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AUSTRALIAN ATOMIC ENERGY COMMISSION

RESEARCH ESTABLISHMENT

LUCAS HEIGHTS

DEVELOPMENT OF SOLVENT EXTRACTION PROCESSES FOR THE H.T.G.C.R. FUEL CYCLE

PART 1. DESIGN OF A FLOWSHEET FOR THE RECOVERY OF ACTINIDES

by

M.G. BAILLIE

R.K. RYAN

Issued Sydney, June 1965

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PREFACE

This report is part 1 of a series of reports on the development of solvent extraction processes for the High Temperature Gas-Cooled Reactor fuel cycle.

Companion series deal with :

Laboratory Development of the Grind-Leach Process for the H.T.G.C.R. Fuel Cycle,

and :

Economics of the H.T.G.C.R. Fuel Cycle.

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ABSTRACT

A preliminary flowsheet for the recovery and decontamination of residual actinides in spent H.T.G.C.R. fuel has been designed. A calculation method for the multicomponent liquid–liquid extraction system $H_2O - UO_2(NO_3)_2 - Th(NO_3)_4 - Be(NO_3)_2 - HNO_3 -$ TBP – Kerosene using the limited equilibrium data available is described. The flowsheet which has been designed incorporates a novel "split contactor" concept which is necessary to prevent third phase problems.

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1. INTRODUCTION

As part of the Australian Atomic Energy Commission's high temperature gas cooled reactor feasibility study, attention has been given to the development of a suitable process for recovering actinides from spent fuel elements. The purpose of this work is to demonstrate a practical recovery scheme to a point which will enable a cost estimate for the process to be made. A liquid–liquid extraction process was selected as it showed more promise of early development; however it is appreciated that other processing techniques (such as halide volatility) may ultimately prove superior.

Few basic chemical data relating to the type of feed solution to be processed were available in the literature. The approach adopted in developing a flowsheet, therefore, was to use the limited data available, with extrapolation where necessary, to provide a "first guess" flowsheet, which could then be developed further by experiment.

The preliminary calculations, which enabled an estimate of the number of ideal stages for liquid–liquid extraction to be made, are reported here.

2. THE FUEL ELEMENT

The fuel elements under consideration consist of a dispersion of small diameter particles of fissile–fertile oxides in a matrix of beryllia. During the period of development which has occurred since the H.T.G.C.R. study was first undertaken, the composition and shape of the proposed fuel elements have been subjected to many changes. However, these have not significantly affected the actinide separation step of the reprocessing scheme, and the solvent extraction flowsheet development programme is discussed here as it applies to the fuel element proposed at December 1964.

This fuel element is a sphere of approximately one inch diameter. The atomic ratio of elements present is 1:16.5:1650 (fissile:fertile:moderator), and the fuel consists of a dispersion of $(Pu, Th)O_2$ in BeO, the Pu having an initial isotopic content: Pu^{239}, 78%; Pu^{240}, 17%; Pu^{241}, 5%.

3. THE H.T.G.C.R. FUEL CYCLE

The composition of the fuel after reaching the level of burnup at which it is rejected is quite different to the input composition. Table 1 lists the composition of a complete core of an H.T.G.C.R. of the reference design (Ebeling, Unpublished data) after a burnup of 2.5 fissions per initial fissile atom (FIFA).

The quantities of materials discharged, with the exception of fission products, are those calculated when using the FEVER code (E. Hesse, private communication). Total fission product yield was obtained using the data of Frost (1964).

The recovery of uranium, plutonium, and thorium from a nitric acid leach solution of this fuel, and the decontamination of these materials from fission products and beryllium, is one step in the overall fuel cycle for H.T.G.C.R. fuel. Figure 1 shows the relationship of this step (the separation of the actinides) to the others making up the fuel cycle as currently conceived.

Leaching and dissolution studies have shown that the fissile/fertile material cannot be leached from the ground fuel elements without some dissolution of the BeO moderator matrix (Farrell and Shying 1965). For the solvent extraction studies it was assumed that 20 per cent. of the BeO would appear in the feed solution for solvent extraction of the actinides. This appears to be a suitable figure based on inactive leaching studies. Under these circumstances, the composition of the solution resulting from nitric acid leaching of the fuel is as shown in Table 2.

It was decided that the flowsheet to be developed should provide for the co–recovery and decontamination of uranium, plutonium, and thorium. The reasons for recovering thorium have been given previously (Cairns 1964) and include the following :

(1) The fuel cycle under development utilizes a liquid–liquid extraction step, and a sol–gel reconstitution step in which these materials may be recycled together.

(2) If not recovered, thorium appears in a waste stream and must be stored.

(3) Conservation of the fertile material is desirable.

4. THE SOLVENT EXTRACTION STEP

The solvent tributyl phosphate (TBP) was selected for use in the solvent extraction step. This solvent has been used extensively and successfully in similar applications. A Kerosene-type diluent, Odourless Mineral Spirits (OMS) was chosen.

In general, the distribution coefficients (D) for the components present in the feed solution (Benedict and Pigford 1957) vary as shown below:

$$D_{U\,(VI)} > D_{Pu\,(IV)} > D_{Th} > D_{HNO_3} > D_{\text{Fission Products}}$$

However, most fission products extract to some extent.

Conditions in the extraction section of the solvent extraction process can be selected to use these differences in extractability to enable recovery and decontamination of the heavy metals. Greater decontamination from fission products can be obtained by:

(1) Saturation of the organic phase with heavy metals (U, Pu, Th).

(2) Scrubbing the extract

(a) with inorganic nitrates (for example $Be(NO_3)_2$, $Al(NO_3)_3$),

(b) with nitric acid.

(3) Using a number of solvent extraction cycles.

4.1 Calculation Method

The McCabe-Thiele method (Faust et al. 1960) has been used to calculate the number of ideal stages required for separation of uranium and thorium from fission products. No separate calculations have been made for plutonium, as it is assumed to follow the uranium and thorium if maintained in the Pu(IV) state. The system being considered, then, is $H_2O - UO_2(NO_3)_2 - Th(NO_3)_4 - Be(NO_3)_2 - HNO_3$ – TBP – OMS.

The solvent extraction step may be considered in three separate sections for the purpose of calculating ideal stage requirements. These are:

(1) The extraction section.

(2) The scrubbing section.

(3) The stripping section.

Certain approximations are permissible in order to simplify the calculations. They are, however, different for each of the three sections.

4.1.1 Extraction section

In the extraction section, uranium, thorium, and nitric acid are extracted from an acidic beryllium nitrate solution. The distribution coefficients for the components are functions of the aqueous nitrate ion concentration, and extraction takes place according to the following equations:

$$2TBP]_O + UO_2^{2+}]_A + 2NO_3^-]_A \longrightarrow UO_2(NO_3)_2 \,.\, 2TBP]_O \quad * \quad \ldots\ldots (1)$$

$$TBP]_O + HNO_3]_A \longrightarrow HNO_3 \,.\, TBP]_O \quad * \quad \ldots\ldots (2)$$

* Gresky (1956)

In dilute solutions:

$$4TBP]_O + Th^{4+}]_A + 4NO_3^-]_A \longrightarrow Th(NO_3)_4 \cdot 4TBP]_O \quad * \quad \ldots\ldots (3a)$$

In concentrated solutions:

$$2TBP]_O + Th^{4+}]_A + 4NO_3^-]_A \longrightarrow Th(NO_3)_4 \cdot 2TBP]_O \quad ** \quad \ldots\ldots (3b)$$

The distribution coefficients depend on the aqueous nitrate ion concentration, which is seen to be reduced during extraction. However, the proportion of the total nitrate ion concentration supplied by the extractable species is small compared with that associated with the beryllium. Aqueous nitrate ion concentration is therefore essentially constant during extraction, resulting in a constant distribution coefficient for each of the constituents during extraction. Simple McCabe – Thiele calculations can therefore be made to estimate ideal stage requirements. A sample calculation is given in Appendix 1.

4.1.2 Scrub section

In the scrub section, a significant proportion of the aqueous nitrate ion concentration is supplied by extractable species (notably $Th(NO_3)_4$ and HNO_3). This is true even when a non-extractable salting agent is used in the scrub solution. Distribution coefficients will therefore vary with the concentration of nitric acid in the aqueous phase, which in turn varies as nitric acid is extracted. In the system under study, the nitrate ion concentration associated with thorium nitrate in the aqueous phase is also significant and has a similar variable effect on distribution coefficients. The nitrate ion concentration associated with uranium is, however, negligible and can be ignored.

In calculating stage requirements, then, interaction between thorium and nitric acid extraction must be considered. Uranium extraction may be determined later by using the information obtained when making the thorium and nitric acid calculations. An equilibrium relationship is therefore required between these competing reactions; that is between $[Th]_A$, $[Th]_O$, $[HNO_3]_A$, and $[HNO_3]_O$ (Siddall 1956; Wood and Williams 1958). Two methods of representing this relationship are:

(1) A graphical representation of $[Th]_O$ versus $[Th]_A$ with $[HNO_3]_O$ and $[HNO_3]_A$ as parameters.

(2) A graphical network obtained by plotting $[HNO_3]_O$ versus $[Th]_O$ with $[HNO_3]_A$ and $[Th]_A$ as parameters.

Method (1) may be considered as two super-imposed graphs – a linear one of $[Th]_O$ versus $[Th]_A$ and a curvilinear one of $[HNO_3]_O$ versus $[HNO_3]_A$ – on which McCabe-Thiele diagrams can be drawn. These must give the same equilibrium conditions on both linear and curvilinear graphs and, for the concentrations under consideration, the method proved to be impractical.

Method (2) was satisfactory, when used with analytical material balances and trial and error calculations. An example of this type of calculation is given in Appendix 2.

4.1.3 Stripping section

The calculation method necessary in the stripping section is similar to that used in the scrub section and described in Section 4.1.2 above.

4.2 Equilibrium Data

A survey of the available literature revealed that very few equilibrium data were available for systems of the type under study, in which the extractable species are present in very low concentrations. Approximate information can however, be obtained by calculation. It was decided to use such a method to obtain the data for use in the preliminary ideal stage calculations. More accurate equilibrium data may then be obtained by experiment during the developmental programme.

* Gresky (1956)
** Hesford etal. (1957)

Stoller and Richards (1961) state that for the thorium concentrations encountered (<0.2 M), a linear equilibrium relationship is approximately valid at nitric acid molarities greater than 1 M. The distribution coefficient for thorium (D_{Th}) is found to vary with aqueous nitric acid concentration when extracting thorium into 40 per cent. TBP in OMS from a nitric acid solution as shown in Figure 2 (Siddall 1958). This information has been used to prepare Figure 3, which shows their inter-relationship as a family of lines.

Siddall (1956) reports that for nitric acid concentrations > 1 M:

$$\frac{H'_O}{H_O} \approx 1 - \frac{2Th_O}{T_O} \quad \ldots\ldots\ldots\ldots\ldots\ldots \quad (4)$$

where H'_O = HNO_3 molarity in the solvent phase containing thorium and nitric acid,

H_O = HNO_3 molarity in the solvent phase containing nitric acid only,

Th_O = thorium molarity in the solvent phase, and

T_O = concentration of TBP (moles per litre).

The value of H_O is obtained by using the relationship between the nitric acid distribution coefficient (D_{HNO_3}) and the aqueous nitric acid concentration (H_A) as given by Carleson (1958) and plotted in Figure 4.

$$D_{HNO_3} = \frac{[HNO_3]_O}{[HNO_3]_A} = \frac{H_O}{H_A} \quad .$$

$$\text{Hence, } H_O = D_{HNO_3} \cdot H_A \quad . \quad \ldots\ldots\ldots\ldots\ldots\ldots \quad (5)$$

H_O has been plotted against H_A in Figure 5.

Substitution of T_O = 1.55 M (that is, 40 per cent. TBP) in Equation 4 gives a linear relationship between $\frac{H'_O}{H_O}$ and Th_O, which is shown graphically in Figure 6.

Figures 3, 5, and 6 may then be used to obtain points representing an equilibrium relationship between components.

For example, if Th_A = 0.100 g/l (4.31×10^{-4} M) and H_A = 1 M,

then, from Figure 3, Th_O = 0.120 g/l (5.17×10^{-4} M) .

From Figure 6, $\frac{H'_O}{H_O} = 1$.

Thus $H'_O = H_O$,

and, from Figure 5, $H'_O = H_O = 2.8 \times 10^{-1}$ M .

One point of the equilibrium relationship is:

Th_O = 0.120 g/l	Th_A = 0.100 g/l
H'_O = 0.28 M	H_A = 1 M

Further points may be calculated in the same manner and plotted as shown in Figures 7 and 8.

Siddall (1956) has given an equilibrium diagram for the system $Th(NO_3)_4 - HNO_3 - 42.5\%$ TBP – Ultrasene at higher thorium concentrations. This information, for aqueous nitric acid concentrations of 0.5 M and O M,is shown on Figures 3 and 7, the extrapolated values being used in the calculations.

5. FLOWSHEET DESIGN

By using the calculation methods outlined in Section 4.1 and the equilibrium data developed in Section 4.2, a flowsheet to permit recovery of the actinides from H.T.G.C.R. fuel may be designed. Permissible losses for the various components, however, must first be defined. These losses, which represent a 0.1 per cent. loss of U and Th to each raffinate stream, are:

0.2% loss for Th.
0.2% loss for U.

A decontamination factor of 10^3 for actinides from fission products and beryllium is also required.

5.1 The Feed Solution

The solution resulting from the nitric acid leaching of fuel (see Table 2) may, if required, be adjusted before the solvent extraction operations by:

(1) Concentration or dilution.

(2) Addition of nitric acid or removal of nitric acid by steam stripping.

Factors which influence the decision to make a feed adjustment are as follows:

(1) High concentrations assist in achieving high recoveries. The upper concentration limits are set by beryllium, which is present at a much higher molarity than other components.

(2) High concentrations minimize waste volumes and waste evaporation requirements.

(3) Low acidity in this system enhances thorium and uranium extraction (Farrell, unpublished work). Greatest D values are obtained with near neutral solutions.

(4) Low acidity results in hydrolysis of metal ions to form a precipitate. This is especially true during extraction when nitric acid may be strongly extracted into the organic phase.

(5) High acidity may result in extraction of nitric acid into the organic phase to such an extent as to produce a third phase (Ewing etal. 1954).

Factors (1) and (2) led to a decision to use a feed solution with the highest practical concentration. A beryllium molarity of 2.5 M was selected, as higher concentrations tend to be viscous and unstable (Ryan, unpublished work).

Factors (3) and (4) indicated that the lowest acidity which could be used without introducing hydrolysis problems should be selected. Preliminary work indicated that this acidity would be in the region of 0.5 to 1 M HNO_3.

The solution obtained from the nitric acid leach step (Table 2) is steam stripped to remove excess nitric acid giving the adjusted feed solution compositions shown in Table 3.

5.2 The Split Contactor

Calculations were made, which confirmed that a third phase problem would occur if such a feed solution was used in a conventional solvent extraction flowsheet (Figure 9a). This is due to the extremely high salting strength of the feed solution, which produces very high distribution coefficients for all extractable species. Some scouting experiments also verified this, and it was found that the acidity at the feed point quickly built up to such an extent that nitric acid fumes were given off. Severe third phase problems occurred.

It was therefore necessary to develop a "split contactor" concept in the extraction-scrub sections of the flowsheet. In such a system, the scrub solution raffinate is not combined with the feed solution at the feed point as in the conventional flowsheet, but is rejected separately (Figure 9b). High acid concentrations therefore cannot build up at the feed point.

The unscrubbed extract is fed to an intermediate point of the scrub section, the top half of this section providing the scrubbing required and a second solvent stream being used in the lower half to control actinide losses. This type of flowsheet allows close process control, as beryllium decontamination is provided by the extraction section, fission product decontamination is provided by the upper half of the scrub section, and losses are controlled by conditions in the lower half. The nitric acid content of the extract can be minimized, producing a scrubbed extract ideally suited for the stripping section.

A scrub solution of water, nitric acid, or inorganic nitrate could be used in the design of the flowsheet. Water was chosen for the initial calculations. The salting effect is thus wholly supplied by the nitric acid present in the unscrubbed extract. This has the advantage that it produces a non-salt raffinate which can be concentrated by evaporation.

Final selection of a scrub solution must await experimental work, particularly as no data are available to permit an estimate of the decontamination factors from fission products which can be obtained.

5.3 The Flowsheet

Using a feed solution similar to that given in Table 3, but omitting the plutonium, and adopting the split contactor concept, the preliminary calculations to enable a flowsheet to be designed have been made. Appendices 1 and 2 contain examples of these calculations and Figure 10 summarizes the results, which apply to a feed solution having the following composition:

Be – 2.5 M	Th – 0.125 M
HNO_3 – 0.5 M	U – 0.8 g/l

6. CONCLUSIONS

By using incomplete and extrapolated equilibrium data, coupled with simplified calculational methods, a preliminary flowsheet to permit the recovery and decontamination of actinides in spent H.T.G.C.R. fuel has been developed. This flowsheet incorporates a novel "split contactor" concept, which is necessary because of the unusually high salting strength of the feed solution. The calculation method presented has led to a flowsheet which has been the starting point of experimental work. This flowsheet is to be checked at full levels of activity.

7. ACKNOWLEDGEMENT

The authors wish to record their appreciation of the valuable guidance given by Dr. R.C. Cairns, Head of Fuel Cycle Development Section, who was also responsible for initiating the work.

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* Referred to in Appendix 2.

TABLE 1

QUANTITIES OF MATERIALS CONTAINED IN AN EQUILIBRIUM CORE OF THE H.T.G.C.R, REFERENCE DESIGN

Material	Quantity in Equilibrium Core (kg)
$Th^{232}O_2$	5,140
$Pa^{233}O_2$	21.0
$U^{233}O_2$	145
$U^{234}O_2$	49.5
$U^{235}O_2$	12.2
$U^{236}O_2$	4.2
$Pu^{239}O_2$	0.2
$Pu^{240}O_2$	0.4
$Pu^{241}O_2$	5.3
$Pu^{242}O_2$	34.2
BeO	56,200
Fission Products *	8.2×10^7 curies

*After 120 days cooling

TABLE 2

COMPOSITION OF SOLUTION RESULTING FROM NITRIC ACID LEACH

Leach Conditions:

13 M HNO_3. 0.05 M F'. 0.1 M Al^{3+}.
Temperature 115°. Mechanical agitation. Particle size −70 +100 BSS.

Component	Concentration
Be	20.55 g/l
U	1.02 g/l
Pu	0.178 g/l
Th	22.9 g/l
HNO_3 (free acid)	8.0 M
Fission products	412 curies/litre

TABLE 3

SOLVENT EXTRACTION FEED SOLUTION

Component	Concentration
Be	22.5 g/l (2.5 M)
U	1.13 g/l
Pu	0.20 g/l
Th	25.3 g/l
HNO_3 (free acid)	0.5 – 1 M
Fission Products	460 curies/litre

APPENDIX 1

CALCULATIONS FOR THE EXTRACTION SECTION

The large excess of nitrate ion not associated with the extractable species permits the assumption that the distribution coefficients for uranium, thorium, and nitric acid remain constant during extraction.

Thorium Calculation

This sample calculation considers thorium only, as the calculation methods for uranium and nitric acid are identical. The complete results of these calculations are tabulated in Appendix 3.

The assumed feed composition is:

Be	–	2.5 M (22.5 g/l)
U	–	0.0034 M (0.8 g/l)
Th	–	0.125 M (29.0 g/l)
HNO_3	–	0.5 M

Other assumptions are:

(1) A loss of 0.1 per cent. thorium and uranium to the raffinate.

(2) A feed-to -solvent flow ratio of 1:0.8.

If:

L is the flow rate of the heavy (aqueous) phase,

V is the flow rate of the light (organic) phase,

x^m_A is the concentration of component A in the heavy phase in stage m, and

y^m_A is the concentration of component A in the light phase in stage m,

then, by simple mass balance, the concentration of thorium in the extract (y^1_{Th}) can be found. The overall mass balance is given in the diagram on the next page.

Equilibrium data may now be used to determine the number of ideal stages required to achieve these conditions.

Unpublished experimental work by Farrell indicates that essentially all the free acid and possibly some acid produced by hydrolysis of $Be(NO_3)_2$ is extracted. In the present calculations, therefore, it is assumed that the free acid content of the aqueous phase in the first stage is zero.

The distribution coefficient (Farrell, unpublished work) for thorium in neutral beryllium nitrate solutions (Be = 2.5 M) is:

$$D_{Th} = 2{,}500 = \frac{y^1_{Th}}{x^1_{Th}} \quad \text{in stage 1.}$$

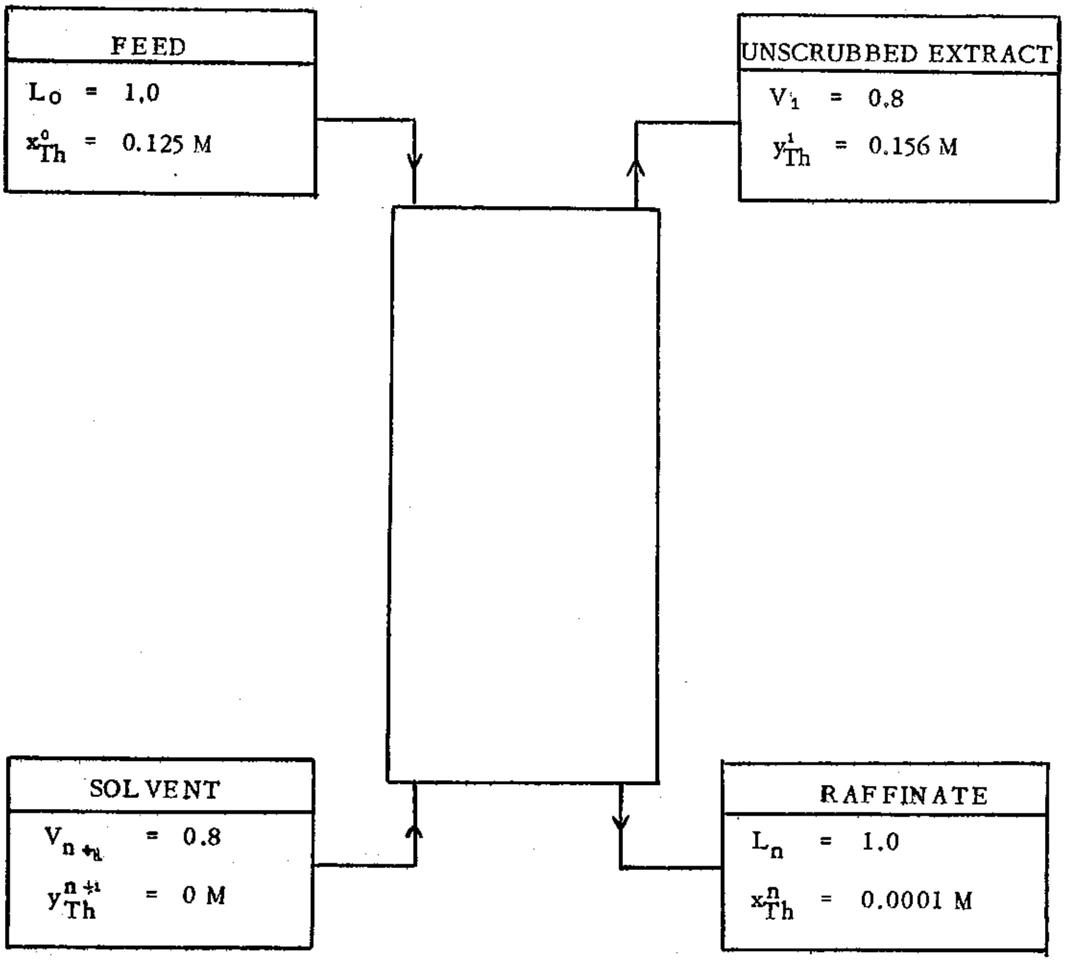
$$\text{Thus } x^1_{Th} = \frac{0.156}{2500} = 0.00006$$

This concentration is lower than that required (x^n_{Th}) to keep losses below 0.1 per cent. Therefore the required conditions can be achieved by the use of one extraction stage.

APPENDIX 1 (continued)

Calculations for uranium also indicate the need for only one ideal extraction stage. The appropriate equilibrium data must be used in each case (Figure 11).

OVERALL THORIUM MASS BALANCE



APPENDIX 2

CALCULATIONS FOR SCRUB AND STRIP SECTIONS

In the scrub section, the distribution coefficients for thorium, uranium, and nitric acid are functions of the thorium nitrate and nitric acid concentrations in the aqueous phase.

Sample calculations are presented for thorium, nitric acid, and uranium over two stages to illustrate the method, and the complete results are given in Figure 12 and Appendix 3.

Thorium and Nitric Acid Calculation

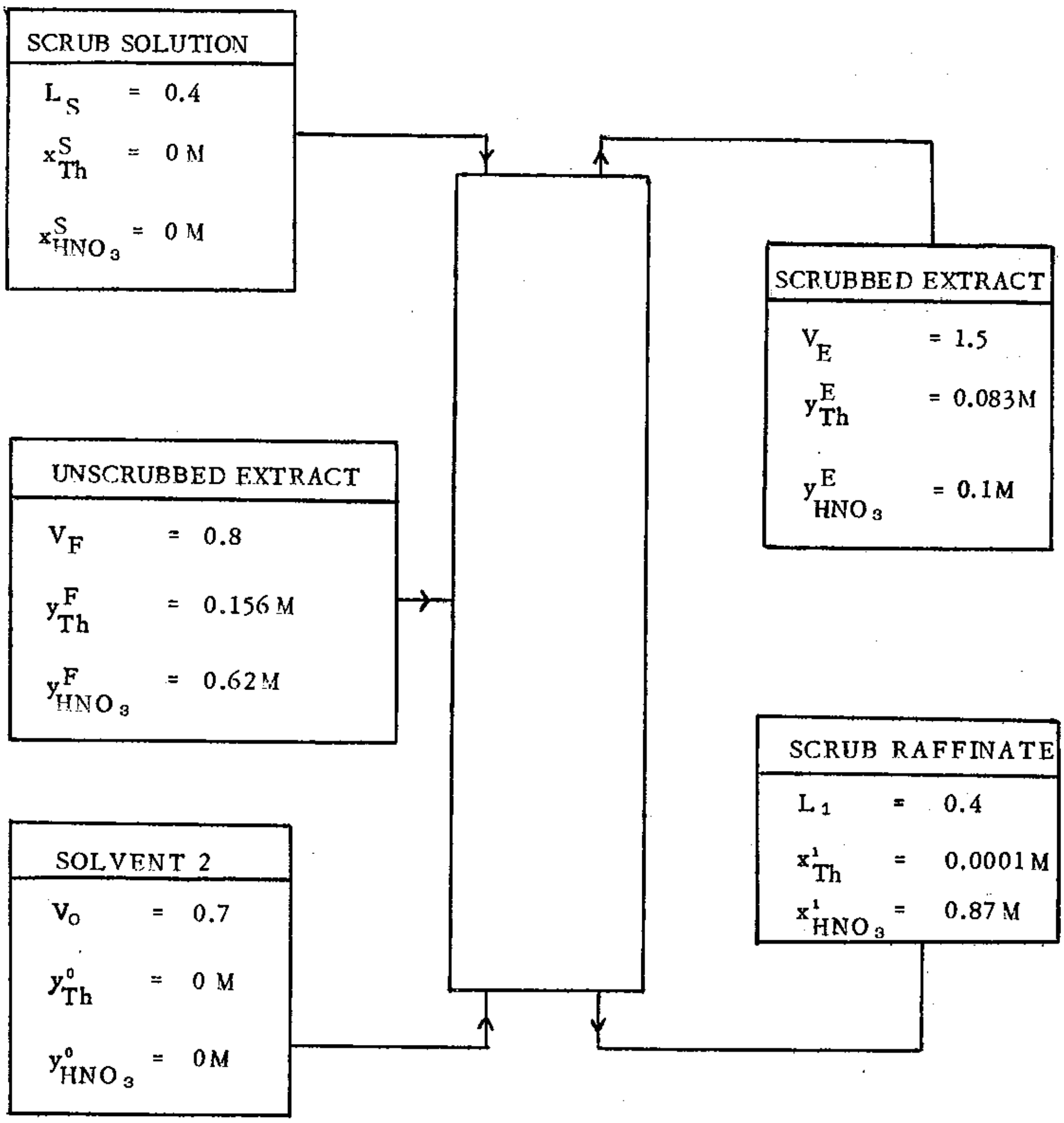
The assumptions made are:

(1) 0.1 per cent. loss of thorium.

(2) Acid concentration in the scrubbed extract is 0.1 M. This should permit easy stripping.

Using the split contactor concept with input flow ratios:

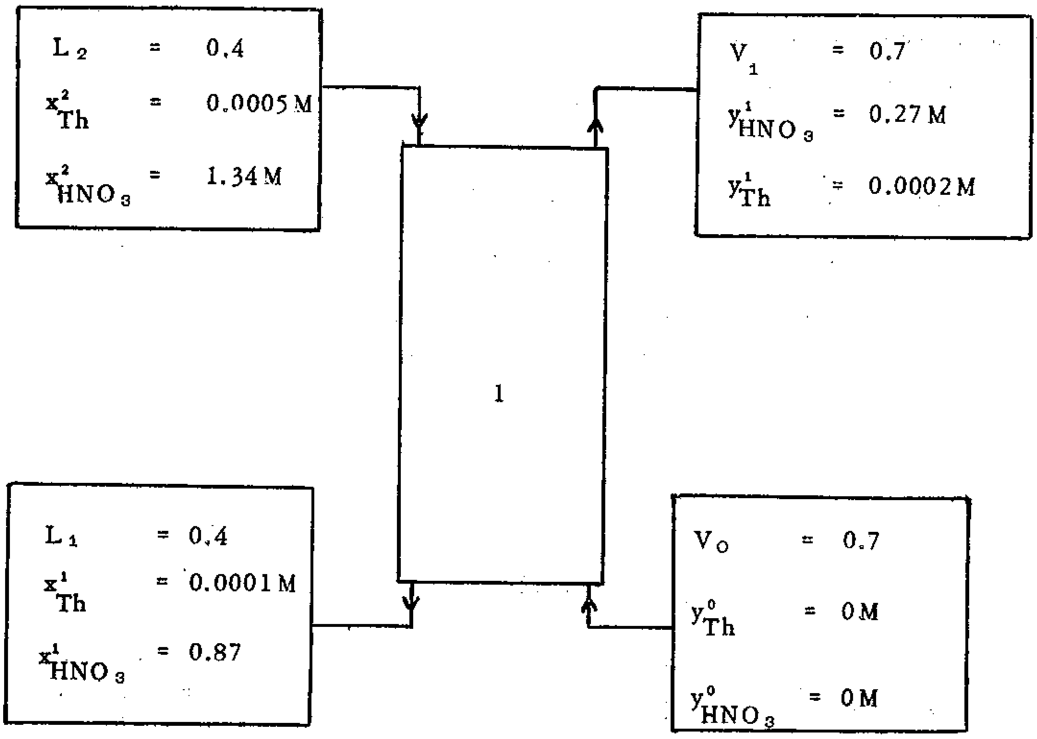
unscrubbed extract: scrub solution: fresh solvent = 0.8: 0.4: 0.7,
an overall mass balance can be made. This is set out in the following diagram, which also serves to define symbols.



APPENDIX 2 (Continued)

The number of ideal stages required to achieve this set of conditions may now be calculated.

If the last scrub stage before rejection of the scrub solution as raffinate is numbered 1, then conditions in this stage may be represented as below:



This information is obtained in the following manner. Using the equilibrium data as in Figure 8:

$$y^1_{Th} = 0.0002\,M$$

$$y^1_{HNO_3} = 0.27\,M \quad ,$$

a material balance can then be made to obtain the values:

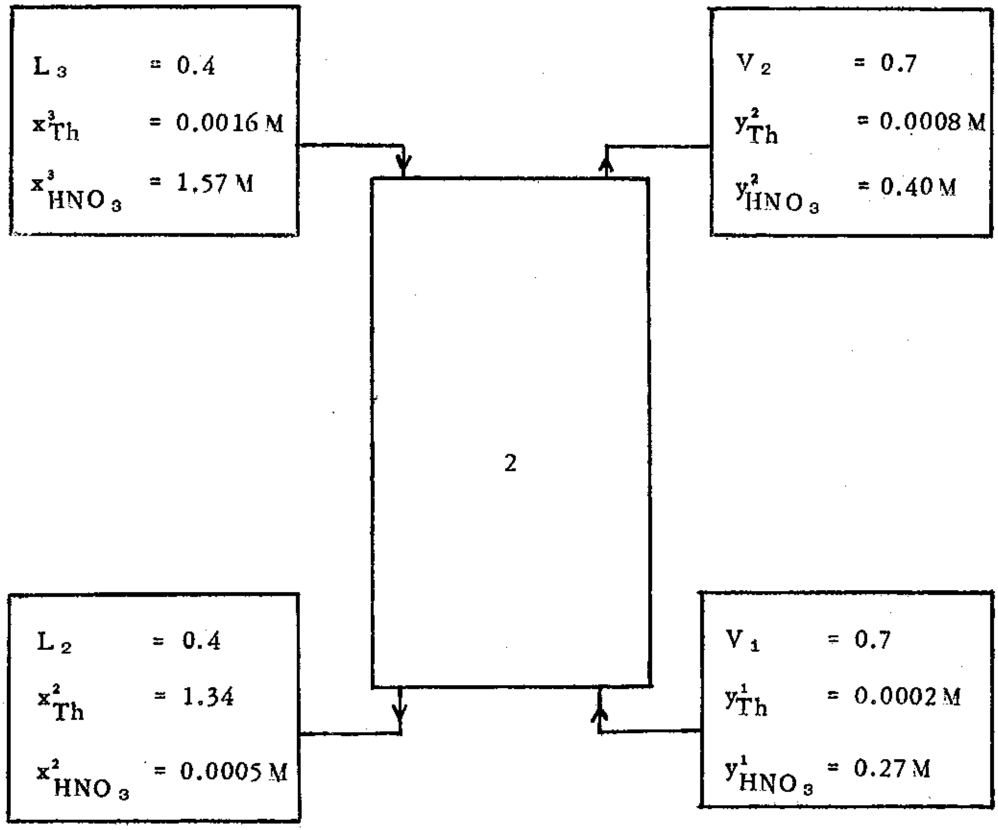
$$x^2_{Th} = 0.0005\,M$$

$$x^2_{HNO_3} = 1.34\,M \quad .$$

This same procedure may be repeated to enable conditions in the next stage to be defined. These are summarized in the diagram on the next page.

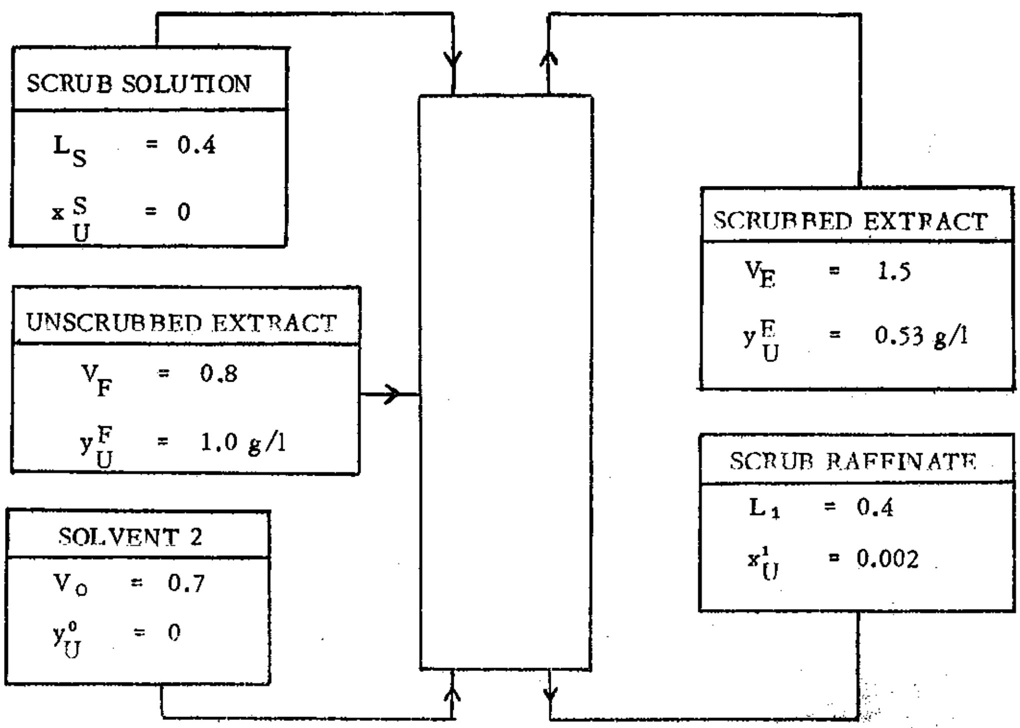
This procedure is simply repeated until a condition equal to that at the feed point is reached. The number of stages required is then known. Figure 12 and Appendix 3 summarize the calculations.

APPENDIX 2 (Continued)



Uranium Calculation

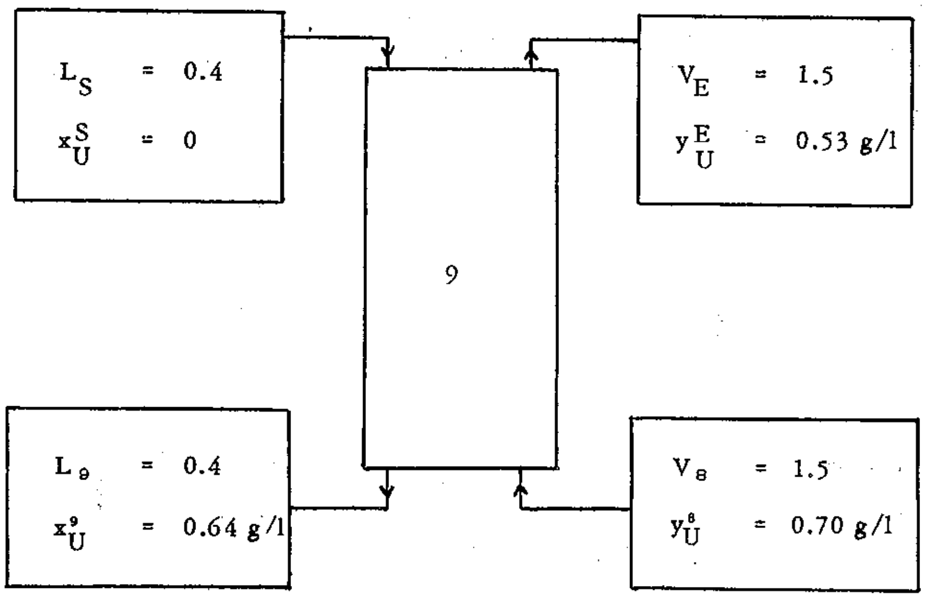
Using the same flowsheet and the assumption that a 0.1 per cent. uranium loss can be tolerated, an overall mass balance for uranium can be made. This information is presented below:



APPENDIX 2 (Continued)

By stage calculations, starting at the first scrub stage (stage 9) and using the stage conditions as calculated previously, the behaviour of uranium can be checked to determine whether these conditions are met.

The behaviour of uranium in stage 9 is summarized in the following diagram:



These values are obtained in the following manner.

From the equilibrium data given in Figure 13 (Blanco 1960a, 1960b),

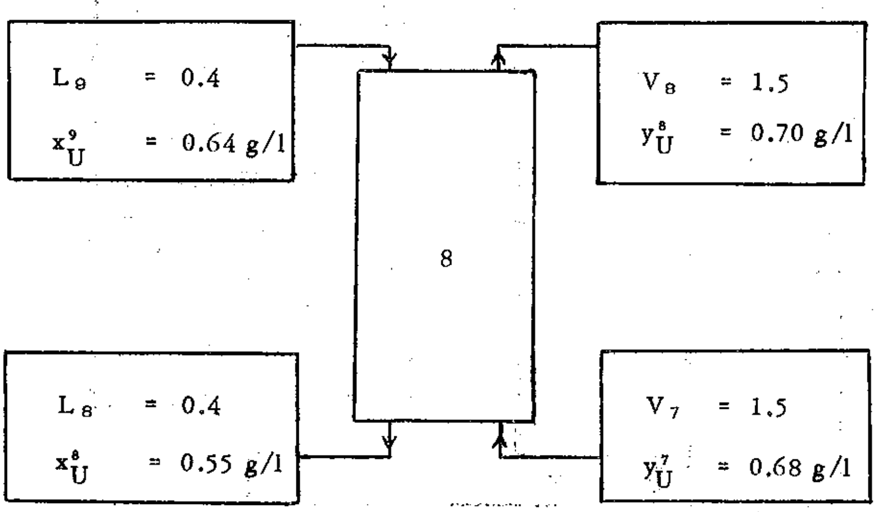
if $\quad y_U^E = 0.53$ g/l, $x_{Th}^9 = 0.065$, and $x_{HNO_3}^9 = 0.28$ M,

then $\quad x_U^9 = 0.64$ g/l

A material balance for uranium then gives

$$y_U^8 = 0.70 \text{ g/l} .$$

This same procedure may be repeated to enable conditions in the next stage to be defined. These are summarized in the following diagram:



APPENDIX 2 (Continued)

By repeating this calculation over the remaining stages, the uranium concentrations throughout the scrub section can be found. The procedure must, of course, be repeated in its entirety until the same extract and raffinate concentrations as the overall mass balance are obtained. Figure 12 and Appendix 3 summarize the results.

Strip Section Calculations

These calculations are made in the same manner as those outlined above for the scrub section, using the relevant equilibrium data (Figure 13 (Blanco 1960a, 1960b) and Figure 14 (Lerner and Petretic 1951). The complete set of results is given in Appendix 3.

APPENDIX 3

SUMMARY OF THE RESULTS OF THE CALCULATIONS

Flow Ratios Used in Calculation

FEED	1.0
SOLVENT 1 (40% TBP)	0.8
SCRUB	0.4
SOLVENT 2 (40% TBP)	0.7
STRIP	10.5
UNSCRUBBED EXTRACT	0.8
SCRUBBED EXTRACT	1.5
SCRUB RAFFINATE	0.4
RAFFINATE	1.0
ORGANIC RAFFINATE	1.5
PRODUCT	10.5

Stage No.	Section	Concentrations					
		Light Phase			Heavy Phase		
		U (g/l)	Th (M)	HNO_3 (M)	U (g/l)	Th (M)	HNO_3 (M)
SOLVENT 1	EXTRACTION	0	0	0			
1	EXTRACTION	1.0	0.156	0.62	0.0002	0.00008	0.0
FEED	EXTRACTION				0.8	0.125	0.5
SOLVENT 2	SCRUB	0	0	0			
1	SCRUB	0.006	0.0002	0.27	0.0015	0.0001	0.87
2	SCRUB	0.006	0.0008	0.40	0.0006	0.0005	1.34
3	SCRUB	0.006	0.0035	0.46	0.0006	0.0016	1.57
4	SCRUB	0.008	0.013	0.46	0.0008	0.0063	1.68
5	SCRUB	0.04	0.05	0.41	0.004	0.0229	1.68
UNSCRUBBED EXTRACT	SCRUB	1.0	0.156	0.62			
6	SCRUB	0.57	0.13	0.36	0.06	0.089	1.59
7	SCRUB	0.68	0.13	0.20	0.13	0.177	0.99
8	SCRUB	0.70	0.10	0.17	0.55	0.177	0.39
9	SCRUB	0.53	0.050	0.13	0.64	0.065	0.28
SCRUB	SCRUB				0	0	0
EXTRACT	STRIP	0.53	0.050	0.13			
1	STRIP	0.000008	0.0001	0.0	0.124	0.031	0.031
STRIP	STRIP				0	0	0

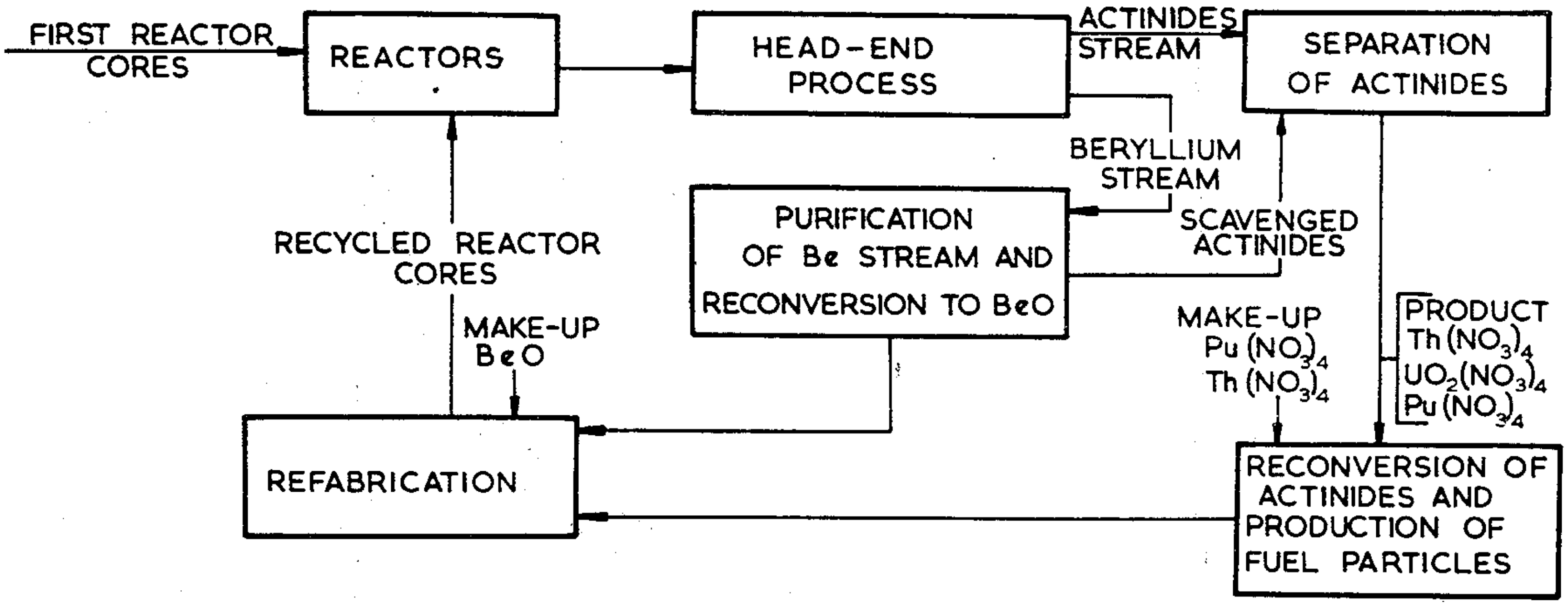


FIGURE I H.T.G.C.R. FUEL CYCLE JULY, 1964

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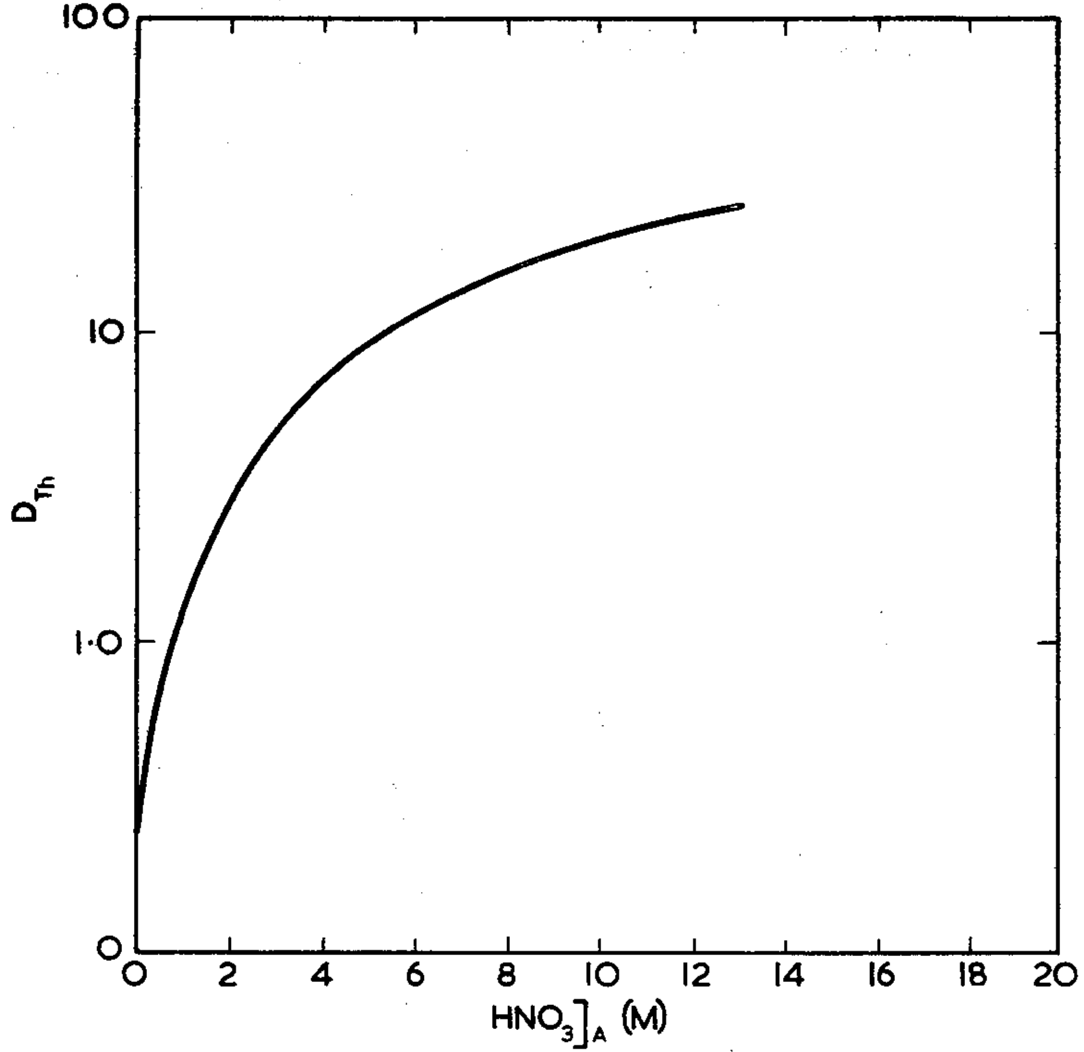

FIGURE 2 VARIATION OF THE DISTRIBUTION COEFFICIENT OF THORIUM NITRATE INTO 40% TBP WITH INCREASING NITRIC ACID MOLARITY (Siddall 1958)

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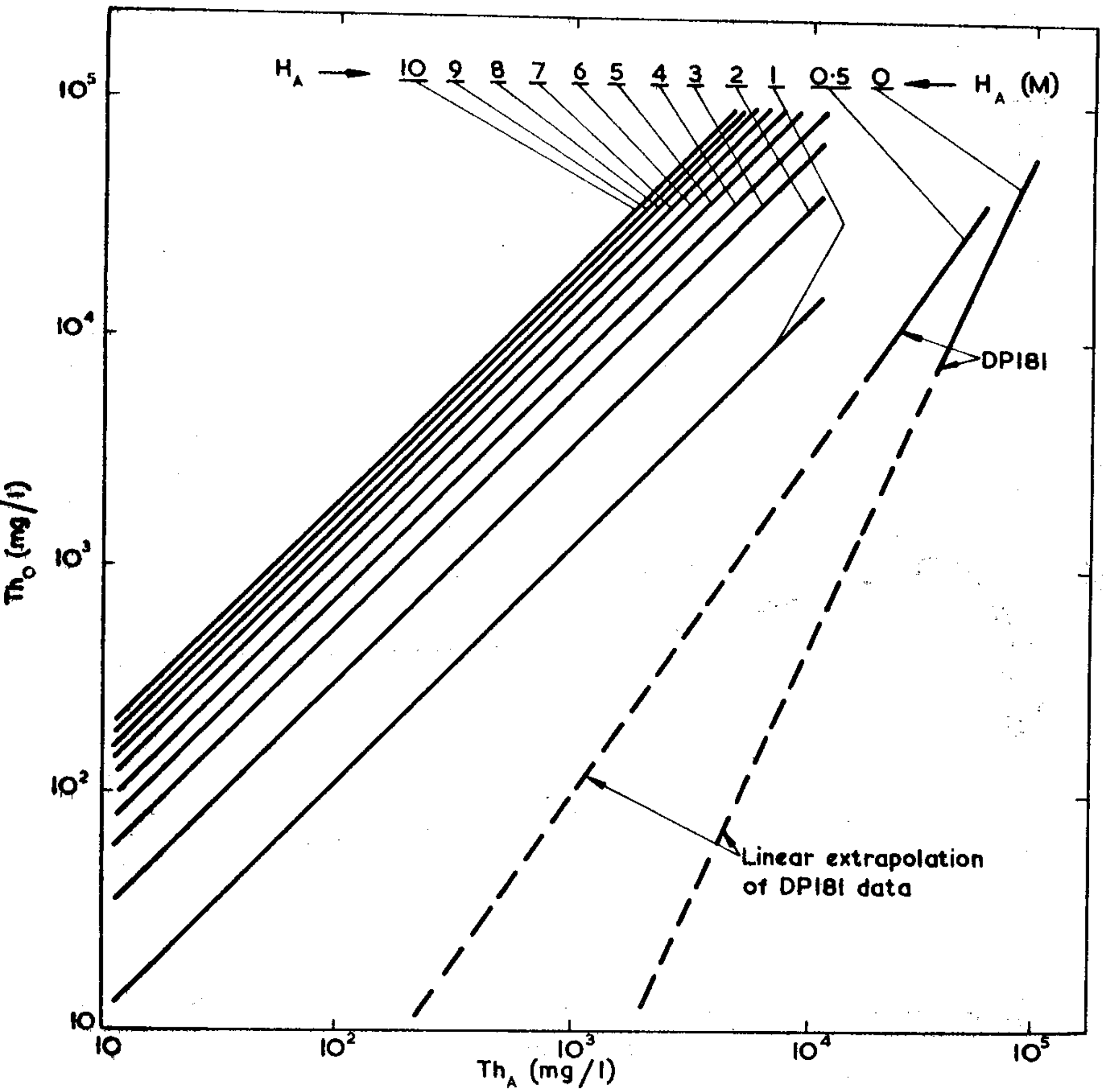

URE 3 EQUILIBRIUM DATA FOR $Th(NO_3)_4$ EXTRACTION INTO 40% TBP AT DIFFERENT AQUEOUS PHASE ACIDITIES

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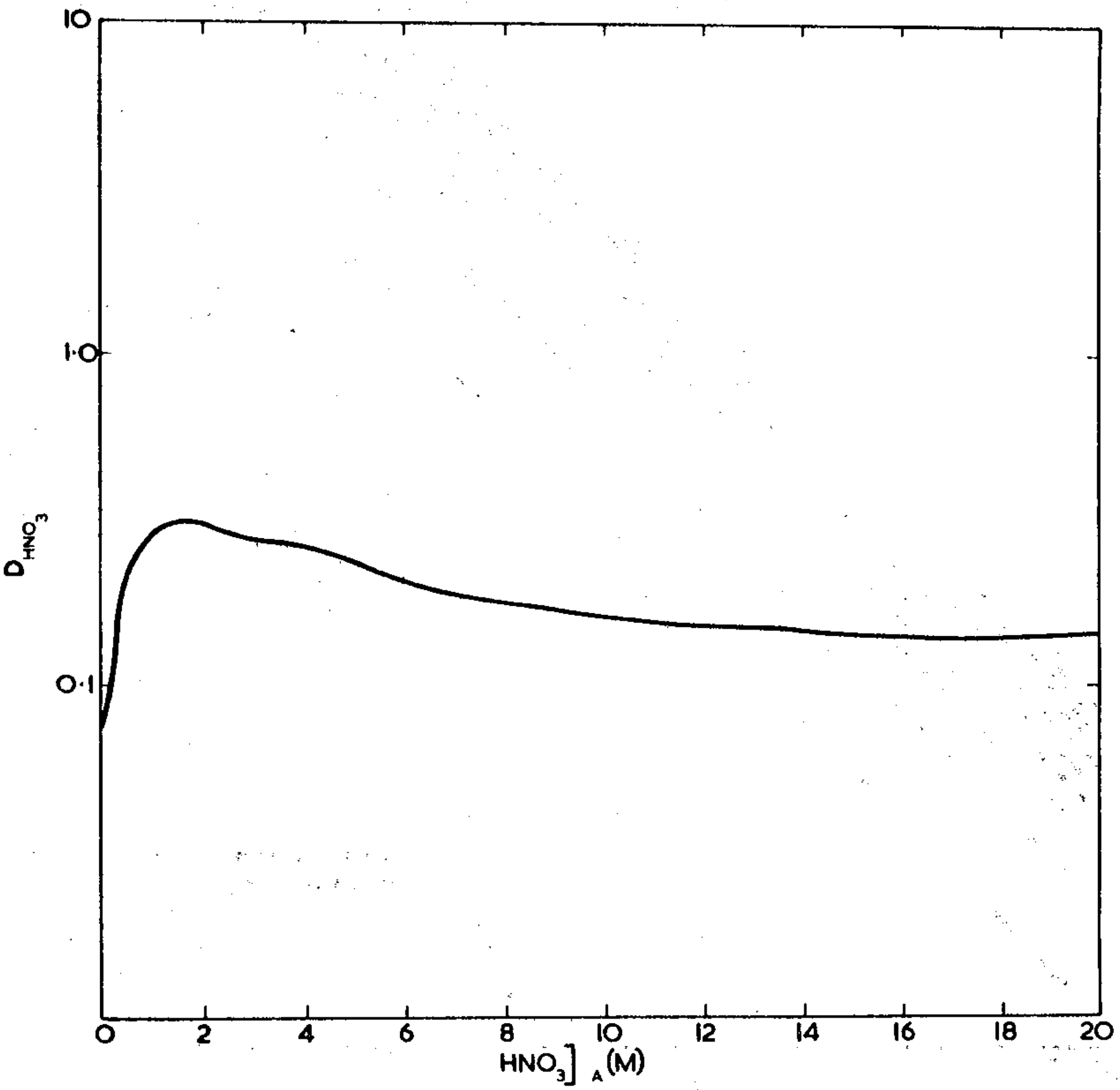

FIGURE 4 VARIATION OF THE DISTRIBUTION COEFFICIENT OF NITRIC ACID INTO 40% TBP WITH INCREASING NITRIC ACID MOLARITY (Carleson 1958)

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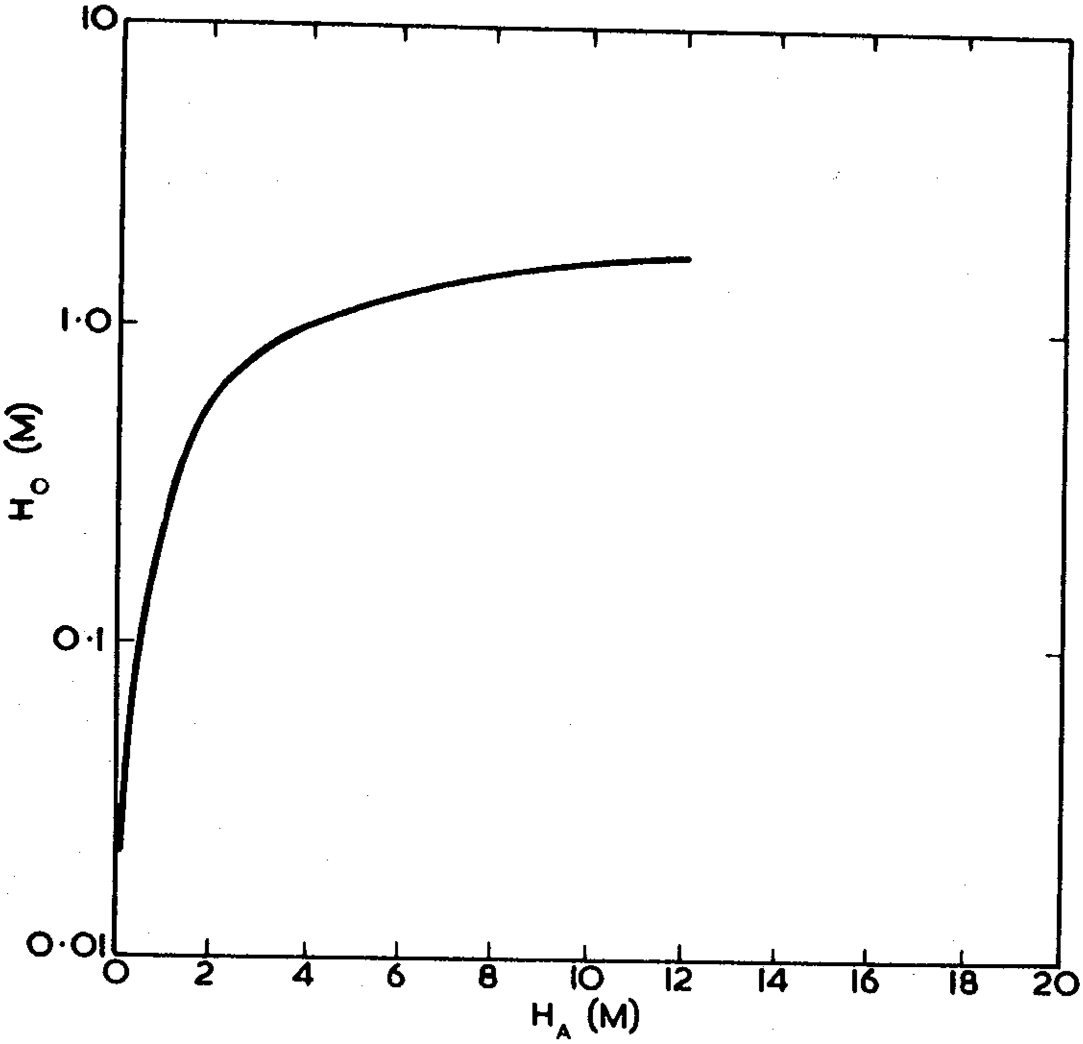

IGURE 5 EQUILIBRIUM DATA FOR THE EXTRACTION OF NITRIC ACID INTO 40% TBP FROM A NITRIC ACID SOLUTION AT VARIOUS MOLARITIES

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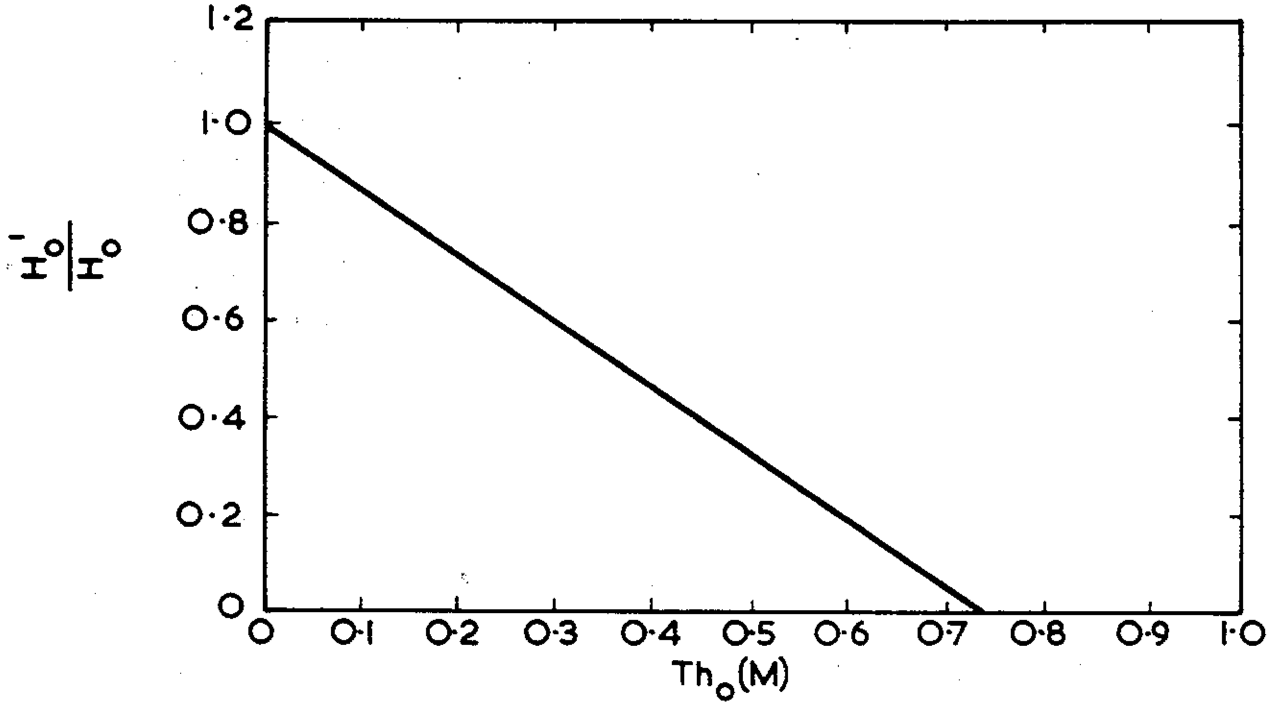

FIGURE 6 GRAPHICAL REPRESENTATION OF THE EQUATION

$$\frac{H_o'}{H_o} = 1 - \frac{2Th_o}{T]_o}$$

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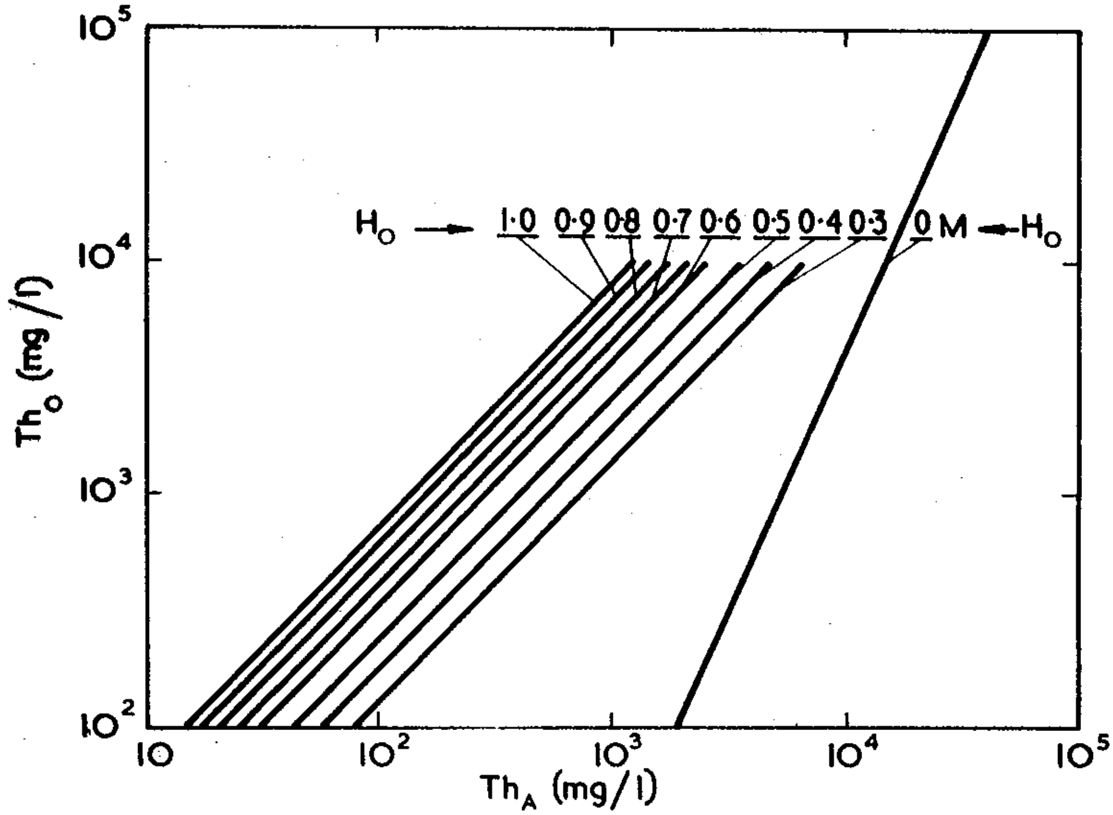

FIGURE 7 EQUILIBRIUM DATA FOR THE EXTRACTION OF $Th(NO_3)_4$ INTO 40% TBP AT DIFFERENT ORGANIC PHASE ACIDITIES

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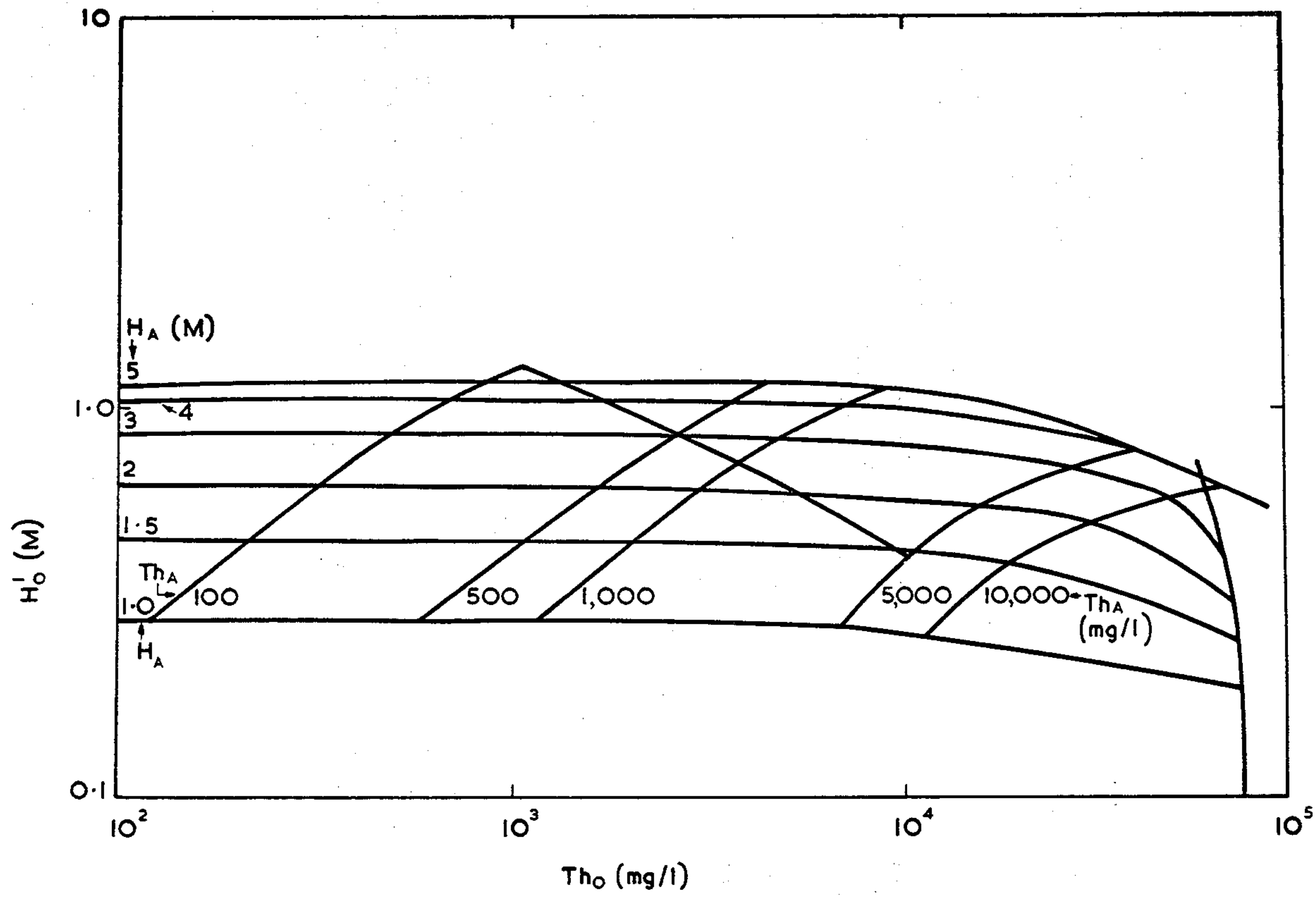

FIGURE 8 EQUILIBRIUM DATA FOR THE SYSTEM THORIUM — NITRIC ACID — 40% TBP IN OMS

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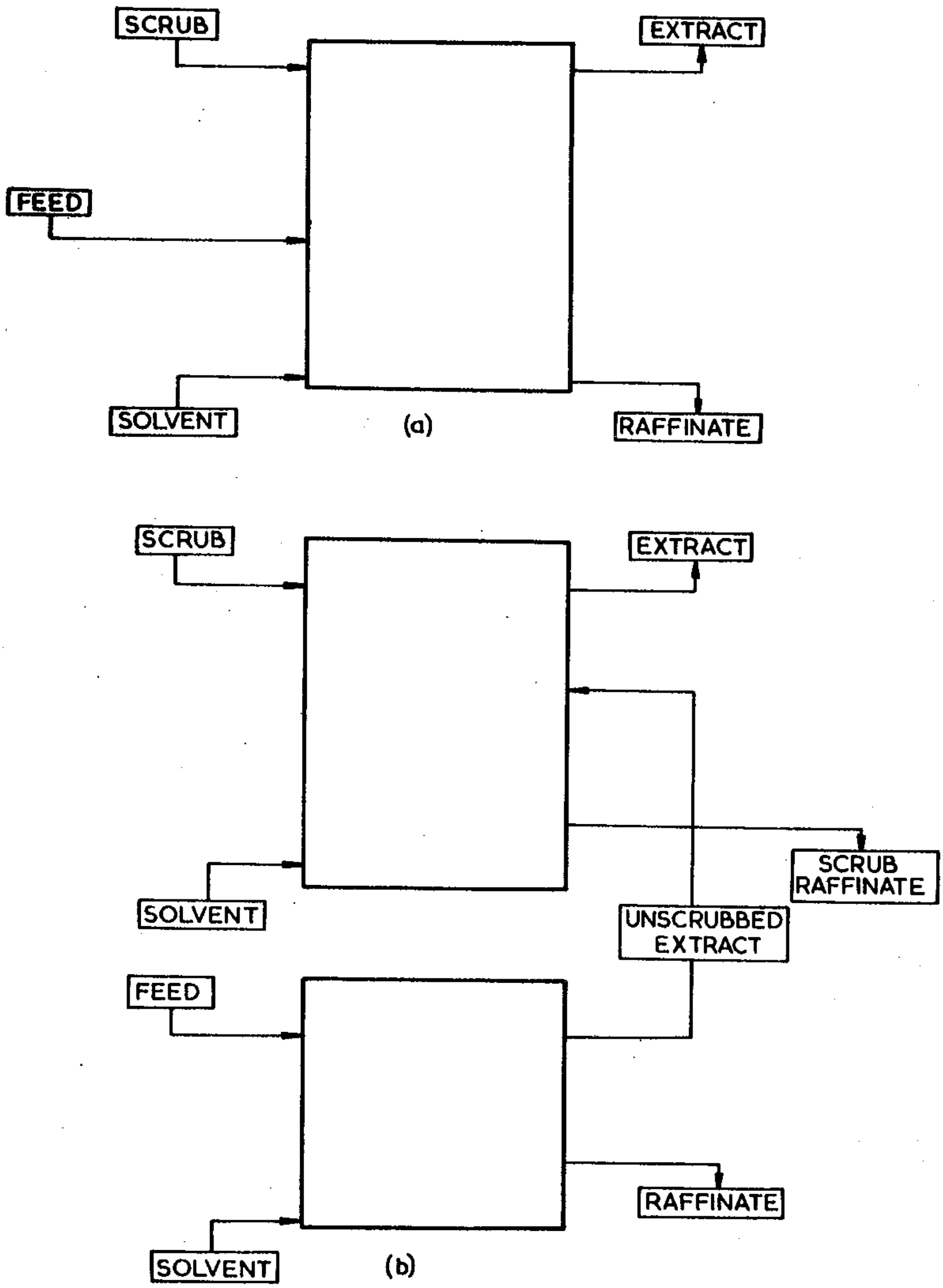

FIGURE 9 STANDARD AND SPLIT CONTACTOR CONCEPTS
(a) Standard Contactor (b) Split Contactor

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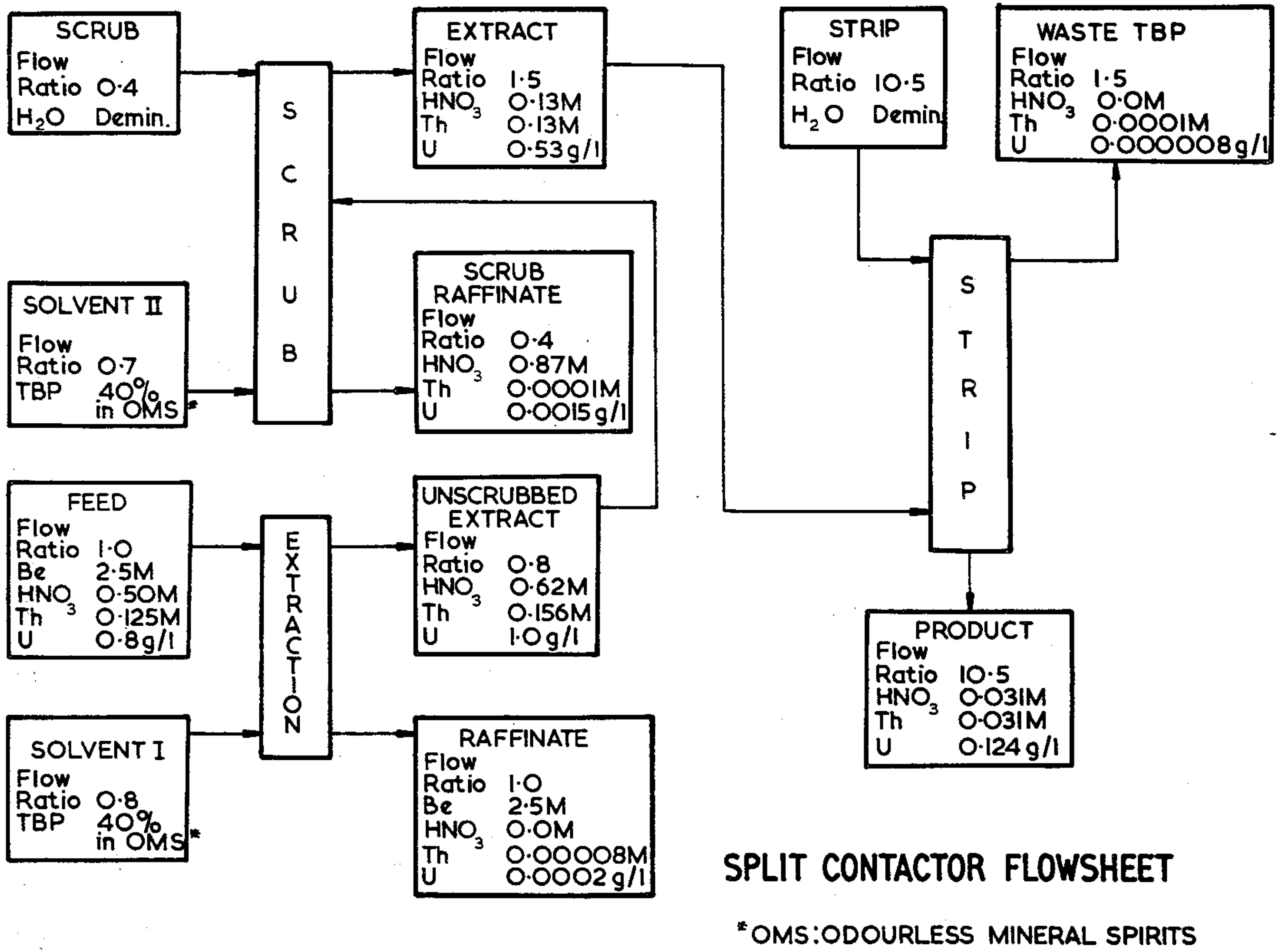

FIGURE 10 END-STREAM CONCENTRATIONS FROM FLOWSHEET DESIGN

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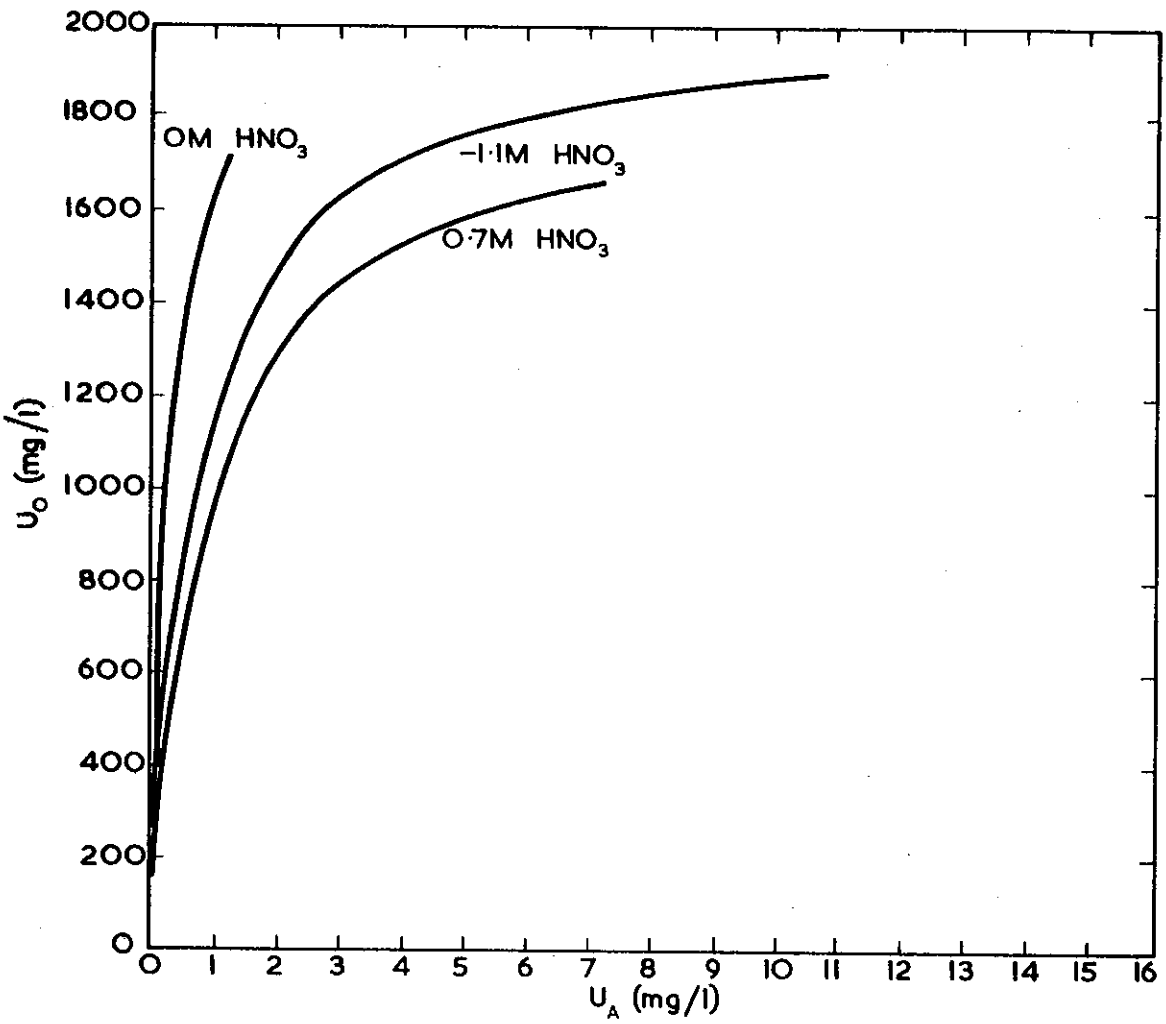

FIGURE II EQUILIBRIUM DATA FOR THE EXTRACTION OF URANIUM FROM A $Be(NO_3)_2 - UO_2(NO_3)_2 - Th(NO_3)_4 - HNO_3$ SYSTEM (Farrell—Private Communication)

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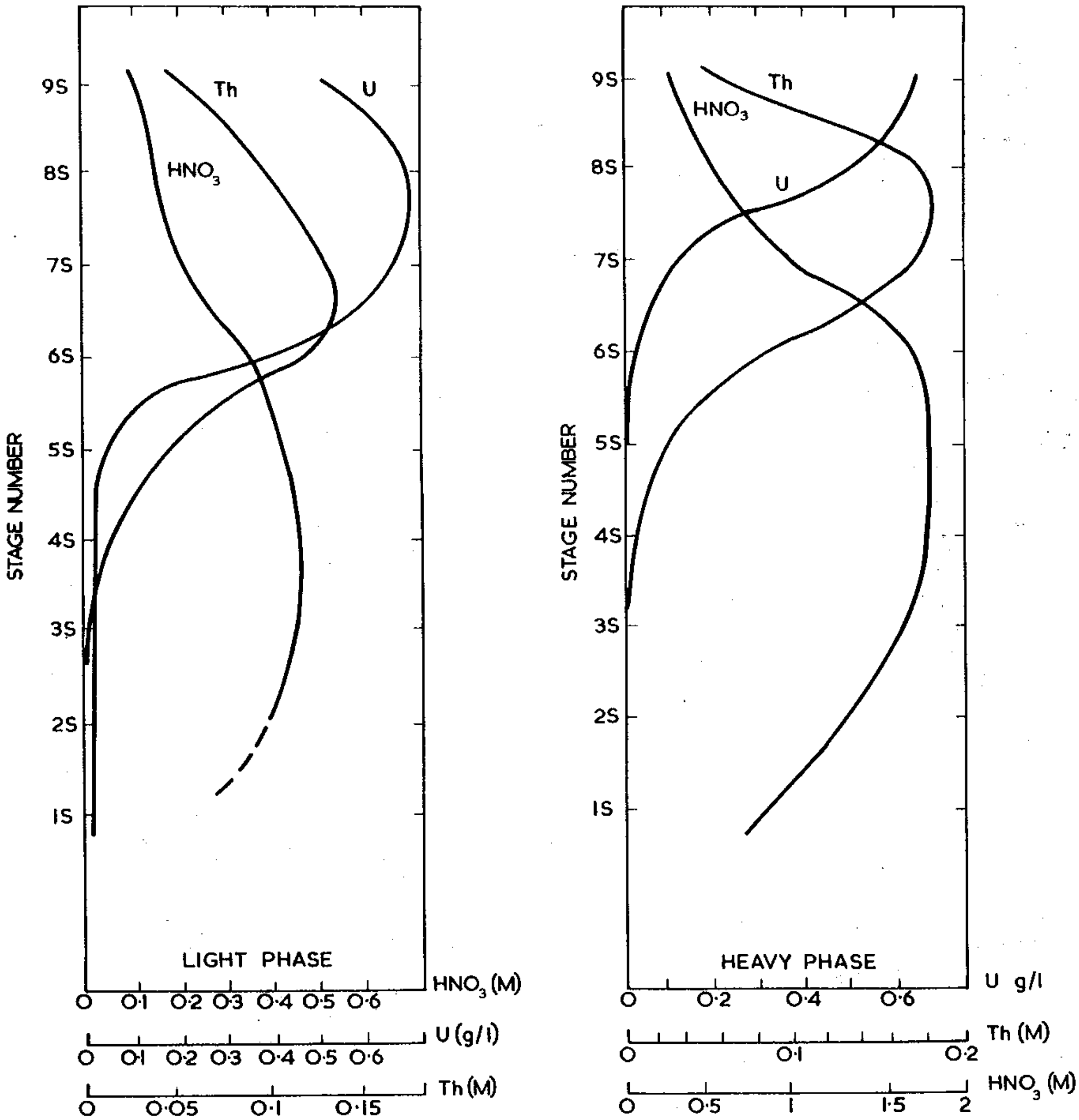

FIGURE 12 DIAGRAM SHOWING THE STAGE CONCENTRATIONS OF U, Th, AND HNO_3 IN THE SCRUB CONTACTOR

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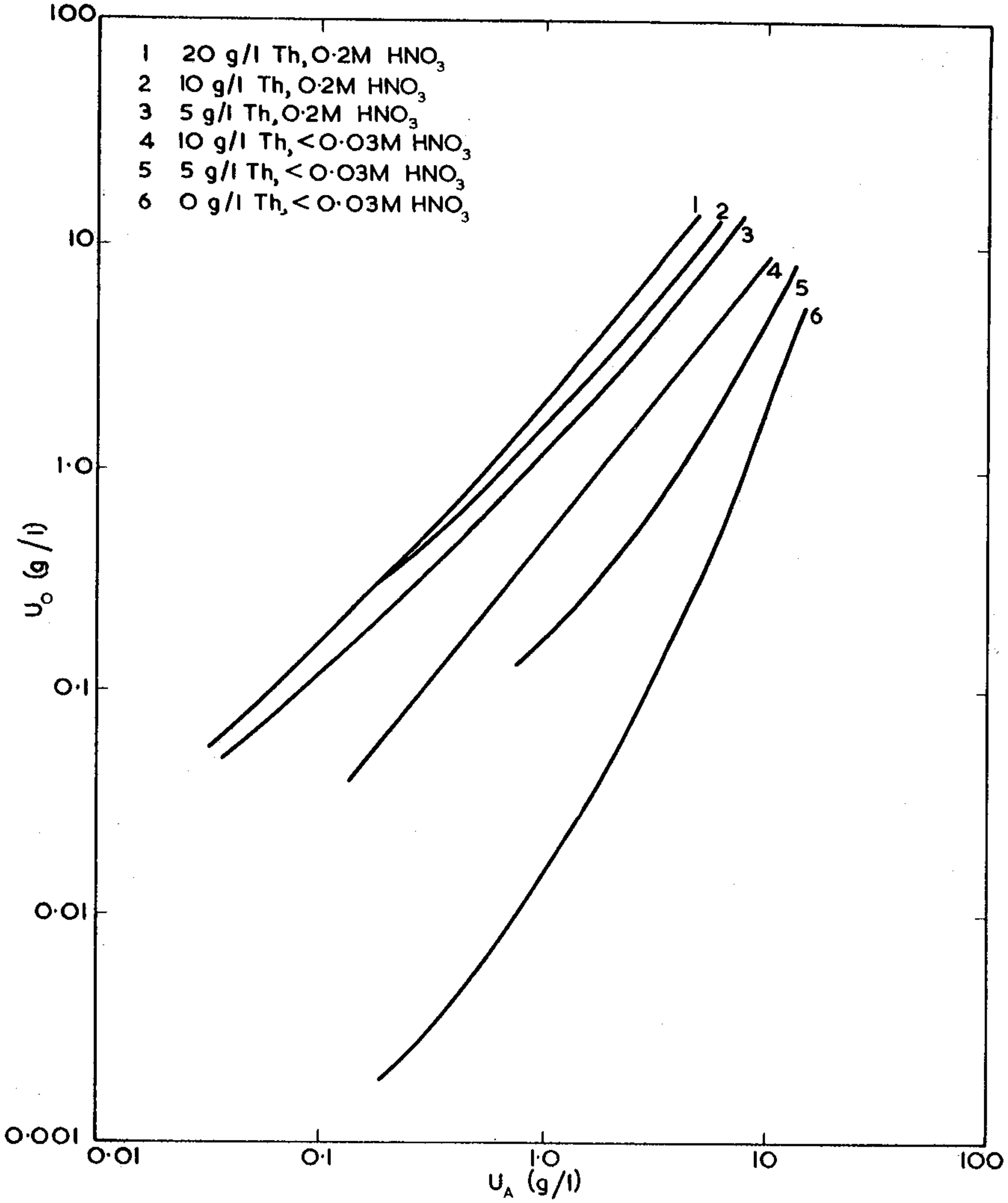

IGURE 13 EQUILIBRIUM DATA FOR THE TBP EXTRACTION OF URANIUM FROM A $UO_2(NO_3)_2 - HNO_3 - Th(NO_3)_4$ SYSTEM (Blanco 1960a, 1960b)

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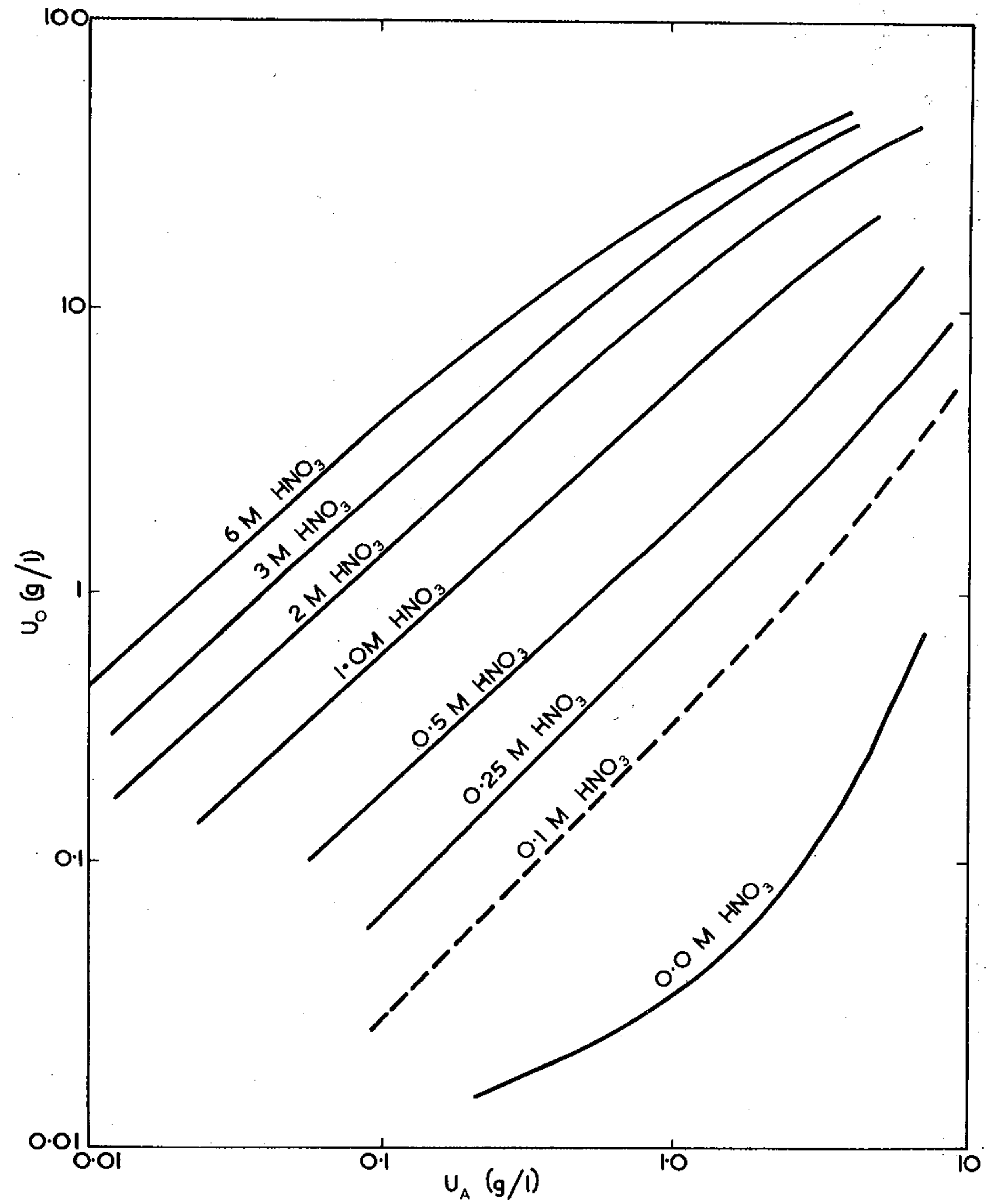

FIGURE 14 EQUILIBRIUM DATA FOR THE EXTRACTION OF URANIUM FROM HNO_3 SOLUTION BY TBP (Lerner and Petretic 1951)

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