeneously reduced, in spite of their widely distributed particle-size range. Davis tube tester determinations showed iron recoveries as high as 97% with 68% Fe in the concentrates. In Table II, the chemical and particle size analyses of unreduced and reduced calcines are presented. Table III shows reduction performance data on natural hematite compared to pyrite calcine. As predicted from the thermogravimetric work, the pyrite calcines can be converted essentially to 100% magnetite under milder reducing conditions than natural ores. It was also found that reoxidation takes place readily at temperatures above 250°C (482°F). ►►

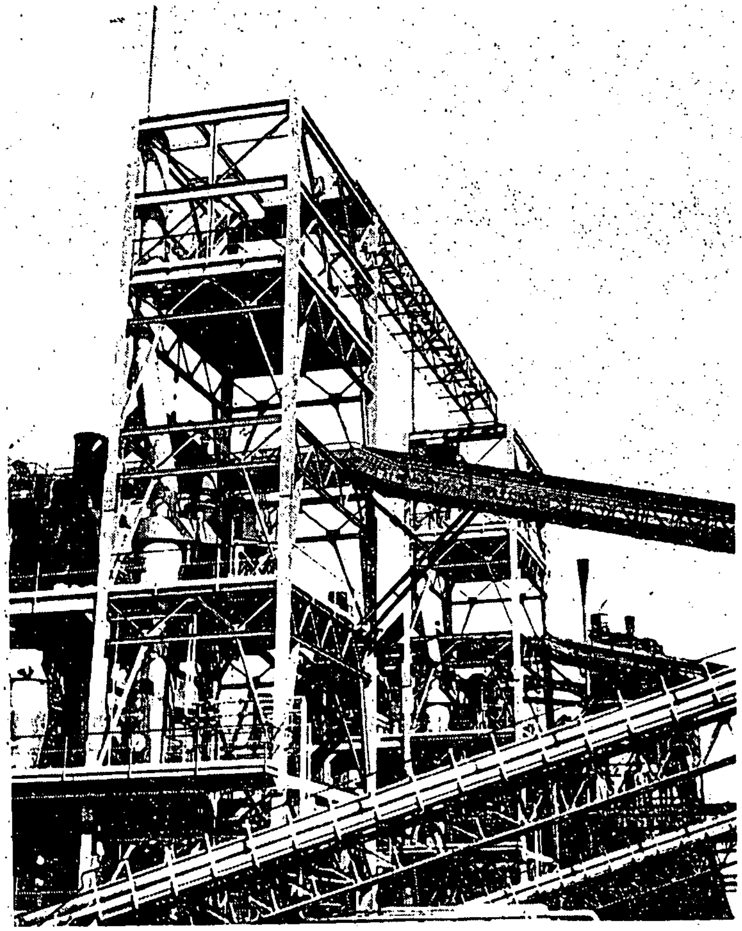


Fig. 3—A view of the magnetizing reduction plant.

Table III—Continuous Single-Stage Tests in a 4-in. Diam FluoSolids Reactor: Reduction with Separately Generated Reducing Gases.

	Hematitic Pyrite Calcine	Natural Hematite
Stages of Reduction	One	One
Size	−4 mesh	−10 mesh
% total Fe in feed	68.14	42.4
Reduction temp, °C	650	650
Heat requirement (high heat) accounting for gas generator efficiency (Million Btu/ton feed)	0.90	0.95
% Excess reducing gas over that required for reduction only	60	80
Inlet gas composition		
% CO	16.4	18.0
% H_2	17.2	14.0
% CO_2	Tr	5.0
Bed product:		
Detention time (minutes)	25	27
% by wt.	88.2	71.0
% reduction	97.7	95.0
Davis Tube results:		
% total Fe in concentrate	68.0	64.8
% Fe recovery	99.8	97.0
Cyclone product:		
% by wt	11.8	29.0
% reduction	98.6	97.4
Davis Tube results:		
% total Fe in concentrate	68.5	65.5
% Fe recovery	96.9	96.8

Tests With DFI

This method of reducing hematite to magnetite is based on the direct injection of oil into a bed fluidized by an amount of air less than that theoretically required to burn the oil completely. The gases from this partial combustion (N_2, CO, H_2, CO_2, H_2O, and hydrocarbons) create a sufficiently reducing atmosphere to convert the hematite in the feed to magnetite.

The 4-in. pilot plant reactor was equipped with a direct fuel injection device, and reduction tests on pyrite calcines and on natural hematites were made. Both single-stage and two-stage reduction tests were made on the natural hematites. Some typical results are presented in Table IV. The conclusions of the tests can be summarized as follows:

1.—On pyrite calcines, satisfactory reduction values were obtained with DFI, but somewhat higher fuel requirements were indicated for DFI than for gaseous reductants.

2.—On natural hematites, satisfactory reduction was obtained in a single-stage DFI operation only by use of a relatively high temperature (900°C) and higher fuel input (1.23 million Btu per ton ore).

3.—By the use of a two-stage DFI reduction on natural hematites, satisfactory reduction results were obtained at a temperature (650°C) and fuel input approaching those required when using gaseous reductants. Furthermore, the milder reduction temperature not only minimizes oil requirements but also prevents the formation of high-temperature iron compounds which are not amenable to subsequent magnetic concentration.

The overall results were extremely encouraging, in view of the great simplification which could be anticipated for commercial operation by elimination of gas generators. However, it was felt that problems might arise in connection with scale-up to commercial size, and the possibility of undesirable effects on the performance of the magnetic separation. It was decided for these reasons that the industrial reactors for the Montecatini plant at Follonica should be designed for operation with either gaseous reductants or liquid fuel by the DFI method.

Table IV—Continuous Tests in a 4-in. Diam FluoSolids Reactor: Reduction by DFI

	Hematitic Pyrite Calcine	Natural Hematite A	Natural Hematite B
Stages of reduction	One	One	Two
Size	−4 Mesh	−10 Mesh	
% total Fe in feed	47.8	41.2	39.0
Reduction Temp, °C	675	900	650
Heat requirement (high heat), million Btu/ton feed	1.07	1.23	1.09
Bed product:			
Detention time, (min)	25	40	66
% by wt	75.3	76.0	62.5
% reduction	99.8	100.0	98.2
Davis Tube results			
% total Fe in concentrate	68.3	69.6	70.2
% Fe recovery	96.3	97.5	98.0
Cyclone product			
% by wt	24.7	24.0	29.8
% reduction	100.7	>100.0	95.0
Davis Tube results			
% total Fe in concentrate	68.9	72.6	69.9
% Fe recovery	97.8	92.4	95.9

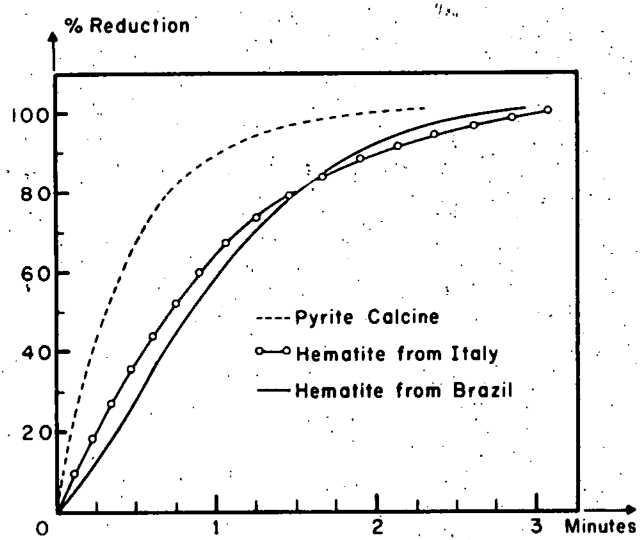


Fig. 4—Reduction % vs time of hematite at 650°C.

Concurrent with reduction testing, other extensive laboratory and pilot work was carried out on the subsequent magnetic separation and pelletizing steps of the Montecatini process. The results were most satisfactory and led to the decision to proceed with the design and construction of the industrial plants.

The Industrial Installation

Roasting of crude pyrites and reduction of their calcines are carried out in four separate lines, each of which has a roasting and a reduction unit.

Each roasting line consists of a FluoSolids reactor with an inside diameter of about 20 ft, a LaMont type waste-heat boiler, one stage of cyclones, and a gas scrubber. Approximately 600 metric tpd of 35-39% S-pyrite, crushed to ¼ in., are metered by an electronic weighing belt feeder and delivered to the reactor. There the ore is extensively desulfurized in a dead roast, yielding a gas with 12% SO_2. The gas is discharged through the waste heat boiler, which yields 750 metric ons of steam at 590 psig and 450°C. The calcines, amounting to 440 metric tpd line, are obtained as three different products: coarse fraction at 950°C (1742°F) from the fluid bed; fine-sized fraction at 300°C (572°F) from the boiler and cyclones; and ultrafine fraction, collected wet in the gas scrubber.

This latter fraction, thanks to the selective liberation of the minerals during roasting, has an Fe content of about 66% and therefore does not require further up-grading. It will be recovered as a sludge from a thickener and sent directly to the pelletizing plant.

In order to minimize heat losses, calcines from the roaster, boiler and cyclones are immediately transferred to the adjacent reduction unit. In Table V, chemical analyses and particle size distribution of pyrite and of corresponding calcines are shown.

Two reduction lines were put on stream early in 1964, one using gaseous reductant, and the other using DFI. For several months, comparative data were collected on the performance of the two systems. At the end of this period, it was concluded that the DFI system is by far superior, and so it was adopted for all four reducing lines.

It is of interest at this point to summarize the characteristics of the two reduction systems and to evaluate their respective performance data.

Reduction at Follonica is basically a three-stage operation: preheating, reduction proper, and cooling of the calcines. In the case where reducing gas produced in a separate oil cracking unit is utilized, the flowsheet shown in Fig. 6 was adopted. It incorporates a three-compartment FluoSolids reactor where solids flow countercurrently to the gases.

Hot calcines from the roasting plant are fed to the preheating compartment, where their temperature is further raised by the hot gases from the reducing section and by the heat released in the combustion of

Table V—Industrial Operations at the Follonica Plant

Particle size distribution: cumulative % by wt

	Wt Distribution[a] %	Mesh (Tyler) +4	8	14	35	65	100	200	325	−325
Pyrite	127.4	10.1	28.0	47.2	73.0	80.1	90.7	96.2	97.8	2.2
Calcines from bed	52.4	5.0	21.2	43.5	77.2	97.2	99.3	—	—	—
Calcines from boiler and cyclones	43.3	—	—	—	—	6.1	23.3	55.6	71.0	29.0
Calcines from scrubber	4.3	—	—	—	—	—	0.4	0.7	0.9	99.1

Chemical analyses: % by wt

	Fe Total	Fe Ferrous	S Total	S Sulfate	S Sulfide	CaO	MgO	SiO_2	Al_2O_3	CO_2
Pyrite	38.53	—	38.26	0.21	38.05	5.97	0.96	10.29	1.19	3.04
Calcines from bed	40.59	1.50	3.34	3.01	0.33	9.71	1.89	19.91	2.76	0.36
Calcines from boiler and cyclones	57.70	1.53	2.34	1.99	0.35	5.59	0.50	6.00	0.18	0.41
Calcines from scrubber	66.12	0.22	0.08	0.01	0.07	2.43	0.11	2.56	0.23	Tr

[a] 100 tons of pyrites yield 78.5 tons of hematitic calcines.

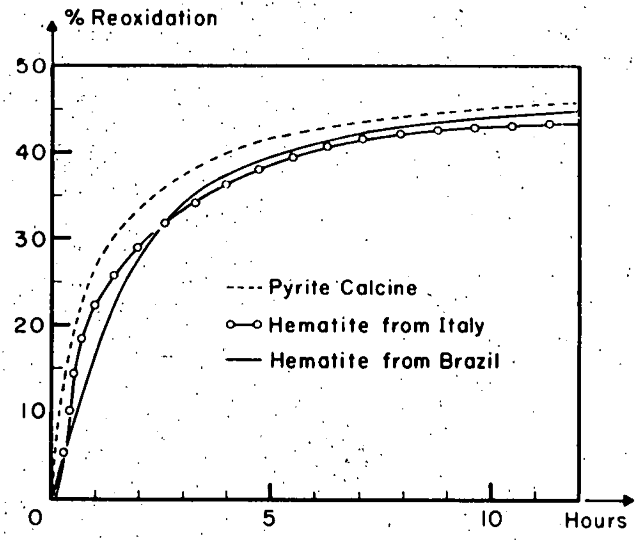


Fig. 5—Reoxidation % vs time of artificial magnetites.

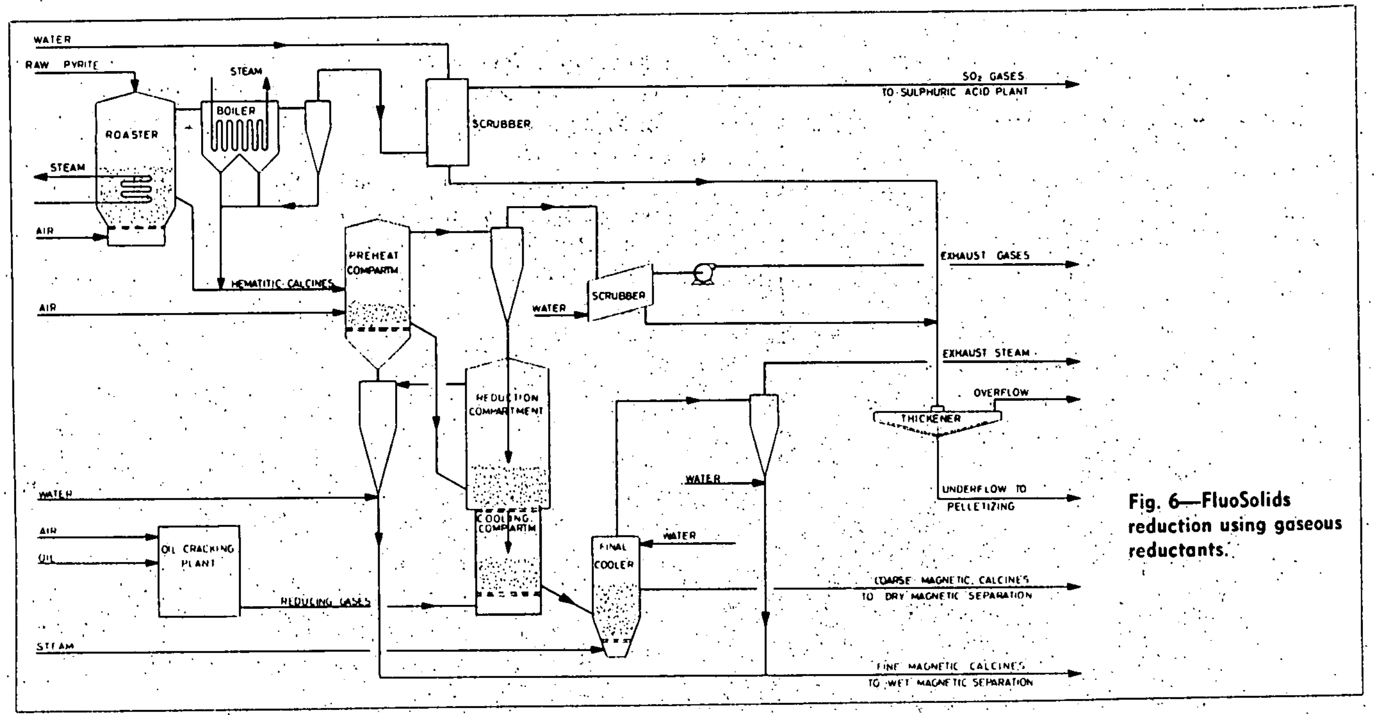


Fig. 6—FluoSolids reduction using gaseous reductants.

the residual reducing gases, with a controlled amount of injected air.

The preheated hematitic calcines are transferred to the 5-ft diameter reduction compartment, where they are fluidized by the reducing gases at proper temperature and held there long enough for reduction to take place.

For cooling, the reduced calcines are discharged into a bed fluidized with cold reducing gas, which is, in turn, preheated. Reduced calcines are further cooled below reoxidation temperature in an external bed fluidized with steam and cooled by a water spray. The cooled bed product is sent to dry magnetic separation. The reduced fines, carried over by the gases, are collected in a cyclone, quenched in water, and sent to wet magnetic separation. The reducing gas, produced in a separate cracking plant, is fed under pressure into the bottom of the FluoSolids reactor.

Operating data for the gaseous reduction system are reported in Table VI, together with chemical analyses, particle-size distribution of the reduced calcines, and Davis tube-test results.

As illustrated in Fig. 7, the use of DFI technique permits substantial simplifications in the system. Heat balance is such that calcines need not be preheated above the temperature at which they are delivered from the roasting plant. Instead, they are fed directly into the reduction reactor. Fluidization is maintained by the reducing gas produced by partial combustion of oil with air. Oil is injected directly into the fluid bed through a number of guns, while air is admitted from under the constriction plate through a number of tuyeres. Bed temperature, detention time of solids, air-to-oil, and oil-to-feed ratios are the governing factors of the process. The hot-reduced calcines are discharged into the steam-fluidized bed of a separate reactor and cooled below

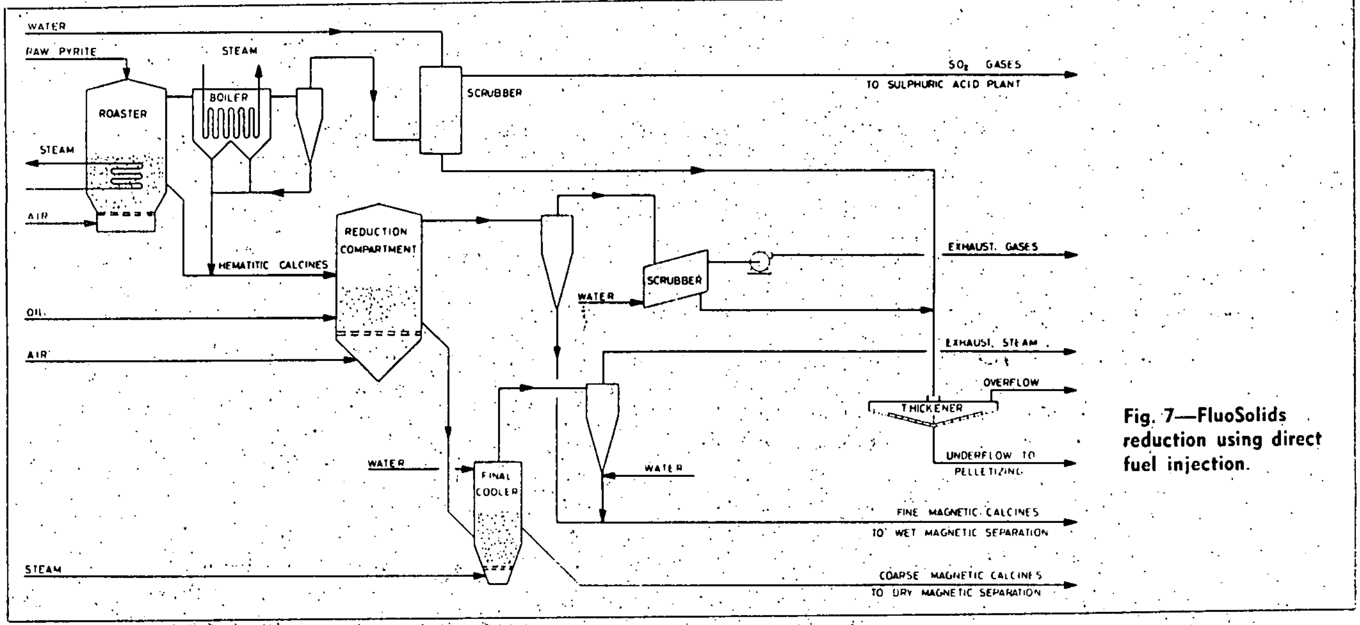


Fig. 7—FluoSolids reduction using direct fuel injection.

Table VI—Industrial Plant at Follonica: Reduction with Separately Generated Reducing Bases

Operating conditions	
Feed rate, metric tpd	400
Total Fe in feed, %	48.4
Reducing gas composition	
CO, % by vol	19.0
H_2, % by vol	12.1
CO_2, % by vol	4.2
N_2, % by vol	64.7
Avg temp of feed, °C	530
Preheat compartment temp, °C	680
Reduction compartment temp, °C	640
Cooling compartment temp, °C	330
Final cooler temp, °C	190
Heat requirement, accounting for gas generator efficiency, million Btu/ton feed	1.0
Power requirement, including gas generator, kwhr/ton feed	13.5
Man-hour requirement, including gas generator, man hr per ton feed	0.2

Particle size distribution, cumulative % by wt

	Wt Distribution %	Mesh (Tyler)								
		+4	8	14	35	65	100	200	325	−325
Calcines from bed	93.7	2.9	12.4	25.4	45.1	59.7	69.3	85.2	92.2	7.8
Calcines from cyclone	4.7	—	—	—	—	—	0.3	0.9	8.3	91.7
Calcines from scrubber	1.6	—	—	—	—	—	0.1	0.6	5.0	95.0

Chemical analyses and Davis Tube Tester results, % by wt

	Calcines					Davis Tube Concentrate	
	Fe Total	Reduction %	S Total	S Sulfate	S Sulfide	Fe Total	Fe Recovery
Calcines from bed	48.53	99.4	3.12	2.75	0.37	68.53	95.80
Calcines from cyclone	61.17	100.0	0.68	0.53	0.15	69.48	98.41
Calcines from scrubber	66.83	—	0.62	0.61	0.01	—	—

Table VII—Industrial Plant at Follonica: Reduction with DFI Technique

Operating conditions	
Feed rate, metric tpd	400
Total Fe in feed, %	48.4
Avg temp of feed, °C	530
Reduction compartment temp, °C	670
Final cooler temp, °C	180
Heat requirement, million Btu/ton feed	1.05
Power requirement, kw-hr/ton feed	4
Man-hr requirement, man-hr/ton feed	0.1

Particle size distribution, cumulative % by wt

	Wt Distribution %	Mesh (Tyler)								
		+4	8	14	35	65	100	200	325	−325
Calcines from bed	77.0	3.2	14.0	22.5	38.3	54.0	65.4	91.5	97.1	2.9
Calcines from cyclone	21.3	—	—	—	—	—	0.9	20.5	66.2	33.8
Calcines from scrubber	1.7	—	—	—	—	—	0.2	1.2	11.1	88.9

Chemical analyses and Davis Tube results, % by wt

	Calcines					Davis Tube Concentrate	
	Fe Total	Reduction %	S Total	S Sulfate	S Sulfide	Fe Total	Fe Recovery
Calcines from bed	45.35	98.6	3.51	3.11	0.40	66.92	95.63
Calcines from cyclone	62.81	100.2	1.22	1.03	0.19	69.88	98.53
Calcines from scrubber	65.41	75.1	0.53	0.51	0.02	—	—

the reoxidation temperature by a water spray. Fines are collected in a cyclone and quenched in water.

In Table VII, DFI operating data are given along with chemical and particle-size analyses.

The performance of the industrial plant has shown good agreement with the result of the pilot tests, and has proved that a satisfactorily reduced material can be produced, suitable for subsequent magnetic separation. In the commercial DFI reduction reactor, the heat requirements amount to about 1.1 million Btu per ton. Commercial operation has shown that the DFI system has these inherent advantages over the use of gaseous reductants:

1.—Simplified, less expensive installation and reduced operating expenses. Reductant gas generators become unnecessary and are replaced by a conventional oil pumping station. Need for production of inert gas for purging the system on start-ups and shut-downs disappears. Use of Bunker C oil presents no problems.

2.—Simpler, easier and safer operation and increased flexibility.

3.—Substantial savings in the overall electric energy consumption and manpower requirements.

The use of fluidized beds for reduction has specific advantages. Intensive gas-solid contact at high mass-flow and at high heat-transfer rates, typical of fluidization, produces particularly favorable kinetic conditions. The reduction reaction is easily controlled by changing calcine detention time, bed temperature, and strength of reducing atmosphere.

The reduction chamber of the DFI reactor is designed and operated to produce a 150 mesh classification of the products. This is optimum for subsequent dry magnetic separation of the coarse bed product and wet magnetic separation of the fine cyclone product. (See Fig. 1.)

The fine-sized particles, which represent 20-25% of the total reduced product, are separated from the reactor stack gas by a hot cyclone, quenched in water, and pumped as a slurry at about 25% solids to the wet magnetic separation without any additional screening and grinding.

The bed and cyclone products are both very close to 100% magnetite. In the case of the carried-over particles, this is made possible, in spite of the very short detention time, by the great reactivity of the finer particles. This is related, in turn, to their high specific surface area.

The calcines from the bed of the reduction reactor, having lost the finer fraction, are essentially dust-free, and they can therefore be handled in dry magnetic separators with most satisfactory results.

It has been found that a certain amount of carbon, corresponding to 4-5% of the heat value of the fuel used, is deposited on the magnetic calcines. About 60% of this carbon is recovered in the concentrates from the magnetic separations, and is available as a source of heat in the pelletizing process.

The steps which follow DFI reduction are magnetic separation and pelletizing of the concentrate. They are not described at length in this article, since data has been provided in another US publication.[10]

Nevertheless, a schematic flowsheet of these operations is shown in Fig. 8.

The magnetic separation plant incorporates two sections, one for the dry up-grading of the coarser fraction (by means of Mortsell-Sala separators), the other for the wet separation of the fine cyclone product and of the ground middlings from the dry separation (by low-intensity permanent magnet Sala drums). The dry and wet concentrates are mixed after dewatering of the wet product, resulting in a final concentrate with about 12% moisture. Since green balling requires a moisture content of about 14%, the additional 2% is added directly to the balling drum of the pelletizing plant.

The pelletizing plant is of the straight grate type based on the McKee design. The effective grate area is 540 ft^2, and the specific capacity is on the order of two tons of fired pellets per ft^2 per day. Pellets are first dried, then preheated and fired to a maximum temperature of 1320°C (2408°F), and finally cooled to 90°C (194°F). At this temperature, they are discharged from the grate and conveyed to storage. Total fuel oil consumption in pelletizing is in the order of 0.73 million Btu per ton of pellets. Energy consumption of magnetic separation and pelletizing plants is about 50 kw-hr per ton.

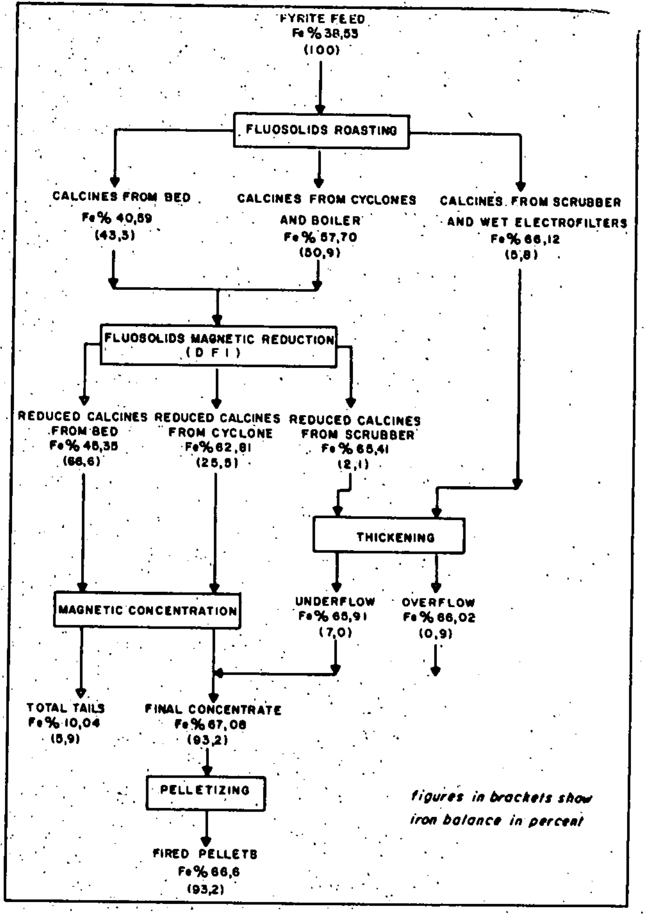
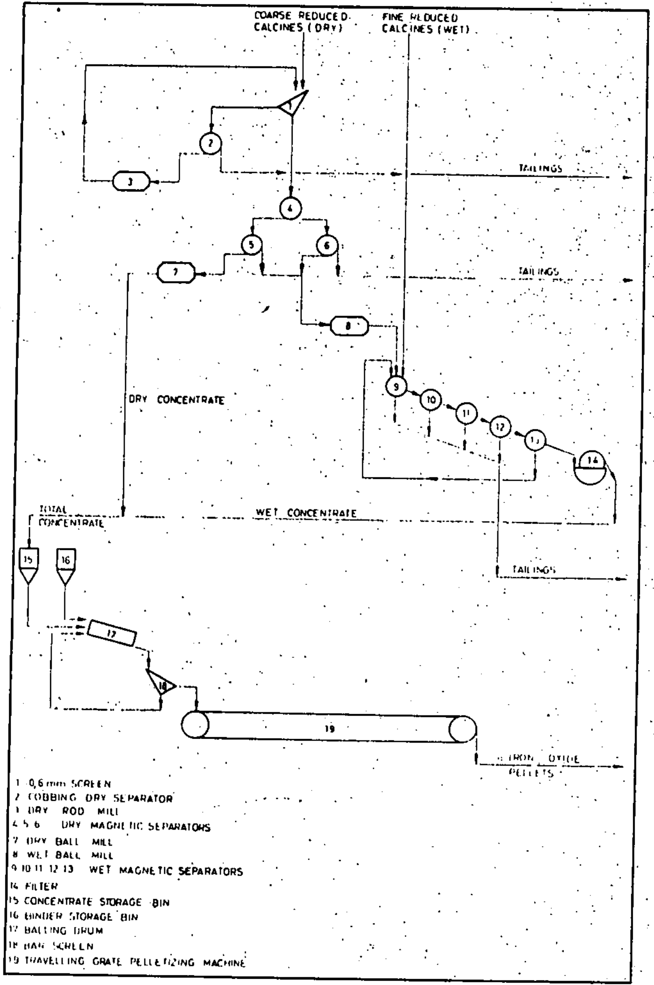


Fig. 9—Right—Iron balance in the Montecatini process.

Fig. 8—Below—Flow sheet of magnetic concentration and pelletizing units.



A comparison of the fuel requirement for reducing and pelletizing pyrite calcines by the Montecatini process with that for pelletizing a hematitic ore is revealing. The preparation of 1 metric ton of pellets from pyrite calcines at Follonica requires 2.18 million Btu, of which 1.43 million Btu are used for the reduction (equivalent to 1.05 million Btu per ton of pyrite calcines), and the remaining 0.75 are used for pellet firing. Under similar conditions, the estimated heat requirement for a natural hematitic ore or concentrate is about 1.25 million Btu per metric ton of pellets. Thus, from an overall process standpoint, the net heat cost of the reduction step is only 0.93 million Btu per metric ton of pellets (or 0.68 million Btu per metric ton of pyrite calcine). With magnetite, moreover, control of the pellet firing operation is much easier because of a more uniform temperature profile through the pellets.

Analytical data of magnetic concentrate and of pellets are presented in Table VIII.

The overall iron balance of the Montecatini process at the Follonica works is shown in Fig. 9. Total Fe recovery is over 93%. The fired pellets average 66% Fe and 0.01% S, with only minor traces of nonferrous metal impurities.

The operations at Follonica are closely controlled by means of instrumental analytical facilities of the same scientific standard as those employed for research work. For instance, the determination of iron in pyrite and calcines in the various phases of the process is made by a Philips X-ray fluorescence spectrometer. Non-ferrous elements are determined spectrographically by a direct reading Jarrell-Ash equipment, Atom-counter model. Sulfur is determined by combustion technique, using a Dietert apparatus. Equipment for differential thermal analysis, thermogravimetric analysis, gas chromatography and polarography is also available in the plant laboratory. Current research in the Montecatini laboratories aims at the further development of this process, in view of its applicability to other pyrite ores, containing greater amounts of non-ferrous metals such as copper, zinc, lead, and arsenic.

CONCLUSIONS

Full scale operations at Follonica have shown that the beneficiation of pyrite calcines by FluoSolids magnetizing reduction is technically sound and attractive from an economic point of view. Application of the DFI technique has further contributed to improve these favorable results by permitting greater ease and safety of operation as well as important savings in the operating costs. Compared to the use of separately generated reducing gases, the DFI technique resulted in a decrease (per metric ton of calcine) of power consumption from 13.5 to 4 kw-hr, and of man-hours from 0.2 to 0.1, while heat requirements were 1.05 instead of 1 million Btu. Elimination of the reducing gas generator and of inert gas for purging purposes along with simpler reactor construction means a substantial reduction of first cost, on the order of 50%.

These results show that the DFI process is superior to other reduction techniques employing gaseous reductants, and also competitive with such conventional methods for up-grading natural ores as concentration by gravity and flotation. It is possible to anticipate treatment of many low-grade hematite ores such as the taconites of the North Central states, the Canadian specular hematites, and the European carbonate ores. By-product fractions obtained from the concentration of non-magnetic ores, such as semi-taconites, are also amenable to concentration by magnetizing reduction followed by magnetic concentration.

Based on the results of extensive pilot test work on a great number of iron ores, and on the commercial operations at Follonica, it is anticipated that the reduction of 1100 net tpd of natural hematite with 40% Fe would require a FluoSolids reactor incorporating three fluidized beds in series, of which two are for a two-stage reduction step and one is for the preheating of the cold ore by means of the hot outgoing gases. The unit, shown schematically in Fig. 10, has an overall height of about 80 ft and a maximum outside diameter of about 12 ft. The operating requirements, per metric ton of ore, are estimated at about 1 million Btu, 0.07 man-hours, and 3.5 kw-hr.

Large-capacity units will justify the recovery of the sensible heat of the reduced ore and of the stack gases, thereby further improving the thermal efficiency of the system.

The cooperative efforts of Montecatini and Dorr-Oliver continue towards further improvements of the FluoSolids magnetizing reduction technology for new and wider applications.

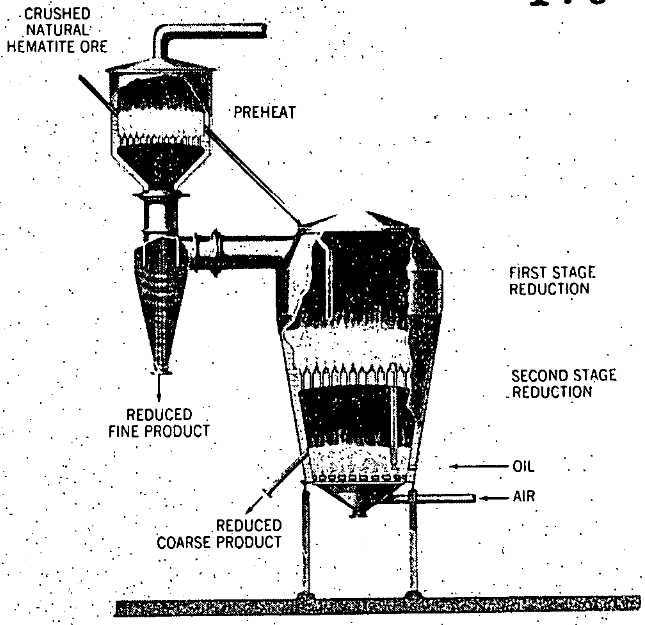


Fig. 10—Hematite magnetic reduction with direct fuel injection.

Table VIII—Final Concentrates and Pellets Produced at the Follonica Plant

Particle size distribution: cumulative % by wt

Concentrates		Pellets[a]	
Mesh	% cum.	mm	% cum
+100	2.5	+15	0.5
150	5.3	12	12.4
200	10.2	8	94.0
325	31.2	3	98.5
−325	68.8	−3	1.5

Chemical analyses: % by wt

	Concentrate	Pellets
Fe total	67.08	66.60
S total	1.10	0.015
S sulfate	0.88	0.015
S sulfide	0.22	—
CaO	2.06	2.04
MgO	0.35	0.34
SiO_2	2.09	2.07
Al_2O_3	0.42	0.41
Cu	0.021	0.020
Zn	0.025	0.021
Pb	0.016	0.013
As	0.010	0.010
P	0.001	0.001

[a] Compressive strength: 150 kg
Resistance to abrasion (Tumble test): 7%—28 mesh
True density (kg/dm^3) 4.7
Porosity 23%

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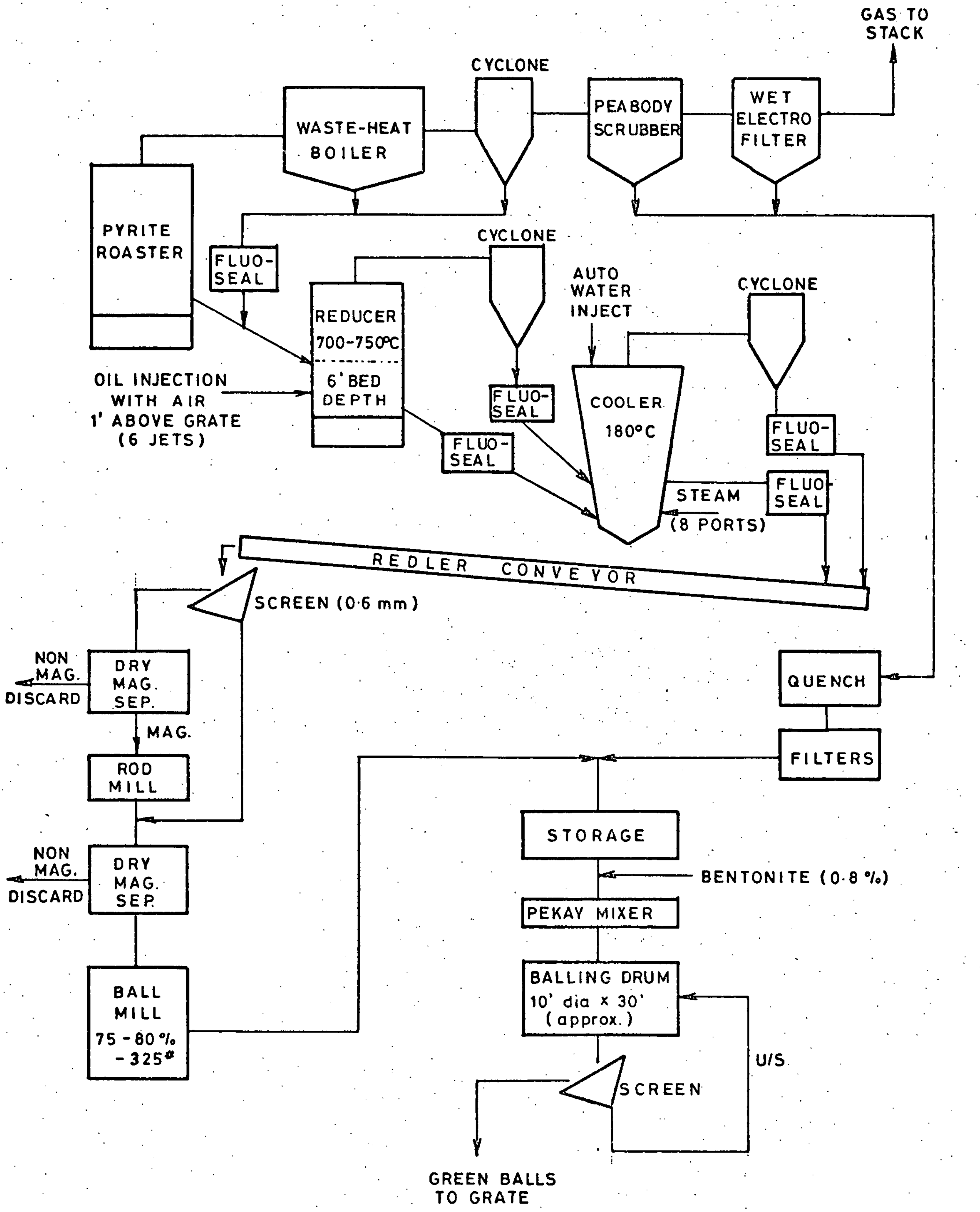


FIGURE A 19.1: DIAGRAMMATIC FLOWSHEET OF CALCINE UPGRADING AND PELLETISING AT FOLLONICA

REMOVAL OF NON FERROUS METALS, ARSENIC AND SULPHUR FROM PYRITE CALCINES (MONTECATINI-EDISON PROCESS)

1. The Process

The process is based on a high temperature selective chlorination of the non ferrous metals (Cu, Zn, Pb, Ag, Au, Co, Ni) present in the pyrite calcines. The chlorides thus formed are given off as vapours and recovered by subsequent scrubbing with water.
During the chlorination, also arsenic and sulphur are extensively removed. The purified calcines represent now an iron ore, suitable for the iron and steel industry. The non ferrous metals values are recovered from the chloride solution by hydrometallurgical extraction. In practice, the process includes the following steps:

1.1 Pre-heating and magnetizing reduction of the calcines in a fluidized bed reactor by the D.F.I., system (Direct Fuel Injection).
This step can be avoided if the calcines are available at high temperature, as for instance when they are directly withdrawn from the pyrite roaster operated under specific conditions.

1.2 Fluidized bed chlorination of the calcines coming from step 1, using gaseous chlorine (or HCl). The reaction temperature is held making use of the sensible and chemical heat available in the solids under treatment.
In the course of this step the non ferrous metals are volatilized as chlorides. Also sulphur and arsenic are conveyed in the gaseous phase.

1.3 Cooling of the purified calcines. The convenience of recovering the sensible heat of the solids depends upon the economics of each individual case.

1.4 Recovery of the metal values. The chloride vapours are cooled and scrubbed with water. From the solution thus obtained copper is recovered as high grade, arsenic free cement (>90% Cu). Moreover, the copper cement contains most of the silver and gold initially present in the calcines. Lead can be recovered as cement (>90% Pb); zinc as oxide. The opportunity of recovering this latter metal is dependent upon economic considerations. The arsenic removed from the calcines is present in the chloride solution. Eventually it is disposed from the system.

2. The Pilot Plant

After extensive research work carried out at the "G. Donegani" Research Institute, Novara, a 50 mtpd pilot plant is operating at the Montecatini-Edison works at Scarlino-Follonica (Fig. 1). The pilot plant is flexible enough to be able to treat cold stockpiled calcines from different sources (Spain, Portugal, USSR, etc.). A simplified flow-sheet of the plant is shown in Fig. 2, while Fig. 3 shows the pilot plant itself.

The research work is being carried out in cooperation with Dorr Oliver Inc. (U.S.A.), who have engineered the fluidized bed reactors.

From the test work carried out so far, it appears that by using not more that 1,1 - 1,3 times the stoichiometric amount of chlorine necessary for the chlorination of the non ferrous metals, the amount of impurities left in processed calcines is reduced to satisfactory values, such as : % Cu 0,02 - 0,04; % Zn 0,03 - 0,06; % Pb 0,03 - 0,06; % As 0,03 - 0,04; % S 0,2 - 0,02, depending on the nature and composition of the feed. These low -sulphur calcines can be used as feed to a cold pelletizing process cheaper than high temperature ones.

3. Economics

An economic evaluation of the process had been made based on the data ob tained in the laboratory. The operations of the pilot plant have so far confirmed that the initial assumptions, based on pre-pilot data, were realistic and sound. Further testing on different calcines is being carried out and it is expected that sets of operating data under widely diversified conditions shall be available in the near future. These data will enable Montecatini-Edison to asses the feasibility of the process for different types of feed and under varying local economic conditions.

It is worth mentioning right now that, in comparison with other processes, the present one is highly flexible, technically effective and economically attractive also with highly contaminated calcines.

January 1970

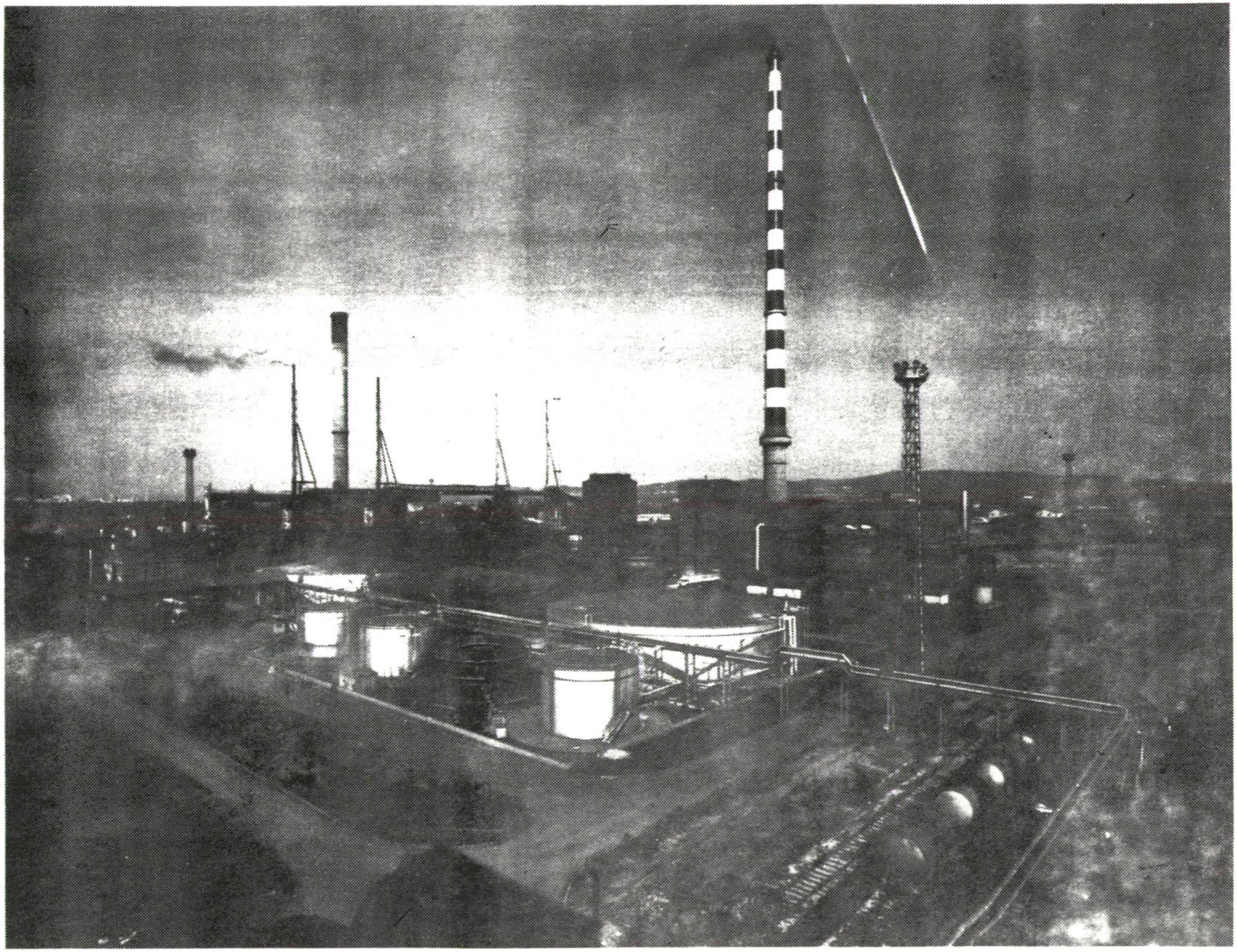
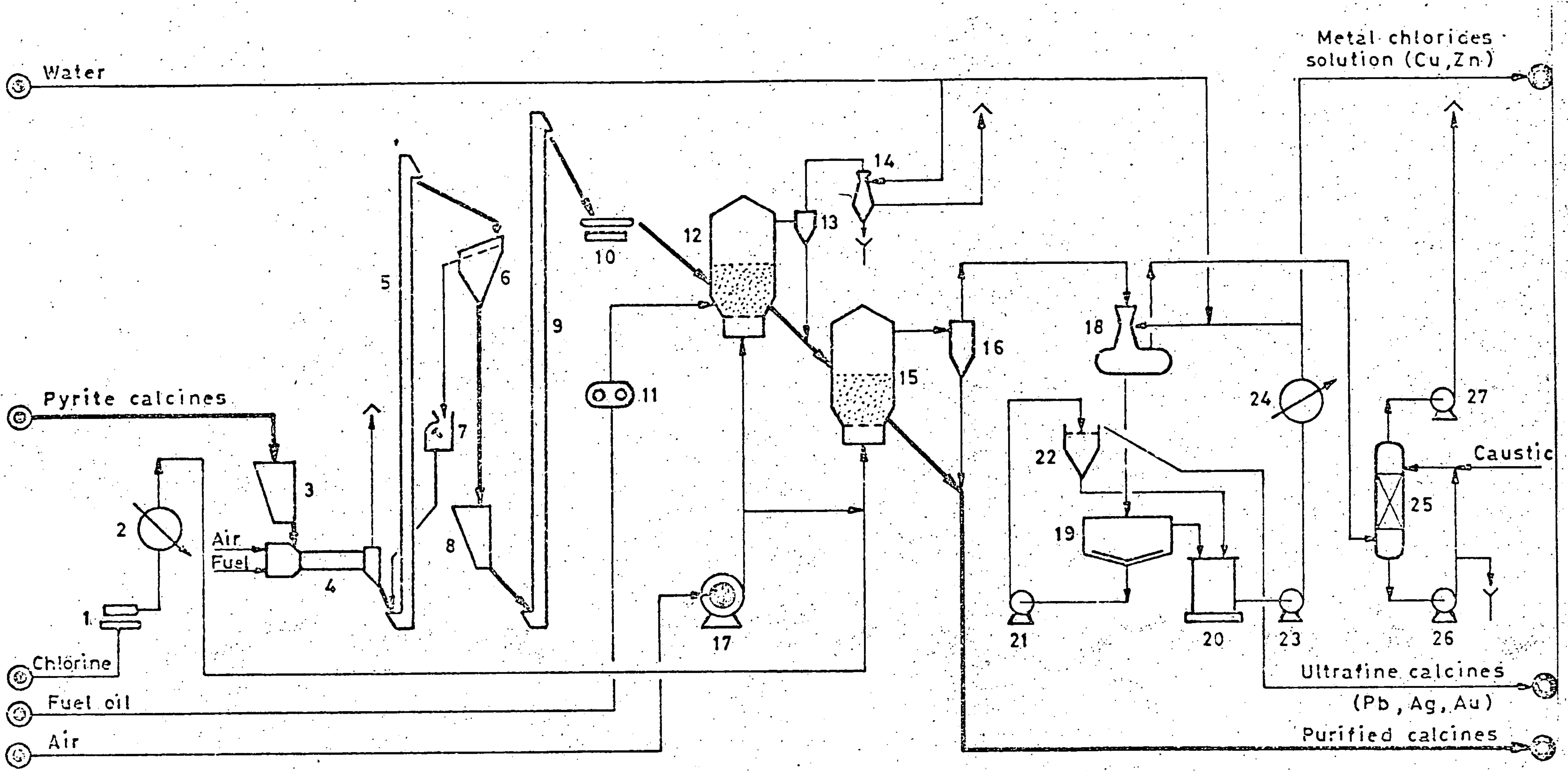


FIG. 1 - GENERAL VIEW OF THE SCARLINO-FOLLONICA WORKS WHERE THE PILOT PLANT IS LOCATED



1 - Chlorine scale
2 - Chlorine vaporizer
3 - Raw calcines bin
4 - Rotary drier
5 - Bucket elevator
6 - Vibrating screen
7 - Hammer mill
8 - Dry calcines bin
9 - Bucket elevator
10 - Feed conveyor scale
11 - Metering pump
12 - Prereducer
13 - Prereducer cyclone
14 - Waste gas scrubber
15 - Chlorinator
16 - Chlorinator cyclone
17 - Air compressor
18 - Chlorination gas scrubber
19 - Thickener
20 - Solution tank
21 - Underflow pump
22 - Vacuum filter
23 - Solution pump
24 - Solution cooler
25 - Neutralization column
26 - Caustic pump
27 - Exhaust fan

FIG. 2 - PYRITES CALCINES PURIFICATION PILOT PLANT - FLOW SHEET

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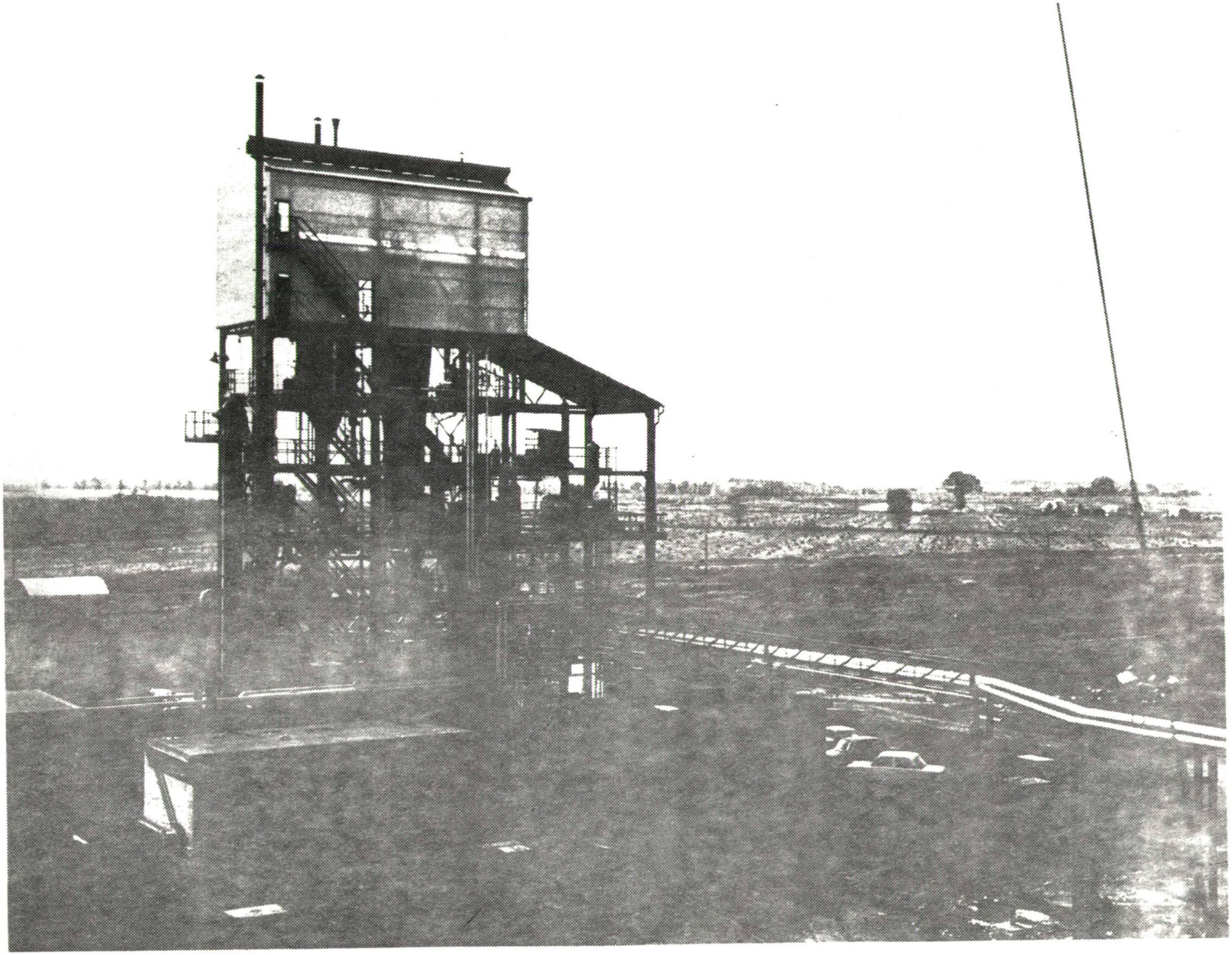

FIG. 3 - VIEW OF THE 50 MTPD PILOT PLANT AT THE SCARLINO-FOLLONICA WORKS

APPENDIX B

IRON ORE RESOURCES IN THE OLARY PROVINCE

(From a letter from the Director of Mines)

I acknowledge your letter of 9 October 1970 referring to sources of iron ore for a mini-steel works.

The Olary Province was mapped in the early 1950's and is being remapped at present. While current mapping may define more closely the stratigraphic units and the geological structures, past prospecting has been intense and it is unlikely that any new sizeable iron ore outcrops will be discovered during this mapping programme.

Of the three iron ore types found in the Olary Province, those associated with Lower Proterozoic (previously called Archaean) rocks are considered most likely to be economic. These are described in Summary Report RB.54/86:

Deposit	One Mile Sheet	Reserves Tons Per 100' Depth	Grade % Fe	Miles From Olary R.S.
Koolka	Kalabity	350,000	66.3	37
Plumbago	Plumbago	185,000	66.6	35
Old Toraminga	Kalabity	200,000?	64.6	28
Alconie	Kalabity	small	67.3	25
Billeroo	Glenorchy	100,000	65.5	40
Billeroo	Glenorchy	1.5 million	48	40
Total i. High grade ore		800,000?	65-66	25-40
ii. Concentrates from Billeroo		1 million?	60+	40

This information is based on detailed 100 scale mapping and limited surface sampling. For all deposits the P_2O_3 content may be a concern. Dry magnetic separation has produced a 60.6% Fe concentrate from Billeroo low grade with 88.7% recovery. Humphrey spirals have produced a 61.8% Fe concentrate with 85.9% recovery. Summary Report RB.54/86 is available at cost of copying as is also Report RB.54/96 on "Beneficiation Tests on Billeroo Iron Ore" (by L. Bollen).

APPENDIX C

IRON ORE DEPOSITS, OLARY PROVINCE

1. HIGH-GRADE ORE

1.1 Koolka

1.1.1 North Outcrop

Assume: (a) 4:1 Barren rock:ore ratio
(b) mining to 200 ft below plain level
whence (c) ore reserve 550,000 tons

1.1.2 Main South Outcrop

Assume: (a) 4:1 Barren rock:ore ratio
(b) mining to 140 ft below plain level
whence (c) ore reserve 95,000 tons

1.2 Billeroo Dam

Assume: (a) 4:1 Barren rock:ore ratio
(b) mining to 140 ft below plain level
whence (c) ore reserve 200,000 tons

1.3 Plumbago

Outcrop is of similar size to Koolka (North). Assume same data.

1.4 Old Toraminga Dam

Assuming continuity between outcrops, deposit would be of similar area to Koolka (North) but of negligible height.

Assume: (a) 4:1 Barren rock:ore ratio
(b) mining to 160 ft below plain level
whence (c) ore reserve 400,000 tons

2. LOWER-GRADE ORE

2.1 Billeroo Peak

Assume: (a) 1:1 Barren rock:ore ratio
(b) mining to 160 ft below plain level
whence (c) ore reserve 1.7 million tons

2.2 Billeroo Centre

Assume: (a) 1:1 Barren rock:ore ratio
(b) mining to 140 ft below plain level
whence (c) ore reserve 1.3 million tons

APPENDIX D

DATA PROVIDED BY WOODALL-DUCKHAM LTD

DUTY

Plant to produce 5,000 tons per annum of iron powder from scrap, working two shifts per day for 250 days per year.

PROCESS DESCRIPTION

Melting

The scrap metal is transferred from storage and melted in an electric induction furnace. The molten metal is transferred to a ladle for pouring into the atomisation unit.

Atomisation

The molten metal is poured continuously from the ladle into a heated tundish situated over the atomising chamber. The metal stream pouring into the chamber is subjected to a controlled high pressure water blast which converts it into powder particles that are cooled very rapidly. These collect in a large tank of water.

Drying

When the whole of the furnace charge has been atomised, the powder and water mixture is partially dewatered, and the slurry which is left is transferred to a drier where the remaining water is removed.

Annealing

The dry powder is transferred to trays and passed through a conveyor furnace in a reducing atmosphere at a temperature suitable for softening the particles. The annealing time is pre-set to allow for the removal of impurities.

After annealing, the lightly sintered powder cakes are broken up in a slow-speed hammer mill so that the particles are restored to their as atomised size.

The soft powder passes through a suitable screen, and the small percentage of oversize is fed back to the melting furnace.

Finishing and Packing

The screened powder is now ready for blending or storage.

For in-plant use, large storage containers or drums may be used. A bagging machine can be incorporated if desired.

Suitable inert atmosphere conditions are maintained in the finishing section of the plant.

COST ESTIMATES

Capital Cost (Budget Price)

Cost of plant erected on U.K. greenfield site = £500,000 (including all process plant, buildings and utilities).

Operating Costs

a. Raw Materials

1.1 tons low carbon steel scrap per ton of product. Packing bags.

b. Services - per ton of product basis

Power	800 KW
Hydrogen or Cracked Ammonia	200 cu ft
Towns Gas	6 therms
Water	200 gal
Sundries (incl. Nitrogen)	Equivalent to 25¢

c. Labour Requirements

Labour (2 shifts, 250 days per year), 20 operators

1 General Foreman

2 Maintenance Men

d. Indirect Costs

Insurance

Maintenance - 5% of capital cost

Depreciation

Interest on borrowings

APPENDIX E

AN APPRAISAL OF
EXPORT MARKETS FOR

A. IRON AND FERROUS-ALLOY POWDERS

B. COPPER AND COPPER-ALLOY POWDERS

for

THE AUSTRALIAN MINERAL DEVELOPMENT LABORATORIES

by

INTERNATIONAL TECHNICAL SERVICES LIMITED

17 February, 1971

INTRODUCTION

On December 30, 1970, The Australian Mineral Development Laboratories requested that International Technical Services Limited undertake an appraisal of export markets for two classes of metal powders:

a. Iron and ferrous-alloy powders
b. Copper and copper-alloy powders

It was suggested that in the appraisal Japan, Europe and North America be excluded for iron powder and North America for copper powder.

The limited budget and the short time allowed for data gathering mean this appraisal should be considered as a "best attempt" under the circumstances.

The appraisal is based on information and comment from Australian Government Trade Commissioners and powder metallurgy industry sources overseas. Unfortunately, the time factor together with the QANTAS Strike which delayed overseas mail has meant that replies containing information relating to potential markets in various parts of the world have not yet been received. This has resulted in gaps in the appraisal, particularly in South America.*

SUMMARY AND CONCLUSIONS

1. No potentially unsatisfied markets for iron, copper, ferrous-alloy or copper-alloy powders were found.
2. In the markets appraised it was notable that suppliers originated from several sources. Sweden, West Germany, U.K. and U.S.A. appear regularly as countries of origin in the import statistics and in several cases up to twelve countries are listed as suppliers. This indicates that an Australian exporter would face strong competition in almost all potential markets from these established exporting nations.
3. In certain markets, such as those in European Economic Community countries, Australian exporters would have to compete with suppliers enjoying preferential tariff treatment.
4. Many developing nations face severe foreign exchange difficulties which limit potential imports of metal powders.
5. The appraisal was based mainly on import statistics from countries considered potential export markets. The broad classification in all the statistics does not allow the identification of specialised markets such as those for a particular metal alloy powder. Further study of

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* Information received later is given in the Addendum (page E-33 *et seq.*)

types and usage of imports could reveal potentially profitable markets.

6. The applications and usage of powder metallurgy require a level or stage of industrialisation not yet reached by a majority of countries outside Europe, North America and Japan. Therefore it is suggested that any further study of export opportunities for iron and copper powders should place emphasis on opportunities for market development in addition to identifying existing markets.

1. SOUTH-EAST ASIA - IRON AND COPPER POWDERS

1.1 Hong Kong

1.1.1 General

Imports of iron and ferrous-alloy powders totalled 1,159.2 tons in 1968, 1,278.8 tons in 1969 and 593.6 tons for the first nine months of 1970. Main suppliers were Japan for iron and steel powders, Norway and Japan for manganese alloys and Canada and Mozambique for other ferro-alloys.

1.1.2 Import Statistics

Import statistics for 1968, 1969 and the first nine months of 1970 are contained in Appendix EA.

1.1.3 Tariffs and Import Duties

The Australian Government Trade Commissioner in Hong Kong has advised that there are no import duties or tariffs relating to copper, iron and ferrous-alloy powders.

1.1.4 Summary

The market is small involving approximately 1,200 tons annually. The annual amount spent by Hong Kong consumers is in the order of $120,000 which is shared between six or seven exporting countries.

1.2 India

1.2.1 General

Imports supply the bulk of the Indian market as local production is insignificant. Sintered parts manufacturers are the main consumers. In view of the continuing import restrictions and the emphasis on import substitution there is little doubt that dependence on imports will gradually be decreased by increasing local production.

1.2.2 Import Statistics - Iron Powder

	April 1969/March 1970		April 1969/July 1970	
	Quantity (Tons)	Value (Rupees)	Quantity (Tons)	Value (Rupees)
West Germany	4	10,553	14	15,324
Italy	1	2,189	-	-
Japan	2	7,515	3	11,102
Sweden	214	359,153	62	116,985
U.K.	1	3,175	2	2,801
U.S.A.	104	197,763	15	28,685
Total	326	580,348 ($A71,700)	96	174,897 ($A21,500)

1.2.3 Tariffs and Import Duties

Imports of iron powder are subject to import licensing.

Under the regulations for April 1970/March 1971, only actual users are eligible to obtain import licences for iron and copper powders for the manufacture of sintered parts.

1.2.4 Summary

The Indian market is small in volume and value. An Australian exporter would be faced with import licensing restrictions and with competition from other countries at present supplying the Indian powder-metal market.

1.3 Indonesia

1.3.1 General

Unfortunately Indonesian import statistics include imports of copper, iron and ferrous-alloy powders under the classification "chemical elements, not elsewhere specified". The total value of this item in 1969 was $45,421.

1.3.2 Import Statistics

	1968		1969	
Source of Imports	Gross Weight Kilogrammes	Value CIF US$	Gross Weight Kilogrammes	Value CIF US$
Japan	2,099	382	53,559	7,418
Pakistan	-	-	2	18
U.S.A.	-	-	212	2,980
U.K.	34	106	70	82
Netherlands	30,443	2,362	625	1,148
West Germany	-	-	5,799	33,168
Austria	79	641	1	171
Switzerland	-	-	2	436
Total	32,655	3,491	60,271	45,421

1.3.3 Tariffs and Import Duties

Iron powders are imported free of duty under Tariff Item 524. They are classified as essential goods under Category "B"

1.3.4 Summary

Metal powder used is limited in Indonesia. This situation will continue until significant industrial development takes place.

1.4 Singapore

1.4.1 General

Over the past three years there has been an increasing trend in usage of iron powders by industries in Singapore. However, the quantities imported are still insignificant.

1.4.2 Import Statistics

Description of Goods and Countries of Origin	1969		1968		1967	
	Qty. (Cwt)	Cif (S$)	Qty. (Cwt)	Cif (S$)	Qty. (Cwt)	Cif (S$)
Copper Powders and Flakes (682240)						
Germany Fed. Rep.	245.34	72,153	265.74	70,883	355.80	98,998
Hong Kong	3.72	4,698	4.61	2,103	1.98	1,023
Japan	6.42	2,921	15.19	9,979	-	-
Sweden	21.42	6,849	17.83	5,322	-	-
United Kingdom	109.16	39,643	30.50	13,619	10.15	5,818
U.S.A.	1.78	1,214	11.40	10,747	2.67	1,850
W. Malaysia	38.16	5,172	0.03	35	0.89	300
Australia	-	-	0.13	36	-	-
China	-	-	-	-	8.30	1,919
Total	426.00	132,650	345.43	112,724	379.79	109,908

	1969		1968		1967	
	Qty. (ton)	Cif (S$)	Qty. (ton)	Cif (S$)	Qty. (ton)	Cif (S$)
Iron or Steel Shot Powders or Sponge (671300)						
Australia	37.48	16,816	68.83	31,529	5.00	1,755
Germany Fed. Rep.	0.02	129	-	-	-	26
India	12.47	4,911	27.00	13,355	-	-
Italy and San Marino	24.60	11,482	0.01	31	-	-
Japan	6.88	5,401	0.89	3,269	6.95	4,186
United Kingdom	154.34	56,878	51.18	16,990	100.13	37,726
U.S.A.	0.98	1,000	-	-	1.50	1,641
W. Malaysia	7.60	3,150	1.11	54	-	-

	1969		1968		1967	
	Qty. (ton)	Cif (S$)	Qty. (ton)	Cif (S$)	Qty. (ton)	Cif (S$)
Thailand	-	-	0.90	1,240	-	-
Canada	-	-	-	-	0.98	1,360
Sweden	-	-	-	-	5.00	5,390
	244.37	99,767	149.92	67,497	119.47	52,084

1.4.3 Import Duties and Tariffs

There are no import duties or tariffs imposed on import of iron powders into Singapore. Similarly there are no import duties or tariffs relating to copper powders.

1.4.4 Summary

The market for these powders in Singapore is small and would have to be shared with at least ten other exporting countries as can be seen from the import statistics above.

1.5 Pakistan

1.5.1 General

Pakistan has an acute shortage of foreign exchange and import trade is subject to strict controls. The current import policy valid to June 30, 1971, permits importation of metal powders under Free List. Items on this list may be imported only from specified countries in terms of barter aid, loan or credit agreements which Pakistan has concluded with some countries. Australia does not have an agreement of this kind so that under present conditions the Pakistan iron powder market is closed to an Australian exporter.

However, ferrous alloys can be imported under the cash-cum-bonus list which in effect means that Australian exporters could export ferrous alloy powders to industrial users and established importers according to their quotas and entitlements.

1.5.2 Import Statistics

Import Statistics for these powders are not classified by type, usage or country of origin. The following statistics show the value of imports for the last three years.

	* 1967-68	* 1968-69	* 1969-70
Iron and Steel Powder	11,553	2,030	3,053
Copper Powder and Flakes	10,194	88,360	214,236

	* 1967-68	* 1968-69	* 1969-70
Ferro Manganese	1,489,607	732,568	1,944,948
Ferro Aluminium	6,478	6,212	10,160
Ferro Silicon Alloys	743,165	712,959	1,256,098
Ferro Alloys N.S.	163,328	70,317	846,952
Ferro Chrome Alloys NES	-	16,295	10,174

* Value in Rupees - (5.38 rupees = A$1.00)

1.5.3 Tariffs and Import Duties

The following import duties are applicable to metal powders:

Iron and Steel Powder	25% ad. val.
Copper Powder	50% ad. val.
Ferro Alloys Powder	25% ad. val.

1.5.4 Summary

The import statistics and the current import restrictions mean that only ferro-manganese could possibly be exported to Pakistan. However, further investigation is obviously warranted.

1.6 Philippines

1.6.1 General

The potential market for these powders in the Philippines is not promising. There are no trends or planned developments which could see a significant increase in local usage. In addition, imports have been heavily curtailed by Government restrictions during 1969 and 1970 as a direct consequence of an acute shortage of foreign exchange.

1.6.2 Import Statistics

The latest statistics received are those for 1967. Later statistics are being obtained.

1967 Imports of Iron and Steel Powder

Country of Origin	Quantity (Kilos)	F.O.B. Value U.S. Dollars
United States	92,608	18,228
Federal Republic of Germany	4,000	1,342

Country of Origin	Quantity (Kilos)	F.O.B. Value U.S. Dollars
Sweden	80,100	12,204
Japan	108,975	25,563
	285,683	57,337

1967 Imports of Copper Powder and Flakes

Country of Origin	Quantity (Kilos)	Value U.S. Dollars
United Kingdom	226	259
Japan	10	61
	236	320

1.6.3 Tariffs and Import Duties

Imports of iron powders are free of duty. The applicable rate of duty on copper powder is 20% ad valorem which is assessed on the C.I.F. value.

1.6.4 Summary

The market for metal powders in the Philippines is small and shows little indication of growing.

Competition from existing suppliers and the import restrictions make the prospects even less encouraging.

1.7 Thailand

1.7.1 General

The market is small as the usage of metal powders is related to the stage of Thailand's industrialisation.

1.7.2 Import Statistics

The statistics for iron and steel powders and "ferro alloys" imports follow. However, the proportion of powders in the "ferro-alloy" statistics are not recorded.

Iron and Ferrous-Alloy Powders

a. Quantity is in kilograms

b. Value is in Australian Dollars

	1967		1968		1969	
Iron and Steel Powders	Quantity	Value	Quantity	Value	Quantity	Value
Japan	67	52	14,874	5,152	250	58
Sweden	-	-	-	-	5,000	1,482
United Kingdom	100	24	731	60	-	-
Australia	2,000	282	-	-	-	-
Total	2,168	364	15,607	5,223	5,349	1,763
Ferro-Alloys						
India	-	-	365,847	61,798	614,707	134,284
Japan	220,360	55,906	975,571	178,910	637,000	109,976
Norway	294,416	49,485	291,835	52,539	338,850	52,171
U.S. Africa	1,260,161	197,130	1,413,161	227,808	1,790,000	280,832
Total	2,332,092	417,983	3,487,383	638,328	3,604,633	635,444

1.7.3 Tariffs and Import Duties

Import duty on iron and ferrous alloy powders would be 5% on the c.i.f. Bangkok price. A business tax of 1.65% on the aggregate of the c.i.f. price plus duty would also be levied.

1.7.4 Summary

The Thai market is small and does not appear to have prospects of rapid growth. The size of the ferrous alloy market cannot be gauged from import statistics.

1.8 Other Possible Markets in South-East Asia

There are possible markets in Malaysia and Taiwan for iron and copper powders. Unfortunately, Trade Commissioners responsible for these areas have not yet forwarded information requested on local markets for these metal powders.

1.9 General Summary of South-East Asian Markets

The existing markets for metal powders in South-East Asia are small. Most of the countries in this region have foreign exchange problems which curb potential imports. In addition, these countries have not reached the stage and rate of industrial development that results in significant demand for metal powders.

The small market that exists in these developing nations of South-East

Asia is supplied by six or seven industrialised countries. Consequently, an Australian exporter would meet considerable competition in these markets. In particular, Japan which imports copper alloy powders but exports copper powder, is expected to increase its exports of copper powder to South-East Asia. Appendix Two contains details of Japan's exports of copper powder for the past three years.

2. SOUTH AFRICA - IRON AND COPPER POWDERS

2.1 General

In the limited time available it has not been established with certainty whether all copper, iron, and associated alloy powders used in South Africa are imported. Supreme Metal Works Pty.Ltd. in Transvaal, which is the only South African associate member of the U.S. Metal Powders Producers Association, produces aluminium powder only.

2.2 Import Statistics

Product - Iron or steel powders

Tariff Item	Main Suppliers	Imports 1969 (lbs.)	Value 1969 (f.o.b. - Rand)
73.05.10	Sweden	2,179,160	104,613
	U.K.	69,186	10,411
	France	162,095	10,266
		2,453,276	136,565
		Imports 1970	Value 1970
73.05.10	Sweden	2,808,158	135,687
	U.K.	18,403	11,193
	Belgium	144,842	7,433
	W. Germany	35,939	10,931
	Canada	200	871
	U.S.A.	830	3,197
		3,008,372	169,312

Product - Copper Powder & Flakes

Tariff Item	Main Suppliers	Imports 1969	Value 1969
74.06	U.K.	213,878	161,940
	W. Germany	48,582	36,006
		268,013	203,188
		Imports 1970	**Value 1970**
	Sweden	2,016	1,438
	U.K.	208,462	177,149
	W. Germany	36,457	30,641
	France	110	188
	U.S.A.	2,350	3,339
		249,395	212,755

2.3 Tariff and Import Duties

The metal powders in question are imported free of duty into South Africa.

2.4 Summary

The import statistics show that the imports of iron and steel powders grew from 1,095.2 tons in 1969 to 1,343 tons in 1970. Taking an exchange rate of one Rand to 1.2513 Australian dollars these imports were worth $A170,884 in 1969 and $A211,860 in 1970.

While at least 5 years' statistics are needed to establish any trend, the two years' statistics available indicate the South African market for iron and steel powders is growing.

The import statistics for copper powder and flakes show that imports declined from 119.6 tons in 1969 to 111.3 tons in 1970. These imports were valued at $A254,249 in 1969 and $A266,220 in 1970.

3. JAPAN - COPPER POWDERS ONLY

3.1 General

As mentioned in Section 1.9 Japan exports copper powder but imports copper-alloy powders. The Australian Embassy Trade Office has advised in Japan copper metal supply is currently in a surplus situation. The three major copper powder manufacturing companies are Fukuta Metal Foil and

Powder Company, Mitsui Mining and Smelting Company and Nippon Mining Company. These companies are supplying copper powder to Japanese and overseas markets particularly in South-East Asia. In 1969 Japan exported copper powder worth $A7,200 approximately to Australia.

It is understood that copper alloy flakes and powders are expected to be produced in Japan by the end of the year. The producing company will be a joint-venture between the Asaki Chemical Industry Company of Japan and a German manufacturer, Eckart-Werke.

3.2 Import Statistics

Import statistics relating to imports of copper alloy powders and exports statistics relating to the export of copper powder can be found in Appendix Two.

3.3 Tariffs and Import Duties

The present Japanese tariff on copper powders and flakes is 16%. This will be reduced to 15% later in 1971.

3.4 Summary

The Japanese market for copper powders and copper alloy powders does not look promising for an Australian exporter. Local production and tariff protection, together with competition from established suppliers, indicates only a small share of the market for copper alloy powders could be anticipated.

Japanese imports of copper alloy powders were worth approximately $A340,000 in 1969.

4. EUROPE AND UNITED KINGDOM - COPPER AND COPPER ALLOY POWDERS

4.1 Austria

4.1.1 General

The only information received from Austria is that in 1969 imports of copper powder and flakes were 124.5 metric tons valued at $A240,000. This limited information would indicate little market potential in Austria for an Australian exporter of copper powders.

4.2 France

4.2.1 General

France is both an exporter and importer of copper powders. However, the tonnage of imports is far greater than the export tonnage which was 98 m tons in 1967, 162 m tons in 1968 and 249 m tons in 1969.

4.2.2 Import Statistics

French imports for the years 1967, 1968 and 1969 are given in the table below. Quantities are in metric tons and values in Australian dollars.

	1967	1968	1969
74.06.12 Copper powder, lamellar and flakes for dye manufacture. Alloyed or non-alloyed.			
Total	325 ($550,000)	343 ($582,000)	Not detailed
of which E.E.C.	260	284	
U.K.	61	54	group in next
U.S.A.	4	5	item
74.06.13 Copper powder, lamellar and flakes for other uses. Alloyed or non-alloyed.			
Total (origin not given)	4 ($13,500)	15 ($36,000)	448 ($900,000)
of which E.E.C.	-	-	356
U.K.	-	-	81
U.S.A.	-	-	11
74.06.15 Other copper powders or flakes, having at least 90% (in weight) of 0.315 mm diameter			
Total	487 ($590,000)	486 ($628,000)	989 ($1,580,000)
of which E.E.C.	118	107	309
U.K.	324	359	517
U.S.A.	45	20	151

	1967	1968	1969
74.06.17			
Other copper powders n.e.i.			
Total	246 ($263,000)	297 ($347,000)	432 ($626,000)
of which E.E.C	55	67	158
U.K.	191	229	267

4.2.3 Tariffs and Import Duties

Copper powders, lamellar and flakes, under tariff number 74.06 of the Brussels nomenclature are subject to a duty rate of 10.8% when entering France from countries other than members or associates of the Common Market. Other copper powders are subject to 1.8% duty. Duties are applied to c.i.f. landed prices.

4.2.4. Summary

The French market for copper powders appears relatively large on the basis of import statistics. However, existing suppliers have distinctive advantages over an Australian exporter. E.E.C. member suppliers have a tariff and geographic advantage while U.K. suppliers have a geographic advantage and possibly a tariff advantage if the U.K. joins the E.E.C.

4.3 West Germany

4.3.1. General

Copper powder production, imports, exports and local consumption from 1959-69 in West Germany in metric tons are as follows:

Year	Production	Imports	Exports	Apparent Local Consumption
1959	4.006	31	1.489	2.548
1960	4.749	107	1.477	3.379
1961	4.839	80	1.750	3.169
1962	4.895	69	1.882	3.082
1963	5.786	109	2.277	3.618
1964	6.167	93	2.774	3.986
1965	6.668	173	2.574	4.267

Year	Production	Imports	Exports	Apparent Local Consumption
1966	6.880	110	3.132	3.858
1967	7.109	122	3.588	3.643
1969	10.095	220	4.560	5.535

These figures show in particular:

a. the consumption of copper powder in 1969 was more than twice as high as in 1959.

b. local consumption is mainly covered by local production. The rate of growth of production is higher than that of consumption.

c. imports have grown rapidly over the last years but are small in relation to the market.

d. a little less than half of the copper powder production is for export.

Copper powder production for pigment purposes is larger than for sintering purposes and more than half the production for pigment purposes is exported. This export figure may be divided into one half exported to E.E.C. countries and the other half to U.S.A. and Switzerland.

In the case of copper powder for sintering purposes one-third goes to E.E.C. countries and two-thirds to the Soviet Union and Sweden. Imports of copper powder for sintering purposes come mainly from the U.S.A. and the U.K. whereas imports of copper powder for pigment use come mainly from Switzerland and the E.E.C. countries.

Prices vary and are based on London Metal Exchange quotations. At present German FOB/Free German customer prices are:

DM 5.40 ($A1.33)	per kg for sintering copper powder (coarse grained)
DM 8.70 ($A2.16)	per kg for pigment copper powder (medium fine grained)

The largest producer of sintering copper powder in Germany is Norddeutsche Affinerie, Metallhutten-und Schmelzwerke, 200 Hamburg 36, Alsterterrasse 2.

The largest producer of pigment copper powder is Eckart-Werke, 8510 Furth/Bay., Kaiserstr. 30. (See Section 3.1).

4.3.2 Import Statistics

These can be found in the table in the preceding Section (4.3.1).

4.3.3. Tariffs and Import Duties

Import duties for copper powder are:

740610T	10.8% for copper powder with lamellar structure or flaked
740620T	1.8% for copper powder other than above

4.3.4 Summary

The current trend in the West German market for copper powder is toward a higher rate of domestic production and falling prices. This situation is indicated by the statistics in the table in Section 4.3.1.

This trend, together with the tariff barrier, would make the marketing of Australian copper powders in West Germany difficult.

4.4 Italy

4.4.1 General

Copper powders are produced in Italy but the bulk of Italian consumption is supplied by imports from the United Kingdom, Sweden, West Germany.

4.4.2 Import Statistics

Copper Powders		Quantity (m tons)	Value ($A)
	1968	108	207,000
	1969	103	396,500
Jan/Sept.	1970	56	142,500
Copper Alloy Powders			
	1968	176	256,000
	1969	133	272,000
Jan/Sept.	1970	152	534,000

4.4.3 Tariffs and Import Duties

Import duties on copper and copper alloy powders is 11.06%. In addition there are other taxes payable amounting to about 9%.

4.4.4 Summary

The Italian market for copper and copper alloy powders is relatively small and E.E.C. suppliers have a tariff advantage over a potential Australian supplier.

4.5 The United Kingdom

4.5.1 General

The United Kingdom is one of the leading exporters of copper and copper alloy powders to all parts of the world.

4.5.2 Import Statistics

In 1969 imports of copper powders and flakes totalled 77.5 tons valued at $A188,101.

4.5.3 Tariffs and Import Duties

Imports from Commonwealth and European Free Trade Association countries are free from duty. Imports from other countries are subject to 12% duty.

4.5.4 Summary

As the United Kingdom is a major exporter of copper and copper alloy powders strong competition from local producers from the U.K. market could be expected. The relatively small volume of imports is further evidence of the strong hold local producers have on the British market for copper and copper-alloy powders.

4.6 Other Western European Countries

To date information requested from Greece, Spain, the Netherlands and Sweden has not been received. It is known that Sweden is an exporter of copper powders. Belgium and the Netherlands, being members of the European Economic Community, would have similar tariff barriers to Australian imports of copper powders as were described for France, West Germany and Italy.

The markets in Greece and Spain would probably not be as large as those in more industrialised Western European countries.

4.7 General Summary of Western European Markets

An initial appraisal of markets for copper and copper alloy powders in Western Europe does not reveal any major export opportunities. In addition, there appears to be a trend toward production growing at a faster rate than consumption in the industrialised nations of Western Europe. The only statistical evidence for this received is the table relating to West Germany reproduced in Section 4.3.1. However, this trend is also evident in North Amercia and Japan according to metal powder industry sources in those countries.

5. NEW ZEALAND - IRON AND COPPER POWDERS

5.1 General

The New Zealand market for these powders is dependent on welding as the use for sintered parts is considered insignificant. Market growth is therefore not expected to be rapid.

5.2 Import Statistics

The latest import statistics available are those relating to 1967-68. Statistics relating to imports of ferrous alloy and copper alloy powders are unfortunately not separated from imports of ingots and bars of ferrous alloys and copper alloys.

Imports into New Zealand 1967-68

Country of Origin	Quantity (cwt)	c.i.f. Value (NZ Dollars)
Iron or Steel Powders		
Australia	864	10,655
West Germany	-	13
Sweden	787	5,704
United Kingdom	-	314
	1,651	16,676
Copper Powders/Flakes		
Australia	-	378
Canada	-	1,060
West Germany	-	19,252
Norway	-	2,618
Sweden		1,152
Switzerland		145
United Kingdom		19,866
U.S.A.		2,533
		47,004

5.3 Import Duties and Tariffs

The metal powders in question are free of duty and import licensing restrictions.

5.4 Summary

New Zealand currency is officially at parity with Australian currency. The 1967-68 import statistics listed in Section 5.2 indicate that the annual

New Zealand markets for iron and copper powders are very small. These markets would presumably have grown marginally since 1967-68 but would still present a poor outlet for an Australian metal powder producer.

6. SOUTH AMERICA - IRON AND COPPER POWDERS

6.1 Mexico

6.1.1 General

The following import statistics show significant markets for both copper and iron powders. Unfortunately ferrous-alloy and copper alloy powders are not separated from ingots and bars in the Mexican import statistics.

6.1.2 Import Statistics

Item	Origin	Weight (Kilos)	Value (Pesos)
		1969	
Iron or Steel Powders	Canada	2,472	6,891
	U.S.A.	448,601	2,203,876
	W. Germany	76,521	251,309
	Belgium	119	2,016
	Spain	5	14
		527,718	2,464,106
		1968	
	U.S.A.	428,952	1,646,831
	W. Germany	35,581	102,874
	Belgium	123	2,063
	Italy	57	505
		464,713	1,752,273
		1969	
Copper Powders	U.S.A.	195,810	4,931,008
	G.F.R.	147,955	4,075,731
	Belgium	167	3,906
	Spain	13	36
	France	9	438
	United Kingdom	1,257	38,834
	Italy	1,721	50,767

		1969	
Item	Origin	Weight (Kilos)	Value (Pesos)
Copper Powders	Netherlands	219	17,279
	Sweden	6,060	133,057
	Switzerland	35	1,509
	Japan	2	28
	Total	353,238	9,252,593
		1968	
	U.S.A.	123,376	2,508,498
	G.F.R.	135,701	3,255,750
	Austria	1,625	38,094
	Belgium	1,339	30,416
	France	1,199	18,756
	United Kingdom	388	15,265
	Italy	1,684	47,747
	Netherlands	1,109	23,665
	Sweden	23,936	541,209
	Total	290,357	6,479,400

6.1.3 Tariffs and Import Duties

Import duties are 15% levied on the official price which is 18 pesos per gross kilo in the case of copper powders and 3 pesos per gross kilo for iron powders.

6.1.4 Summary

Using an exchange rate of 1 peso to 7 cents Australian, imports of iron powders were valued at $A122,659 in 1968 and $A172,487 in 1969. Copper powder imports were valued at $A453,558 in 1968 and $A647,682 in 1969.

6.2 Other South American Countries

Replies to our request for information relating to markets in other South American countries have not been received to date.

APPENDIX EA

Imports into Hong Kong

	1968		1969		Jan/Sept 1970	
	Cwt.	HK$	Cwt.	HK$	Cwt.	HK$
(671320)						
Iron, Steel Powders						
Germany Fed. Rep.	18	1,067	-	-	-	-
Japan	16,949	482,129	18,133	531,752	5,032	150,858
	16,967	483,196	18,133	531,752	5,032	150,858
		(A$69,028)		(A$75,965)		(A$21,551)
(671400)						
Fe Manganese Alloys						
Norway	3,000	139,991	1,574	71,898	-	-
Japan	59	5,135	653	37,294	-	-
South Africa	600	27,078	1,800	122,078	4,089	209,590
Taiwan	-	-	590	31,927	-	-
Mozambique	-	-	600	46,000	-	-
Germany Fed. Rep.	-	-	-	-	590	52,604
France	-	-	-	-	23	2,853
	3,659	172,204	5,217	309,197	4,702	265,047
		(A$24,601)		(A$44,171)		(A$37,864)
(671500)						
Ferro Alloys, n.e.s.						
Canada	1,968	120,665	984	58,551	1,139	99,436
Taiwan	590	35,633	-	-	-	-
U.S.A.	-	-	40	16,710	-	-
Germany Fed. Rep.	-	-	2	2,380	-	-
United Kingdom	-	-	96	7,127	-	-
Japan	-	-	104	10,841	-	-
Mozambique	-	-	1,000	56,610	1,000	55,974
	2,558	156,298	2,226	152,219	2,139	155,410
		(A$22,328)		(A$21,746)		(A$22,201)

	1968		1969		Jan/Sept 1970	
	Cwt.	HK$	Cwt.	HK$	Cwt.	HK$
(682240)						
Copper Powders						
U.S.A.	35	43,215	23	28,962	56	56,583
Canada	9	7,300	19	13,300	-	-
Sweden	8	6,300	8	4,932	5	3,120
Germany Fed. Rep.	1,329	830,005	1,159	765,705	1,086	799,026
United Kingdom	428	341,864	518	394,400	710	697,467
Japan	84	59,119	78	68,615	47	31,831
	1,893	1,287,803	1,805	1,275,914	1,904	1,588,027
		(A$183,972)		(A$182,273)		(A$226,861)

(A$1.- to HK$7.-)

APPENDIX EB

Japanese Imports of Copper-Alloy Powders and Exports of Copper Powders

Source: Ministry of Finance

	1967			1968	
Imported from	Quantity Kg	Value 1,000 yen	Imported from	Quantity Kg	Value 1,000 yen
United Kingdom	286	308	United Kingdom	28,002	23,698
West Germany	64,920	50,388	West Germany	100,876	77,289
Canada	50	328	Canada	–	–
U.S.A.	27,222	14,855	U.S.A.	13,578	10,826
Total	92,478	65,879	Total	142,456	111,813

Exported to	Quantity Kg	Value 1,000 yen	Exported to	Quantity Kg	Value 1,000 yen
Ryukyu	445	457	Ryukyu	300	319
South Korea	9,683	8,499	South Korea	6,535	6,092
Communist China	14,945	13,568	Communist China	14,710	13,350
Republic of China	11,510	10,947	Republic of China	20,122	17,919
Hong Kong	6,474	6,044	Hong Kong	12,046	9,434
Thailand	750	813	Thailand	820	702
Philippines	871	838	Philippines	1,503	1,271
Pakistan	197	168	Pakistan	203	176
Iraq	376	396	Iraq	749	580
Turkey	250	430	Turkey	200	336
Nigeria	100	134	Nigeria	200	258
Australia	7,644	3,989	Australia	12,500	7,569
Indonesia	100	199	Indonesia	–	–
Ceylon	38	41	Ceylon	–	–
Afganistan	226	244	Afganistan	–	–
Mexico	700	870	Mexico	–	–
Brazil	1,000	751	Brazil	–	–

	1967			1968	
Exported to	Quantity Kg	Value 1,000 yen	Exported to	Quantity Kg	Value 1,000 yen
Tanzania	50	53	Tanzania	-	-
North Korea	-	-	North Korea	75	92
Rep. of Vietnam	-	-	Rep. of Vietnam	1,000	1,048
Singapore	-	-	Singapore	390	571
Malaysia	-	-	Malaysia	40	72
Cambodia	-	-	Cambodia	360	291
India	-	-	India	150	132
Iran	-	-	Iran	300	163
Israel	-	-	Israel	50	107
U.S.S.R	-	-	U.S.S.R	100	225
U.S.A.	-	-	U.S.A.	132	502
Chile	-	-	Chile	500	523
Argentina	-	-	Argentina	1,200	762
Total	55,359	48,441	Total	74,185	62,494

	1969		1970 (January-October)		
Imported from	Quantity Kg	Value 1,000 yen	Imported from	Quantity Kg	Value 1,000 yen
United Kingdom	42,726	35,524	United Kingdom	34,203	31,892
West Germany	83,136	66,605	West Germany	77,502	75,644
U.S.A.	40,785	32,679	U.S.A.	56,747	48,252
France	50	89	France	-	-
Italy	270	555	Italy	-	-
Canada	-	-	Canada	13	110
Total	166,967	135,452	Total	168,465	155,898

	1969			1970 (January-October)	
Exported to	Quantity Kg	Value 1,000 yen	Exported to	Quantity Kg	Value 1,000 yen
Ryukyu	544	741	Ryukyu	480	615
South Korea	8,112	7,392	South Korea	18,792	16,518
North Korea	65	102	North Korea	690	872
Communist China	11,290	10,832	Communist China	1,300	1,334
Republic of China	28,834	18,575	Republic of China	36,529	31,159
Hong Kong	7,501	6,240	Hong Kong	7,257	8,248
Thailand	630	514	Thailand	1,113	1,857
Singapore	514	646	Singapore	968	1,048
Malaya	0	142	Malaya	50	85
Philippines	826	999	Philippines	7,697	10,458
Indonesia	427	187	Indonesia	1,217	1,907
India	199	121	India	13,400	11,645
Israel	100	205	Israel	100	205
Nigeria	100	130	Nigeria	200	285
Australia	4,727	2,892	Australia	2,000	1,378
Rep. of Vietnam	500	415	Rep of Vietnam	-	-
Cambodia	381	333	Cambodia	-	-
Pakistan	254	265	Pakistan	-	-
Iraq	218	289	Iraq	-	-
U.S.S.R	224	487	U.S.S.R.	-	-
U.S.A.	695	708	U.S.A.	-	-
Panama	90	94	Panama	-	-
Lebanon	-	-	Lebanon	30	217
Canada	-	-	Canada	20	392
Chile	-	-	Chile	30	98
Ceylon	-	-	Ceylon	16	43
Total	66,231	52,309	Total	91,889	88,364

APPENDIX EC

Notes On The Iron Powder Industry In North America Extracted From I.T.S. Files

The powder metallurgy process derived its basic impetus from two quite different sources. The first, and probably most important, was a need for soft sintered iron rotary bands for German artillery shells during World War, 1. The critical availability of copper, which was the conventional material, prompted the search for a substitute material. Rotating bands made from iron powder were developed and used. This requirement reappeared during World War 11 and again the powder process was employed. The acceptance of iron and copper parts fabricated from powders has shown a strong rate of growth since that time.

The second important impetus for the powder process was the increasing demands for the refractory metals such as tungsten and molybdenum. While both tungsten and tungsten carbides were conventionally produced by the powder process, the requirements for improved grades of these materials prompted considerable research on the powder approach. This led to the development of improved compacting and sintering techniques which were in turn applied to iron, copper, aluminum, and more recently, titanium parts fabricated from powder. In addition to ingots and mechanical parts, the powder process is widely used in the production of porous elements and currently is receiving considerable attention for the production of wrought strip and sheet and for seamless tubing blanks. However, the major application of the powder process is in the production of molded parts of iron and steel. The following discussion is based on previous studies conducted by Battelle-Columbus on iron and stainless steel powder parts problems.

The iron powder industry in the United States showed a decade of grandual development during the period 1945 to 1955. Since 1955, there has been a sharp increase in the growth of the industry. For example, the shipments of iron powder increased from 32,000 tons in 1960, to 100,000 tons in 1966. During this same period, the shipments of iron powder parts increased from 12,000 tons in 1960, to 62,000 tons in 1966. Iron powder parts production consumed approximately 30 percent of the total iron powder produced during the period 1955 to 1961. Since that time, this percentage has increased, and in 1966, powder parts consumed approximately 60 percent of the total powder shipments. However, it must be noted that the total powder shipments shown include all grades of iron powder, including those used for welding and cutting applications.

Traditionally the production capacity for iron powders in the United States has exceeded the shipments by about 60 percent. This ratio of excess capacity decreased during the period 1963 to 1966, as the shipments increased at a greater rate than did the production capacity. However, additional capacity brought into production during 1968, together with more than 130,000 tons of new capacity planned for 1969, assures a more than adequate supply for present and near future requirements.

In addition, there has been a significant increase in imports of powder to the United States from Italy and Japan within the past 3 years. However, imports represent less than 3 percent of the total powder market at present.

In 1965, powder metallurgy parts accounted for 65 percent of the total iron powder shipments; welding fluxes 23 percent; flame cutting 7 percent; electronic and magnetic applications 2 percent; and miscellaneous uses 3 percent. It is expected that the proportion of parts to the total production will continue to increase in relation to the overall iron powder consumption, particularly with the advent of increased captive production capacity in iron powder parts producing facilities.

The distribution of powder metallurgy parts, by the percentage of total weight, to the consuming industry is as follows. As would be expected, the automotive industry is the largest single user, representing in 1965 about 60 percent of the total demand. It is expected that this proportion of the market will increase. Present automotive consumption of powdered parts represents about 10 to 12 pounds per car, and it is expected that, by 1970, this will increase to about 20 pounds per vehicle. This weight, however, includes the use of not only iron, but also copper, aluminum, and some stainless steel parts.

The appliance industry is second to the automotive field in total weight of iron parts consumed, mainly because of the utilization of relatively small parts in these applications. Other important segments of the market include farm and garden equipment at 8 percent, hardware at 7 percent, and business machines at 6 percent. These latter segments are expected to grow at a higher rate than the automotive portion of the pressnt market.

The increasing acceptance of the powder process is being reflected in a marked upward trend in the acceptance of stainless steel powders. In 1968 the U.S. shipments of stainless powder amounted to 4 million pounds. In 1969 the shipments of stainless powder are expected to increase to slightly more than 5 million pounds. The increasing acceptance of stainless powders in the 310, 316, and 410 grades can be attributed to the fact that the oxide coatings of the particles can be reduced during subsequent heat treatment by use of proper furnace atmospheres. The primary reason for the increased acceptance of stainless powder is the marked saving in machining costs.

While the current U.S. capacity for producing iron powder is well in excess of the demand, there is no adequate supply of premium quality stainless powders at present. However, most of the producers of stainless powder are conducting research on improved production processes.

Technical Considerations

At present there are a number of technical developments that will exert a strong influence on the future demand for molding-grade iron powders. The inherent advantages of the powder process include high material yield; controlled uniformity; control of properties by blending; grain-size control, and readily automated processing. An additional advantage currently receiving increased attention is the improved wear resistance of properly made powder parts as compared with machined or forged components in a number of applications where sliding wear is encountered. Another hidden advantage is the frequently reduced scrap or chip disposition cost. While initially limited to relatively small parts, usually less than 6 ounces, recent improvements in press design and capacity will accelerate application of parts in the 2 to 4 pound range. Many of the parts fabricators interviewed are producing increasing numbers of parts in the 1 to 2 pound sizes. They are installing new, heavier equipment that will provide high speed production of parts up to 4 pounds. In addition, while the original presses were single action with severe limitations as to parts design, several of the new press designs offer both vertical and horizontal action. Thus, the wider design tolerance, increased capacity, and higher rate presses will accelerate the trend to larger parts. At the present time, parts up to 12, 20, and 40 pounds have been produced on a development basis. This trend to larger parts is expected to provide a strong impetus to the increasing demand for iron powders.

The high energy rate forming practices such as the Dynapak (and similar equipment) explosive compacting, isostatic pressing, and hydrostatic and hydraulic extrusion are being applied to the compaction of powder parts. Several high production installations are working on increasing the speed and automation of the Dynapak. It is said that a resultant increase in green density frequently results in a reduction in sintering time or permits a reduction in sintering temperature. An interesting innovation is that of compacting, heating and hot forging rather than sintering after compaction. This is claimed to provide better densities with resultant higher strengths and improved fatigue properties. In addition to improvements in compacting technology, the furnace producers are now designing more sophisticated equipment for the powder parts producers. These improvements collectively accelerate the growing acceptance of powder parts through both improved quality and lowered production costs.

One of the most critical problems in the powder metallurgy field is the lack of standardized quality specifications with regard to mechanical properties. Here the German industry appears to have led the way in establishing acceptable specifications and classifications for various grades of processed parts. In this respect, the Metal Powder Institute in the United States is aware of this problem and is attempting to promote such standardization. Another critical problem is a lack of designers with experience in designing for the powder process. These difficulties can be attributed to the fact that the powder process, as an accepted production method, is too new to have acquired a substantial background and reserve of trained personnel.

In addition to the above-mentioned aspects, the questions of quality control and inspection methods remain to be answered. Here again, powder metallurgy, as a new process, is at a disadvantage with regard to having established inspection and control procedures. However, the increasing number of parts being made can be considered indicative of the progress in these fields, and reliability does not appear to be a major concern in iron and stainless applications.

Present Producers

While at the present time there are 22 commercial sources of iron powder, most of today's market demands are being met by six producers. These producers, together with the processes employed and the present production capacity, are shown in the following table. The second table shows the estimated capital equipment and production costs for current processes.

Major Producers, Processes, and Present Capacity (1967)

Producers	Process	Installed Capacity in 1967, tons/year
Hoeganaes	Carbon reduction of enriched ore	90,000
Pyron Div. of AMAX	Hydrogen reduction of mill scale	20,000
Domtar	Atomized high-carbon metal	25,000[(a)]
Glidden	Electrolytic; and carbon reduction of mill scale	10,000
Easton	Carbon reduction of oxidized, water-atomized, steel scrap	20,000
A.O. Smith	Water atomization of melted scrap	5,000[(b)]
	Total	170,000

(a) Represents combined capacity of Canadian and U.S. production facilities.

(b) Can be readily expanded to 20,000 tons annually.

Estimated Capital Equipment and Production Costs
for Producing 50,050 tons of Iron Powder
per year by three processes

	Carbon Reduction of Iron Ore Concentrates (SL-RN Process) (a)	Hydrogen Reduction of Mill Scale	Atomization of Melted Steel Scrap
Total Capital Investment, Dollars	3,915,840	7,305,000	2,150,000
Investment Cost per ton annual capacity, dollars	78.22	145.95	42.96
Fixed Charge, dollars per net ton (12% of Capital)	9.39	17.51	5.15
Production Cost, dollars per net ton	68.50	79.02	68.87
Powder Cost, cents per pound	3.53	4.00	3.44

(a) Steel Company of Canada - Lurgi Gesellschaft fur Chemie und Huttenwesen m.b.H; Frankfurt (Main), West Germany - Republic Steel Co. - National Steel Company patented direct reduction kiln process. Stelco and Lurgi developed the SL process and took over the patents of the RN process after the RN

Corporation was dissolved.

Pricing

The question of the effect of pricing on the volume of sales has been a serious issue in the powder producing industry for 15 years. The consensus of the consumers suggests that the present price structure could be reduced by 2 cents per pound without jeopardizing investment requirements and that any price reductions would increase the demand. On the other side of the coin, the producers are quick to emphasize that powder prices have remained stable while prices of other forms of steel have shown a number of increases. Further, the quality of the powders has been consistently improved without increasing the price.

Basically the problem can be partially attributed to the overall state of the art with regard to the present position of the industry on the growth or learning curve. As mentioned earlier, present markets for iron powder represent less than 1 percent of the total steel market, and as a volume market it is relatively new. While production capacity has consistently exceeded the market demand, it is only recently that any of the producers have operated at a level of maximum efficiency. An example of the effect of level of operation on pricing is Hoeganaes recent announcement of a price reduction of 1 cent per pound which is attributed to the improved efficiency of a new kiln with a capacity of 50,000 tons per annum. Conversely, a new producer, Fina Metal Ltd., plans to enter the market with an initial capacity of 15,000 tons per annum at current price levels. If the new product meets with the anticipated acceptance, the present plans call for price reductions as the volume increases.

An interesting anomaly in contract price versus demand considerations is the wide range of accepted prices. While the larger consumers are frequently of the opinion that the price of powder is 2 cents per pound higher than it should be in the 10.5 to 11.5 cents per pound grades A.O. Smith Company had little difficulty penetrating the market with a premium priced (14 cents per pound powder.) Further, several consumers reported using electrolytic grades at 27 cents per pound exclusively on the basis that over-all production costs were lower for premium quality parts. This then introduces the quality or grade consideration. In order to improve the metallurgical grades of powder to more closely approach the electrolytic quality, the aggressive producers have invested in product oriented research and development programs.

Increasing demands on quality further necessitate added quality control operations, with a resultant increase in costs. The present markets and production rates may not be sufficient to support extensive research and development programs in both process and production development, and a review of the production processes suggests that this area may have been slighted. However, irrespective of the direction of application, it can be reasonably assumed that the cost of research and development has also contributed to the difference between price and cost.

Examination of the difference between cost and prices requires consideration of both technical and economic factors. One of the important contributing factors is that of lack of standardization. While such basic specifications as chemistry, particle size distribution, flow rate, and hydrogen weight loss are commonly used to specify powder grades, in many instances there are not sufficient to permit using alternative sources of supply for a given application. Such factors as particle shape, porosity, and hardness, which may be determined by both starting material and manufacturing process, have a pronounced effect on dimensional characteristics, density, and physical properties of the finished parts. The larger producers of powder parts prefer not to be dependent upon a single source of supply for any one important grade of powder. Dies are usually designed for use with a specific brand of powder. Only within the past year has it been possible to interchange certain grades from different producers without penalty, and this frequently requires custom blending. The development work required to obtain this "match" is added to the cost of the powder.

A further contributing factor to the difference between cost and price is that of technical sales and service charges. Because the technology is relatively new, many of the potential users frequently depend heavily upon their suppliers for technical assistance. While such services are common to the metal market, there is a higher degree than normal in the powder-metals-supply business. Another indirect factor is that the supply or availability of metallurgists skilled in powder technology is not sufficient to meet the current demand. This, then, places more dependence upon the producers of powder to supply the required technical guidance and services to their customers. This is particularly true for the smaller shops and where a producer is attempting to introduce a new grade of powder.

ADDENDUM TO APPENDIX E

INTERNATIONAL TECHNICAL SERVICES LIMITED

31st March, 1971

The Director,
The Australian Mineral Development Laboratories,
Conyngham Street,
FREWVILLE. S.A. 5063

Attention: Mr J. Gooden

Dear Sir,

Export Markets - Iron and Copper Powders.

Thank you for your letter of 12th March, 1971, acknowledging receipt of our preliminary appraisal of export markets for Iron and Copper powders.

We have since received replies from the Trade Commissioners in Peru, Sweden, Iran, Malaysia and Lebanon.

Photo-copies of their replies and enclosed statistics are attached.

Yours faithfully,
INTERNATIONAL TECHNICAL SERVICES LIMITED,

John Pash,
Senior Consultant.

OFFICE OF SENIOR AUSTRALIAN GOVERNMENT TRADE COMMISSIONER

Australian Embassy, Stockholm, Sweden

5th March 1971

Mr J.H. Pash,
International Technical Services Limited,
Manhattan Centre,
107 Walker Street,
NORTH SYDNEY. 2060
Australia.

Dear Mr Pash,

Thank you for your letter of January 6th, 1971 concerning copper and copper alloy powders.

I regret that I have been unable to reply earlier to your enquiry but the information I required from prospective Scandinavian consumers of the above mentioned commodities was not available until recently.

(a - b) Total imports into Sweden of copper classified by type and usage and values of these imports (your items a - b), are shown in the attached photo-copies, tables 1 - 2. A translation of all the Swedish headings and sub-headings appears in tables 3 - 5.

(c) The present import duties are shown in tables 3 - 5.

(d) Interviews have shown that there is only one enterprise in Scandinavia using copper powder in its production, i.e. Gränges S M AB. Gränges S M AB were earlier known as Svenska Metallverken. In late 1969 they merged with the world-wide known company, Grängesbergsbolaget, now called Gränges.

Their technical manager, Mr L. Pantzerhielm, Gränges S M, 721 88 Västeras, who has just returned from a world trip, has informed me that they use copper powder in steadily increasing quantities, exclusively in connection with their production of bearings.

The copper powder is delivered containing a lubricant and other components in accordance with a special formula thus enabling them to press the coatings onto the bearings in one process, instead of two as previously.

Special mixtures of this type have been processed by e.g. Berks, England; Amax, USA; N.A. and Deutscher Affinaderie of Germany. Mr Pantzerhielm also said that they get copper powder from Alcan Aluminium in Canada which owns 20% of Gränges S M.

Finally, Mr Pantzerhielm stated that prospects for suppliers, other than those mentioned above, to deliver copper, ferrous or other powders, at present must be considered as non-existent.

Should you require further technical information regarding copper powder and its usage, Mr Pantzerhielm will be pleased to assist you.

Yours sincerely

K.F. McKernan
Senior Trade Commissioner

Stat.nr	Varuposter	Imports		Exports	
		Införsel		Utförsel	
		Kvant	Tkr 000	Kvant	Tkr 000
		Tons	Swe Kro	Tons	Swe Kro
690	— — andra	51	450	367	7 302
	— andra slag:				
	— — emaljerade:				
	— — — av plåt:				
910	— — — — badkar	494	1 333	1 450	4 488
921	— — — — andra	39	176	171	1 020
929	— — — av annat material	1	3	3	18
990	— — andra	291	911	124	773
73.39.000	**Stålull, putsbollar, putsvantar och liknande artiklar av järn eller stål, för skurning, polering e d (697.9.1)**	318	1 015	13	91
73.40	**Andra varor av järn eller stål:**				
200	gjutjärnsgods, obearbetat (679.1.0)	5 090	4 372	1 545	3 061
300	stålgjutgods, obearbetat (679.2.0)	815	1 999	365	1 716
400	smidda varor, ej vidare bearbetade (679.3.0)	27	114	1 111	2 857
900	andra slag (698.9.1)	19 990	84 575	15 771	88 871
	74 KAP. *Koppar*				
74.01	**Kopparskärsten; koppar, obearbetad, även raffinerad; samt avfall och skrot av koppar:**				
100	kopparskärsten (283.1.2)	25 039	20 379	1 800	1 853
	obearbetad koppar:				
150	— oraffinerad (682.1.1)	8 759	59 110	—	—
	— raffinerad (682.1.2):				
210	— — olegerad	62 347	446 035	33 818	252 395
	— — legerad:				
301	— — — mässing	927	4 226	1 800	8 197
309	— — — bronser och andra kopparlegeringar	2 080	13 575	3 555	21 460
	avfall och skrot (284.0.2):				
601	— av olegerad koppar	346	2 876	32	196
603	— av legerad koppar	12 567	30 947	550	2 257
74.02.000	**Koppartillsatslegeringar (682.1.3)**	240	2 404	1	8
74.03	**Stång och tråd av koppar (682.2.1):**				
	av olegerad koppar:				
100	— s k valstråd	845	6 546	14 456	109 709
	— annan stång och tråd:				
	— — tråd:				
201	— — — i ringar, på spolar o d	440	3 606	677	5 305
209	— — — annan	21	216	41	315
303	— — stång, annan än profilstång	626	4 845	536	4 573
309	— — profilstång	12	105	74	740
	av legerad koppar:				
500	— s k valstråd	312	2 958	1	5
	— annan stång och tråd:				
	— — tråd:				
	— — — av mässing:				
602	— — — — i ringar, på spolar o d	709	6 514	181	1 190
603	— — — — annan	144	986	9	63
	— — — av bronser eller andra kopparlegeringar:				
607	— — — — i ringar, på spolar o d	803	9 472	25	207
608	— — — — annan	36	548	3	46
	— — stång, annan än profilstång:				
703	— — — av mässing	3 525	18 157	1 883	10 159
706	— — — av bronser eller andra kopparlegeringar	390	3 602	82	677
	— — profilstång:				
707	— — — av mässing	117	822	107	711
709	— — — av bronser eller andra kopparlegeringar	18	268	5	64
74.04	**Plåt och band av koppar (682.2.2):**				
	av olegerad koppar:				
101	— släta, rätvinkliga, ej vidare bearbetade	2 302	17 752	2 432	19 683
105	— rondeller och ovaler	99	817	65	583
109	— andra slag	43	411	1	33
	av mässing:				
201	— släta, rätvinkliga, ej vidare bearbetade	1 471	10 199	2 492	15 980
203	— rondeller och ovaler	59	433	223	1 327
205	— andra slag	39	521	14	115
209	av bronser och andra kopparlegeringar	160	1 841	233	1 971
74.05	**Folier av koppar, även mönsterpressade, tillskurna, perforerade, försedda med överdrag, tryckta eller fästade på papper, papp, plast eller dylikt förstärkningsmaterial, med en tjocklek av högst 0 15 mm (förstärkningsmaterial ick inräknat) (682.2.3):**				
001	av olegerad koppar	375	6 376	5 348	42 111
002	av mässing	237	1 696	1 269	8 083
009	av andra kopparlegeringar	253	2 400	3	43
74.06	**Pulver och fjäll av koppar (682.2.4):**				
001	användbara till bronsering	28	380	214	2 176
009	andra slag	291	2 699	1	21

Stat.nr	Varuposter	Införsel		Utförsel	
		Kvant	Tkr	Kvant	Tkr
74.07	**Rör (hålade rörämnen och ihålig stång härunder inbegripna) av koppar (682.2.5):**				
	av olegerad koppar:				
	— av rund genomskärning:				
401	— — i ringar, på spolar o d	195	1 762	432	3 821
102	— — andra	3 136	24 593	3 009	25 848
109	— av annan genomskärning	89	924	3	102
	av mässing, annan än specialmässing:				
203	— av rund genomskärning	720	5 668	675	5 253
205	— av annan genomskärning	269	1 507	42	287
207	av specialmässing	1 177	9 512	58	688
209	av bronser eller andra kopparlegeringar	379	4 392	358	3 337
74.08	**Rördelar av koppar, såsom skarvrör, knärör, muffar och flänsar (682.2.6):**				
100	av olegerad koppar	296	5 811	23	484
201	av mässing	657	11 168	105	2 281
209	av bronser eller andra kopparlegeringar	56	1 071	18	208
74.09.000	**Cisterner, kar och liknande behållare av koppar, för alla slags ämnen, med en rymd av mer än 300 liter, även med inre beklädnad eller med värmeisolerande beklädnad men utan maskinell utrustning och utan anordning för uppvärmning** (692.1.2)	0	0	—	—
74.10	**Linor, kablur, flätade band och liknande varor av koppartråd, utan elektrisk isolering (693.1.2):**				
100	av olegerad koppar	290	3 016	89	813
200	av legerad koppar	46	615	33	450
74.11	**Duk, nät och galler, även ändlösa, av koppartråd** (693.3.2):				
100	ändlös duk för pappersmaskiner (viror)	267	9 198	718	24 338
900	andra slag	15	499	9	290
74.12.000	**Klippnät av koppar** (693.4.2)	5	79	12	428
74.13.000	**Kedjor och delar därtill av koppar** (698.8.1)	24	543	0	9
74.14.000	**Spik och stift, häftstift härunder inbegripna, märlor och krampor samt spetsade krokar och öglor, av koppar eller av järn eller stål och med huvud av koppar** (694.1.2)	12	204	9	146
74.15	**Bultar och muttrar, ogängade eller gängade, skruvar, skruvkrokar, skruvöglor, nitar (andra än rör- och tvåspetsnitar) sprintar, kilar och liknande artiklar av koppar; underläggsbrickor (låsbrickor och fjäderbrickor härunder inbegripna) av koppar (694.2.2):**				
	skruvar och gängade bultar, även med tillhörande muttrar:				
110	— träskruvar	129	1 287	44	518
190	— andra	196	2 096	27	263
900	andra slag	121	1 760	41	671
74.16.000	**Fjädrar av koppar** (698.6.2)	1	109	—	—
74.17	**Kokapparater och andra uppvärmningsapparater av koppar, icke elektriska, av sådana slag, som huvudsakligen användes i hushåll, ävensom delar därtill, av koppar (697.1.2):**				
001	kokapparater samt delar därtill	9	164	622	10 606
009	andra slag	15	470	2	144
74.18	**Hushålls- och sanitetsartiklar samt delar därtill, av koppar (697.2.2):**				
001	förgyllda eller försilvrade	23	1 003	123	3 216
009	andra	127	2 975	30	858
74.19	**Andra varor av koppar (698.9.2):**				
100	gjutna eller smidda ämnen, obearbetade	12	124	3	44
200	fönsterbågar och fönsterkarmar, dörrar och dörrkarmar	9	73	1	23
901	knappnålar och säkerhetsnålar	9	179	2	29
909	andra varor	517	9 748	154	2 755
	75 KAP. *Nickel*				
75.01	**Skärsten, speis och andra mellanprodukter vid framställning av nickel; nickel, obearbetad (med undantag av anoder, hänförliga till nr 75.05), samt avfall och skrot av nickel:**				
100	skärsten, speis och andra mellanprodukter (283.2.2)	1 280	15 959	—	—
	obearbetad nickel (683.1.0):				
200	— olegerad	12 561	175 882	160	4 355
300	— legerad	459	5 005	179	1 354
	avfall och skrot (284.0.3):				
601	— av olegerad nickel	85	2 275	12	159
602	— av legerad nickel	3 305	27 123	493	2 659
75.02	**Stång och tråd av nickel (683.2.1):**				
001	av olegerad nickel	51	997	2	13
002	av legerad nickel	384	6 389	782	18 412

73.40		Other articles of iron or steel :	
	200	Of cast iron, unworked	8 % E
	300	Of cast steel, unworked	8 % E
	400	Of forged steel, not subsequently worked	8 % E
	900	Other	8 % E

CHAPTER 74. — *Copper and articles thereof.*

Notes. — 1. For the purposes of heading No. 74.02, the expression « master alloys » means alloys (except copper phosphide [phosphor copper] containing more than 8 % by weight of phosphorus) of copper with other materials in any proportion, not usefully malleable and commonly used as raw material in the manufacture of other alloys or as de-oxidants, de-sulphurising agents or for similar uses in the metallurgy of non-ferrous metals.

(Copper phosphide [phosphor copper] containing more than 8 % by weight of phosphorus falls within heading No. 28.55 and not within this Chapter.)

2. In this Chapter the following expressions have the meanings hereby assigned to them :

(*a*) *Wire* (heading No. 74.03) :

Rolled, extruded or drawn products of solid section of any cross-sectional shape, of which no cross-sectional dimension exceeds 6 millimetres;

(*b*) *Wrought bars, rods, angles, shapes and sections* (heading No. 74.03) :

Rolled, extruded, drawn or forged products of solid section, of which the maximum cross-sectional dimension exceeds 6 millimetres and which, if they are flat, have a thickness exceeding one tenth of the width.

Also cast or sintered products, of the same forms and dimensions, which have been subsequently machined (otherwise than by simple trimming or de-scaling);

(*c*) *Wrought plates, sheets and strip* (heading No. 74.04) :

Flat-surfaced, wrought products (coiled or not), of which the maximum cross-sectional dimension exceeds 6 millimetres, and of which the thickness exceeds 0.15 millimetre but does not exceed one tenth of the width.

Heading No. 74.04 is to be taken to apply, *inter alia*, to such products, whether or not cut to shape, perforated, corrugated, ribbed, channelled, polished or coated, provided that they do not thereby assume the character of articles or of products falling within other headings.

3. Heading No. 74.07 is to be taken to apply, *inter alia*, to tubes, pipes and hollow bars which have been polished or coated, or which have been shaped or worked, such as bent, coiled, threaded, drilled, waisted, cone-shaped or finned. Heading No. 74.08 is to be taken to apply, *inter alia*, to tube and pipe fittings which have been similarly treated.

74.01		Copper matte; unwrought copper (refined or not); copper waste and scrap :	
	*100	Copper matte	Free E
		Unwrought copper :	
		Unalloyed :	
	210	Refined	Free E
	290	Other (blister and cement copper)	Free E
		Alloyed :	
	301	Ordinary brass	Free E
	305	Special brass	Free E
	309	Bronze and other copper alloys	Free E
		Waste and scrap :	
	400	Of unalloyed copper	Free E
	500	Of copper alloy	Free E
74.02	000	Master alloys	Free E
74.03		Wrought bars, rods, angles, shapes and sections, of copper; copper wire :	
		Of unalloyed copper :	
	100	Wire rod	3 % E
		Other bars, rods, angles, shapes and sections, and wire :	
		Wire :	
	201	In coils and the like	3 % E
	209	Other	3 % E

---

* For statistical purposes, it is necessary to declare the weight in tons and value in Kronor.

74.03	(*Continued*) :	
	Bars and rods, other than angles, shapes and sections:	
301	In coils and the like	3 % E
302	Other	3 % E
309	Angles, shapes and sections	3 % E
	Of copper alloy :	
	Wire rod :	
501	Of brass	3 % E
509	Of other copper alloys	3 % E
	Other bars, rods, angles, shapes and sections, and wire :	
	Wire :	
	Of ordinary brass :	
602	In coils and the like	3 % E
603	Other	3 % E
605	Of special brass	3 % E
609	Of bronze and other copper alloys	3 % E
	Bars and rods, other than angles, shapes and sections :	
	Of ordinary brass :	
701	In coils and the like	3 % E
702	Other	3 % E
705	Of special brass	3 % E
706	Of bronze or other copper alloys	3 % E
	Angles, shapes and sections :	
707	Of ordinary brass	3 % E
708	Of special brass	3 % E
709	Of bronze or other copper alloys	3 % E
74.04	Wrought pates, sheets and strip, of copper :	
	Of unalloyed copper :	
101	Plane, simply cut to square or rectangular shape but not otherwise worked	3 % E
105	Of circular or oval shape	3 % E
109	Other	3 % E
	Of ordinary brass :	
201	Plane, simply cut to square or rectangular shape but not otherwise worked	3 % E
203	Of circular or oval shape	3 % E
205	Other	3 % E
207	Of special brass	3 % E
209	Of bronze and other copper alloys	3 % E
74.05	Copper foil (whether or not embossed, cut to shape, perforated, coated, printed, or backed with paper or other reinforcing material), of a thickness (excluding any backing) not exceeding 0.15 mm. :	
001	Of unalloyed copper	3.4 % E
002	Of ordinary brass	3.4 % E
009	Of other copper alloys	3.4 % E
74.06	Copper powders and flakes :	
100	Bronzing powders and flakes	6 % E
900	Other	Free E
74.07	Tubes and pipes and blanks therefor, of copper; hollow bars of copper :	
	Of unalloyed copper :	
	Of circular cross-section :	
101	In coils and the like	3 % E
102	Other	3 % E
109	Of other cross-section	3 % E
	Of ordinary brass :	
	Of circular cross-section :	
201	In coils and the like	3 % E
202	Other	3 % E

74.07		*(Continued)*:		
	205	Of other cross-section		3 % E
	207	Of special brass		3 % E
	209	Of bronze and other copper alloys		3 % E
74.08		Tube and pipe fittings (for example, joints, elbows, sockets and flanges), of copper:		
	100	Of unalloyed copper	4.8	~~8~~ % E
	201	Of ordinary brass	4.8	~~8~~ % E
	205	Of special brass	4.8	~~8~~ % E
	209	Of bronze and other copper alloys	4.8	~~8~~ % E
74.09	000	Reservoirs, tanks, vats and similar containers, for any material, of copper, of a capacity exceeding 300 litres, whether or not lined or heat-insulated, but not fitted with mechanical or thermal equipment	4.8	~~8~~ % E
74.10		Stranded wire, cables, cordage, ropes, plaited bands and the like, of copper wire, but excluding insulated electric cables:		
	100	Of unalloyed copper wire	3.6	~~8~~ % E
	200	Of copper alloy wire	3.6	~~8~~ % E
74.11		Gauze, cloth, grill, netting, fencing, reinforcing fabric and similar materials (including endless bands), of copper wire:		
	100	Endless bands of wire cloth for paper-making machines	4.8	~~8~~ % E
		Other:		
	901	Of unalloyed copper	4.8	~~8~~ % E
	905	Of ordinary brass	4.8	~~8~~ % E
	909	Of other copper alloys	4.8	~~8~~ % E
74.12	000	Expanded metal, of copper	4.8	~~8~~ % E
74.13	000	Chain and parts thereof, of copper	4.8	~~8~~ % E
74.14		Nails, tacks, staples, hook-nails, spiked cramps, studs, spikes and drawing pins, of copper, or of iron or steel with heads of copper		
	001	Drawing pins	4.4	~~6~~ % E
	009	Other	4.4	~~6~~ % E
74.15		Bolts and nut (including bolt ends and screw studs), whether or not threaded or tapped, and screws (including screw hooks and screw rings), of copper; rivets, cotters, cotter-pins, washers and spring washers, of copper:		
		Screws and threaded bolts, with or without nuts:		
	110	Wood screws	4.4	~~6~~ % E
	190	Other	4.4	~~6~~ % E
	900	Other	4.4	~~6~~ % E
74.16	000	Springs, of copper	4.8	8 % E
74.17		Cooking and heating apparatus of a kind used for domestic purposes, not electrically operated, and parts thereof, of copper:		
	100	Cooking apparatus and parts thereof	4.8	~~8~~ % E
	900	Other	4.8	~~8~~ % E
74.18		Other articles of a kind commonly used for domestic purposes, builders' sanitary ware for indoor use, and parts of such articles and ware, of copper:		
	100	Gilt or silvered	4.8	~~8~~ % E
	900	Other	4.8	~~8~~ % E
74.19		Other articles of copper:		
	100	Unworked cast or forged blanks	4.8	~~8~~ % E
	200	Doors, window frames and other similar metal structures	4.8	~~8~~ % E
	410	Ordinary pins and safety pins	6	10 % E
	900	Other	4.8	8 % E

CHAPTER 75. — *Nickel and articles thereof.*

Notes. — 1. In this Chapter the following expressions have the meanings hereby assigned to them:

(*a*) *Wire* (heading No. 75.02):

Rolled, extruded or drawn products of solid section of any cross-sectional shape, of which no cross-sectional dimension exceeds 6 millimetres;

AUSTRALIAN GOVERNMENT TRADE COMMISSION
MANILA

Australian Embassy
7th Floor, L & S Bldg.
1414 Roxas Blvd.,
PO Box 1948
Manila. Philippines

25th February 1971

Mr J.H. Pash
Senior Consultant
International Technical Services Limited
Manhattan Centre
107 Walker St
North Sydney, N.S.W. 2060

Dear Mr Pash,

Further to my letter of January 18, 1971, herewith find the figures on imports of iron or steel powders and copper and flakes for 1968, 1969 and the first half of 1970.

Yours sincerely,

F.W. Gluth
Trade Commissioner
MANILA

KWG/mmm

PHILIPPINE IMPORT STATISTICS

	1968		1969		January to June 1970	
	Quantity	FOB Value	Quantity	FOB Value	Quantity	FOB Value
	Kilo	US$	Kilo	US$	Kilo	US$
Iron and Steel Powder	209526	40819	404983	70598	37613	6579
United States	68695	11230	122401	18675	27543	4210
Sweden	36000	5155	156000	23919	-	-
Japan	104831	24434	119748	26239	9979	2239
United Kingdom	-	-	30	7	-	-
Hong Kong	-	-	6804	1758	-	-
Australia	-	-	-	-	91	130
Copper Powder and flakes	2432	6787	5813	13322	1002	1655
United States	118	560	-	-	2	5
Germany	695	1537	2800	6679	1000	1650
United Kingdom	1609	4641	759	1942	-	-
Japan	10	49	454	687	-	-
Belgium	-	-	1300	2547	-	-
Netherlands	-	-	500	1467	-	-

OFFICE OF COMMERCIAL COUNSELLOR AND
AUSTRALIAN GOVERNMENT TRADE COMMISSIONER
BEIRUT

9th February, 1971

Mr J.H. Pash
Senior Consultant,
International Technical Services Ltd,
Manhattan Centre,
107 Walker Street,
North Sydney. 2060

Dear Mr Pash

Iron and Ferrous-Alloy Powders - Lebanon

Thank you for your letter dated the 5th January, 1971.

In reply to your request, I regret to inform you that the quantity and value of imported iron and copper powder in 1968 (the latest available) were very small. Figures were:

Iron or Steel Powder:

W. Germany -	5 Kgs.	LL. 75
U.K. -	32 Kgs.	LL. 120
	37 Kgs.	LL. 195

Copper Powder:

W. Germany -	1310 Kgs.	LL. 10.345	
Italy -	290 Kgs.	LL. 3.345	
	1600 Kgs.	LL. 13.690	(A$ 1 = LL. 3.56 approx.)

No figures are available on ferrous-alloy powder.

The above commodities are exempted from customs duty, but are subject up to 3.5% ad valorem municipal tax.

Yours sincerely,

I.M. Haig
Australian Government
Trade Commissioner

AUSTRALIAN GOVERNMENT TRADE COMMISSION
Buenos Aires, Argentina

24th March, 1971

Mr J.H. Pash
Senior Consultant
International Technical Services Ltd.,
Manhattan Centre
107 Walker Street,
North Sydney. N.S.W. 2060

Dear Sir,

I acknowledge receipt of your letter dated 5th January, 1971, requesting information on the possibilities of finding export markets in Argentina for iron and ferrous-alloy powders.

Argentina is not a producer of ferrous or copper powders; these products are imported and, according to available figures, imports have been as follows:

Iron or Steel Powers:

a Powder for separating seeds electromagnetically
13 tons - u$s 9,690

b Other powders of iron or steel
803 tons - u$s 225,754

No import duty is applied for the importation of these mineral powders.

The principal supplier is U.S.A. (70%), followed by Germany (15%) and Italy (10%).

A local company, which imports these powders mentioned that consumption would increase in the near future with national production growth.

If you wish to contact this importer please let us know as he will be only too glad to advise you on this subject.

Yours faithfully,

(Hugh P. Thompson)
Australian Government
Trade Commissioner
BUENOS AIRES.

AUSTRALIAN EMBASSY

LIMA

March 3rd., 1971

Mr J.H. Pash
Senior Consultant
International Technical Services Ltd
Manhattan Centre
107 Walker Street
North Sydney 2060
Australia

Dear Sir

We received your letter of January 5th, in which you asked us to provide information relating to Peru regarding iron and ferrous-alloy powders to assist your investigations on this product.

Please find below the information required.

Imports 1968

(The last year for which statistics are published)

Ferro Alloys

Duties GK $0.20 30% Ad Val.

		Kilos	Value $A
73.02	Ferro Alloys		
0.01	Ferro Manganese		
	N. Germany	8,070	2,705
	Norway	15,870	4,461
	U.K.	319	66
	Rep. of South Africa	651,368	86,765
		675,627	93,997
0.02	Ferro Chromium		
	France	1,970	1,455
	U.K.	639	285
	Rep. of South Africa	179,689	44,348
	Sweden	1,080	321
	Switzerland	678	943
		184,056	47,352

0.04	Ferro Silicon	Kilos	Value $A
	Denmark	505	187
	U.S.	46,706	10,095
	Norway	4,488	983
	U.K.	6,736	1,103
	South Africa	356,990	58,769
	Switzerland	10.088	2,253
		425,513	73,660

Future trends in local use of this product seem to be promising. It is estimated that Sogesa, main producing steel plant in Peru, will increase its production 30% in 1971 and that their requirements will amount for more than 1,200 tons of Ferro-Manganese and 800 tons Ferro-Silicon.

Price is the determining factor in placing orders with South Africa and it is suggested that this important fact be taken into account when making the appraisal of this market.

Yours faithfully,

R.F. Shaw
Australian Government
Trade Commission
LIMA

OFFICE OF THE COMMERCIAL COUNSELLOR
AND
AUSTRALIAN GOVERNMENT TRADE COMMISSIONER

Australian Embassy
23 Avenue Arak,
Tehran
Iran

February 9, 1971

Mr J.H. Pash
International Technical Service Ltd
Manhattan Centre
107 Walker Street
North Sydney. 2060
Australia

Dear Sir

Iron and Ferrous-Alloy Powders

We refer to your letter of 5th January 1971, concerning your studies on export markets for ferrous and copper alloy powders in Iran.

The Iranian Tariff Classification does not have a separate item for ferrous and copper powders, hence Iranian import statistics shows no data for them.

However imports of ferrous and copper powder fall under the following Tariff Classification and the relevant duties are shown:

Iranian Tariff Item 698
Iron/Steel Scraps, Chips, Fitting

Customs Duty	0.20 Rials per kilo ($A1 = 85.01 Rls)
Commercial Benefit Tax	0.10 " " "

Iranian Tariff Item 756 - B
Copper shavings, filings, scraps

Customs Duty	6 Rials per kilo
Commercial Benefit Tax	NIL

The following is Iran's combined imports under these two Tariff items during the past 3 years, however, it does not indicate separate figures on imports of ferrous and copper powders:

	Iron and Steel Chips, Scraps, Powder etc.		Copper Shaving, Filing, Powder, Scraps etc.	
	Quantity Kilos	Value $A	Quantity Kilos	C & F Value $A
21.3.67 - 20.3.68	5,170,000	308,173	71,849	28,776
21.3.68 - 21.3.69	6,029,997	538,092	41,962	30,004
21.3.69 - 20.3.70	3,483,376	151,595	55,547	21,367

Listed below for your information are the Iranian imports of ferrous and copper powders shown in a report released by the United States Bureau of International Commerce. This is not, however, an official information:

Iron & Steel Powders

1967	US $	49,000
1968	US $	101,000

Copper Powders, Flakes

1967	US $	9,000
1968	US $	30,000

Yours faithfully,

N.C. Carroll
Trade Commissioner
Tehran.

AUSTRALIAN GOVERNMENT TRADE COMMISSION
KUALA LUMPUR
MALAYSIA

117 Jalan Ampang
PO Box 934
Kuala Lumpur
Malaysia

8th February 1971

Mr J.M. Pash
Senior Consultant,
International Technical Services Ltd
Manhattan Centre,
107 Walker Street,
North Sydney. 2060

Dear Sir,

Iron and Ferrous Alloy Powders

Thank you for your letter of 6th January asking for market information on ferrous and copper powders. We apologise for the delay in replying to your letter.

The following are answers to your questions:

(a) Attached are import figures showing the imports of iron and steel powders, copper powder and flakes and other ferrous-alloys into West Malaysia for the years 1967 to 1969.

(b) See attached.

(c) There is no import duty on these powders. However, there is a 4% surtax applied to all imports, irrespective of country of origin.

(d) There is no local production of these powders in Malaysia and end-users have to import all their requirements from overseas countries. During 1969, total imports of copper powder and flakes and other ferrous-alloys were 311 tons and 285 tons respectively.

The main consumers of ferrous-alloys are steel foundries and casting firms. The market for this powder is expected to increase the next few years in view of the expansion of existing foundries and the implementation of new projects. (e.g. Malayawatta Iron & Steel Works) (Iron and Steel Foundry), Bradkan (M) Sdn. Bhd., (steel foundry complex), Malleable Castings (M) Sdn. Bhd., (castings).

Only a few electro-plating companies reported using copper powder and in very small quantities. We could not locate other users of this powder although the Malaysian import figures show a substantial import for the year 1969.

We trust the foregoing is of assistance to you.

Yours faithfully,

R.W. Holberton
Trade Commissioner
Kuala Lumpur

Attach.

IMPORTS INTO WEST MALAYSIA - BY COUNTRIES

Description of Commodity and Country	January - December 1967		January - December 1968		January - December 1969	
	Quantity	Value M$ CIF	Quantity	Value M$ CIF	Quantity	Value M$ CIF
Iron or Steel Powders BTN No. 73.05.100 (SITC No. 671.320)						
Total Imports	-	-	-	-	2.21	2,678
India	-	-	-	-	0.01	10
Singapore	-	-	-	-	-	3
United Kingdom	-	-	-	-	2.19	2,225
U.S.A.	-	-	-	-	0.01	440
Copper Powder and Flakes BTN No. 74.06.000 (SITC No. 682.240)						
Total Imports	132.73	30,106	187.99	55,318	311.00	78,494
China	6.39	1,167	5.99	798	-	-
Fed. Republic of Germany	71.97	17,218	141.55	39,948	222.69	50,673
Hong Kong	10.80	1,879	6.98	1,459	3.48	753
Japan	0.17	66	3.70	1,193	0.75	314
United Kingdom	37.45	8,886	26.73	9,207	81.94	26,259
Other Ferro-Alloys BTN No. 73.02.900 (SITC No. 671.500)						
Total Imports	192.12	142,484	34.17	19,432	295.10	211,241
Australia	-	-	-	-	1.70	1,234
Formosa	29.12	21,875	-	-	38.44	25,184
Fed. Republic of Germany	-	-	0.30	276	28.86	34,015
India	-	-	30.82	16,994	204.56	130,190
Sweden	-	-	2.00	1,417	2.60	5,379
United Kingdom	5.52	3,855	1.05	745	13.76	11,122

TABLES 1 to 9

FIGURES 1 to 7

TABLE 1: HyL PLANT — DIRECT AND INDIRECT OPERATING COST FOR REDUCTION

Item	Unit Cost, $	Days/annum	Quantity	$/annum
Natural gas	0.30*/thousand cu ft	330	2,639,000 thousand cu ft	790,000
Water make-up	0.25/thousand gal	330	118,000 thousand gal	15,000
Electricity	0.02/kWh	330	1,040 thousand kWh	20,800
Catalyst and chemicals		330		15,000
Operating Labour	1.50/man hour	365 + 10	49,500 man hours	74,300
Supervision	2.50/man hour	365 + 10	9,000 man hours	22,500
Maintenance	4% capital investment	365 + 10		320,000
General overhead	100% labour and supervision			96,800
Miscellaneous supplies				22,000
Royalty	0.90/ton Fe			117,000
Direct Operating Cost	$11.50/ton		130,000 Fe in pellets	1,493,400

* This is assumed as a final negotiated price

TABLE 2: PRODUCTION OF IRON POWDER FROM MOLTEN METAL BY ATOMISATION

	Unit Cost $	Quantity/ton Product	$/ton Product
Electricity	0.02/kWh	150 kWh	3.00
Hydrogen or cracked ammonia	200/ton NH_3	200 cu ft	(say) 0.50
Gas	0.30/thousand cu ft	600 cu ft	0.18
Water	0.20/thousand gal	200 gal	0.05
Sundries (including nitrogen)			0.25
Operating labour	1.50/man hour	8 man hours	12.00
Supervision	2.50/man hour	0.8 man hour	2.00
Maintenance (3% FCI)			7.80
Direct Operating Cost			25.78

TABLE 3: IRON AND STEEL MINI-COMPLEX — ESTIMATED COST OF PRODUCTS AT SALISBURY, AND CAPITAL INVESTMENT

Product	Total Production Cost*				Fixed Capital Investment	
	$/ton Feed	$/ton Concentrate or Pellets	$/ton Fe in Reduced Pellets	$/ton Metal	Stage $ million	Cumulative $ million
Olary ore feed on trucks at Salisbury:						
High grade	8.30	9.70	-	-	nil	nil
Low grade	6.80					
Concentrate	-	12.80	-	-	1.56	1.56
Semi-indurated pellets	-	15.60	24.40	-	2.00	3.56
Reduced pellets (Sponge Iron)	-	-	36.95	-	8.00	11.56
Steel billets	-	-	63.90	67.30	4.00	15.56
Iron powder	-	-	-	144.30	1.3	16.86

* Rounded to nearest $0.05.

TABLE 4: ANACONDA-TREADWELL PROCESS
Manufacturing Cost Data for an 80,000 tpa Plant

		Cost, $\$10^6$ pa
Chemicals		
H_2SO_4:	assumed nil	
HCN:	680 tpa cost at 3 x molar cost of NH_3 ie. $\$160 \times 3 \times \frac{17}{27} = \$303$/ton	0.206
Utilities		
Power:	39.7×10^6 kWh Cost @ 1.5 ¢/kWh	0.595
Natural gas:	$5{,}900{,}000 \times 10^6$ Btu/annum Cost @ 3.5 ¢/therm	2.065
Water:	108×10^6 gal/annum Cost @ 50 ¢/1000 gal	0.054
Labour and Supervision		
The Anaconda-Treadwell estimate allows for 32 shift operators, 32 shift labourers, 8 shift supervisors. 14 day-shift operators, 8 day-shift labourers and 2 day-shift supervisors:		
	32 @ \$4,500 = 144,000	
	32 @ \$3,500 = 112,000	
	14 @ \$4,000 = 56,000	
	8 @ \$3,000 = 24,000	
	8 @ \$6,500 = 52,000	
	2 @ \$7,000 = 14,000	
	Total Labour	0.402
Maintenance		
Assume as 5% pa of capital cost		2.105
Overheads		
Operating supplies:	assume 10% direct labour	
Payroll overhead:	assume 20% direct labour	
Plant overhead:	assume 125% direct labour	
Process control:	assume 25% direct labour	
Total:	180% of direct labour	0.725
	c/f	6.152

Continued

TABLE 4: CONTINUED

		Cost, $\$10^6$ pa
		b/f 6.152
Depreciation		
Assume straight line depreciation at 10% pa		4.210
Taxes and Insurance		
Assume 1% pa of capital cost		0.421
	Sub-Total	10.783
Interest		
Assume 6% pa on total capital cost		2.940
	Total	13.723
	ie.	$\$13.723 \times 10^6$ pa

TABLE 5: ANACONDA-TREADWELL PROCESS
Manufacturing Cost Data for a 10,000 tpa Plant

		Cost, $\$10^6$ pa
Chemicals		
HCN:	1/8 x 0.206	0.026
Utilities		
Power:	1/8 x 0.595	0.075
Natural gas:	1/8 x 2.065	0.259
Water:	1/8 x 0.054	0.007
Labour and Supervision		
Direct labour:	1/6 x 0.402	0.067
Maintenance		
5% pa of capital		0.520
Overheads		
180% of direct labour		0.121
Depreciation		
10% pa of capital		1.040
Taxes and Insurance		
1% pa of capital		0.104
	Sub-Total	2.219
Interest		
6% pa on total capital cost		0.730
	Total	2.949
	ie.	$\$2.949 \times 10^6$ pa

TABLE 6: ZINC OXIDE REDUCTION

Temperature °C	Equilibrium Partial Pressures, atm				
	Zn	CO	CO_2	H_2	H_2O
900	0.269	0.236	0.006	0.467	0.021
950	0.297	0.222	0.012	0.425	0.046
1,000	0.325	0.200	0.023	0.372	0.075
1,050	0.367	0.175	0.034	0.300	0.121
1,100	0.414	0.144	0.051	0.225	0.165

TABLE 7: ZINC SILICATE REDUCTION — THERMOCHEMICAL VALUES FOR REACTION 6

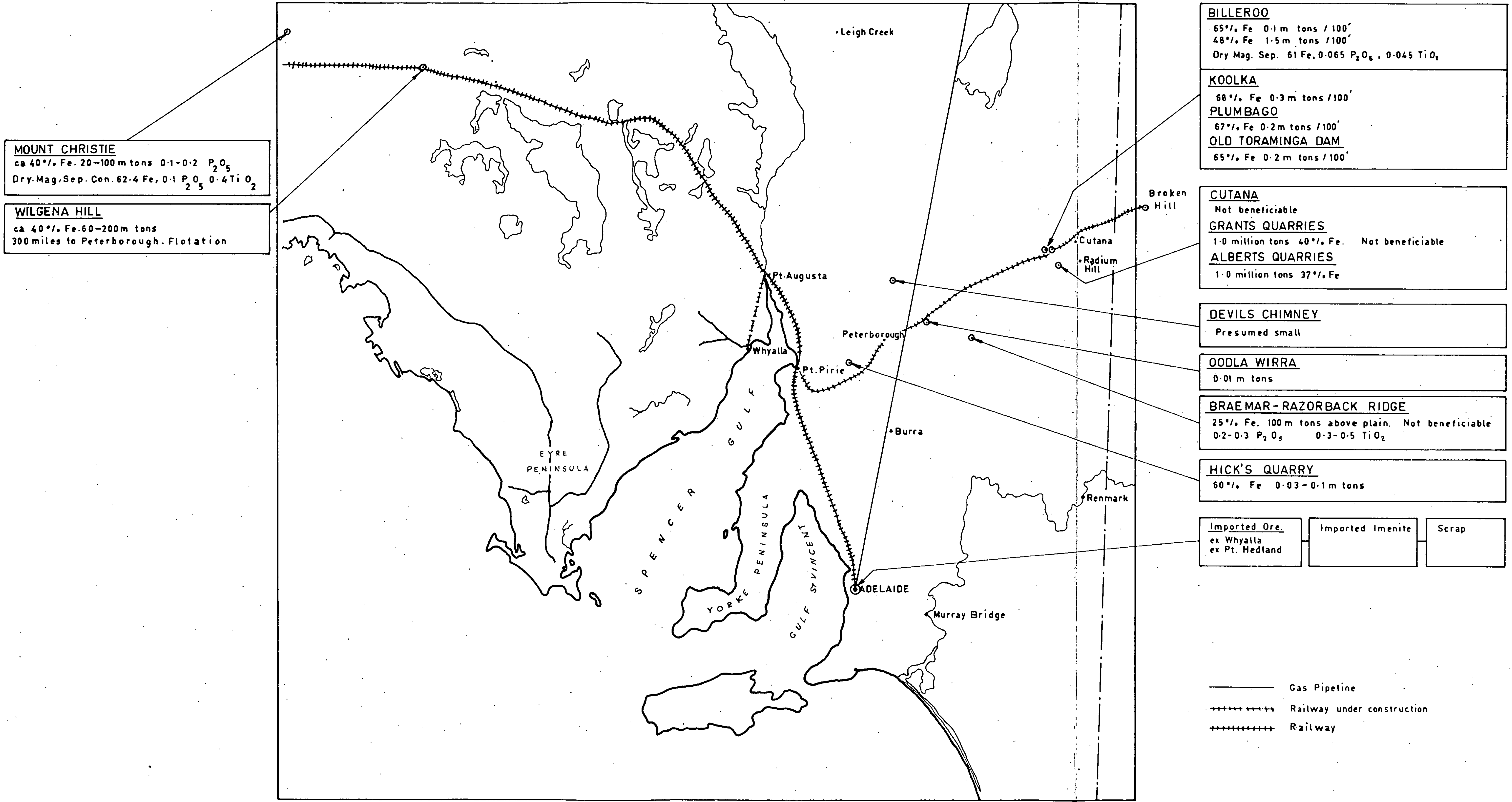
Temp °C	ΔG^o_T kcal	Log Kp	Kp	Equilibrium Gas Partial Pressures, atm			
				CO	Zn	H_2	CH_4
907	+26.4	-4.89	12.88×10^{-6}	0.141	0.141	0.282	0.440
950	+19.1	-3.41	3.89×10^{-4}	0.187	0.187	0.374	0.248
1000	+11.1	-1.91	1.23×10^{-2}	0.240	0.240	0.480	0.038
1050	+ 3.1	-0.51	30.9×10^{-2}	0.244	0.244	0.488	0.025
1100	- 5.2	+0.83	6.76	0.249	0.249	0.498	5.92×10^{-3}
1150	-13.4	+2.06	1.15×10^{2}	0.25	0.25	0.50	1.46×10^{-3}
1200	-21.4	+3.17	14.79×10^{2}	0.25	0.25	0.50	4.07×10^{-4}
1300	-37.6	+5.22	16.6×10^{4}	0.25	0.25	0.50	3.84×10^{-5}
1400	-53.8	+7.03	10.72×10^{6}	0.25	0.25	0.50	4.76×10^{-6}
1500	-69.8	+8.60	3.98×10^{8}	0.25	0.25	0.50	7.8×10^{-7}

TABLE 8: ZINC SILICATE REDUCTION — THERMOCHEMICAL VALUES FOR REACTIONS 7 AND 8

Temp	Reaction 7			Reaction 8		
°C	ΔG^{o}_{T} kcal	Log Kp	Kp	ΔG^{o}_{T} kcal	Log Kp	Kp
907	+62.2	-11.50	3.16×10^{-12}	+60.8	-11.22	6.026×10^{-12}
950	+60.9	-10.87	13.49×10^{-12}	+59.3	-10.59	25.70×10^{-12}
1000	+59.9	-10.28	52.48×10^{-12}	+57.5	- 9.86	1.38×10^{-10}
1050	+58.9	- 9.70	1.995×10^{-10}	+55.9	- 9.21	6.166×10^{-10}
1100	+57.6	- 9.16	6.918×10^{-10}	+53.8	- 8.56	27.54×10^{-10}
1150	+56.4	- 8.65	22.39×10^{-10}	+51.8	- 7.95	1.122×10^{-8}
1200	+55.4	- 8.20	63.10×10^{-10}	+50.0	- 7.40	3.981×10^{-8}
1300	+53.2	- 7.40	3.981×10^{-8}	+46.2	- 6.42	38.02×10^{-8}
1400	+51.2	- 6.70	19.95×10^{-8}	+42.4	- 5.54	2.884×10^{-6}
1500	+49.4	- 6.09	81.28×10^{-8}	+38.8	- 4.78	16.60×10^{-6}

TABLE 9: ZINC SILICATE REDUCTION

Temp	Equilibrium Partial Pressures, atm					
°C	Zn	CH_4	CO	CO_2	H_2	H_2O
907	0.141	0.440	0.141	1.78×10^{-6}	0.282	4.91×10^{-6}
950	0.187	0.248	0.187	3.68×10^{-6}	0.374	1.01×10^{-5}
1000	0.240	0.038	0.240	7.25×10^{-6}	0.480	2.36×10^{-5}
1050	0.244	0.025	0.244	1.41×10^{-6}	0.488	4.96×10^{-5}
1100	0.249	5.92×10^{-3}	0.249	2.64×10^{-5}	0.498	1.05×10^{-4}
1150	0.250	1.46×10^{-3}	0.250	4.71×10^{-5}	0.500	2.12×10^{-4}
1200	0.250	4.07×10^{-4}	0.250	7.95×10^{-5}	0.500	3.99×10^{-4}
1300	0.252	3.85×10^{-5}	0.250	1.99×10^{-4}	0.499	1.23×10^{-3}
1400	0.253	4.72×10^{-6}	0.249	4.38×10^{-4}	0.496	3.32×10^{-3}
1500	0.256	7.51×10^{-7}	0.247	8.68×10^{-4}	0.487	7.75×10^{-3}



MOUNT CHRISTIE
ca 40% Fe. 20–100 m tons 0·1–0·2 P_2O_5
Dry·Mag/Sep. Con. 62·4 Fe, 0·1 P_2O_5 0·4 TiO_2

WILGENA HILL
ca 40% Fe. 60–200m tons
300 miles to Peterborough. Flotation

BILLEROO
65% Fe 0·1 m tons / 100′
48% Fe 1·5 m tons / 100′
Dry Mag. Sep. 61 Fe, 0·065 P_2O_5, 0·045 TiO_2

KOOLKA
68% Fe 0·3 m tons / 100′

PLUMBAGO
67% Fe 0·2 m tons / 100′

OLD TORAMINGA DAM
65% Fe 0·2 m tons / 100′

CUTANA
Not beneficiable

GRANTS QUARRIES
1·0 million tons 40% Fe. Not beneficiable

ALBERTS QUARRIES
1·0 million tons 37% Fe

DEVILS CHIMNEY
Presumed small

OODLA WIRRA
0·01 m tons

BRAEMAR - RAZORBACK RIDGE
25% Fe. 100 m tons above plain. Not beneficiable
0·2–0·3 P_2O_5 0·3–0·5 TiO_2

HICK'S QUARRY
60% Fe 0·03–0·1 m tons

Imported Ore.
ex Whyalla
ex Pt. Hedland

Imported Imenite

Scrap

FIGURE 1: IRON ORE SOURCES

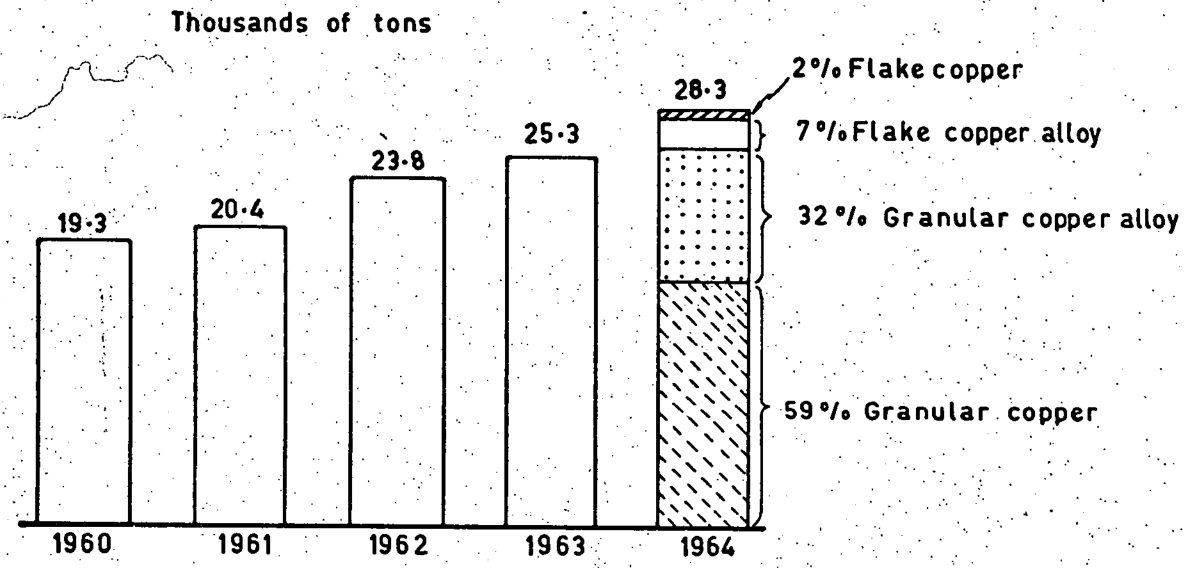


FIGURE 2 : SHIPMENT OF COPPER AND COPPER ALLOY POWDERS (USA)

(Atter Poster 1966)

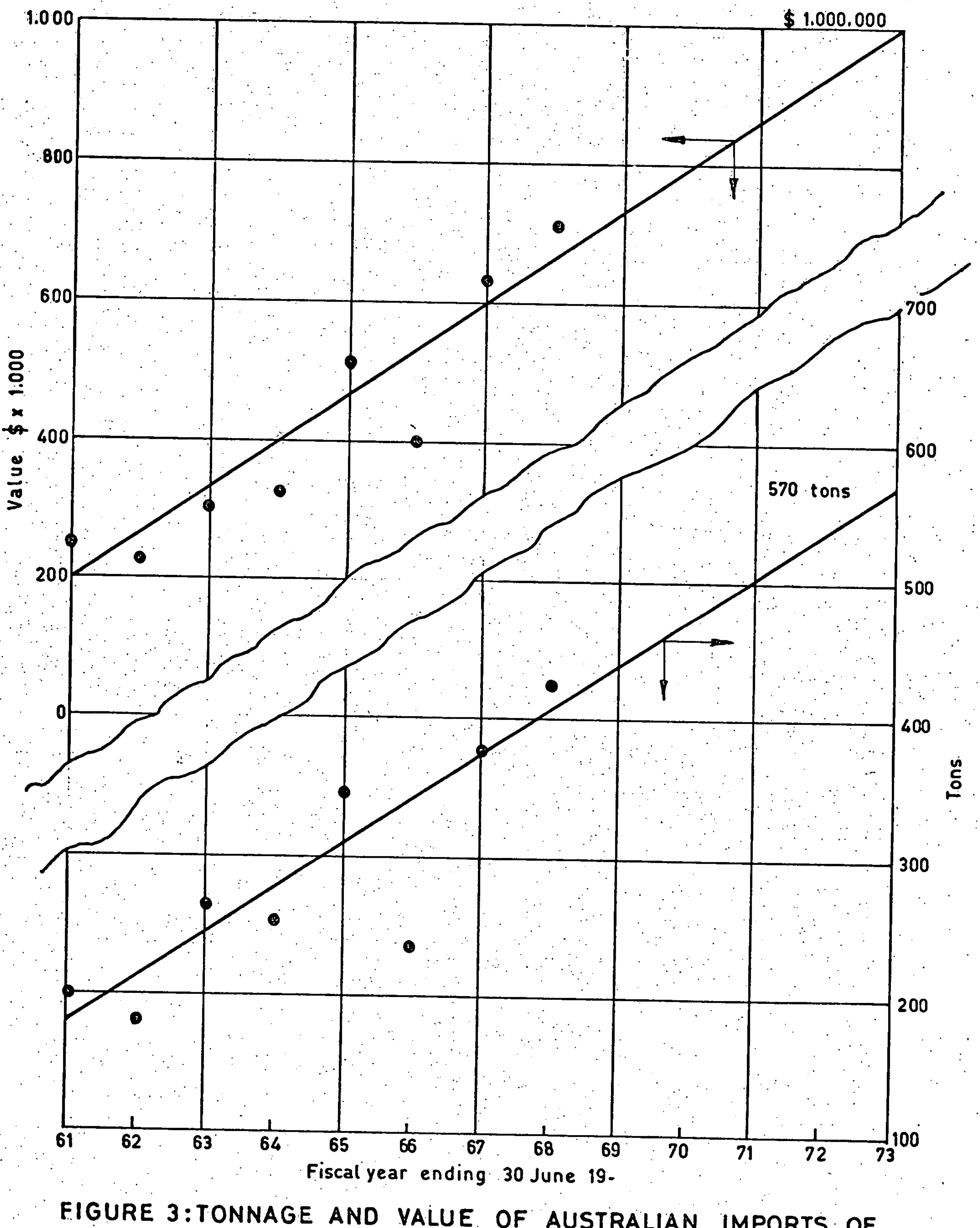


FIGURE 3: TONNAGE AND VALUE OF AUSTRALIAN IMPORTS OF COPPER AND COPPER BASE-ALLOY POWDERS AND FLAKES

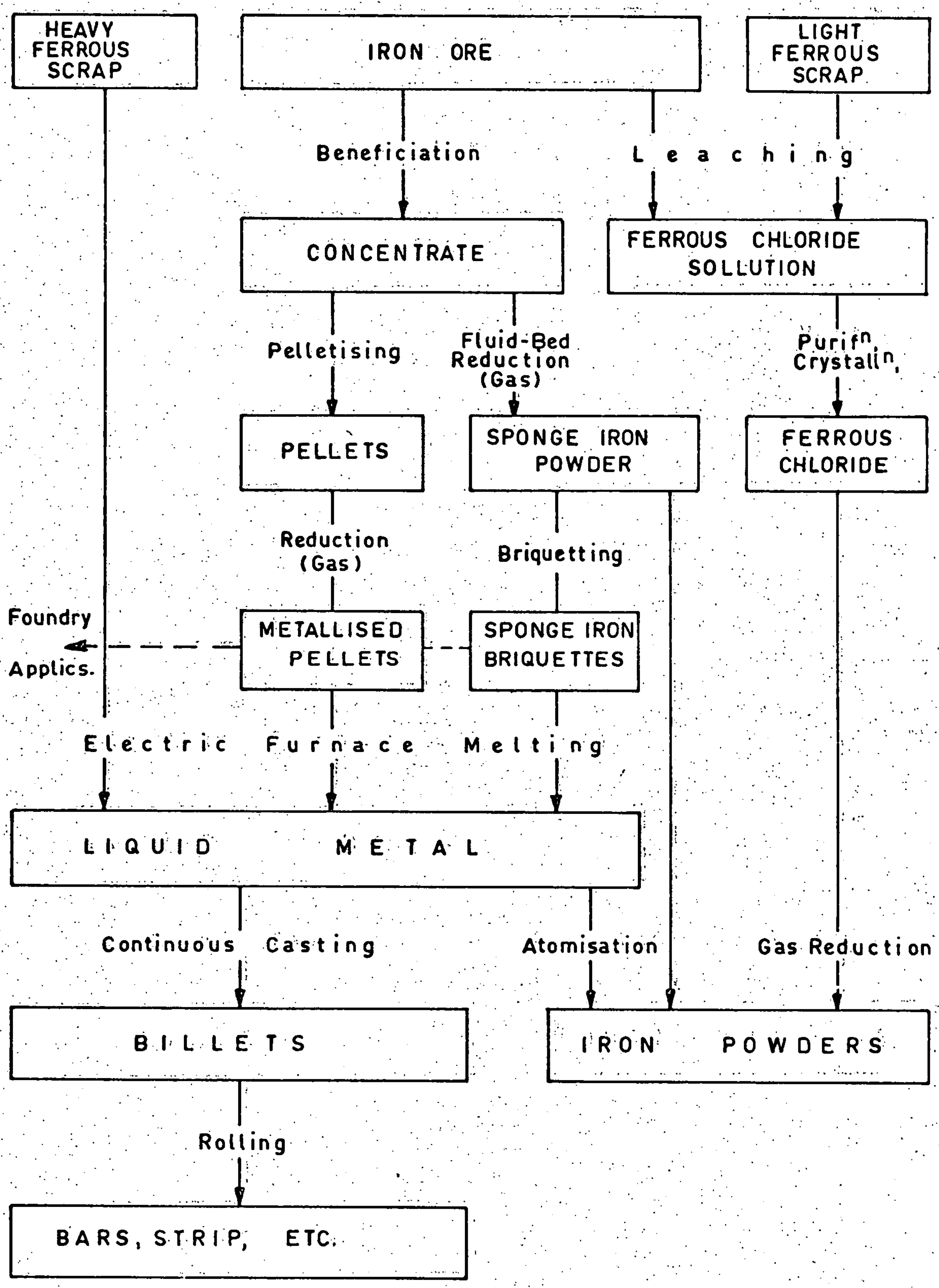


FIGURE 4: GENERALISED ROUTES EXAMINED TO STEEL AND IRON POWDER

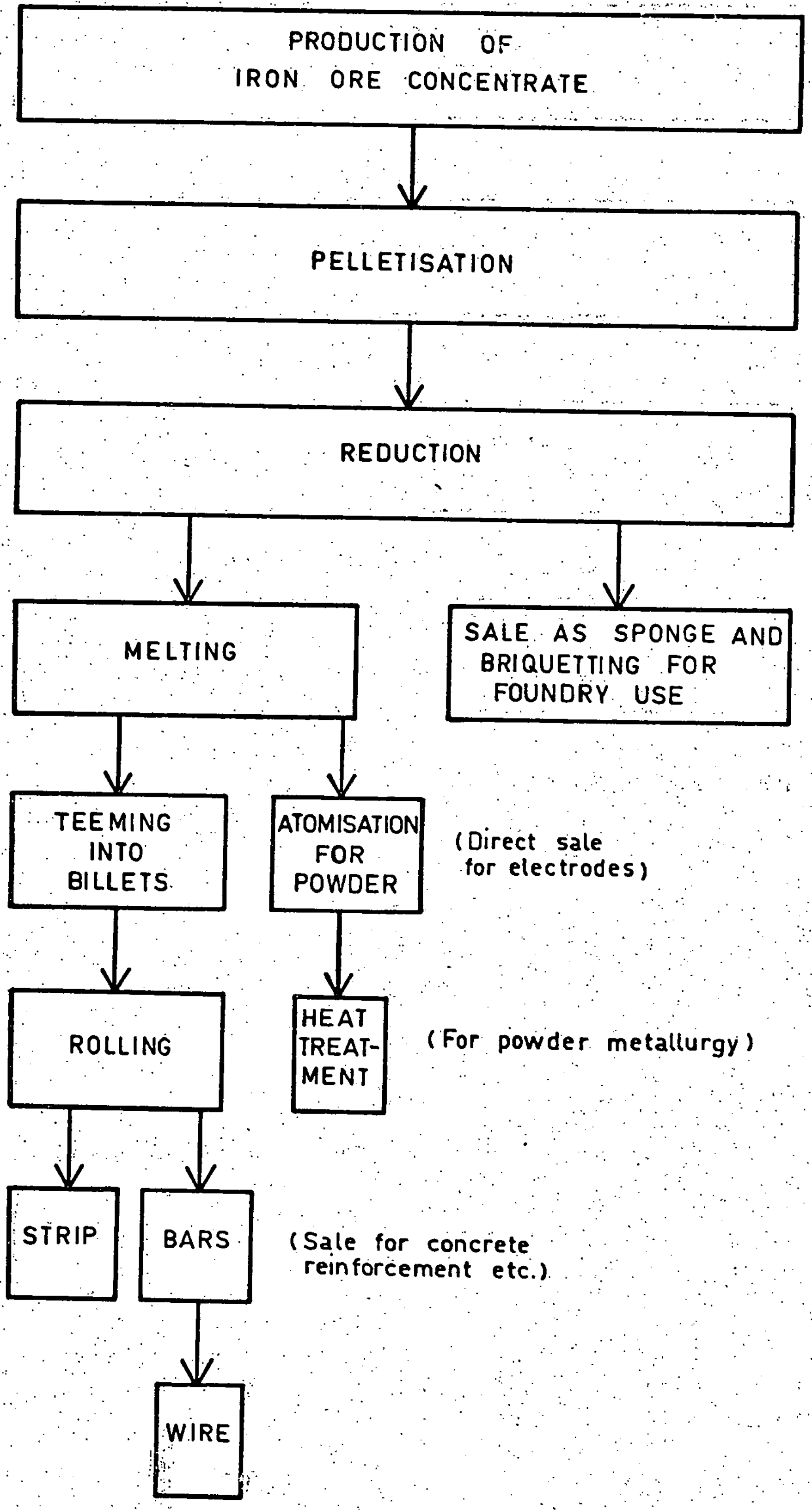


FIGURE 5 : DIAGRAMMATIC REPRESENTATION OF PROCESSING STAGES PROPOSED FOR S.A. OPERATION.

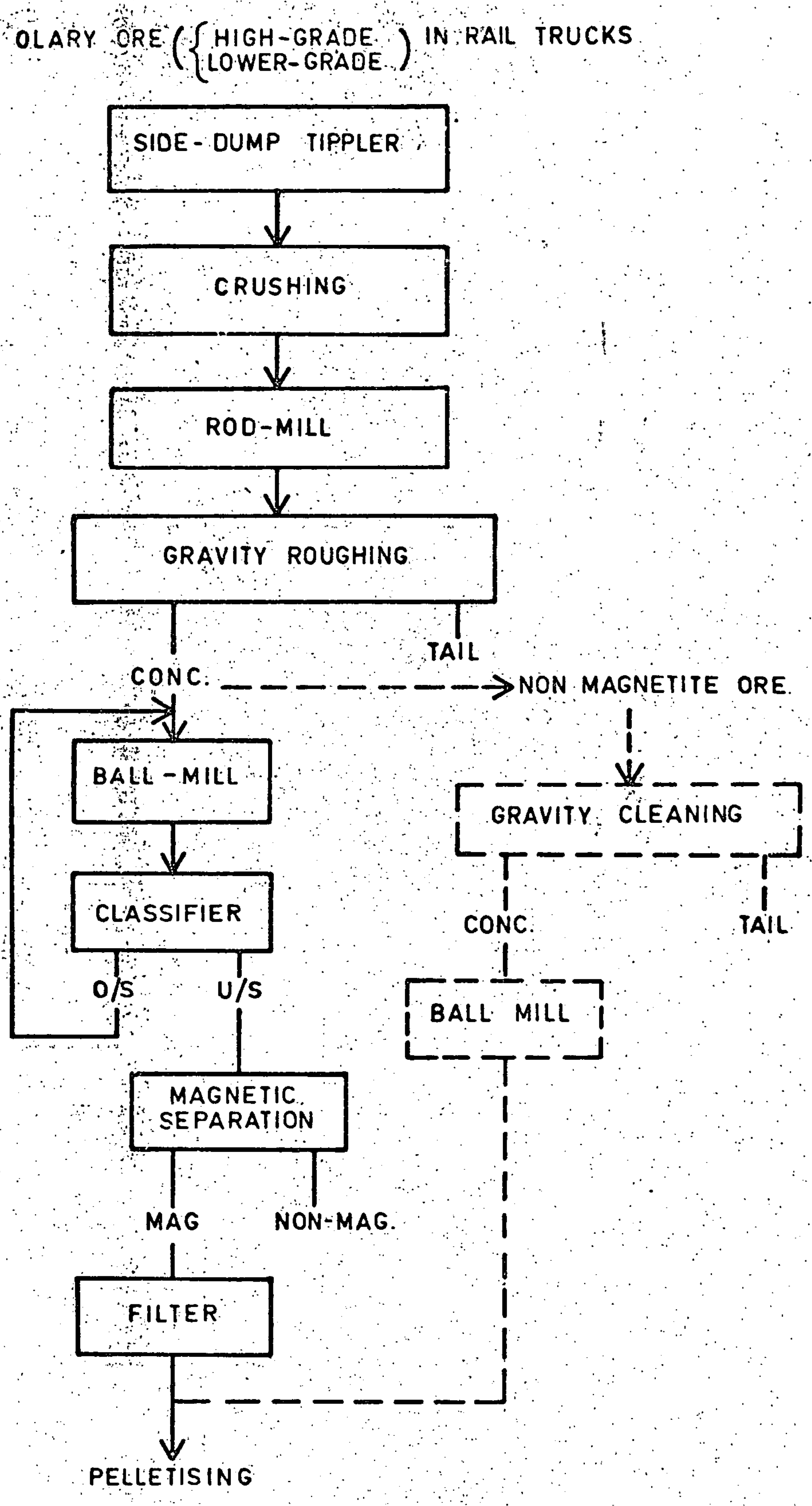
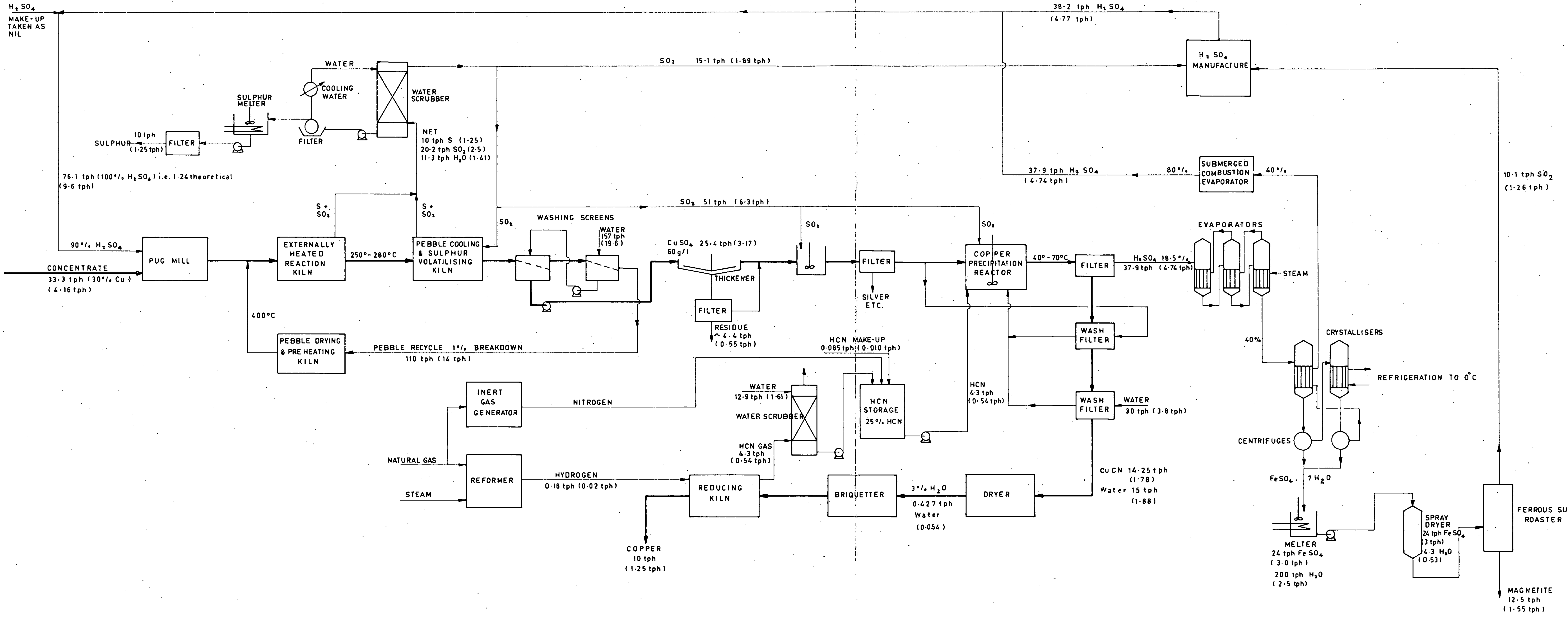


FIGURE 6 : DIAGRAMMATIC FIRST-ESTIMATE FLOWSHEET OF BENEFICIATION PLANT



Legend: Quantities not in parenthesis refer to a plant producing 80,000 tpa copper
Quantities in parenthesis 10,000 tpa ״

FIGURE 7: ANACONDA - TREADWELL PROCESS

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