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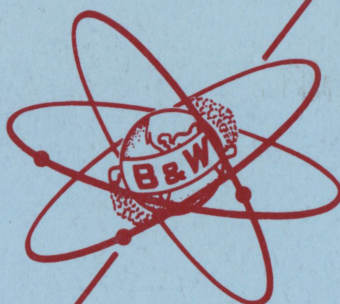
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CHEMICAL PROCESSING  
METHODS AND ECONOMICS  
FOR VARIOUS  
LIQUID METAL FUEL REACTORS

MARCH 1958



**THE BABCOCK & WILCOX COMPANY**

**ATOMIC ENERGY DIVISION**

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CHEMICAL PROCESSING  
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MARCH 1958

Prepared By

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BY  
THE BABCOCK & WILCOX COMPANY  
ATOMIC ENERGY DIVISION







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## I. ABSTRACT

The economic factors of chemical processing of the fuel and blanket from liquid bismuth fuel reactors are described and evaluated. Cost data are presented as a function of pertinent reactor parameters, permitting the estimation of costs for any particular reactor and mode of operation. Costs are presented for on- and off-site processing for single- and two-region reactors. Processes are considered for solutions and slurries of both oxides and intermetallic compounds.

Technical details useful in interpreting and evaluating cost data are presented in the appendices.

These costs are as general as possible, and are essentially independent of other parts of the reactor system. Therefore they are useful in estimating the over-all power costs.



## II. INTRODUCTION

The LMFR design manifesting the most economic promise is uncertain at present. Alternate reactors have differences in both fuel and blanket composition and reactor arrangement. The following possibilities are being considered, all utilizing liquid bismuth.

1. A two-region reactor using a uranium solution as fuel and a thorium bismuthide slurry blanket.
2. A two-region reactor using a uranium solution as fuel and a thorium oxide slurry blanket.
3. A two-region reactor using a uranium oxide slurry as fuel and thorium oxide slurry blanket.
4. A one-region reactor using a mixed uranium oxide and thorium oxide slurry.

All of these utilize the thorium-fuel conversion cycle; however only the two-region reactors can produce as much fuel as they burn. All can improve their conversion ratios by increasing fuel processing, thus controlling nuclear poisons. The conversion ratio is also improved by rapid processing, moving uranium from the blanket of a two-region reactor to the core.

Reactor evaluation requires a complex economic analysis of fuel costs, and also many mechanical, nuclear, and chemical factors. This report presents the chemical processing economics. It is as general as possible, to ease application of the information to all reactors to be considered.

Two processing schemes were studied. The first was the conventional aqueous (solvent extraction) process. The second was a pyro-process incorporating fluoride volatility and zinc slagging techniques. The latter scheme was considered for on-site processing for a reactor with a solution fuel and a thorium bismuthide blanket. With existing technology, the other fuel and blanket combinations require aqueous

processing. The aqueous plant was designed for application to both on-site and central station operation. All economics are presented as a function of processing rate.

The aqueous processing study was done by the Union Carbide Nuclear Company. Argonne National Laboratory (ANL) furnished much information used in the fluoride volatility study.

### III. AQUEOUS PROCESSES

Aqueous processes are more suitable for some reactor systems. Figure 1 describes the aqueous processing method. The flow sheet is a modified Pennsylvania Advanced Reactor (PAR) fuel reprocessing scheme, using a tributyl phosphate solvent. A special head-end step has been added to strip the bismuth of all minor constituents.

For on-site processing of a one-region reactor, the plant would be as shown in Figure 1. This plant will adapt to other modes of operation with only minor modifications to the basic flow sheet. In the off-site processing scheme, the core fluid oxidation, separation, and slurry makeup steps are all done on-site; the rest of the process is done at the central, off-site processing facility. To process a two-region reactor, one more modification must be made, a separation of the thorium and uranium in the primary extraction stage. This can be accomplished by addition of an extra extraction column and possibly another cleanup column to remove the fission products prior to the calcination step. Addition of these units would cause only a nominal (less than 5%) increase in plant costs. Therefore these additions were not explicitly accounted for in the cost estimates.

The slurry from the reactor is sparged with a stream of oxygen. Since bismuth resists oxidation, fission product oxides will form preferentially. The oxidation also will de-wet the oxide particles, making subsequent filtering or centrifuging easier. Without the oxidation step, many soluble fission products would remain in the liquid phase and would not be processed; with oxidation, insoluble oxides are formed.

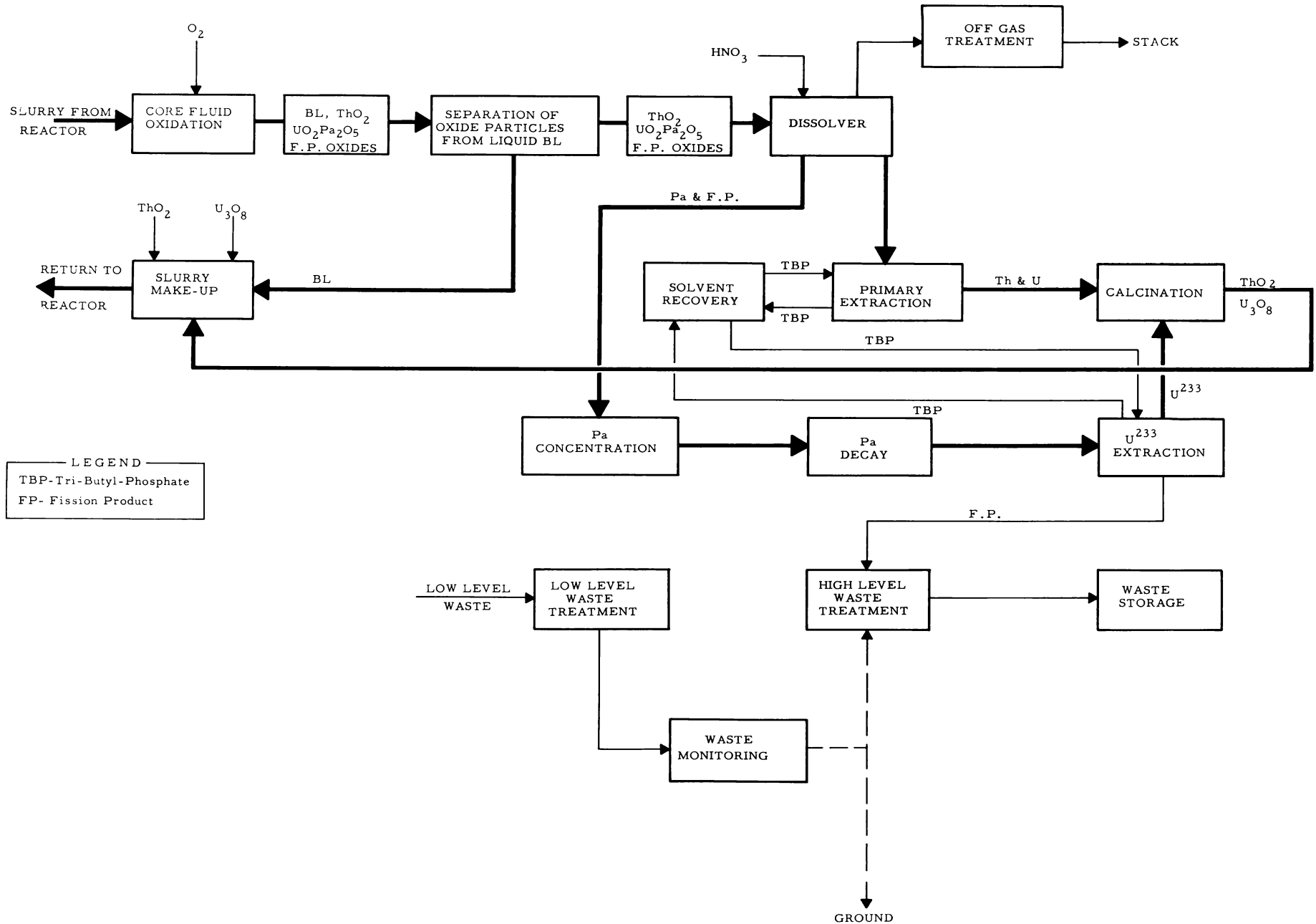
A liquid-solid separation step is done after oxidation, possibly by centrifuging or filtering. After most of the bismuth has been removed some will still adhere to the solids, to be removed by a mercury wash.



The clean particles are then dissolved in nitric acid for on-site aqueous processing or are packaged for shipment to a central plant. The mercury is distilled from the bismuth for re-use; the bismuth is returned to the reactor.

After the clean thoria and urania particles are returned from the processing plant, they are reconstituted with the bismuth from the solids-liquid separation step and from the mercury-bismuth distillation. Mixed uranium oxides (mainly  $\text{U}_3\text{O}_8$ ), not  $\text{UO}_2$ , are the end product of this process. The reconstituted slurry is returned to the reactor.

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#### IV. PYROMETALLURGICAL PROCESSING

A fluoride volatility-zinc slagging method was chosen for this study. Development of the volatility process for application to bismuth fuel processing is underway at Argonne National Laboratory.

Figure 2 shows the main steps in the volatility process. The process as shown may be applied to either blanket or core liquid. When the plant is processing core liquid the basic steps are: (1) hydrofluorination to oxidize most of the fission products, all of the uranium (to  $UF_4$ ), and to change the oxidized material to a fused salt phase; (2) fluorination of the salt to further oxidize the uranium to volatile  $UF_6$ ; (3)  $H_2$  reduction of the  $UF_6$  to  $UF_4$ ; (4) dissolution of the  $UF_4$  in a fused salt phase; and (5) electro-reduction of  $UF_4$ , depositing uranium directly into bismuth.

Uranium, protactinium, and fission products are separated from most of the thorium before applying volatility techniques to the blanket. This separation, referred to as the head-end treatment, is done by heating and diluting the slurry, thus dissolving the particles either partially or completely, and following this by cooling to reform the particles. Much of the uranium, protactinium, and fission products remain in solution throughout the recrystallization. (See Fig. 3.) After separation of the solid and liquid phases the thorium bismuthide is returned to the reactor and the solution can be processed as U-Bi fuel solution.

The head-end process is not 100% efficient, since the uranium and protactinium are not completely removed from the slurry before it is reconstituted and returned to the blanket region.

The fluoride flow sheet is slightly modified for the blanket fluid: (1) salt effluent from the hydrofluorination step is stored, permitting the decay of protactinium to uranium, and (2) the bismuth liquid is returned to the blanket head-end process without the addition of uranium.



Certain fission products are not removed by volatility processing. These are normally not considered serious, and they can be removed by zinc slagging. (See Fig. 4.) Zinc forms stable intermetallic compounds with the noble metals and these compounds can be skimmed from the bismuth. Excess zinc can be removed from bismuth by several procedures; crystallization followed by vacuum distillation was selected for this study.

The processing cycle times for the blanket of a two-region reactor are determined largely by the rate of uranium production. Uranium concentration must be held at the design point to maintain the desired blanket power level and not exceed the capabilities of the blanket's heat removal system.

FIG. 2: FLUORIDE VOLATILITY PROCESS FOR CORE AND BLANKET

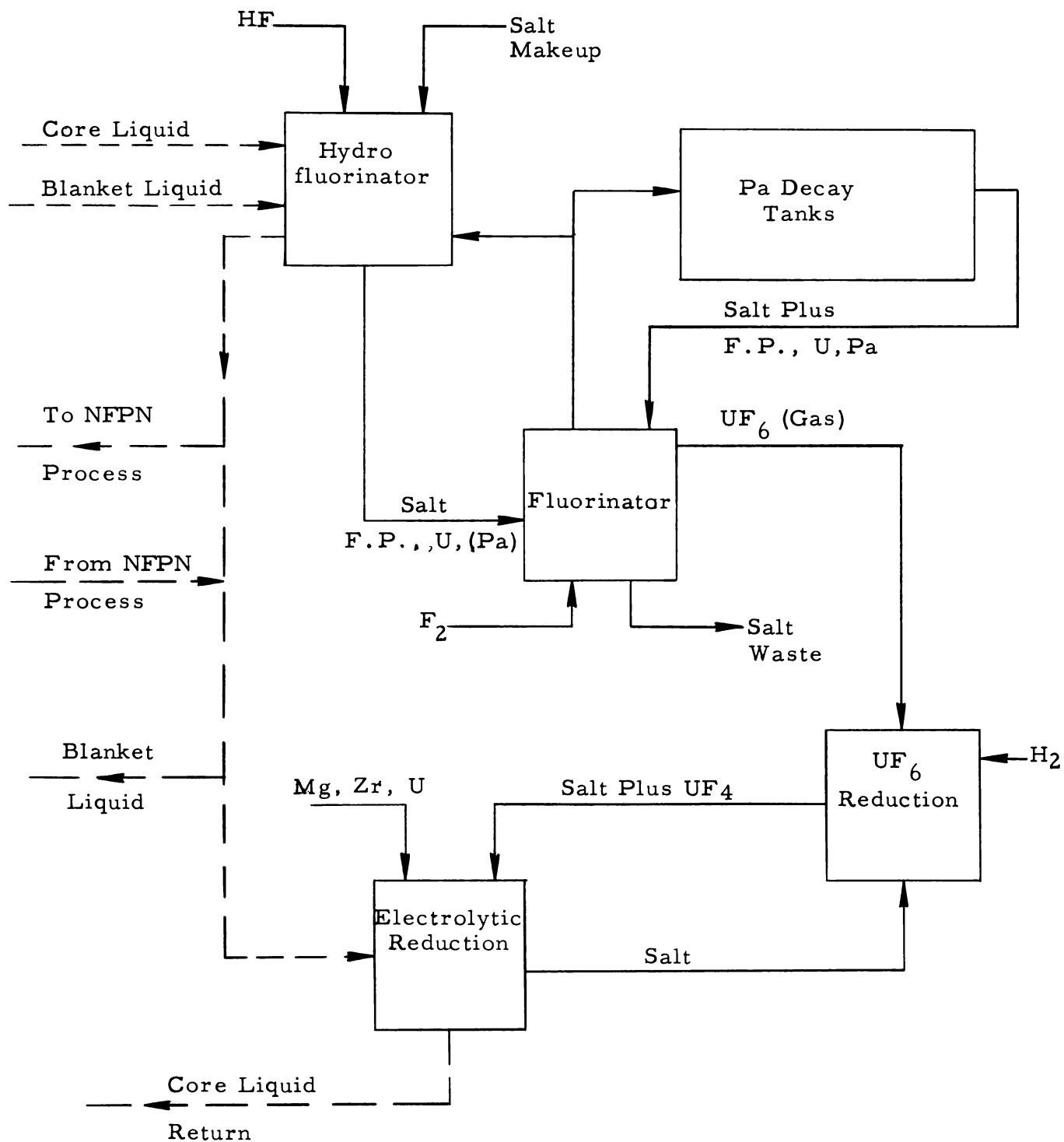


FIG. 3: HEAD-END PROCESSING BISMUTHIDE SLURRY

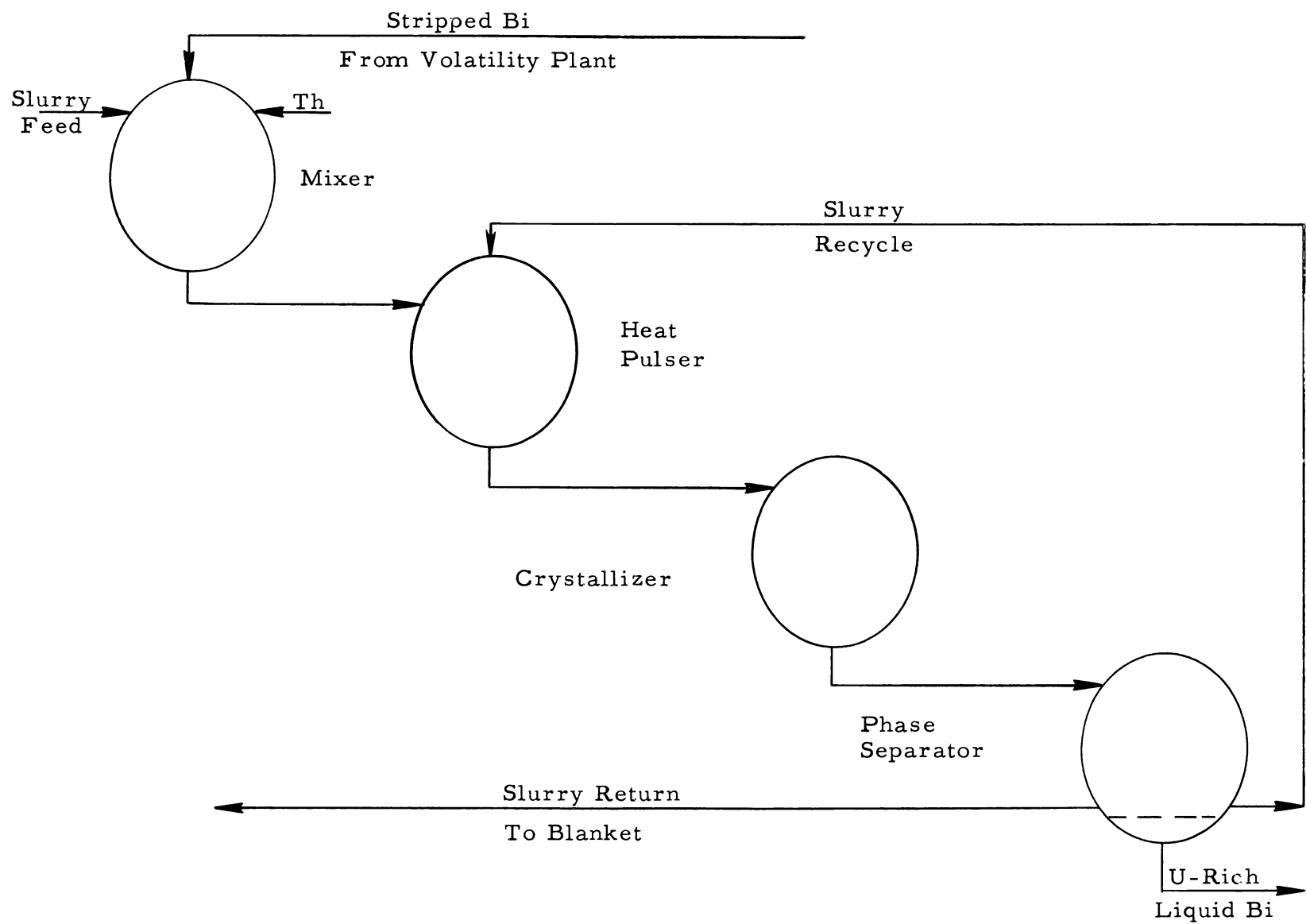
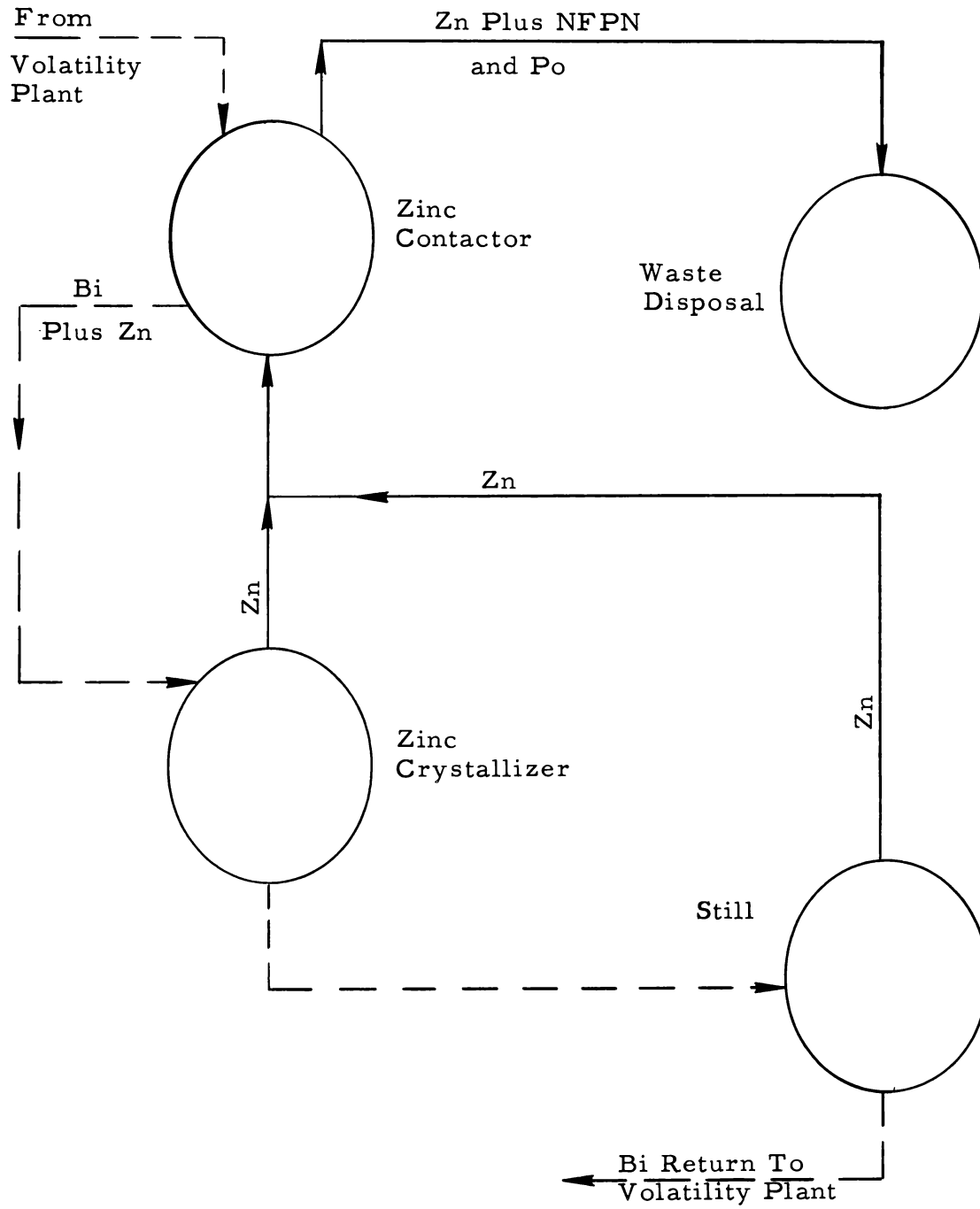




FIG. 4: NFPN FISSION PRODUCT REMOVAL



## V. RESULTS

The processing cycle for a single-region reactor should be very long, perhaps more than ten years. The most expedient processing approach for these long cycles is to send the fuel (in batches) to a central processing facility. The economics of a central aqueous processing plant were studied, considering this point. The cost of this processing is open to speculation since it depends on central plant capacity, which, in turn, depends on the number of reactors supplying spent fuel. It was assumed that the processing is done at a profit by a chemical company. The resulting processing charge per kilogram of thorium as a function of plant size is presented by the top curve in Figure 5. These charges are presented as a function of the thorium throughput because the thorium concentration of the slurry is the determining factor, since it is much greater than the uranium concentration. Figure 5 shows the effect of capacity on unit processing charge.

For this study the capacity of the central plant was assumed to be 200 kilograms per day, for a unit cost of \$86 per kilogram of thorium.

The off-site processing cost does not represent the total processing charges. On-site operations will be required to separate the solids from the bismuth and to prepare them for shipping. On-site storage must be provided to permit radioactive decay prior to shipment. Charges for shipping and inventory in process must also be included as part of the chemical processing cost.

Processing costs for an aqueous processing plant located on-site are presented in Figure 6. On-site operation becomes practical when high processing rates are required.

Annual processing costs for a volatility-zinc extraction plant are presented in Figure 7.

Both of the on-site chemical plants were amortized according to utility procedures, assuming that the LMFR operator would also operate the chemical plant. The aqueous plant was amortized over a 10-yr expected life and the fluoride-zinc plant over a seven-year life. The buildings are included in the total annual costs presented in Figures 6 and 7.

FIG. 5: AQUEOUS OFF-SITE PROCESSING COST

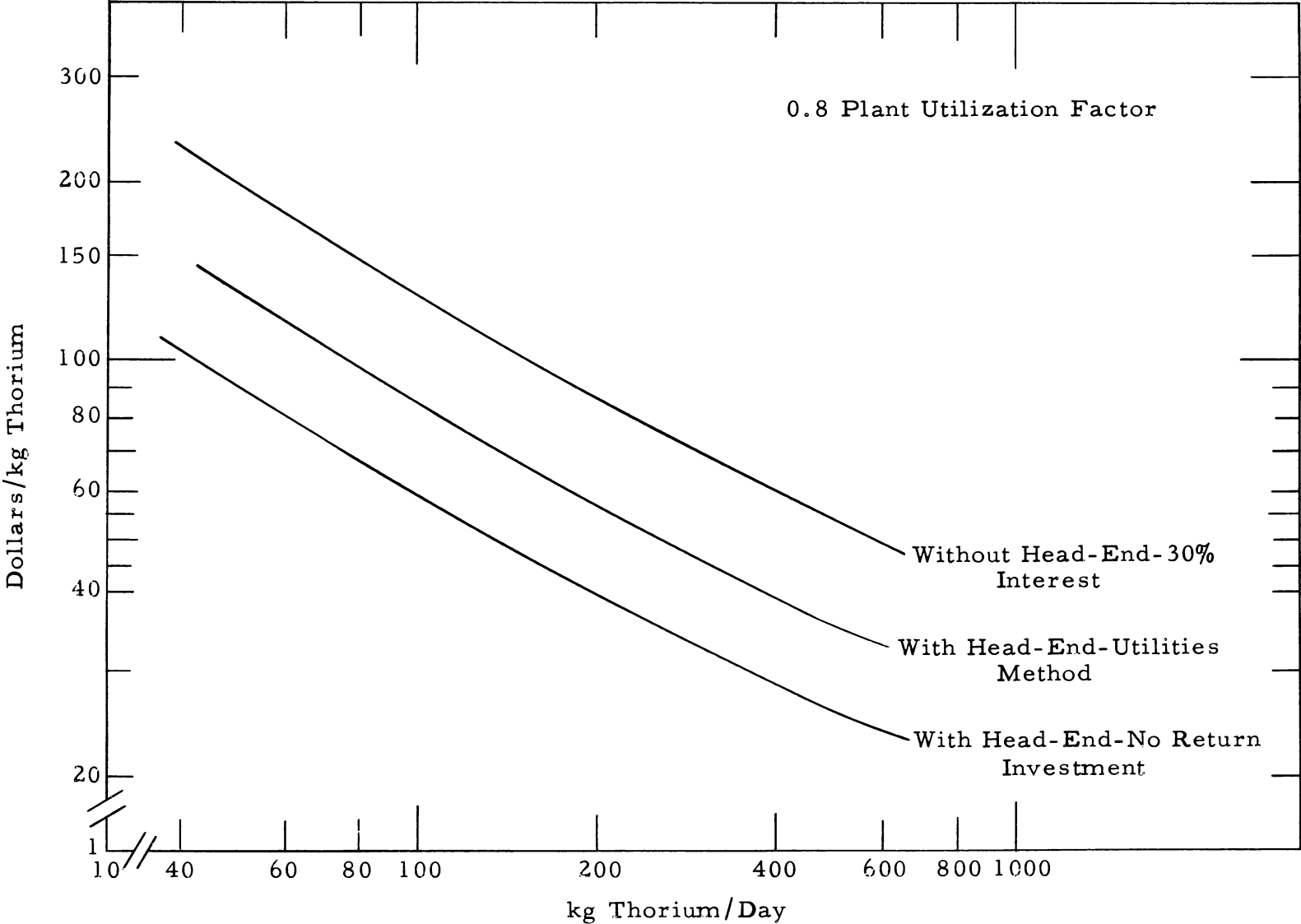




FIG. 6: ON-SITE AQUEOUS PROCESSING COST

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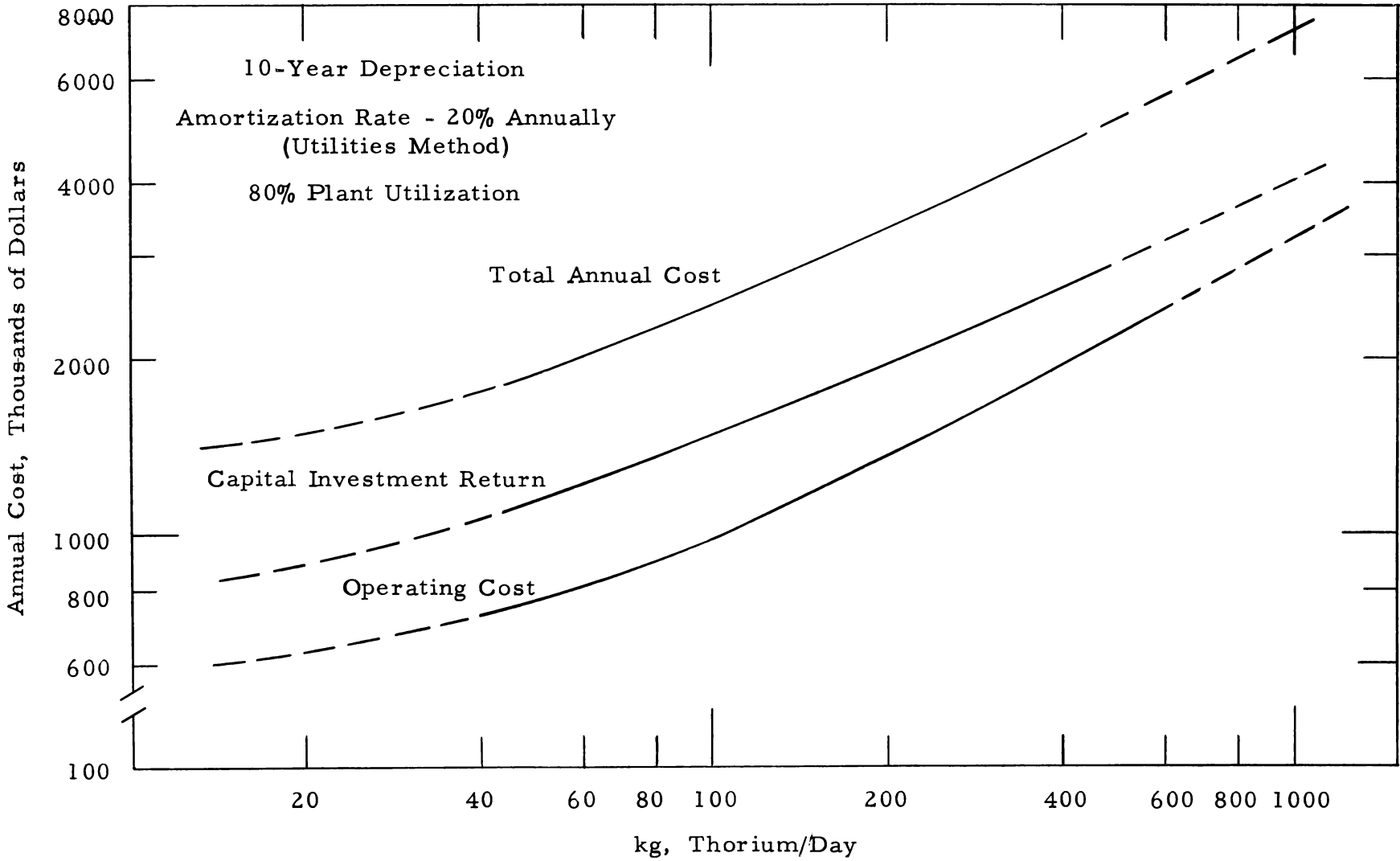
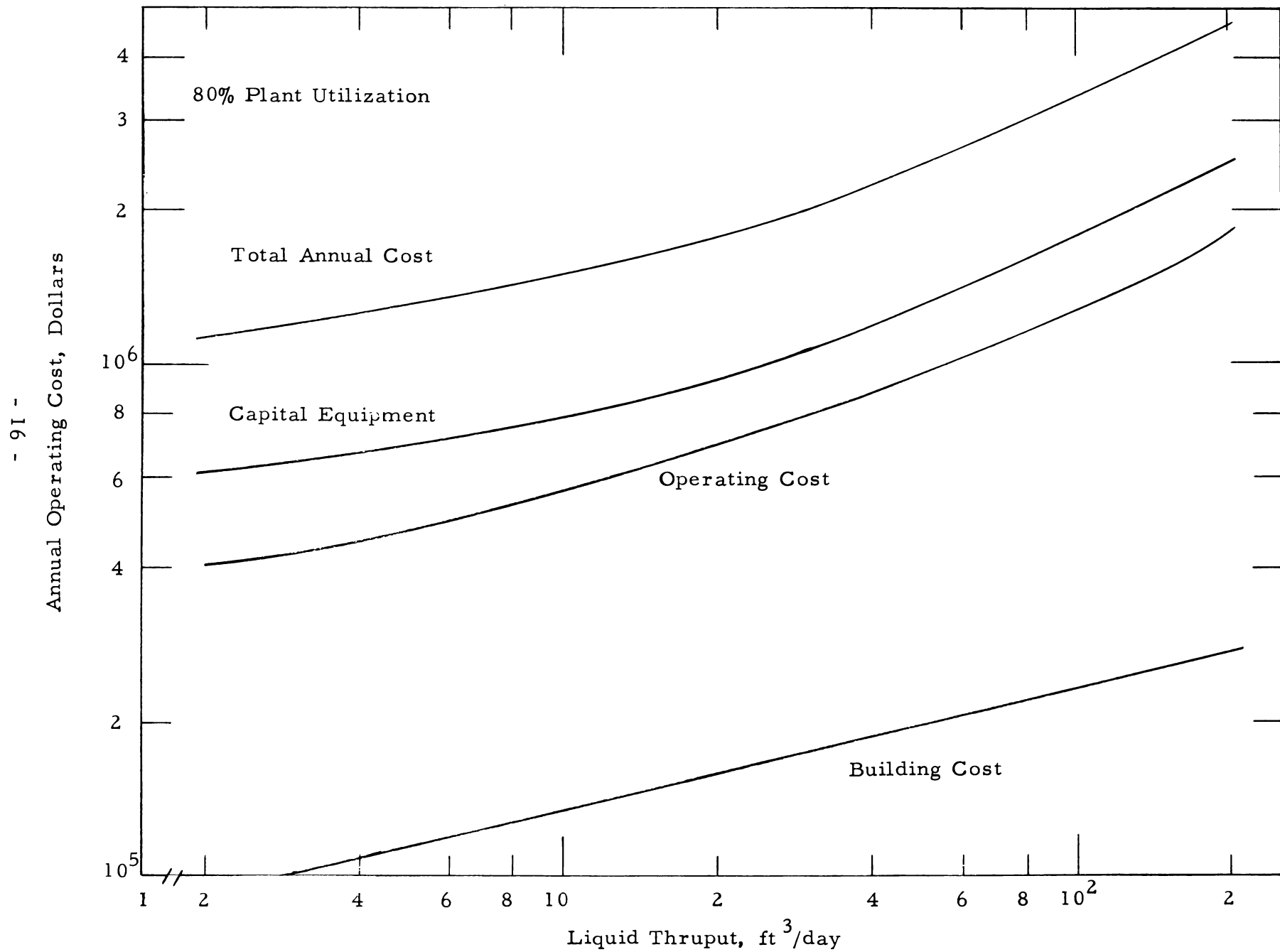


FIG. 7: COST OF FLUORIDE VOLATILITY PROCESSING  
(825 MW, 2-REGION LMFR)



## APPENDICES



## APPENDIX A

### PROCESS ECONOMICS AND COST ESTIMATES FOR AQUEOUS PROCESSING OF AN OXIDE SLURRY

#### 1. Introduction

In the following presentation of aqueous processing costs, various methods of plant operation and amortization will be considered.

##### a. Types of Plant Operation

Two main types of plant operation are considered. These are on- and off-site processing. The essential difference between these two is that for off-site processing, the head-end separation step is performed at the plant and is therefore not included in the off-site cost estimate. For the on-site plant, the head-end step is considered as an integral part of plant operation and is therefore included in the cost estimates.

The on-site process may be sub-divided into two types; processing a two-region reactor, and processing a one-region reactor. The only difference is that for the two-region reactor an additional Th-U separation stage must be included. Otherwise, the plants are almost identical. Figure 9 shows a flow sheet for a plant servicing a one-region reactor.

##### b. Methods of Amortizing Capital Investment

Three methods of amortizing the capital investment were considered, each based on an assumed 10-yr capital recovery period. Only two of these are of real importance, one being applicable to the cost of a central, off-site plant operated by a chemical company, the other being applicable to on-site processing of a reactor operated by a utility company.

The three methods are:

(1) Straight line depreciation with no return on investment. This is mainly of academic interest to determine the bare minimum annual costs for an on-site plant.

(2) Straight line depreciation with 30% annual return on capital investment. This method of amortization is used by chemical companies in computing annual costs of a comparatively new process. It is assumed that the off-site central plant (without head-end processing) will be operated by a chemical company which uses this method of determining the process costs.

(3) Charging 20% of the capital costs annually for a 10-yr capital recovery period according to the practice of utilities in writing off their investment. This "utilities" method will be applied to the on-site plants with head-end processing, which will be associated with reactors operating for a utility company. (See Fig. 8.)

## 2. Estimation of "Basic Plant" Costs

The "basic plant," upon which all the further cost estimates are based, is a modified PAR reactor processing plant with a capacity of 47.5 kg thorium/day. These estimates are based on Union Carbide Nuclear Company's (UCNC) experience in the field, and are taken from a UCNC report (UCN-51) by Rebholz and Zellnik. The basic plant flow sheet is shown in Figure 9. No differentiation in cost was made between the plant that will serve a two-region reactor and one that will serve a one-region reactor. The main difference would be the inclusion of a U-Th separation stage; this would result in less than 5% increase in over-all final costs of the process. The difference in costs was not determined by detailed cost estimates, therefore the same basic costs have been used for estimating both types of plants.

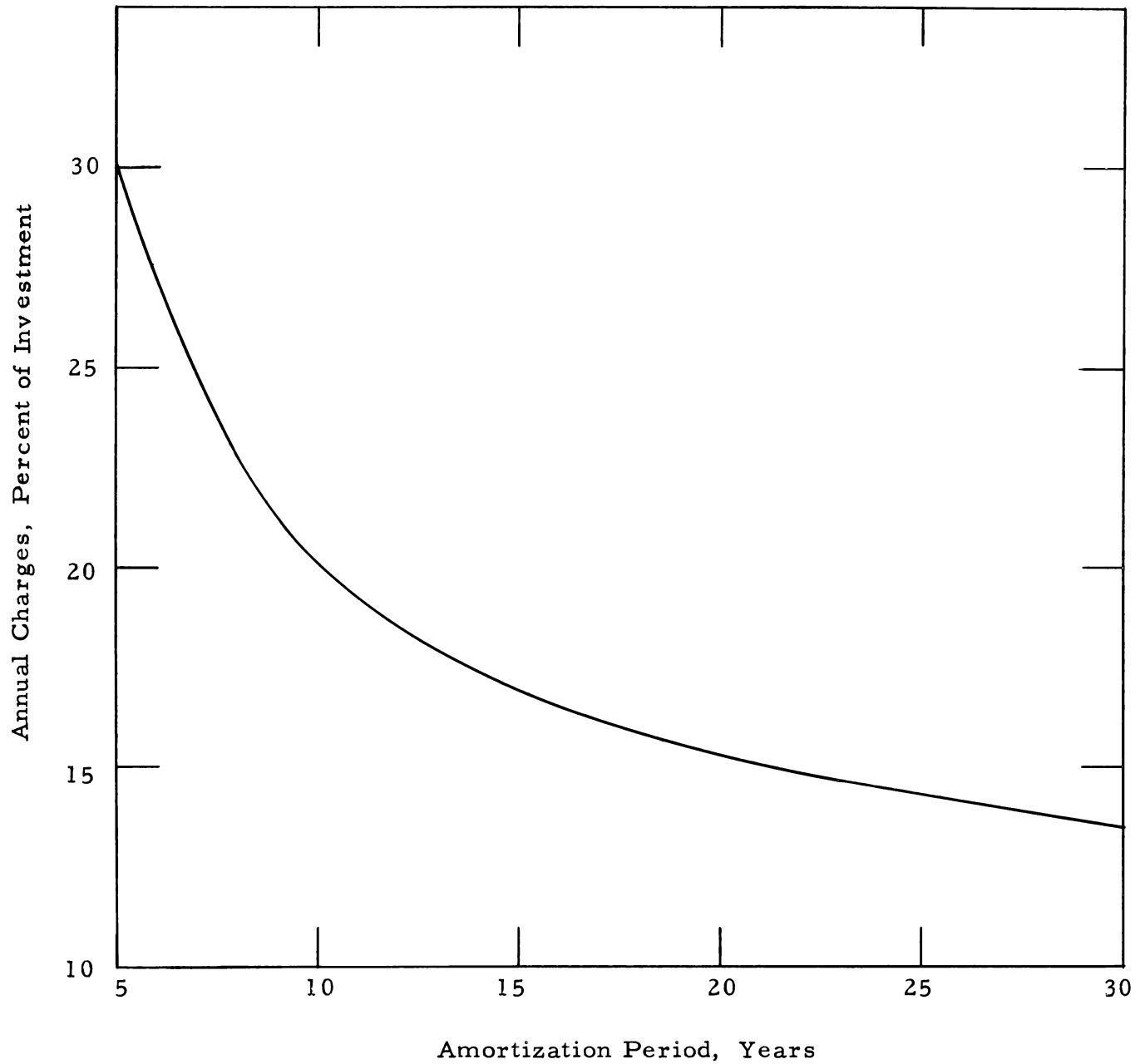
The groups into which the estimates are divided were chosen on the basis of future ease in scaling up the costs for larger plants.

The plant is designed for 80% utilization, or an operating year of 292 days at the indicated capacity.

To determine the actual capital investment required, the grand total of the group costs in Table I must be multiplied by 2.17, a factor accounting for engineering and design, construction overhead, contingencies, and preoperation and startup.



FIG. 8: UTILITIES AMORTIZATION PRACTICE FOR  
CAPITAL INVESTMENT WRITE-OFF



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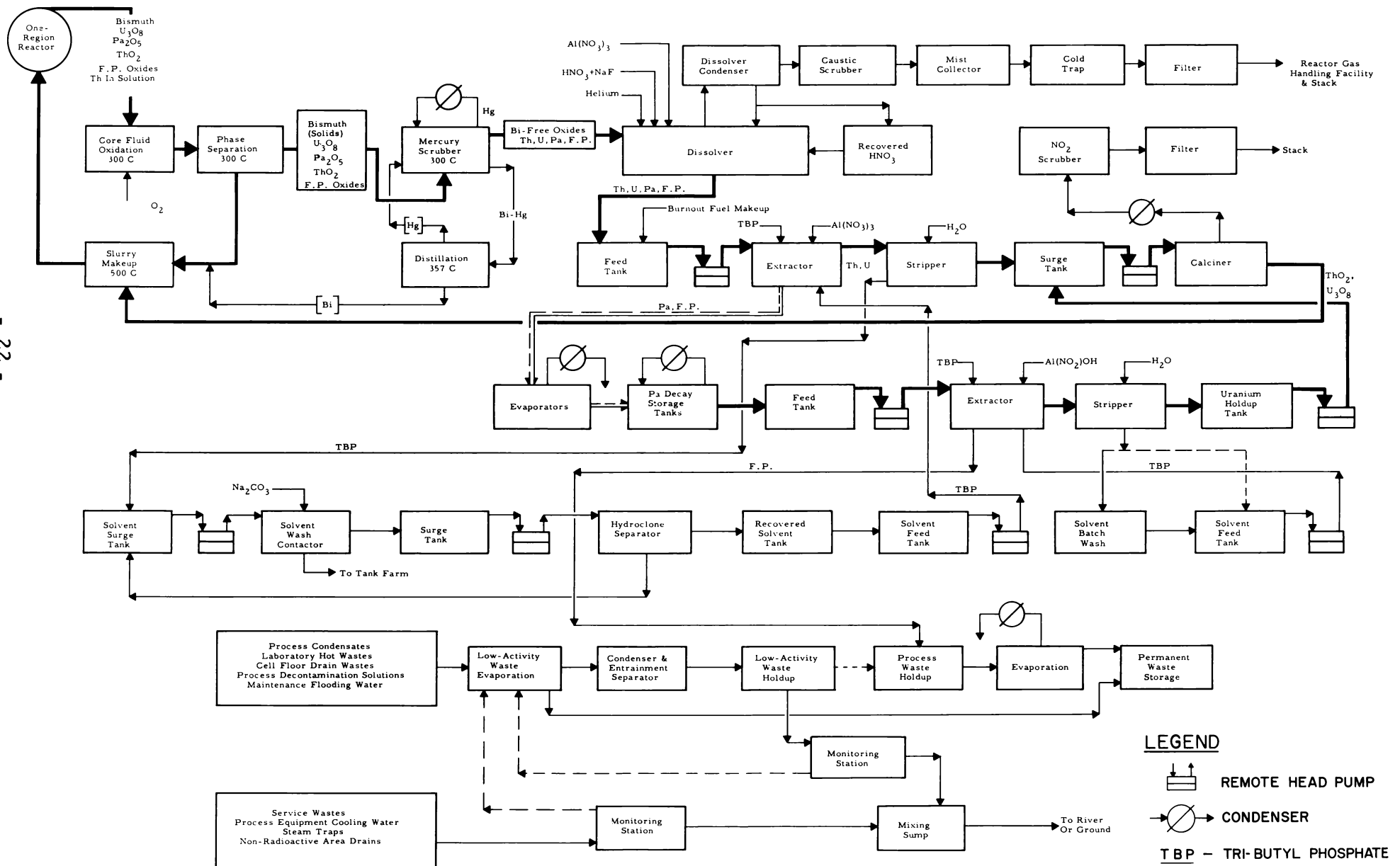


TABLE I  
EQUIPMENT AND BUILDING COSTS FOR A MODIFIED  
PAR REACTOR PROCESSING PLANT<sup>1</sup>  
(Capacity - 47.5 kg Th/day)

<u>CLASSIFICATION</u>	<u>COST</u>
Group I	
Section 1 Dissolver	\$ 11,000
Section 2 Dissolver off-gas	23,430
Section 3 Thorium-uranium extraction	23,650
Section 4 Slurry reconstitution	34,320
Section 5 Thorium-uranium extraction raffinate evaporator	5,500
Section 6 Solvent recovery	18,370
Section 7 Rerun system	5,290
Section 8 Makeup and chemical storage	53,680
Total	<u>\$175,240</u>
Group II	
Section 1 Protactinium decay and U-233 processing	<u>\$ 65,010</u>
Total	<u>\$ 65,010</u>
Group III	
Section 1 Low level liquid-waste evaporator system	\$ 31,460
Section 2 Liquid-waste surge and storage system	<u>220,550</u>
Total	<u>\$252,010</u>
Group IV	
Section 1 Laboratory equipment	\$267,850
Section 2 Special equipment	<u>158,400</u>
Total	<u>\$426,250</u>
Group V	
Section 1 Process buildings	\$338,030
Section 2 Laboratory	166,320
Section 3 Offices, health physics, change area, etc.	<u>98,340</u>
Total	<u>\$602,690</u>
Group VI <sup>2</sup>	
Section 1 Head-end equipment	<u>\$ 35,000</u>
Total	<u>\$ 35,000</u>
Total for Basic Plant	<u>\$1,556,200</u>

<sup>1</sup>Data from UCN-10 corrected to current cost index.

<sup>2</sup>The cost of this equipment was based on a separate estimate shown in Table II and scaled down to a 47.5 kg Th/day plant by a seven-tenths factor based on the thorium throughput.

TABLE II  
HEAD-END SEPARATION STEP  
(Rough Estimate of Equipment Cost for Separating  
Oxide Particles from Liquid Bismuth<sup>3</sup>)

<u>QUANTITY</u>	<u>ITEM</u>	<u>COST BLS 125<sup>4</sup></u>
1	Reaction vessel	\$ 2,000
1	Centrifuge	25,000
1	Mercury scrubber	3,000
1	Dryer	5,000
2	Mercury condensers	4,000
1	Mercury storage tank	500
1	Mercury distillation column	3,000
1	Conveyor	1,000
600 lb	Mercury	2,200
8	Pumps	4,000
8	Motors	2,400
	Total	<u>\$52,100</u>
	Total Corrected to BLS <sup>4</sup> 160	\$67,000

<sup>3</sup>For a bismuth throughput of approximately 2500 kg/day. This amount is associated with the 121 kg of Th/day plant when Th concentration is 10% by weight. This was scaled down to \$35,000 for the "basic plant" of 47.5 kg/day.

<sup>4</sup>Bureau of Labor Statistics Index

TABLE III  
"BASIC PLANT" OPERATING COSTS<sup>5</sup>

1.	<u>Direct Labor</u> (4 shifts with 4 operators each)	\$95,000
2.	<u>Departmental Expense</u>	
	Supervision (1 technical assistant/shift at \$7,000/yr)	28,000
	Wage premiums and supplemental labor costs (20% of direct labor)	19,000
	Equipment and building repairs (2% of fixed investment)	87,800*
	Miscellaneous supplies	5,000*
	Sundry invoices charges	5,000*
	Rebuilding and rearranging (1% of fixed investment)	43,900*
	Miscellaneous charges not itemized	2,000*
	Instrument and process safety expense (1 health physicist, 2 instrument mechanics x 2 for materials)	42,000

<sup>5</sup>Costs must be increased by 10%, correcting then to the current cost index.

TABLE III (CONT'D)

Maintenance expense (2 electricians, 1 millwright, 1 pipefitter x 2 for materials)	40,000*
Laboratory expense (2 analysts/shift + \$1/analysis)	50,000
Spectroscopists (mass and emission, one each)	16,000
Counter room operator	5,200
Total	<u>\$343,900</u>
3. <u>Auxiliary Expense</u>	
Plant superintendent	12,000
Plant engineer (assistant superintendent)	9,000
Secretary	4,500
Clerk	3,600
Chief analyst	7,600
Health physics (plant safety engineer)	6,300
Janitor	3,000
Departmental expense	<u>40,000</u>
Total	\$ 86,000
4. <u>Total Plant Expense</u> (Items 2 + 3)	\$429,900
5. <u>Utilities</u>	
Cooling water	7,800*
Electricity	5,100*
Steam	2,000*
Compressed air	<u>2,100*</u>
Total	\$ 17,000
6. <u>Total Raw Materials</u>	\$ 15,000*
7. <u>Inventory and Losses</u>	\$ 39,000*
8. <u>Plant Operating Cost</u> (Items 1 + 4 + 5 + 6 + 7)	\$595,900
9. <u>Contingencies</u> (10% of Item 8)	\$ 59,600
10. <u>Total Plant Operating Cost</u> (Items 8 + 9)	\$655,500

\*Treated separately in scale-up of operating costs.

### 3. Effect of Plant Size on Costs

These estimates are based on the "basic plant" costs for a plant having a capacity of 47.5 kg of thorium per day. These data are recorded in Table I where they have been arranged in convenient groups to facilitate extrapolation by the six-tenths-factor rule. The cost of head-end equipment for handling liquid bismuth was obtained from a separate estimate, shown in Table II. The total cost for each group was extrapolated to higher thorium capacity levels.

Group I, consisting of process equipment varying in size with thorium throughput, was extrapolated on the basis of the 0.7 power of the ratio of new-to-original capacity.

Group II, consisting of process equipment varying in size with the protactinium throughput, was assumed constant over the range of capacities considered. This assumption is possible because the UCN-10 plant is necessarily oversized to insure operational characteristics.

Group III consists of waste handling equipment which was extrapolated on the basis of the 0.7 power of the change in volume through the system. The total volume through the system is the sum of a fixed volume plus a volume that varies directly with the thorium throughput.

Group IV, consisting of laboratory and special equipment, was extrapolated on the basis of the 0.25 power of the thorium capacity change.

Group V, consisting of buildings, was extrapolated on the basis of the 0.25 power of the change in thorium capacity.

Group VI consists of equipment for separating oxide particles from liquid bismuth. The size of the equipment will vary as a function of the liquid bismuth throughput. In this case, the concentration of thorium in bismuth was assumed to be 5%, and the cost of equipment was factored on the basis of the 0.7 power of the thorium capacity change. This equipment, associated piping, and instrumentation is not included in the calculation of off-site cleanup charges.

Process piping was estimated for the modified 47.5 kg of Th/day plant 1.1 times the delivered cost of process equipment. Waste system piping was estimated for the modified plant as 0.9 times delivered cost of waste equipment. The total piping cost was extrapolated to each capacity level on the basis of the 0.7 power of the change in thorium capacity.

Instrumentation was estimated for the modified plant using the UCN-10 factors; that is, process instrumentation cost equals 0.75 times delivered process equipment cost; waste instrumentation cost equals 0.1 times delivered waste equipment cost. Total instrumentation cost was extrapolated to each capacity level on the basis of the 0.3 power of the change in thorium capacity.

Installation charges were estimated as 1.15 times the delivered process equipment cost at each capacity level.

In UCN-10, total fixed investment was estimated by including allowances for construction overhead, engineering and design, contingency, and startup. The factors gave a total fixed investment of approximately 2.17 times the total of buildings and equipment. In these estimates, the 2.17 factor is used.

Operating charges were estimated for each capacity level on the basis of UCN-10 data in Table III. The items listed have been classified into two groups, for extrapolation. Items in Group A, designated by (\*), are assumed factorable on the basis of the 0.7 power of the thorium capacity change. The unmarked items, Group B, are assumed factorable on the basis of the 0.25 power of the thorium capacity change. The totals for both groups, increased by 10% to correct to the current cost index, are: A = \$375,000 and B = \$260,000. Allowing for a 10% contingency, operating costs for the various plant capacities were calculated according to the following relationship:

$$\text{Operating Cost} = 1.1 (375.00 [x]^{0.25} + 260,000 [x]^{0.7}), \quad (1)$$

where x is the appropriate thorium ratio.

The extrapolated data obtained were used to estimate the cost of processing as a function of chemical plant capacity for two conditions: (1) continuous on-site processing using utility amortization, and (2) off-site cleanup of the fuel after extended operation, where the head end separation costs are not included and a 30% annual return on capital investment is included. For academic interest, the no interest, straight line, 10-yr depreciation costs have also been computed and plotted in some of the subsequent figures. Summaries of the costs for various plant capacities are given in Tables IV-VIII.

Figure 5 shows the processing cost per kg thorium for the three methods in Tables VI-VIII. Figure 6 shows the data of Table VII, and Figure 10 shows both Tables VI and VII for comparison. Both figures 6 and 10 refer to on-site processing. Figure 11 shows Table VIII, the off-site processing cost as a function of the central plant capacity when it is operating at full capacity. The off-site central plant will not operate a full year to process a single core charge of a one-region reactor. A 12,500-kg thorium reactor charge was assumed as representative of the inventory of a one-region LMFR system. The time to process such a charge and the associated costs are given in Table VIII.



FIG. 10: ANNUAL COST OF AQUEOUS PROCESSING VERSUS KG TH/DAY

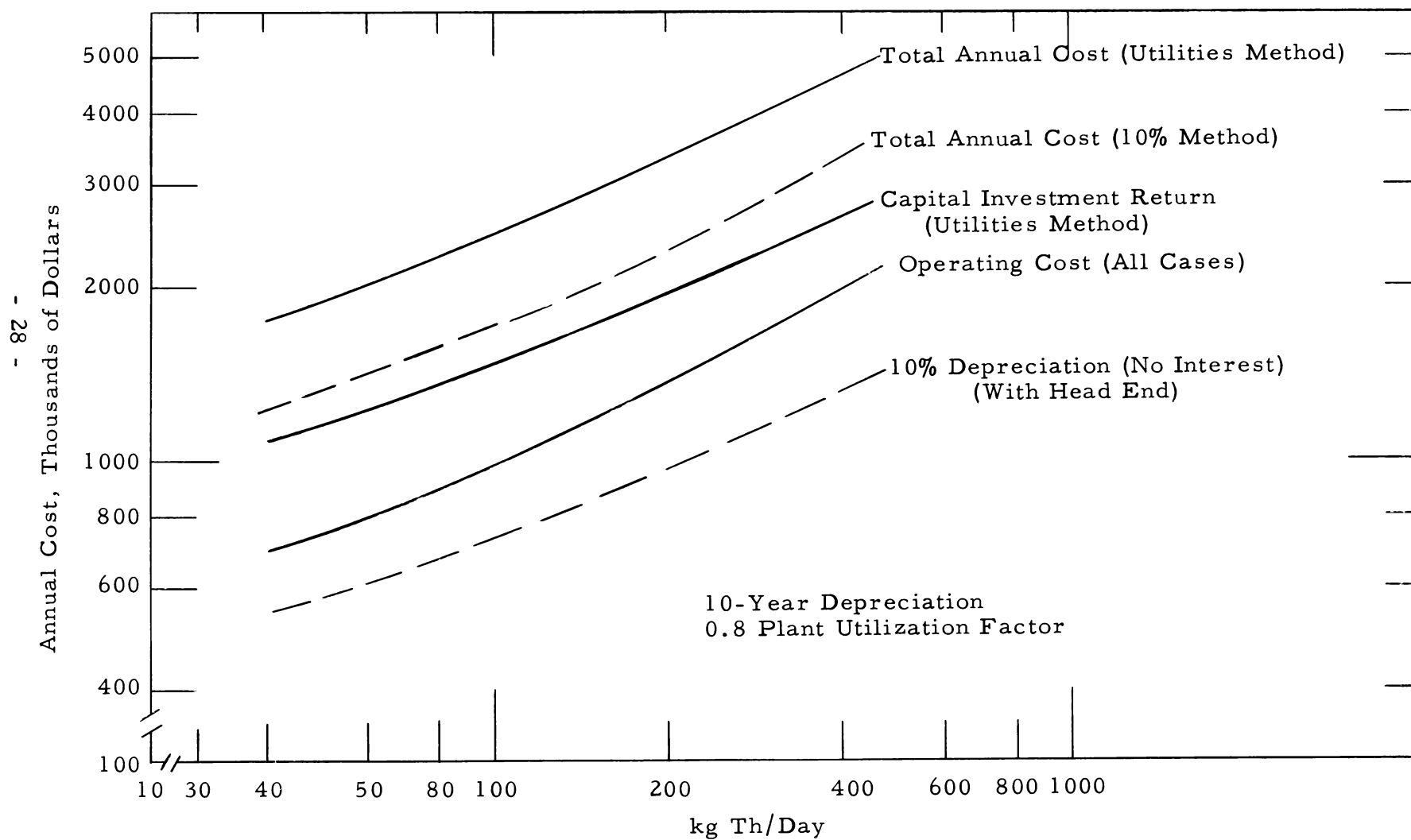


FIG. 11: COST OF OFF-SITE AQUEOUS PROCESSING

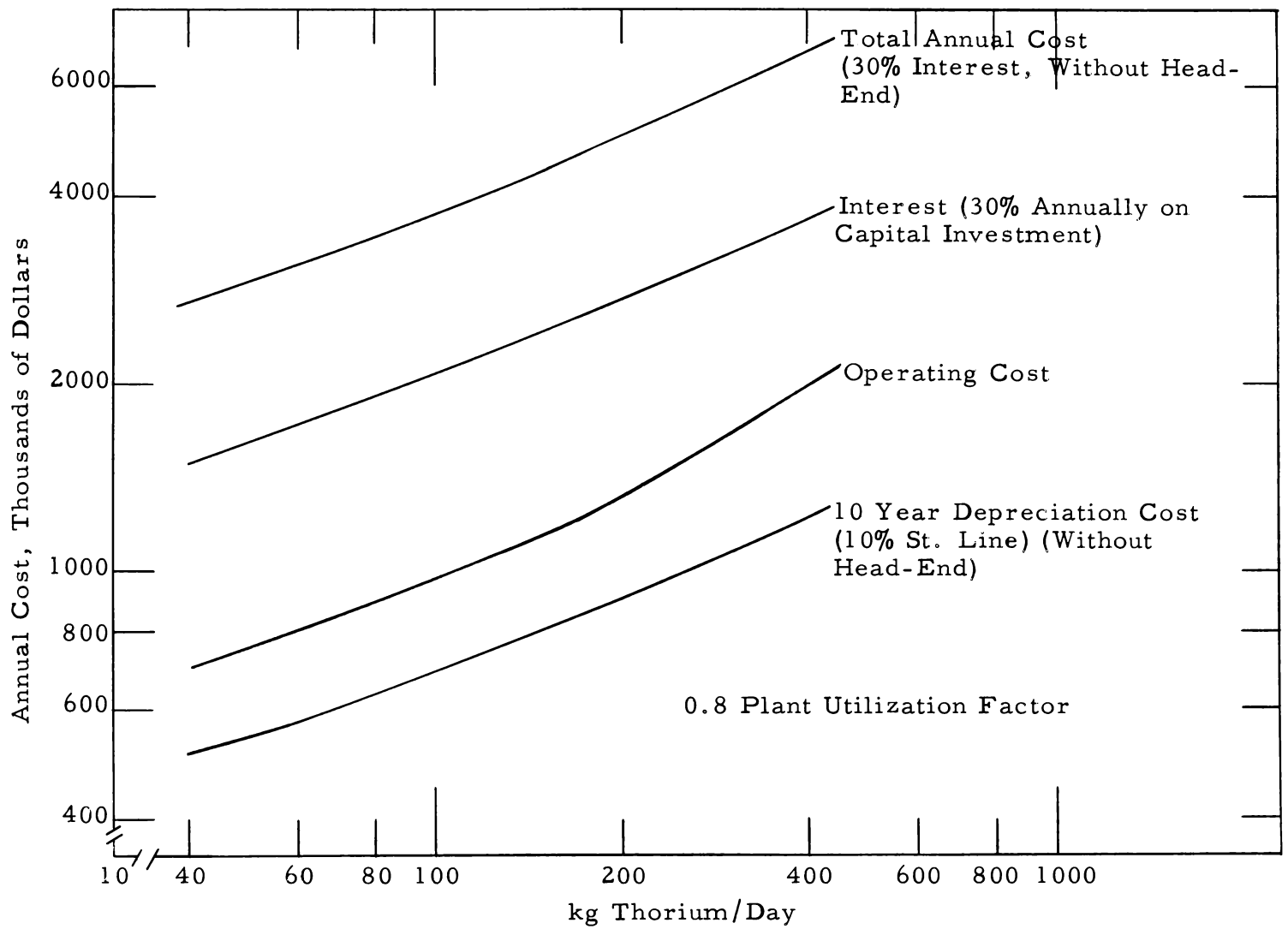
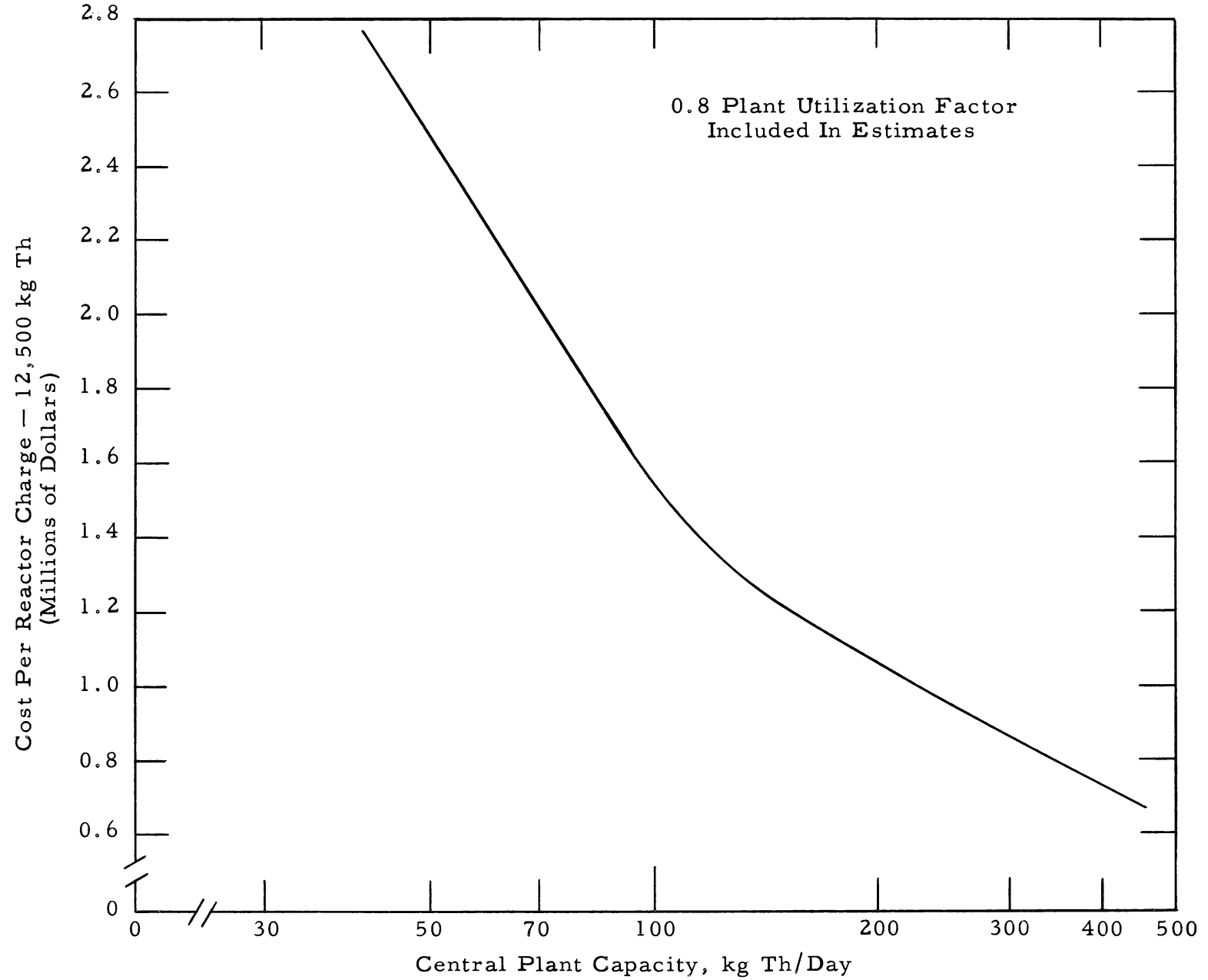


FIG. 12: OFF-SITE PROCESSING COST



The cost of processing such a charge as a function of central plant capacity is also shown in Figure 12. These values will be used to compute the over-all cost of off-site processing.

To obtain the cost for reactor charges other than 12,500 kg of thorium, the cost of a 12,500-kg charge is read from Figure 12. The cost is then scaled up or down in direct proportion to the actual reactor charge size. Another method of obtaining this cost is to read the cost per kg thorium for off-site processing from the top curve of Figure 5 for an assumed central plant size and then to multiply this cost by the size of the charge.

TABLE IV  
TOTAL FIXED INVESTMENT, ON-SITE CONTINUOUS PROCESSING

<u>ITEM</u>	<u>Plant Capacity (kg Th/day)</u>				
	<u>425</u>	<u>218</u>	<u>121</u>	<u>85</u>	<u>47.5</u>
	Cost (in thousands of dollars)				
Installed process and waste equipment, piping, and instrumentation	\$4,425	\$3,166	\$2,336	\$1,976	\$1,517
Laboratory and special equipment	737	622	541	494	426
Buildings	1,100	928	801	737	636
Subtotal	6,262	4,716	3,678	3,207	2,579
Total Fixed Investment (2.17 x subtotal)	\$13,588	\$10,234	\$7,981	\$6,959	\$5,596

TABLE V  
TOTAL FIXED INVESTMENT, OFF-SITE CLEANUP

<u>ITEM</u>	<u>Plant Capacity (kg Th/day)</u>				
	<u>425</u>	<u>218</u>	<u>121</u>	<u>85</u>	<u>47.5</u>
	Cost (in thousands of dollars)				
Installed process and waste equipment, piping and instrumentation	\$3,989	\$2,860	\$2,109	\$1,783	\$1,379
Laboratory and special equipment	737	622	541	494	426
Buildings	1,100	928	801	737	636
Subtotal	\$5,826	4,410	3,451	3,014	2,441
Total Fixed Investment (2.17 x subtotal)	\$12,642	\$9,570	\$7,489	\$6,540	\$5,297

TABLE VI  
CHEMICAL PROCESSING CHARGES

(Continuous On-Site Processing, No Return on Investment)

<u>ITEM</u>	Plant Capacity (kg Th/day)				
	425	218	121	85	47.5
	Cost (in thousands of dollars)				
Annual operating costs	\$2,040	\$1,435	\$1,074	\$ 907	\$ 732
Depreciation (10%)	1,359	1,023	798	696	560
Total	\$3,399	\$2,458	\$1,872	\$1,603	\$1,292
<hr/>					
kg of Th processed annually at 80% plant factor	124,100	63,656	35,449	24,820	13,870
\$ /kg of Th processed	\$27.39	\$38.61	\$52.81	\$64.58	\$93.15

TABLE VII  
CHEMICAL PROCESSING CHARGES

(Continuous On-Site Processing Using Utility Amortization of 20%  
Annually on Capital Investment 10-Yr Capital Recovery Period)

<u>ITEM</u>	Plant Capacity (kg Th/day)				
	425	218	121	85	47.5
	Cost (in thousands of dollars)				
Annual operating costs	\$2,040	\$1,435	\$1,074	\$ 907	\$ 732
Utilities Amortization	2,718	2,046	1,596	1,392	1,120
Total	\$4,758	\$3,481	\$2,670	\$2,299	\$1,852
<hr/>					
kg of Th processed annually at 80% plant factor	124,100	63,656	35,449	24,820	13,870
\$ /kg of Th processed	\$38.40	\$54.70	\$75.30	\$92.60	\$134.00

TABLE VIII  
COST FOR REACTOR CHARGE CLEANUP AT AN OFF-SITE PLANT

<u>ITEM</u>	Plant Capacity (kg Th/day)				
	425	218	121	65	47.5
	Cost (in thousands of dollars)				
Annual operating costs	\$2,040	\$1,435	\$1,074	\$ 907	\$ 732
Depreciation	1,264	957	749	654	530
30% return on investment	3,793	2,871	2,247	1,962	1,589
Total	\$7,097	\$5,263	\$4,070	\$3,523	\$2,851

TABLE VIII (CONT'D)

ITEM	Plant Capacity (kg Th/day)				
	425	218	121	65	47.5
kg of Th processed annually at 80% plant factor	124,100	63,656	35,449	24,820	13,870
\$/kg of Th processed	\$57.19	\$82.68	\$114.81	\$141.94	\$205.00
Charge for processing reactor containing 12,500 kg Th (in thousands of dollars)	\$715	\$1,033	\$1,435	\$1,774	\$2,569
Days to process 12,500 kg Th	29	57	103	147	263

#### 4. Factors Influencing the Cost of Off-Site Processing

##### a. Off-Site Central Plant Capacity

The cost of off-site processing is greatly affected by the capacity of the central plant, since unit costs decrease with higher capacities. (See Fig. 12.) The actual size of a central plant which might be expected to exist in 1980 is a matter of pure conjecture, though some estimates can be made.

The size of central plant to be expected in 1980 will be a function of the demand for its services. To estimate this plant size, the expected plant capacity was computed as a function of the number of reactors the plant would service, assuming that:

- (1) The average reactor charge would be about 12,500 kg thorium; and
- (2) The reactors serviced by the central plant would be on a 10-yr average cycle.

The results of such a study were partially tabulated in Table VIII, and are fully presented in Table IX.

The results in Table IX are shown in Figures 13 and 14. Figure 13 shows the expected plant capacity vs the number of reactors serviced, and Figure 14 shows the cost per 12,500 kg thorium reactor charge based on the number of reactors serviced by the central plant.

FIG. 13: REACTORS REQUIRED FOR CENTRAL PLANT ECONOMY

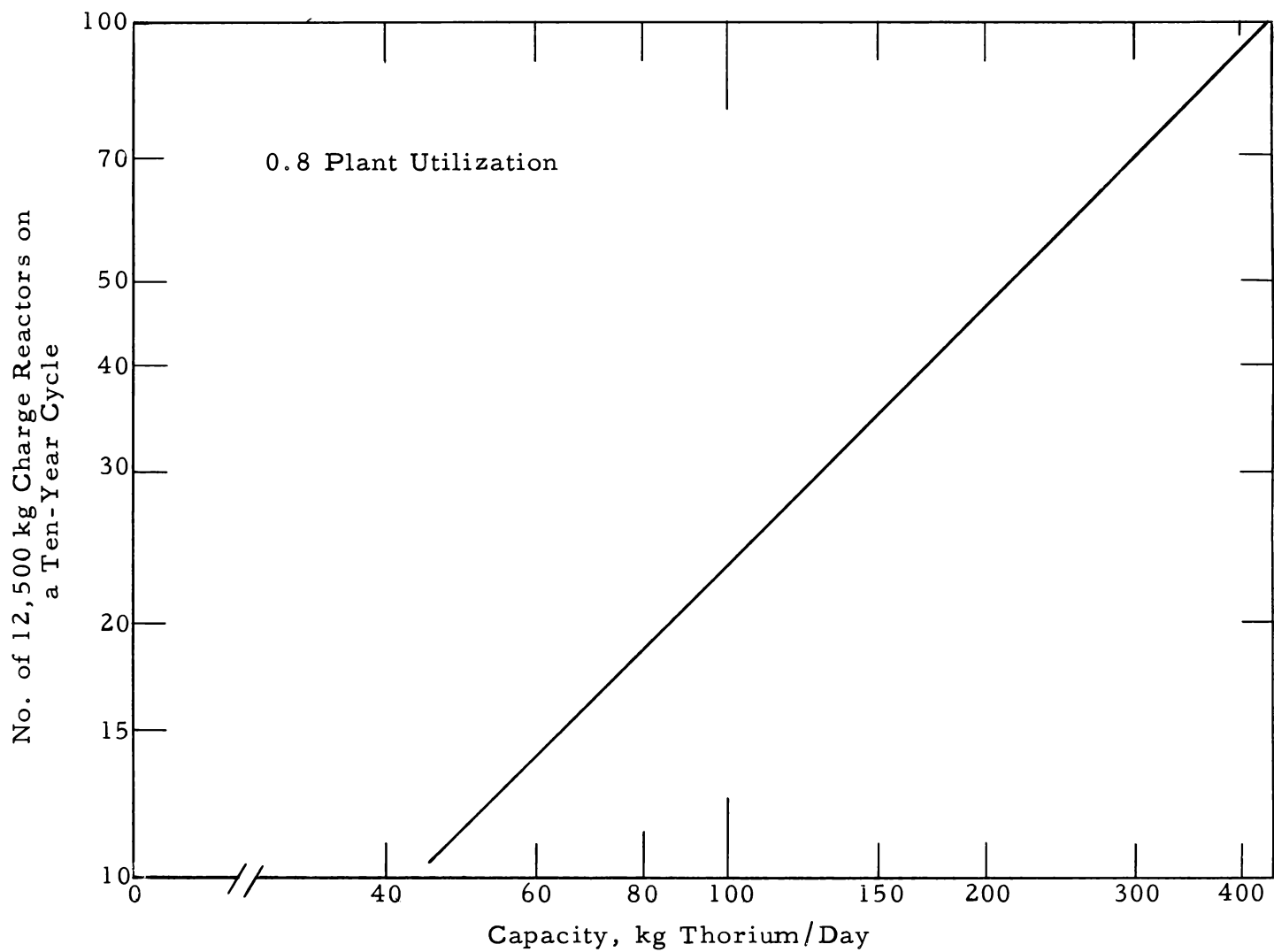




FIG. 14: COST VERSUS REACTORS SERVICED

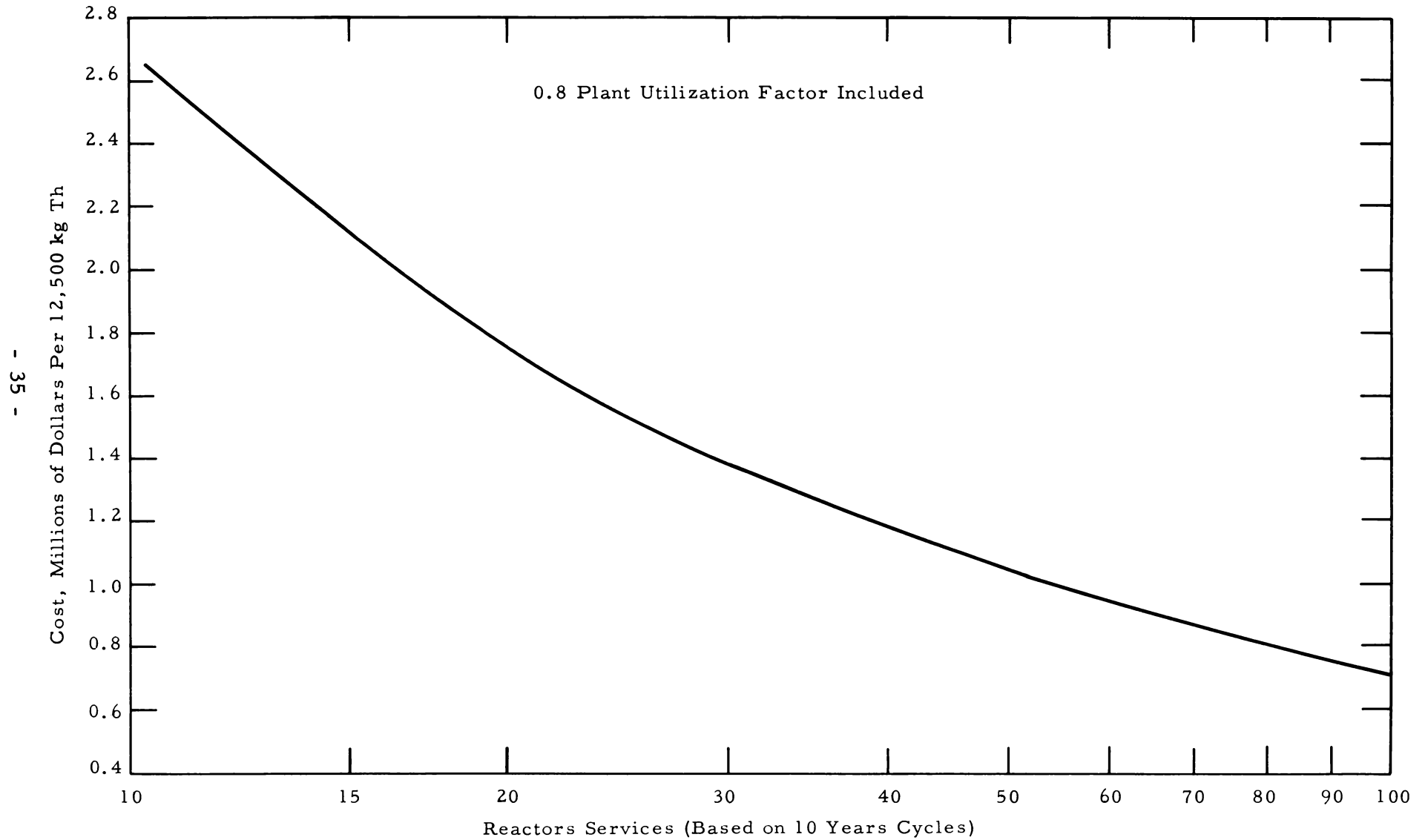


TABLE IX  
PROCESSING COSTS AS A FUNCTION OF CENTRAL  
PLANT CAPACITY, AND PLANT CAPACITY  
VERSUS NUMBER OF REACTORS SERVICED

ITEM	Central Plant Capacity (kg Th/day)				
	425	218	121	85	47.5
Annual plant capacity at 80% plant factor	\$124,100	\$63,656	\$35,449	\$24,820	\$13,870
\$/kg Th processed	\$57.19	\$82.68	\$114.81	\$141.94	\$205.00
Cost of processing reactor charge of 12,500 kg Th (in thousands of dollars)	\$715	\$1,033	\$1,435	\$1,774	\$2,569
Days to process 12,500 kg Th	29	57	103	147	263
Number of 12,500-kg charges/yr which can be processed in plant at 80% plant factor	9.94	5.10	2.84	1.987	1.100
Number of reactors of 12,500-kg charge operating on a 10-yr cycle which are required to keep central plant working at capacity	99.4	51.0	28.4	19.87	11.00

A 12,500-kg reactor operating on a two-year cycle is equivalent to five of the assumed reactors, and a reactor with a charge of 25,000 kg of thorium and operating on a 10-yr cycle is equal to two of the assumed reactors. On this basis, a sufficient demand will exist by 1980 to require the services of a 200 kg/day central plant.

b. On-Site Facilities for Off-Site Processing

Another factor entering in the cost of off-site processing is the on-site facility cost per charge for head-end separation and storage. The annual cost of this facility remains almost constant as a function of cycle time, but the proportional charge per batch decreases with increasing cycle time.

Rough cost estimates of these on-site facilities for off-site processing are as follows.

	<u>Capital Investment</u>
Oxide slagging, phase separator and associated plant	\$ 700,000
Storage facility	<u>500,000</u>
Total	\$1,200,000

Also, there is an operating cost for each charge removed from the bismuth, estimated as \$250,000/charge. These costs were not worked out in detail, but they are felt to be conservative. This problem must be studied further if a one-region reactor, with off-site processing, is chosen. Also, transportation costs must be considered.

APPENDIX B  
PROCESS ECONOMICS AND COST ESTIMATES  
FOR FUEL PROCESSING BY FLUORIDE VOLATILITY  
AND ZINC SLAGGING TECHNIQUES

1. Detailed Equipment Cost Estimates (See Fig. 15.)

a. FPS and FPN Poison Removal by Fluoride Volatility Processing  
(Processing rate - 20 ft<sup>3</sup>/day of liquid Bi)

<u>ITEM</u>	<u>QUANTITY</u>	<u>SIZE</u>	<u>DESCRIPTION</u>	<u>COST</u>
C4T-1	1	450 gal	Caustic supply	\$2,500
C4T-2	1	300 gal	Fuel holding tank	4,400
C4T-3	1	20 gal	Sampling tank	600
C4T-4	1	300 gal	Fuel holding tank	4,400
C4T-5	1	20 gal	Sampling tank	600
C4T-6	1	150 gal	Salt holding tank	3,300
C4T-7	1	150 gal	Salt column feed tank	3,300
C4T-8	1	10 in. x 6 ft	Gas scrub tower	4,000
C4T-9	1	7 in. x 6 ft	Hydrofluorinator	4,600
C4T-10	1	7 in. x 6 ft	Fluorinator	4,000
C4T-11	1	5 in. x 3 ft	Trap	2,300
C4T-12	1	7 in. x 3 ft	UF <sub>6</sub> reducer	4,000
C4T-13	1	250 ft <sup>3</sup>	HF feed cylinder	300
C4T-14	1	250 ft <sup>3</sup>	F <sub>2</sub> feed cylinder	300
C4T-15	1	250 ft <sup>3</sup>	H <sub>2</sub> feed cylinder	300
C4T-16	1	250 ft <sup>3</sup>	He supply	300
C4T-17	1	3 ft <sup>3</sup>	Salt receiver	500
C4T-18	1	300 gal	Blanket liquid holdup tank	4,400
C4-C-1	1	25 ft <sup>2</sup>	UF <sub>6</sub> condenser	3,000
C4-C-2	1	25 ft <sup>2</sup>	HF condenser	3,000

<u>ITEM</u>	<u>QUANTITY</u>	<u>SIZE</u>	<u>DESCRIPTION</u>	<u>COST</u>
C4-E-1	1	25 l	U cell	\$4,500
C2-E-1	1	25 l	Mg cell	4,500
C2-E-2	1	25 l	Zr cell	4,500
C4-P-1	1	10 gpm	Caustic scrub pump	800
C4-P-2	1	10 gpm	HF recycle pump	1,500
C4-P-3	1	10 gpm	Cell feed pump	1,000
C4-P-4	1	10 gpm	Bi feed EM pump	2,000
C4-P-5	1	10 gpm	Bi feed EM pump	2,000
C4-P-6	1	10 gpm	Salt storage pump	800
C4-P-7	1	10 gpm	Salt storage pump	800
C4-P-8	1	10 gpm	Column salt feed	800
C4-P-9	1	10 gpm	Column salt feed	800
C4-P-10	1	10 gpm	Column salt feed	800
C4-P-11	1	10 gpm	Cell salt feed	800
			Tank heaters	12,000
			Samplers	54,000
			Subtotal	\$141,700
			Unlisted equip- ment - 20% of subtotal	\$ 28,400
			Total	\$170,100

b. Poison Removal by Zinc Precipitation Processing Rate -  
2 ft<sup>3</sup>/day (See Fig. 16.)

<u>ITEM</u>	<u>QUANTITY</u>	<u>SIZE</u>	<u>DESCRIPTION</u>	<u>COST</u>
C5T-10	1	4 in. x 3 ft	Zn contactor	\$1,000
C7T-3	1	50 gal	Bi receiver	800
C7T-4	1	50 gal	Distillation column	800
C3-T-3	1	4 in. x 5 ft	Feed tank distillation column	3,000
C5-T-1	1	50 gal	Process feed tank	800
C5-T-14	1	10 gal	Column feed weir box	500
C7-T-2	1	50 gal	Zn extract receiver	800
C3-T-1	1	50 gal	Bi receiver	800
C7-C-1	1	50 gal	Zn crystallizer	2,000
C7-C-2	1	50 gal	Zn crystallizer	2,000

<u>ITEM</u>	<u>QUANTITY</u>	<u>SIZE</u>	<u>DESCRIPTION</u>	<u>COST</u>
C7-H-6	1	5 kw	Gas heater	\$1,500
C7-H-7	1	5 kw	Gas heater	1,500
C3-C-1	1	10 ft <sup>2</sup>	Cooling coil	1,000
C3-H-3	1	10 kw	Heating coils	3,000
C5-P-1	1	10 gpm	Bi feed EM pump	2,000
C7-P-2	1	5 cfm at STP	Air blower	500
C7-P-3	1	5 cfm at STP	Air blower	500
C7-P-4	1	10 gpm	Zn feed pump	800
C7-T-5	1	50 gal	Zn supply tank	800
C3-P-1	1	3/4 hp	Vacuum pump	1,000
			Tank heaters	4,000
			Samplers at \$6,000	24,000
			Total	<u>\$53,100</u>

c. Head End (Bismuthide Blanket Phase Separation and  
Reconstitution) - 20 ft<sup>3</sup>/day to Volatility Plant (See Fig 17.)

<u>ITEM</u>	<u>QUANTITY</u>	<u>SIZE</u>	<u>DESCRIPTION</u>	<u>COST</u>
C8-T-2	1	150 gal	Bi hold tank with heaters	\$ 4,000
C8-T-3	1	400 gal	Slurry mixer agitator heater	12,400
C8-T-1	1	50 gal	Salt supply with heaters	1,800
C8-H-1	1	100 gal	Heat pulser induction heater agitator	12,000
C8-C-1	1	200 gal	Cooler agitator	9,000
C8-F-1	1	50 gal	Filter agitator	8,000
C8-E-1	1	25 l	Th feed electrolytic cell	4,500
C8-P-1,2,3, 4,5,6		10 gpm	Process pumps	6,000
			Heaters	2,000
			Samplers	<u>24,000</u>
			Subtotal	\$83,700
			Unlisted equip- ment 20% of subtotal	<u>16,800</u>
			Total	\$100,500

FIG. 15:  
LMFR FLUORIDE VOLATILITY PROCESS

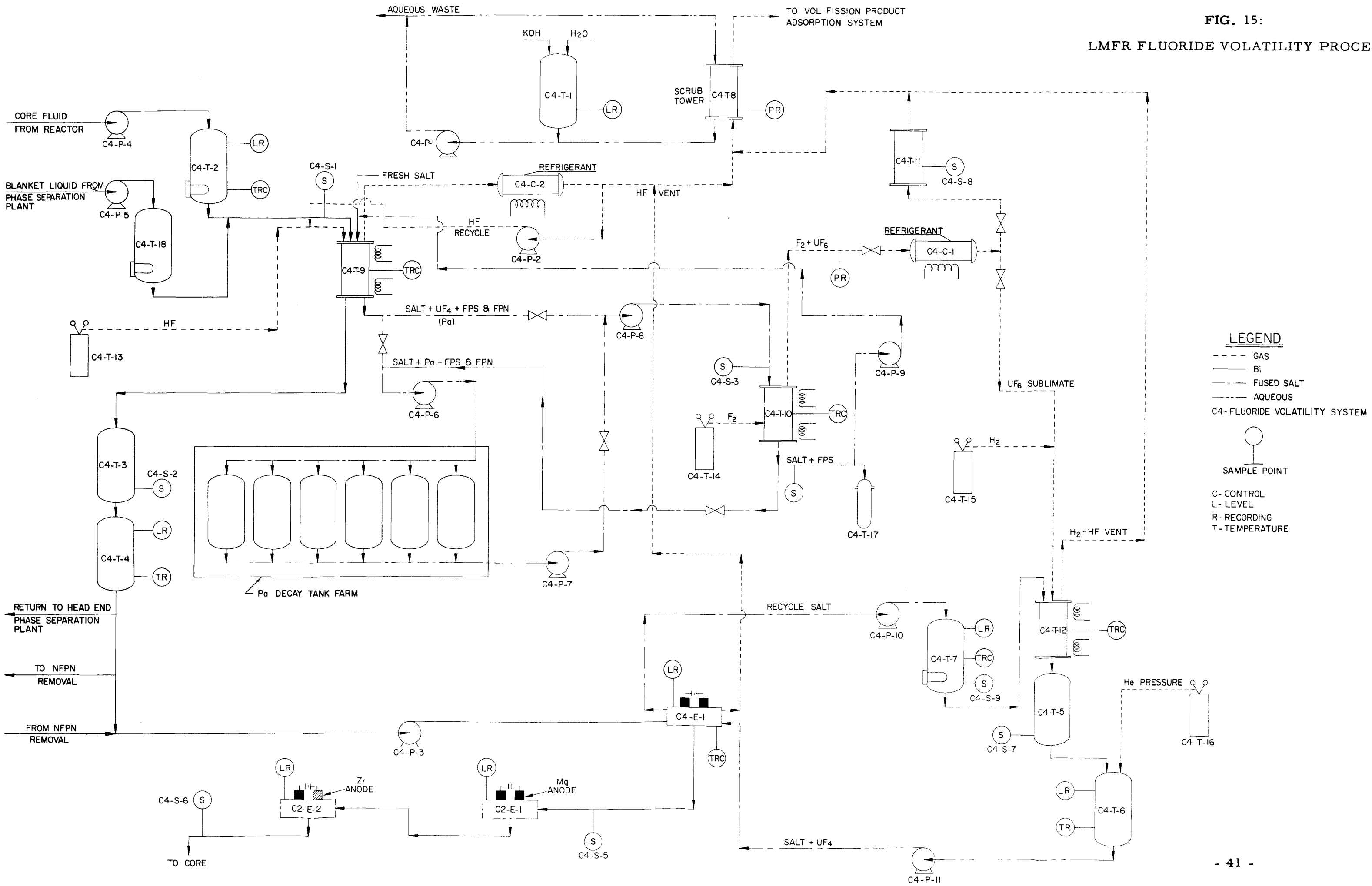


FIG. 16: ZINC EXTRACTION SYSTEM, NFPN REMOVAL

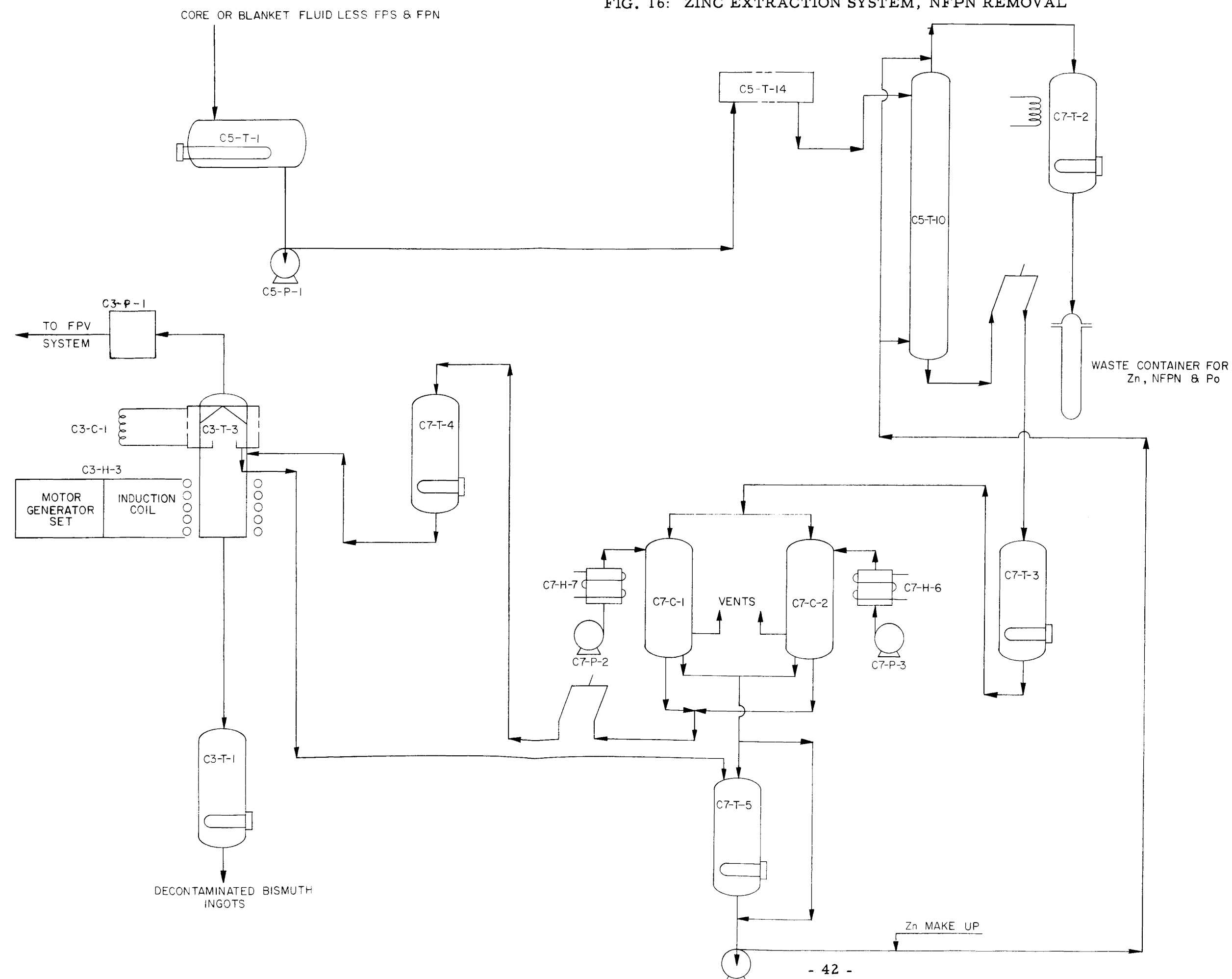
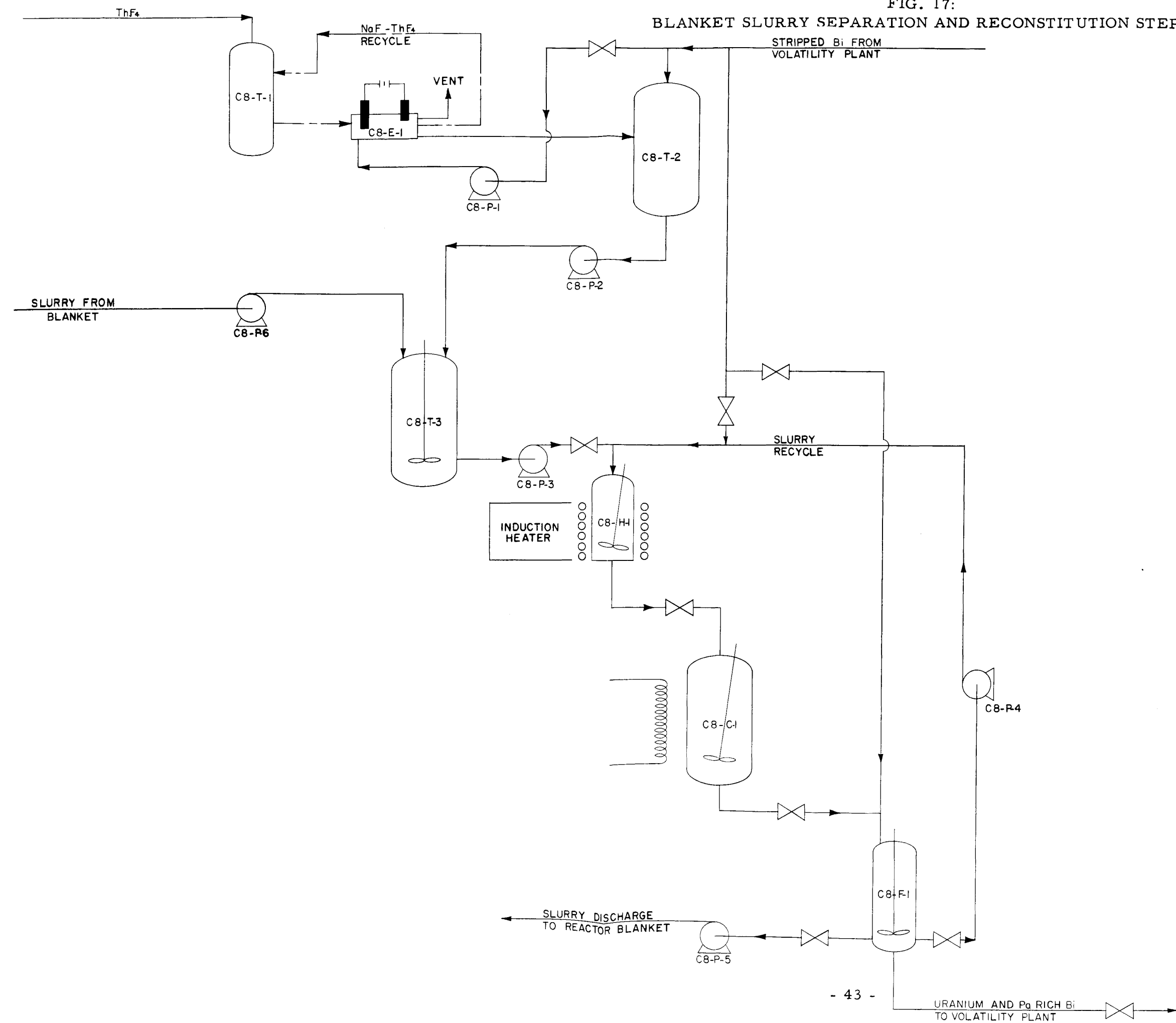




FIG. 17:

## BLANKET SLURRY SEPARATION AND RECONSTITUTION STEP



d. Other Chemical Plant Items (Bi Processing)

Salt purification system	\$ 27,000
Auxiliary fuel makeup	85,000
Protactinium decay tank farm	68,000
Chemistry laboratory equipment	300,000
Total	\$480,000

2. Building Cost Estimates

a. Process Buildings Cost Estimates

(1) Space Requirements, Main Processing Building

- (a) Process equipment requires 2 cells, each 20 by 20 by 15 ft.
- (b) The protactinium decay tanks require a cell 18 by 12 by 7 ft.
- (c) A space 20 by 20 by 15 ft is required for cold area equipment.
- (d) Plant control area requirements are about 400 sq ft.

(2) Direct Construction Cost (Main Processing Building)

HOT AREA	LABOR	MATERIALS	CONSTRUCTION COST
Wall construction (Includes piping fixtures)	\$102,000	\$152,000	\$254,000
Cell liners	19,000	19,000	38,000
Windows	2,000	10,000	12,000
COLD AREA	16,000	16,000	32,000
	\$139,000	\$197,000	\$336,000

b. Direct Construction Costs (Processing Plant Total Building)

	LABOR	MATERIALS	CONSTRUCTION COST
Hot processing area	\$139,000	\$197,000	\$336,000
Hot laboratory	75,000	105,000	180,000
Offices, health physics, etc.	40,000	40,000	80,000
Maintenance tools	8,000	20,000	28,000
Cold laboratory	25,000	25,000	50,000
	\$287,000	\$387,000	\$674,000

c. Blanket Slurry Separation and Reconstitution (Head-End Process) Basis 20 ft<sup>3</sup>/day Liquid Processing Rate

	<u>BASE LABOR</u>	<u>BASE MATERIAL</u>	<u>CONSTRUCTION COST</u>
Material cost		100	
Installation labor cost (15% of material cost)	15		
Installed equipment			115
Piping materials		66	
Piping installation	49		
Instruments		23	
Installation of instruments	23		
Construction cost	87	189	276

d. Other Plant Items (Salt Purification, Fuel Makeup, Protactinium Decay Tanks, Laboratory Equipment)

	<u>BASE LABOR</u>	<u>BASE MATERIAL</u>	<u>CONSTRUCTION COST</u>
Material cost		480	
Installation labor cost (15% of material cost)	72		
Installed equipment			552
Piping materials		120	
Piping installation	90		
Instruments		42	
Installation of instruments	40		
Construction costs	202	642	<u>844</u>

3. Direct Construction Cost (Thousands of Dollars)

a. FPS and FPN Removal (Basis 20 ft<sup>3</sup>/day Processing Rate)  
By Fluoride Volatility Process

	<u>BASE LABOR</u>	<u>BASE MATERIAL</u>	<u>CONSTRUCTION COST</u>
Material cost		170	
Installed labor cost (15% of material cost)	26		
Installed equipment			196

	<u>BASE LABOR</u>	<u>BASE MATERIAL</u>	<u>CONSTRUCTION COST</u>
Piping material		112	
Piping installation	84		
Instruments		39	
Installation of instruments	39		
Construction costs	149	321	470

b. NFPN Processing (Zn Precipitation) Based on 2 ft<sup>3</sup>/day  
Processing Rate

	<u>BASE LABOR</u>	<u>BASE MATERIAL</u>	<u>CONSTRUCTION COST</u>
Material cost		53	
Installation labor cost (15% of material cost)	8		
Installed equipment			61
Piping materials		35	
Piping installation	26		
Instruments		23	
Installation of instruments	23		
Construction costs	57	111	168

4. Direct Construction Cost Variation with Processing Rate  
Labor and Materials Breakdown

a. Cost Categories (Thousands of Dollars)

	<u>Fixed</u>		<u>Variable</u>		<u>Variable</u>	
	<u>Labor</u>	<u>Materials</u>	<u>Labor 0.25 Scale Factor</u>	<u>Materials Scale Factor</u>	<u>Labor 0.7 Scale Factor</u>	<u>Materials Scale Factor</u>
FPS and FPN	39	39			110	282
NFPN	23	25			34	86
Head-end	23	23			64	166
Other plant items	202	642				
Building	<u>25</u>	<u>25</u>	<u>262</u>	<u>362</u>	<u>        </u>	<u>        </u>
Totals	312	754	262	362	208	534

b. Variation with Processing Rate in Thousands of Dollars  
(Basic Rate 20 ft<sup>3</sup>/day to Volatility Plant and 2 ft<sup>3</sup>/day  
to NFPN Removal Plant)

Process- ing Rate ft <sup>3</sup> /day to Vola- tility Plant	Fixed		Variable (0.25 Scale Factor)		Variable (0.7 Scale Factor)		Total	
	Labor	Mat'ls	Labor	Mat'ls	Labor	Mat'ls	Labor	Mat'ls
2	312	754	148	204	42	107	502	1065
5	312	754	188	258	79	203	579	1215
10	312	754	221	306	129	331	662	1391
20	312	754	262	362	208	534	782	1650
50	312	754	320	440	395	1010	1027	2204
100	312	754	396	546	643	1650	1351	2950
200	312	754	471	649	1040	2670	1823	4073

5. Capital Investment, Annual Charges

a. All Chemical Plant Buildings (Basic Processing Rate)

Direct labor	\$ 287,000
Direct material	<u>387,000</u>
Direct construction	674,000
Indirect construction, 73% of direct labor - \$209,000	
Total construction cost	\$ 883,000
Engineering, 15% of total construction	\$132,000
Engineering and total construction	<u>\$1,015,000</u>
Contingency, 15% of engineering plus total construction	\$152,000
Total capital investment	<u>\$1,167,000</u>

b. Scaleup of Building Cost with Processing Rate

For buildings, the ratio of total capital investment to direct construction cost is 1.73. This factor is used to convert direct construction costs to total capital investment for a range of plant throughputs.

<u>Processing Rate ft<sup>3</sup>/day to Volatility Plant</u>	<u>Direct Construction Cost</u>	<u>Total Capital Investment</u>	<u>Annual Costs at 13.5%/Yr (30-Yr Amortization)</u>
2	\$ 402,000	\$ 695,000	\$ 94,000
5	496,000	857,000	116,000
10	577,000	1,000,000	135,000
20	674,000	1,168,000	158,000
50	810,000	1,400,000	189,000
100	992,000	1,716,000	231,000
200	1,170,000	2,020,000	273,000

c. All Chemical Plant Processes (Basic Processing Rate)

Direct labor		\$ 495,000
Direct materials		<u>1,263,000</u>
Direct construction cost		\$1,758,000
Indirect construction cost		
73% of direct labor	\$362,000	
Plant startup		
20% of direct construction	352,000	
Total construction and startup turn-key plant		2,472,000
Engineering, 20% of turn-key plant	494,000	
Engineering and turn-key plant		2,966,000
Contingency 28% of engineering and turn-key plant	840,000	
Total capital investment		<u>\$3,806,000</u>

d. Scaleup of Capital Equipment with Processing Rate

For plant equipment, the ratio of total capital investment of direct construction is 2.17. This factor is used to convert direct construction costs to total capital investment for a range of plant throughputs.

<u>Processing Rate ft<sup>3</sup>/day to Volatility Plant</u>	<u>Direct Construction Cost</u>	<u>Total Capital Investment</u>	<u>Annual Cost at 24.5%/Yr (7-Yr Life)</u>
2	\$1,165,000	\$2,530,000	\$ 620,000
5	1,298,000	2,815,000	690,000
10	1,476,000	3,200,000	784,000

<u>Processing Rate ft<sup>3</sup>/day to Volatility Plant</u>	<u>Direct Construction Cost</u>	<u>Total Capital Investment</u>	<u>Annual Cost at 24.5%/Yr (7-Yr Life)</u>
20	\$1,758,000	\$ 3,810,000	\$ 933,000
50	2,418,000	5,240,000	1,285,000
100	3,409,000	7,400,000	1,815,000
200	4,726,000	10,270,000	2,510,000

## 6. Operating Costs

### a. Plant Labor and Associated Cost

Operating labor 3 men/shift	\$ 66,000/yr
Maintenance labor 0.5 operating labor	33,000/yr
Supervision 0.5 operating labor	33,000/yr
Overhead 100% of operating, maintenance labor, and supervision	132,000/yr
Maintenance materials and operating supplies - Maintenance materials ~ 1.25 times maintenance labor	42,000/yr
Operating supplies = 0.1 times operating + maintenance labor	10,000/yr
Total	<u>\$316,000/yr</u>

### b. Chemical Laboratories, Labor and Associated Cost

Chemists 2 at \$9,000/yr	\$ 18,000/yr
Technicians 8 at \$7,000/yr	56,000/yr
Chemicals	25,000/yr
Maintenance and Operating Supplies 0.1 x labor	7,400/yr
Overhead 100% x labor	74,000/yr
Total	<u>\$180,400/yr</u>

### c. Total Plant Utilities

The chemical plant utilities cost depends on throughput. Most of the cost is for electricity required to melt process fluid, maintain these fluids in the molten state and pump them. The head-end process will require additional electricity. This process is necessary to separate liquid from solid, and to provide heat pulsing to establish equilibrium of protactinium, uranium, and fission products between solids and liquid.

The power requirements were estimated from similar processes using electricity.

$$\frac{\text{Core liquid}}{\text{at 8 mil power, } 0.75 \frac{\text{kwh}}{\text{lb of Bi}}} \times \frac{\$.008}{\text{kwh}} = \$.006 \text{ lb processed}$$

$$\frac{\text{Blanket liquid}}{\text{at 8 mil power, } 1.0 \frac{\text{kwh}}{\text{lb}}} \times \frac{\$.008}{\text{kwh}} = \$.008 \text{ lb processed}$$

$$\begin{aligned} \frac{\text{On a volume basis}}{\text{Core liquid}} &= \$3.6/\text{ft}^3 \text{ of Bi solution} \\ \text{Blanket liquid} &= \$4.8/\text{ft}^3 \text{ of slurry} \end{aligned}$$

### d. Chemical Costs for Fuel Processing

No requirement for fission products is included, since these chemical requirements depend only on reactor power level.

For fuel cleanup in a fluoride volatility system:

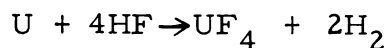
$$\begin{aligned} \text{Dollars}/\text{ft}^3 \text{ of fuel} = & \quad (1) \text{ cost for uranium oxidation to UF}_4 \\ & + (2) \text{ cost for Zr oxidation} \\ & + (3) \text{ cost for Mg oxidation} \\ & + (4) \text{ cost for Zr consumption} \\ & + (5) \text{ cost for Mg consumption} \\ & + (6) \text{ cost for UF}_4 \text{ oxidation to UF}_6 \\ & + (7) \text{ cost for uranium reduction (H}_2\text{)} \\ & + (8) \text{ cost for uranium reduction UF}_4 \text{ to} \\ & \quad \text{U metal, electrolysis} \\ & + (9) \text{ cost for oxidation of corrosion} \\ & \quad \text{products} \end{aligned}$$

Let  $c$  = ppm uranium in bismuth.



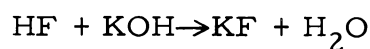
Item (1)

(a) HF, hydrofluorination at 10% utilization



$$\$/\text{ft}^3 = (610 \text{ c}/10^6) (4/233) (20/0.1) \times \$ .25/\text{lb} = 5.23 \times 10^{-4} \text{ c}$$

(b) KOH, neutralization of HF effluent at 90% utilization

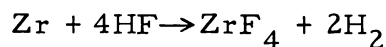


$$\$/\text{ft}^3 = (610 \text{ c}/10^6) (4/233) (9/0.9) (56.1) \times \$ .12/\text{lb} = 7.05 \times 10^{-4} \text{ c}$$

$$\text{Total (a) + (b) = } 1.23 \times 10^{-3}$$

Item (2) Based on 80 ppm Zr in fuel

(a) HF, hydrofluorination at 10% utilization



$$\$/\text{ft}^3 = (610 \times 80/10^6) (4/91.2) (20/0.1) \times \$ .25/\text{lb} = 0.107$$

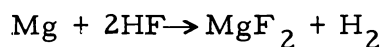
(b) KOH, neutralization at 90% utilization

$$\$/\text{ft}^3 = (610 \times 80/10^6) (4/91.2) (9/0.9) (56.1) \times \$ .12/\text{lb} = 0.144$$

$$\text{Total (a) + (b) = } 0.251$$

Item (3) Based on 350 ppm Mg in fuel

(a) HF, hydrofluorination at 10% utilization



$$\$/\text{ft}^3 = (610 \times 350/10^6) (2/24.3) (20/0.1) \times \$ .25/\text{lb} = 0.880$$

(b) KOH, neutralization at 90% utilization

$$\$/\text{ft}^3 = (610 \times 350/10^6) (2/24.3) (9/0.9) (56.1) \times \$ .12/\text{lb} = 1.18$$

$$\text{Total (a) + (b) = } 2.06$$

Item (4)

$$\$/\text{ft}^3 = (610 \times 80/10^6) \times \$10.00/\text{lb} = 0.487$$

Item (5)

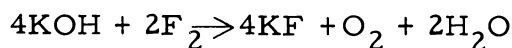
$$\$/\text{ft}^3 = (610 \times 350/10^6) \times \$ .27/\text{lb} = 576 \times 10^{-4} = 0.057$$

Item (6)  $\text{UF}_4 + \text{F}_2 \rightarrow \text{UF}_6$

(a)  $\text{F}_2$ , fluorination at 10% utilization

$$\$/\text{ft}^3 = (610 \text{ c}/106) (1/233) (38/0.1) \times \$2.00/\text{lb} = 1.99 \times 10^{-3} \text{ c}$$

(b) KOH, neutralization of excess  $\text{F}_2$  at 90% utilization



$$\$/\text{ft}^3 = (610 \text{ c}/10^6) (2/233) (9/0.9) (56.1) \times \$.12/\text{lb} = 3.54 \times 10^{-4} \text{ c}$$

$$\text{Total (a) + (b) = } 2.34 \times 10^{-3} \text{ c}$$

Item (7)

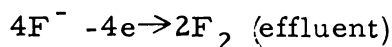
(a)  $\text{H}_2$  reduction  $\text{UF}_6 + \text{H}_2 \rightarrow \text{UF}_4 + 2\text{HF}$

Hydrogen cost is negligible.

(b) KOH neutralization

$$\$/\text{ft}^3 = (610 \text{ c}/10^6) (2/233) (56.1/0.9) \times \$.12/\text{lb} = 3.91 \times 10^{-5} \text{ c}$$

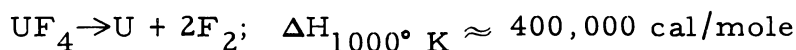
Item (8)  $\text{U}^{++++} + 4\text{e}^- \rightarrow \text{U}(\text{metal})$



(a) KOH, neutralization of  $\text{F}_2$  effluent

$$\$/\text{ft}^3 = (610 \text{ c}/10^6) (4/233) (56.1/0.9) \times \$.12/\text{lb} = 7.82 \times 10^{-5} \text{ c}$$

(b) Power charges for  $\text{UF}_4$  reduction



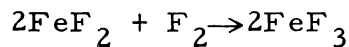
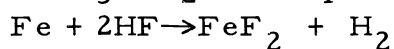
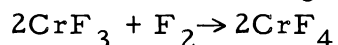
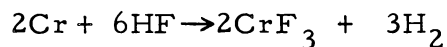
$$400 \text{ K-cal/mole} = 400 \times 1.163 \times 10^{-3} \times 454/233 = 0.91 \text{ kwh/lb of U}$$

If the cell operates at 20% energy efficiency

$$\$/\text{ft}^3 = (0.91/0.2) (610 \text{ c}/10^6) \times \$.01/\text{kwh} = 2.78 \times 10^{-5} \text{ c},$$

Item (9) Assume liquid saturation at 550 C, equivalent to about 30 ppm Fe and 80 ppm Cr.

(a) HF and  $\text{F}_2$  requirements



HF requirement

$$\$/\text{ft}^3 = 610/10^6 \left[ (3)80/52 + (2) (30)/55.8 \right] 20/0.1 \times \$ .25/\text{lb} = 0.174$$

F<sub>2</sub> requirement

$$\$/\text{ft}^3 = 610/10^6 \left[ 80/(52)(2) + 30/(2)(55.8) \right] 38/0.1 \times \$2.00/\text{lb} = 0.482$$

(b) KOH, neutralization requirement

HF neutralization

$$\$/\text{ft}^3 = (0.174/0.25) (0.9/20) 56.1/0.9 \times \$ .12/\text{lb} = 0.235$$

F<sub>2</sub> neutralization

$$\$/\text{ft}^3 = (0.482/2.0) (0.9/38) (2) (56.1)/0.9 \times \$ .12/\text{lb} = 0.085$$

$$\text{Total (a) + (b) = 0.976}$$

Summary

Cost of chemicals/cu ft of bismuth core liquid processed,  
where c is the concentration of U in ppm, is as follows.

$$\$/\text{ft}^3 = 3.72 \times 10^{-3} c + 3.83$$

e. Chemical Costs for Blanket Processing

After liquid-solids separation it is assumed that the liquid will be saturated with thorium and that this will be a process loss. Any thorium recovery would involve a step to separate thorium fluorides from fission product fluorides.

The chemical requirements for protactinium and uranium are assumed to be identical, since the protactinium will be held as  $\text{PaF}_4$  in a salt and then fluorinated to  $\text{UF}_6$  after protactinium decay.

The liquid is assumed to contain corrosion products and zirconium, in addition to the fertile and fissile material.

For blanket fluid cleanup in a fluoride volatility system:

$$\begin{aligned} \text{Dollars}/\text{ft}^3 \text{ of fuel} = & \quad (1) \text{ cost for Pa and U oxidation to } \text{UF}_4 \\ & + (2) \text{ cost for Zr oxidation} \\ & + (3) \text{ cost for Th oxidation} \\ & + (4) \text{ cost for Zr consumption} \\ & + (5) \text{ cost for Th consumption} \\ & + (6) \text{ cost for } \text{UF}_4 \text{ oxidation to } \text{UF}_6 \\ & + (7) \text{ cost for } \text{UF}_6 \text{ reduction to } \text{UF}_4 \\ & + (8) \text{ cost for } \text{UF}_4 \text{ reduction to U} \\ & + (9) \text{ cost for oxidation of corrosion} \\ & \quad \text{products} \end{aligned}$$

Let  $c'$  = ppm uranium plus protactinium in the liquid

Item (1) Same as for core processing except  $c'$  is substituted for  $c$

$$\text{Total} = 1.23 \times 10^{-3} c' \text{ \$/ft}^3$$

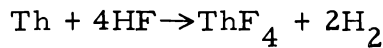
Item (2) Same as for core processing,

$$\text{Total} = 0.251 \text{ \$/ft}^3$$

Item (3) The amount of thorium present in the liquid phase will depend on the temperature of the final solid-liquid separation step.

In a volatility system, all this thorium will be oxidized and lost. If the separation is done at 350 C, the concentration of thorium is about 60 ppm.

(a) HF, hydrofluorination at 10% utilization



$$\text{\$/ft}^3 = (610 \times 60/10^6) (4/232) (20/0.1) \times \$ .25/\text{lb} = 3.14 \times 10^{-2}$$

(b) KOH, neutralization of HF effluent at 90% utilization

$$\text{\$/ft}^3 = (610 \times 60/10^6) (4/232) (9/0.9) 56.1 \times \$ .12/\text{lb} = 4.23 \times 10^{-2}$$

$$\text{Total} = 7.37 \times 10^{-2}$$

Item (4) Same as for core processing

$$\text{\$/ft}^3 = 0.487$$

Item (5)

$$\text{\$/ft}^3 = (610 \times 60/10^6) \times \$20.00/\text{lb} = 0.732$$

Item (6) Same as for core processing except  $c'$  is substituted for  $c$

$$\text{Total} = 2.34 \times 10^{-3} c', \text{ \$/ft}^3$$

Item (7) Same as for core processing except  $c'$  is substituted for  $c$

$$\text{Total} = 3.91 \times 10^{-5} c'$$

Item (8) Same as for core processing except  $c'$  is substituted for  $c$

Item (9) Same as for core processing

$$\text{Total} = 0.976 \text{ \$/ft}^3$$

Summary

Cost of chemicals/cu ft of bismuth liquid processed where  $c'$  is the concentration of Pa plus U in ppm is as follows.

$$\text{\$/ft}^3 = 3.72 \times 10^{-3} c' + 2.52$$

f. Chemical and Waste Disposal Costs Directly Attributable  
to the Removal and Handling of Fission Products

(1) Oxidation - Reduction Costs - FPS and FPN Processing

Fission product production and hence the chemical costs involved in their removal are largely independent of the cycle time. This is also true of the waste charges resulting from concentration of these fission products into disposable form.

The fission products have been divided into four groups (BAW-2): FPV, FPS, FPN, and NFPN. The fluoride volatility process removes all elements in the FPS and FPN groups. The volatile, FPV, group will be degassed from the fuel but the degassing efficiency will affect the number of FPS's in the fuel, since many of these are decay daughters of FPV's. For this analysis all of the Kr-85 and Xe-133, and half of the Xe-135, is considered removed before decay. The daughter products of all other unstable gases are considered to be in the fuel.

$$(825 \times 1.06) = 870 \text{ gms total FP/day}$$

From BAW-2,

NFPN: 226 NFPN atoms are formed per 1000 fissions.

Since the average NFPN atomic weight is 100

$$[100 (226)/233 (1000)] \quad 870 = 84 \text{ gm NFPN/day.}$$

Similarly,

$$\text{FPV: } [127 (395)/233 (1000)] \quad 870 = 187 \text{ gm FPV/day,}$$

$$\text{FPS: } [124 (951)/233 (1000)] \quad 870 = 442 \text{ gm FPS/day, and}$$

$$\text{FPN: } [98 (428)/233 (1000)] \quad 870 = 157 \text{ gm FPN/day.}$$

If the average FPS and FPN valence is assumed to be +3, the HF requirement for oxidation is

$$\begin{aligned} [442/124 + 157/98] \quad 3 &= 1.55 \text{ equivalents/day,} \\ &= 0.03 + \text{lb-equivalents/day.} \end{aligned}$$

Daily HF cost at 10% utilization is

$$0.034 \times 20 \times 1/0.1 \times \$ .25 = \$1.70/\text{day}$$

Daily KOH cost at 90% utilization is

$$0.034 \times 0.9/0.1 \times 1/0.9 \times 56 \times 0.12 = \$2.30/\text{day.}$$

Total reagent cost is \$4.00/day

(2) Waste Disposal Cost - FPS and FPN Processing

(a) High Level Wastes

Cost of salt matrix  $\text{ZrF}_4\text{-NaF}$

Assume a 1.5% fission product concentration  
in the waste salt.

Therefore  $(599/0.015) (1/454) = 88 \text{ lb salt/day}$   
at  $\$2.85/\text{lb} = \$250.00/\text{day}$ .

Cost of disposal at  $\$50.00/\text{ft}^3$  is  
 $(88/200) 50 = \$22.00/\text{day}$ .

(b) Low Level Waste

This item depends on cycle time and uranium concentration.  
There are two costs involved, (1) evaporation of the  
scrub caustic solution to dryness, and (2) disposal of the  
resulting solids.

Item (1) For 20% KOH solution at  $\$.10/\text{gal}$  total processing cost  
including amortization,

$$\$/\text{ft}^3_{\text{Bi}} = \left( \frac{9.85 \times 10^{-3} c + 13.67}{0.20} \right) \left( \frac{7.5}{(62.4)(1.18)} \right) \times$$
$$$.10 = 5.04 \times 10^{-4} c + 0.70.$$

Item (2) At  $\$10.00/\text{cu ft}$  of solid formed,  
KOH required =  $9.85 \times 10^{-3} c + 13.67$ ,

$$\$/\text{ft}^3_{\text{Bi}} = \left( \frac{9.85 \times 10 c + 13.67}{(62.4) (2)} \right) \times \$10.00/\text{ft}^3 =$$
$$7.9 \times 10^{-4} c + 1.1.$$

Total Item (1) plus Item (2) =  $1.29 \times 10^{-3} c + 1.8$

This is the low level waste disposal cost, dollars/  
cu ft of fuel processed, with concentration of uranium  
(c) as a parameter.

### (3) NFPN and Polonium Processing Costs

#### (a) Chemical and Waste Disposal Costs

The exact process for NFPN removal is not firm. UCN-42 shows a zinc extraction process. This will not be economical unless there is a high fission product concentration in the exact phase, resulting from bismuth losses in the zinc-rich extract. Other processes are under development (H. Howe, ASR-private communication) which use zinc in such a way that a liquid - liquid phase separation is not involved. This will result in low bismuth losses. A material cost will be involved for the zinc matrix to hold the polonium and NFPN fission products.

The amount of zinc required per gram of NFPN is a function of the heat dissipating properties of the waste product ingot.

#### (b) Estimated heat release from NFPN fission products

Let  $N$  = total nuclides of a particular isotope in the reactor,

$k$  = fraction of  $N$  removed per second into the NFPN processing system,

$y$  = yield of nuclide at 825 MW power level, then

$$\frac{dN}{dt} = y - \lambda N - kN = 0, \text{ at equilibrium, and}$$

$$N = \frac{y}{\lambda + k}.$$

If the NFPN's are collected continuously for one day before zinc processing, then the maximum decay rate for that batch is

$$kN(1 - e^{-\lambda t}), \text{ disintegrations/sec, where } t = 1 \text{ day.}$$

Since the NFPN group does not represent a severe poison, a suitable cycle time might be 300 days.

$$k = \frac{1}{(300)(24)(3.6 \times 10^3)} = 3.86 \times 10^{-8} \frac{1}{\text{sec}}$$

TABLE X  
FISSION RATE AT 825 MW =  $2.56 \times 10^{19}$  FISSIONS/SEC

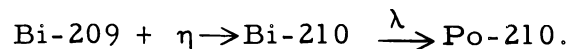
Nuclide	Yield, $\frac{\text{Atom}}{\text{Sec}}$ at 825 MW	$\lambda$	N	
Mo-99	$1.2 \times 10^{18}$	$2.9 \times 10^{-6}$	$4.2 \times 10^{23}$	
Ru-103	$2.2 \times 10^{17}$	$2.0 \times 10^{-7}$	$9.3 \times 10^{23}$	
Rh-105	$1.2 \times 10^{17}$	$5.3 \times 10^{-6}$	$2.2 \times 10^{22}$	
Ru-106	$8.5 \times 10^{16}$	$2.2 \times 10^{-8}$	$1.4 \times 10^{25}$	
Rh-108	$3.3 \times 10^{16}$	$2.1 \times 10^{-5}$	$1.5 \times 10^{21}$	
Pd-109	$8.0 \times 10^{15}$	$1.4 \times 10^{-5}$	$5.7 \times 10^{20}$	
Pd-112	$5.1 \times 10^{15}$	$9.2 \times 10^{-6}$	$5.5 \times 10^{20}$	

Nuclide	$e^{-\lambda t}$	Decay Rate $kN(1 - e^{-t})$	$\frac{\text{Mev}}{\text{Dis}}$	$\frac{\text{Mev}}{\text{Sec}}$
Mo-99	0.78	$3.6 \times 10^{15}$	0.61	$2.3 \times 10^{15}$
Ru-103	0.98	$7.2 \times 10^{14}$	0.57	$4 \times 10^{15}$
Rh-105	0.63	$3.3 \times 10^{14}$	0.21	
Ru-106	$1 - e^{-\lambda t} =$ 0.0019	$1.0 \times 10^{15}$	1.31	$1.3 \times 10^{15}$
Rh-108	0.16	$4.8 \times 10^{13}$	0.50	
Pd-109	0.30	$1.5 \times 10^{13}$	0.32	
Pd-112	0.45	$1.2 \times 10^{13}$	0.09	
Total =				$4.0 \times 10^{15} =$ $0.61 \frac{\text{Btu}}{\text{sec}}$

(c) Estimated heat release from Po-210, since in the fuel processing system it is likely to act as an NFPN.

Estimating the number of Po-210 atoms in an 825 MW reactor employing 800 ppm U-233 in bismuth, for the Po-210 chain,



Equilibrium Bi-210 decay rate or Po-210 production rate is given by

$$c \sigma^{209} \phi V$$

where

$c$  = atoms Bi-209/cu cm,

and

$(\phi V)$  is the product of the flux and total volume in the flux.



If Po-210 is withdrawn from the reactor at the same rate as NFPN's, then for Po-210 in the reactor, neglecting neutron captures in Po-210,

$$\frac{dN}{dt} = c \sigma^{209} \phi V - \lambda N - kN = 0,$$

at equilibrium

$$N = \frac{c \sigma^{209} \phi V}{\lambda + k}.$$

The maximum decay rate from one day's chemical processing collection is

$$k \frac{c \sigma^{209} \phi V}{\lambda + k} (1 - e^{-\lambda t}),$$

where

$$\begin{aligned} t &= 1 \text{ day}, \\ k &= 3.9 \times 10^{-8}, \\ \phi V &= 3.6 \times 10^{21}, \text{ for 800 ppm U and 825 MW,} \\ \lambda &= 5.8 \times 10^{-8}, \\ c &= 2.9 \times 10^{22}, \\ \sigma^{209} &\approx 10^{-26}, \text{ and} \\ N &= 1.1 \times 10^{25}. \end{aligned}$$

$$\begin{aligned} \text{Decay rate, dis/sec} &= (3.9 \times 10^{-8}) (1.1 \times 10^{25}) \\ &= 4.3 \times 10^{17} \end{aligned}$$

$$\begin{aligned} \text{Energy release from 1 day's collection} \\ (2.1 \times 10^{15}) (5.3) (1.5) (10^{-16}) &= 1.6 \frac{\text{Btu}}{\text{sec}} \end{aligned}$$

(d) To estimate the waste disposal costs for NFPN plus Po-210, assume that the NFPN plus Po-210 are cast into a zinc slab, which is unclad and which is cooled by natural air convection.

$$\begin{aligned} \text{The total heat to be dissipated by this volumetric source is} \\ 0.61 + 1.6 &= 2.25 \frac{\text{Btu}}{\text{sec}} = 8100 \frac{\text{Btu}}{\text{hr}}. \end{aligned}$$

For an infinite slab, cooled by air convection on both sides the heat dissipating capability is

$$Q = \frac{\Delta t}{a \left( \frac{a}{2k} + \frac{1}{h} \right)},$$

where

Q = source strength,

$\Delta t$  = temperature drop from slab midpoint to air,

a = slab half thickness,

h = natural convection coefficient,

$\Delta t$  = 300 F (giving a maximum Zn temperature of 380 F at 80 F ambient),

a = 0.083 ft (for a 2-in. thick slab),

k = 55  $\frac{\text{Btu}}{\text{hr ft}^2 \text{ } ^\circ\text{F/ft}}$ ,

h = 2  $\frac{\text{Btu}}{\text{hr ft}^2 \text{ } ^\circ\text{F}}$ , and

$$Q = \frac{300}{0.083 \left( \frac{0.083}{(2) 55} + 0.5 \right)} = 7,150 \frac{\text{Btu}}{\text{hr ft}^3}.$$

Zn matrix requirement is  $8,100/7,150 = 1.12 \text{ ft}^3$   
at \$.18/lb, \$/day for Zn = (62.4) (6.8) (1.12) (0.18) = \$87.00.

Additional charge for waste disposal at  
\$50.00/  $\text{ft}^3$  = \$56.00.

Total daily charge = \$143.00.

#### (4) Summary of Waste Costs

Salt matrix cost	\$250.00/day
Salt disposal cost	\$ 22.00/day
Low level waste disposal	
C = ppm uranium in bismuth feed	$\$(1.29 \times 10^{-3} C + 1.8)/\text{ft}^3$
Zn matrix cost	\$ 87.00/day
Zn disposal cost	<u>\$ 56.00/day</u>
Total -	<u>\$415.00/day</u> $\$(1.29 \times 10^{-3} C + 1.8)/\text{ft}^3$

#### g. Summary of Chemical Plant Operating Costs

The operating charges can be divided into several basic categories, (1) constant costs, (2) costs varying with bismuth throughput but independent of uranium concentration, (3) costs varying with throughput of uranium and (4) costs varying with reactor heat output. Table XI gives these charges.

TABLE XI  
COST CATEGORY

<u>Item</u>	<u>(1) (\$/yr)</u>	<u>(2) (\$/ft)</u>	<u>(3) (\$/ft)</u>	<u>(4) (\$/full power-day)</u>
1. Plant labor	316,000			
2. Laboratory	180,400			
3. Utilities		Core liquid 3.6		
		Blanket liquid 4.8		
4. Core liquid processing		3.8	$(3.7 \times 10^{-3}) c$	
5. Blanket liquid processing		2.5	$(3.7 \times 10^{-3}) c'$	
6. Waste disposal		1.8	$(1.3 \times 10^{-3}) c$	\$415.00
7. Totals				
Based on blanket liquid	496,400	9.1	$(5.0 \times 10^{-3}) c$	\$415.00

$c$  = ppm uranium in core liquid

$c'$  = ppm uranium plus protactinium in blanket liquid

#### h. Annual Operating Cost (See Fig. 7.)

Labor costs are assumed to vary as the 0.25 power of the plant throughput.

Basic labor cost is for a plant throughput of 20 cu ft/day.

Curve is based on blanket liquid with a concentration of 1000 ppm uranium plus protactinium.

Curve is based on 292 days/yr reactor operation at full power, 825 MW.

Curve is based on 295 days/yr chemical plant on-stream operation.

## APPENDIX C

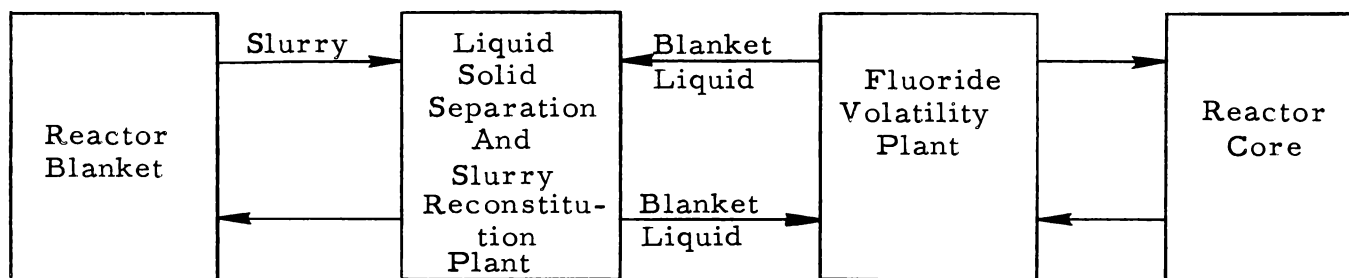
### HEAD-END SEPARATION PROBLEMS IN THE VOLATILITY PROCESS

#### 1. Introduction

In a two-region reactor with thorium in the blanket, Pa-233 and U-233 are generated from neutron capture in thorium followed by beta decay. It will be necessary to transfer bred U-233 and its precursor Pa-233 out of the blanket region into the core region so that better over-all neutron economy may be achieved. It will also be necessary to achieve this transfer to hold the power level of the blanket at the design point.

There is a possibility that a fraction of the liquid from the blanket could be enriched with uranium and sent into the core system. This alternate, when carried to the limit, ends in a single-region reactor, so its possibilities are not examined here.

A number of alternatives are possible in blanket processing. This arises from a consideration of the head-end part of the process. The over-all system is shown below.



The method of handling a slurry in this type of processing is to separate the solids from the liquid, and then process the liquid. However, bred uranium and protactinium does not transfer completely to the liquid phase. The solids may be repeatedly washed with fresh batches of bismuth and heat pulsed to remove bred material and fission products, but as greater volumes of bismuth wash are used, plant size and cost increases accordingly. Each subsequent wash is less

effective than the first. The amount of liquid processed for a slurry batch should be equivalent to the volume of liquid phase of the slurry batch.

The slurry from the blanket contains thorium, uranium, protactinium and fission products. These are distributed between the solid and liquid phases of the slurry. BNL work has indicated a distribution coefficient of about 6 for uranium and 60 for protactinium. No data are available for fission products. Work on the effect of various variables on the distribution coefficient for uranium and protactinium is not complete. The distribution coefficient ( $k$ ) as defined by BNL for this work is pounds of U-233 per pound of solid divided by pounds of U-233 per pound of liquid.

To transfer bred material at a given rate from the blanket to the volatility plant, both the blanket cycle time (flow rate) and the flow rate of the stream from the volatility plant to the separator can be varied. The respective flow rates are not independent but are related by the separator efficiency which is itself a function of the ratio of the two flow rates. To remove a given amount of materials from a blanket of given U-233 concentration a certain slurry flow rate is required at a given blanket liquid rate. If the slurry rate is reduced, blanket liquid rate must be increased. If slurry rate is increased, blanket liquid rate will be decreased.

The situation is further complicated by the fact that the liquid-solid ultimate separation can be done by at least four different methods. The only methods considered in detail here will be the ones thought to be most feasible:

- a. One stage continuous mixing and filtration
- b. Semicontinuous batchwise elution
- c. Semicontinuous draining of liquid from slurry mixture
- d. Multistage counter-current continuous mixing and filtration

Methods a, b, and c will be treated in the discussion and derivations presented below. Method d will not be considered in detail because the advantages it offers are of negligible value because:

- a. Very few stages can be utilized until limiting conditions are reached.
- b. From a process maintenance viewpoint, as few items of equipment as possible should be installed.

- c. With thermal cycling, high efficiencies are hoped to be attained in one stage alone, operating as in methods a, b, or c.

In the following discussion the equations to be developed apply only to equilibrium conditions at the separation temperature. If a thermal cycling step is included in the head-end separation process, then the separation efficiencies as defined here will not apply. This is because during thermal cycling, all or most of the solid particles will go into solution during the hot part of the cycle; upon cooling the particles will reform. The distribution coefficient will not apply immediately since the rate of approach to equilibrium will be dictated by the diffusion rates of the uranium and protactinium back into or onto the solid particles. Hence, if the separation step is performed rapidly after the reconstitution of the slurry in the cold leg of thermal cycling, then much more uranium and protactinium are in solution than would be dictated by the equilibrium distribution coefficients, and the resulting apparent separation efficiencies would be much higher. The limiting condition in this direction would be for none of the uranium or protactinium to go back into the particle, in which case the distribution coefficient is equal to zero. This case represents the highest theoretical removal attained by a given method, while the other case of assuming the original equilibrium  $k$  value delimits the lower bound of separation efficiency to be expected.

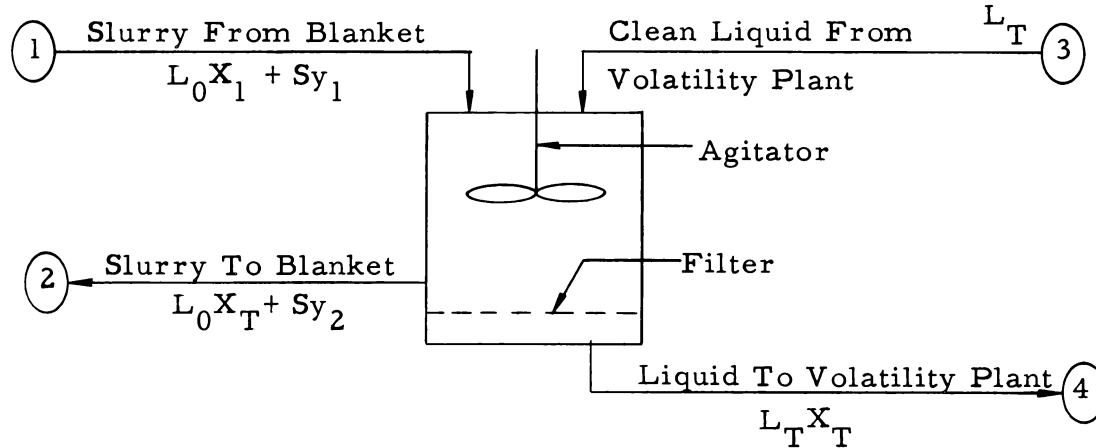
Therefore, the equations developed in the descriptions of the methods may be used to compute both the absolute upper and lower bounds to the separation efficiencies to be expected.

## 2. Description of Methods of Head-End Operation

### a. Method "a"

This method consists of a continuous filter. The blanket slurry inlet feed would be subjected to the proper heat treatments prior to introduction to the filter, thus alternating particle shape and size and establishing equilibrium conditions between the bred material in the solid and liquid.

The filter would operate as shown in the sketch below.



Let

- $L_0$  = lb/hr liquid bismuth from blanket,
- $L_T$  = lb/hr clean liquid bismuth from volatility plant,
- $S$  = lb/hr solid in slurry,
- $X$  = concentration of component in the liquid phase, lb/lb,
- $Y$  = concentration of component in the solid phase, lb/lb,
- $k$  = distribution coefficient,  $Y/X$ ,
- $Z$  = fraction of component removed, and
- $N$  = liquid ratio,  $(L_T/L_0)$ .

The amount entering in stream 1 is  $L_0 X_1 + S Y_1$ , which in turn is  $L_0 X + S k X_1$ , since  $Y_1 = k X_1$ .

Therefore, the inlet amount is  $L_0 X_1 (1 + \frac{S k}{L_0})$ .

Similarly, the amount returned to the blanket with stream 2, (where it is assumed that all solids are contained and that the original  $S/L_0$  ratio is maintained) is  $L_0 X_T (1 + \frac{S k}{L_0})$ .

$X_T$  is the average "stirred pot" liquid concentration existing in the separation, and the terminal condition.

Taking a material balance around the unit, and using the definition  $\alpha \equiv (1 + k S/L_0)$ ,

$$a L_0 X_1 = L_T X_T + a L_0 X_T = X_T (L_T + a L_0),$$

$$\text{or } X_T = \frac{X_1 (a L_0)}{(L_T + a L_0)} = \frac{X_1}{\frac{L_T}{a L_0} + 1} = \frac{X_1}{1 + N/a} \quad (1)$$

$Z \equiv$  fraction of component that is removed,

$$Z = \frac{L_0 X_1 + S Y_1 - (L_0 X_T + S Y_2)}{L_0 X_1 + S Y_1} = \frac{a L_0 X_1 - L_0 X_T}{a L_0 X_1},$$

$$\text{or } Z = 1 - \frac{X_T}{X_1} \quad (2)$$

Substituting equation (1) into equation (2),

$$Z = 1 - \frac{1}{1 + N/a} = \frac{N/a}{1 + N/a} \quad (3)$$

For a 10 w/o thorium slurry,  $S/L_0 = 1/3^*$ .

$$k_U = 6, k_{Pa} = 60,$$

therefore

$$a_U = 1 + k_U S/L = 1 + 6(1/3) = 3, \text{ and}$$

similarly,

$$a_{Pa} = 1 + 60(1/3) = 21.$$

Therefore, the working equations for method a are as follows for the case of equilibrium conditions, which represent the lower boundary to the separation efficiency expected:

$$Z_U = \frac{N/3}{1 + N/3}, \text{ and} \quad (4)$$

$$Z_{Pa} = \frac{N/21}{1 + N/21}, \quad (5)$$

where  $N$  is the liquid ratio used; i.e.,  $L_T/L_0$ .

The upper theoretical boundary to the separation efficiencies, both for uranium and protactinium, is the case  $k = 0$ , where all the possible constituent in the solid has been eluted. For this case,

$$a_{k=0} = 1,$$

and

$$Z_{Max} = \frac{N}{N+1}. \quad (6)$$

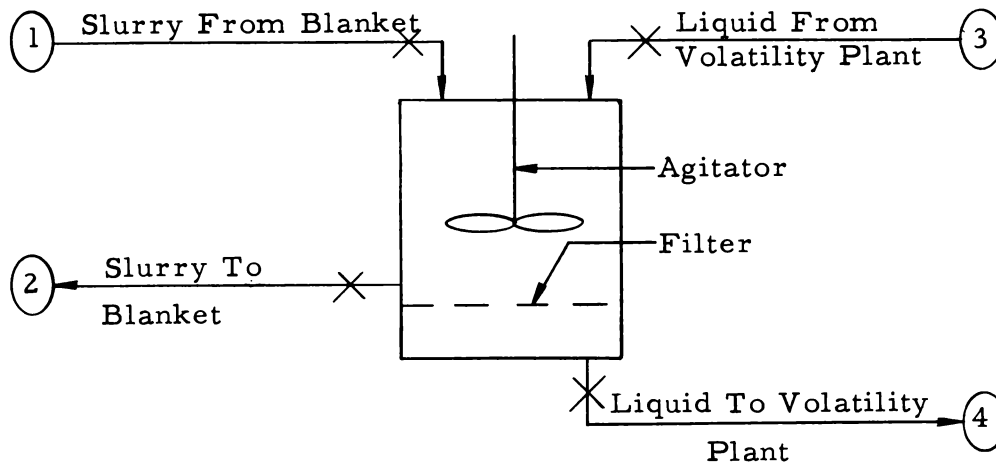
\*In a thorium bismuthide slurry, experimental evidence indicates that a 10 w/o thorium slurry consists of about 25% solids, so  $S/L = 25/75 = 1/3$ .



The uranium curves based on equations (4) and (6) are plotted as the solid lines on Figure 18 and represent the upper and lower possible boundaries of head end separation by method a. It should be noted that any process where the liquids and solids are continuously mixed and then separated will obey the maximum limitations imposed by equation (6), and the minimum will be represented by equation (4). Equation (5) for  $Z_{Pa}$  is plotted in Figure 19.

b. Method "b"

This method consists of a semicontinuous filter. A charge of slurry is brought to the filter from the blanket, and the flow is then shut off. Bismuth flow is started through the filter by pumping the stream from the volatility plant. After elution, the slurry at the original L/S ratio is returned to the blanket. The filter would operate as shown below.



Note: Streams 1 and 2 operate when streams 3 and 4 are off.  
Streams 3 and 4 operate when streams 1 and 2 are off.

Let  $L$  = lb/hr liquid, streams 3 and 4,  
 $L_0$  = initial lb of liquid on the filter,  
 $S_0$  = initial lb of solid on the filter,  
 $X$  = concentration of component in liquid,  
 $Y$  = concentration of component in solid, and  
 $k$  = distribution coefficient =  $Y/X$ .

FIG. 18: URANIUM SEPARATION EFFICIENCY

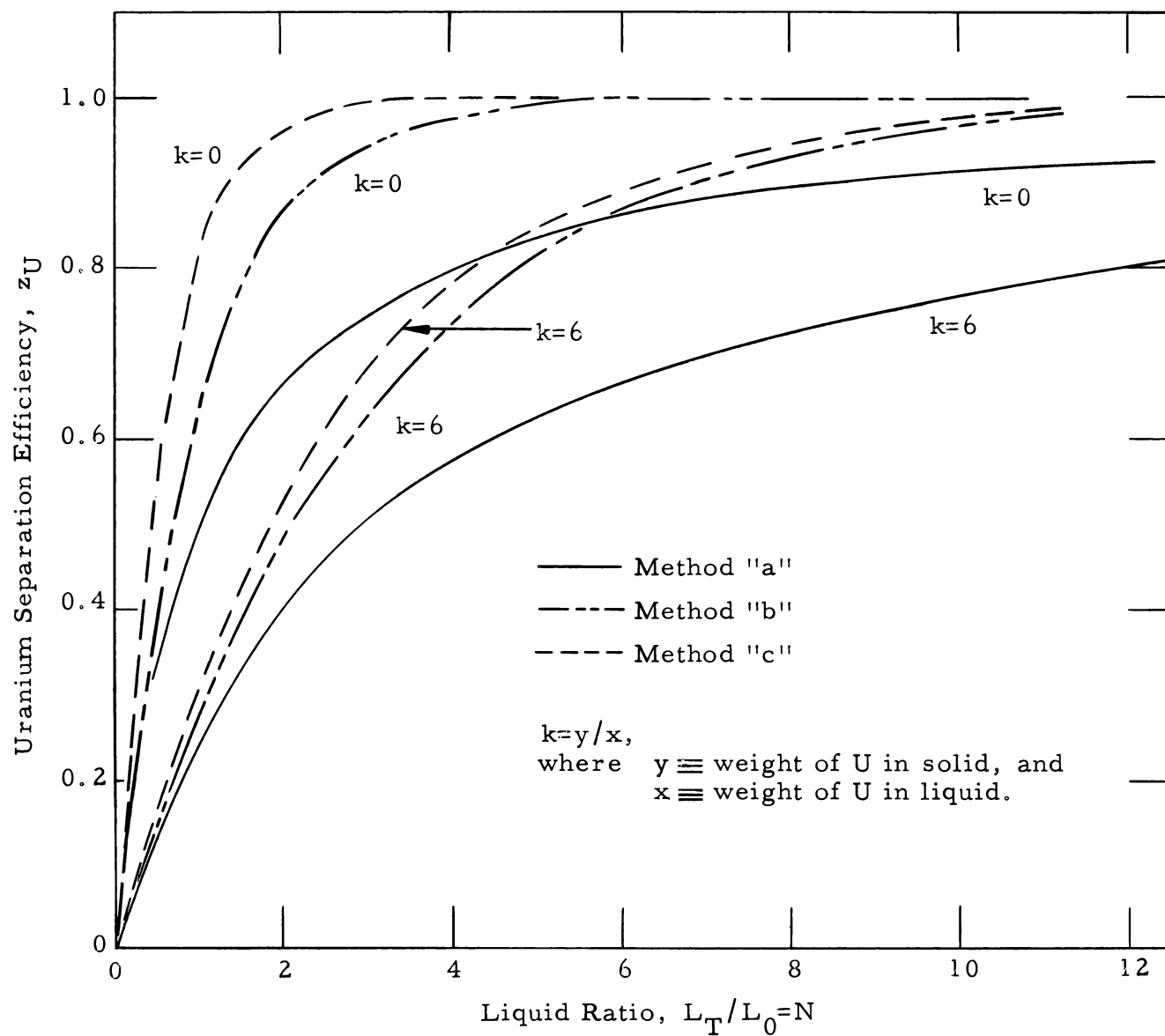
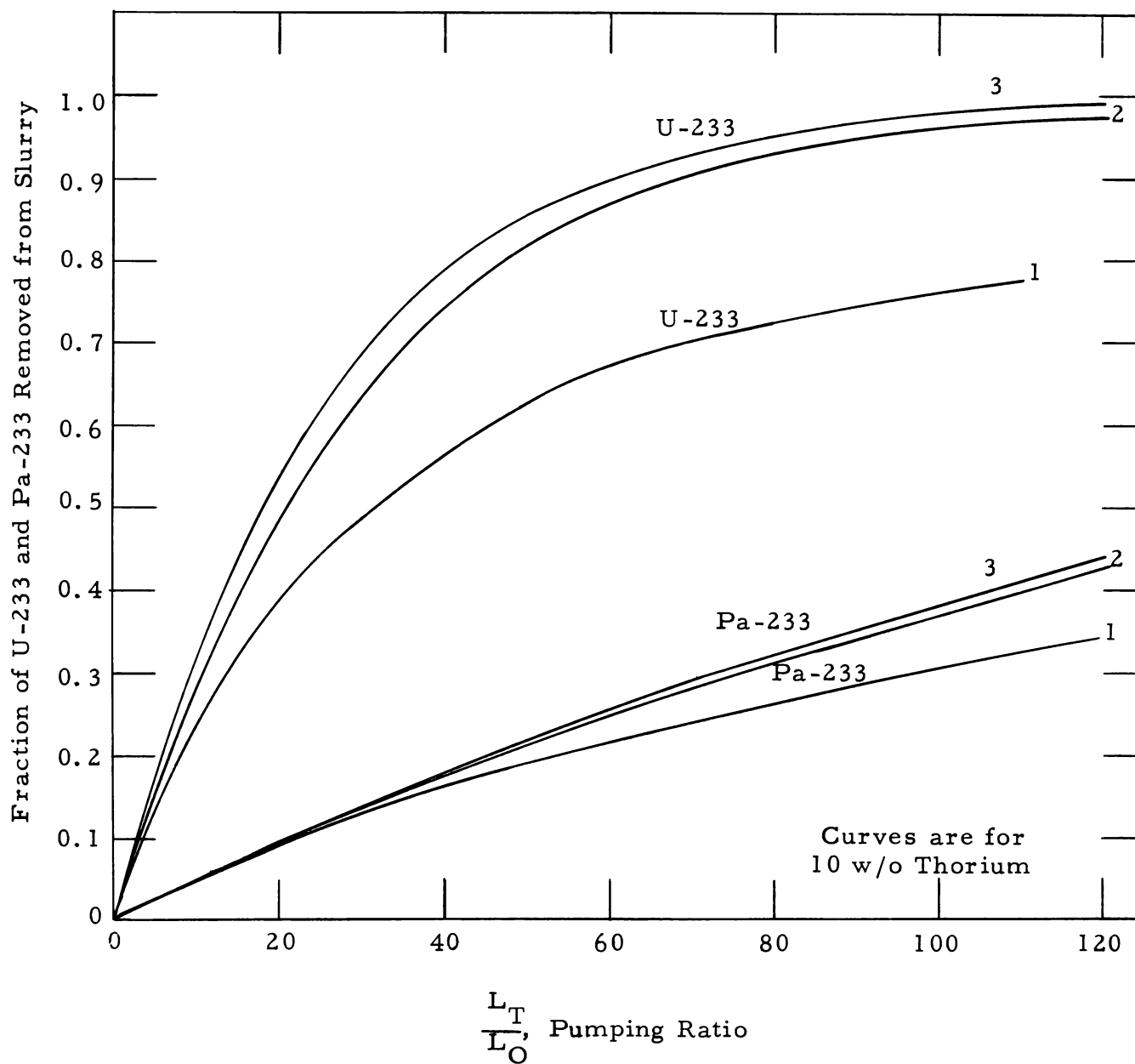


FIG. 19: EFFICIENCY OF U-233 and Pa-233 TRANSFER  
IN HEAD-END PROCESS



During the entire stripping operation the amount of liquid and solid on the filter is unchanged, then where T refers to end conditions,

$$Z = \text{fraction component removed} = \frac{L_0 X_0 + S_0 Y_0 - (L_0 X_T + S_0 Y_T)}{L_0 X_0 + S_0 Y_0},$$

$$\text{or } Z = 1 - \frac{X_T}{X_0}. \quad (2')$$

Let Q = total amount of component on the filter at any time,

$$\text{then } Q = L_0 X + S_0 Y = L_0 X \left(1 + \frac{S_0}{L_0} k\right) = a L_0 X, \quad (7)$$

where  $a \equiv 1 + Sk/L$ .

$$\frac{dQ}{dt} = LX$$

and since

$$\begin{aligned} \frac{dQ}{dt} &= a L \frac{dX}{dt}, \\ a L_0 \frac{dX}{dt} &= LX. \end{aligned} \quad (8)$$

The solution is

$$\frac{X}{X_0} = e^{-\frac{L}{L_0 a t}}. \quad (9)$$

If  $L_T$  is the total volume pumped in stream 3 and 4 during elution time t, i.e.,  $L_T = L_t$ , then

$$\frac{X_T}{X_0} = e^{-\frac{L_T}{L_0} \frac{1}{a}} = e^{-N/a} \quad (10)$$

and, by substituting equation (10) in equation (2),

$$\begin{aligned} 1 - Z &= e^{-N/a}, \\ \text{or } Z &= 1 - e^{-N/a}. \end{aligned} \quad (11)$$

Equation (11) may now be applied as was equation (3) for method a. By substituting  $a$ 's for  $k = 6$  (U-233,  $k = 60$  (Pa-233), and  $k = 0$  (maximum limiting case for complete elution of solids), the values of Z corresponding to those of equations (4), (5), and (6) may now be obtained. The upper and lower bounding values of Z for uranium for method b as a function of the liquid ratio are presented in Figure 18 by the dash and dot curves; Figure 19 gives the lower boundary values for  $Z_{Pa}$ .

### c. Method "c"

This is essentially a batch method, in which an original charge of slurry is brought into the separator. After thermal cycling, the liquid is drained from the agitated slurry mixture. After draining the "rich" liquid, fresh liquid is added to the remaining, solids-rich slurry, and the draining may be repeated or the resultant slurry of the original S/L ratio may be sent back to the blanket. Unfortunately, not all of the liquid can be withdrawn if the slurry is to be in reconstitutable form; BNL feels that a 50-50 solid-to-liquid ratio is the lower limit. This implies that for a 75% liquid, 25% solid entering slurry, no more than two-thirds of the liquid can be drained without adding fresh liquid to reconstitute the slurry.

Taking a material balance around the unit, as for methods a and b, the following working equation is derived for the case of successive drains of two-thirds of the liquid present in the separator at any time

$$Z = 1 - \left( \frac{a - 2/3}{a} \right)^{(3/2)N} \quad (12)$$

For one drain,  $N = 2/3$ , for two drains,  $N = 4/3$ , etc. The upper and lower expected boundaries for uranium, corresponding to  $a_k = 0$  and  $a_k = 6$  are depicted by the plain dashed lines in Figure 18.

The method selected for plant use will depend on many factors, though it is felt that method a is the easiest. It will yield a sufficiently high uranium separation efficiency which will still be under the maximum limiting conditions imposed by uranium solubility in the exit stream.

The equilibrium protactinium separation efficiencies are much lower than the uranium efficiencies for all liquid ratios,  $N$ ; for  $k = 0$  (complete extraction of the solid),  $Z_{Pa_{Max}} = Z_{U_{Max}}$ , shown in Figure 18.

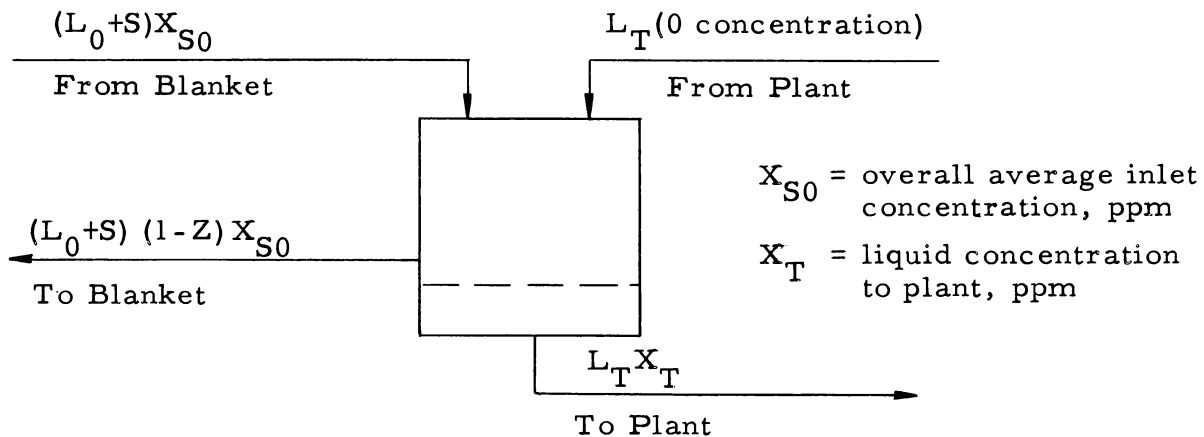
A complete comparison between the equilibrium (lower bound) uranium and protactinium extraction efficiencies as a function of  $N$  is shown in Figure 19.

### 3. Limiting Aspects of Head-End Separation Step

At first, it may appear desirable to attain as high a separation efficiency as possible by any and all means available. However, a limiting factor is reached since it is stipulated that the exit stream from the separator to the volatility plant shall be held below the solubility limit of uranium in bismuth at the separation temperature. If this criterion is to be met, a high separation efficiency, even if attainable, is unusable. (Naturally, by raising the separation temperature, higher concentrations can be tolerated in the exit stream and higher  $Z_U$ 's, possibly attainable by certain methods, can be utilized; however, higher temperature operation causes a much higher thorium loss since its solubility also increases, increasing operating costs.) Because only a lower  $Z_U$  may be required, it will be shown that method a (the continuous, one stage mixing and filtration, the easiest to effect) should deliver these permissible separation efficiencies, especially if the effect of thermal cycling will give close to the maximum theoretical  $Z_U$ ; i. e., for  $k = 0$ . Therefore, this method of head-end operation is suggested as being the most practical and advantageous.

#### a. Calculation of Permissible Separation Efficiencies

The permissible removal efficiencies are dictated by the solubility of uranium in the exit stream at the separation temperature. This permissible  $Z_U$  is a function of inlet slurry concentration; it applies to any continuous method of operation. (During the course of a semicontinuous or batch process, the instantaneous calculated exit liquid stream concentrations could exceed the solubility restrictions, even though at the end of processing a batch the final "average" concentration would be described by the following equations.



### Component Balance

$$(L_0 + S)X_{S0} = (L_0 + S)(1 - Z)X_{S0} + L_T X_T, \quad (13)$$

therefore

$$X_T = \frac{(L_0 + S) Z X_{S0}}{L_T},$$

$$\text{or } \frac{X_T}{X_{S0}} = \frac{(1 + S/L_0) Z}{N}, \quad (14)$$

where

$$N = \frac{L_T}{L_0}.$$

For a 25% slurry, (10% by weight thorium),

$$\frac{1}{(1 + S/L_0)} = 0.75.$$

Now  $X_{T_{\text{permissible}}} = f(T)$  as follows

T	400 C	350 C	300 C
$X_T$	1500 ppm	900 ppm	360 ppm.

(Concentrations given are approximately 80% of pure solubility. This accounts for solubility losses due to presence of Zr, Mg, fission products, etc.)

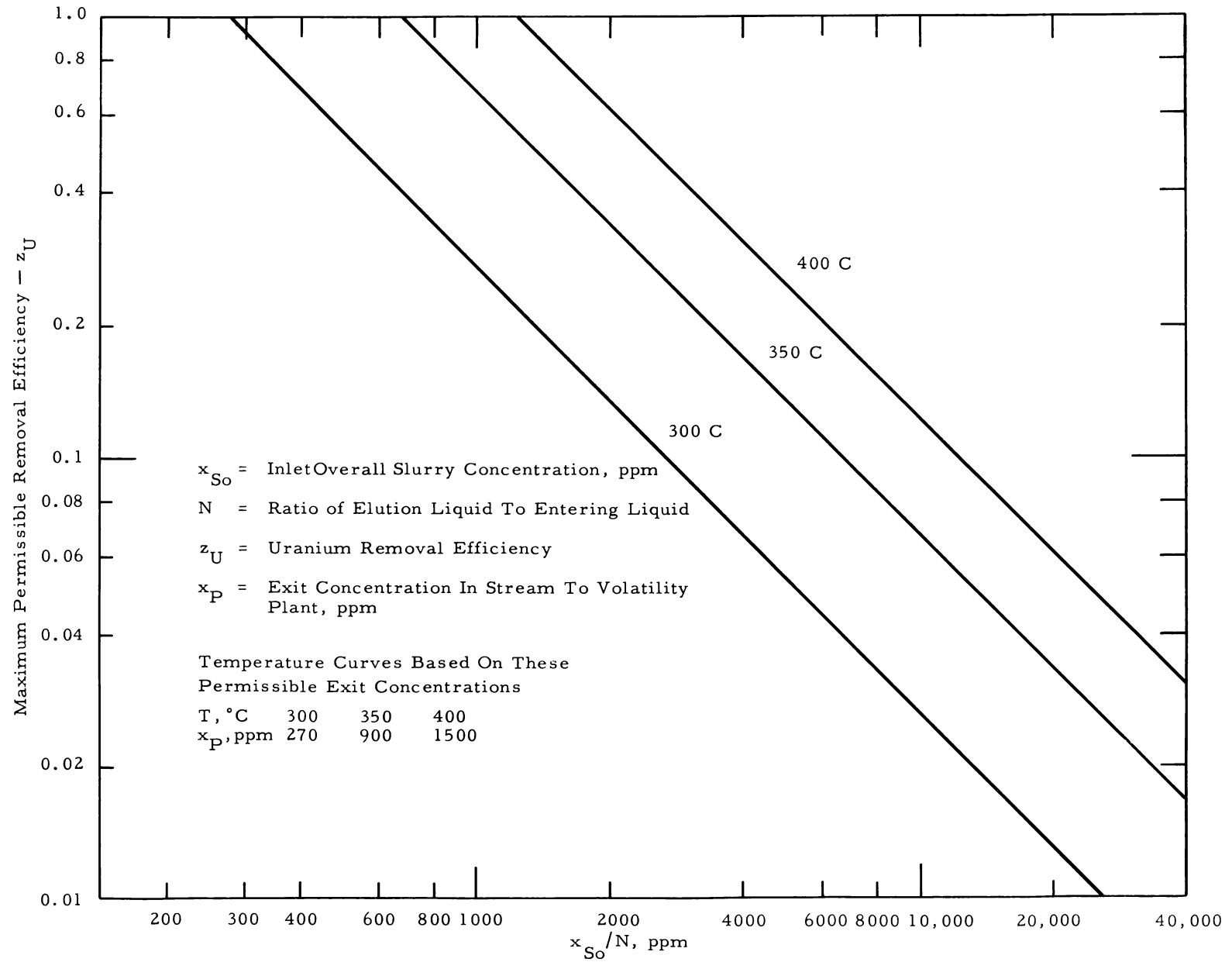
Rearranging equation (14) (only for a 25% slurry)

$$Z_{\text{permissible}} = \frac{0.75 X_T(T)}{(X_{S0}/N)}. \quad (15)$$

Equation (15) has been solved for the three temperature levels indicated, and the resulting plot of  $Z_{\text{permissible}}$  vs  $X_{S0}/N$  is shown in Figure 20.

Any Z attainable in a process which exceeds this  $Z_{\text{permissible}}$  cannot be utilized. Therefore, it is useless to try to improve the separation efficiency unless such an efficiency can be utilized with the expected entering slurry concentration.

FIG. 20: PERMISSIBLE REMOVAL EFFICIENCY





In order to show  $Z_{\text{permissible}}$  and  $Z_{\text{max}}$  attainable on one plot, equation (6), giving the maximum attainable  $Z_U$  for any continuous process, was rewritten by substituting  $X_S(N/X_S)$  for  $N$ .

Then

$$Z_{\text{max}} = \frac{X_S}{X_S + (X_S/N)} \quad (16)$$

The above equation is now  $Z = f(X_S, N)$  whereas the previous equation was a function of  $N$  alone. The additional variable  $X_S$  was introduced so that equations (15) and (16) could both be plotted on the same graph. (See Fig. 21.) Figure 21 shows both the theoretical and permissible limits of the separation efficiency. Given on inlet concentration of  $X_{S0}$  ppm, a liquid ratio  $N$ , and a temperature  $T$ , the maximum  $Z$  for a process may be found on Figure 21 since this  $Z_{\text{max}}$  must lie below the lowest of either of the appropriate curves; i. e., the solubility or the maximum feasible separation curves for that particular operating condition.

#### b. Discussion of the Limiting Aspects

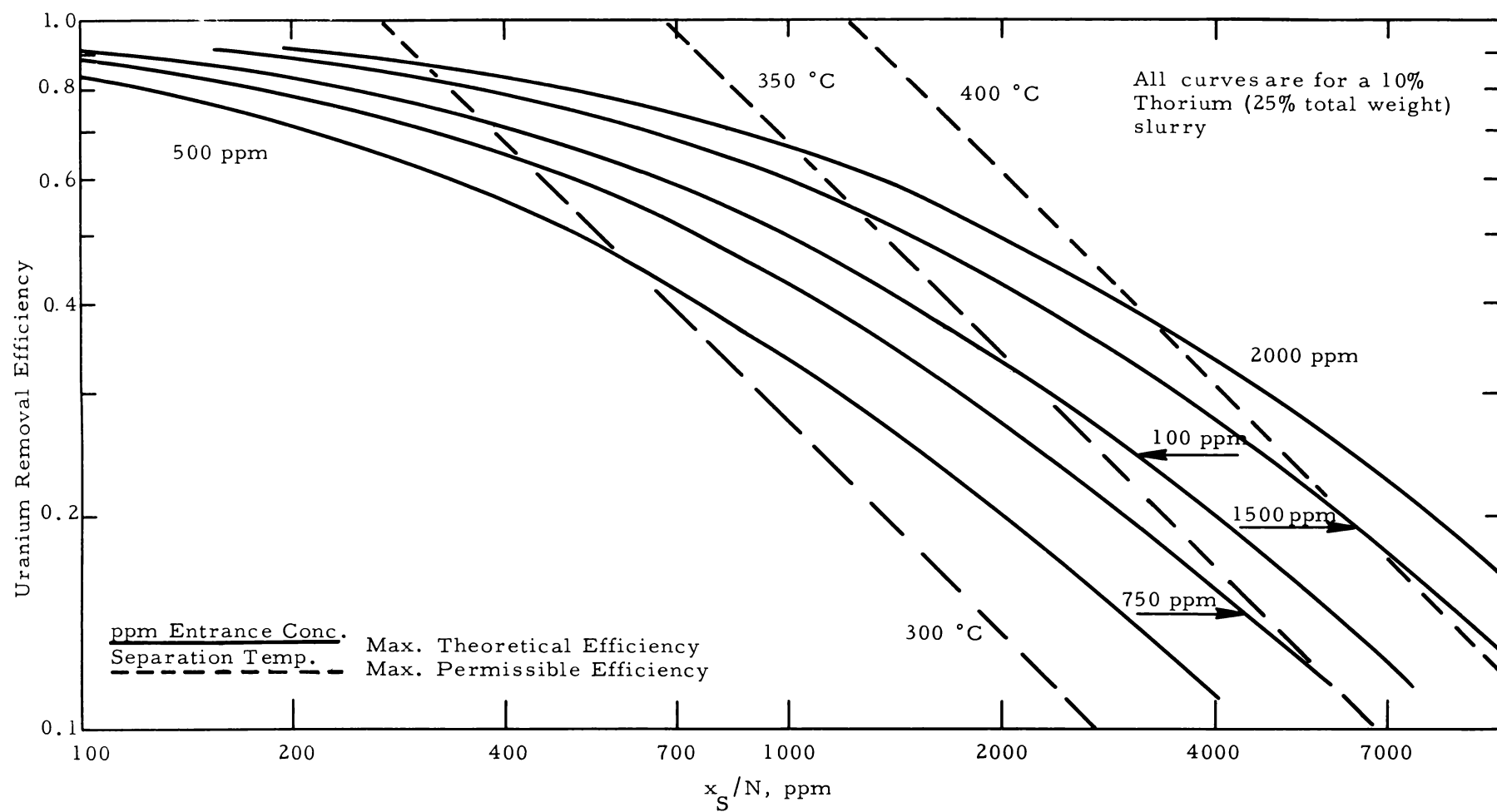
The results of the previous section, shown in Figures 20 and 21, indicate that the continuous method should yield sufficiently high separation efficiencies which can still be utilized with the given solubility restrictions. Therefore, this easier method of plant operation is recommended, unless other factors interfere.

To overcome the solubility restrictions, Figure 20 shows that raising the temperature would permit the higher  $Z_U$ 's which might be attainable. However, many factors influence the decision against raising the separation temperature above 350 C, the contemplated operating temperature. Some factors which would be adversely affected by an increase in separation temperature are as follows.

- (1) Increase of thorium loss due to the increased solubility of thorium in the exit stream to the volatility plant
- (2) Greater heating and insulation requirements
- (3) Possible requirements of better suited materials of construction

No economic balance was made where these undesirable temperature effects are balanced against the reduction in volatility plant size, due to the lower flow rates caused by the higher attainable and/or permissible separation efficiencies. However, independent of the results of such an economic study, intangible factors such as simplicity of operation and ease of maintenance will dictate the final choice of separation temperature and method.

FIG. 21: MAXIMUM THEORETICAL AND PERMISSIBLE  
REMOVAL EFFICIENCIES



## APPENDIX D

### CALCULATION OF CYCLE TIMES

#### 1. Introduction

Calculation of the equilibrium cycle times required to maintain a given blanket power-to-total power split is very important, since this will dictate the capacity of an on-site processing plant. For off-site processing, only the influence of the fission product poisoning on the conversion ratio (and therefore fuel costs) need be considered. This type of calculation should be done by a physicist.

Equilibrium processing cycles to maintain a given power split and the consequent amount of liquid to be processed per day in a two-region reactor can be resolved by either the "material balance" or the "nuclear" method. The method used depends on the information most readily available. Results calculated by the two methods should be identical.

In the "material balance" method, the amount of potential uranium (U and Pa) formed (in excess of the amount consumed by producing power in the blanket) must be removed to maintain a constant uranium concentration in the blanket, consistent with the desired blanket power level and flux. In the "nuclear" method, only the nuclear parameters of the system are used to predict the cycle times required to maintain a given U/Th ratio. In the "nuclear" method, the percent power split, conversion ratio, and power level, (all explicit in the "material balance" method) are now contained implicitly in the nuclear parameters such as effective cross sections, thermal fluxes, and uranium concentrations.

Derivations of the equations for these two methods follow. To use the equations, basic parameters must be obtained from nuclear calculations giving the effective cross sections, average system fluxes, conversion ratios, and concentrations of uranium and thorium as a function of poison level. When these are known, either method may be used.

## 2. Derivation of Methods for Obtaining Equilibrium Cycle Times

In the derivations below, the following nomenclature is used.

- $\sigma_a^i$  - effective absorption cross section for isotope i.
- $\phi_T$  - average thermal system flux of blanket
- $\lambda$  - decay constant of protactinium
- $\nu$  - neutrons per fission in U-233
- $\eta$  - neutrons produced per neutron absorbed in fissile material  
 $\frac{\nu \sigma_f}{\sigma_a}$  (Note -  $[\nu/\eta] \cong 1.12$  for all cases considered)
- $Z_U$  - fraction of uranium that is removed per pass in separator
- $Z_{Pa}$  - fraction of protactinium that is removed per pass in separator
- $N_U$  - atoms uranium/cc
- $N_{Th}$  - atoms thorium/cc
- $U$  - uranium inventory of loop, gm
- $Pa$  - protactinium inventory of loop, gm
- $T$  - equilibrium cycle time in days

### a. Calculation of Cycle Time by Material Balance Method

Gm fissioned for 825 MW are

$$\left( \frac{24 \text{ hr} \times 825 \times 10^3 \text{ kwh}}{\text{day}} \right) \left( \frac{\text{gm mole}}{5.3 \times 10^6 \text{ kwh}} \right) \left( \frac{233 \text{ gm}}{\text{gm mole}} \right) = \frac{870 \text{ gm}}{\text{day}} .$$

$$\frac{\text{Amt of U-233 consumed}}{\text{day}} = \frac{870 \text{ gm fissioned}}{\text{day}} \left( \frac{\text{gm consumed}}{\text{gm fissioned}} \right)$$

Now  $\left( \frac{\text{gm consumed}}{\text{gm fissioned}} \right) = \frac{\sigma_a}{\sigma_f} ,$

therefore

$$\frac{\text{Amt of U-233 consumed}}{\text{day}} = 870 \frac{\sigma_a}{\sigma_f} .$$

Since

$$\eta = \frac{\sigma_f}{\sigma_a} , \text{ then } \frac{\sigma_f}{\sigma_a} = \frac{\eta}{\nu} ,$$

so that

$$\text{Amt of U-233 consumed/day} = 870 (\nu/\eta) .$$

The definition of the conversion ratio is

$$CR \equiv \frac{\text{atoms of specie formed}}{\text{atom of specie consumed}},$$

(Note: Since we are dealing with a single specie, gm may be substituted for atoms)

hence

$$\frac{\text{amt of U-233 produced}}{\text{day}} = \left( \frac{\text{gm fissioned}}{\text{day}} \right) \left( \frac{\text{gm consumed}}{\text{gm fissioned}} \right)$$

$$\left( \frac{\text{gm formed}}{\text{gm consumed}} \right) = 870 \frac{\nu}{\eta} CR$$

The above represents the total amount formed daily in the blanket.

However, not all of the above is "excess" which needs to be removed to maintain a constant concentration, since some is burned up directly in the blanket.

$$\frac{\text{Amt consumed in blanket}}{\text{day}} = \frac{870 \text{ gm fissioned}}{\text{day}} \left( \frac{\text{Fraction of power}}{\text{produced in blanket}} \right)^x$$

$$\left( \frac{\text{gm absorbed}}{\text{gm fissioned}} \right) = 870 \left( \frac{\text{Blanket power}}{\text{Total power}} \right) \frac{\nu}{\eta}$$

Hence, the excess daily uranium production,  $p_U$ , which has to be removed to maintain a constant concentration in the blanket is

$$p_U = \frac{\text{Amt U-233 produced}}{\text{day}} - \frac{\text{Amt consumed in blanket}}{\text{day}}$$

$$= 870 \frac{\nu}{\eta} CR - 870 \frac{BP}{TP} \frac{\nu}{\eta},$$

or

$$p_U = 870 (\nu/\eta) [CR - (BP/TP)]. \quad (17)$$

Equation (17) assumes that  $\frac{\nu}{\eta} \text{ blanket} = \frac{\nu}{\eta} \text{ core} = \frac{\nu}{\eta} \text{ average}$ .

This condition is met for the required calculational accuracy.

The amount of potential uranium (U + Pa) removed per day is

$$q_U = \frac{Z_U U + Z_{Pa} Pa}{T} = \frac{Z_U U [1 + (Z_{Pa}/Z_U) (Pa/U)]}{T},$$

or

$$q_U = \frac{U [1 + (Z_{Pa}/Z_U) (Pa/U)]}{(T/Z_U)}. \quad (18)$$

At equilibrium conditions

$$q_U = P_U,$$

or

$$\frac{U [1 + (Z_{Pa}/Z_U) (Pa/U)]}{(T/Z_U)} = 870 (\nu/\eta) [CR - (BP/TP)], \quad (19)$$

which may conveniently be solved for  $(T/Z_U)$ .

(It should be realized, however, that the ratio  $(Pa/U)$  is also a function of  $(T/Z_U)$  as well as flux level, and this relationship is given by equation (5) of the next subsection.)

b. Calculation of Cycle Times by "Nuclear" Method

$$\frac{dN_U}{dt} = \lambda N_{Pa} - N_U \sigma_a U \bar{\phi} - \frac{Z_U}{T} N_U \quad (20)$$

which at steady state is

$$\frac{N_U}{N_{Pa}} = \frac{\lambda}{\sigma_a U \bar{\phi} + \frac{Z_U}{T}} \quad (21)$$

A balance for Pa at steady state leads to

$$\frac{N_{Pa}}{N_{Th}} = \frac{\sigma_a^{Th} \bar{\phi}}{\lambda + \frac{Z_{Pa}}{T} + \sigma_a^{Pa} \bar{\phi}} = \frac{\sigma_a^{Th} \bar{\phi}}{\lambda + \frac{Z_{Pa}}{Z_U} \frac{Z_U}{T} + \sigma_a^{Pa} \bar{\phi}} \quad (22)$$

In the above,  $\sigma_a^{Th}$  and  $\sigma_a^{Pa}$  must be effective thermal cross sections in which resonance captures have already been accounted for.

Combining the above two equations,

$$\frac{N_U}{N_{Th}} = \frac{\lambda \sigma_a^{Th} \bar{\phi}}{\left[ \lambda + \sigma_a^{Pa} \bar{\phi} \left( \frac{Z_{Pa}}{Z_U} \right) \left( \frac{Z}{T} \right) \right] (\sigma_a^{Th} U \bar{\phi} + \frac{Z_U}{T})} = \text{constant}. \quad (23)$$

If the above is set equal to the desired  $N_U/N_{Th}$  ratio, i.e., the uranium concentration, we have an implicit relationship for  $T$  which may be solved to yield  $T_{\text{equil}}$ . This should be the same figure as that obtained by the "material balance" method.

### 3. Discussion

Both equations (19) and (23) may be solved for  $T/Z$ , the "adjusted" equilibrium cycle time required to maintain a given power split. The ratio of the protactinium to the uranium removal efficiencies was retained as a parameter in the studies, because previous experimental data indicated that different removal efficiencies are to be expected in the processing of a bismuthide slurry. For processing of an oxide slurry, it may be assumed that  $Z_U = Z_{Pa} = (Z_{Pa}/Z_U) = 1$ , for which case  $T/Z_U$  also equals  $T$ .

The higher isotope concentrations in the blanket need not be explicitly considered as part of the load on the processing plant as far as preliminary studies are concerned, since these higher isotope concentrations at steady state are only about 0.1% of the U-233 concentration.

If  $W$  is the total weight of slurry in the whole blanket loop, then  $L_0$ , the liquid flow rate from the reactor is  $3/4 W/T_{\text{equil}}$  lb/day for a 25% thorium bismuthide slurry.  $L_T$ , the flow rate to the volatility plant is  $NL_0$ , where  $N$  is amount of elution liquid volume per batch of main slurry to stream liquid volume. Therefore

$$L_T = NL = 3/4 NW/T_{\text{equil}} \text{ lb/day.}$$

Similarly, if uranium is defined as the total uranium loop inventory in the blanket, then the amount of potential uranium to be processed per day is:

$$\begin{aligned} \frac{\text{Amt of Pa + U}}{\text{day}} &= \frac{U Z_U}{T} \left[ 1 + \frac{Z_{Pa}}{Z_U} \left( \frac{Pa}{U} \right) \right] \\ &= 4/3 Z_U L_0 a \left[ 1 + \frac{Z_{Pa}}{Z_U} \left( \frac{Pa}{U} \right) \right] = c', \end{aligned}$$

where

$a$  = weight fraction U-233 in slurry.

$C'$  is the amount of potential uranium removed per day; this determines part of the operating costs of the plant.



## APPENDIX E

### CALCULATION OF PROCESSING COSTS

#### 1. Introduction

At present, probable LMFR fuels and blankets are (1) U-Bi solution fuel, (2) thorium-bismuthide slurry blanket, (3) uranium oxide dispersed in bismuth fuel, (4) thorium oxide dispersed in bismuth blanket, and (5) mixed thorium oxide and uranium oxide dispersed in bismuth for a single-region reactor fuel. Aqueous processing could be applied to any or all of these types of fuels; fluoride volatility processing could be applied only to (1) and (2) above.

Aqueous processing plants can be located on or away from the reactor site. If the plant is located away from the site, it is still necessary to locate a phase separation plant at the reactor site, and therefore a certain amount of capital investment and operating expenditures are required.

#### 2. Processing Costs for Reactor Economics

##### a. Fluoride Volatility (Processing of Bismuth Solution Fuels and Bismuthide Slurries)

The fluoride volatility method of processing has its main use in processing fuel from the two-region reactor. The processing method is convenient for transferring bred uranium from the blanket of a two-region reactor to the core. To estimate these processing costs the following procedure is used:

(1) Compute the blanket cycle time ( $T_{\text{days}}$ ) by one of the methods in Appendix D. Conservative values of  $Z_U$  and  $Z_{Pa}$  are 0.25 and 0.04, respectively. The rest of the parameters required must come from physics considerations for the particular computer.

(2) From the cycle time and blanket system inventory the slurry volume processed per day is obtained. To obtain the liquid volume in the slurry, multiply slurry volume by  $(1-2.5X)$ , where  $X$  is the weight-fraction thorium in the slurry. Call this processing rate  $L_T \text{ ft}^3/\text{day}$ .

(3) Compute the core cycle times from poison considerations. Obtain the volume of core fluid processed per day. Call this  $C_T \text{ ft}^3/\text{day}$ .

(4) Obtain the total annual cost from Figure 7 at the processing rate  $(C_T + L_T)$ .

(5) Certain errors are inherent in the use of the simplified cost curve Figure 7, and certain costs not included are discussed below.

#### Capital Cost

- (a) Equipment and building scale-up factors described in Appendix B have been applied.
- (b) NFPN processing rate is set at  $1/10$  the FPS + FPN rate.
- (c) Head-end processing is included. An error will be introduced if these curves are applied to a situation involving high core fluid processing rates as compared to slurry rate. About 15.5% of the direct construction cost is in the head-end plant, and this has been scaled as the total plant flow, including both core and blanket fluid. The maximum error in the total annual charges is therefore 8.2% for the extreme case of core fluid processing only.
- (d) A factor of 2.17 is applied to the direct construction investment for process equipment to obtain the total capital investment. This factor includes construction overhead fees, engineering, contingency, and preoperation and startup.
- (e) A factor of 1.73 is applied to the direct construction investment for building to obtain this total capital investment. This factor includes construction overhead fees, engineering, and contingency.

- (f) Curves can be applied for cases of any percent thorium, where liquid flow has been computed.
- (g) The equipment amortization period is 7 yr. The depreciation rate is 24.5%/yr on the capital investment. (See Fig. 8.)
- (h) To account for differing plant and reactor utilization factors, see below.

#### Operating Cost

(a) Curve is for blanket liquid containing 1000 ppm U-233 + Pa-233. At 20 cu ft/day the maximum error is \$29,200/292-day operating year. This amounts to + 1.7% maximum error in the total annual charges.

(b) No figure is included in the curve for Pa-233 or U-233 inventory charges. An approximation would be 5-day inventory for U-233, and 135-day inventory for Pa-233. These inventories would be based on the processing rate being considered.

Pa inventory = lb of blanket slurry processed per day x Pa concentration x efficiency of removal in head end plant, (0.04) x 135

U inventory = [ lb of blanket slurry processed per day x U concentration x efficiency of removal in head end plant (0.25) + lb of core fluid processed per day x U concentration ] x 5

(c) Curve is based on labor charges for 365 days/yr. The chemicals and waste disposal charges are for 292 full power days of operation per year.

#### b. Method of Obtaining On-Site Processing Cost for Aqueous Processing of an Oxide Slurry

The on-site costs calculated by the following method are for amortization as used by utilities.

To calculate the cost of aqueous chemical processing for an on-site, two-region reactor, the following is done.

(1) Compute the equilibrium cycle time, (T), based on nuclear parameters furnished by the reactor physicists. For aqueous processing,  $Z_U = Z_{Pa} = (Z_U/Z_{Pa}) = 1$ .

(2) From the known thorium inventory of the system, compute the daily quantity of thorium to be processed. Let this quantity be  $p_T$ , so that

$$p_T = \frac{\text{kg Th inventory}}{T_{\text{equil}}} .$$

The curve in Figure 6 (or the middle curve of Fig. 5) should be entered at a quantity  $P_T$  kg thorium per day where this  $P_T$  has been adjusted for reactor and plant utilization factors as follows

$$P_T = p_T \frac{f_r}{f_p} ,$$

where

$f_r$  = reactor utilization factor, and

$f_p$  = plant utilization factor.

This is required to adjust for the possibility of differing plant and reactor plant utilizations, and to bring  $P_T$  in line with the cost curves drawn for an 80% plant utilization factor. For the presently contemplated reactor,  $f_r = f_p = 80\%$  so that

$$P_T = p_T .$$

(3) Having calculated  $P_T$ , the annual processing cost is read from Figure 6 (or the middle curve of Fig. 5).

#### c. Method of Obtaining Cost Per Batch for Off-Site Processing

Off-site processing costs are a function of the cost per charge of off-site processing in the central plant, and the on-site head-end processing and storage costs.

To obtain the cost of off-site processing the following is done.

(1) An estimate of the cycle time is obtained from physicists and/or nuclear reactor engineers. This cycle time is determined by balancing the cost of processing against the increased fuel costs caused by a decrease in conversion ratio, which in turn is caused by an increase in fission product poisoning. Since the cycle time is a function of the costs, solution of cycle times and costs must be done simultaneously. Therefore, this computation of cost per charge will be for an arbitrarily assumed cycle time.

(2) Having assumed a cycle time, the cost per batch of off-site processing is obtained from the upper curve of Figure 5 if a central plant capacity has been decided on. For an assumed 200-kg/day capacity central plant, the unit cost is about \$86.00/kg Th. Therefore the cost per charge is \$/charge = 86 (kg/charge).

(3) Use of on-site facilities will be charged to each batch of fuel sent to off-site processing. When the annual capital return cost for on-site processing has been determined,

$$\frac{\text{On-site capital cost}}{\text{charge}} = \frac{\text{Annual capital}}{\text{return cost}} \times \frac{\text{Cycle time}}{\text{in years}}.$$

(4) An operating cost must be estimated for the on-site plant.

(5) Transportation costs per charge to the central plant must be found.

(6) Inventory charges while processing, and loss in value when selling must be computed.

(7) The annual cost of off-site processing is then

$$\text{Annual cost of off-site processing} = \frac{\text{Items (2)+(3)+(4)+(5)+(6)}}{\text{Cycle time in years}}.$$







