

The conversion of a uranium plant to the recovery of zinc

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SYNOPSIS

The conversion of the Vogelstruisbult Uranium Plant into a plant for the production of electrolytic zinc is described. The reasons for the conversion, the preliminary investigations, and the basic design principles are all outlined, and a brief account is given of operating experience, during the commissioning of the plant, with equipment that had been put to a use for which it was not originally intended.

SAMEVATTING

Die omskepping van die Vogelstruisbult-uraanaanleg in 'n aanleg vir die produksie van elektrolitiese sink word beskryf. Die redes vir die omskepping, die voorlopige ondersoek en die basiese ontwerpbeginnels word almal in hooftrekke aangegee en daar word 'n kort beskrywing gegee van die bedryfservaring, tydens die indiensstelling van die aanleg, met uitrusting wat gebruik is vir doeleindes waarvoor dit nie oorspronklik ontwerp is nie.

The Zinc Corporation of South Africa has been established with the principal objective of producing South Africa's requirements of virgin zinc metal as far as possible from local raw materials. A number of circumstances occurred in 1965 that combined to initiate the concept of establishing a local smelter as a replacement industry for a redundant uranium plant. These circumstances were, in the main, as follows.

- (1) The over-riding principle that large domestic industries, such as the South African Iron and Steel Corporation (Isco), which produces strategic products, should become as independent as possible of imported raw materials. Zinc metal was regarded as a strategic material required by the steel and other industries, and at that time was a wholly imported product. At the same time, zinc concentrates were being produced in South Africa and South West Africa and were being exported for processing elsewhere. Isco therefore examined the possibilities of establishing a zinc smelter, and approached Gold Fields about the availability of the South West Africa Company's concentrates to augment the concentrates that they intended to produce at the Rosh Pinah Mine in southern South West Africa.

- (2) The imminent closure of the uranium plant at Vogelstruisbult

Gold Mining Areas. Vogelstruisbult had not participated in the stretch-out agreements in respect of the uranium contracts and would be required to close down a comparatively new plant, which had been in commission for only ten years. The Chairman of the Company, Mr R. A. Hope, therefore asked that the use of the plant or equipment for alternative processes should be considered.

- (3) Berg Aukas, the Gold Fields mine in South West Africa, produced zinc concentrates in addition to its principal output of lead-zinc vanadates. Two types of zinc concentrates were produced: conventional zinc blende and willemite, a zinc silicate. Tonnage-wise the willemite was the more important product, and examination of the mine's ore reserves indicated that large tonnages could be produced for a considerable period of time. One problem, however, was the unpopularity of zinc silicates with conventional zinc smelters. The request was therefore made that consideration should be given to the establishment of a zinc-recovery plant in South Africa that would be able to treat zinc silicates in large quantities.

Visits to producing zinc smelters drew attention to similarities between an electrolytic-zinc smelter and a uranium plant, and a project was initiated that had as its specific objective the conversion of the

Vogelstruisbult Uranium Plant to an electrolytic-zinc smelter based on local concentrates.

ZINC TECHNOLOGY

A conventional electrolytic-zinc plant purchases zinc sulphide concentrates and processes them to produce zinc slab, according to the following basic technology.

Concentrates are roasted to produce sulphur dioxide and calcine, in which the zinc occurs as zinc oxide. The sulphur dioxide is converted to sulphuric acid.

Calcine is leached in dilute sulphuric acid under near-neutral conditions so that the zinc is dissolved but only a limited, though necessary, quantity of iron is taken into solution. The leach pulp is neutralized with excess calcine, thus precipitating iron hydroxide, and is then thickened and filtered. Washing of the filter cake is necessary to reduce soluble zinc losses. The washed cake is discharged as residue.

Solution purification is undertaken to prevent interference with effective electrolysis by substances that seriously hinder electroplating when present in concentrations as low as 0,01 mg/l. Essentially, purification consists in precipitation of impurities by activated zinc dust and is usually carried out in two stages: the first at about 90 °C, and the second at a lower temperature. Filtration for the removal of excess zinc dust and precipitated sludge is carried out after each operation.

Electrowinning of dissolved zinc is

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the final process in the chemical sequence, and is normally carried out in lead-lined concrete cells, using lead anodes and aluminium cathodes. An emf of $3\frac{1}{2}$ V is applied across electrodes, and current densities vary from about 300 to 800 A/m^2 . Under these conditions, considerable heat is developed in the solution, and it is necessary to provide cooling towers in the circuit. The electroplating of the zinc results in regeneration of sulphuric acid, and spent electrolyte is recycled to the leaching section for the dissolution of calcine, thus completely closing the chemical circuit. Zinc-metal sheets are stripped from the cathodes.

Melting of cathodes into salable ingots is the final operation prior to marketing.

This brief description will help to show the similarities between the recovery of dissolved uranium by acid leaching and the electrolytic recovery of zinc. It will also help to explain the thinking behind the concept to convert the redundant uranium plant to the recovery of zinc.

Once the stage of a complete feasibility report had been reached, it became necessary to examine all the relevant factors more closely. In broad outline, these could be classified into two major groups: technical feasibility and economic feasibility. This paper deals principally with the technical considerations.

FEEDSTOCK

Basically, the project was investigated on the assumption that the raw feed to the plant would consist of the following.

(a) Zinc blende concentrates from Iscor's Rosh Pinah Mine in South West Africa, of which a probable analysis (in percentages) would be Zn 53, S 27, Fe 2,5, and Cd 0,08, with the following deleterious constituents: Ca 1,5, and Mg 0,75.

(b) Willemite concentrate from Berg Aukas, the Gold Fields mine in South West Africa. Typical significant constituents (in percentages) would probably be Zn 47, SiO₂ 25, Ca 0,80, Mg 0,54, and Ge 0,009.

Of these two raw materials, the

sulphide concentrate is typical of clean zinc concentrates that are treated successfully elsewhere and presented no foreseeable problems. The willemite concentrate, on the other hand, was known to be unacceptable to electrolytic plants owing to the solubility of the silica in sulphuric acid, which results in the production of a difficult-to-filter, voluminous gelatinous precipitate.

However, a programme of laboratory investigations was instituted, and possible techniques for the treatment of this material were established. These were confirmed on both laboratory and pilot-plant scales at the East St. Louis works of the American Zinc Company, and were later confirmed on a pilot-plant scale by Zincor. The translation of these results into a production process was never pursued because the Berg Aukas Mine indicated that it had decided to market its non-sulphidic zinc in the form of Waelz oxide. Samples of oxide produced on its behalf in a pilot plant in Germany were made available for testing, and extraction tests on this material had the following approximate analysis (in percentages): Zn 63, Pb 11, SiO₂ 1,0, Ca 1,0, Mg 0,5, and Ge 0,002. The fluorine content was 265 p.p.m.

At that stage it was decided to erect a pilot plant at Vogelstruisbult for larger-than-laboratory scale investigations of the recoverability of zinc from Rosh Pinah concentrates and Waelz oxide.

PILOT PLANT

The pilot-plant units were chosen to enable charges of 110 kg of calcine or Waelz oxide to be treated. The plant consisted of a rubber-lined conical leaching tank, mechanically stirred and provided with lead coils for steam heating. This was followed by a stainless-steel rotary-drum filter and two purification tanks. A plate-and-frame filter was provided for purification. Finally, storage tanks for purified solution and two electrolysis cells, each containing two production-size anodes and one cathode, were provided.

It was possible either to run single-batch tests with this equipment, or to recycle spent electrolyte and thus simulate continuous oper-

ation by the repetition of cycles as often as required. Twenty-five cycles were considered sufficient to simulate continuous closed-circuit operations.

For the roasting of sulphide concentrates, a multiple-hearth type of roaster was available, which, although not ideal, produced calcine sufficiently well roasted to give meaningful results.

DISCUSSION OF ZINC TECHNOLOGY

The simple sequence of operations previously enumerated appears to be straight forward. In practice, however, the unit operations are far more complex, and a more thorough understanding is necessary to appreciate the complexities of operating plants. Essentially, the objective is to dissolve the zinc content of the feed materials and to electrowin the zinc as pure metal.

For a proper understanding of the processes involved, it is advisable to examine the electrolysis operation in the first instance. The preferred electrolyte is a zinc sulphate solution in which are located lead anodes and aluminium cathodes, with an imposed emf of $3\frac{1}{2}$ V. Under normal circumstances, this emf will cause water to dissociate, as well as plate out zinc. However, with very pure electrolyte, a hydrogen over-voltage is set up and zinc is plated out in preference to hydrogen. This fortunate mechanism makes it possible to achieve a current utilization for the plating out of zinc of as high as 92 per cent. However, if impurities are present, even in the most minute concentrations, they reduce the hydrogen over-voltage, and current is wasted on the electrolysis of water.

A further factor to be allowed for is that the successful use of aluminium cathodes depends on the natural passivity at the surface of aluminium, which prevents adhesion of deposited zinc to the aluminium and permits ready stripping of the deposited sheet. If, however, this passive oxide film is removed, the zinc becomes firmly welded to the aluminium and cannot be stripped. One of the worst offenders here is fluorine, which causes sticking when present in concentrations of as low as 20 p.p.m. As is well known, fluorine

cannot readily be precipitated from sulphate solutions, and it is a trouble-maker in most zinc plants. Yet another impurity that is usually avoided is magnesium, which, too, cannot readily be precipitated from sulphate solutions and when present builds up to undesirable concentrations in a cyclic operation. It has a twofold effect. In the first place, it displaces zinc in saturated solution. With an atomic weight of 24, compared with 65 for zinc, 1 g of magnesium in solution will displace 2.7 g of zinc. Secondly, the presence of magnesium increases the viscosity of the solution, which has the effect of decreasing current efficiency and of increasing voltage drop. The removal of impurities from zinc sulphate solution thus becomes the overriding consideration in plant technology.

With this general background, a more detailed discussion of the implications of the unit process follows.

ROASTING

This is an area to which major effort is directed. The primary objective is to produce acid-soluble zinc oxide and sulphur dioxide by roasting at about 930 °C. There are, however, some side reactions that affect the subsequent leaching process. Most sulphide ores contain iron minerals and some base-metal minerals that participate in the roasting reactions. Some iron minerals unfortunately react with zinc and base metals to form ferrites, which are insoluble in leach liquor. The formation of ferrites can be reduced by roasting at a lower temperature, but under this condition a large proportion of the sulphur burns off as sulphur trioxide, and either persists as a gas or reacts with base metals to form sulphates. On the face of it, this should be advantageous in that the load on the sulphuric acid plant will be reduced. However, this is not the case because, in the cyclic zinc-leaching and electrowinning operation, the process *per se* is not a net consumer of acid, and zinc sulphate introduced into the circuit can result in a build-up of acid that would have to be bled from the circuit unless there are sufficient acid-consuming con-

stituents in the calcine to form insoluble precipitates and to be removed as such. Two such constituents are lead and calcium.

Roasting therefore calls for a measure of tight control, and both the design of the roasters and the operating parameters play an important role in the achievement of maximum efficiency.

HOT-GAS TREATMENT

The calcine leaves the roasters at about 900 °C, and, in the case of fluidized-bed reactors, a portion of the calcine is discharged as coarse material via a bed-overflow weir, while the remainder is entrained in the sulphur dioxide-bearing gases, from which it has to be separated and cooled. The solution-purification process uses a large quantity of steam, and it has become common practice to provide waste-heat boilers that serve the dual purpose of cooling the gases with their entrained solids and of providing the steam required on the plant. Rapid cooling of the gases has the beneficial effect of reducing the autogenous oxidation of sulphur dioxide to sulphur trioxide. A problem with boilers is the accumulation of dust on the tube bundles, and highly efficient rapping gear must be provided. The quality of the calcine, too, can affect the nature of the dust deposit, and copper and lead in the concentrates can be harmful, particularly if they occur together. In fact, these two metals also exercise a deleterious influence in the roasters and can cause freezing of the bed at high operating temperatures.

Some of the gas-borne solids are recovered from the boilers, and the remainder are removed by cyclones and hot-gas electrostatic precipitators before they enter the usual train of vessels in a contact sulphuric acid plant. The solids collected from the bed overflow, boiler hopper, cyclones, and electrostatic precipitator are further cooled and de-nodulized in a dry ball mill prior to being stored dry ahead of leaching.

LEACHING

The object of leaching is to dissolve the zinc oxide into a near-saturated solution of zinc sulphate, without

dissolving other soluble constituents of the calcine. The most soluble of these constituents is the iron oxide, which, on neutralizing, forms the usual voluminous, difficult-to-filter, ferric hydroxide. However, some dissolved iron is desirable for the following reason. Whereas the major removal of impurities is achieved by precipitation with zinc dust, it is difficult to remove all impurities by this technique. Such impurities can, however, be controlled by adsorption onto ferric hydroxide precipitate, which is discharged as leach residue.

One of two leaching procedures is normally practised: batch leaching or continuous leaching.

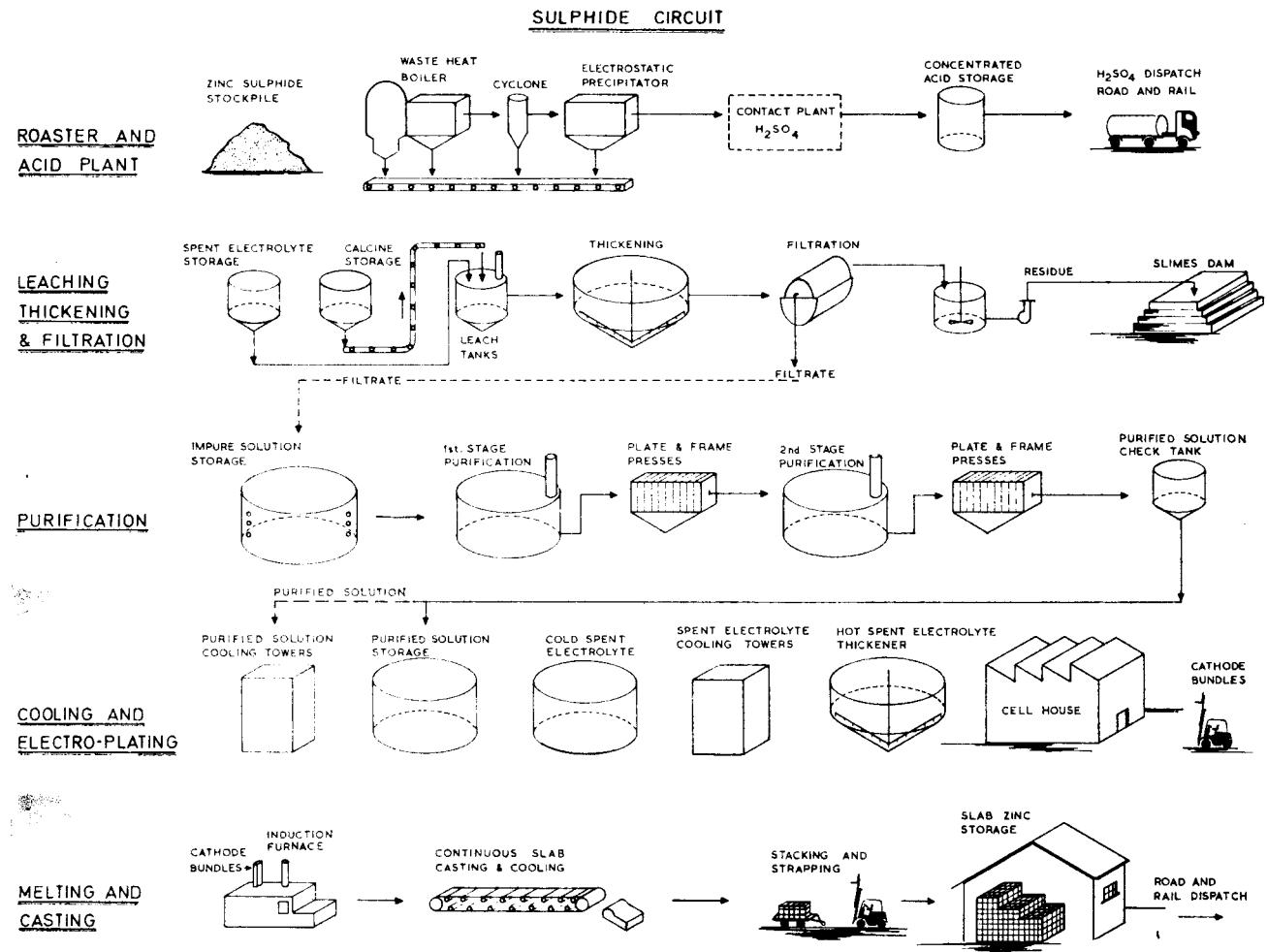
In batch leaching, an accurately measured quantity of cell acid is introduced into the agitator, followed by a charge of calcine, accurately weighed to represent the stoichiometric quantity of zinc to be dissolved. The calcine is introduced as rapidly as possible into the acid, and the free-acid strength falls quickly to a level at which the attack on iron minerals is minimized. Agitation is continued until the readily available zinc has been dissolved and the pulp is just basic. The desired degree of basicity is achieved by fine additions of either acid or calcine.

In continuous leaching, dissolution takes place in a train of agitators. Into the first of these agitators, continuous additions of both acid and calcine are made in the correct proportion to enable complete dissolution of the zinc to take place while at the same time ending up with the desired basicity. By this means the dissolution of iron is maintained at the desired level.

Liquid-solids separation of the leached pulp finally takes place in a combination of thickeners and filters, which range from the pressure leaf batch type to continuous drum machines.

PURIFICATION

Although neither filtrate nor thickener overflow is normally a clear solution, it is unusual to provide clarifiers ahead of purification. The usual procedure is to store impure solution in tanks fitted with steam coils in which the solution temperature is raised to between



70 and 80 °C. Thereafter, the solution is transferred to the purification tanks, and here, again, the operation is either on a batch basis or continuous.

The usual, although not invariable, technique is two-stage purification, zinc dust being added to each stage for the precipitation of impurities. The first-stage tanks are fitted with steam coils, and the temperature of the solution is raised to 90 °C and higher. The solution is then dosed with copper sulphate and arsenic trioxide, after which coarse zinc dust is added in batches until the impurities have dropped to the desired level. This is detected by continuous sampling and testing for a pilot impurity, which is usually cobalt. The solution is mechanically agitated in a manner that maintains the solids in suspension without introducing air. The suspension of partially purified solution is then pumped through plate-and-frame presses for the removal of the pre-

cipitate, which contains all the impurities except cadmium. Cadmium is removed in the second stage of purification, which is carried out at the now ambient temperature by the addition of fine zinc dust only. After the second stage, the suspension is again filtered in plate-and-frame presses and is transferred to a holding tank, from which a sample is taken for analysis by the control laboratory. When the all-clear is given, the solution is available for transfer to the electrolysis section.

The two filter cakes are salable byproducts. The first, known as copper cake, is sold for its copper content, while the second is sold for its cadmium and zinc content. Most plants usually provide a cadmium-treatment section in which cadmium metal is recovered from the second cake and is sold as such.

The reason for the use of plate-and-frame presses, which have to be cleaned manually, is that cleaning

must be thorough. Any precipitate left in the press or on the cloth will immediately oxidize and redissolve when the press is brought on-stream, thus contaminating the next batch of solution.

ELECTROWINNING

As discussed previously, the high standard of purity required of solutions for the efficient electrowinning of zinc metal is the overriding factor influencing techniques in the leaching, filtration, and purification processes. The comparative purity of the solutions will also determine the optimum current density for any circuit. The purer the solution, the greater the current density that can be used. Current density, in turn, determines the area (and number) of electrodes it will be necessary to provide for a particular duty. A low current density will be about 300 A/m², while a high figure could be about 800 to 1000 A/m². Whereas it is essential to eliminate most

metals from the cell acid, iron can be tolerated to a limited extent, and a limited amount of manganese is actually a necessary constituent of the electrolyte. Iron finds its way into the solution via the suspended solids present in the impure filtrate and thickener overflow, and is redissolved in the purification cycle. Manganese is deliberately added to the leach circuit as manganese dioxide and is not removed by zinc purification. The electrolyte in a healthy cell circuit is given a deep wine colour by manganese, which has a twofold effect: to assist electrodeposition of the zinc, and to deposit as a manganese dioxide scale on the anodes and a heavy sludge in the bottom of the cell. This necessitates a regular programme of anode cleaning and sludge removal. This sludge is usually returned to the leaching circuit, where the manganese is redissolved and participates in yet another closed cycle.

As previously mentioned, the anodes consist of lead sheets that are usually perforated to assist the circulation of electrolyte. These anodes are cast from lead alloyed with a small percentage of silver (1 per cent or less) and have cast-in copper header bars. The cathodes, on the other hand, are usually fabricated from aluminium sheet and are welded to cast aluminium-alloy header bars fitted with copper contacts.

Zinc is deposited into the cathode as a smooth sheet, which is allowed to reach a thickness of 4 to 6 mm before being stripped. Adhesion to the cathode is firm, but the sheet can readily be removed by loosening a corner and prising it free. The lengths of plating-stripping cycles will depend on the current density employed, and usually last from 16 to 48 hours, 24 hours being common. Stripping is assisted by prevention of the growth of zinc round the vertical edges of the cathode sheets. This can be done by covering these edges with rubber, plastic, or wooden edge-strips.

MELTING

The stripped cathode sheets are accumulated and melted in furnaces, which in the older plants are fired by

gas or oil. In the new plants, induction furnaces are used and the molten metal is cast into suitably sized ingots or slabs in mechanical casting machines.

The usual grade of electrolytic zinc metal is Zn 2 with a zinc purity of 99.95 per cent or higher. The use of modifying agents in the cell circuit makes it possible to produce Zn 1, or special high-grade (SHG), which assays more than 99.995 per cent zinc.

MATERIALS OF CONSTRUCTION

The range of materials that can be used for the construction of equipment in an electrolytic-zinc plant is limited. Almost all items of equipment that are in contact with solutions or pulps must be capable of withstanding attack by hot, dilute sulphuric acid and saturated zinc sulphate solutions. In the cell house, it is essential that non-conductive materials are employed to prevent current leakage and to safeguard the operators.

In many of the older plants, wooden vats are used in the leaching and purification sections, and the cooling towers, too, are made of timber. Electrowinning cells are usually cast out of concrete and lined with lead, while wood is used extensively for the launders.

A factor that has continually to be borne in mind in the design of a plant is that many of the solutions are saturated, not only with zinc plus magnesium sulphate, but also with calcium sulphate, and that the scaling of gypsum becomes a real problem and calls for frequent cleaning of pipes, cooling towers, and other items of equipment.

Modern zinc plants make extensive use of stainless steel, fibre glass, and other modern industrial plastics. Care must be taken to prevent electrolytic corrosion of stainless steel when it is used for equipment in the purification and electrolytic sections. Any impurities, such as nickel, that might enter the circuit solutions as a result of stainless-steel corrosion could cause havoc in the cell house.

FLOW CIRCUITS

The pilot-plant results showed that the use of conventional circuits

would result in good recoveries from the concentrates tested. However, one problem emerged; namely, that the Waelz oxide tested contained sufficient fluorine to give rise to serious doubts about whether cathode zinc would stick on the aluminium plates. It therefore became necessary to make provision, either to eliminate the fluorine from the Waelz oxide by installing a second-stage densifying kiln, or to deal with the fluorine in the Zincor plant. It was eventually decided to deal with the problem at Zincor in the following manner. Two completely separate circuits would be provided, one to treat low-fluorine concentrates and the other to handle high-fluorine oxides. In the cell house, provision would be made, first, to plate the cathodes in the fluorine-free solution and, when a thick layer of zinc had been deposited on the aluminium cathode, to transfer the latter to the high-fluorine circuit for plating to be completed on the zinc base. It was argued that the fluorine present in the solution used for the second stage of electrode position would not cause the initial layer of zinc to stick to the aluminium sheet. This argument has, in practice, been found to be correct.

A strange feature of the pilot-plant testwork was that sticking never took place, even with solutions having as high a concentration of fluorine as 300 p.p.m. Only when these solutions were artificially spiked to raise the concentration to 1000 p.p.m. could sticking be induced.

DECISION TO PROCEED WITH CONVERSION

After the conclusion of the pilot plant testwork, a technical and financial feasibility report was issued (February, 1967), and this was followed by the decision to proceed with the formation of the Company and with the conversion of the redundant uranium plant to an electrolytic-zinc smelter.

In broad outline, it was agreed that action should proceed according to the following general principles.

The new Company would apply to the Departments of Planning, Commerce, and Bantu Administration for the necessary permission to

erect a zinc plant at Vogelstruisbult on the basis (a) that this was to be a replacement industry for a dying mine, and (b) that it would be a strategic industry providing the country's requirements of a previously imported commodity.

The principal advantages of the conversion were as follows:

- (a) services such as power, road and rail access, water, housing, accommodation for Bantu, and a near-by labour force were available;
- (b) Vogelstruisbult was within a 75-mile radius of over 90 per cent of the inland consumers of zinc; and
- (c) items of plant equipment and buildings were available for conversion to use in the new plant.

The main disadvantages were that the items of equipment referred to were not of the type normally employed in zinc plants, and that a layout taking full advantage of the equipment and building would result in a somewhat unsatisfactory flow path of materials.

However, the advantages appeared to heavily outweigh the disadvantages, and a Consultant was engaged to assist with the planning of the conversion.

BASIC DESIGN PRINCIPLES

The following broad principles were to be used as the basis for the design of the new plant.

- (1) Initially, the daily capacity would be 100 tons of slab zinc of electrolytic grade.
- (2) There would be two completely parallel plants, each with a capacity of 50 tons of slabs per day. One circuit would treat high-fluorine material (Waelz oxide), and the other would treat material low in fluorine (sulphides). Sticking of cathode sheets in the oxide circuit would be prevented by pre-plating of cathodes in the calcine (sulphide) cells.

ROASTING AND ACID PRODUCTION

The new roasting plant would employ the most efficient fluidized-bed roasters for zinc concentrates, and would be required to produce a maximum of 0,30 per cent sulphide

sulphur in the calcine with, at the same time, a minimum of sulphate production. The plant would incorporate waste-heat boilers and hot electrostatic precipitators. It would be followed by a contact acid plant and would make use of the latest double-absorption techniques to ensure maximum conversion of sulphur dioxide to sulphur trioxide (more than 99,5 per cent conversion). When the contract for the roasting and acid plant was eventually issued, an acid capacity of 170 tons per day of monohydrate was stipulated and two roasters were specified, each to be capable of roasting sufficient concentrate to produce 85 tons of acid per day (equivalent to a capacity of more than 50 tons of slab zinc each per day). Initially, one roaster was to be used to burn zinc blonde, while the other would start up on pyrite but would be required to switch over to blonde at some future date. The latter requirement presented interesting problems, as the usual zinc concentrate contains about 30 per cent sulphur, compared with 40 per cent and more in pyrite. The flexibility was eventually achieved by the provision of removable cooling coils in the bed of the pyrite roaster.

It was anticipated that the pyrite required for the second roaster would be massive ore from a mine to be opened up in the Northern Cape and that it would be necessary to make provision to mill the ore at Zincor.

RAW-MATERIALS HANDLING

Concentrate would be delivered by rail, and the final zinc and acid produced would leave the property by either rail or road.

There was a rail spur to the plant site, but it needed to be doubled and rerouted to a certain extent. In addition, the siding at Struisbult would have to be redesigned and rebuilt, in collaboration with the railway administration.

The existing all-concrete gold-plant mill building, which had previously housed four milling units (each consisting of one 9 ft by 10 ft ball mill and two 6 ft 6 in by 20 ft pebble mills), was to be cleared and used for the storage of concentrates.

The existing mill ore-feed bins were to remain intact to receive pyrite ore, and one 9 ft by 10 ft ball mill was to be retained in position for the dry-grinding of massive pyrite ore.

A suitable arrangement of conveyor belts would be provided to feed concentrate from the store to the roasters.

Calcine and Waelz oxide were to be stored in available pachuca agitators (33 ft in diameter and 48 ft in height), two for calcine and four for Waelz oxide. Only one of the calcine tanks (previously a uranium-pulp stock tank) was correctly sited. The remaining five were gold-plant agitators that would have to be moved to new positions. Provision was made to cool and dry-mill the calcine and then to transfer it pneumatically in a continuous pressure system to the stock tanks. The mill earmarked for this purpose was to be transferred from the manganese-preparation section to a position adjoining the roasters. Further handling of the calcine from stock tanks to agitators was to be effected by bucket elevators and screw feeders via weigher-feeders.

The handling system for Waelz oxide was designed to incorporate a receiving bin below the rail track, followed by a series of screw feeders and bucket elevators.

LEACHING SECTION

Batch leaching would be practised in four existing rubber-lined pachuca tanks (22 ft 6 in by 45 ft high), two being required for the leaching of calcine and two for the leaching of oxide. In addition, two existing similar tanks, which adjoined the agitators, would be used for storing recycled spent electrolyte required for leaching, and one further existing agitator was to be used as a surge tank for Waelz oxide between the true storage tanks and the agitators.

The existing manganese and limestone preparation section, which consisted of two 25 ft by 12 ft thickeners, was to be acid-proofed by the addition of a rubber lining, and augmented by two existing 10 ft by 10 ft acid-proofed agitators and one 15 ft by 10 ft acid-proofed dissolving tank for the preparation and feeding of manganese dioxide

sludge to the leaching tanks.

The leaching of calcine called for the partial filling of leaching tanks with spent cell acid, followed by the simultaneous rapid transfer of calcine and the remaining quantity of cell acid in measured quantities designed so that the leaching cycle finished with neutral or slightly basic pulp. It was planned to control the dissolution of iron in this way. Waelz oxide, being free from iron and having been received as a pelletized material, would undergo a somewhat different leaching procedure. Pellets and cell acid would be fed simultaneously into a scrubber (mounted above the leaching tanks), in which disintegration of the pellets would occur and from which the pulp would flow into the agitators.

SOLIDS-LIQUID SEPARATION

The flow in this section was to be countercurrent separation, utilizing two thickeners and two filters in series for each of the two circuits. Use was to be made of three existing concrete thickeners and one existing tank, which it would be necessary to convert to a thickener. Four of the existing ten acid-proofed rotary filters were to be recommissioned.

Two remaining 50 ft diameter tanks would be acid-proofed and fitted with steam-heating coils for storage of filtrate prior to purification. All thickeners and storage tanks would require acid-proofing.

The filter building would be retained, and the portion previously occupied by non-acid filters would be cleared and used as a plant workshop.

Compressors and vacuum pumps were to be recommissioned.

PURIFICATION SECTION

This section was designed to incorporate pressure clarification followed by batch purification in two stages, the first at temperatures in excess of 90°C and the second at ambient solution temperature. Pressure filtration in plate-and-frame filters would take place after each purification.

The building that had housed the ion-exchange cum precipitation plant was to house the purification section,

and three existing eluate- and effluent-storage tanks, 30 ft in diameter by 12 ft in height, were to form the nucleus of the purification tanks. Eight tanks in all would be required, four in each circuit, of which two would be first-stage tanks and two second-stage tanks. The first-stage tanks would be heated with steam coils, and would be lead-lined and lagged with insulation against heat losses. The unheated second-stage tanks would be rubber-lined. All tanks would be fitted with agitator mechanisms capable of imparting satisfactory agitation without effecting aeration. It was necessary to clear the building and to erect an elevated, acid-proofed working floor. In addition, the building required extending in order to accommodate two pressure leaf clarifiers, four thirty-plate 30 in by 30 in pressure filters, a control cabinet, and the eight 30 ft diameter tanks. The clarifiers consisted of two 60 in diameter horizontal stainless-steel pressure filters, each containing twenty 41 in by 44 in canvas-covered filter leaves. Steam for heating the impure-filtrate storages and the first-stage purification tanks was to be derived from the waste-heat boilers in the roaster plant.

In addition to the basic equipment mentioned under this section, provision had to be made for adequate venting of the purification tanks and safe discharge of the gases produced during purification, the most dangerous of which could be arsenic. Provision had also to be made for zinc dust (coarse for first-stage purification and finer for second-stage) to be stored at each purification tank and added in accurately measured quantities by means of weigh-feeders.

Provision was also to be made for a conveniently placed information centre and control room on the operating floor.

Before purified solution is transferred to the cell house, it is necessary to give it a final check for purity, and for this purpose two holding tanks would have to be provided. Existing 15 ft by 45 ft pachuca tanks were to be acid-proofed and resited for this purpose.

CELL HOUSE

The building that had housed the flotation plant attached to the uranium plant was set aside for use as the new cell house.

It was badly placed geographically, and the moderately hot saturated solution would have to be transferred across the area occupied by the thickeners, the filtration building, the leaching plant, the acid plant, and the cell house itself, to the tanks that would be provided as storage for purified solution. On the other hand, it was conveniently situated for future extensions, for the erection of peripheral equipment, and for the shipping of slab zinc by rail. It would be necessary to clear the building completely, to demolish the existing raised floors and erect a new working floor at the correct level, to move certain pillars, and to open up the roof to provide for free ventilation. In addition, the brick-panelled walls would have to be replaced by ventilation louvres at ground level to provide for through-draught. The cell house was designed to incorporate two completely separate circuits. The basic units decided on were as follows: aluminium cathodes made up of 3/16 in plate welded to aluminium header bars, the submerged surface measuring 20 in by 30 in; lead cathodes, 5/16 in thick and cast in lead alloyed with 0.75 per cent silver, the lead to be cast round copper header bars and the submerged area to measure 19½ in by 28 in, and to be perforated with 1 in diameter holes spaced at 3 in centres; individual cells to be cast in concrete and lined with lead and to accommodate 40 cathodes and 41 anodes, the cells to be arranged side by side in rows of twelve cells per row and each circuit to be provided with ten rows of cells. Thus, each circuit would consist of a total of 120 cells containing 4800 cathodes, and the plating area in each cell would be 332 ft² and the total plating area per circuit would be 39 840 ft².

The electric circuit provided for parallel flow across each cell, with cells and banks in series. Thus, the voltage drop across each cell would be 3½ V and across the entire circuit 420 V. The current density would be about 70 A/ft², and pro-

vision would have to be made for a total load of 23 240 A per circuit. When the orders were finally placed, they were for two rectifiers, each with a capacity of 27 000 A and 680 V.

It was decided that all launders in the cell house should be of resin-impregnated fibre glass.

COOLING TOWERS AND ELECTROLYTE STORAGE

In the layout of the solution circuit, it was decided that the purified solution should be cooled, after it had been piped to the cell-house area, in three evaporative cooling towers, and that the solution should be pumped from the holding tanks at the purification section to those cooling towers. The cold solution, prior to joining the cell circuit, would then be stored in two rubber-lined steel tanks 50 ft in diameter by 8 ft.

To achieve a high flowrate of electrolyte through the cells, and thus to ensure a minimum rise in the temperature of the solution as it passes through the cells, and also to ensure uniform distribution of solution to each cell, provision was made for a high flowrate through the cells—equivalent to the circulation of 4,5 tons of cell acid for every ton of purified solution. Thus, the purified solution would be added to 4,5 times its volume of cell acid immediately before entering the cell house. It would then be distributed by a suitable system of launders to each of the 120 cells in a circuit. The flowrate through each cell was designed to be approximately 17 gal/min, and it was anticipated that there would be a rise in temperature of 8 to 10°C in each cell. It was necessary to dissipate this additional heat, and a further series of five cooling towers was provided for this duty.

To cater for the manganese dioxide precipitate that usually forms during electrolysis, provision was to be made for the hot spent electrolyte to flow through sedimentation tanks prior to being pumped to the spent-acid cooling towers. For this purpose, it was decided to use two acid-proofed 50 ft diameter thickeners, and to continuously draw off a dilute suspension of settled sludge.

This suspension would be recycled to the plant as the solvent for incoming Waelz oxide or calcine. The relatively clear thickener overflow would then be pumped to the cooling towers and would gravitate to two 50 ft by 8 ft 'cold spent' storage tanks prior to being recycled through the electrolytic circuit.

Except for the thickeners and storage tanks, most of the equipment in the cell house and its peripherals would be entirely new.

HANDLING AND MELTING OF ZINC METAL

Provision for the manual stripping of cathode sheets every 24 hours was to be made by the installation of stripping racks on the cell-house working floor and the provision of a somewhat elaborate system of cranes and crawls to handle 'baskets' of electrodes for both preplating and stripping. This provision for preplating complicated the layout in the cell house fairly considerably.

Provision was to be made to transport the bundles of stripped cathode sheets to one end of the cell house and to deliver them to an adjoining melt house, where they would be melted and cast into salable zinc ingots.

The melt house was to be laid out to accommodate two 650 kVA induction furnaces for melting cathode sheets, feeding one continuous casting machine designed to produce slabs weighing approximately 25 kg each. Slabs were to be tied together into bundles of 1 tonne each.

Provision would also be made for one small alloying furnace and a facility for the manufacture of zinc dust. In addition, arrangements for the handling of dross would be required.

ZINC DUST AND DROSS

The plant for the manufacture of zinc dust was designed as follows.

Zinc metal would be melted in one 150 kVA induction furnace, and the molten metal would be transferred into a graphite crucible having a 4 mm hole in its base. A stream of molten zinc falling through the hole would be blasted by a jet of air and atomized to spherical zinc dust, to be caught in a settling

chamber. The zinc dust thus formed would be sieved on a three-deck screen into plus 6 mm, plus 35 mesh, plus 100 mesh, and minus 100 mesh products, the last mentioned being collected in a bag filter. The two coarse products would be returned for remelting, while the plus 100 mesh and minus 100 mesh products would be used in first- and second-stage purification respectively.

A considerable quantity of dross, equivalent to between 2 and 4 per cent of the total quantity of zinc produced, would be skimmed off the molten zinc in the melting furnace, and provision was made to deal with this product as follows.

The dross would be skimmed directly into bottom-discharge hoppers, which would be hoisted to an elevated dry-grinding rod mill designed to discharge onto a 2 mm vibrating screen. The metallic oversize would be returned to the furnaces for remelting, and the fine dross would be fed into the roasters, where chlorine (derived from the fluxes used) would be eliminated and the oxide would re-enter the dissolution cycle with calcine.

WHERE ZINCOR IS DIFFERENT

Whereas there are many technical areas of similarity between the leaching and solid-liquid separation techniques employed in a uranium plant and those in an electrolytic zinc plant, there are very few items of equipment in conventional plants that are common to both. In the case of Zincor, equipment was adapted to uses for which it appeared to be perfectly satisfactory but for which it had not previously been used elsewhere. The following is an account of operating experience with such equipment over the period in which the plant was being commissioned.

Handling of Waelz Oxide

Waelz oxide as produced by Kiln Products in South West Africa is an extremely fine-grained powder produced by the fuming of zinc and the simultaneous burning of the zinc fume to zinc oxide in a rotary kiln. It is collected in electrostatic precipitators, and in the present case is pelletized on disc pelletizers to a product containing approximately

8 to 10 per cent moisture. This product can be shipped without undue dusting and, when accidentally wetted *en route*, still arrives at its destination in a relatively manageable condition. However, when it was fed into the system at Zincor, which consisted of screw feeders and bucket elevators, it broke down into a non-flowable powder that rapidly built up on the walls of screw-feeder chutes and blades and on elevator buckets, and was the cause of endless stoppages. This material could just not be handled by equipment of this type.

On the other hand, the system was satisfactorily able to handle dry, unpelletized fume or dry-cured pellets. However, once the pellets had been dried, they became extremely hard and could not readily be disintegrated, either by the scrubber originally provided for the purpose, or by a hammer mill or crushing rolls. The only material that the handling system could effectively cope with was unpelletized fume, and it became necessary to request Kiln Products to despatch its product in this form until an effective means of handling pellets could be provided. This was a very unsatisfactory compromise that resulted in dusting losses and in a health hazard both in transit and at the Kiln and Zincor plants.

There was a further problem that caused considerable difficulty. When dry powder was fed into the scrubber simultaneously with spent cell-acid (which analysed approximately 35 g/l zinc and 130 g/l free sulphuric acid), it frequently formed a paste and choked the scrubber. This was due to the formation of basic zinc sulphate, which formed a cement with the zinc oxide. On the other hand, when dry zinc oxide powder was fed simultaneously with cell-acid directly into the pachuca tanks, a large quantity of granules formed, and this choked the thickeners and filter pans when material was discharged from the agitators after a batch-leaching cycle.

The solution to the problem eventually proved to be the grinding of pelletized material in a wet ball mill in open circuit in water. It has

been found that it is possible to pump a finely ground pulp containing 67 per cent solids, and that a suspension of this consistency can readily be stored in a pachuca tank. Furthermore, a suspension is a highly satisfactory product to feed into a leaching circuit.

Leaching with Large Pachucas

Large pachuca tanks are commonly employed for gold but are not used as zinc-leach agitators, probably because the degree of agitation is not sufficiently violent. As recorded in the previous section, the simultaneous addition of dry oxide powder and solution to the pachuca agitators resulted in the formation of nodules, which choked the thickeners and caused filter rakes to jam. Because of this, it was decided to convert from batch agitation to running the tanks in series, and to allow the nodules, or 'grit' as they are known at Zincor, to accumulate in the conical portion of the tanks. There are only two agitators in series, and the additions of oxide or calcine and acid are regulated to give an acidity in the first agitator of about 5 g/l. With no further additions of calcine, further dissolution occurs in the second agitator, and the pulp in this tank is very slightly basic, resulting in the precipitation of ferric hydroxide. The pulp now flows continuously to the thickeners and filters, and the previously encountered problems no longer occur. However, the formation of grit has continued to take place and has had to be dealt with in various ways.

The following mechanism is believed to be that by which grit forms. When dry calcine is added to a near-neutral solution, it forms small dry spherules into which the surrounding solution seeps. The environment in the outer surface of the spherule rapidly becomes basic, and basic zinc sulphate forms and cements the spherule into a hard nodule that is not soluble in weak acid.

This problem was largely overcome in the oxide circuit after the wet-milling of pellets had been

introduced and a water suspension was fed to the agitators. Some grit is still formed, but this is believed to be due to unground particles that as such have entered the circuit.

With high-speed agitation and a higher acid strength in the first agitator, the formation of grit would be considerably reduced.

Thickeners and Filters

Since the initial problems were solved, these items have been giving satisfactory service.

Purification, Electrowinning, and Melting

Equipment in these divisions has, to all intents and purposes, been provided *de novo*.

CONCLUSION

The scope of this paper has been limited to the considerations and investigations that resulted in the decision to convert the old Vogelstruisbult Uranium Plant into an electrolytic-zinc smelter, and to the action taken to give effect to that decision.

The plant experienced many teeth-ing problems owing, to some extent, to the putting to unfamiliar use of equipment and materials of construction that had originally been provided for a different purpose. The story of how these problems have been overcome makes fascinating reading, and the hope is expressed that the Zincor staff will be able to present a further paper on their experiences while these are still fresh in their memories.

Zincor is now a highly successful producer of zinc, and the original plant has been pushed to produce a consistent 145 tons per day, or 4350 tons per month, of slab zinc. An expansion programme is already under way and provides for a systematic increase in capacity to 400 tons per day.

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