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FINAL REPORT

Research and Development in the Field of Thorium Chemistry & Metallurgy

Volume III.

Cost Estimate for 1,000 Ton/Yr. Thorium Metal Production Plant

To: Manager
U. S. Atomic Energy Commission
Savannah River Operations Office
P. O. Box A, Aiken, South Carolina
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Date: June 30, 1956

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A. EQUIPMENT

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ABSTRACT

In two separate volumes a complete description of techniques for preparing electrolytic cell feed, and producing granular thorium metal powder by electrolysis, have been presented on a pilot plant basis. This report presents a brief description of the development program leading to the pilot operation and describes in some detail a 1,000 ton per year production plant based on data resulting from the pilot activity. The costs are compared directly with those developed by R. B. MacMullin Associates some two years ago, in which aqueous techniques for the preparation of electrolyte were assumed.

Cost estimates appended hereto illustrate that the turn-key plant cost for a production facility capable of producing 1,000 tons of high quality metal per year is $4,623,000. Amortization of this sum over a five-year period in its entirety yields a metal production price of $1.07 per pound, not including the cost of thorium nitrate.

It is concluded that thorium metal powder can be produced in a most economical fashion by the medium of preparing electrolytic cell feed by fused salt chlorination techniques and converting this material to high quality metal by high temperature electrolysis. Serious consideration should be given to this process should occasion arise in the future for the necessity of producing large quantities of thorium metal.

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COST ESTIMATE
FOR
PRODUCTION OF THORIUM METAL
BY
FUSED SALT ELECTROLYSIS

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I. FOREWORD

During a 4-year period ending June 30, 1956, Horizons Incorporated, under A.E.C. Contract No. AT(30-1)-1335, conducted a research and development program to determine the most economic method for production of thorium metal. The first two years of this operation were spent primarily in determinations of feasible production approaches, the relative merits of several candidate processes, and the selection of a single system which seemed to merit consideration. An intensive investigation of the fundamental properties and variables associated with the process ensued, leading to a rather well-delineated electrolytic process for thorium metal preparation.

Most of the emphasis prior to the summer of 1954 was placed on the fused salt electrolysis of thorium tetrachloride, with comparatively nominal efforts being applied to methods of producing a satisfactory cell feed material. Research quantities of the chloride were prepared by wet chemical techniques in the chemical laboratory on a bench scale.

When interest heightened in 1954 in thorium metal, a shift in program scope was made; almost overnight it was necessary to set up a small scale production unit in order to produce several hundred pounds of metal for evaluation. It was necessary to expand the wet method for cell feed

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preparation and place in service many equipment items that were expediently available, not necessarily well-designed for the particular job. This was done, and for six months efforts were devoted solely to quantity production on a pilot scale of cell feed and metal.

During this period a cost estimate was prepared by R. G. MacMullin Associates, consulting engineers of Niagara Falls, to determine the probable cost of metal production in a plant with an annual rated capacity of 1,000 tons. A 5-day week was proposed. Gross assumptions had to be made, using the wet method for cell feed preparation, and basing operations on very small research scale cells and auxiliaries. A capital investment in excess $6,350,000 was arrived at, together with conversion costs of $1.43 per pound of metal in transforming thorium nitrate to granular metal. It should be noted that a large fraction of the total costs was associated with cell feed preparation.

Preparation of cell feed followed the steps enumerated below:

1) Dissolve thorium nitrate crystals in hot water.
2) Dissolve sodium carbonate in hot water.
3) Mix solutions (1) and (2), precipitating thorium oxycarbonate.
4) Settle the precipitate, decant supernatant liquor, re-pulp in water and repeat the process twice.
5) Filter the thorium oxycarbonate, producing a thixotropic cake.
6) Dissolve filter cake in hydrochloric acid.
7) Add ammonium chloride, 2 mols per mol of thorium.
8) Evaporate excess liquor, forming a thick syrup.

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9) Cast syrup into chill molds and allow to solidify.

10) Crush the cake, charge to dryer and dehydrate at a temperature of 260°C. This is carried out under a hydrochloric acid gas atmosphere, and eliminates most of the water. The product is a thorium ammonium complex chloride, which can be produced water-free.

11) Crush the complex chloride and mix with sodium chloride in proper proportions to constitute a suitable cell feed.

12) Decompose the complex by heating to 600° - 650°C under hydrogen chloride or inert gas atmosphere. This process volatilizes the ammonium chloride, leaving a feed material of thorium tetrachloride and sodium chloride.

13) Fuse the residue and cast into chill molds. Make a mechanical separation between oxide-free cell feed and oxygen-bearing salt; store the former for electrolysis and consign the latter to scrap.

The preparation of cell feed by this method on a small tonnage basis was carried out with reasonable success; however, it is difficult to envision a production plant incorporating this process, especially in view of the materials of construction required in the evaporation phase and others.

In October, 1954, the program emphasis again shifted; larger scale pilot operations were placed secondary to development of a more feasible method of cell feed preparation and engineering of

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prototype electrolytic cell equipment. The activity met with marked success, culminating in the operation of a cell feed preparation system with batch capacity of 1,000 pounds, and a prototype electrolytic cell capable of continuous operation at a current input of 7,500 amperes.

The purpose of this estimate is to provide cost data comparable to those now available, eliminating the wet chemical method of cell feed preparation, substituting a direct chlorination technique, and modifying electrolytic plant and operating costs in accordance with pilot plant experience, the latter not having been available to MacMullin Associates. It is our belief that the costs developed below more nearly represent probable costs for production of thorium than those presented by MacMullin, since they reflect major improvements in processes, equipment design and operating techniques.

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II. THE PRODUCTION PROCESS

A. Preparation of Cell Feed

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A chlorination process has been developed which would most economically operate with a feed consisting of thorium oxide and carbon. However, in order to integrate the process with spent fuel element recovery products and encompass a method of returning process wastes to the operation, it is assumed that crystalline thorium nitrate will be the starting raw material.* The following procedures are required in order to convert nitrate into cell feed:

1. Dissolve thorium nitrate crystals in hot water.
2. Dissolve sodium carbonate in hot water.
3. Add lampblack to sodium carbonate solution and maintain in suspension with suitable wetting agents.
4. Pump solution (1) into solution (3), precipitating thorium oxycarbonate on the lampblack.
5. Filter thorium oxycarbonate-carbon precipitate, discharging to continuous belt or tunnel dryer.
6. Continuously dry, to maximum temperature of 300°C; discharge into storage hopper for blending.
7. Mix (6) with sodium chloride-potassium chloride and feed continuously into fused salt chlorinator.
8. Chlorinate the mix at 725°-750°C with chlorine, using compressed chlorine from the cells and some make-up chlorine as required.
9. Tap cell feed directly into holding buggy, or cast into into chill molds, crush and store in hopper as feed for cells.

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* Note: If associated with a spent fuel element recovery plant, a solution of 10% thorium as nitrate would be available, eliminating one of the dissolving tanks, one or two crystallizers in the nitrate dehydration operation and much of the solids handling costs. However, these savings are not reflected in the cost data presented herewith.

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B. Metal Production

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A nearly continuous fused salt electrolytic process for metal production has been developed. Cells of all-nickel construction, lined with graphite, are rectangular in shape with approximate dimensions 24" x 48" x 48". A cathode 10" in diameter, water- and air-cooled, is centered in each cell. The crucible serves as anode, metal being plated on the tip of the cathode as a sponge-like agglomerate of granular metal crystals interspersed with salt.

Deposits weighing more than 200 lbs. (including salt adhering to the metal) have been removed from a cell operated at 7,500 amperes; it is expected that each cathode will contain 100 lbs. of useable metal in a 10,000 ampere cell, and that 8 deposits will be produced in each 24-hour day.

Make-up cell feed will be charged continuously to the cells at a rate of 1-1/2 to 2 lbs. per minute in order to maintain bath level and provide constant thorium content in the bath.

Cathode deposit removal is effected by raising the cathode into a removal chamber area, stripping it mechanically and dropping the deposit into a water-cooled receiver. The cathode is then returned to the bath and electrolysis resumed. This differs from the technique used in pilot operations.

Metal deposits are removed after cooling by valving off the receiver, disengaging it and dumping. The chamber is replaced, evacuated and flushed with argon, and is then ready for the next deposit.

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Start-up and operating procedures are as follows:

1. After complete cell assembly, evacuate with vacuum pumps; eliminate leaks and bake under vacuum at 800°C.
2. Break vacuum with argon; charge 2,000 lbs. sodium chloride, melt and hold molten at 850°C for 24 hours; tap the cell. (This soak removes many impurities in the graphite and saturates the furnace with salt.)
3. Charge 1,500 lbs. sodium chloride; melt; add cell feed to desired thorium concentration (5 to 15%). Fuse the entire charge and lower temperature to 800°C.
4. Pre-heat cathode 2" above bath for 20 to 30 minutes, insert cathode into bath under polarizing voltage and hold for 20 to 30 minutes to attain thermal equilibrium

5. Initiate full D.C. current on cathode; start cathode cooling air within 5 minutes.
6. Electrolyze at 10,000 amperes for 27,500 ampere hours. Cut D.C. power and cooling air; apply polarizing potential.
7. Raise cathode to removal position, strip deposit and lower cathode into bath with polarizing potential still applied.
8. Repeat steps (5), (6) and (7) ad infinitum.
9. Maintain continuous feed to cell during electrolysis. Compress chlorine evolved in electrolysis for use in the chlorinator.
10. When deposit is cold (less than 80°C - 1-1/2 to 2 hours), remove receiver, dump deposit, replace receiver, evacuate and flush receiver with argon.

C. Metal Recovery

Aqueous techniques are applied to the problem of metal recovery. As a result of the coarse, crystalline nature of the thorium granules it is possible to remove occluded salt by the expedient of counter-current or fluid bed washing in water.

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Some thorium salts are present in the metal deposits and hydrolyze when contacted with water, probably forming a water-insoluble hydrated thorium hydroxide. If not removed, the coating reports in metal ingots as the oxide after melting. An acid treatment on completion of the water wash therefore is resorted to in order that oxide content of the metal is held to a minimum.

Acid treatment is followed by a further water wash and vacuum drying. The granular metal product should be packaged and stored under inert atmosphere, since it oxidizes slowly on standing in air.

Sequential metal recovery operations are presented herewith:

1) Crush cathode deposit to 20 mesh maximum size.
2) Charge crushed deposit into counter-current, pulse bed washer, recirculating a suitable quantity of water to remove salt content.
3) Acid treat metal to remove oxide-forming coatings.
4) Rinse metal with water to remove traces of acid.
5) Vacuum-dry.
6) Package under inert atmosphere.

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III. PRODUCTION PLANT

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A. Plant Description

1. Cell Feed Preparation

Figure I in the Appendix presents a schematic layout of the entire production plant and auxiliaries.

Raw Materials will be received by truck shipment in fiber drums and paper bags, and will be stacked on pallets by fork truck. At periodic intervals the thorium nitrate will be weighed out in proper proportions sufficient to make 2 days' operating batch. Twenty tons will be required. A sample for thorium content will be submitted to the laboratory, and should assay 40 to 42 per cent.

A corresponding quantity of soda ash will also be weighed out, some 17,500 lbs., approximately.

Three tanks will be used to prepare thorium oxycarbonate in accordance with the reaction,

$$Th(NO_3)_4 \times H_2O + 2Na_2CO_3 \xrightarrow{(C)} ThOCO_3 + 4NaNO_3 + CO_2\uparrow + XH_2O$$

The first, Tank "A", will be used to dissolve thorium nitrate in hot water, and will have a volume of 25,000 gallons. Tanks "B_1" and "B_2" will be used alternately to dissolve soda ash and precipitate thorium oxycarbonate. Each tank will be constructed of cypress with a reinforced concrete, resin-coated base, and will have a capacity of 50,000 gallons.

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In sequential operation, Tank "A" will be filled with 18,700 gallons of hot water, to which will be charged 40,000 lbs. of thorium nitrate. Tank "B_1" will be filled with 22,700 gallons of hot water, 17,500 lbs. of sodium carbonate being added during the fill cycle. A small quantity of wetting agent, such as Pluronic,* will then be added and 1,300 lbs. of lampblack will be charged. Agitation will produce a uniform suspension.

With temperature maintained at 80°C by steam sparging, the thorium nitrate solution from "A" tank will be pumped into "B," where precipitation of thorium oxycarbonate on carbon will occur.

The entire contents of "A" tank will be transferred, after which another batch of thorium nitrate solution will be prepared in proper time sequence in Tank "A."

The liquor in "B_1" tank is adjusted to a pH of 7.6 to 8.0 by addition of small quantities of soda ash, then allowed to settle until all precipitate is out of suspension. The supernatant liquid is pumped to waste disposal, it representing an 8 per cent solution of sodium nitrate by weight. This may have a finite value, but on a temporary basis would be pumped to a solution disposal pond.

Additional water is added and the precipitate re-pulped to wash out residual sodium nitrate solution, the volume ranging from 5,000 to 15,000 gallons. Again the precipitate is settled and the liquor decanted to a sump pond. This process may be repeated once, but is not presently deemed necessary.

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* Commercial trade name, Pluronic L61, Wyandotte Chemical Corp.

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Five thousand gallons of cold water are added to Tank "B_1," the slurry agitated and fed by gravity to a continuous string filter as required.

During the time Tank "B_1" serves as feed to the filter, Tank "B_2" will be filled, sodium carbonate dissolved and thorium oxycarbonate precipitated, following the same steps outlined above. In this manner there will be a continuous supply of feed to the filter despite batch-type operations in the oxycarbonate precipitate stage.

Tanks "B_1" and "B_2" will discharge to one of two drum-type string filters, each 8 ft. in diameter and 8 ft. wide. The filter cake will be fed continuously to two tunnel-type woven belt dryers, each 8 ft. wide and 35 ft. long, with belt speeds synchronized to match the string speed on the filters.

Heat supplied to the dryer will be of two types: pre-heat in air blown through the dryer in a counter-current manner to remove moisture, and radiant heat from electrical resistance elements mounted in a shield above the belt. Glo-bars will be used for the latter purpose.

An alternative method of heating will encompass muffle-type, gas-fired heating units, the choice being a function of fuel availability and power economics.

Filter and dryer capacities are designed such that one system can handle the entire plant requirements. The tandem system will be used when filter or dryer repairs are required, and for additional capacity in the event of plant expansion.

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Dry thorium oxycarbonate in granular form discharges into a bin feeding a bucket elevator. The elevator discharges into one of two storage bins which in turn feed weigh skips that convey the feed directly to the chlorinators. Use of two separate storage bins is considered desirable in order to provide for thorium accountability on a batch basis, checking the over-all efficiency prior to chlorination.

Adjacent to the thorium oxycarbonate bins will be situated bins for storage of sodium chloride and potassium chloride. Each of the three hoppers will be serviced by a weigh-conveyor which will proportionate the correct amount of salt onto a cross conveyor. The mixer-conveyor will discharge into one of two storage bins which feed the chlorinators. The blend will be fed by gravity into the chlorinator, using a star-type metering valve driven by a motorized speed reducer to control feeding rates.

Two chlorinators, the outer dimensions of which will be 8 ft. in diameter by 40 ft. high, will provide chlorinating capacity well in excess of the plant requirement. Two units are necessary because of the batch-type of operation. On completion of a batch the chlorinator will be used as a hot storage vessel for molten cell feed, the latter being withdrawn as required for cell operation. Electrolyte will be tapped until the volume of the chlorinator has been depleted 50 per cent, at which point the other chlorinator will be tapped and operations resumed in the half-filled unit. A tap-hole half-way up the chlorinator will be

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used for normal tapping operations. During the chlorination cycle which consumes 20 hours, there will be charged 19,000 lbs. of solids (3,800 lbs. sodium chloride, 2,200 lbs. potassium chloride, 12,300 lbs. thorium oxycarbonate and 700 lbs. lampblack). Rate of charging will be 1,000 lbs. per hour until the entire charge has been converted; following this will be a 5-hour period of bath adjustment with carbon or oxycarbonate in order to clarify the bath. The chlorinator may then be tapped and duplicate operation initiated on the alternate unit.

Chlorine will be fed to the base of the chlorinator at the equivalent rate of 2,500 to 3,000 cu. ft. per hr., in liquid form. Up to 10,000 lbs. of chlorine will be consumed per day. Gases produced in the chlorinator, together with unreacted chlorine, will be ducted through a scrubber system for neutralization and entrained solids removal. These off-gases will consist predominantly of carbon monoxide and carbon dioxide, with minor quantities of air, chlorine, moisture and phosgene. Acid-forming constituents will be neutralized with caustic soda.

Should chlorine conversion be lower than anticipated, provision will be made for refrigeration of the gases to condense chlorine for recycling.

Fused cell feed will be tapped from the chlorinator into a mixer ladle car with a capacity of 5,000 lbs. The ladle car will then transfer molten salt to a holding furnace adjacent to each electrolytic cell once every 6 to 8 hours, returning to the chlorinator to be re-filled after completion of the circuit. The capacity of each holding furnace will be 1,000 lbs. of molten salt.

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Each holding furnace will be heated by bath resistance, using D.C. power across embedded resistors during non-electrolyzing periods as a power source; i.e., the holding furnaces will be heated only during a 15 to 30 minute period every 3 hours. Insulation will be provided to maintain the bath with minimal heat loss during the remainder of the time. Auxiliary connections to A.C. power will be provided for emergencies as a secondary source of heat. Copper bus jumper bars to by-pass the furnaces also will be at hand should it be necessary to take a unit out of service or shut down a cell.

Transfer of molten salt from holding furnaces to electrolytic cells will be on a continuous argon pressure surge basis. Gradually increasing pressure will displace a small volume of salt which will be forced through a multiplicity of holes in a common wall behind a baffle plate between the holding furnace and cell, thus providing a uniform, continuous feed of make-up electrolyte.

B. Electrolytic Operations

The key metal production unit is an electrolytic cell operating at a D.C. power input level of 10,000 amperes semi-continuously. Each cell, constructed of nickel and graphite, holds 2,500 lbs. of molten electrolyte. Internal dimensions provide an electrolyzing chamber 24" x 24" x 48" deep; on each end is a heater compartment 12" x 24" which houses two heating assemblies. On one side is a make-up salt compartment 24" x 24" which also houses two heating assemblies. The

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make-up salt compartment is separated from the cell proper by a graphite baffle plate. Heating elements are bayonet-type, constructed of graphite and encased in fused quartz tubes.

Cell lining consists of graphite slabs 2-1/2" to 3" thick, segments being joined with lap, step or tongue-and-groove joints cemented together with graphite paste. The grade of graphite is immaterial, except that it should be of reasonably high purity and uniform in structure.

Between the outer nickel shell and the graphite crucible a layer of lampblack is tamped in, to a density of 35 to 45 lbs./ft.^{3}. Layer thickness is 2" to 2-1/2" on sides and ends, and 4-1/2" on the bottom. Acid-proof ceramic bricks, spaced properly, 4 to 8 in number, are bedded into the bottom to support the crucible and eliminate problems associated with non-uniform packing of lampblack.

Wedges of graphite between the shell and the crucible provide good electrical contact, anode leads being attached to the outer shell for convenience in design. Shell sections in areas where graphite wedges and bricks are located should be provided with cooling jackets through which cooling air is blown.

At one end of the cell a tap-hole is provided. This consists of a Graph-i-tite* pipe surrounded by an insulating sleeve and nickel cylinder, and a graphite tee to which a power lead may be attached. A water-cooled safety rod is also included to prevent premature tapping. By applying an electrical current to the tap tube, it acts as

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* Carbon impregnated graphite

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a resistor, heating up and allowing molten salt to pour out. Salt flow may be interrupted by inserting the safety rod and cutting off the heating current. An A.C. welder adequately serves as the power source for heating the tapping device.

A lid is provided for each cell which forms a vacuum-tight closure. Graphite heat shields and water cooling of flanges are adequate to prevent thermal damage. The lid contains ports for heating elements, sampling, thermocouples and metal removal.

Cathodes are 10" diameter tubes 14 ft. long, water-cooled except for the lower 36". Walls are nickel forgings or centrifugal castings 1-1/2" thick, and house 3" copper piping which ducts cooling air to the lower tip. A blower provides cooling air at up to 20,000 cfh under 10" of water head.

The cathode tip is fabricated of Hastelloy-C. Dimensions are 10" O.D. x 20" length, closed on the end and partially filled with copper to improve heat extraction.

The cathode shaft above the tip has an 8" diameter and is provided with a split graphite protective cylinder. Sections of graphite are held in place by a split ring of nickel which bolts to the main shaft.

A thermocouple well is positioned inside the shaft and cast into the copper bottom.

Removal of cathode deposits is effected by raising them into a removal chamber attached to the lid, stripping the hot deposit mechanically and directing the deposit to a cooling chamber which can be valved off for deposit removal at the proper time. As soon as the deposit is stripped the cathode is returned to the bath to begin another electrolysis.

Electronic switch systems will throw the D.C. current into a parallel circuit consisting of the cell and the make-up salt furnaces when the cathode is moved upward. When the deposit clears the bath, all current will flow through the make-up furnace, none through the cathode. On lowering the cathode, reverse switching will again place full current on the cathode circuit. It is expected that no more than 10 per cent of the total day will be consumed in deposit removal, and that 90 per cent or more of the operating day will actually be utilized in full-scale electrolyses.

There will be nine electrolytic cell sites in the plant, each equipped with a complete cell and all necessary auxiliaries. Only seven will be required in order to produce the requisite tonnage, two being held in stand-by to provide for such contingencies as over-estimations on production rate, taking cells out of service for major rebuilding, expansion of plant capacity, etc.

A single D.C. power source will be provided, all cells being connected in series. The power supply will have an output capacity of 11,000 amperes at 135 volts and will be of the constant amperage control type. Rectification by germanium is specified. This unit has an operating capacity capable of powering all nine cells simultaneously under normal statistical operating cycles, i.e., not all cells would initiate electrolysis simultaneously.

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Separate A.C. power systems for each cell installation are provided. These consist of multiple tap transformers with infinite variation between taps, field-controlled. Switchgear will include facility for diverting a portion or all of this power from the cell heaters to the make-up salt furnaces if necessary. No A.C. heat will be required during one-third of the electrolysis cycle, only a limited amount will be required during the remainder of the cycle, and up to 60 K.W. will be needed during the cathode extraction cycle.

Expected minimum cell life is six months, after which it may be necessary to tap out the molten salt, cool the cell to room temperature and rebuild the shell. Replacement of the graphite crucible, repacking of lampblack, renovation of auxiliary units and repairs on the cathode may be necessary. Renovation time is conservatively estimated at two weeks with a two-man crew.

C. Metal Recovery Operations

The proposed production operation, based on continuous electrolysis in seven electrolytic cells, will yield 56 deposits, each containing approximately 100 lbs. of recoverable metal, per 24-hour day. Total weight of each deposit will approximate 250 lbs., including unrecoverable metal fines and salt dragged out with the metal deposits.

Metal recovery equipment is designed to process the cathode deposits continuously.

Cathode deposit receiving chambers will be equipped with hydraulically closed and sealed bottom covers. Once deposits are sufficiently cool to

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allow removal, the chamber cover will be opened, allowing the cathode deposit to drop into a waiting shuttle car. Deposits will be taken individually to a chute leading to a crushing table. A short,continuous conveyor belt will deliver cathode deposits to the crushing site which will be equipped with a steam powered impact hammer. The purpose of this table is to crush the deposits from an initial size of 18" to 20" diameter and 36" to 40" length into smaller chunks not exceeding 4" across. A 4" grizzly will be built into the crushing table, and chunks passing through fall by gravity into a hopper feeding a Fitzmill. Two such comminuters are being provided, one being reserved for use in the event of equipment failure or breakdown.

The comminuted product discharged from the Fitzmill will have a maximum size of 20 mesh. Approximately 10 minutes will be required to comminute each 250-lb. deposit.

The Fitzmill discharge resulting from crushing cathode deposits is caught in a surge bin which feeds the material onto a conveyor. The conveyor discharges directly into a counter-current flighted drum washer which removes most of the salt adhering to the metal granules. The washer is continuously fed, at a rate of 15 lbs. per minute, and discharges onto a mesh drain conveyor which is sprayed from above with fresh water. The drain conveyor deposits the partially washed metal in a surge bin which in turn feeds a group of pulse tables, (jigs), each jig being comprised of 12 compartments in a cascade arrangement. Fresh water is fed to the jig system, recycled counter-currently and

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ultimately pumped to the drum washer for the removal of gross salt content. Exhaust hoods are positioned over each washer in order to withdraw any gaseous products which might result from the action of water on small quantities of thorium carbide should there be any present.

Sequential water flow in the washing process is described below.

The cascade pulse washer is divided into 12 compartments, each having a surface grid 24" square and operating with a tapered metal depth averaging 1". Washed metal discharges from compartment 12, and metal to be washed is charged onto compartment 1. Metal movement is by levitation and gravity from 1 toward 12.

Every other compartment will discharge its water into its own sump, alternate compartments being designed to overflow into adjacent units. For example, fresh water will enter sump 11, pulse through the bed and overflow with suspended metal into compartment 12. Nearly-dewatered metal discharges from compartment 12, pulse water passing through a side screen, flowing to sump 12 and being pumped from sump 12 to sump 11. Thus, sump 11 has two water sources: fresh water and recycle water from sump 12.

The pump having sump 12 as feed also pumps to sump 9, at a rate equal to the fresh water influx to sump 11. The system is therefore balanced volume-wise. Water pulsing through compartment 9 overflows to compartment 10 and thence into sump 10. A pump takes water from sump 10 and recirculates a portion to sump 9, the remainder going to sump 7. Thus, the system operates as a group of six 2-compartment tanks, recycling most of their water volume but continuously supplying make-up water to the next higher tank pair at the same rate fresh water is fed to the system. The result is a recycle system incorporating counter-current advantages as well as pulse extraction principles.

The pump fed by sump 2 by-passes a portion of the water to a spray system above the drain conveyor, washing the metal in transit from the drum washer to the jig washer. A collector tank below the drain conveyor acts as a sump for the fresh water, it being pumped to the drum washer for the major salt removal step in the process.

Water flow rates for fresh make-up will be determined primarily by the rate of deposit feed. Final discharge water from the drum washer should not exceed 10 per cent in its salt content. Approximate water consumption for the washing system therefore will be about 1,000 gallons per hour, allowing for vaporization losses, spillages, etc.

Water discharged from the drum washer, containing between 5 and 10 per cent salts by weight, will carry over a quantity of finely divided thorium metal powder as well, mostly particles 100 microns or smaller. This suspension will flow to a small Dorr thickener to allow the metal to settle out prior to neutralizing the wash waters and precipitating soluble thorium values. Metal fines are continuously discharged, dewatered by filtration and recycled to the dryer for re-chlorination in the feed preparation department.

The liquor overflowing from the fines settling tank is pumped to one of two 250,000 gallon Dorr tanks, being partially neutralized with sodium carbonate in transit.

Each Dorr tank has a liquid capacity of approximately one week's throughput. Following the week's receipt of wash waters the alternate tank is placed in service and the filled tank completely precipitated of its thorium values. This is accomplished by the introduction of additional sodium carbonate into the tank, agitating for 8 hours with the addition of a coagulating and precipitating agent, and allowing the material to settle for 24 to 48 hours. Paddle scraping of the precipitated solids then accomplishes removal through the bottom port in the tank, which discharges to a continuous drum filter, 4 ft. in diameter and 6 ft. across. This material, predominantly thorium oxycarbonate interspersed with finely divided thorium metal powder and oxides, is caught on a continuous belt conveyor and transported to the drum filters which serve the feed materials preparation area. At this point it is intermixed with the filter cake from the feed materials filter drums and becomes a portion of the feed into continuous belt driers.

Fully washed metal discharging from the jig washer is chuted into an acid pickling tank maintained automatically with a 5 per cent concentration by weight of sulphuric acid. Make-up acid from a feed tank is metered in as demanded by a hydrometer control system.

Vigorous agitation of the acid tank is accomplished with a series of motor-driven stirrers positioned along the tank, thus providing a suspension of the metal. The tank itself has a sloping bottom and is designed to provide passage down its length in a period of 30 to 40 minutes.

At the discharge end of the acid tank a rake mechanism loads the treated metal onto a washing conveyor which brings the metal out of acid and under a series of jets which spray slightly acidified water over it to remove adherent acid. The partially rinsed metal then passes through a drum washer similar to the one used for salt removal, fresh water being introduced for final acid removal. Overflow water is pumped to the spray head which rinses the metal as it emerges from the acid tank.

The acid treated and rinsed metal is discharged directly onto a coarse filter for gross moisture removal, from which it is charged to a jet-aspirated, continuous vacuum dryer. On passage through the dryer the metal powder, fully dried, discharges into a packaging device which is operated under an argon atmosphere. Metal powder is packaged in polyethylene-lined steel drums for storage pending further disposition.

On a "once through" basis approximately 70 per cent of the total thorium will be recovered as useful metal. The remaining 30 per cent will be reprocessed through the drying, re-chlorination and electrolyzing cycle in order to convert it back to useful metal. The 30 per cent recycle is constituted of thorium in the form of chlorides in salt dragged out with the deposits, and in the form of metal fines which are not recovered as useful metal in the washing operation.

IV. ASSUMPTIONS REFLECTED IN COST ESTIMATES

The assumptions incorporated in cost estimation for a thousand ton per year plant are comparable in general to those assumed by MacMullin Associates in their estimate of May, 1954. For example, identical building costs per sq. foot, erection and installation costs, etc., have been reflected in this report, in order that the two might be compared on an identical basis.

It may be possible to criticize this approach, since MacMullin's report was based on a production plant which would be built in conjunction with the Fernald facility. For any other type of plant this might not necessarily be the most desirable location, nor might the production and operating costs be comparable to those experienced at the Fernald location. However, it appears to be the general consensus of opinion that any type of radioactive metal production plant would be associated with another plant which would concern itself with the recovery of spent fuel elements, fabrication of fuel element components and associated operations. In essence, therefore, the production of thorium metal as outlined in this cost estimate would be one step in a closed circuit containing a number of steps, and hence the assumptions made would be valid. For example, existing facilities would be assumed for a steam supply, warehouse and office space, a change house, etc.

Because of the nature of the radioactivity associated with the plant product and desirability of close control over the operation, certain facilities which normally would be available to units of an overall integrated

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plant have been duplicated, these including a complete analytical laboratory, metallographic and physical metallurgy evaluations of the product, minor first aid facilities, etc.

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V. RESULTS AND CONCLUSIONS [redacted]

Based on the pilot plant developments under Contract AT(30-1)-1335 during the years 1954-1956 a potential production process for the manufacture of granular thorium metal has been outlined. Assumptions generally comparable to those utilized by R. B. MacMullin Associates in their cost estimate of May, 1954, have been reflected throughout the present estimate in order that there might be a direct comparison of resulting manufacturing costs, and hence some indication of the process improvement which has arisen during the past two years.

With reference to the appended cost estimate it may be seen that a conversion cost per pound of $1.07 has been arrived at for converting thorium nitrate to granular thorium metal powder of extremely high purity, acceptable to the Atomic Energy Commission for nuclear purposes. This compares with a figure of $1.43 which was arrived at by Mac Mullin Associates.

It is interesting to note that of the $1.07 total conversion cost 46 cents are reflected in depreciation of the plant itself. Should a longer amortization period be selected than obviously the conversion costs would be reduced thereby.

It has been demonstrated in prototype pilot plant equipment that a high quality granular thorium metal powder can be produced in accordance with the rates and efficiencies assumed in this report. Nearly two tons of metal have been produced during the past year in two short campaigns in a 7,500 ampere electrolytic cell. It has been proven without question that it is possible to operate a large-sized electrolytic cell substantially continuously, withdrawing

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deposits in excess of 200 lbs. in total weight without difficulty. Every aspect of the electrolytic production technique has been proven completely, and large quantities of cell feed have been prepared by direct fused salt chlorination approaches. The only area in which unproven assumptions have been made is in the metal recovery section, where continuous processes have been substituted for batch-type systems which were used in the pilot operations. It is probable that much lesser costs actually will be realized in the metal washing section than have been shown in the estimate, but for contingency purposes the costs have been presented in a pessimistic manner.

Based on actual pilot plant activities and estimates of costs it is concluded that the production of thorium metal of acceptable Atomic Energy Commission grade can be achieved in a most economical manner by utilizing fused salt electrolytic techniques. Consideration should be given to the utilization of this type of operation should occasion arise in the future wherein large quantities of thorium metal will be required.

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APPENDIX

DETAILED COST ESTIMATE

1,000 TON PER YEAR

THORIUM METAL PRODUCTION

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A. EQUIPMENT

(1) Cell Feed Preparation

2 bins, feed, thorium nitrate	$3,800.00
2 bins, feed, sodium carbonate	3,200.00
2 bins, feed, lampblack	600.00
1 tank, wood, thorium nitrate dissolver, 25,000 gal.	2,500.00
2 tanks, wood, oxycarbonate precipitation, 50,000 gal.	8,000.00
2 agitators, tank, thorium nitrate	1,900.00
3 agitators, tank, oxycarbonate precipitation	2,850.00
3 steam spargers, tank, oxycarbonate precipitation	750.00
2 pumps, water booster pressure, 10,000 gph	2,570.00
4 pumps, transfer, thorium oxycarbonate, 5,000 gph	1,520.00
4 control valves, liquid level, filter tank	4,000.00
1 scale, metering, Pluronic	900.00
4 pumps, decant, tank, 5,000 gph	2,600.00
2 filters, drum, 8 x 8, complete	25,000.00
2 feeders, vibrator	3,000.00
2 filter scraper modifications	1,000.00
2 dryers, tunnel	80,000.00
2 blowers, Roots-Connorsville,	1,000.00
2 preheaters, air	5,000.00
2 elevators, bucket	15,360.00
2 bins, storage, thorium oxycarbonate	3,400.00
1 bin, storage, sodium chloride	1,000.00
1 bin, storage, potassium chloride	800.00
4 weigh conveyors, bin discharge	10,000.00
1 conveyor, collector, chlorinator feed	2,500.00
1 discharge, conveyor chute, swivel	2,000.00
2 hoppers, mixer, elevator feed	2,500.00
2 elevators, bucket	10,000.00
2 hoppers, feed, chlorinator	1,600.00
2 chlorinators, 8' dia. x 40'	200,000.00
2 vessels, tapping, pressure-controlled	3,000.00
2 transformers, 150 KVA each, power, multiple tap, 3 phase	8,000.00
1 portable welder	1,000.00
2 compressors, chlorine, including refrigeration unit	50,000.00
1 manifold, chlorine	1,000.00
1 tank, storage, chlorine	5,600.00

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1 sump, water 1 system, circulator, cooling water 2 pumps, cooling water	$3,000.00
2 traps, dust, chlorinator	10,000.00
2 tanks, neutralization, chlorinator gas	3,900.00
2 systems, caustic jets and pumps	6,500.00
2 bubblers, seal, liquid	500.00
1 car, ladle, mixer, 5,000 lb. cap.	10,000.00
1 truck, ladle transfer	4,500.00
1 system, manifold, argon	1,000.00
misc. special tools and rigging @ 1% of equipment	5,000.00
plant instrumentation @ 5% of equipment cost	25,700.00
3 pumps, decant, oxycarbonate tank	1,200.00
	$539,830.00

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(2) Electrolytic Plant Equipment

9 cells, electrolytic	$351,000.00
9 tanks, surge, argon	2,400.00
2 pumps, vacuum, large	5,000.00
1 control, D.C., 1500 KVA, 11,000 amp., 135 volt	100,000.00
2 shunts, 12,000 amp.	1,500.00
9 systems, hoist and monorail	6,000.00
1 system, hydraulic, servo	3,000.00
2 pumps, hydraulic	5,000.00
9 supports, metal receiver, hydraulic	5,000.00
1 manifold, argon	2,200.00
manifold valving, hydraulic	3,000.00
1 manifold, vacuum	3,000.00
9 compressors, air, cathode cooling	11,000.00
9 systems, water cooling, circulatory	5,000.00
1 manifold, cooling air, cell, w/blower	3,500.00
9 transformers, multiple tap, 75 KW	15,000.00
2 receivers, molten salt	3,500.00
18 receiver cans, deposit	7,500.00
1 truck, hydraulic lift, motorized	5,000.00
1 welder, 1000 amp., portable	1,000.00
1 compressor, service	1,000.00
instrumentation @ 5% of equipment	27,250.00
miscellaneous standard & special tools @1%	5,400.00
miscellaneous handling equipment	7,750.00
	$580,000.00

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(2a) Electrolytic Plant Equipment

Electrolytic Cell: (Component Costs per Cell)

Cell base	$ 800.00
Shell, lower, nickel, air-cooled, 1/2" side plate, 1" bottom	9,000.00
Shell, upper, nickel	6,000.00
Chamber, removal, cathode	3,500.00
Valve, removal, deposit	4,000.00
Receiver, deposit	1,000.00
Cathode, nickel, jacketed	4,000.00
Connector, power, cathode	250.00
Stripper, cathode	750.00
System, stripper, hydraulic	400.00
Crucible, cell	1,600.00
Insulation, crucible	100.00
Shields, heat, upper shell	600.00
Wells, thermocouple	100.00
Wells, heater, cell	300.00
Elements, heater, cell	1,000.00
Tubes, graphite, heater well protection	500.00
Tubes, graphite, thermocouple well protection	100.00
Connector, power, anode	200.00
Switchgear, D.C.	3,000.00
Heaters, immersion, pre-melt	300.00
Control, level, electrolyte	250.00
Assembly, tap-hole	750.00
Bubbler, chlorine	200.00
Chamber, dust	200.00
System, probe, bath depth, D.C.	100.00
	$39,000.00

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(3) Metal Recovery Operations

4	dollies, tilting, cathode deposit	$3,600.00
1	conveyor, crusher table	600.00
1	table, crusher	2,100.00
1	grizzly, 4"	600.00
1	chute, grizzly	300.00
1	hammer, impact, steam-powered	1,500.00
2	Fitzmills, knife-equipped, stainless steel	3,600.00
1	hopper, crushed deposit	750.00
1	elevator, bucket	5,000.00
1	chute, discharge	500.00
1	washer, drum, flighted	10,000.00
1	thickener, metal fines, Dorr type	7,500.00
1	pump, Dorr underflow	400.00
1	conveyor, mesh. water-sprayed	10,000.00
1	tank, collector, spray water, conveyor	1,000.00
1	pump, sump	350.00
1	hopper, surge, jig feed, including)	1,500.00
1	feeder, vibratory, jig)	
1	feeder, vibratory, jig)	
1	jig, 12-compartment	45,000.00
7	pumps, transfer, wash water	2,500.00
1	receiver, washed metal)	1,500.00
1	feeder, vibratory, acid tank)	
1	trough, acid	5,000.00
6	agitators, rubber-coated	3,000.00
1	conveyor, discharge, acid tank	1,500.00
1	conveyor, mesh, water-sprayed	10,000.00
1	washer, drum, flighted	10,000.00
1	centrifuge, dewatering	4,500.00
1	bin, feed, vacuum dryer	1,750.00
1	dryer, continuous, vacuum	40,000.00
1	hopper, dried metal, inert atmosphere	1,500.00
1	machine, packing, inert atmosphere	7,500.00
1	conveyor, roller	200.00
	miscellaneous standard & special tools @1%	1,830.00
	plant instrumentation, metal recovery @5%	9,050.00
	miscellaneous handling equipment	3,000.00
2,000	finished product containers	2,000.00
		$199,030.00

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(4) Auxiliary Processing Equipment

2	tanks, Dorr, 250,000 gal. cap.	$50,000.00
1	tank, surge, oxycarbonate decant, 1,000 gal. cap.	1,000.00
2	pumps, surge tank transfer	2,400.00
1	collector, dust, oxycarbonate dryer	3,600.00
1	system, scrubber, chlorine, emergency	5,000.00
1	system, fresh air	10,000.00
1	power source, stand-by, 100 KVA, diesel	25,000.00
1	welder, gasoline-powered, 1000 amp.	2,000.00
1	furnace, melting, sample	1,700.00
1	power source, furnace	800.00
1	system, vacuum, furnace	700.00
1	pelletizer, powder	250.00
4	pumps, transfer, Dorr tank	1,600.00
		$104,050.00

(5) Materials Handling & Service Equipment

1	scale, platform, 15 ton cap.	$2,500.00
1	truck, fork, 5,000 lb. cap.	6,000.00
1	scale, platform, 1,000 lb. cap.	750.00
1	scale, platform, 100 lb. cap.	200.00
200	pallets, wood	1,000.00
2	hoists, Budgit, 4,000 lb. cap. w/rails	1,000.00
4	elevators, bucket type	20,000.00
3	conveyors, roller, salt-handling	1,500.00
1	rig, salt-dumping, mechanical	1,500.00
9	disconnects, AC-DC, electrolytic cell	27,000.00
1	console, control, electrolytic cell	10,000.00
12	hoods, exhaust, scrubber serviced	15,000.00
6	exhausters, dust collector	3,000.00
1	system, water filter	25,000.00
	machine shop equipment	100,000.00
	maintenance shop equipment	25,000.00
	weld shop equipment	20,000.00
1	system, vacuum, floor	9,000.00
1	overhead crane	4,000.00
3	switches, high voltage,	45,000.00
6	pads, transformer	25,000.00
1	motor control center	20,000.00
1	charger, battery	5,000.00
	Total	$367,450.00

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(6) Quality Control, Health and Safety

1	tester, hardness, Brinell	$ 600.00
1	tester, hardness, Rockwell	450.00
1	lathe, shop	700.00
3	sanders, belt, metallographic	200.00
1	wheel, cut-off, metallographic	150.00
3	wheels, polishing, metallographic	250.00
1	microscope, metallographic	1,500.00
1	camera, 4" x 5"	500.00
1	laboratory, analytical, complete	25,000.00
	fire extinguishers	5,000.00
	fire blankets	100.00
	sand pails, shovels, etc.	200.00
	first aid kits	100.00
	first aid station	1,000.00
	radiation monitoring equipment	25,000.00
	masks, fresh air	5,000.00
	system, safety alarm	3,500.00
	Total	$69,250.00

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Equipment Cost Summary

Cell Feed Preparation		$539,830.00
Electrolytic Plant Equipment		580,000.00
Metal Recovery		199,030.00
Auxiliary Processing Equipment		104,050.00
Material Handling & Service Equipment		367,450.00
Quality Control, Health & Safety		69,250.00
Total Equipment Cost		$1,859,610.00
Equipment Erection, Piping & Wiring		1,487,690.00
Buildings, 890,000 ft.^{3} @ \$0.60/ft.^{3}		534,000.00
Roadways, 15,000 ft.^{2} @ \$9.00/yd.^{2}		15,000.00
Sewers		25,000.00
Total Plant Cost		$3,921,300.00
Process Engineering @ 1%	$39,200.00	
Design Engineering @2%	78,500.00	
Mechanical & Architectural Eng. Fee @7%	274,500.00	
Contractor's Fee @3%	117,650.00	
Engineering & Contracting Expense	35,000.00	
Purchasing & Procurement @ 4%	156,850.00	
Total Fees		$ 701,700.00
Grand Total, Turn-key Plant Cost		$4,623,000.00

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B. MANUFACTURING COST ESTIMATE

Metal Production Rate - 2,000,000 lbs. per year

1. Raw Materials

Raw Materials	Quantity	Annual Cost
thorium nitrate	4,825,000 lbs.	no cost
argon	500 cyls.	$16,750.00
sodium carbonate	2,110,000 lbs.	34,650.00
sodium chloride	2,000,000 lbs.	29,900.00
sulfuric acid	300 tons	7,200.00
hydrated lime	50 tons	550.00
caustic soda	20 tons	1,050.00
lampblack	157,000 lbs.	7,300.00
Total, Raw Materials		97,400.00

2. Utilities

steam	30,000,000 lbs.	$22,600.00
electricity	14,500,000 kwh	130,500.00
water	91,000,000 gal.	2,750.00
natural gas	10,000,000 $ft.^3$	5,000.00
Total, Utilities		$160,850.00

3. Repair Materials and Maintenance

Estimated @ 3% of capital investment annually, .03 x $4,623,000	$138,690.00
Cell lining replacements	105,000.00
Supplies, $100/day	36,500.00
Total, Repair Mat'ls & Maint.	$280,190.00

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4. Labor, Operating

(a) Electrolytic Plant

No./Shift	Job	Total Men	Annual Cost
1	shift supervisor	4	$ 21,620
1	feed make-up tender	4	16,440
1	deposit remover	4	16,440
1	shift helper	4	18,140
	Total	16	$ 72,640

(b) Metal Recovery

No./Shift	Job	Total Men	Annual Cost
1	metal crusher	4	$ 19,400
1	metal washer	4	19,400
1	helper	4	18,140
1	drying & packout	4	18,140
	Total	16	$ 75,080

(c) Cell Feed Preparation

No./Shift	Job	Total Men	Annual Cost
1	shift supervisor	4	$ 21,620
2	truckers	8	32,880
1	oxycarbonate prep.	4	19,380
1	chlorinator feed	4	16,440
1	chlorinator operator	4	19,400
2	helpers	8	36,280
	Total	32	$ 146,000

(d) Monthly Labor

No./Shift	Job	Total Men	Annual Cost
1	general foreman	1	$ 6,000
2	shift engineers	2	10,000
	Total	3	$ 16,000

(e) Maintenance

No./Shift	Job	Total Men	Annual Cost
2	shift mechanics	8	$ 41,600
10	mechanics (days)	10	52,000
11	helpers (days)	11	64,800
	Total	29	$ 158,400

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Key to Fig. 1.

A	Shipping & Receiving Area
B	Thorium Oxycarbonate Precipitation Area
C	Chlorinator Area
D	Electrolytic Area
E	Metal Recovery Area

1	Fork Truck
2	Pallet Storage
3	Freight Elevator
4	Thorium Nitrate Storage Bin
5	Thorium Nitrate Dissolving Tank
6	Thorium Oxycarbonate Precipitation Tanks
7	Precipitate Slurry Transfer Pumps
8	Drum Filters
9	Tunnel Dryers
10	Discharge Conveyor
11	Bucket Elevator
12	Thorium Oxycarbonate Storage Bin
13	Sodium Chloride Storage Bin
14	Potassium Chloride Storage Bin
15	Proportioning Feeder-Conveyors
16	Mixing Conveyor
17	Chlorinator Feed Elevator
18	Chlorinator Feed Surge Bins
19	Chlorinators
20	Molten Salt Transfer Buggy
21	Transfer Rail Line
22	Electrolytic Cells
23	Cathode Deposit Transfer Buggy
24	Cathode Deposit Chute
25	Crushing Table
26	Drum Washer
27	Dewatering Conveyor
28	Jig Washer
29	Transfer Conveyor
30	Acid Leach Tank
31	Dewatering Conveyor
32	Drum Washer
33	Vacuum Dryers
34	Packaging Machine
35	Roller Conveyor
36	Dust Collectors
37	Solution Transfer Pumps
38	Thickener-Settler Tanks
39	Fines Thickener Tank
40	Caustic Scrubbers

HORIZONS INCORPORATED

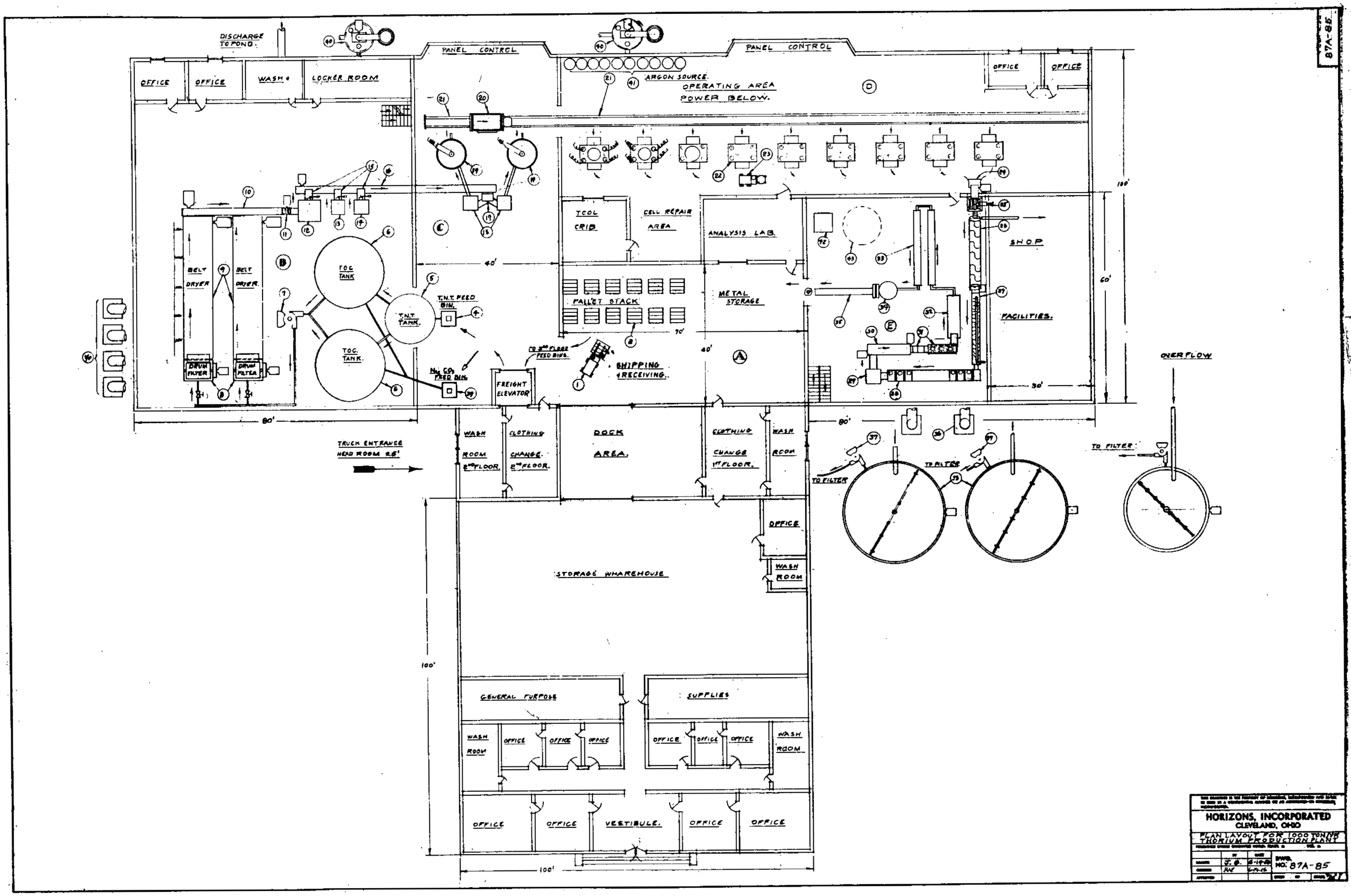

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Fig. 1.
Plan Layout for 1,000 Tons Per Day
Thorium Production Plant

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