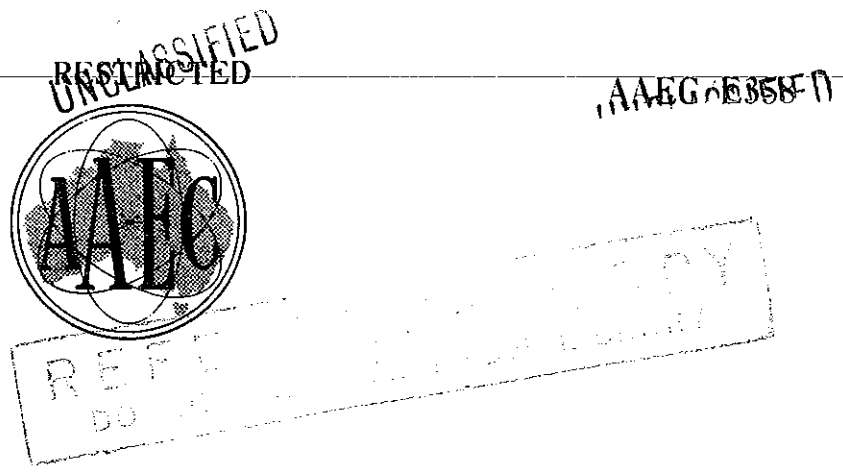


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**AUSTRALIAN ATOMIC ENERGY COMMISSION**  
**RESEARCH ESTABLISHMENT**  
**LUCAS HEIGHTS**

**COMPARISON OF COSTS OF CONVENTIONAL FLUORINATION  
AND IMPROVED FLUOROX PROCESSES FOR THE PRODUCTION  
OF URANIUM HEXAFLUORIDE**

by

**B. G. CHARLTON**

May 1975

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COMPARISON OF COSTS OF CONVENTIONAL FLUORINATION AND IMPROVED  
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ABSTRACT

Capital and operating costs for direct fluorination and two schemes of improved Fluorox plants are compared for a uranium throughput of  $3,000 \text{ Mg y}^{-1}$ . Scheme I involves the recycle of uranyl fluoride ( $\text{UO}_2\text{F}_2$ ) through separate reduction and hydrofluorination steps, while in Scheme II the reduction and hydrofluorination of recycle  $\text{UO}_2\text{F}_2$  is carried out in one step. There are savings in total capital requirement of approximately \$982,000 for the Scheme I Fluorox plant and approximately \$2,156,000 for the Scheme II Fluorox plant compared with a total capital requirement of approximately  $\$14.6 \times 10^6$  for the conversion of yellow cake to uranium hexafluoride via direct fluorination. Annual expenditures (including capital charges) are reduced by \$480,000 and

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\$848,000 for the two Fluorox plants respectively, saving \$0.16 kg<sup>-1</sup> and \$0.28 kg<sup>-1</sup> respectively in the cost of conversion of yellow cake to uranium hexafluoride compared with \$2.43 kg<sup>-1</sup> reported for a direct fluorination plant.

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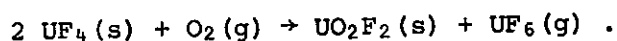
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## 1. INTRODUCTION

The method of conversion of uranium tetrafluoride ( $UF_4$ ) to uranium hexafluoride ( $UF_6$ ) in production plants throughout the world is the contacting of powdered  $UF_4$  with elemental fluorine in a fluidised bed or flame reactor. Elemental fluorine is expensive as it is produced by the electrolysis of anhydrous hydrofluoric acid, a process which uses electricity inefficiently and requires a relatively large building to house the numerous electrolytic cells. Since fluorine is very corrosive and toxic, fluorine plants require stringent ventilation and air cleaning systems.

The Fluorox process which uses oxygen to perform the conversion is described by the equation:



The uranyl fluoride ( $UO_2F_2$ ) is recycled for reduction to uranium dioxide ( $UO_2$ ) and reconversion to  $UF_4$ . This reaction is attractive since it avoids the use of elemental fluorine and it has been examined on a pilot plant scale in the US [Scott et al. 1960] and South Africa [Geertsma et al. 1965]. The reaction requires a temperature of 800 to 850°C for a suitable reaction rate. Since at this temperature the corrosion of the reactor is prohibitively fast, this process was not developed beyond the pilot plant stage.

Recently, Batley et al. [1974] described an improved Fluorox process in which the oxidation is catalysed with platinum on alumina resulting in a reaction rate at 600°C comparable to the uncatalysed rate at 800°C. Corrosion rates should therefore be considerably reduced. Batley et al. also reported development of the process in a small scale (42 mm diameter) fluidised bed reactor with promising results.

Production cost of uranium hexafluoride by the conventional fluorination route was compared with that for the improved Fluorox process at a uranium throughput of 3,000 Mg  $y^{-1}$ . Costello's [1972, 1974] estimated capital and operating costs for production of uranium hexafluoride in Australia by the conventional fluorination route, have been used as a basis for this comparison. All costs are on the basis of 1972 Australian prices.

## 2. DESCRIPTION OF PROCESSES

Conversion of yellow cake to  $UF_6$  by the *direct fluorination route* (DF) [Costello 1972] involves dissolution in nitric acid, solvent extraction with tributyl phosphate in kerosene, concentration of uranyl nitrate, fluidised bed thermal denitration to uranium trioxide ( $UO_3$ ), reduction with hydrogen to  $UO_2$ , hydrofluorination to  $UF_4$ , and fluorination to  $UF_6$ . This report

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examines only the conversion of  $UO_3$  to  $UF_6$  (Figure 1).

Two improved Fluorox plants are considered, Schemes I and II. In Scheme I (Figure 2)  $UO_2F_2$ , together with some  $UF_4$  and catalyst, from the oxidation reactor is recycled through a complete train of reduction and hydrofluorination reactors separate from that used to provide the primary feed of  $UF_4$ . In Scheme II (Figure 3) reduction and hydrofluorination of the recycled  $UO_2F_2$  stream takes place in one step in a separate reaction system and the hydrogen fluoride produced in the reduction reaction is consumed in the hydrofluorination reaction along with some fresh material. In both schemes, a portion of the recycled  $UO_2F_2$  stream is removed for processing to clean the catalyst.

### 3. ESTIMATION OF CAPITAL COSTS

The steps of reduction hydrofluorination, fluorination, hydrofluoric acid recovery and waste recovery and disposal in the data of Costello [1972] were dissected into costs for individual process streams and plant items (Tables 1 to 3) using the equipment flow diagrams (Figures 4 and 5) as described in the Appendix. An independent cost estimate was made and anomalies considered before the differential cost exercise was carried out. These anomalies were within 6 per cent of the total costs. These cost data were then used to estimate the capital cost of plants using the improved Fluorox process at a uranium throughput of  $3,000 \text{ Mg y}^{-1}$ . The equipment flow diagrams for the two schemes of Fluorox plant considered are shown in Figures 6 and 7. The flowsheet for catalyst cleaning, common to both schemes, is shown in Figure 8.

### 4. ESTIMATION OF OPERATING COSTS AND ANNUAL EXPENDITURES

Operating costs of the Fluorox processes were assessed using the same bases as those used by Costello in 1972 and incorporated his 1974 revised estimate of electrical power costs. The principal costs considered were process chemicals and labour based on the materials flowsheets in Figures 2 and 3. The overall capital and operating costs were then used to derive the annual expenditures for the plant using the same bases as Costello [1972]. Details of these calculations and the bases used are given in the Appendix.

### 5. RESULTS

Comparison of the estimated capital costs indicates that the installed capital cost saving of Scheme I over DF plant is \$459,000 (Table 7), and that Scheme II shows a further installed capital cost saving of \$662,000 over DF plant. The capital cost of Scheme II is approximately 75 per cent of that of the DF plant for conversion of  $UO_3$  to  $UF_6$ .

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The cost of process-feed materials is greater for both Fluorox schemes than for DF, though only by 0.79 cents  $\text{kg}^{-1}$  for Scheme II (Table 8). There is a reduction in the costs of services of approximately 20 per cent for both schemes with a slight advantage to Scheme II. There is also a reduction in annual labour costs of approximately 13 per cent, mainly from elimination of fluorine generation, the most labour intensive section of the plant, and its replacement by a duplicate line of gas-solid reactors which is considerably less labour intensive.

When capital cost, starting costs and operating capital are taken into account (Table 12), there is a saving in total capital of \$982,000 for Fluorox Scheme I and \$2,156,000 for Fluorox Scheme II compared with DF, while annual expenditures (Table 13) are reduced by \$480,000 and \$848,000 respectively. These reductions represent savings of \$0.16  $\text{kg}^{-1}$  and \$0.28  $\text{kg}^{-1}$  respectively in the conversion cost of \$2.43  $\text{kg}^{-1}$  reported by Costello [1972, 1974] for a DF plant with an uranium throughput of 3,000  $\text{Mg y}^{-1}$ .

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APPENDIX A

A1. INTRODUCTION

This appendix first outlines the bases of cost data derived from the information presented by Costello [1972, 1974] and presents the process flowsheets and corresponding estimated costs for a DF plant and two schemes for improved Fluorox plants. Second, the appendix outlines the bases for estimates of operating costs and presents materials flowsheets for these plants. These estimates are then combined to obtain comparisons of the total annual expenditure for the three plants. All costs are based on 1972 Australian dollars.

A2. BREAKDOWN OF DIRECT FLUORINATION PLANT COSTS

Costello [1972] referred to the use of fluidised beds for all processes used to convert concentrated uranyl nitrate to uranium hexafluoride, but costs were derived from Paynter & Geertsma [1965] who assumed a moving bed technique for the combined reduction and hydrofluorination reactions; the difference was not considered in this study. Costello also gave a flow diagram of a fluorination plant with compressors in the line between fluorinator and primary UF<sub>6</sub> collection traps; this is not usual practice and in this study it was assumed that circulation compressors are used to return unused fluorine to the process after the cold traps. 'Fluorine Recovery' in his report should have read 'HF Recovery'.

Costello's flow diagrams (Figures 4 and 5) were broken down into sections and the cost of each individual plant item estimated mainly from the information given by Buchanan & Sinclair [1964], Peters & Timmerhaus [1968], Guthrie [1969], and The Institution of Chemical Engineers [1969]. These costs, updated as necessary, included installation, piping, instrumentation and lagging, which were estimated by the modular approach described by Guthrie [1969] and The Institution of Chemical Engineers [1969]. The resultant total costs were within 6 per cent of Costello's and individual plant item costs were adjusted to make the totals equal to his totals. The cost breakdowns obtained are shown in Tables 1 to 4.

A2.1 Basis for Costs

A2.1.1 Reduction

In reduction reactors of the standard USAEC design (throughput 2,900 tonnes (U) per year), Smiley [1961] showed that the time for 90 per cent

conversion of fluid-bed denitrated  $\text{UO}_3$  was three times that for pot-denitrated  $\text{UO}_3$ . However, fluid-bed  $\text{UO}_3$  is approximately 60 per cent denser than pot-denitrated  $\text{UO}_3$  so a bed volume 1.8 times that of the USAEC design was used. The comparative throughputs for the USAEC design are  $35.8 \text{ kg h}^{-1}\text{m}^{-2}$  ( $850 \text{ lb h}^{-2} \text{ft}^{-2}$ ) or  $2.57 \text{ kg h}^{-1}\text{m}^{-3}$  ( $200 \text{ lb h}^{-1}\text{ft}^{-3}$ ) against  $19.0 \text{ kg h}^{-1}\text{m}^{-2}$  ( $450 \text{ lb h}^{-1} \text{ft}^{-2}$ ) or  $1.38 \text{ kg h}^{-1}\text{m}^{-3}$  ( $107 \text{ lb h}^{-1}\text{ft}^{-3}$ ) for this study.

#### A2.1.2 Hydrofluorination

The two-stage hydrofluorination reactors in the Allied Chemicals plant are designed for a 6,000 tonnes (U) per year throughput [Sutton *et al.* 1966]. It was assumed that half their bed volume would be required. An equal evaporative load on the dilute (azeotropic) hydrofluoric acid (DHF) and anhydrous hydrofluoric acid (AHF) boilers was assumed and the required surface area was calculated by reference to the HF vaporiser in the experimental fluorine plant [Royston *et al.* 1975]. The DHF and AHF superheaters also carry the same loads and raise the vapour temperature to  $\sim 400^\circ\text{C}$ .

#### A2.1.3 Fluorination

Fluorination reactor size was based on the report of Rogan [1972] who stated that the UK reactor was 760 mm (30 inches) diameter for a 3,000 tonnes (U) per year throughput. Fluorine preheater size was based on the volume of those in the Fluidised Bed Experimental Facility (FBEF) [Janov 1973]. The fluorine cleanup reactor was sized to have a 1-hour solids residence time in both moving bed and screw sections. The ash cleanup reactor was sized to have a 1-hour solids residence time. The surge drum was sized to contain 300 seconds of  $\text{UF}_6$ -flow as vapour at atmospheric pressure and  $65.5^\circ\text{C}$ . The alumina traps were based on the FBEF traps.

#### A2.1.4 Waste recovery and treatment

Costello's [1972] materials flowsheet shows that the high fluoride disposal equipment receives an equal load from hydrofluorination and fluorination processes. Table 5 summarises the waste disposal and recovery costs associated with that section of the plant under examination.

### A3. ASSESSMENT OF CAPITAL COSTS FOR A FLUOROX SCHEME I PLANT

#### A3.1 General

Uranyl fluoride ( $\text{UO}_2\text{F}_2$ ), together with catalyst and some uranium tetrafluoride ( $\text{UF}_4$ ), is discharged from the oxidation reactor, recycled through a separate and complete train of reduction and hydrofluorination (Scheme I), and mixed with the primary  $\text{UF}_4$  stream before entry into the oxidation reactor (Figure 6). A portion of the recycle stream is removed for processing to clean the catalyst for refining by the suppliers. The catalyst was assumed

to be 5 per cent by weight platinum on an alumina support with a concentration 10 per cent of the weight of  $UF_4$  in the bed [Batley et al. 1974]. Tables 1 to 3 show costs derived for each item of equipment for reduction, hydro-fluorination and fluorination respectively.

### A3.2 Reduction

The primary stream for reduction of uranium trioxide ( $UO_3$ ) was the same as that for DF. Batley et al. [1974] showed that the catalysed reduction of  $UO_2F_2$  is 40 per cent faster than the uncatalysed reduction of fluidised bed denitrated  $UO_3$ . However,  $UO_2F_2$  has a density of about one third that of fluid-bed  $UO_3$ ; and so a reactor volume 2.2 times greater was assumed. Only one cracking furnace was required to provide hydrogen for reduction in the two streams, sufficient spare capacity being available in the standard size unit assumed for DF to accommodate the additional load. However, separate heat exchangers for the off-gases were required, since the recycle reactor produces HF in the reduction of  $UO_2F_2$ .

### A3.3 Hydrofluorination

The hydrofluorination rate was assumed unaffected by the presence of catalyst. Therefore all lines were duplicated except for the final storage hopper (with doubled volume and flow capacities), and heat exchanger (with doubled surface area). HF storage had to be added since it was included in the DF fluorine generation section.

### A3.4 Oxidation

Double the throughput of  $UF_4$  was required in the Fluorox process. The size of reactor required depended upon reaction rate,  $UF_4$  concentration and bed density (assuming the catalyst concentration is negligible).

The weight of  $UF_4$  in the bed was assessed from the equation

$$\frac{dW}{dt} = -kW$$

where, using any consistent set of units,  $\frac{dW}{dt}$ , is the conversion rate,  $k$  the rate constant and  $W$  the weight of  $UF_4$  in the bed.

Batley et al. [1974] reported the fluidised bed reaction rate,  $k'$ , at 650°C as  $9.9 \times 10^{-3} \text{ mol m}^{-2}\text{h}^{-1}$ , where  $k'$  is the reaction rate constant  $k$  divided by surface area. They used  $UF_4$  with surface areas of 1.25 and 1.9  $\text{m}^2\text{g}^{-1}$ . Hawthorn et al. [1960] reported that the surface area of  $UF_4$  produced from fluid-bed  $UO_3$  was 0.1 to 0.5  $\text{m}^2\text{g}^{-1}$ . A mean surface area of 0.75  $\text{m}^2\text{g}^{-1}$  between  $UF_4$  (from the two streams of fluid-bed  $UO_3$  and recycled  $UO_2F_2$ ) and the reaction rate ( $k'$ ) above gave  $k = 2.33 \text{ h}^{-1}$ . With the assumption that the

particle size, catalyst concentrations and proportions were the same as those of Batley et al. [1974], the weight of  $UF_4$  in the bed was calculated to be 240 kg.

Janov [1973] reported that no sintering had been encountered in one experiment using a 100 per cent  $UF_4$  starting bed. For the same heat generation as in the DF reaction, 36 per cent  $UF_4$  could be used. However, the oxidation reactor operates at a lower temperature, and should produce less sintering. Moreover,  $UF_4$  passes out of the bed with the recycle  $UO_2F_2$  stream, increasing its total weight and the quantity of  $UF_4$  in the catalyst regeneration stream. A reasonable value of  $UF_4$  concentration was assumed to be 20 per cent.

Calculation of the bed volume and the assumption of a bed height-to-diameter ratio of three gave a bed diameter of 0.76 m, the same as that for DF. The corresponding cleanup reactor, its filter and intermediate screw feeder (required for DF), the ash cleanup reactor, ash discharge screw, and fluorine recycle blowers were not required for the Fluorox process. Discharge of the recycle  $UO_2F_2$  is by overflow to an additional hopper and screw feeder.

The oxygen preheater was assumed to be the same size as the fluorine preheater, as were the other items of associated equipment. Compressed air was used to supply 50 per cent of the oxygen required, the remaining oxygen being supplied from a bulk storage tank of capacity sufficient for one month's operation. The Fluorox catalyst storage vessel is considerably smaller than the DF calcium fluoride ( $CaF_2$ ) storage hopper.

### A3.5 Waste Recovery and Treatment

There is no electrolyte or  $CaF_2$  waste for disposal in the Fluorox process. Costello [1972] showed that ~ 8.5 per cent of the HF supplied to hydrofluorination appeared in the waste stream from HF recovery, but did not say what excess was assumed in calculating the quantity of HF to be recovered. However, Turner [1964] reported that at Allied Chemicals the ratio of weight HF to uranium fed to the hydrofluorination reactors was 0.4:1. From this it was concluded that Costello had assumed a minimum of 50 per cent efficiency of HF recovery and this was used here. In the recycle reduction of uranyl fluoride, essentially anhydrous HF is produced. Approximately 93 per cent of this HF can be recovered as anhydrous HF by cooling the off-gas to  $-40^\circ C$ , the remaining off-gas then passing to the HF recovery plant. A separate refrigeration plant to that in the  $UF_6$  production section was assumed to be required in conjunction with an additional heat exchanger, making the load on the HF recovery plant 2.19 times that of DF. Examination of the equipment

diagram for HF recovery [Costello 1972] shows that the major items of equipment are distillation towers and heat exchangers, with an appropriate scale factor of 0.65 in each case [Guthrie 1969].

The low fluoride waste stream also has a throughput 2.19 times greater and the high fluoride waste stream has a throughput 1.1 times greater, since only half the high fluoride waste in DF arises from HF recovery.

#### A3.6 Catalyst Regeneration

Catalyst has been used over four cycles of oxidation and reduction without showing any poisoning [Batley et al. 1974]. It was assumed that the catalyst would be significantly poisoned and require refining every 100 cycles, and that a continuous bleed would be cleaned (Figure 8) to separate the catalyst and recover the associated uranium. The cleaned catalyst would then be shipped to the suppliers for refining and platinum metal recovery. Costs of main plant items for catalyst cleaning are shown in Table 6. Note that all operations are batch, except for evaporation and drying of the  $UO_2F_2$  extracted in the first stage.

More frequent catalyst cleaning (up to every 20 cycles) would cost approximately double, because two centrifuges would operate continuously in series, evaporator and spray drier costs would double, and  $UF_4$  removal would remain a small scale batch process. These increases are negligible in the overall cost of the plant.

#### A4. ASSESSMENT OF CAPITAL COSTS FOR A FLUOROX SCHEME II PLANT

Batley et al. [1974] have proposed that reduction and hydrofluorination of recycle  $UO_2F_2$  can take place in the same reactor (Scheme II), using HF produced in the reduction. The two-stage hydrofluorination system proposed here lends itself particularly to this approach (Figure 7). Note that in this scheme the location of the water product of the hydrofluorination reaction is changed from the first reactor to the second, probably decreasing the degree of conversion of  $UF_4$ . This is not critical in this system, since unconverted material appears in the recycle stream after oxidation and its effect on processing or capital costs is small.

This scheme eliminates the recycle reduction reactors together with associated seal hoppers, filter and storage hopper, and the HF condenser and refrigeration requirement. The HF recovery requirement is double that for DF but the high fluoride disposal duty is equal to that of DF.

#### A5. MATERIALS FLOWSHEETS AND COSTS OF PROCESS FEED MATERIALS

Figure 1 shows the relevant portion of the materials flowsheet presented for DF by Costello [1972]. Figures 2 and 3 show corresponding materials

flowsheets for the Fluorox Schemes I and II respectively. Note that the recycle stream in Scheme I contains 1.2 times the uranium content of the DF primary stream, and in Scheme II a further 5 per cent because of the 20 per cent  $UF_4$  content assumed for the oxidation reactor in both Schemes, together with the assumption that the conversion in the combined reduction/hydrofluorination step in Scheme II is 95 per cent (see Section A4).

Table 8 shows the comparison of process feed requirements and costs. Compared with DF, ammonia usage (for the reduction reactors by cracking to  $N_2 + H_2$ ) is doubled for both Fluorox schemes, anhydrous HF usage is increased slightly in Scheme I, Fremy's salt (KH.HF) is not required in either scheme, while oxygen is an additional requirement for both. Lime usage is greater in Scheme I, but less in Scheme II compared with DF. Nitrogen usage is doubled in Scheme I compared with both DF and Scheme II since all the fluidisation gas required at operating feed rates is supplied by HF [Sutton et al. 1966] as can be confirmed by calculation.

The cost of catalyst refining has been quoted as  $\pounds 75 \text{ kg}^{-1}$  of contained platinum metal, less a credit for recovered platinum at the current metal price, assumed to be  $\pounds 2.267 \text{ g}^{-1}$  [Engelhardt Industries 1974] (Note:  $\$A \approx \pounds \text{stg } 0.6$ ). At the assumed throughput  $\$5,000 \text{ y}^{-1}$  is added to the processing cost. A 10 per cent loss of platinum metal adds  $\$12,000 \text{ y}^{-1}$  to the processing cost.

These assumptions result in net increases over DF in the cost of process feed materials of 1.9 cents  $\text{kg}^{-1}$  for Scheme I and 0.79 cents  $\text{kg}^{-1}$  for Scheme II.

## A6. SERVICES

### A6.1 Installation Costs

Installation costs of services for the Fluorox plants are lower than those for DF plant (see Table 9) because of two main factors associated with fluorine production; first, electrolytic fluorine cells are inefficient in their use of electricity, making the total installation large in comparison with the Fluorox recycle stream; and second, the floor area required for fluorine production is much greater than for the Fluorox recycle stream.

The calculated electrical installation required for the fluorine plant was based on a cell voltage of 10 V, and increased by 50 per cent to allow for other associated installations [Rudge 1971]. The power requirements for reduction and hydrofluorination were calculated by reference to the Allied Chemicals reactors [Sutton et al. 1966]. The floor area of the fluorine plant was calculated from Costello's [1972] plant layouts and used in the area calculation for electrical and plumbing installation.

### A6.2 Operating Costs

Costello [1974] has suggested on the basis of more recent information that the total electrical usage of  $16.2 \times 10^6$  kWh per annum assumed earlier is an underestimate and that it could be as high as  $33 \times 10^6$  kWh per annum. For the Fluorox plants, the data also suggest that  $10 \times 10^6$  kWh per annum could be saved by eliminating the direct current electrolysis component necessary to produce fluorine. This is  $4.6 \times 10^6$  kWh per annum more than the electrolysis usage calculated from the data of Huber et al. [1958] and Rudge [1971]. Reduction usage was based on sensible heat requirements and heat losses, and Smiley's statement [1961] that 65 per cent of the sensible heat requirements for cold powder and gas feeds are met by the heat of reaction. The hydrofluorination usage was based on an estimate that one fluidised bed of each two-stage hydrofluorination system would be exothermic [Sutton et al. 1966] and a reasonable assumption of the heating requirements of each system's second fluidised bed. Reduction in water usage for the Fluorox process was assessed on assumptions that fluorine cells are only about 18 per cent efficient [Rudge 1971], and waste heat is dispersed by water evaporation in a cooling tower, with its associated additional water usages e.g. mist blow-by, blow-down, etc. The effect of these factors is shown in Table 10.

### A7. LABOUR

Table 11 shows labour rates and annual labour costs estimated for the three processes, and the reductions obtained by use of either of the Fluorox schemes.

The labour requirement for fluorine cell operation was estimated to be 20 men on the basis of the data of Huber et al. [1958]. This was apportioned into the data of Costello [1972] as 17 process operators, 2 maintenance staff and 1 analytical services staff. These were replaced in the Fluorox Scheme I by 5 process operators working on the reduction/hydrofluorination recycle reactors and in Scheme II by 3 process operators also working on the recycle stream.

### A8. CAPITAL COSTS

Table 5 summarises the capital costs for DF and both Fluorox schemes for plants for complete conversion of yellow cake to  $UF_6$  using the data of Costello [1972] together with his capital charge components and percentages for process buildings, contingency, engineering design, procurement and installation, architect/layout and insurance/services.

A9. SUMMARY

Tables 1 to 3 compare the installed capital costs for DF with those for the Fluorox process Scheme I for the reduction, hydrofluorination and fluorination steps respectively. Table 5 compares the installed capital costs for waste recovery and disposal for the two processes, and Table 7 summarises the installed capital costs of the direct fluorination route and the two Fluorox schemes assessed.

Table 13 summarises the annual expenditures for conversion of yellow cake to  $UF_6$  for the three processes considered. All factors used in this table are the same as those used by Costello [1972]. Note that the unit toll conversion cost calculated from the sum of the annual expenditures in Costello's [1972] Table 6 is  $\$2.286 \text{ kg}^{-1}$  and not  $\$2.27 \text{ kg}^{-1}$  as quoted by him; however when the later data on electrical costs are taken into account this rises to  $\$2.434 \text{ kg}^{-1}$ .

TABLE 1  
COMPARISON OF REDUCTION COSTS BETWEEN DIRECT FLUORINATION  
AND IMPROVED FLUOROX PROCESSES

Item	Direct Fluorination			Fluorox Scheme I				
	Critical Dimension	Materials of Construction	No.	Cost (\$A)	Critical Dimension	Materials of Construction	No.	Cost (\$A)
Main hopper and filter	1.42 m <sup>3</sup>	SS	1	23,000	1.42 m <sup>3</sup>	SS	1	23,000
Seal hopper and feeder	1.42 m <sup>3</sup>	SS	1	27,500	473 kg h <sup>-1</sup>	SS	1	27,500
Reduction reactors	0.43 m dia	SS	2	34,000	0.43 m dia	SS	2	34,000
UO <sub>2</sub> seal hopper, feeder and filter	1.42 m <sup>3</sup>	SS	1	38,500	1.42 m <sup>3</sup>	SS	1	38,500
Heat exchanger	473 kg h <sup>-1</sup>	SS	1	8,000	473 kg h <sup>-1</sup>	SS	1	8,000
Cracking furnace	1.25 m <sup>2</sup>	SS	1	3,800	1.25 m <sup>2</sup>	CS/SS	1	3,800
UO <sub>2</sub> storage hopper and feeder	2.14 m <sup>2</sup>	CS/SS	1	5,100	2.14 m <sup>2</sup>	CS	1	5,100
Gas burner	1.42 m <sup>3</sup>	CS	1	27,500	1.42 m <sup>3</sup>	SS	1	27,500
<u>Recycle Stream</u> (not used in Scheme II)	473 kg h <sup>-1</sup>	CS	1	2,000	473 kg h <sup>-1</sup>	CS	1	2,000
Main hopper and filter					1.42 m <sup>3</sup>	Monel	1	39,400
Seal hopper and feeder					1.42 m <sup>3</sup>	Monel	1	47,300
Reduction reactors					540 kg h <sup>-1</sup>	Monel	2	116,000
UO <sub>2</sub> seal hopper, feeder and filter					0.53 m dia	Inconel/Monel	1	62,500
Heat exchanger					473 kg h <sup>-1</sup>	Inconel/Monel	1	15,200
UO <sub>2</sub> storage hopper and feeder					1.25 m <sup>2</sup>	CS/Monel	1	5,000
					2.14 m <sup>2</sup>			
					1.42 m <sup>3</sup>			
					473 kg h <sup>-1</sup>			
<b>TOTAL</b>				<b>169,400</b>				<b>502,100*</b>

\* Fluorox Scheme II cost = \$203,300

SS = Stainless Steel

CS = Carbon Steel.

TABLE 2  
 COMPARISON OF HYDROFLUORINATION COSTS BETWEEN DIRECT FLUORINATION  
 AND IMPROVED FLUOROX PROCESSES

Item	Direct Fluorination			Fluorox Scheme I				
	Critical Dimension	Materials of Construction	No.	Cost (\$A)	Critical Dimension	Materials of Construction	No.	Cost (\$A)
UO <sub>2</sub> seal hopper and feeder	1.42 m <sup>3</sup> 473 kg h <sup>-1</sup>	Monel	1	47,300	1.42 m <sup>3</sup> 473 kg h <sup>-1</sup>	Monel	2	94,600
Hydrofluorination reactors	0.91 m dia	Monel	2	232,000	0.91 m dia	Monel	4	464,000
UF <sub>4</sub> intermediate seal hopper and feeder	1.42 m <sup>3</sup> 550 kg h <sup>-1</sup>	Inconel/Monel	1	62,500	1.42 m <sup>3</sup> 550 kg h <sup>-1</sup>	Inconel/Monel	2	115,000
UF <sub>4</sub> product hopper and feeder	1.42 m <sup>3</sup> 550 kg h <sup>-1</sup>	Monel	1	47,300	1.42 m <sup>3</sup> 550 kg h <sup>-1</sup>	Monel	2	94,600
DHF superheater	18.6 m <sup>2</sup>	Monel	1	29,000	18.6 m <sup>2</sup>	Monel	2	58,000
AHF superheater	18.6 m <sup>2</sup>	CS	1	13,700	18.6 m <sup>2</sup>	CS	2	27,400
DHF boiler	16.7 m <sup>2</sup>	Monel	1	29,500	16.7 m <sup>2</sup>	Monel	2	59,000
AHF boiler	16.7 m <sup>2</sup>	CS	1	14,000	16.7 m <sup>2</sup>	CS	2	28,000
UF <sub>4</sub> storage hopper and feeder	1.42 m <sup>3</sup> 550 kg h <sup>-1</sup>	Monel	1	47,300	1.42 m <sup>3</sup> 550 kg h <sup>-1</sup>	Monel	1	55,000
Heat exchanger	2.32 m <sup>2</sup>	CS/Monel	1	5,000	4.64 m <sup>2</sup>	CS/Monel	1	5,500
HF storage	Included in Fluorine Generation				100 tonnes	CS	2	54,000
HF condenser					5 m <sup>2</sup>	CS/Monel	1	17,600*
Refrigeration (-40°C)					74 kW		1	110,800*
TOTAL				527,600				1,191,500

\* Not required in Scheme II  
 CS = Carbon steel.

**TABLE 3**  
**COMPARISON OF FLUORINATION COSTS BETWEEN DIRECT FLUORINATION**  
**AND IMPROVED FLUOROX PROCESSES**

Item	Direct Fluorination			Fluorox Scheme I and II				
	Critical Dimension	Materials of Construction	No.	Cost (\$A)	Critical Dimension	Materials of Construction	No.	Cost (\$A)
UF <sub>4</sub> feed hopper and feeder	1.42 m <sup>3</sup>	Monel	1	47,300	2.84 m <sup>3</sup>	Monel	1	55,000
UF <sub>4</sub> cleanup reactor	550 kg h <sup>-1</sup>	Monel	1	37,500	1,100 kg h <sup>-1</sup>	Monel	1	
Cleanup reactor filter	1 h res time	Monel	1	7,500				
UF <sub>4</sub> screw feeder to fluorinator	0.6 m <sup>2</sup>	Inconel/Monel	1	15,000				
Fluorinator and furnace	150 mm x 1.5 m	Monel	1					
	0.76 m dia				0.76 m dia			
	43 m <sup>2</sup>	Inconel	1	235,000	43 m <sup>2</sup>	Inconel	1	235,000
Caf <sub>2</sub> (or catalyst) feeder and hopper	1.42 m <sup>3</sup>	Monel	1	47,300	0.14 m <sup>3</sup>	Monel	1	14,000
Bed discharge screw	150 mm x 1.5 m	Monel	1	15,200				
Filter discharge screw	150 mm x 1.5 m	Monel	2	20,000	150 mm x 1.5 m	Monel	2	20,000
F <sub>2</sub> (or O <sub>2</sub> ) preheater	0.05 m <sup>3</sup>	Monel	1	20,000	0.05 m <sup>3</sup>	SS	1	13,000
Ash cleanup reactor	300 mm x 3 m	Monel	1	30,800				
Bed off-gas filter	1.25 m <sup>2</sup>	Inconel/Monel	2	30,400	1.25 m <sup>2</sup>	Inconel/Monel	2	30,400
Ash discharge screw	150 mm x 1.5 m	Monel	1	10,000				
F <sub>2</sub> recycle blower	2.6 m <sup>3</sup> h <sup>-1</sup>	Monel	2	200,000				
Roughing cold traps	4.65 m <sup>2</sup>	Monel	6	126,000	4.65 m <sup>2</sup>	Monel	6	126,000
Cleanup cold traps	4.65 m <sup>2</sup>	Monel	6	126,000	4.65 m <sup>2</sup>	Monel	6	126,000
Alumina traps	0.14 m <sup>3</sup>	Monel	12	62,500	0.14 m <sup>3</sup>	Monel	12	62,500
Scrubber	610 mm x 4.9 m	CS*	1	20,000	460 mm x 3.0 m	CS*	1	10,000
Pumps	7.5 t s <sup>-1</sup>				7.5 t s <sup>-1</sup>			
	5 MN m <sup>-2</sup>	various	9	31,500	5 MN m <sup>-2</sup>	various	9	31,500
Surge drum	2.8 m <sup>3</sup>	Monel	1	35,000	2.8 m <sup>3</sup>	MS	1	5,000
Cooling unit	75 t s <sup>-1</sup>		1	78,000	75 t s <sup>-1</sup>		1	78,000
Refrigeration unit	28.8 kW		1	60,000	28.8 kW		1	60,000
O <sub>2</sub> supply					10 tonnes			8,700
Air compressor					12 t s <sup>-1</sup>			11,500
UO <sub>2</sub> F <sub>2</sub> product seal hopper and feeder					1.42 m <sup>3</sup>			
					550 kg h <sup>-1</sup>	Monel		47,300
<b>TOTAL</b>				<b>1,255,000</b>				<b>933,900</b>

SS = Stainless steel  
CS = Carbon steel

MS = Mild steel  
\* = rubber lined.

TABLE 4  
BREAKDOWN OF WASTE RECOVERY AND DISPOSAL COSTS

Item	Cost (\$A)	Assignment
Nitric acid recovery	258,000	
High nitrate disposal	190,000	
High fluoride disposal	94,000	Equal quantities hydrofluorination and fluorination
Low fluoride streams	10,000	Hydrofluorination only
Low nitrate disposal	5,000	
Electrolyte from F <sub>2</sub> cells	5,000	Fluorination only
CaF <sub>2</sub> waste from fluorinator	5,000	Fluorination only
SUB TOTAL	567,000	
HF recovery	340,000	Equal quantities hydrofluorination and fluorination
TOTAL	907,000	

TABLE 5  
COMPARISON OF WASTE RECOVERY AND DISPOSAL COSTS BETWEEN  
DIRECT FLUORINATION AND IMPROVED FLUOROX PROCESSES

Item	Direct Fluorination		Fluorox Scheme I	
	Throughput	Cost (\$A)	Throughput	Cost (\$A)
Low fluoride streams		10,000		16,000
Electrolyte from F <sub>2</sub> cells		5,000		
CaF <sub>2</sub> waste from fluorinator		5,000		
High fluoride disposal	68 kg h <sup>-1</sup>	94,000	1,820 kg h <sup>-1</sup>	95,000
HF recovery		340,000		554,000
Catalyst cleaning				70,000
TOTAL		454,000		735,000

**TABLE 6**  
**CATALYST CLEANING, EQUIPMENT AND COST**

Equipment	Operation	Capacity	Cost (\$A)
<b>UO<sub>2</sub>F<sub>2</sub> Removal</b>			
Mixing tank	3 x batch	150 l	500
Solid bowl centrifuge	2-cycle batch	150 mm	6,000
Evaporator	continuous	0.09 m <sup>2</sup>	8,000
Spray drier	continuous	9 kg h <sup>-1</sup>	6,000
Intermediate tanks	-	2 x 150 l	500
Pumps	-	2 x 1 l s <sup>-1</sup>	800
<b>UF<sub>4</sub> Removal</b>			
Furnace 600°C	3 x batch	0.25 m <sup>2</sup> bed	10,000
Tanks	batch	2 x 100 l	500
Filter press	batch	0.65 m <sup>2</sup> P & F	1,500
Oven	batch		500
Pump	-	1 l s <sup>-1</sup>	400
<b>TOTAL (main plant items)</b>			<b>34,700</b>
<b>Cost with installation (2 x main plant items)</b>			<b>70,000</b>

TABLE 7  
SUMMARY OF COMPARISON OF CAPITAL COSTS BETWEEN DIRECT  
FLUORINATION AND TWO IMPROVED FLUOROX PROCESSES

Operation	Building Factor (%)	Direct Fluorination		Fluorox Scheme I		Fluorox Scheme II	
		Equipment Cost (\$A)	Installed Cost (\$A)	Equipment Cost (\$A)	Installed Cost (\$A)	Equipment Cost (\$A)	Installed Cost (\$A)
Reduction	20	169,400	203,300	502,100	602,500	169,400	203,300
Hydrofluorination	20	527,600	633,100	1,191,500	1,429,800	1,046,900	1,256,300
Fluorination	35	1,255,000	1,694,300	933,900	1,260,800	933,900	1,260,800
Fluorine generation	35	1,184,000	1,598,400	-	-	-	-
Waste recovery and disposal	20	454,000	544,900	735,000	882,000	694,000	832,500
TOTAL installed capital cost			4,674,000		4,215,100		3,553,200

TABLE 8  
COST OF PROCESS FEED MATERIALS

Feed Material	Material Cost \$/Mg	Direct Fluorination		Fluorox Scheme I		Fluorox Scheme II	
		Usage Mg/Mg U	Cost \$/Mg U	Usage Mg/Mg U	Cost \$/Mg U	Usage Mg/Mg U	Cost \$/Mg U
NH <sub>3</sub> anhydrous	162.5	0.048	7.8	0.0952	15.5	0.0952	15.5
HF anhydrous	530.0	0.57	302.1	0.58	307.4	0.57	302.1
KHF <sub>2</sub>	673.0	0.01	6.7	-	-	-	-
Lime	16.0	0.5	8.0	0.52	8.3	0.39	6.2
Oxygen	80.0	-	-	0.037	3.0	0.037	3.0
Nitrogen	84.7	0.053	4.5	0.106	9.0	0.053	4.5
Catalyst	-	-	-	-	5.7	-	5.7
Miscellaneous	-	-	3.2	-	3.2	-	3.2
<b>TOTAL Materials Cost (\$/Mg U)</b>			<b>332.3</b>		<b>352.1</b>		<b>340.2</b>
<b>Difference from Direct Fluorination</b>			<b>-</b>		<b>+18.8</b>		<b>+7.9</b>

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TABLE 9  
SERVICES - COST OF INSTALLATION

Service	Cost Basis (\$A)	Direct Fluorination		Fluorox Scheme I		Fluorox Scheme II	
		Capacity	Cost (\$A)	Capacity	Cost (\$A)	Capacity	Cost (\$A)
Electricity	{ 100/kVA +10.76 m <sup>-2</sup>	4,600 kVA	460,000	3,900 kVA	390,000	3,600 kVA	360,000
Plumbing		7,740 m <sup>2</sup>	83,300	6,640 m <sup>2</sup>	71,500	6,540 m <sup>2</sup>	70,400
Others*	10.76 m <sup>-2</sup>	7,740 m <sup>2</sup>	83,300	6,640 m <sup>2</sup>	71,500	6,540 m <sup>2</sup>	70,400
TOTAL			928,640		835,040		802,840
Reduction over Direct Fluorination			-		93,600		125,800

\* See Costello (1972) for details

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TABLE 10  
SERVICES - ANNUAL OPERATING COSTS

Service	Rate Assumed	Direct Fluorination		Fluorox Scheme I		Fluorox Scheme II	
		Usage	Cost (\$A)	Usage	Cost (\$A)	Usage	Cost (\$A)
Electricity	2¢/kWh	33.0 x 10 <sup>6</sup> kWh	660,000	23.7 x 10 <sup>6</sup> kWh	474,000	23.35 x 10 <sup>6</sup> kWh	467,000
Water	0.088 \$A m <sup>-3</sup>	4.43 x 10 <sup>5</sup> m <sup>3</sup>	39,000	4.16 x 10 <sup>5</sup> m <sup>3</sup>	36,700	4.16 x 10 <sup>5</sup> m <sup>3</sup>	36,700
Steam	0.0022 \$A kg <sup>-1</sup>	56.35 x 10 <sup>6</sup> kg	124,000	56.35 x 10 <sup>6</sup> kg	124,000	56.35 x 10 <sup>6</sup> kg	124,000
Annual Cost			823,000		634,700		627,700
Reduction in Annual Cost			-		188,300		195,300

TABLE 11  
LABOUR RATES AND ANNUAL COSTS

Classification	Rate \$A Y <sup>-1</sup>	Direct Fluorination		Fluorox Scheme I		Fluorox Scheme II	
		No.	Cost (\$A)	No.	Cost (\$A)	No.	Cost (\$A)
Analytical services	5,337	5	26,685	4	21,348	4	21,348
Maintenance staff	3,947	10	39,470	8	31,576	8	31,576
Process operator							
1. reduction and hydro-fluorination	5,337	11	58,707	16	85,392	14	74,718
2. direct fluorination or Fluorox oxidation	5,337	27	144,099	10	53,370	10	53,370
3. others*	*	16	85,392	16	85,392	16	85,392
Other staff*	*		240,467		240,467		240,467
<b>SUB-TOTALS</b>			594,820		517,545		506,871
+ overheads (25 per cent)*			148,705		129,385		126,719
<b>TOTAL \$A Y<sup>-1</sup></b>			743,525		646,930		633,590
Reduction over direct fluorination					96,595		109,935

\* See Costello (1972) for details.

TABLE 12  
CAPITAL SUMMARY

Item	Direct Fluorination Cost (\$A)	Fluorox Scheme I Cost (\$A)	Fluorox Scheme II Cost (\$A)
<b>1. <u>SITE AND SITE DEVELOPMENT</u></b>			
1.1 Land \$6,000/acre (30 acres)	180,000	180,000	180,000
1.2 Land services \$8,000/acre	240,000	240,000	240,000
1.3 Process buildings 35% of 2.1.4 & 2.1.5	854,000	327,000	327,000
20% of 2.1 & 2.2 excluding 2.1.4 & 2.1.5	540,000	797,000	688,000
1.4 Process services	929,000	835,000	803,000
<b>2. <u>EQUIPMENT</u></b>			
2.1 Process equipment			
2.1.1 Dissolver/solvent Extraction/denitration	871,000	871,000	871,000
2.1.2 Reduction	169,400	502,100	169,400
2.1.3 UF <sub>4</sub> conversion	527,600	1,191,500	1,046,900
2.1.4 UF <sub>6</sub> conversion	1,255,000	933,900	933,900
2.1.5 Fluorine production	1,184,000	-	-
2.1.6 HF recovery	340,000	554,000	515,000
2.1.7 Waste recovery	567,000	634,000	632,000
2.2 Contingency on equipment 5%	227,000	234,000	208,000
<b>3. <u>FEES AND SERVICES</u></b>			
3.1 Architect/layout, 10% of 1.2, 1.3 & 1.4	256,000	220,000	206,000
3.2 Engineering design, procurement and installation, 55% of 2	2,827,000	2,706,000	2,407,000
3.3 Insurance/services, 1% of 1 & 2	76,000	71,000	64,000
<b>4. <u>PRE-OPERATIONAL COSTS</u></b>			
4.1 Interest on capital 1 year at 8.75% on debt component (50% fixed capital)	561,000	502,500	458,000
4.2 Labour 6 months' annual cost	371,000	324,000	317,000
4.3 Consumables 6 months' annual cost	784,000	812,000	795,000
4.4 Services 6 months' annual cost	412,000	318,000	314,000
4.5 Losses 0.5% of throughput during commissioning	75,000	75,000	75,000
<b>5. <u>TOTAL FIXED CAPITAL</u></b>	13,246,000	12,328,000	11,250,000
<b>6. <u>OPERATING CAPITAL</u></b>			
6.1 Work in hand, 1 month's supply of process chemicals	109,000	112,000	110,000
6.2 Stores and spares 10% of 2	514,000	492,000	438,000
6.3 Funds 10% of annual operating costs	725,000	680,000	640,000
<b>7. <u>TOTAL CAPITAL</u></b>	14,594,000	13,612,000	12,438,000
7.1 10 year capital (2.1 + 2.2 + 7/11 x 3.2)	6,940,000	6,642,000	5,908,000
7.2 15 year capital (7 - 6 - 7.1)	6,306,000	5,686,000	5,342,000
7.3 Returned capital	1,348,000	1,284,000	1,188,000

TABLE 13

ANNUAL EXPENDITURES

Item	Direct Fluorination Cost (\$A)	Fluorox Scheme I Cost (\$A)	Fluorox Scheme II Cost (\$A)
<b>1. MANUFACTURING COSTS</b>			
1.1 Depreciation (8.75% interest yield)			
(a) 10 year capital (0.083)	576,000	551,000	490,000
(b) 15 year capital (0.050)	315,000	284,000	267,000
1.2 Labour	743,000	647,000	634,000
1.3 Consumables			
(a) Process chemicals	1,303,000	1,359,000	1,327,000
(b) Maintenance (2% fixed capital)	255,000	240,000	218,000
1.4 Services	823,000	635,000	628,000
1.5 Fees (1% fixed capital)	127,000	120,000	109,000
1.6 Contingency (10% of above)	372,000	354,000	337,000
<b>2. SAMPLING AND TRANSPORT COSTS</b>	480,000	480,000	480,000
<b>3. FINANCIAL EXPENDITURES</b>			
3.1 Interest payment of debt capital (8.75% of 50% of total capital)	640,000	596,000	544,000
3.2 Return to equity (12% of 50% of total capital)	876,000	817,000	746,000
3.3 Implied taxation $(3.2 \times \frac{0.475}{1-0.475})$	793,000	740,000	675,000
<b>COST OF TOLL CONVERSION</b>	<b>7,303,000</b>	<b>6,823,000</b>	<b>6,455,000</b>
<b>UNIT COST OF TOLL CONVERSION</b> (3,000 Mg U y <sup>-1</sup> , \$A kg <sup>-1</sup> )	<b>2.434</b>	<b>2.274</b>	<b>2.152</b>
Reduction from direct fluorination (\$A)		480,000	848,000
Reduction in toll conversion cost (\$A kg <sup>-1</sup> )		0.16	0.28

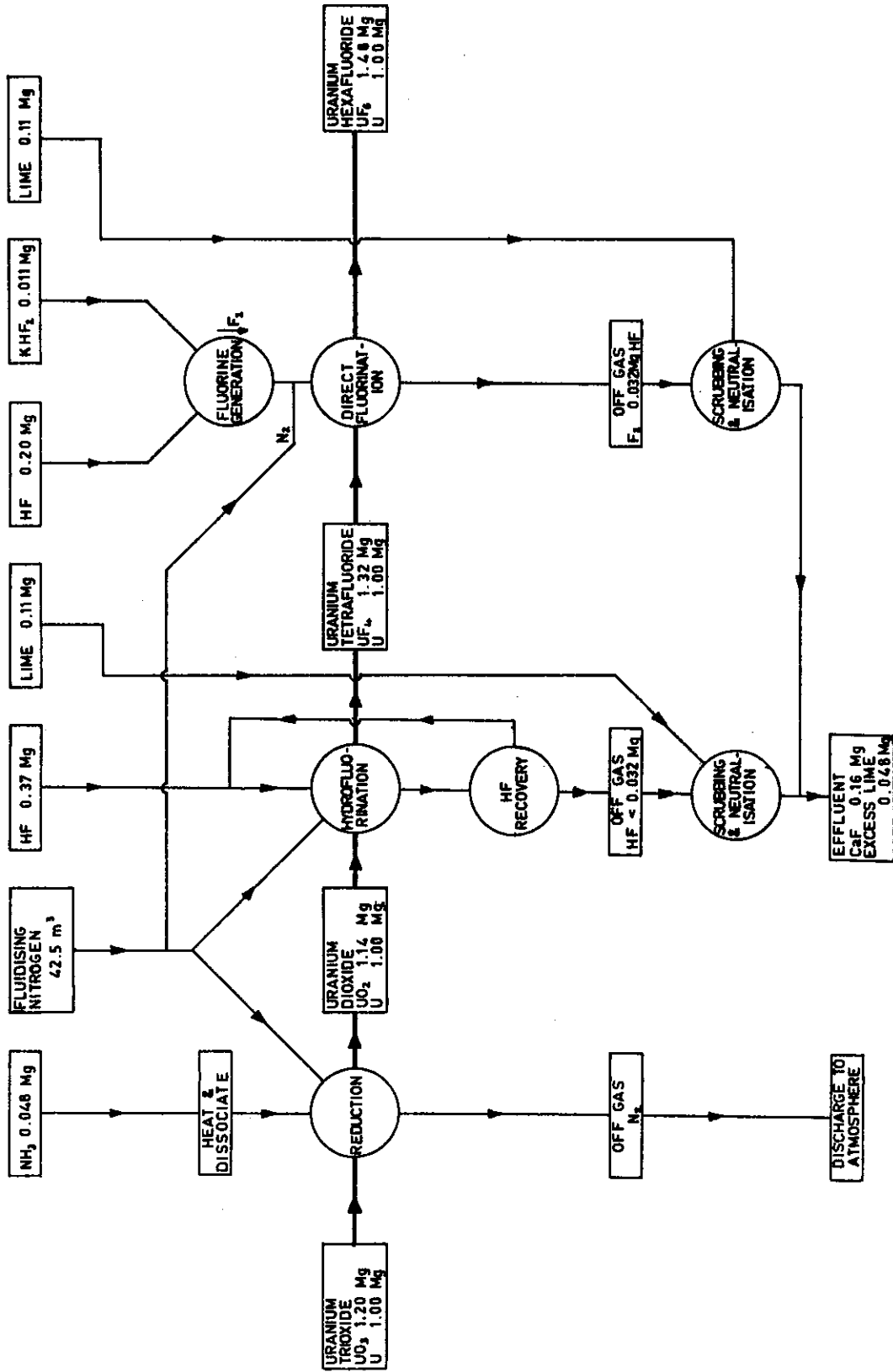


FIGURE 1. MATERIALS FLOWSHEET FOR DIRECT FLUORINATION (after Costello 1972)

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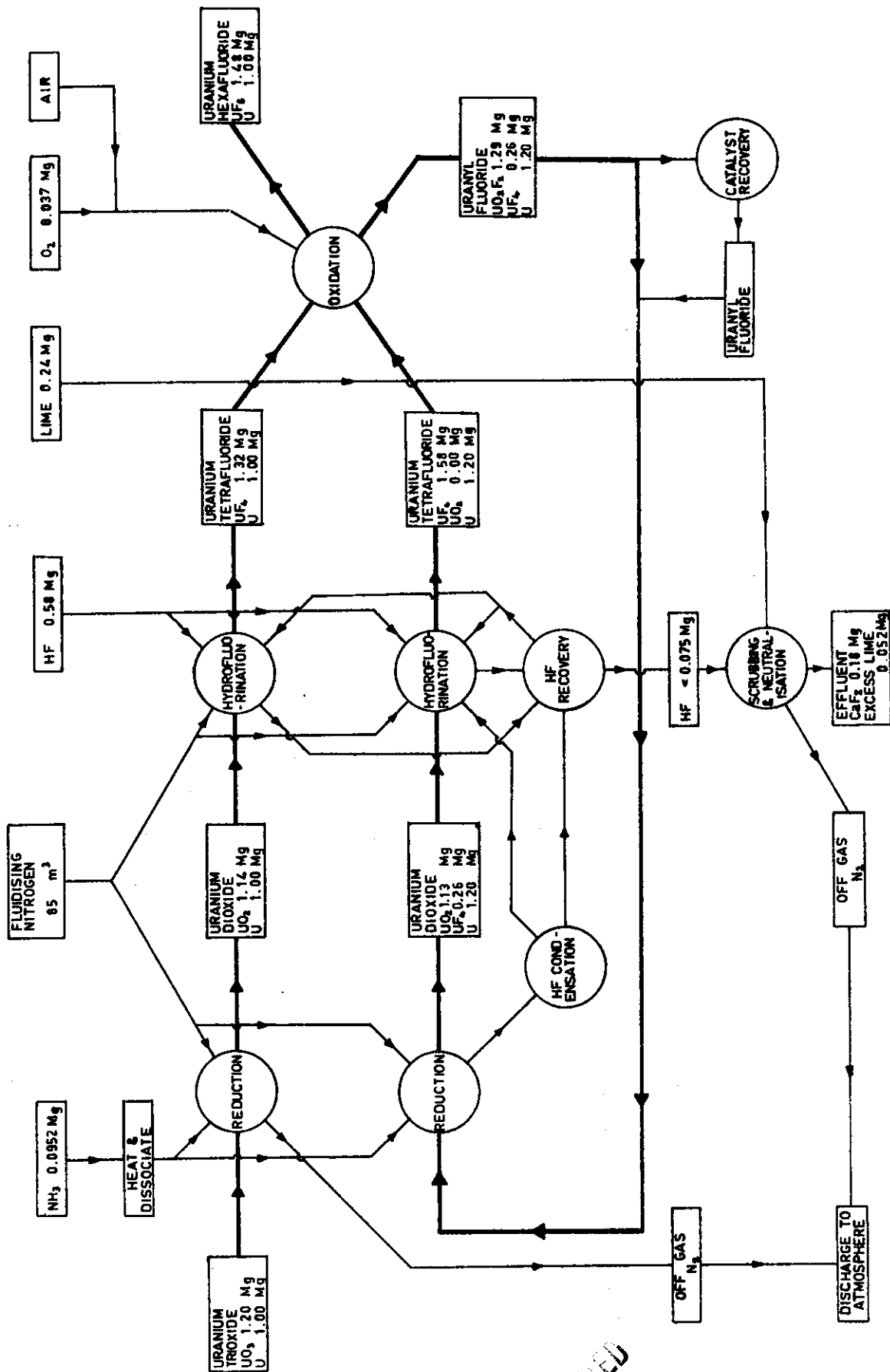


FIGURE 2. MATERIALS FLOWSHEET FOR FLUOROX SCHEME I

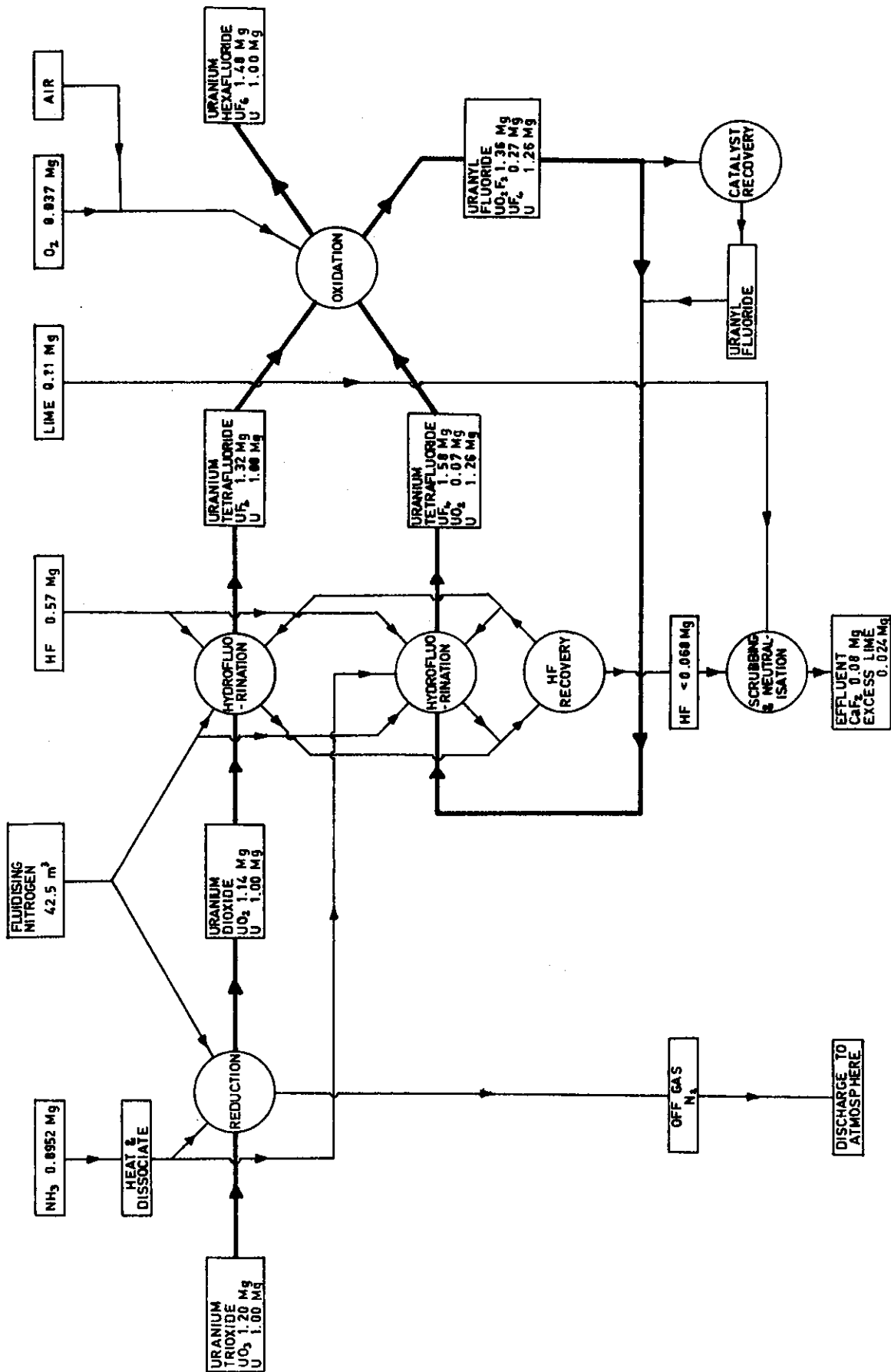


FIGURE 3. MATERIALS FLOWSHEET FOR FLUOROX SCHEME II

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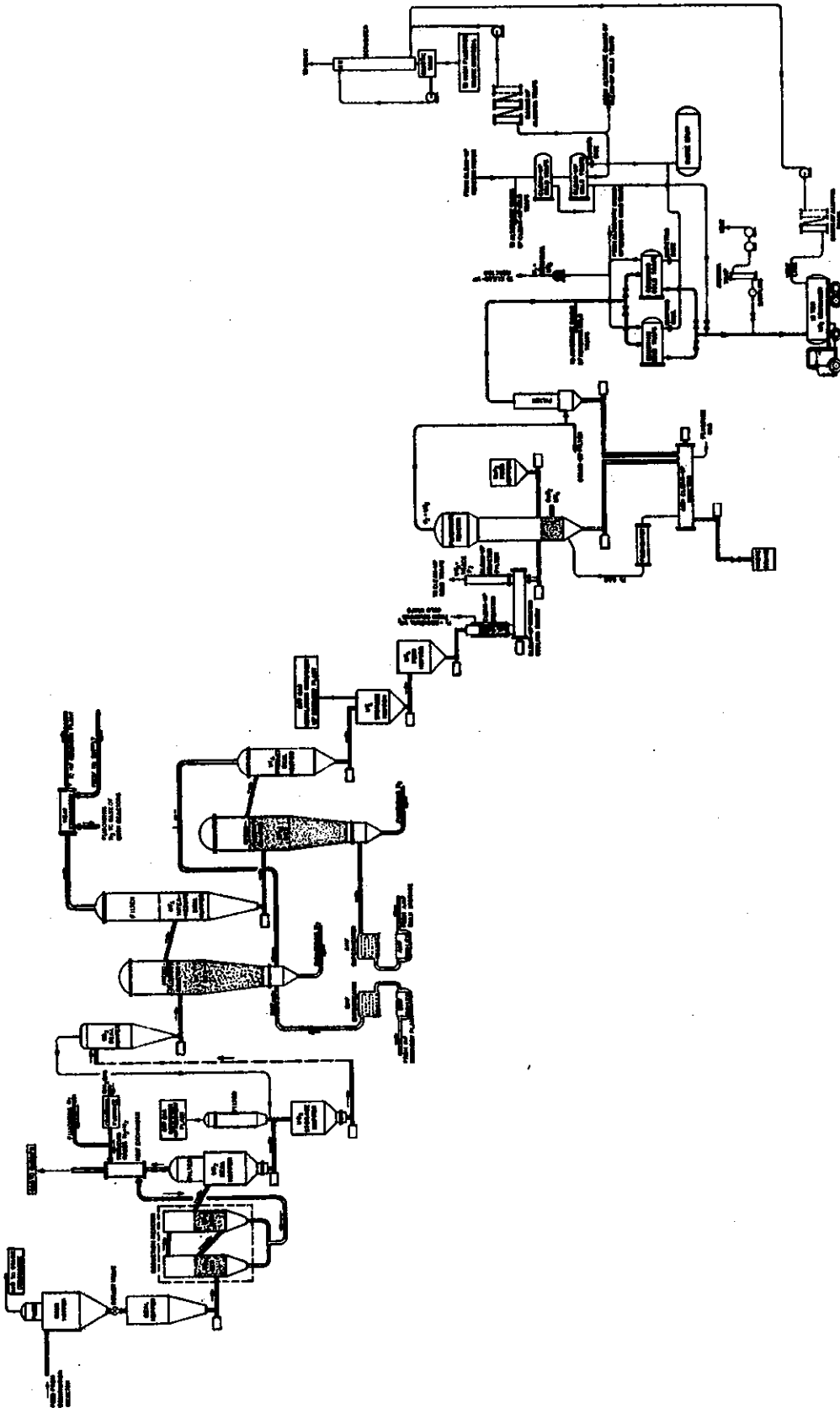


FIGURE 4. DIRECT FLUORINATION FLOW DIAGRAM (after Costello 1972)

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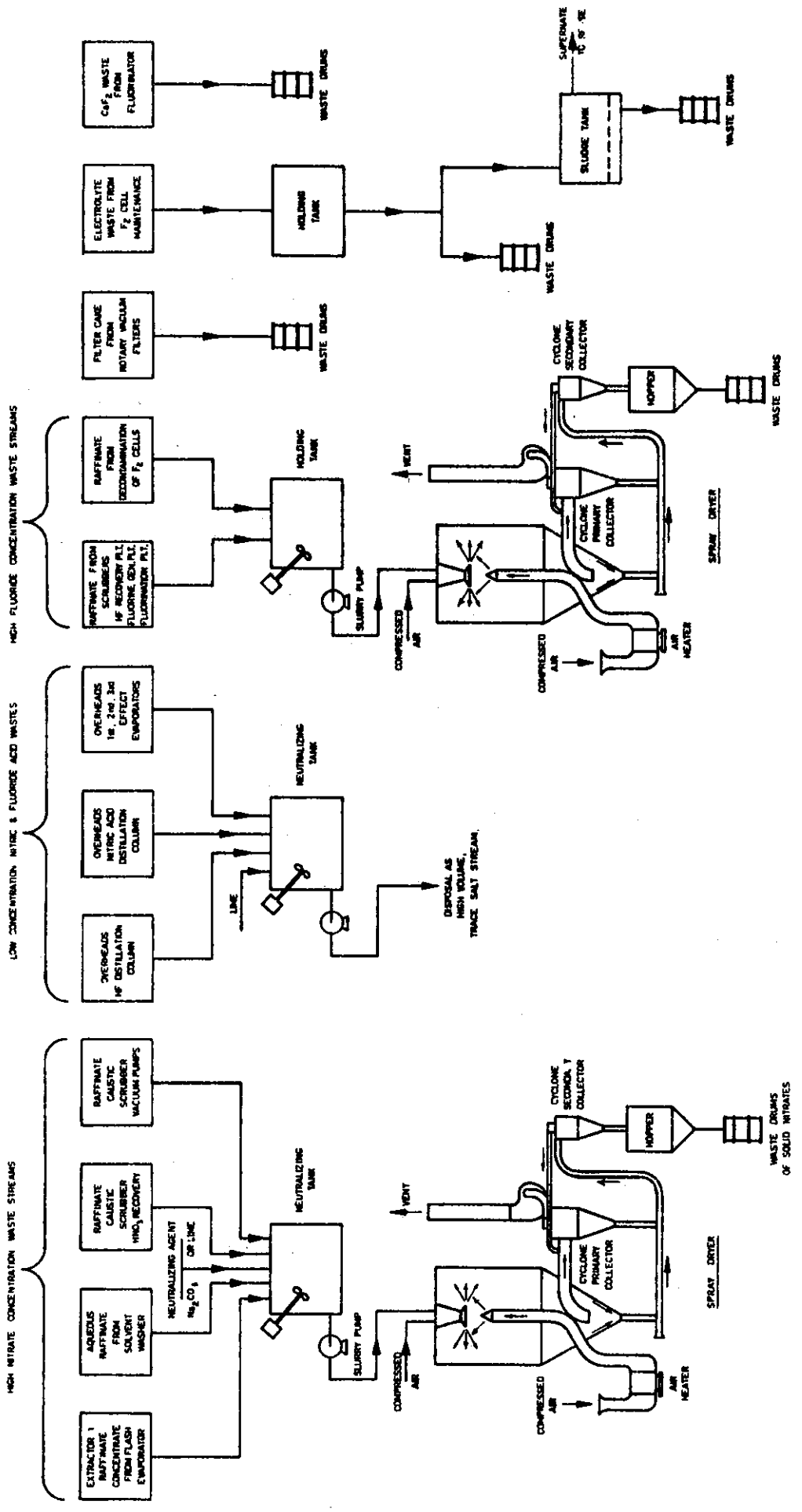


FIGURE 5. WASTE DISPOSAL FLOWSHEET (after Costello 1972)

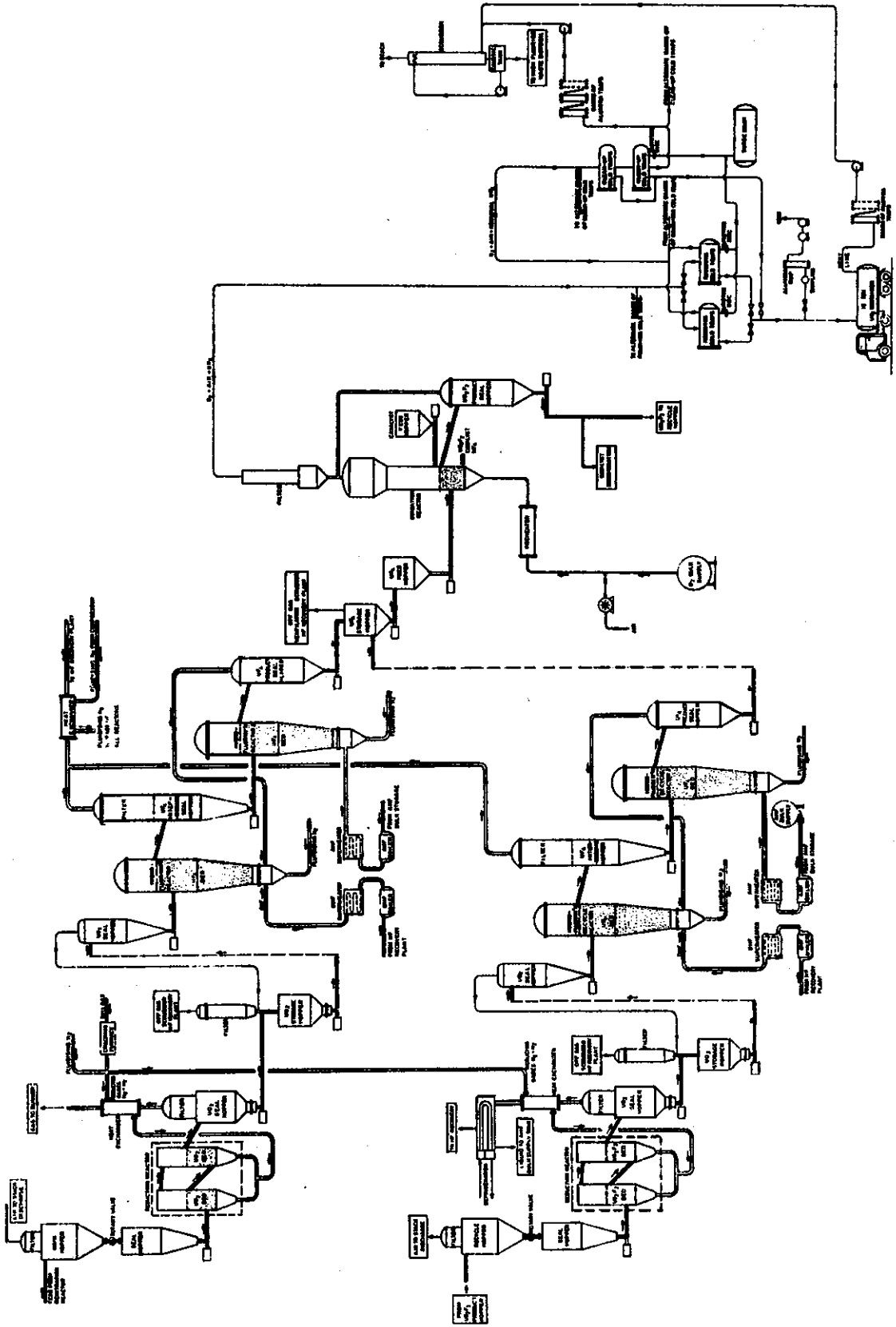


FIGURE 6. FLUOROX SCHEME I FLOW DIAGRAM

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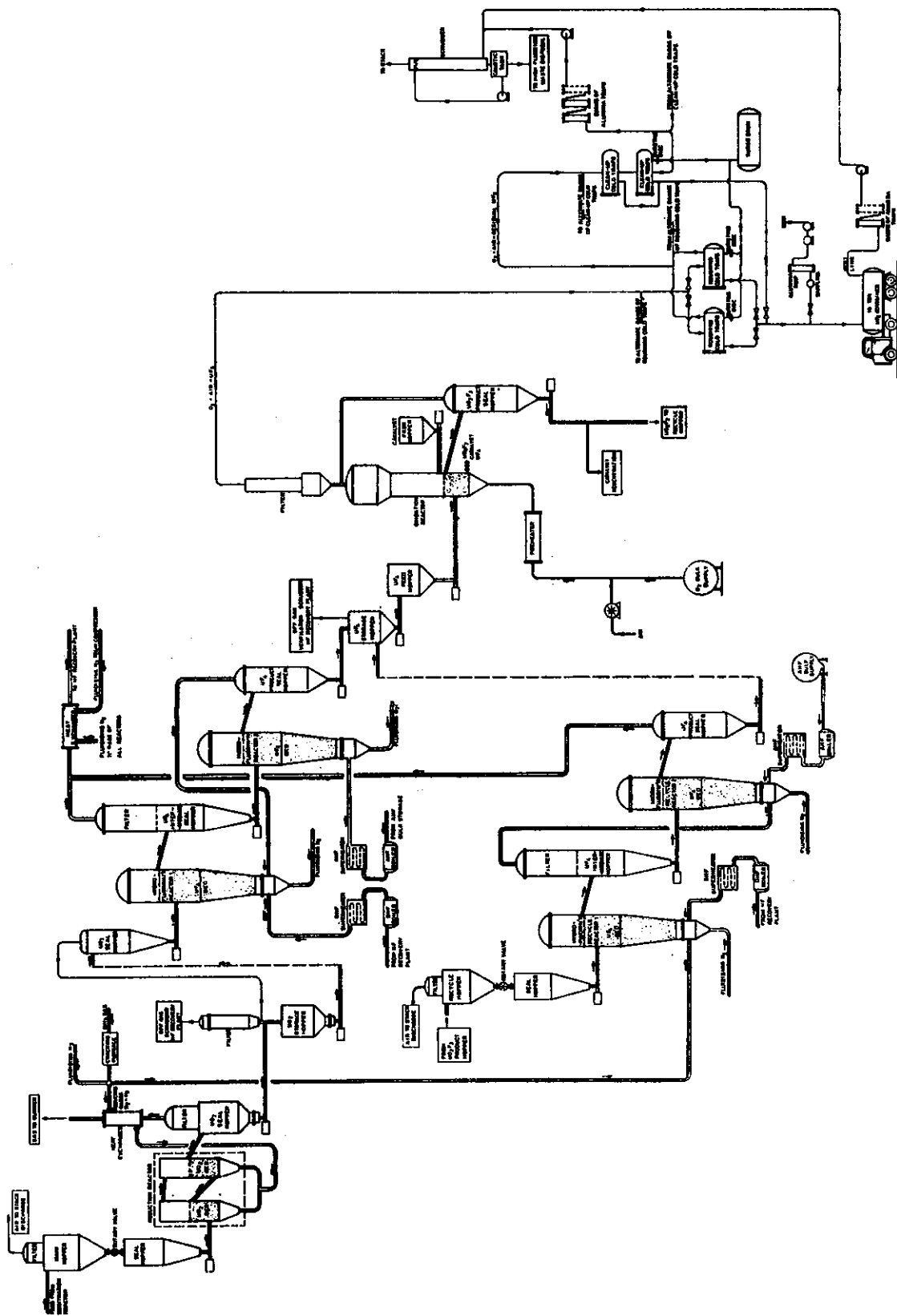


FIGURE 7. FLUOROX SCHEME II FLOW DIAGRAM

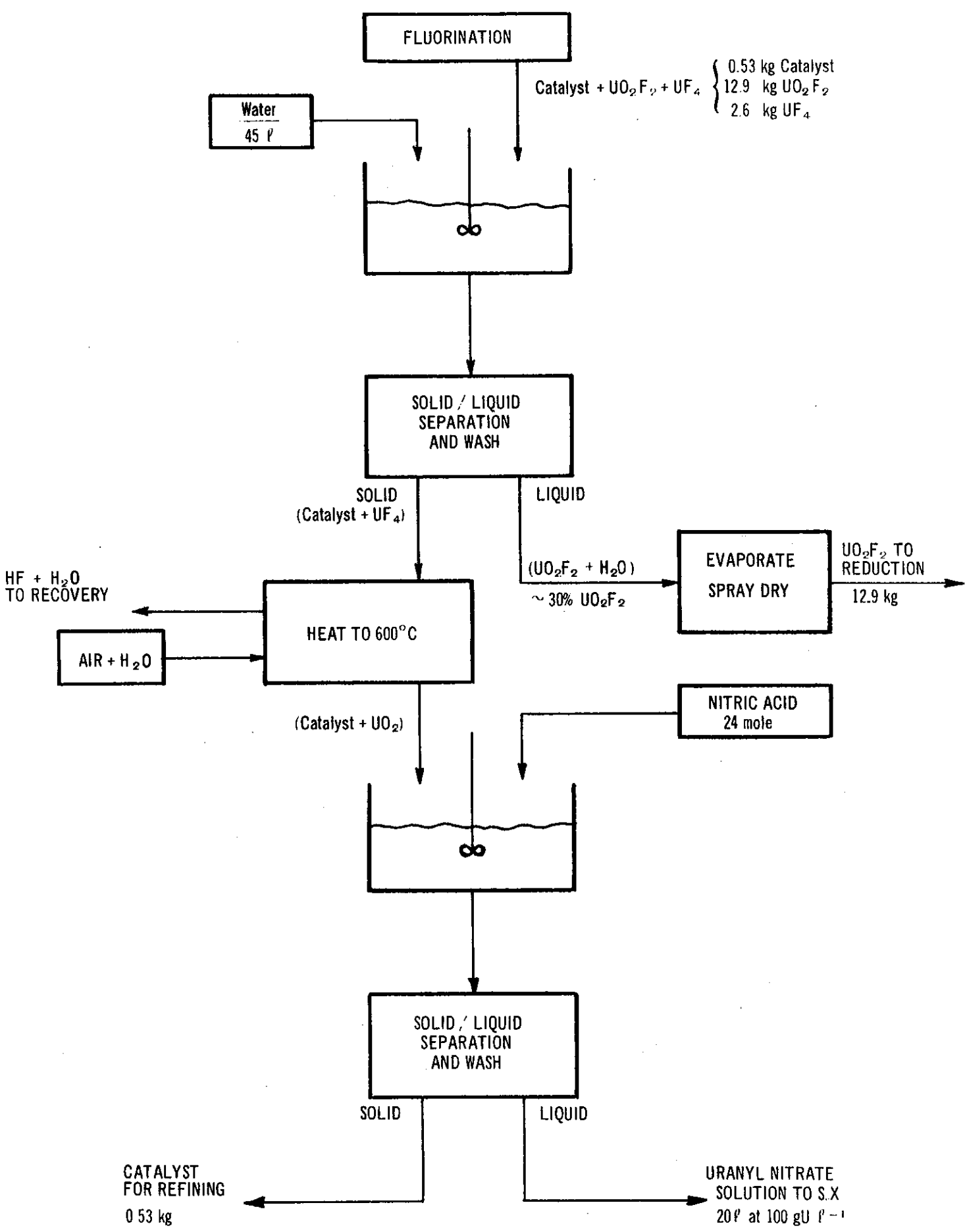


FIGURE 8. CATALYST CLEANING FLOWSHEET