GEOLOGIE EN MIJNBOUW VOLUME 51 (3), p. 423–448 1972

METALLURGICAL PRACTICE AT THE BROKEN HILL DIVISION OF NCHANGA CONSOLIDATED COPPER MINES LIMITED

B. BARLIN¹)

ABSTRACT

A detailed description is given of the metallurgical complex comprising a concentrator, electrolytic zinc plant, sinter plant, Imperial Smelting Furnace and lead refinery used to produce refined lead and two grades of zinc at the Broken Hill mine in Kabwe, Zambia.

Flow sheets and 1970 production details are included.

INTRODUCTION

On a plateau, 1,180 metres above sea level, in Zambia, Central Africa, are found the deposits of lead, zinc and vanadium ores of the Broken Hill Division of Nchanga Consolidated Copper Mines Limited, formerly the Zambia Broken Hill Development Company Limited. The mine is situated at Kabwe 2,137 kilometres by rail from the nearest port of Beira in Mozambique and about 14 degrees south of the Equator.

The ore deposits discovered by T.G. Davey in 1902 occur as pipe-like bodies within a massive featureless dolomite. At the outcrop, the orebodies are in the main, completely oxidised whilst at depth they contain a core of massive sulphide ore surrounded by an envelope of zinc silicate ore containing some oxidised lead minerals. Between the zinc silicate and the dolomite walls is a zone of more or less broken ground containing a hard clay with vanadium minerals. The main ore minerals are galena, sphalerite, willemite and cerussite with minor amounts of smithsonite, hemimorphite, anglesite, pyromorphite and the rarer zinc phosphates – hopeite, parahopeite and tarbuttite. The major vanadium minerals are descloizite and vanadinite but these are no longer available in economic quantities. The gangue consists of dolomite and pyrite and iron oxides.

The history of metallurgical practice developed over the past 67 years, has been described in detail (Barlin, 1970). The early years saw the recovery of lead from cerussite outcrops by blast furnace smelting, followed later by the recovery of zinc by electrolysis from predominantly zinc silicate ores. In 1938 underground mining was started to exploit the sulphide ores at depth and in 1946 a major plant construction programme was completed to produce both lead and zinc from these ores.

The existing plants treat a mixture of underground ores together with large tonnages of materials from dumps that have accumulated in former years during the development of the mine. To treat this combination of sulphide and oxidised ores the metallurgical treatment plants consist of: (1) a concentrator for the separation of lead and zinc sulphides from the zinc silicate tailing, (2) an electrolytic plant for grade 2 zinc recovery, including a flash roasting and acid

---

¹) Broken Hill Division, Nchanga Consolidated Copper Mines Limited.

Photo 1
An aerial view of the metallurgical plants.

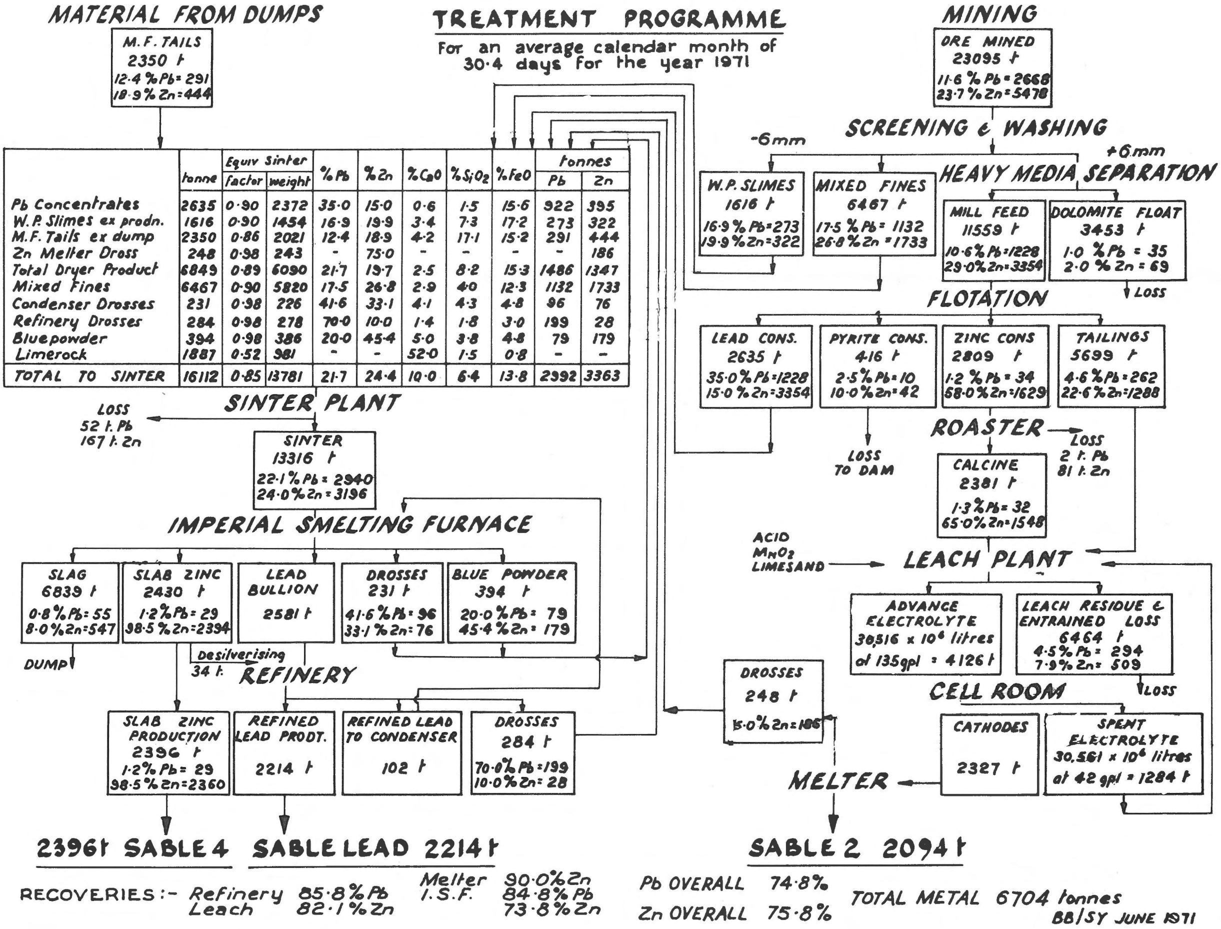
plant and a zinc leach plant, (3) a zinc smelter, (4) a cadmium plant, (5) an Imperial Smelting Furnace with sinter plant for the recovery of lead and grade 4 zinc and (6) a lead refinery. By-products cadmium and silver are also recovered. The 1971 treatment programme is given in fig. 1.

CONCENTRATOR

The concentrator flowsheet is shown in fig. 2.

Crushing

The ore hoisted from underground is generally



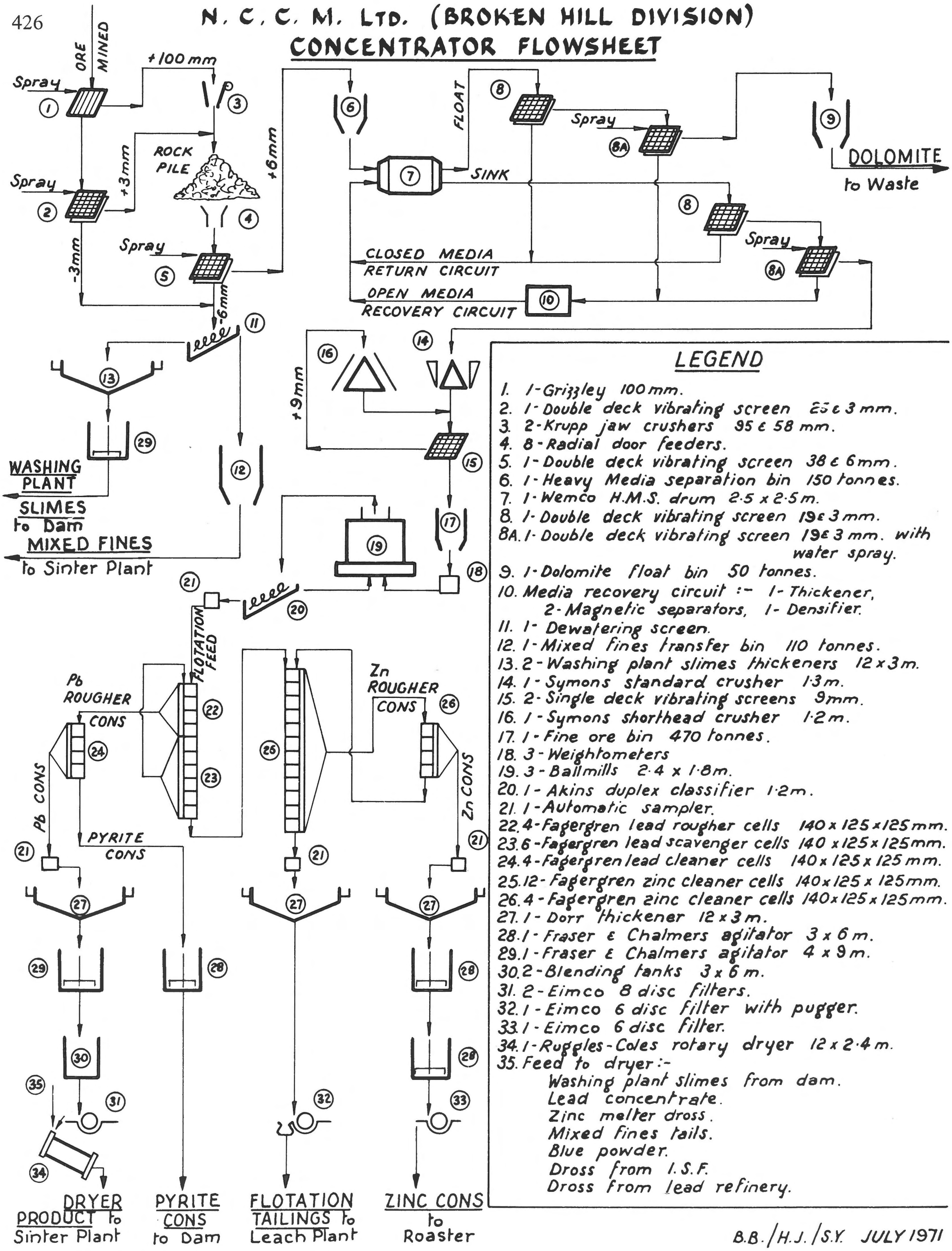
	tonne	Equiv Sinter factor	Equiv Sinter weight	% Pb	% Zn	% CaO	% SiO₂	% FeO	tonnes Pb	tonnes Zn
Pb Concentrates	2635	0·90	2372	35·0	15·0	0·6	1·5	15·6	922	395
W.P. Slimes ex prodn.	1616	0·90	1454	16·9	19·9	3·4	7·3	17·2	273	322
M.F. Tails ex dump	2350	0·86	2021	12·4	18·9	4·2	17·1	15·2	291	444
Zn Melter Dross	248	0·98	243	-	75·0	-	-	-	-	186
Total Dryer Product	6849	0·89	6090	21·7	19·7	2·5	8·2	15·3	1486	1347
Mixed Fines	6467	0·90	5820	17·5	26·8	2·9	4·0	12·3	1132	1733
Condenser Drosses	231	0·98	226	41·6	33·1	4·1	4·3	4·8	96	76
Refinery Drosses	284	0·98	278	70·0	10·0	1·4	1·8	3·0	199	28
Bluepowder	394	0·98	386	20·0	45·4	5·0	3·8	4·8	79	179
Limerock	1887	0·52	981	-	-	52·0	1·5	0·8	-	-
TOTAL TO SINTER	16112	0·85	13781	21·7	24·4	10·0	6·4	13·8	2392	3363

Fig. 1
Treatment Programme 1971.

damp and at times sticky, making handling without washing difficult. The -250 mm run-of-mine ore is hoisted in 4-tonne bottom dump skips at the rate of 1100 tonnes per day to a surge bin. Mine and concentrator operate a 6-day week, hoisting 8 hours per day while the concentrator operates 24 hours a day. The ore from the surge bin is conveyed to a stationary sloping-bar grizzly set at 100 mm and sprayed with water. The grizzly oversize passes through a 79 cm x 48 cm Krupp jaw crusher and is conveyed to a storage pile, which provides a surge capacity of 1000 tonnes. Grizzly undersize is washed over a double deck vibrating screen equipped with high-pressure water sprays. The screen oversize (+ 3 mm) joins the crusher discharge at the storage pile. The undersize is pumped to the secondary washing plant cyclones.

Washing

The -100 mm ore from the storage pile is drawn on to a conveyor, through chutes, into a vibrator feeder and conveyed to a second double deck screen, sprayed with wash water. The thoroughly washed (+ 6 mm) oversize is transported to a surge bin ahead of the heavy media separation plant. The -6 mm mixed fines and the -3 mm fines from the primary washing screen are dewatered in a 60 cm diameter hydrocyclone and Robins Gyrex dewatering screen. Cyclone overflow consisting of the slime fraction is sampled, measured and pumped to a storage dam for



LEGEND

1. 1-Grizzley 100mm.
2. 1-Double deck vibrating screen 25 & 3mm.
3. 2-Krupp jaw crushers 95 & 58 mm.
4. 8-Radial door feeders.
5. 1-Double deck vibrating screen 38 & 6mm.
6. 1-Heavy Media separation bin 150 tonnes.
7. 1-Wemco H.M.S. drum 2·5 x 2·5m.
8. 1-Double deck vibrating screen 19 & 3mm.
8A. 1-Double deck vibrating screen 19 & 3 mm. with water spray.
9. 1-Dolomite float bin 50 tonnes.
10. Media recovery circuit :- 1-Thickener, 2-Magnetic separators, 1-Densifier.
11. 1-Dewatering screen.
12. 1-Mixed fines transfer bin 110 tonnes.
13. 2-Washing plant slimes thickeners 12 x 3m.
14. 1-Symons standard crusher 1·3m.
15. 2-Single deck vibrating screens 9mm.
16. 1-Symons shorthead crusher 1·2m.
17. 1-Fine ore bin 470 tonnes.
18. 3-Weightometers
19. 3-Ballmills 2·4 x 1·8m.
20. 1-Akins duplex classifier 1·2m.
21. 1-Automatic sampler.
22. 4-Fagergren lead rougher cells 140 x 125 x 125mm.
23. 6-Fagergren lead scavenger cells 140 x 125 x 125mm.
24. 4-Fagergren lead cleaner cells 140 x 125 x 125 mm.
25. 12-Fagergren zinc cleaner cells 140 x 125 x 125mm.
26. 4-Fagergren zinc cleaner cells 140 x 125 x 125mm.
27. 1-Dorr thickener 12 x 3m.
28. 1-Fraser & Chalmers agitator 3 x 6 m.
29. 1-Fraser & Chalmers agitator 4 x 9 m.
30. 2-Blending tanks 3 x 6 m.
31. 2-Eimco 8 disc filters.
32. 1-Eimco 6 disc filter with pugger.
33. 1-Eimco 6 disc filter.
34. 1-Ruggles-Coles rotary dryer 12 x 2·4 m.
35. Feed to dryer :-
    Washing plant slimes from dam.
    Lead concentrate.
    Zinc melter dross.
    Mixed fines tails.
    Blue powder.
    Dross from I.S.F.
    Dross from lead refinery.

Fig. 2 Concentrator flowsheet.

drying purposes. The dewatered -6 mm mixed fines are conveyed to an intermediate storage bin and shipped to proportioning bins in the sinter plant.

Heavy Media Separation

The + 6 mm -100 mm ore is conveyed to a 2.5 m x 2.5 m Wemco heavy media drum separator using a ferro-silicon media controlled at 2.97 density. The float overflows the drum and the sink product is elevated and discharged through a launder chute onto the medium recovery screens. These are Allis Chalmers double-deck screens with a 19 mm top deck and 3 mm bottom deck. The medium that drains from the product screens returns to a densifier. The medium adhering to the products is washed off by high pressure sprays and pumped to the medium recovery magnetic separators. All reclaimed medium is returned to the densifier. The sink product contains all the lead and zinc values and is stored ahead of the secondary crushing plant. The float product consists of dolomite with minor proportions of lithified muds and low grade oxidised iron minerals. The separation is highly efficient and the ore minerals are almost completely recovered.

The H.M.S. plant is controlled by ensuring perfect washing of the feed, maintaining a constant medium density and frequent slime removal from the medium.

Crushing Plant

The sink from the heavy media plant is crushed and ground. The secondary and tertiary crushing installations are of conventional design and consist of a 1.3 m Symons standard cone crusher in open circuit with a 9 mm single deck Allis Chalmers low head screen fitted with 9 mm square wire mesh screen cloths, followed by a 1.2 m Symons shorthead cone crusher in closed circuit with the same screen. The crushed ore is conveyed to the grinding plant storage bins.

Grinding Plant

The grinding plant consists of three 2.4 m x 1.8 m ball mills independently fed from the storage bin. The ore is weighed before being fed to the mill. The discharge from each mill is pumped independently by a 3 x 2 Warman sand pump to a common 1.2 m simplex Akins Classifier. Classifier sands are returned to the feed end of the mill via a 3 way splitter ensuring an even feed distribution. The classifier overflow discharges at 42% solids, corresponding to a grind of 65% passing 325 mesh. The grinding circuit is controlled by automatically maintaining a constant 7 tonnes per hour feed to the mills and a constant classifier overflow density of 42% solids. The mill discharge pulp density is also maintained constant at 78% solids. Minor variations in flotation feed sizing are of little significance, as long as sliming of galena is avoided. For optimum flotation, the pH of the pulp is maintained at 7.5 by the addition of 630 g./tonne of ore of soda ash solution to the milling circuit.

Flotation

The flotation circuit is arranged for the differential flotation of galena, pyrite and sphalerite. The "tailing" which is comparatively rich in willemite and also contains some oxidised lead minerals, is transferred to the leach plant for zinc recovery. The concentrator products are utilised in their entirety in other plants for metal recovery, so that the recovery of metals in the flotation plant is virtually 100%. Flotation of the sulphide minerals is relatively simple. However, the selectivity of flotation between galena, sphalerite, pyrite and the oxidised mineral tailing is important. To this end, galena and pyrite are floated in a bulk concentrate in 10 x 1.4 m Fagergren cells with 90 g/tonne of iso-propyl xanthate and 55 g./tonne of T.E.B. frother. No special precautions are taken to prevent the flotation of sphalerite – which normally does not float without activation – and the fraction which does reports in the galena concentrate which is treated for the recovery of both lead and zinc in the Imperial Smelting Furnace plant. The bulk galena-pyrite concentrate is laundered to the cleaner circuit where 250 g./tonne of cyanide solution is added for pyrite depression. Pyrite reports in the cleaner tailing while galena and little sphalerite report in the concentrate.

The tailing from the lead rougher circuit is pumped, after the addition of 400 g./tonne of copper sulphate solution, to 12 x 1.4 m Fagergren zinc rougher cells. The sphalerite floats with 120 g/tonne of isopropyl xanthate and 30 g./tonne of T.E.B. frother. The rougher concentrate is cleaned in 4 x 1.4 m Fagergren cleaner cells and the cleaner tails are returned to the head of the rougher circuit. All rougher cell agitator mechanisms rotate at 360 r.p.m. and the cleaner cells at 440 r.p.m.

The pyrite concentrate is dumped at present; the galena concentrate is thickened, filtered and dried, and blended with dump materials for shipment to the sinter plant.

Control of the flotation plant has the objective of ensuring that a very clean zinc concentrate, free of lead and iron, is produced for subsequent flash roasting and electrolytic zinc recovery. The pyrite concentrate, if not discarded, would form part of the I.S.F. feed and would report in the Imperial Smelting Furnace slag.

Thickening, Filtering, Blending and Drying

The pyrite concentrate is sampled, weighed by means of a gamma density gauge and magnetic flowmeter and pumped to a storage dam.

The zinc concentrate is thickened in a 12 m diameter x 3 m deep Bowley lift thickener, sampled, weighed, pumped to a Goldfields storage tank, filtered on a six-leaf Eimco disc filter and conveyed to the roaster storage shed.

The lead concentrate is thickened to 45% solids and pumped to a 6 m x 3 m Goldfields storage tank prior to filtration on an eight-leaf Eimco disc filter. At this stage dump tailings materials, dry washing plant slimes and filtered lead concentrate are blended. The dump materials and dump-dried washing plant slimes are simultaneously fed into a hopper and withdrawn at a predetermined rate and join the filtered lead concentrate; the final blended mixture is fed to a Ruggles-Coles dryer 12 m long by 2.4 m diameter, dried to 4-5% moisture and perfectly blended. It is transferred to a storage bin, then to railway trucks, sampled, weighed and transported to the sinter plant proportioning bunkers.

The tonnage of the flotation "tailing" is computed by means of a gamma density gauge and magnetic flowmeter on pumping to a 12 m x 3 m Dorr thickener and filtered on a six-leaf Eimco disc filter. The cake is repulped with impure zinc sulphate solution in a pugger and pumped to the leach plant storage tanks.

Details of the concentrator metallurgy are given in tables 1A and 1B.

TABLE 1 A
Concentrator metallurgy 1971.

Material	% Wt.	Assays							Distribution %						
		T/Pb	PbS	T/Zn	ZnS	T/Fe	CaO	SiO_2	T/Pb	PbS	T/Zn	ZnS	T/Fe	CaO	SiO_2
Ore Hoisted	100.0	11.9	8.0	24.1	14.7	14.5	4.7	10.5							
Washing Plant															
Washing Plant Slimes	11.6	18.3	10.7	21.6	14.0	17.7	3.8	10.1							
Mixed Fines	26.0	17.2	11.7	29.2	22.1	12.9	3.3	5.3							
Heavy Media Plant															
Feed	62.4	8.5	5.8	22.4	11.5	14.8	5.7	13.5	100.0	100.0	100.0	100.0	100.0	100.0	100.0
Sink	45.9	11.0	7.5	29.1	15.2	17.6	0.7	12.2	95.2	95.8	95.3	97.3	87.4	8.7	66.8
Float	16.5	1.5	0.9	4.0	1.1	7.0	19.8	16.9	4.8	4.2	4.7	2.7	12.6	91.3	33.2
Flotation Plant															
Feed	100.0	11.0	7.6	29.1	15.2	17.6	0.7	12.1	100.0	100.0	100.0	100.0	100.0	100.0	100.0
Lead Concs.	20.5	39.3	34.0	15.3	11.2	14.8	0.6	2.3	73.1	92.3	10.8	15.0	17.2	19.3	3.8
Pyrite Concs.	4.9	2.2	0.8	14.4	8.5	34.3	0.2	4.1	1.0	0.5	2.4	2.7	9.5	1.8	1.7
Zinc Concs.	22.3	1.2	0.8	58.6	53.6	4.7	0.2	1.5	2.5	2.4	45.0	78.5	5.9	5.7	2.8
Flotation Tailings	52.3	4.9	0.7	23.2	1.1	22.6	0.9	21.2	23.4	4.8	41.8	3.8	67.4	73.2	91.7

TABLE 1 B
Flotation Agents

	Reagents grams per tonne	
	Pb Circuit	Zn Circuit
Soda Ash	631	
Isopropyl Xanthate	92	124
T.E.B. Frother	55	31
Cyanide	254	
Copper Sulphate		409

ELECTROLYTIC ZINC PRODUCTION

Flash Roaster

The roaster and acid plant flow sheet is shown in fig. 3. The roaster was installed in 1946, at the same time as the concentrator, to treat zinc concentrates for leaching in the electrolytic zinc plant. It consists of a cylindrical steel shell 7.5 m in diameter and 12 m high, lined with firebrick. It has two drying hearths and a large combustion chamber. Raking of the material across the hearths is carried out by rabble blades carried on alloy steel arms and supported on a centrally rotated column. A third, calcine cooling, hearth was dispensed with in 1965 when repeated failures of the hearth brickwork drastically curtailed operations. The two drying hearths are used prior to fine grinding and roasting. They are heated to 250°C by hot gases drawn from the combustion chamber. The dried concentrate then passes to a Hardinge ball mill in closed circuit with an air classifier to provide a satisfactorily fine size of material for roasting (90% – 325 mesh). The ground concentrate is injected with compressed air into the combustion chamber with excess combustion air from the burner fan. Oxidation of the zinc sulphide takes place with the evolution of sulphur dioxide. Roasting proceeds autogenously at a temperature of approximately 1070°C. The calcine settles on the 4th hearth or roaster floor and is rabbled to the outlet drophole from where it is transported in water-cooled screw conveyors to another conveyor system leading to the leach plant storage bins. Roaster gases at 1100°C containing between 9 and 13% SO_2 pass through a series of two water-cooled heat exchanger columns, 1 m diameter x 5 m high, followed by 3 air-cooled cooling columns. The gas temperature is reduced to 500°C before entering two Cottrell electrostatic precipitators in parallel, for dust precipitation. Calcine collected in the precipitator joins the main stream via a system of water-cooled screw conveyors.

Sulphuric Acid Plants (see fig. 3)

The acid plant was built in 1926 to provide acid for the leaching of willemite ores. The source of sulphur was pyrite ore which was roasted in multiple hearth Herreshoff roasters. These were replaced by the flash roaster in 1946. The plant is a conventional lead chamber installation with a capacity of 65 tonnes of sulphuric acid per operating day. It consists of two brick packed Glover towers, 3.5 m dia. x 10 m high, eleven Mills Packard chambers 14 m high, with a volume of 440 m^3 and four bricked-packed Gay Lussac towers 4.5 m x 14 m high. Oxides of nitrogen, produced by the oxidation of ammonia, are dissolved in the circuit acid which is sprayed over the brick-work of the Glover towers. The oxides of nitrogen are driven off and join the gas stream passing up the tower. The gas stream passes to the Mills Packard chambers where a fine spray of water rains over the gases. The reactions taking place, simply stated, are as follows:

$$SO_2 + NO_2 = SO_3^+ + NO$$
$$SO_3^+ + H_2O = H_2SO_4$$
$$2NO + O_2 = 2NO_2$$

The oxides of nitrogen act as the catalyst converting sulphur dioxide to trioxide which, when dissolved in water, produces sulphuric acid. The regenerated oxides of nitrogen are now redissolved from the gas stream in the circuit acid emerging from the Glover towers, cooled, and recirculated over the Gay Lussac towers. The regenerated acid is again passed over the Glover towers and the cycle repeats itself. Losses of oxides of nitrogen are replenished by continuously converting ammonia gas to the oxides of nitrogen in a specially designed converter using platinum gauze as the catalyst. Roaster and acid plant productions for 1970 given in table 2.

Leaching section

The leach and electrolytic zinc plant flow sheet is shown in fig. 4. The leach plant first came into operation in 1928 for leaching of zinc silicate ores exclusively. In 1946, however, calcine from the

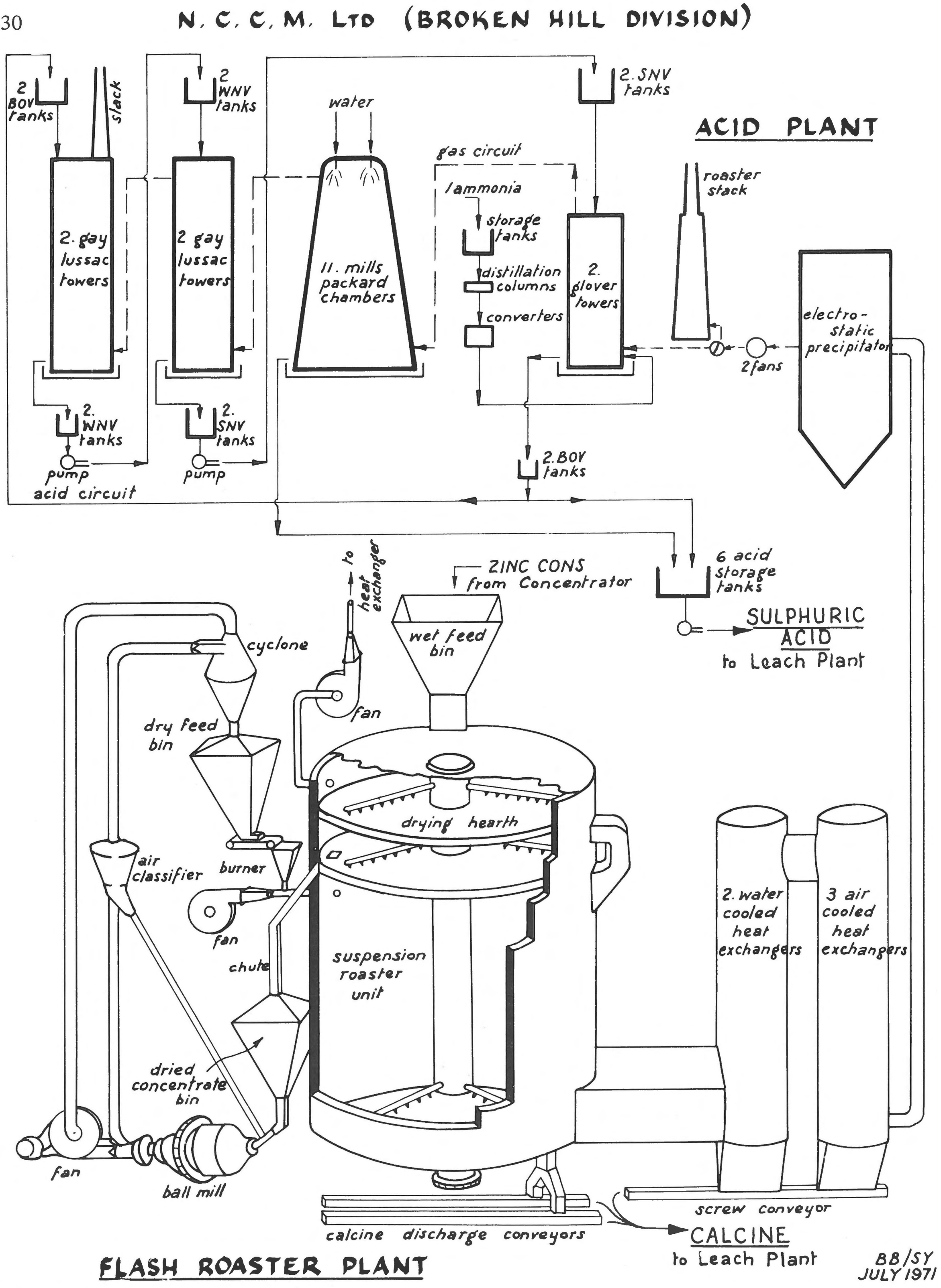


Fig. 3 Roaster and Acid Plant flowsheet.

TABLE 2
Roaster and Acid Plant – 1970 production.

Material	Wt. Tonnes	Assays				Distribution%	
		T/Pb	PbS	T/Zn	ZnS	Pb	Zn
Zinc Concentrate Roasted	26 341	1.2	0.8	58.6	53.6	100.0	100.0
Calcine Produced	22 316	1.4	0.1	67.4	3.2	98.7	97.5
Unaccounted loss	–	–	–	–	–	1.3	2.5

100% H_2SO_4 produced in 1970 – 13 460 tonnes.

TABLE 3
Leach and Electro Zinc Plant – 1970 production.

Material	Wt. Tonnes	Assay		Distribution	
		Pb	Zn	Pb	Zn
Leach Plant					
Input					
Calcine	21 684	1.4	67.4	9.1	48.9
Flotation Tailings	61 909	4.9	23.2	90.9	48.0
Zinc Dust	924	–	100.0	–	3.1
Limesand	1 620	–	–	–	–
Manganese Ore	4 079	–	–	–	–
Output					
Leach residue to dam	66 536	5.0	9.4	100.0	20.8
Cathode Zinc	23 680	–	100.0	–	79.2
Melter					
Input					
Cathode Zinc	23 680	–	100.0	–	100.0
Output					
Sable 2 Zinc	21 092	–	100.0	–	89.1
Zinc Dust	1 017	–	100.0	–	4.3
Dross and Stack loss	–	–	–	–	6.6

Overall leach recovery 73.6%

roaster became available for leaching, boosting zinc production considerably. The 1970 production details are given in table 3.

A unique leaching technique, which overcomes the problem of filtration of pulps containing soluble silica was developed by the company's metallurgists in the early 1920's. Leaching is carried out, batch-wise, in mild steel, rubber-lined, tanks 5 m dia. x 5 m deep of 90,000 litre capacity, fitted with Denver high speed agitators. A leach is started by filling the tank to the level of the agitator with spent electrolyte from the cell room and a measured quantity of strong acid from the acid plant, followed by additional 12 to 15 tonnes of repulped flotation tailings. This procedure ensures an initial acid concentration in the tank of more than 300 g.p.l. sulphuric acid which "dehydrates" the silica. Leaching of silicates at low acid concentrations renders the silica gelatinous and filtration is impossible. When the reaction is completed and effervescence ceases, spent electrolyte is added to fill the tank to about 90,000 litres. Calculated quantities of calcine are added to reduce the free acid to 12-15 g.p.l. and a slurry of ground manganese ore is added in predetermined proportions to oxidise all ferrous iron to the ferric state. After sufficient agitation the residual free acid is neutralised with

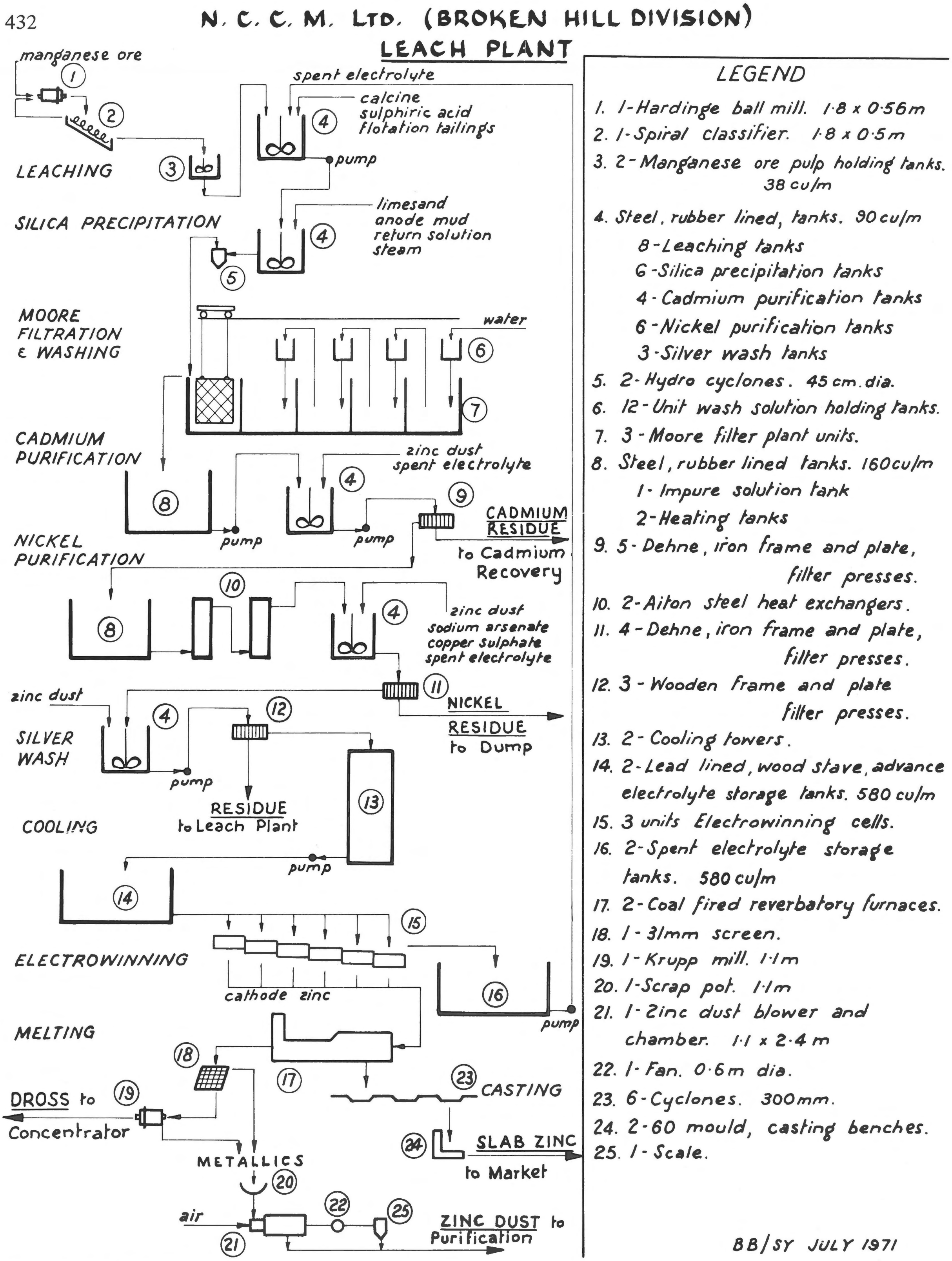


Fig. 4 Leach Plant flowsheet.

further calcine additions to a residual free acid tenor of 0.5 to 1.0 g.p.l. The leach is now complete after a period of 8 hours and is pumped to the silica precipitation section by means of a 3 x 4 Warman rubber lined pump.

Silica Precipitation

The silica precipitation tanks are identical in construction to the leach tanks, but they are fitted with live steam lines. Any residual ferrous iron in the charge is oxidised by the addition of anode mud from the cell room, which is slightly acid.

Limesand, a locally obtained limestone, is repulped in impure electrolyte and calculated quantities are added to neutralise the residual acid in the charge. A measured excess, usually 90 kg, is then added while the temperature of the pulp is raised to +65°C. It is this operation which is so important in granulating the silica and rendering the pulp filterable. The filterability is tested empirically by timing the filtration of 200 cc. of hot pulp on a 12 cm Buchner funnel at 50 cm Hg vacuum. A solution recovery of better than 80% of the volume filtered within 60 seconds is usually considered satisfactory. If necessary, further additions of limesand and higher temperatures are utilised to ensure that these standards are maintained, because otherwise subsequent Moore filter operations can be seriously impeded. At this stage the pH is 5.0-5.2 and iron, As, Sb, Sn, P, V_2O_5 and Ni are hydrolysed and precipitated with the silica.

Moore Filtration

Prior to filtration, the pulp – after silica precipitation – is pumped through a 45 cm dia. cyclone to remove the coarser granular particles which would otherwise settle in the Moore filter tanks. The cyclone underflow is washed with water and allowed to settle. The supernatant liquor is decanted and joins the main circuit, while the solids are pumped with the Moore filter residues to the residue dam. The cyclone overflow is stored in a tank before being dropped to the "form tank" of the filters. The filter basket lowered into the tank consists of 40 leaves each 3 m x 2 m of wooden framing covered with cotton twill cloth. Vacuum at 30 cm Hg is applied for a fixed period, usually 60 minutes of forming time after which a cake of washable thickness (2.5 cm) has formed on the leaves and 40 000 litres of solution has been filtered. Four stages of countercurrent washing follow using the filtrate from the succeeding wash tank. A measured quantity of wash water is added in the last wash tank. The washed cake is now hosed off the basket leaves with high pressure water jets and pumped, via the residue disposal pumps, to the residue dams. A cyclic system of cloth washing and caustic soda soaking to remove blinding silica gel from the cloths is carried out. Solutions from the wash tanks at +130 g.p.l. Zn are combined and clarified through an additional filter basket before being pumped to the purification impure solution storage tanks.

At this stage a very careful check is kept on the leach plant solution balance. Should the volume of circulating solution increase, a calculated reduction in wash water is made. Should the zinc tenor of the solution drop below the accepted limits, the acid additions to the leach tanks are increased and vice versa.

Purification of Electrolyte

Impurities which interfere with efficient electrolysis are removed as well as those elements which plate out with the zinc and would affect its final purity. Nickel, cobalt and copper are electrolytically toxic while cadmium which is removed for its value as a by-product, would normally plate out with zinc.

Cadmium purification

The impure zinc sulphate solution containing about 130 g.p.l. zinc, is pumped to the cadmium impure solution storage tank. The purification process is carried out batch-wise in 4 rubber lined steel tanks fitted with slow speed wooden agitators. To each batch of 80 000 litres of solution a weighed quantity of fine zinc dust is added with gentle agitation. Solution temperature is 45°C and is kept low. The zinc replaces the cadmium which is precipitated as the metal. After one hour's agitation, the cadmium content is reduced from 0.05 g.p.l. to less than 0.001 g.p.l. At the same time the copper content is reduced to 0.0005 g.p.l. The charge is filtered in a Dehne filter press with iron frames. The residue when washed from the filter press is transferred to the cadmium plant for eventual recovery of cadmium.

Nickel purification

The solution from the cadmium presses is transferred to one of two 160 000 litre nickel impure solution storage tanks. The process is again carried out batch-wise in similarly designed tanks. Each charge is drawn from the nickel impure tanks through one of two Aiton heat exchangers where the solution temperature is raised to +85°C in the purification tank. To each charge sufficient spent electrolyte is added to just acidify the solution and 1.5 kg sodium arsenite, 11 kg of Cu as $CuSO_4$ solution and 35 kg of coarse zinc dust are stirred in. The nickel is reduced from 0.0035 to 0.0011 g.p.l. and cobalt from 0.002 g.p.l. to less than 0.0008 g.p.l. After a similar agitation period the solids are filtered off through one of 5 Dehne presses and the purified solution is transferred to the last stage of purification. The solids removed from the press frames are transferred to a dump and sold when a sufficient quantity has accumulated. No attempt is made to recover the valuable metals.

Silver Wash purification

In former years at this stage, chlorine was removed from solution by the use of silver sulphate. This practice was discontinued in 1967. Chlorine tenor of the solution increased steadily to the 0.7 g.p.l. mark; current efficiency or lead tenor of the zinc were not affected and a considerable saving in silver resulted. Hence the last stage of purification was called the silver wash, signifying a final clean up of any residual silver that may have been present in the solution. However, this last phase was not discontinued and is specifically aimed at the recovery of any residual cadmium and the removal of any remaining copper. The process is carried out batchwise in similar rubber-lined steel tanks with 50 kg of fine zinc dust and with gentle agitation.

The silver wash residue is recovered in Dehne presses fitted with timber plates and frames. The solids are releached and returned to the circuit, while the solution is pumped over two cooling towers in series where the temperature is reduced from 50-55°C to 30°C and then pumped over to the cell room feed tanks. Each stage of purification is carefully controlled by submitting samples at regular intervals for analysis. Trace elements in these solutions are determined by atomic absorption techniques.

Cell room feed solution in 1970 averaged 131.8 g.p.l. Zn, 0.003 g.p.l., Fe, 10.3 g.p.l. Mn, 0.0008 g.p.l. Cd, 0.0003 g.p.l. Cu, 0.645 g.p.l. Cl, 0.0011 g.p.l. Ni, 0.0008 g.p.l. Co, 0.015 g.p.l. F and 15.5 g.p.l. MgO.

The electrolytic zinc plant (see fig. 4)

The plant consists of two identical units and a third, smaller unit, installed in 1953. The two original units consists of six double cascades of six cells each, making a total of 72 cells per unit. Each cell contains 40 aluminium cathodes and 42 lead anodes. The cells are of wood, lead-lined and set on insulated blocks on concrete foundations. The third unit contains four double cascades of 9 cells each, making a total of 72 cells each with 24 cathodes and 25 anodes. Each cell is fitted with lead cooling coils through which some 20 million litres of water are circulated daily to maintain a temperature of approximately 36°C in the cell electrolyte.

The main solution feed lines distribute the solution from the main storage tanks to the head of each bank of cells in 15-cm pipes and a 5-cm pipe down the centre of each double cascade. All cells are fed with solution from carefully controlled plastic taps. The solution flows from one cell to the next and the overflow from the last cell in each cascade discharges into the spent electrolyte launder and flows into two 580 000-litre storage tanks before returning to the leach tanks to complete the cycle. The current for units one and two is provided by five rotary converters each 1750/2000 kW, 220/285 V and 8000 A.D.C. The average operating voltage is 262 V and the average amperage per unit is 20 000 A. Unit 3 is served by 12 1000 A. mercury arc rectifiers and operates at an average voltage of 258, taking 1 000 amps. Current density is 280 amps per square metre. Current efficiency in 1970 averaged 83.5%.

Cathodes are 4.8 mm thick, hard-rolled, aluminium welded to aluminium header bars with welded aluminium lugs for lifting purposes. The submerged area is 71 cm by 102 cm, the spacing 10.8 cm between centres. The anodes are made of lead containing 0.025% Ag and 0.013% Cu. They are 8 mm to 9 mm thick with copper header bars cast in.

The submerged area is 64 cm by 94 cm; the life two years. The acid tenor of the electrolyte is controlled by frequent titration and adjusted by varying the flow of new solution into each cell to maintain a concentration of 42 g.p.l. Zn and 150 g.p.l. H_2SO_4.

A cascade of cells is stripped on alternate days giving a zinc deposition time of 48 hours. Cathodes are lifted in groups of ten by block and tackle on an overhead crawl beam and transferred to the stripping racks where some 80 tonnes per day of zinc is stripped. The stripped zinc is stacked in bundles, weighed and transferred to the melting section. The cathodes, after stripping, are examined and scrubbed if necessary, straightened and fitted with grooved rubber edging strips to facilitate stripping. On a regular basis, the cells are cleaned manually of accumulated anode mud by manipulation with long-handled scoops, after removing cathodes and anodes in 10 batch lots from the cell being cleaned. The anode mud, mainly manganese dioxide is reused in the leach section for iron oxidation.

ZINC MELTER

The melter section consists of two hand-fired, coal-burning, reverberatory furnaces, used alternately. Each furnace is 5.8 metres long by 3.7 m wide by 2.1 m high, measured internally. Bundles of cathodes weighing approximately 90 kgs are fed manually into the charging door of the furnace. The furnace has a bath capacity of 120 tonnes. Cathodes are fed in three batches, one batch per shift. Feeding is carried out as quickly as possible, maintaining the bath molten and under reducing conditions to limit oxidation of zinc. The bath is then raised to 650°C and the furnace is drossed. Dross containing some 6.5% of the zinc fed to the furnace is skimmed off into cars and transferred to a Krupp mill for the recovery of metallics. The metal-free dross is transferred to the concentrator drying plant for inclusion in the dryer product before shipment to the sinter plant.

Molten metal is tapped into ladles, suspended from a chain block on a crawl beam, to a row of tilting moulds. The slabs, cast by hand, are skimmed, allowed to cool and tipped out of the moulds under a spray of water. The slabs, each 25 kg, are stacked in one tonne lots, strapped, weighed and railed to market. Details of the process are somewhat academic, because at time of writing the installation is being replaced by an 8 tonne/hour Ajax Magnathermic induction melter and Shepperd straight line casting machine.

The Sable 2 zinc produced confirms to B.S.S. 3436, Zinc 2, and its assay in 1970 averaged 0.0216% Pb, 0.0008% Cu, 0.0010% Fe, 0.0007% Cd, with Zn (by difference) 99.9759%.

Zinc dust production is carried out in the melter building. Scrap zinc is melted into conical pots fitted with a 5 mm hole at the bottom. The molten zinc gravitates through the hole in front of a blast of compressed air and is atomised. The dust so formed is caught in a cooling chamber and a series of grain cyclones which catch the finer zinc dust for use in cadmium purification.

CADMIUM PLANT

The cadmium plant flowsheet is shown in fig. 5.

Dissolution of Precipitate

The cadmium precipitate from the Dehne presses in the leach plant purification section is ground in a small Hardinge ball mill with water and fed to a 36 000-litre wood-stave, lead-lined, leach tank containing acid. Some 90% of the soluble material dissolves in a very short period. Most of the remaining, undissolved, cadmium and zinc can however be brought into solution by allowing a limited quantity of the copper in the precipitate to oxidise and dissolve. As soon as copper dissolves, the undissolved zinc and cadmium displace it from solution, precipitating the copper again and themselves going into solution. This process is continued until a stable showing of dissolved copper indicates that all available zinc and cadmium have passed into solution. At this stage a small quantity of fresh precipitate is added in an amount just sufficient to precipitate the dissolved copper. This results in a pulp which contains a maximum of zinc and cadmium in solution and a maximum of copper as a precipitate. The pulp is filtered and the filtrate transferred to cadmium sponge precipitation.

Filtration and sponge precipitation

The pulp is pumped through a Dehne press, with the aid of some "Superfloc", to the leach tank. The

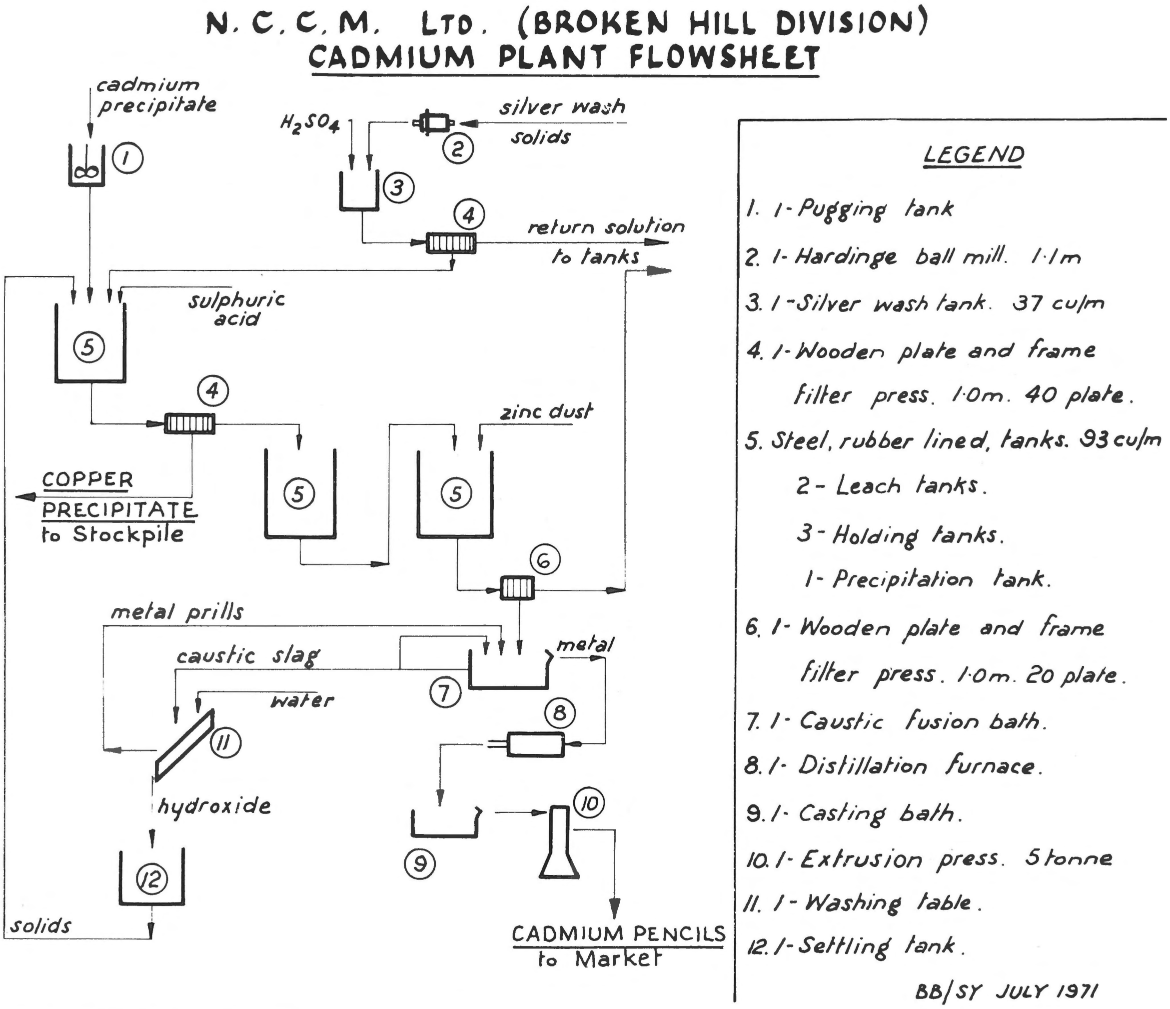


Fig. 5 Cadmium Plant flowsheet.

filtrate is circulated until it is free of solids and contains only traces of copper. Once copper-free, the filtrate is delivered to a holding tank. The residue in the press is washed with water until free of cadmium bearing solution. The copper bearing residue is removed from the press, air dried and sold for its copper content.

A measured quantity of the cadmium bearing solution is now pumped to the precipitation tank. For efficient precipitation, conditions are controlled within very narrow limits. The solution is diluted to about 7 to 9 g.p.l. cadmium. Acid is then added to bring the acidity to 3.5-4.5 g.p.l. The acidity must not lie outside these limits. Higher starting acid causes unnecessary acid and zinc consumption and resolution of the cadmium sponge. Low starting acids may result in a colloidal cadmium precipitate, or a basic solution, either of which may cause poor filtration and result in the inclusion of undissolved zinc in the cadmium sponge.

Zinc dust is added to the solution in the tank in stages, little by little, to precipitate the cadmium as a sponge. Zinc dust additions are continued until a sample solution after standing a few minutes shows a white haze when treated with sodium sulphide. Care is taken not to add an excess of zinc dust. When precipitation is complete the precipitate is allowed to settle. Filtration is then carried out in a Dehne press which is first filled with water to prevent oxidation of the sponge. The filtrate is returned to the main zinc leach circuit.

The precipitate in the press is thoroughly washed with water and the wash solution is returned to the precipitation tank. The precipitate is then allowed to drain.

Melting

The drained precipitate, still containing some 50% moisture, is removed from the press and transferred to the melting section. Here it is mixed intimately with flake caustic soda before being introduced into the melting bath. The surface of the bath is sprinkled liberally with caustic soda. The bath is heated to 460-480°C and the mixture is rabbled vigorously in order to enhance fusion and reduce cadmium loss. When the reaction ceases the metal is poured into oil-brushed moulds and the slag skimmed off. The ingots of crude metal are thoroughly wire brushed to remove any caustic which might otherwise cause distillation difficulties. The slag is returned to the next melt.

Distillation

When sufficient crude metal has accumulated, (usually some 7000 kilos) a distillation campaign is started. Distillation is carried out at 780-800°C; the ingots of distilled metal are remelted under caustic soda and cast into billets. The risers from the billets are removed and returned to the melting bath, while the billets are extruded into pencil form. The extrusion temperature is 305°C; the pressure is 175 kg/cm^2. The final pencils measure 25 cm long x 1.25 cm/diameter and are packed in boxes containing 50 kg. The purity of the cadmium produced is better than 99.95% Refined cadmium production in 1970 was 12 242 kg.

THE IMPERIAL SMELTING FURNACE PLANT

The Imperial Smelting Process was introduced into the Broken Hill metallurgical complex in early 1962 with the specific purpose of producing both lead and zinc from current concentrator products and higher grade dump materials which up to that time could not be treated in the existing plants. The plant was the second to be built outside Britain, where the process originated, and the first to treat comparatively low grade materials. The principles of the process have been described in detail (Morgan, 1957). The following sections describe the sintering and Imperial Smelting processes for the treatment of concentrator products and dump materials. Details of sinter plant feed materials in 1970 are shown in tables 4A and 4B.

SINTER PLANT

The sinter plant flowsheet is shown in fig. 6.

Charge proportioning

Feed materials are transported to the proportioning plant from the concentrator in 40 tonne

TABLE 4 A
Sinter Plant – materials treated 1970

Material	% Wt.	Assays					
		Pb	Zn	CaO	SiO_2	FeO	T/S
Dryer Product	43.4	24.2	20.2	2.4	8.2	13.0	12.0
Mixed Fines	35.9	17.4	29.5	3.4	5.2	16.7	17.7
Plant Residuals	5.9	42.3	28.0	3.1	4.6	1.2	12.6
Blue Powder	2.4	39.5	28.1	3.6	4.4	3.2	3.7
Lime Rock	10.3	–	–	51.5	1.5	0.7	–
Coke Breeze	2.1	–	–	1.0	5.6	0.4	1.2
Total New Feed	31.6	20.2	21.6	7.9	5.9	12.4	11.7
Return Sinter	68.4	24.2	25.1	11.1	7.3	15.1	2.6
Sinter Produced	168 931 Tonnes	23.6	26.6	11.0	8.3	15.3	0.6

TABLE 4 B
Dryer product – 1970 production

Material	% Wt.	Assays				
		Pb	Zn	CaO	SiO_2	FeO
Washing Plant Slimes and Mixed Fines Tailings from dump	49.5	18.0	20.4	4.2	11.4	13.7
Lead Concentrate	27.8	39.3	15.3	0.6	2.1	14.8
Melter Dross and Calcine Cleanings	2.3	1.6	76.4	0.4	2.6	3.8
Zinc Concentrate	2.6	7.5	44.2	5.6	5.3	7.8
Blue Powder and Mixed Fines Tailings	17.8	30.5	21.3	5.1	7.9	6.6
Dryer Product	100.0	24.2	20.2	2.4	8.2	13.0

railway cars and introduced into their respective storage bins via a tippler system. The limestone flux is received by rail truck from the Ndola Lime Plant situated some 175 kilometres north of Kabwe. The tippler system transfers the limerock to a hammer mill where it is crushed from 60 mm down to 95% minus 28 mesh, before being conveyed pneumatically to a proportioning plant bin. Metallurgical coke is received in railway wagons from the Wankie Colliery in Rhodesia and introduced into the plant via the same tippler system. It is screened on 19 mm and the plus 19 mm fraction is used for blast furnace fuel, while the minus 19 mm fraction is crushed through a small roll crusher to minus 5 mm for use as additional fuel in the sinter charge. The coke contains 85% fixed carbon, 1.5% volatile matter and 12% ash.

The materials are fed from the proportioning bins, which are fitted with independently supported truncated base cones, onto one metre wide endless belts which pass over Schenck weighers. The individuel feeds are controlled to the required constant weight by automatically varying the belt speeds. The weigh feeders discharge onto a collector conveyor. Blue powder from the furnace and sinter plant ventilation dust are thickened and filtered and the cake discharged onto the conveyor immediately before the mixing drum.

Charge mixing and pelletising

The proportioned sinter feed is fed into a 2.5 metre by 6 metre long mixing drum rotating at 5.5 r.p.m. Water is added by fine sprays towards the discharge end of the drum, the addition is controlled by a signal from a moisture probe on the conveyor belt immediately after the mixing drum discharge. This probe was developed by the company's instrument engineers; it is based on the measurement of the thermal conductivity of the machine feed. An additional control depends on the measurement of the back pressure on the first windbox of one machine, when blowing a constant volume of air through the bed, using a separate fan for the purpose. The back pressure is used as a measure of the bed permeability, which is of greater significance than moisture content. The single drum provides both mixing and pelletising, necessary for good sintering. However, with increased production rates, constantly being required, it is intended to install a second drum in series with the first to improve the control of mixing and pelletisation by carrying out these operations in separate drums.

Sinter machines

The plant houses two updraught sinter machines each 1.5 metres wide by 18.6 metres long giving a total grate area of 56 m^2. Each machine has 12 updraught windboxes and one downdraught windbox for ignition. The entire length of the updraught section is enclosed with an unlagged, trapezoidal shaped, mild steel hood with appropriate flue connections for the collecting of all SO_2 bearing gases. Each machine is driven by a 30 h.p. electric motor through reduction gearing to give an infinitely variable strand speed which is usually varied between 0.6 and 1 metre per minute. The pelletised feed from the mixing drum is split between the two machines. A moving shuttle feeder directs the material into two rotary table feeders. Load cells are fitted to each

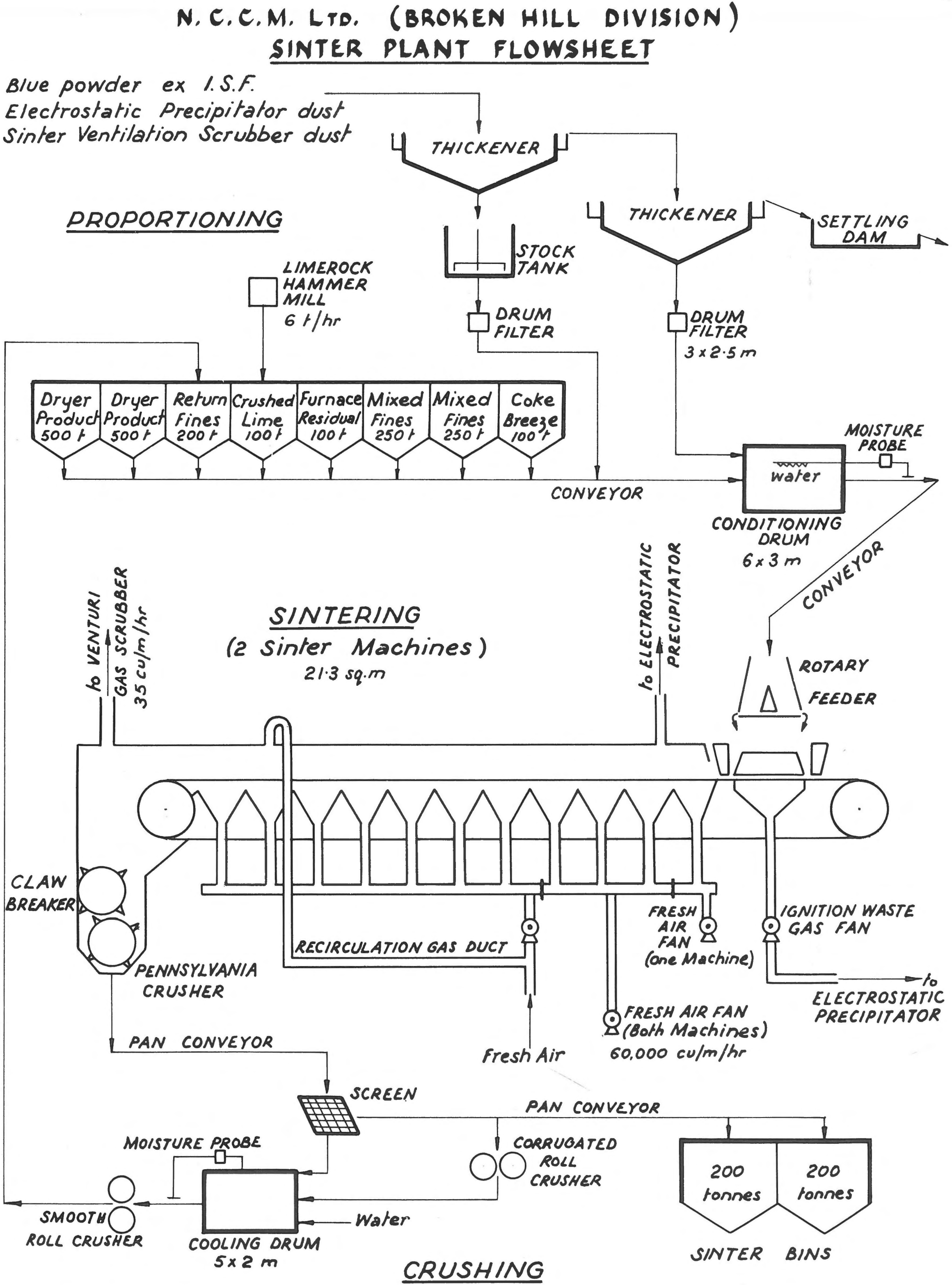


Fig. 6 Sinter Plant flowsheet.

rotary table and the shuttle feeder is actuated by the weight difference between the two rotary tables.

The feed to the ignition and main layers is accomplished by means of adjustable ploughs on opposite sides of the rotary table and controlled by power cylinders actuated by level probes fitted to the feed hoppers. Swinging chutes distribute the material across the feedboards. The ignition layer approx. 35 mm thick, is ignited under an oil fired ignition hood fitted with three No. 4A Scheildrop oil burners. The height of the main layer is set by the feed board to give an overall depth of 300 mm. After ignition by downdraughting, updraught fresh air is provided by 18 000 Nm^3/hr fans to the first four windboxes. The remaining windboxes are provided with updraught fresh air by the main fans (Deutsche-Babcock, variable speed, capacity 65 000 Nm^3/hr). The charge is sintered as it moves along the strand underneath the hood until it discharges at the end of the machine. The discharged sinter cake passes through a claw breaker with grate bars set at 100 mm into a Pennsylvania crusher which reduces the top size to minus 90 mm.

Sinter crushing and screening

Hot, crushed, sinter is conveyed by a steel pan conveyor to a vibrating screen with a 19 mm punched plate deck. The undersize is fed to a 2 m dia. by 5 m long cooling drum fitted with water sprays. Lump sinter required to make up the necessary quantity of return sinter is split from the screen oversize and crushed through a corrugated roll crusher (Esch-Werke, 1.3 m dia. by 0.55 m wide) before passing into the cooling drum. Cooled return sinter is further reduced to approx. 24% plus 4 mesh in a 1.5 m dia. by 1.0,m wide Esch-Werke smooth roll crusher before being returned by conveyor belt to the proportioning plant. The finished sinter or screen oversize containing 0.5% sulphur is conveyed by means of a steel pan conveyor to the sinter bunkers ahead of the furnace.

Waste gas handling

Waste gases from the sinter machines pass through simple dust cyclones and thence to Lodge-Cottrell conditioning towers where fine water sprays reduce the gas temperature to 110°C. The cooled gases are cleaned by a Lodge-Cottrell electrostatic precipitator. Flue dust from the towers and the precipitator is pulped with water and pumped to a 12 m dia. thickener, filtered and returned to circuit.

The tip ends of the machines are ventilated by means of a Venturi gas scrubbing system, while the screen, crushers and cooling drum are ventilated by means of Meduse scrubbers. The effluent from all these sources is pumped to a 12 m dia. thickener and the underflow filtered on a rotary drum filter and returned to the sinter feed collector belt together with blue powder from the furnace.

Control of sinter composition

Two-hourly samples of output sinter are analysed for lead, zinc, lime, silica, iron and sulphur. Lead and zinc tenors are controlled by altering the proportions of mixed fines and dryer product. The lead content can also be controlled by varying the addition rate of plant residuals. The lime/silica ratio is controlled at 1.35 in sinter by adjusting the limestone feed. Control of sinter quality is maintained by careful fuel additions and moisture and machine speed control to maintain the breakthrough point about midway down the length of the machine. The high lead and silica content of the sinter ensures a rattle hardness index of well over 90% which is ideal for subsequent smelting in the I.S.F.

Imperial Smelting Process

The Imperial Smelting Process and lead refinery flowsheet is shown in fig. 7 and 1970 production details are given in table 5.

The basic furnace charge consists of the lead zinc sinter from the sinter plant, preheated metallurgical coke and recirculated flux bath and condenser oversize drosses. The lump sinter is delivered from the sinter plant to two storage bins each of 250 tonne capacity. Cold coke is transferred to the coke preheaters by a conveyor and pocket elevator system after screening to plus 19 mm.

Three coke preheater units are provided, each of refractory lined, mild steel, construction. The coke is heated by an ascending mixture of hot gases produced by burning low calorific value furnace waste gas in a separate combustion chamber, into which are also admitted scrubbed exhaust gases recirculated

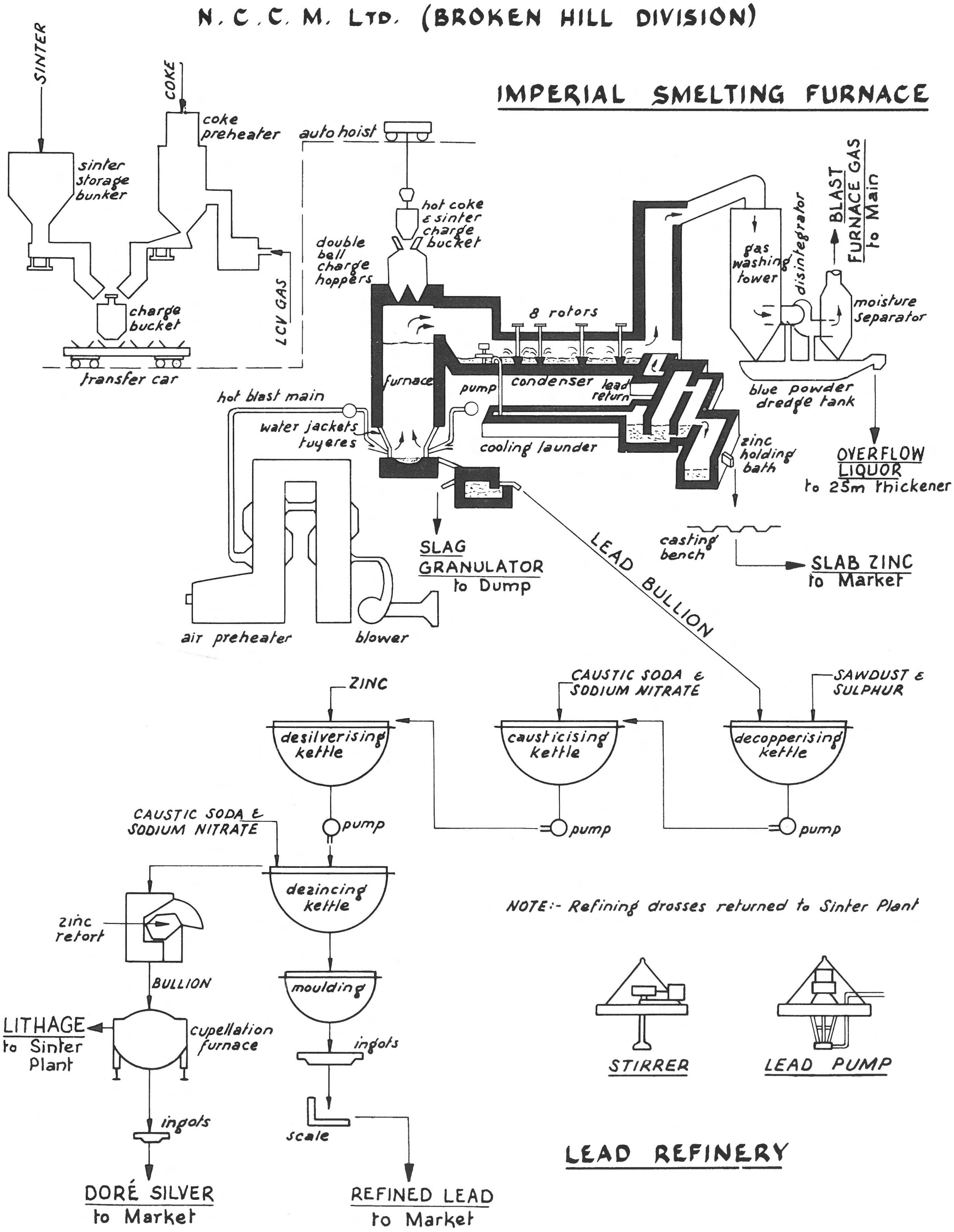


Fig. 7 Imperial Smelting Furnace and Lead Refinery.

BB/SY JULY 1971

TABLE 5
Imperial Smelting Furnace – 1970 production

Material	Wt.	Assay		Distribution	
		Pb	Zn	Pb	Zn
Input					
Sinter	170 198	23.6	26.6	94.9	100.0
Lead to Condenser	2 171	100.0	–	5.1	–
Coke	50 619	–	–	–	–
Output					
Lead Bullion	32 171	99.8	–	76.8	–
Sable 4 Zinc Slabs	32 726	1.2	98.8	0.9	71.7
Slag	82 719	0.7	8.6	1.3	15.7
Furnace Drosses *	14 805	38.0	35.0	13.3	11.5
Unaccounted losses	–	–	–	8.6	1.1

* Furnace drosses are recycled to the sinter plant.

from the preheater. The proportion of combustion air is adjusted to give less than 0.5% free oxygen in the heating gases. Hot coke at 600°C is discharged by a cast iron extractor roll at the base of the preheater. A level controller controls an automatic sequence for the replenishment of the unit after each charge of hot coke has been withdrawn. The sinter bins and coke preheaters are fitted with vibratory screen discharge units and loadcell type weigh hoppers which are automatically controlled to fill two charge buckets with preset quantities of coke and sinter. The two buckets are carried on rotating turntables on a charge transfer car which moves on a rail track running between the charge preparation plant and the furnace auto hoist.

The charging cycle is fully automatic except for the initiation, which is determined by the charge level taken at the furnace top. Individual charge weights are preset in the charge preparation control room according to furnace requirements and are recorded on a print-out system for metallurgical accounting and control purposes.

Imperial Smelting Furnace

The charge buckets are hoisted and transferred to the furnace top, which is fitted with two double bell charging hoppers. When the buckets have been lowered into position the top bell opens under the weight of the charge and the bottom bell opens automatically under the influence of an air cylinder, when the top bell has closed.

The blast furnace itself consists of a reinforced steel casing lined with high grade alumina brick, backed by insulating material. The roof is cast in refractory concrete and the hearth is lined with chrome magnesite brick. The furnace shaft is approximately 9 metres high from the hearth to the roof and is 7 m long and 1.55 m wide at the hearth. The bottom of the furnace is fitted with water jackets containing 18 tuyeres evenly spaced on the long sides of the furnace. Each tuyere, 100 mm diameter, is water cooled by softened water circulated through a bank of four forced draught cooling towers. Blast air is delivered by two Demag blowers each rated at 24 000 Nm^3/hr at 380 millibars. The air is preheated to 650°C by a Thermal Efficiency air preheater, fired on low calorific value furnace waste gas.

Zinc vapour produced from the reduction of the sinter is carried with the furnace gases through the furnace offtake and enters the condenser. The rectangular condenser is fitted with eight rotors which are partially immersed in a pool of molten lead. The lead flows countercurrently to the gas flow and is vigorously agitated by the rotors to create an intense shower of lead droplets throughout the condenser. The furnace gases enter the condenser at a temperature of 1020°C and are shock cooled by the lead spray. The zinc vapour is cooled rapidly through the critical temperature range in which reversion to ZnO occurs and the liquid zinc so formed is taken up by the molten lead. The circulating lead is pumped to an external separation system. The mixture of CO, CO_2 and N_2 gases leaving the condensor passes to a wet scrubbing system which recovers solids (principally oxides of zinc and lead) carried over from the furnace shaft and condenser, as blue powder. The scrubbed gas has a low calorific value and is used for preheating both blast air and coke.

The furnace is tapped continuously through a water cooled copper tapping breast. The mixture of slag and lead bullion tapped from the furnace runs into a refractory-lined forehearth in which the lead bullion is separated from the slag, which flows over the top of the lead into a granulating launder fitted with high pressure water sprays. The granulated slag is laundered to a pump sump and pumped to the slag dump. The lead bullion is transported to the lead refinery in 4 tonne ladles by tractor and trailer.

Zinc recovery

The circulating lead containing the zinc taken into solution in the condenser flows from the condenser into the pump sump, from which it is pumped by two Cockle Creek type lead pumps driven by variable speed motors. Each discharges into a long, water-jacketed, cooling launder in which the lead is progressively cooled from 570°C to 450°C. Cooling results in a quantity of zinc being rejected from solution in the circulating lead. The amount of zinc liberated is 0.2 – 0.3% of the weight of the circulating lead which is pumped through the system at a rate of some 2000 tonnes per hour.

The mixture of liquid zinc and circulating lead discharges from the cooling launders over a weir into the flux bath. Here an ammonium chloride flux layer is maintained to minimise dross formation resulting from the turbulent discharge of metal from the cooling launders and to collect dross and other impurities.

The metals then flow via an underflow baffle into a large refractory lined separator bath. In this the zinc separates and floats as a relatively quiescent layer on top of the circulating lead, which flows back via an underflow baffle, into the condenser, through the return launder. The liquid zinc layer overflows continuously into a small liquation bath which is maintained at 440°C to 445°C to keep the lead content of the zinc as low as possible. From here the zinc flows into an oil fired holding bath and then to a second liquation bath maintained at a temperature close to the freezing point of zinc (419°C) to ensure maximum removal of lead. Zinc overflows from the second liquation bath to a reheat bath where its temperature is raised to 470°C prior to hand casting from ladles into 25 kg moulds before palletising, weighing and strapping for market. Plans are well advanced for the installation of a continuous straight line Shepperd casting machine to eliminate hand casting. The Sable 4 zinc produced conforms to B.S.S. 3436, Zinc 4, and the assay in 1970 averaged 1.19% Pb, 0.014% Fe, 0.0010% Sn, 0.03% Cd, 0.0018% As and 0.0024% Cu, with zinc, (by difference) 98.76%.

Blue Powder Handling

The blue powder collected in the condenser exit gas scrubbing system is pumped by means of an 8 x 6 Warman pump to a 25 m diameter thickener. The thickened pulp is elevated by a Dorrco diaphragm pump into a 6 m x 3 m storage tank ahead of a 2.5 m x 3 m rotary drum filter fitted with polypropylene cloth. The blue powder is filtered and the cake rejoins the new feed to the sinter plant mixing drum.

Operating Controls

The furnace capacity is determined by the coke burning rate, which in turn is determined by the blast rate. For the raw materials available at the present time 26 000 Nm^3/hr is the maximum blast rate possible.

Two-hourly assays of slag indicate the trend of zinc in the slag. The quantity of sinter charged per unit of coke is adjusted as necessary to maintain a constant zinc content in slag, at the lowest level consistent with good slagging conditions. Due to the very high proportion of slag forming constituents in the sinter, arising from the nature of the raw materials currently available for treatment, the zinc loss in slag is relatively high (table 5). Particular attention is therefore paid to slag composition. Investigations have shown that mellitite type slags produce the most favourable operating conditions and therefore iron input to the sinter plant is kept as low as possible. This is the reason for the production of a pyrite concentrate in the flotation section, which is at present time stockpiled. The proportion of alumina in the slag is insufficient to maintain mellitite type slags but similar results are obtainable with dicalcium silicate slags, although zinc elimination is not quite as good. High iron, wustite type, slags have proved disastrous due to the overreduction of iron oxides resulting in freezing of metallic iron in the taphole.

The plant is extensively instrumented and automatic control of all operations is carried out where practicable. The main control room is situated on the condensor floor and is equipped with mimic diagrams and indicating, recording, alarm and control equipment for the remote operation of the coke preheaters, the air preheater, the blast air blowers, ancillary equipment and main services.

REFINERY (fig. 7, Perry, 1961)

The lead bullion from the furnace is comparatively low in impurities. The average analysis is as follows: Ag: 0.03%, Cu: 0.8%, As: 0.04%, Sb: 0.03%.

The bullion is fire-refined, by dry drossing, sulphur drossing, desilverising by the Parkes process, and finally dezincing by causticising. The Parkes crusts are enriched and go forward to cupellation for the recovery of their silver and lead content.

The plant consists of 6 hemispherical mild steel kettles of 90 tonne capacity each, set in a brick firebox that are heated by Fairbanks-Morse stokers using pea coal fuel. Two kettles are used for decopperising, two for desilverising, one for dezincing and one for final drossing and holding prior to casting. They are served by an overhead crane system, which handles mechanical stirrers for the various drossing operations, pumps for transferring the lead from one kettle to the next, desilverising crusts and drosses, as well as assisting in the maintenance of equipment.

There is also a 10 ton moulding kettle and bull-ring for casting the final product into pigs for market.

A rotary furnace is provided for the distillation of the zinc from the enriched crusts and a cupellation furnace for the treatment of the retort bullion.

Decopperising

The bullion from the furnace is transferred to the decopperising kettles, in the molten state, from 4 tonne ladles using the overhead crane. When a kettle is full the temperature is allowed to drop to about 370°C, a mechanical stirrer is placed in the kettle and the lead is stirred for a period of 15 minutes.

A copper-rich dross forms on the top of the bullion and this is dried by additions of sawdust and further stirring. The "dry" or "copper" dross is then skimmed and stored for retreatment. In this operation the copper content of the bullion is reduced to 0.06%. The copper content of the dross varies between 10 and 15%. It also carries some 80% lead and must therefore be returned to the sinter plant for the recovery of this metal.

The second stage of copper removal is the sulphur drossing operation. This process depends upon the affinity of sulphur for copper and the low solubility of Cu_2S in lead at the melting point of lead. The bullion is allowed to cool almost to freezing point. A stirrer is placed in the kettle and rotated. Elemental sulphur is then added to the vortex formed in the lead by the action of the stirrer, at the rate of 1 kg per minute. A total of 1.5 kg sulphur per tonne of bullion is normally added. The copper sulphide dross formed by this treatment is skimmed off, leaving a bullion containing 0.005% Cu or less. The dross contains +80% Pb and 2-5% Cu and is returned to the sinter plant.

At this stage the lead is heated to 500°C and treated with caustic soda and sodium nitrate to remove zinc (returned from the desilverising kettle) together with some arsenic and antimony. Interference with desilverising occurs if these elements are not removed at this stage.

Desilverising

The Parkes process for desilverising is used and owing to the low silver content of the bullion, one stage only is normally sufficient. The process is dependent on the solubility of zinc in lead at various temperatures, the preferential solubility of silver in zinc compared with lead and on the relative insolubility of a zinc-silver alloy in lead saturated with zinc at the freezing point of lead.

When zinc is stirred into the molten lead, the latter gradually becomes saturated with zinc. If the temper-

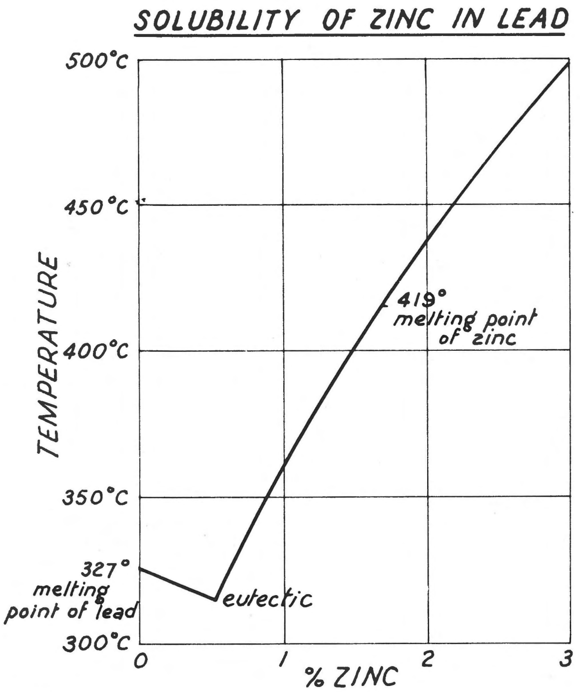


Fig. 8
Solubility of zinc in lead.

ature is raised until the zinc melts, the solubility increases as shown in fig. 8. It will be noted that when the zinc melts at 419°C its solubility in the lead does not increase suddenly and therefore two liquid layers will form if excess zinc is present. The upper layer will be essentially zinc, while the lower layer will be lead saturated with zinc.

When zinc dissolved in lead, it unites with the silver which was dissolved in the lead and forms a compound Ag_2Zn_3 which is virtually insoluble in lead saturated with zinc at its melting point. As the temperature is reduced, the zinc-silver alloy forms crystals which float on the surface of the bullion and these can be skimmed off but will normally carry considerable amounts of entrained lead. If this process is continued to the freezing point of lead, the latter will be almost free of silver but will contain zinc of the lead zinc eutectic composition, i.e. 0.577% Zn.

Before being pumped from the decopperising kettle, the crusts lowest in silver, from the previous charge, are put into the desilverising kettle and the decoppered lead is pumped to the desilverising kettle. The bullion is heated to 540°C, the hooks from the crusts are removed and the bullion is stirred for 10 minutes. A mushy crust forming on the surface is skimmed into two ton moulds and, while still molten, steel hooks are placed in the moulds to facilitate subsequent handling. This is known as the first crust. Since it is normally low in silver and its zinc content is largely oxidised, it is returned to the decopperising kettles.

Approximately 6 tons of crust from the previous charge (containing the next lowest silver content) is now added, the kettle is reheated to 540°C, the hooks are removed and the bullion is stirred for 10 minutes. The second crust is skimmed off as above and held for return to the succeeding charge. The remaining crusts from the preceding charge (some 6 tons) containing the most silver, are now added with the new zinc. The kettle is reheated to 540°C, hooks are removed and the third crust skimmed. The temperature of the bullion is dropped to 440°C by spraying the surface with water. It is then stirred and skimmed. This crust usually carried the highest silver tenor. If the silver assays 3% or more, it is set aside for silver recovery; if below 3% silver it is added to the next charge with the new zinc.

The fires are now removed and the kettle is flooded with water, and further crusts removed. This crust is treated in the same way as the silver crust. The purpose of adding the crusts in stages as described above, is to maintain a comfortable working level in the kettle for skimming and to gradually increase the silver tenor. When the temperature reaches 360°C, the kettle is thoroughly scraped to remove adhering crusts and skimmed. This process is continued right down to freezing point. The bullion is then heated 5°C, sampled and pumped to the dezincing kettle. During desilverising, approximately 20 tons of crusts are removed. These are sampled and held for return to the next charge or for silver recovery. The zinc requirements are normally determined by consideration of the crusts added to and removed from the previous charge or by reference to cooling curves. New zinc consumption normally amounts to about 3.5 kg per tonne of bullion. The desilverised lead contains some 0.0004% Ag but there is usually some redissolution of silver during pumping, so that the final lead usually contains 0.0003% Ag. During desilverising most of the remaining copper is also removed, and the final lead contains about 0.0005% Cu.

Dezincing

Formerly, vacuum dezincing of lead was practiced, as developed by the St. Joseph Lead Company, but since the advent of the Imperial Smelting Process it was found far more expedient to return the zinc rich drosses to the sinter plant, and the vacuum dezincer was abandoned. Removal of the residual zinc and traces of arsenic and antimony is therefore effected by a modified Harris process. Caustic soda and sodium nitrate are stirred into the lead at a temperature of 500°C. The sodium nitrate acts as an oxidising agent and the zinc, arsenic and antimony form sodium salts with the caustic soda. The completion of the operation is judged by observing the film formation time on a sample of the lead and by the well-defined crystal structure of the solidified sample. The resulting dross is skimmed off and returned to the sinter plant. The zinc content of the refined lead is usually 0.0004% and antimony 0.0008%. Caustic soda and sodium nitrate consumption averages 1 kg and 0.5 kg/ton respectively.

Moulding

The refined lead is pumped to an elevated, 10 ton capacity, kettle, provided with an outlet at the bottom, which can be opened by means of a plugcock. The lead is gravitated to a jackpot mounted on a gimbal and from there it is hand-moulded into 25 kg pigs on a semi-circular moulding ring. The pigs are skimmed, stamped with the day and the year and removed from the moulds by hand. They are weighed and stacked ready for market. About 60 tonnes are produced from each cast and at present about 40 charges are moulded per month. The average grade of refined lead in 1970 was 99.9971%. Refined lead is required to conform to B.S.S. 334. 1970 refinery production details are given in table 6.

Treatment of Parkes crusts

As mentioned under desilverising, any crusts containing 3% silver or over are set aside for silver recovery. When approx. 15 tonnes of crusts have accumulated, they are placed in the desilverising kettle, which is filled up to the working level with decopperised lead and the kettle is heated to 540°C with sufficient zinc added to give a zinc content of 5%. The temperature is then allowed to drop and the resulting crust is skimmed and hand pressed to remove some of the entrained lead. While still hot it is broken up to a convenient size for handling. Skimming and pressing is continued until the temperature has dropped to 440°C and then the completion of desilverising of the charge proceeds as usual.

The crusts obtained from this procedure contain some 6% Ag and 70% Zn, most of the remainder being lead. These crusts were formally treated in conventional retort furnaces for the recovery of the contained zinc by distillation. Currently they are fed to a rotary furnace where the zinc is volatilised as the oxide. The furnace bullion from these furnaces is cast into moulds of 1000 troy ozs each and assay 10% Ag, 0.5% Cu, 1.5% Zn, the remainder being lead. The bullion is stored until some 100 000 troys ozs of contained silver have accumulated, ready for cupellation.

Cupellation

The final stage in the recovery of silver has now been reached. In this process the retort bullion is melted, the lead is oxidised to litharge by blowing air

TABLE 6
Lead Refinery – 1970 production

Material	Wt. Tonnes	Pb Assay	Pb Distribution
Input			
Bullion from furnace	32 086	100.0	97.5
Bought in lead scrap	821	100.0	2.5
Output			
Moulded lead	29 736	100.0	90.4
Dross	4 154	76.3	9.6

on the surface of the bath and the litharge is tapped off while still molten. The other base metals are also oxidised and tapped off with the litharge. This leaves Doré silver, which is refined by additions of lead, which is tapped off carrying copper and other impurities with it. Equipment used consists of a small tilting cupellation furnace with a capacity of 100 000 troy ozs of silver. It is lined with Austrian magnesite (Radex) bricks up to the metal level and is provided with oil burners and air nozzles. Litharge is skimmed in the molten state, over a water-cooled cast iron block set in the breast. This same outlet is used for charging, moulding and working the furnace generally. Silver metal is cast into 1000 oz bars for market.

A cupellation is conducted as follows: Sufficient rotary furnace bullion is added to fill the cupel. This is melted and heated to about 1000°C. Air at about 1.5 kg/cm^2 is blown on to the surface of the metal and the nozzles are adjusted until a definite wave motion occurs on the metal surface. The lead slowly oxidises and forms a layer of molten litharge on top of the bath, which is tapped by tilting the cupel so that a fine stream of litharge flows over the breast block. The litharge is collected in cast iron moulds mounted on wheels. When litharge formation slows down considerably or when the cupel cannot be tilted further, it is returned to the horizontal position and topped up with fresh bullion to the working level. When operating temperature is again reached, air blowing and litharge tapping is restarted and the cycle repeated. This is continued (about a week) until the cupel is virtually filled with crude silver. The cupel is then "run down" i.e. litharge formation and removal continues without any fresh bullion additions, until litharge formation virtually ceases. This lasts about a day.

Refining of the silver now commences. Shots of refined lead are added to the bath, which is poled with green timber for agitation and heated to 1050-1100°C. The lead is oxidised to litharge as above and tapped off. This operation gradually removes copper from the silver. The number of shots (ingots) of lead added depends on the final Cu tenor required. (To reduce copper to 0.5 parts per 1000, about 10 shots are usually necessary). When this operation is complete, the bath is heated to over 1100°C and the remaining lead is oxidised by additions of sodium nitrate – about four additions of 13 kg usually suffice. The remaining traces of litharge are soaked-up by throwing cement powder on the bath. A final check is made of the Ag, Cu and Pb content of the bath and if satisfactory, the charge is prepared for moulding. About 45 kg of charcoal is thrown on the bath to deoxidise the silver. This is necessary as silver absorbs over 20 times its own volume of the oxygen at this temperature, and part of this oxygen is given up as the silver cools to solidification point. If the silver is not deoxidised it will sprout in the moulds causing imperfect ingots and possibly endangering personnel. For the latter reason, the moulds are covered after filling. The Doré silver is then moulded into 1000 oz bars for market by attaching a spout to the moulding spoon and thence to the moulds. During moulding, a block of wood is placed on the breast to further remove oxygen. A typical analysis of Doré silver is as follows: Silver: 99.5%, Copper: 0.1%, Lead: 0.3%.

Silver recovery is not good and redesign of the refinery will include reverberatory type furnaces for the recovery of copper matte from the dry and sulphur drosses, conventional retorts for zinc recovery from silver crusts and possibly a Howard press for the enrichment of crusts before silver recovery. Automatic straight line casting of pigs will be introduced as will oil firing of kettles in place of the underfeed stokers used for kettle heating. Details of 1970 silver production are shown in table 7.

TABLE 7
Silver Recovery – 1970 production

Material	Wt. kg.	% Ag	Troy ozs. Ag
Silver Crusts produced	76 474	8.53	209 656
Dore Bullion Produced	52 049	10.15	169 849
Silver Produced	5 752	96.31	178 102

Metallurgical Control

The diversity of the operation and the many and varied types of analysis required call for strict control of analytical techniques and sampling and weighing procedures. To achieve this, a wide variety of weighing devices are used from belt weighers to rail and road weighbridges, as well as overhead scales. Automatic sampling equipment comprises exclusively R.A.M.S. type proportional samplers.

Analytical equipment consists of:
A Philips PW1212 X-ray fluorescence spectrometer for the analysis of ores and metals containing high proportions of metals.
A Philips X-ray diffractometer for the quantitative determination of sphalerite, galena and pyrite in concentrator products.
Three Techtron atomic absorption spectrometers for the determination of trace elements in refined metals as well as the analysis of leach plant solutions.

These techniques have made it possible to enhance the range of analyses carried out and some 20 000 analyses are carried out monthly involving 29 different elements. The methods used require a minimum of skilled labour for their execution.

THE FUTURE

There still remain considerable tonnages of slags and leach residues on dumps and there are a number of ore horizons containing refractory zinc silicate ores. These contain a considerable tonnage of recoverable metal. Recent studies indicate that a fuming process of some sort is necessary to fume the lead and zinc from the silicates so that the resultant fume can be added to the feed materials to the existing plants to improve the grade of feed to these plants. The effect of this will be to enrich the feeds to these plants thus increasing their output and improving the overall recovery of metal for the same tonnage of ore mined. Investigations have revealed that the Waelz Kiln will offer the best means of achieving this objective.

ACKNOWLEDGEMENTS

The author wishes to thank the Managing Director, Nchanga Consolidated Copper Mines Limited for permission to publish this paper, and Mrs. S. Young for compiling the flowsheets.

REFERENCES

Barlin, B. (1970) – The evolution of Lead Smelting Practice at The Zambia Broken Hill Development Co. Ltd. A.I.M.E. Symposium on Mining and Metallurgy of Lead and Zinc, St. Louis, U.S.A., October 1970.

Morgan, S.W.K. (1957) – The Production of Zinc in a Blast Furnace. Trans. Inst. Metall. 66, August 1957, p. 553-65.

Perry, A.J. (1961) – Lead Smelting and Refining at Broken Hill, Northern Rhodesia. Proceedings of Seventh Commonwealth Mining and Metallurgical Congress, May 1961.