SUGGESTED TYPES OF REPROCESSING PLANT

FOR POWER REACTOR FUEL ELEMENTS, IN ARGENTINA

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1. INTRODUCTION

A power reactor of about 300 MW(E) capacity is due to be operating in Argentina by about 1972. The type of reactor is at present under discussion, but presumably full irradiation of the first fuel charge will be achieved about 1975. It is, therefore, advisable to have facilities for reprocessing, either in Argentina or overseas, at about that date. It is possible that a second reactor will be built a few years later, but it is unlikely that a reprocessing plant will be required to treat more fuel elements than the output from these two reactors.

It is probable that the reactors will be fuelled with natural uranium metal, e.g. a Magnox" gas-cooled type, or with slightly enriched uranium oxide, e.g. on advanced gas-cooled or a water-cooled type. The fuel of the first category is likely to reach an average irradiation of about 4,000 MWD/T and the second category about 20,000MWb/T. A 300 MW(E) reactor is approximately 1,000 MW(H), so, the reprocessing capacity requirements will be about four days per ton, or 90 tons/yr for a natural uranium metal type or about 20 days per ton or 18 tons/yr for a slightly enriched oxide type. Reprocessing capacity is therefore required to meet one of three sets of circumstances:-

(a), About 40 tons/yr slightly enriched fuel,

(b), About 200 tons/yr natural fuel,

(c), About 120 tons/yr of mixed fuel.

This report considers only (a) and (b).

As with most chemical engineering operations, the cost per ton decreases markedly with increasing throughput. If an entirely conventional plant were built in Argentina, it would not, therefore, be able to compete economically with the much larger plants which exist in the world. At Windscale in the U.K., for example, a 2,000 ton/yr plant is available, with surplus capacity which is being used to reprocess the fuel elements

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from other countries, cheaper than they could reprocess themselves. The cost of transport, whilst not low, is not a controlling factor.
Considering only conventional plant, therefore, Argentina is faced with the prospect of either (a), having fuel elements reprocessed overseas at economic rates, and losing reprocessing experience or (b), reprocessing in Argentina at several times the world market cost. The repercussions of policy (b) are that the plutonium extracted will probably cost more than its value as a nuclear fuel for thermal reactors (say $7 to $10 per gram) and the loss of interest over many years on the capital investment of a large plutonium stock-pile for a fast reactor, is prohibitively expensive(1).

One alternative exists, which is to examine carefully the possibility of building in Argentina a reprocessing plant to slightly unconventional principles.
The writer is of the opinion that the many years of development devoted to fluoride volatility processes, liquid metal processes, and high temperature "slagging" processes, have only confirmed that these have difficult engineering problems and they are not cheaper, or not significantly cheaper, than the conventional TBP solvent extraction processes.
Alternative solvents have been used, but, at best, they can only just compete with TBP, and a very large amount of data and experience are available with TBP which is of inestimable value to those developing and designing a reprocessing plant at this date. Consequently, mainly TBP extraction processes will be considered in this report.
The writer believes that there are many reasons why existing reprocessing plants based upon TBP have been more expensive to build than is necessary. Some of these reasons are listed below:-

1. Designers have sometimes adopted layouts which are appropriate to a non-nuclear plant with insufficient regard for the fact that space is highly expensive when it is confined behind concrete shielding walls over 1 metre thick.
2. Some plants must be guaranteed to meet important production programmes without delays due to maintenance, so expensive "maintenance free" units have been used and spare sections of plant have been installed for emergencies, which, in practive, have not arisen.

3. "Remote maintenance" plants, where all items can be removed from the active cell and replaced by new ones, require a great deal of specialised equipment for lifting vessels into position, for connecting process lines, and for observing the operations remotely (2). In addition, very stringent equipment and fabrication tolerances are required, together with considerable excess space for movement, over the process vessels. This design principle is stated by the users in the U.S.A. (2) to be mainly suitable for larger plants with throughputs of between 1 and 5 tons per day.
4. The cost of plants has included high development charges, which are not necessary in future, now that the information is available.
5. The required capacity has sometimes been over-estimated for lack of knowledge of future reactor programmes or in anticipation of commercial orders which have not arisen.
6. A high proportion of the equipment in a reprocessing plant is normally used to purify uranium. When the uranium is present in a highly depleted form arising from the irradiation of natural uranium to 4,000 MWD/T, it is a waste product of zero commercial value.

There is, in the writer's opinion, scope for the building of a reprocessing plant of small size which avoids many of the above faults but is still based upon the well-known TBP (or possibly butex) extraction technology. The object of the present report is to suggest ways in which this might be done so that Argentina might be provided with the cheapest small scale reprocessing plant in the world. Only more detailed development of these ideas and accurate costing of the resulting designs could decide whether, in fact, such a plant could be reduced in cost sufficiently to be competitive with the larger plants of other countries.

2. DESIGN PRINCIPLES

It is suggested that no pre-conceived comprehensive design principle should be adopted, such as "remote maintenance", "direct maintenance", or "no maintenance". Instead, the approach should be more flexible than on other plants, so that the type of maintenance would be decided on an individual basis for each item or section of plant. The main aims of the design would, in general, be as follows:-

1. To make use of existing knowledge and experience throughout the world as far as practicable, and so avoid high development expenditure.
2. To design the plant in size and function only for the work it has to do and not to cater for nebulous additional demands which may never arise. (This report, for example, discusses several possible fuel elements and throughputs, but, when the type of power reactor is known, only those which are appropriate to the chosen reactor system should be considered).
3. To achieve a compact layout, bearing in mind the high cost of shielded space.
4. To avoid spare plant items which may not be required, and which might be installed at a later date instead.
5. To design on the assumption that, with only one or two reactors, it is not necessary to provide for a very rigid production programme, with all that this entails regarding guaranteed reliability of equipment or availability of alternatives. In the case of this plant, an occasional unscheduled interruption would not be a catastrophe, provided the decay pond has an adquate capacity to hold more than its normal load of fuel elements.
6. To avoid wasting valuable reprocessing equipment and effort on the uranium stream, where its enrichment is so low that its value is zero.
7. To avoid sophisticated plants for the treatment of radioactive effluents, with evaporation, nitric acid recovery, and the storage of highly active effluent in a very concentrated form.

   Instead, simple alkaline storage in large mild-steel tanks, underground, would be more appropriate in Argentina.

3. REFERENCE FUEL ELEMENTS

In view of the type of fuel element being unknown at the present time, two main reference fuels will be discussed, as follows:-

1. A natural uranium metal rod, 1 metre in lenght and 2.5 cm diameter, weighing about 10 Kg. and canned in magnox, with longitudinal or radial fins.

2. A 1.5% enriched uranium dioxide fuel rod, 1 metre in length, made from 1 cm long by 1 cm diameter sintered pellets of density about 10.6 g/cc, weighing about 0.8 Kg and canned in stainless tubes, or zircalloy tubes, of about 0.4 mm wall thickness.

4. POSSIBLE FLOWSHEETS

Figure 1 shows a suggested flowsheet for the magnox - natural uranium metal fuel elements. It is based upon decay for about 150 days, chemical decanning, batch dissolving, extraction of plutonium alone, leaving the valueless uranium (and dissolved can) behind with the fission products, and subsequent purification of the plutonium stream in equipment of very low volumetric capacity. This is a highly unconventional TBP process and has not, to the writer's knowledge, been used elsewhere. It would, therefore, require development, particularly with regard to the determination of conditions, under which a reasonably high plutonium distribution coefficient may be obtained when the solvent is nearly saturated with uranium. The conditions shown on the flowsheet are highly speculative, but the flowsheet has been presented mainly to indicate the many advantages which arise later, if this stage of extracting only the plutonium can be implemented in practice.

Figure 2 shows a suggested flowsheet for stainless-steel or zirconium clad uranium oxide fuel elements. It includes decay for 150 days, a "chop and leach" stage for decanning and dissolution, followed by three cycle TBP uranium extraction and purification and four cycle TBP plutonium extraction and purification. This process is small in scale because of the use of enriched oxide fuel at high irradiation, with a small amount of fuel per megawatt of electricity generated.

A more complicated flowsheet for 100 tons per year of irradiated natural uranium, together with 40 tons per year of irradiated enriched uranium oxide, is not presented in this report. The main outline of such a plant could be deduced by examination of the other two proposals. It would be more difficult to make a mixed plant of this type competitive with larger plants.

These flowsheets will be referred to throughout the remainder of this report, where the two possible plants are discussed in more detail.

5. SUGGESTED PLANT FOR 200 T/yr NATURAL IRRADIATED URANIUM FUEL ELEMENTS

5.1 Fuel Decay

150 days decay is adequate for the radioactive iodine to be reduced to a negligible amount. Each reactor will have its own cooling pond, but it is suggested that for the whole of the decay to be done there demands a capacity sufficient for about half a year's production, say 45 tons. The reactor pond may have this capacity, or more, but it may be required to be kept empty for the accommodation of large quantities of fuel elements discharged in a possible emergency. A cooling pond is therefore required at the reprocessing plant, with a capacity of at least 2 x 45 tons, or say 100 tons. This pond should be about 12 M by 12 M and 5 M deep, and it should be immediately adjacent to the dissolver end of the reprocessing plant. It should have 100 clearly defined positions for baskets 1 M square, with a few centimetres between each. Each basket would hold about 100 elements of total weight about 1 ton. The baskets would be constructed preferably of stainless-steel (since mild-steel requires frequent painting), in a very simple manner. Movement of the baskets under water would be by means of a 3 ton overhead crane capable of traversing the whole area of the pond. Lifting chains could be attached to four eyes at the corners of the basket, or possibly only one chain attached to an eye on the end of a vertical post attached to the centre of each basket. This attachment would be by the aid of manually operated but hydraulically powered long tongs. The same tongs could be used for any handling of individual elements within the baskets.

A small bay at one corner of the pond would be used for discharging the baskets of elements from their individual shielded container as they come from the reactor site. This shielded container would weigh perhaps 10 to 20 tons. If the walls were hollow steel, filled with lead, the external dimensions would be about 1.5 M square by 1 M high. It might, therefore, be appropriate to lift the whole container from its lorry into the pond, by means of a fixed 30 ton crane. After removal of the lid, under water, the basket could be lifted out quite simply, and the empty shielded container lifted to the surface for washing and decontamination. The baskets would arrive at the pond

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in their shielded containers at the rate of 200 per year, or less than one per day. Regular transport from the reactors would be required, with perhaps three containers and lorries per reactor.

The pond should not become heavily contaminated, since the corrosion of magnox can be considerably reduced by the use of water which has a chloride ion content below 1 p.p.m. This is obtained, if necessary, from an ion-exchange water purifier attached to the pond. Also, it is not proposed to carry out mechanical decanning operations in the pond. However, some contamination will arise from occasional damaged fuel elements and perhaps as a result of cutting away end fittings or braces of massive Magnox, by means of an under-water guillotine. This guillotine could be located in a small bay of the pond at the separation plant end, connected to the main pond by only a small hole through which the elements were passed. A flow of water could be used from pond to bay, to further restrict the contamination to the bay.

With a relatively uncontaminated pond, the immersion of the shielded transport container described above, becomes practicable, although it would need thorough external washing and decontamination before being transported on the public road again. Also, the hazards to the pond operators, and the number of operators, would be minimised. Another possible advantage of a relatively uncontaminated pond is that the hazard arising from a possible leak of water through the pond wall could be minimised. This would probably allow the pond to be sunk into the ground, which, in some locations, can be considerably cheaper than constructing entirely above ground, where leaks can be detected. If a sunken pond were used, a suitable ground drain would be provided below it, with a monitoring facility to detect leakage.

The pond would be constructed of concrete, perhaps 0.5 M thick, with an impervious layer in the middle and an impervious lining of e.g. bitumen or glazed tiles. No building need be provided over such a pond. Even in a windy location near the sea, the chloride content of the water from sea spray can be kept low, without a building, if the ion-exchange plant is used continuously. A flow of new water is also required, to purge the pond of small amounts of fission product activity and to prevent the growth of algae.

There should be an underwater connection between the pond and the separation plant, i.e. a short tunnel, with an inclined base, so that a basket of elements could be pushed down it by means of tongs, once it had

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been positioned by the crane. The basket would enter a small heavily shielded lift shaft of the separation plant, where it would rest on the flat base of a two ton hoist, ready for lifting up to the dissolver charge tube.

5.2 Decanning and Dissolving

Mechanical decanning has been developed for the large U.K. separation plants and, once the many commissioning problems are overcome, this is slightly cheaper than chemical decanning. However, the high cost of the heavy hydraulic equipment and control devices, operating under water or in a shielded cave, is probably not justified if it has to be allocated over only 200 tons/year as in Argentina. It is possible that simpler and cheaper mechanical decanning equipment could be developed, with more manual assistance, to suit the lower throughput of the Argentina plant, but this is a field where the amount of skilled mechanical engineering required, and the cost of development, is very easily under-estimated. The writer prefers to recommend the more straightforward and more conventional chemical decanning process which is used in other countries. The principal disadvantage of the chemical method is the high cost of storage of the liquid magnesium salt effluent, which cannot be evaporated, and is contaminated with fission products. The loss of uranium and plutonium into the decanning liquid is probably about the same proportion as the loss of small pieces of metal in the mechanical decanning process, i.e. about 0.1%. Besides the proportion of fission product activity associated with the uranium and plutonium, there is a further similar quantity which has diffused into the magnox and become separated from uranium and plutonium.

The scrap magnox metal from a mechanical decanning process would still require to be stored as solid active waste, in a special silo, under water, to minimise the fire hazard, and with adequate ventilation to dilute the hydrogen (from corrosion) below the explosive limit. Eventually, such a silo becomes filled with suspended and precipitated magnesium hydroxide, as a result of corrosion, but it is smaller in volume than the corresponding wet storage tank of the chemical decanning method.

It should also be noted that the large U.K. plants have both involved continuous dissolving of the uranium metal, and there is, therefore, no opportunity of carrying out chemical decanning in the dissolver vessel. Instead, a separate vessel in the highly active cell would have been

necessary for chemical decanning. In the suggested Argentina plant, the throughput and simplicity required are such that batch dissolving is appropriate, and it is possible to use the same dissolver for chemical decanning.

The chemical decanning process is simply the dissolution of the magnox in a suitable chemical reagent which does not dissolve the uranium to any appreciable extent. The quantity of magnesium is usually about 10% of the weight of uranium. Various solutions have been used.

If the extraction of plutonium, without uranium, is successful, by the use of a salting-out agent as proposed in Section 5.3, the possibility of using magnesium for this purpose should be examined. Then, it might be possible to dissolve the whole fuel element, including the can, in one operation, in nitric acid, so avoiding completely a separate decanning process. This has been shown on the flowsheet, although it is not a fully proven process.

The dissolving vessel is located in the highly active cell of the separation plant, i.e. the first cell after the pond, and the one of which the shielded lift shaft forms a part. Since batches of 600 Kg are to be dissolved in the vessel, to a concentration of 300 g/l, the vessel should be capable of accommodating a liquid volume of 2 M^3, or a total volume of say 5 M^3. In order to accommodate fuel elements which are about 1 M long, it is suggested that the vessel is 1.5 M diameter by 2.5 M deep, steam jacketted over the base and 0.8 M up the sides. There should be a facility for changing the jacket contents from steam to water cooling. The dissolver should have a substantial condenser, as shown in Figure 3, of a tube and shell type, sufficient to condense all the evolved steam in the magnox dissolution reaction, or the steam and nitric acid in the uranium dissolving. This condenser should be followed by a cooled column in [illegible], about 1 M high and 20 cm diameter, draining through the condenser, in which nitric acid could be re-formed by reaction of the nitric fumes with a calculated flow of oxygen. A little water should be sprayed into the top of this packed column. The oxygen flow would be varied throughout the uranium dissolution, to fit a cycle based upon experience. Any nitric fumes remaining unreacted would pass to a sodium hydroxide scrubber, also in the highly active cell.

The fuel element feed should be through a simple inclined tubular condenser, as shown in Figure 3. The whole charge of 60 fuel elements should be handled by manipulators from the basket into this entry point, after the

the basket had been raised to a position just below the entry point, by means of the hoist. A window would be provided in this cell wall to observe the charging operation, the top part of this lift shaft thus resembling a small standard type of fuel element handling cave.

The fuel element entry point should be sealed tightly by means of a suitable clamp. It might have a lid containing a trapped "O" ring seal made of neoprene, but it should be possible to introduce into the cell a new lid complete with new "O" ring, when required. As an additional safeguard, this entry point should have a small stainless-steel drip tray under it, in case liquor were ever forced through the joint. Any liquid which ever collected in this tray would be removed by "ad hoc" methods (e.g. a suction bottle) using the manipulator. A further safeguard against active liquor ever reaching this point is that a flow of oxygen would be provided through an installed pipe inside the charge tube, near the seal, during dissolving, the oxygen being that required for recombination with nitric fumes.

The engineering construction of this dissolver is most important. It should be made of 18/13/1 niobium stabilized stainless-steel (or equivalent). All welds should be to the highest possible standard, and fully radiographed for inspection. The corrosion rate of this stainless-steel in the dissolver system is about 0.4 mm penetration per year. In view of the vessel's importance, however, and in order to withstand the stresses of charging 10 Kg elements, the walls and base should be at least 2 cm thick. The method of attaching the base should be to use a specially manufactured "dished end", 1.5 M in diameter and 2 cm thick, and to butt-weld this to the vertical wall. Other condensers, entry point, etc. can be reduced in thickness, but they should not be below about 0.5 cm.

Below the dissolver, the highly active cell should contain a receiver for the decanning solution, after use (if separate decanning is used), and two conditioner/feed vessels. All these should be about 3 M^3 capacity, and of say 1 cm. thickness 18/8/1 molybdenum stabilized stainless-steel, but fully radiographed. Liquors should syphon into them, the syphons being started by means of steam ejectors. The lines should, of course, have high lutes to prevent the syphons starting accidentally owing to pressure fluctuations inside the dissolver. Each of these lines, lutes, and ejectors, should be duplicated, and all ejectors should be brought to a

series of lead-brick shielded cavities let into the concrete shield wall of the cell, for easy maintenance.

All active vessels in the cell should be connected to the stainless-steel vessel vent manifold, at a high level, the stainless-steel lines leading to this manifold containining spray traps of a simple type. These might, for instance, take the form of short columns packed with stainless-steel gauze lessing rings, with a facility for spraying down each column with nitric acid when required, for decontamination. All vessels should have "wash-out" facilities, consisting of steam and nitric acid sprays. Wash-out liquors would be conveyed to the decanning liquor receiver or the conditioners, through the ejectors provided.

Non-active feeds to the vessels should be provided from stock tanks in a separate room at high level, serving the whole plant.

The whole of the floor of the cell and the walls up to the level of the top of the dissolver vessel (excluding condensers etc), should be lined with stainless-steel, about 2 mm thick, with sound, radiographed welds.

A duplicate dissolver cell should be provided, as shown in figure 4, but the plant should not be installed. Instead, appropriate connections should be provided, adequately blanked off and shielded, so that the second cell could be commissioned easily if ever it were required. By leaving the equipment until later, the opportunity is provided of perhaps saving its cost entirely, and of possibly rectifying any faults found in the design of the equipment in the first cell.

5.3 Plutonium Extraction

The flowsheet includes the novel process of extracting only the plutonium, at fairly high solvent saturation, with a solvent which is, in fact, already almost saturated with uranium. There is the possibility, under these conditions, of obtaining a high decontamination factor from fission products and of keeping the scale of the remainder of the plant very small indeed.

It is based upon the assumption that uranium obtained by 4,000 MWD/T irradiation of natural uranium has only about 0.4% uranium 235 content, and is of zero value. Taking the world market value of natural uranium as U$S 7 per lb for the 1970's (3) (The present value is only about U$S 4 per lb), and the value of "separative work" as U$S 30 per Kg, (3) (The present US value, unlikely to change) then U 235 concentrations below about 0.4% are

not worth feeding to a separation plant. If, instead, one considers blending the depleted uranium with enriched uranium, U235 concentrations even up to 0.53% have no value.(3)

Under the above circumstances, and since evaporation of the highly active effluent is not being recommended, it is suggested that all the uranium should remain behind with the fission products. Although this will increase the cost of fission product storage, it means that the remaining plutonium extraction stages of the process need only be of a similar size to the Ezeiza pilot plant! Furthermore, the stages of uranium stripping, uranium purification in one or two additional cycles, uranium evaporation, and conversion to solid form, become completely unnecessary.

The key to a process of this type is the plutonium distribution coefficient in a system where the solvent is nearly saturated with uranium. Data available suggest that without a salting-out agent this may be as low as 0.5 in a 20% TBP, 3N nitric acid system at 75 saturation. The figure is known to be about 1.55 with 30% TBP, 60% saturated, or 0.5 when 80% saturated. 30% TBP at 6N acidity and 80% saturated gives a figure of 2.6, which is more promising. The addition of say 1 or 2 M aluminium nitrate at say 3N acidity would be expected to give a large increase, although data are not available. It is also possible that magnesium in solution would cause a substantial increase, but again, there are no data. However, this has been included on the speculative flowsheet to indicate how simple and small would be the whole separation process, if this or a similar stage were successful. The magnesium would be obtained in solution by complete dissolution of the whole fuel element, without separate decanning. All the magnesium and uranium would then pass to the highly active storage plant with the fission products, in one operation, in a single solution.

If conditions cannot be found with TBP as the solvent, which allow extraction with a reasonably high plutonium distribution coefficient when the solvent is nearly saturated with uranium, it is suggested that an attempt is made to use "butex" (dibutoxy diethyleneglycol). This solvent was used very successfully in the first Windscale reprocessing plant, it being more stable to radiation and nitric acid than TBP. The distribution coefficient of tetravalent plutonium is known to be 7 when the solvent is fairly highly saturated. The ruthenium distribution

coefficient is much higher than with TBP, but the TBP cycles on the pilot plant scale could be retained, the cost of an extra TBP cycle if necessary, being negligible.

This process differs from that of a conventional first extraction in that the volume of first cycle solvent can be considerably less, e.g. a ratio of solvent to active feed of 1:1, instead of the more usual 3:1 or more (with the uranium feed concentration 300 g/l). A generous scrub has been shown, of ratio 1:4 with the solvent. Under these conditions, the first extractor above is quite small, with feed, solvent and scrub flow rates of 83 l/hr, 83 l/hr, and 21 l/hr.

The repercussions of the low solvent flow, fairly high scrub flow, and fairly high degree of solvent saturation, on the fission product decontamination factor, should be very favourable, although the influence of a salting-out agent would be unfavourable. A figure of 3,000 has been assumed.

Having obtained the plutonium in this low volume of solvent, the next stage is to strip into an even lower volume of aqueous phase by means of ferrous sulphamate solution. Unlike conventional flowsheets, no account need be taken of the behaviour of uranium and of fission products at this stage. Consequently, full advantage can be taken of the high degree of volume reduction obtainable with this system. The flow rate is, in fact, reduced to only 10 l/hr. Besides concentrating the plutonium, this small volume has the advantage of restricting the amount of uranium, and to some extent of fission products, which can be stripped with it. The flowsheet estimates are 6 Kg uranium and 20 Ci activity, per day, associated with 1.2 Kg plutonium.

After this stage, it is easy to see that all flow rates would only be of the same order as those used in the Ezeiza pilot plant, and the number of stages and process vessels would be similar. Criticality safety can be guaranteed quite easily and the level of fission products is no higher than in the highly active section of the pilot plant (modified for RA-3). Consequently, it can be predicted that the cost of the remainder of the plant should be about the same as the Ezeiza pilot plant (U$S 20,000) or allowing for better engineering, say U$S 100,000. Thus, costs after the first extraction cycle, are very small.

It is suggested that mixed-settlers should be used for this first solvent extraction cycle (as for the later small scale plutonium purification

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cycles). Their size would be sufficiently small for the comparison with the size of pulsed columns to be unimportant. Mixer-settlers, however, do have the important advantage over pulsed columns that they can be shut-down and re-started after hours or even days without appreciable changes in the state of equilibrium in the various extractor stages. This might be important in Argentina with a batch plant, and one constructed as cheaply as possible, where occasional shut-downs might be necessary.

The "simple type" of mixer-settlers (4,5,6) are recommended for this duty, although the more complicated KAPL "pump-mix" type could be used instead if desired. The writer has already made recommendations which have led to the replacement of the "pump-mix" type by the "simple-type" in the Ezeiza pilot plant, (7), and these show promise that they will have a lower maintenance rate. The design principles are the same for the larger units.

The first extractor should have six stages for extraction and six for scrubbing, each stage having mixing compartments of about 3 litres working capacity and settling compartments of about 9 litres working capacity, giving overall dimensions for one stage of 12 cm x 48 cm x 22 cm working depth. Thus, the whole extractor should be about 1,5M x 0.5M x0.3M deep. All stages should be fitted with wash-out pipes in both mixers and settlers. These should have individual small steam ejectors, conveying the contents of their various compartments via a 1 M high lute, into the highly active raffinate tank. This wash-out system may, in practice, only be used about twice per year, to remove emulsions and particulate matter which has accumulated on the bases of the mixers or settlers, or at the interfaces. However, this, in the writer's view, is preferable to a feed filtration system which would remove some of the solids, but not the silica "post-precipitated" on contact of the two phases. Before operation of the wash-out system, the extractors would, of course, be run-down with a little dilute nitric acid or non-irradiated uranium, to make sure that soluble plutonium was not ejected. The small amount of solvent ejected to the tank, and later to the fission product storage tank, can be neglected, since it is smaller than that entering this tank regularly in dissolved and suspended form, over six months.

In order to guarantee that plutonium build-up in this mixer-settler did not take place e.g. adsorbed on precipitated solids, it may be considered desirable to locate two neutron counters directly under the mixer-settlers.

These would respond to tens of grams of plutonium. If so, they could be fitted inside horizontal closed stainless-steel pipes extending through holes in the shielding wall, the holes being normally closed with lead plugs. This would enable the instruments to be kept free from contamination and they could be withdrawn for maintenance after first running down the extractor with non-active solutions.
The plutonium stripping mixer-settler extractor would be made to the same principles, but it would contain only six stages of working capacity, 6 litres each. Each stage would be 9 cm x 36 cm x 18 cm working depth, so the overall dimensions of the six stages would be about 0.6 M x 0.4 M x 0.25 M. Wash-out facilities are not so essential in this case, although they might be advisable in view of the continual recirculation of solvent which is proposed, without alkali treatment. This solvent might, in time, carry small amounts of "crud" from the first extractor to the second. They should again, eject to the highly active raffinate tank. Neutron counters should again be installed if advised by the Health and Safety Department.
The mixer-settlers should be located inside a highly active cell devoted only to first cycle extraction, scrubbing and stripping. It should also contain the 6 M^3 holding tank for the highly active fission product raffinate solution, which arises from this extractor, together with the small solvent holding tank of 0.5 M^3 capacity.

The highly active raffinate tank should have sampling facilities and an ejector to discharge the contents, after analysis, to the highly active storage system, together with a facility for spraying internally with nitric acid and steam. All vessels, mixer-settlers, and pipes should be of 18/8/1 stainless-steel, with good welds, fully radiographed.
The main active feed system must be guaranteed and in the U.K., where no maintenance at all is tolerated on this item, a very expensive rotating wheel device is used. For the Argentine plant, which would be batch operated and would not operate to a highly integrated programme, the writer recommends a simpler and cheaper system. A reliable type of diaphragm metering pump might be satisfacotry if a duplicate installed spare were provided and if each pump had a facility for washing with nitric acid, so that it could be maintained. To allow reasonable access for maintenance, the two feed pumps would be located in "bulges" or

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cavities in the main shield wall, adequately shielded from the remainder of the plant by concrete, and shielded from the operators by a perspex panel and lead bricks. The latter items would be removed for maintenance, after washing out. Experience on the Ezeiza pilot plant would give some knowledge of how satisfactory this system is in practice. It should be noted that the larger size of metering pump, as required for this production plant, is likely to be more reliable than the smaller size used on the pilot plant. The type of pump would be that where the active liquor is contained in a secondary circuit, activated by a primary circuit filled with hydraulic fluid.

The inactive feeds should be by conventional metering pumps or rotameters, located near their feed tanks, in a separate room, outside the active cell.

The active solvent feed pump installation should be similar to that of the main active feed. It has been noted earlier that the uranium loaded solvent would circulate between extractors 1 and 2 without alkali treatment. This would need experimental trials in the pilot plant, but the writer is of the opinion that a considerable number of cycles could be tolerated before this system became unsatisfactory. Then, say after six months, the solvent, together with its uranium load, and any plutonium and fission products retained, would be discharged to the highly active storage system. The volume in the circuit is normally enough to half-fill both extractors (70 l), and keep a reserve in the 0.5 M^3 tank. This volume would gradually decrease in use as a result of solubility and entrainment losses, and it would be made up again as required, with TBP, or diluent, or both. However, the timing of the discharge from the system could usually be arranged to coincide with the volume being at a minimum, i.e. about 70 litres. This would not present any special problem in the highly active effluent plant.

Degradation of the TBP in this solvent, under the action of nitric acid and radiation, takes place to dibutyl phosphate and monobutyl phosphate, and finaly to inorganic phosphate. The main product is the dibutyl ester, but, when uranium is present, this complexes it, so not leaving it free to complex plutonium or fission product zirconium. The inorganic phosphate is similarly eliminated in the aqueous phase. Consequently, unlike a system which requires to extract uranium efficiently, or to extract only plutonium in the absence of uranium, only the build-up of

monobutyl phosphate is important. A little plutonium loss would gradually occur as a result of this build-up and a certain degree of zirconium build-up in the solvent would occur, followed by its partial transfer into the plutonium stream in extractor 2. Fortunately, the plutonium loss is likely to be a very small proportion of the throughput, with such a small volume of solvent. Also, a substantial deterioration of the zirconium decontamination factor can be tolerated in the first cycle because the small size of the later cycles makes decontamination easy and cheap. (An additional cycle at only pilot plant throughputs would involve negligible costs).

The degradation of the diluent also occurs and this is less well understood. However, its effect on the flowsheet is mainly important at higher fission product concentrations and longer residence times than those considered here. In view of the exceptional nature of this first solvent cycle however, without alkali treatment, it is recommended that the diluent should be a pure straight chain paraffin, e.g. Dodecane or n-hexane, instead of the usual crude mixture of paraffine referred as "odourless kerosene". The additional cost is fairly high per ton of diluent, but negligible as a proportion of reprocessing costs.

The extractor would be arranged at a fairly high level in the cell, with the tanks underneath. The main highly raffinate would flow to its tank by gravity, as would the main plutonium product, to its own tank. The solvent flowing between the extractors would, of course, be elevated by means of its feed pump.

5.4 Plutonium Purification

This part of the process could be made to resemble the Ezeiza pilot plant, except that the dissolver system would be replaced by an evaporator to reduce the volume by a factor of 10. There are several cheap alternative processes which are then possible, to remove about 6 Kg or uranium and 20 curies of fission products from 1.2 Kg plutonium, in a volume of 30 litres. The flowsheet illustrates only one of these. It is based upon two cycle 20% TBP extraction, with splitting of the uranium from the plutonium in the first cycle, followed by evaporation, because data for this is immediately available to the writer. A similar 30% process could be used, but the reduction in volume is unimportant in this part of the plant.

It should be noted that it is not essential to achieve the final purity at this stage, since the oxalate precipitation stage gives a good decontamination from uranium and iron, and same decontamination from zirconium (the principal remaining fission product). The 65 day half-life of zirconium, in any case, guarantees its elimination if the plutonium is stored at any stage, including final storage as the oxide product. Handling is only a problem so far as the gamma activity of the plutonium isotopes (and any americium which grows on standing) is concerned.

The first plutonium purification cycle consists of three extractors of 12, 8, and 12 stages respectively, for extraction of plutonium and uranium together (with an aquoeus scrub), plutonium stripping (with a solvent scrub) and uranium stripping. The stripped solvent can then pass through the simple alkali wash system of the Ezeiza pilot plant type before being re-used.

The second cycle involves a simple forward extraction, scrub and strip, to remove iron. Some fission product elimination also occurs.

The same solvent could be used for both cycles, this demanding the use of only one set of solvent washing equipment.

The raffinates from extractors PP1, PP3 and PP4, together with the alkali wash liquors, would all pass to the main fission product storage system, even though the activity levels and concentrations are quite low. The reason for this is that dilution would, in any case, be required in this storage system owing to the presence of a high concentration of uranium.

The extractors for these stages should be quite small. Preferably they should be of a geometrically safe type, i.e. with a cross-sectional area of less than 300 cm^2. If they are difficult to manufacture in this small size, in the normal double-bank arrangement, then an in-line design is available (as used at DERE), which will allow the individual compartments to be a little larger.

The equipment, both the extractors and the stock tanks, would probably best be housed inside a conventional concrete cell, in view of the uncertain fission product activity, rather than having the mixer-settlers in lightly shielded cavities in the wall, as in the Ezeiza pilot plant. Two sides of the cell could be used, with the mixer-settlers on benches against the wall, and tanks underneath. The cell would have windows to

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allow vision of the mixer-settlers.
Access would be provided (after emptying the cell vessels) via a door at the end of the wall, so allowing a man to walk between the two lines of equipment.

5.5 Plutonium Evaporation

A final plutonium evaporation stage is included to give a product solution of about 250 g/l.
The evaporator can be made simply from a long glass tube of about 15 cm diameter, which would be geometrically safe. The glass should be well annealed industrial pipeline and it should preferably be heated by means of a high frequency induction coil, so that it is easy to see the liquor level through the window in the wall of the cell in which it is located. Alternatively, a special 2KW immersion heater could be used, inside a long silica sheath, as has been recommended for the smaller scale plutonium processing of the RA-3 blanket plutonium (8). Operation would be by continuous feed and evaporation at 1 litre/hr, maintaining the liquor at a constant level of 5 litres. The concentrate could finally be discharged into 1 litre bottles, for passing to the plutonium oxide production plant.
The cell should be constructed to floor level but of glove-box standards, and it should be attached to the end of the plutonium purification section.

5.6 Active Effluent Storage System

If the very small plant to extract only the plutonium proved possible, for natural uranium fuel elements, then all the effluent, both high and low activity, including the streams containing uranium, would be made alkaline and stored in a single tank or system of tanks. Plant wash-out liquors would also take this route. This is because the large quantity of uranium would require these solutions for dilution, so that it would form a fairly homogeneous slurry of not too high a viscosity, and not too high a specific activity, on making alkaline. The whole effluent system, therefore, would have the virtue of great simplicity, although it would be relatively expensive in view of the large volume involved.
The system used would resemble those of Savannah River, Hanford and Idaho Falls, in the U.S.A. (9). It is based upon storage alkaline in

mild-steel tanks, supported by concrete vaults.
The main fission product and uranium effluent would have a volume of 2.5 M^3 per day, associated with 0.6 tons uranium and perhaps 60 Kg of aluminium or magnesium, and an acidity of 6N. To this would be added the various other effluents shown in the flowsheet, totalling about 0.7 M^3 or, with wash-out liquors, averaging about a total of 3.5 M^3/day, at an acidity of about 4 N. An addition of 1.5 M^3 of 12N sodium hydroxide solution would precipitate all the uranium as sodium diuranate, precipitate some of the fission products and the magnesium as the hydroxides, convert any aluminium to the soluble sodium aluminate, and leave an excess of almost 1N free alkali. The solids content, mainly uranium and magnesium, would be about 150 grams (dry weight) per litre. This is a higher solids content than is usual in an alkaline storage system, but it is no higher than the 125 g/l to 360 g/l solids obtained in the U.K. acid storage system after evaporation of the fission products by a factor of 200. Furthermore, the possibility might be considered of using a sodium carborate solution for neutralization, instead of sodium hydroxide, which would dissolve some or all of the uranium.
The normal U.S. alkaline storage system when used with fission products from a "Purex" process, involves evaporation before making alkaline (which is possible in the absence of uranium) but by a much smaller factor than in the U.K. acid system (about a factor of 10). The overall cost of the U.S. alkaline system is similar to the U.K. acid system because the smaller evaporation factor is counterbalanced by cheaper plant, e.g. mild-steel tanks instead of stainless-steel, and a less intensive cooling system.

The system proposed for Argentina, (with natural uranium fuel elements) which has an actual increase in volume, would be more expensive than either the normal U.S. or U.K. systems, per ton of uranium reprocessed. Its only virtue is that it might allow a very cheap reprocessing plant to be used, which does not extract the uranium.
There are several types of system used in the U.S., involving water cooling by means of pipes within the tanks, or evaporation or condensation, or the cheapest process, without special heat-removal facilities. In view of the low concentration of the fission products, the system without cooling could probably be used. The specific activity would be about 70 curies of beta plus gamma activity per litre as it entered the storage

tank or tanks. The first solution to enter the tank would have a larger area than normal available for heat transmission, because of the tank being almost empty. If a 5,000 M^3 tank were used, lasting say 3 years, as the last liquor entered the tank, the first liquor would have decayed by a factor of about 10. Consequently, the average specific activity in a full tank would be about 25 curies per litre. The activity level of a full tank would, therefore, be about 1.2×10^8 curies.

The heating effect of the fission products in a full tank is believed to be equivalent to about 400 KW (but this requires more precise calculation). The wetted area of the tank (say 30 M diameter by 7 M deep) is about 1,400 M^2. The heat transfer coefficient required is therefore only about 1 BTU/hr/ft^2/ F, for a temperature of 100 F above ambient. This should be achievable even with a thick steel walled tank, and with a high proportion of solids in the tank, particularly if the solids are kept in motion.

The construction of the tank should take the form of an upright cylindrical vessel 30 M in diameter by about 8 M deep, with a domed roof. The tank thickness should be about 2 cm and the welds should be carefully made and radiographed. This should rest in a concrete vault, lined with steel, the concrete being about 0.5 M thick. The top of the tank should be about 3 M below ground level. The tank would contain a number of large air-lift circulators, to maintain the liquid, and suspended solids, in motion by the use of a relatively low flow rate of air. It is normal practice to use only four of these circulators but, in view of the small additional cost involved, and the rather large quantity of solids present, the writer would use about twenty circulators. The air from the circulators is passed through a cooler, condenser and scrubbing system and filtered before release to atmosphere. When the fission products have decayed for a number of years, it might be found practicable not to replace the condensed water carried by this air stream, but to allow evaporation and an increase in concentration to take place. More solution could then be accommodated in a tank than allowed for in the original design. Ultimately, by the use of air over the surface of the liquor, it might be practicable to allow the well-aged fission products and uranium to become solid. This acts as a protection against leakage due to corrosion of the tanks over many years. The presence of the uranium would be an advantage in this respect, since it would allow

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solidification at a relatively early date.

Vertical dry wells and horizontal laterals around and under the vault are used to detect leaks. It is desirable that the location should be one with a low water table and that the ground should have good ion-exchange properties in case of leakage. Leakage would normally be countered by pumping the liquor from the leaking tank to a new tank. More than one type of remote pump should be installed for this purpose, e.g. airlifts and steam ejectors. In this case, it might be advisable to allow the uranium to settle (if the self-heating effect did not cause convection) and pump only the supernatant liquor to a new tank.

If the reprocessing plant is intended to last ten years, three tanks, each with a capacity of 5,000 M^3 would be required. A fourth spare tank would also be desirable, although the building of this might be postponed, to see if, in the last few years, additional liquor could be placed in the first two tanks, as a result of deliberate evaporation.

5.7 Plant Cells and Main Plant Items

The table below lists the cells, etc. with principal dimensions and main plant items. A very approximate estimate is made of the cost of purchase or fabrication of the main items of equipment inside each cell.

Cells	Internal Dimensions	Wall Thickness	Main Plant Items	Rough First Estimate of Cost of Plant Items - US$
1. Pond	12M x 12M x 5M deep	0.5 M	3 Ton travelling crane 30 ton fixed crane Long hydraulically operated tongs 100 stainless-steel baskets Simple guillotine Decontamination tray	50,000
2. Hoist Cells (2 cells, one only complete, with equipment)	1 M x 1M x 7M high	1.5 M	Hoist Window Manipulator Dissolver seal and tray	20,000

3. Dissolver Cells (2 cells, one only complete with equipment)	6M x 3M x 8M high	1.5 M	Dissolver Condenser Fume reaction column Three 3 M^3 tanks Ejectors Spray traps Wash-out facilities Stainless-steel cell lining	50,000
4. Extraction	3M x 3.5M x 3M high	1.5 M	First extractor Mixer-settler Raffinate tank Strip mixer-settler Uranic solvent stock tank Ejectors Wash-out facilities 2 Main active feed pumps (in cell wall) 2 Active solvent feed pump (in cell wall) Stainless-steel cell lining 4 Neutron counters	40,000
5. Plutonium Purification Cell (Including Pu backwash from uranium loaded solvent)	6M x 5.5M x 3M high	0.5 M	Evaporator 5 plutonium purification extractors Valency conditioning vessel and scrubber 4 active pumps (in cell wall) 2 Solvent washers 3 Small raffinate tanks 5 Force-lift Ejectors Small solvent stock tank Stainless-steel cell lining Portable neutron counter Cell windows	40,000

6. Evaporation Cell	2M x 2M x 3M high	6.2 M	Plutonium evaporator Condenser Condensate receiver Alpha counter Botte filling system Bottle transfer system Product pump Stainless-steel cell lining Portable neutron counter Cell windows	15,000
7. Inactive Reagent Room	13M x 3M x 3M high (elevated 2.5M high)	0.1 M	About 12 tanks, each 1.5M^3 with pneumercator type level gauges 12 feed pumps 12 rotameters 12 filters Nitric acid tank waggon (out-side plant) Oxygen tank (out-side plant)	50,000

A simple sketch of the proposed layout is given in Figure 4.

5.8 Total Cost

The roughly estimated cost of the main plant items totals U$S 265,000. The piping and minor items are difficult to estimate in this manner, without detailed drawings, but a figure equal to the main plant items would probably be appropriate, making U$S 530,000.

In view of the high standards of welding and inspection required for most of the equipment in the active cells, it is proposed to double these estimates, which makes a total plant and equipment cost of U$S 1,060,000.

The volume of concrete used in the cells and pond is estimated as about 1,200 M^3 or say 3,000 tons. If the cost of reinforced concrete, laid in position, is about U$S 70 per ton, the total concrete costs are about U$S 200,000.

The building itself can be constructed quite simply of normal brick,

with plastered interior. The volume of the main part shown on the sketch is about 1,300 M^3, or about 1,500 M^3 if small compartments are added for the rod feed cells. At say U$S 200 per M^3, the building cost would be U$S 260,000. Ventilation systems for vessels, cells and the operating area might double this, making U$S 520,000.

The total cost of the pond and separation plant is therefore estimated at about U$S 1,780,000, or, with design charges, say U$S 2 M.

The highly active effluent storage system adds considerably to these costs however. For a 10 year programme, it is proposed to use four storage tanks, each of 5,000 M^3 capacity. The total cost of each tank of about this size (actually 1,330,000 gallons) at Savannah River, including all installations, was U$S 630,000. Four tanks would, therefore, cost about U$S 2,5 M. It is usual, also, to provide a capital sum, the interest from which would cover all operational and maintenance charges, including the provision of new tanks in future years to replace the corroded old ones. This sum is estimated as about 100% of the initial capital cost.

The total cost of effluent disposal is therefore U$S 5 M. Adding this to the separation plant costs gives U$S 7 M.

By comparison, the cost of conventional reprocessing plants, excluding effluent disposal, (which is normally relatively small) is for plants of 0.1, 0.2, 0.3, 0.5, 1, 2, and5 tons per day, U$S 9 M, 13 M, 15 M, 20 M, 28 M, 40 M, and 60 M respectively.

A conventional plant of the Argentina size, therefore, should cost about U$S 25 M.

The Windscale plant in fact, conforms with the last figure of U$S 60 for a designed throughput of 5 tons/day, but, an actual throughput rather higher, about 2,000 tons/year. This did not include the full cost of effluent disposal, or even the cost of capital for this purpose. Adding say U$S 10 M for this purpose gives a total of U$S 70 M capital. This plant has a throughput about ten times that proposed for the Argentina plant, and a capital cost also ten times as high. Since capital costs constitute 80% of the full costs of reprocessing, it appears that the small unconventional plant proposal for Argentina might just be competitive with a large plant, which at the present time is quoting the lowest reprocessing cost per ton in the world.

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Taking the life of the plant as 10 years, and in interest rate, in stable currency, of 8% per annum, the capital charges for reprocessing would be 15% of the initial capital each year. Thus, the capital charges for reprocessing 180 tons would be 15% of U$S 7 M, or about U$S 10.5 M. This is equivalent to only U$S 6,000 per ton. The figure already includes the operating cost of the effluent storage plant, which is responsible for most of the capital. If, however, another 50% of the total capital charges are added for operational cost, the total reprocessing cost becomes U$S 9,000 per ton. Since 2 Kg of plutonium are anticipated per ton, the cost of extracting plutonium is about U$S 4.5 per gram.

6. SUGGESTED PLANT FOR T/yr IRRADIATED ENRICHED URANIUM FUEL ELEMENTS 40/

6.1 Fuel Decay

Although the irradiation dose is likely to be much higher in this case, 150 days should be sufficient decay for the iodine to be reduced to a negligible amount. Each reactor cooling pond, in this case, needs only be large enough to hold about 10 tons, if cooling is carried out there. However, these thin fuel elements may also require to be disassembled from their stringers or clusters, for more convenient transport to the reprocessing plant, so extra space may be required, if not already present to receive unexpected discharges from the reactor in emergencies. It is possible, with these fuel elements, because of the smaller quantities involved, that a second cooling pond might be avoided at the reprocessing plant. A cave would be used for mechanical operations on the elements, or stringers, after decay, and this would have limited storage space associated with it. The shielded transport flask used for transport from the reactor pond would, therefore, require to feed its contents to a cave. There are standard methods of carrying out operations of this type, using manipulators and remote cranes. The weight of a fuel element stringer, with perhaps up to 100 elements, will probably be about 0.1 tons per metre of length.

6.2 Fuel Preparation for Reprocessing

For fuels of this type, it is generally agreed that the "chop and leach" technique is cheaper than normal decanning before dissolving, which, with oxide fuel elements, could only be carried out by a chemical method. The chop and leach technique is at present being actively developed in

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other countries and it is possible that a guillotine and auxiliary equipment could be bought more cheaply than it could be developed in Argentina.

Normally, a guillotine would chop the whole fuel stringer into short lenghs, say 1 cm. For this purpose, with either stainless steel or zirconium cans, it is best to greatly over-design the power unit. The thrust would depend on the type of fuel stringer being sheared, but it is envisaged that a unit of about 100 tons would be used. The blade should be of hot-work tool-steel, (e.g. English Steel Corporation type "CVM 2", or Firth Brown "cromodie N", or Arthur Balfour "CTU"). The life would be about 10,000 cuts, and facilities would be needed for replacement of the blade, or the whole cutting unit, remotely, by means of manipulators.

The machine would be operated hydraulically, the hydraulic equipment being slow acting, to keep the initial cost and the cost of maintenance low. As much of the non-active hydraulic equipment as possible would be located outside the cave.

For a throughput of 40 tons per year, the number of cuts of a hundred element stringer is only about an average of one every 15 minutes, so any machine of this type would not be in use for the whole 24 hours of a day. An alternative method is to completely dismantle the stringers into individual elements, if this is possible with the particular stringer design. The individual elements could then be cut with a cheap, low-pressure, rapid acting machine, requiring very little development.

With either machine, if the blade moves in a horizontal plane and the fuel is fed vertically, it is easy to arrange for the chopped pieces to fall directly into a stainless-basket. This basket should be complete with an outer container of only slightly larger diameter, to retain any small particles of fuel which fell through the holes in the basket. When the basket alone is finally placed inside the dissolver, it may be possible to invert the outer container and empty the fine material on top of it. If this is mechanically inconvenient, in the dissolver cell, a container from one chopped batch could, more easily, be inverted over the next chopped batch, in its container, in the chopping cell, before transfer to the dissolver cell.

The cutting cell should have a nitrogen or carbon dioxide atmosphere to avoid the fire hazards from finely divided uranium dioxide and from zirconium.

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6.3 Dissolving

For a 40 ton/yr plant operating regularly, daily quantities of about 120 Kg would be required, or about 150 Kg, including the chopped cans. The packed volume of these pieces would be about 35 litres and the plutonium content, at a nominal 5 Kg/ton of fuel, would be 0.6 Kg. A batch operated leaching vessel with only low uranium enrichment might be allowed on criticality grounds without any restriction on shape, owing to the presence of neutron absorbers such as nitric acid, stainless-steel and uranium 238. However, this approach is not recommended in view of the difficulty of guaranteeing the absence of accumulated undissolved plutonium. The plutonium content of the leaching vessel cannot even be determined as a result of a careful mass balance of input minus output, since the plutonium is liable to vary markedly.

The leaching operation takes only a few hours and the daily quantity could be divided into say three 40 Kg batches, each of which would have well below one half of a critical mass of plutonium under normal circumstances. Even so, the writer is of the opinion that even small batches of this type are best dissolved in a vessel of restricted geometry. The maximum allowable diameter for such a dissolver is 9 inches (22.5 cm) and the basket could be about 21 cm diameter. A daily chopping batch would therefore be divided into three dissilving batches, each in a basket of 21 cm diameter an a packed depth of 35 cm, or say a total basket depth of 45 cm.

The basic design of the dissolver would be a vertical tube about 2 M deep by 22.5 cm diameter, with a 1 M condenser above, having side limbs connected to a 100 litre slab tank, size about 2M x 1M x 5 cm thick. A secondary condenser would be attached to the primary condenser, but this would be packed with glass lessing rings (flame annealed with rounded edges) to allow recombination of the nitric fumes with an oxygen feed. The primary condenser would be unpacked to allow access by the basket of chopped fuel. It could have a lid capable of being sealed remotely, perhaps by means of a water lute. Oxygen could be fed at this point to keep this part of the system in a condition relatively free from contamination.

A sketch of the proposed design is given in Figure 5. This should be regarded as little more than diagrammatic, since development would be

needed at this stage.

The dissolver, slab tank, and auxiliary equipment, should be fabricated entirely of 18/13/1 niobium stabilised stainless-steel, and the welds fully radiographed. The steel thickness of the main vessels should be about 1 cm to 1.5 cm, since it is known that, under these conditions, the corrosion rate can be 0.02 to 1 mm/yr at 110 C.

Leaching, to dissolve the uranium and plutonium, would take place once every eight hours, using 133 litres of 5.5N nitric acid. The main part of the dissolution should take about 2 hours so that a further 6 hours is available for charging, discharging, heating, cooling, and rinsing. The heat of dissolution is about 17 K cal/mole, so that the heat of reaction during the two hours is equivalent to about 1.5 KW. A further few hundred watts is contributed by the fission product heat. This total heating rate is too low for for speedy reaction, and a narrow steam jacket should be provided on the main vertical limb. If this takes the total diameter outside that allowable on criticality grounds, then the dissolver diameter could be slightly decreased and the length slightly increased. Alternatively, it might be possible to include the jacket on the bottom limb from the slab tank, if this limb is inclined at an angle. This jacket would also be used for water cooling, wherever its location, but additional air cooling would be desired from the large surface of the slab tank, so as to discharge at a reasonably low temperature. At the end of a leaching operation, 133 litres of 300 g/l solution would be discharged by means of a force-lift steam ejector (duplicated) into one of the two duplicate feed and conditioning tanks. This could be perhaps conventional in shape, with a neutron counter underneath to detect a build-up of solids on the base. Alternatively, if cheaper, another slab tank could be used.

After discharge and draining, the next batch of 5.5N acid would be fed to the dissolver fairly rapidly so that it would wash the drainage liquor from the leached pieces of can, into the slab tank. To make this rinsing operation more efficient, the basket could be raised into the main condenser section and the acid added as a coarse spray, from the top of the dissolver, with the dissolver lid open. Operations such as raising and lowering the lid and basket would be carried out using manipulators which had only a pre-set cycle of movements, since it might not be convenient to arrange for vision within the dissolver cell.

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The old basket should be exchanged for a new one with the acid cold, and the lid quickly replaced. If this operation were found to lead to an undue amount of reaction, and fuming, before the lid could be closed, then the rinsing operation should be with half the volume, of water (or very dilute nitric acid) only. An equal volume of 11N acid should then be added to the dissolver, after the sealing operation.

The basket of leached pieces of stainless-steel or zirconium can, should be sent to an active solid waste silo since each batch would have an activity level of the of a few curies. Owing to the hazards involved in the storage of zirconium metal, this should be first passed through a small concrete mixer, mixed with concrete, and the wet concrete passed to the silo to set hard.

The principles of design of the dissolver cell would be the same as for the natural uranium metal dissolver cell. The dissolver should be at a high level so that the ejectors used for discharging it only have to operate for a few minutes to start a syphon. All vessels should be vented through traps capable of being washed with acid.

The cell should be lined with stainless-steel up to the top of the dissolver. A completely duplicate cell should be provided, but empty of equipment. The size of these enriched uranium oxide dissolver cells should be similar to those for natural uranium metal, since, although the throughput is less, the extended shape of the dissolver, makes any appreciable saving of space impossible.

The ventilation air from the dissolver cell should pass to a chimney on top of the cell. It should also take the final gas stream leaving the dissolver, after passing through a conventional sodium hydroxide scrubber. Besides Krypton and Xenon fission products, about 10 curies per day of tritium will be expected in this gas, i.e. about 25% of that entering the dissolver, but it should not constitute a hazard with a reasonably tall chimney and a high dilution with air.

6.4 Extraction

Although dependent upon the particular type of reactor, it is probable that the uranium after irradiation of enriched uranium oxide will still have a value. Its enrichment may even be greater than that of natural uranium. It is not, therefore, in this case, proposed to extract only the plutonium. The flowsheet can be quite conventional and need not be discussed in great detail, since a number of alternatives are available.

The main saving will arise from the fact that the plant will be sized for a throughput of only 40 tons/yr. Mixer-settlers and the vessels can, therefore, be quite small. Experience thoughout the world suggests that the complications and increase in size of plant needed for remote maintenance to be possible, are rarely justified. In Argentina, particularly, with relatively small vessels, and in the absence of highly integrated production schedules involving a large number of reactors, it is suggested that the plant items should be enclosed in relatively small and simple cells, with little emphasis upon access, for remote maintenance or direct maintenance. The obvious provision for duplicate transfer devices, wash-out facilities, and the use of "bulges" for occasional acces to pumps, motor drive units, samples, etc, should be made. Under these circumstances, the precise flowsheet is less important in controlling the plant costs.

In the U.K., it was decided that the standard TBP process as used for 4,000 MWD/T irradiated natural uranium, did not quite give the desired ruthenium decontamination factor when used with enriched oxide fuels which were to be irradiated to a much higher level. It was, therefore, decided to use a preliminary extraction cycle with "butex", a solvent which is particularly resistent to degradation in the presence of irradiation. This was convenient since an old butex extraction plant was available for use. In Argentina, it is probable that such special measures can be avoided, for the following reasons:-

1. An all-TBP plant can be developed specially for the purpose of reprocessing enriched oxide fuels.
2. The irradiation level will not exceed about 20,000 MWD/T, and may be less.
3. A greater ruthenium concentration of the recycled uranium might be tolerated in Argentina because of the smaller quantity of uranium to be handled in the fuel element fabricacion plant.
4. There is much more opportunity of adjustment of the ruthenium species for low entractability, in a small batch dissolver or as a special pre-treatment after dissolving. For example, a simple nitrite treatment, by passing nitrogen dioxide through the solution, gives a small but definite improvement. Various precipitants could alternatively be used, but it is suggested that the cost and complication of a highly active filter should

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be avoided if at all possible.

be avoided if at all possible. An outline, description of the purification cycles of some well known reprocessing plants is given in the table below:-

Plant	U.K. Windscale	U.S. Purex A	U.S. ORNL	U.S. NFS	Eurochemic
% TBP	20	30	30	5 to 30	30
Early or late split of Pu from U	late	late	Early	Early	Early
Number of U solvent cycles	3	3	2	3	2
Number of Pu sovent cycles	4	3	2	2	2
Total number of solvent cycles	5	4	3	4	3
Number of intercycle evaporators	0	1 Pu/U 1 U	1 U	0	1 U 1 Pu
Additional U purification	None	None	Silica gel	Silica gel	Silica gel
Additional Pu Purification	None	None	Ion-exchange	Ion-exchange	TLA cycle

It is seen that, including the purification cycles which are not with TBP, the total number required is 4, 5, or 6.

The Windscale process has one plutonium cycle which is a reserve, and is not strictly necessary, so it should be classed with Purex A as a 4 cycle process, for natural uranium fuels, ORNL, also for natural uranium fuels, would be in the same category except that it has a fifth cycle because of an early split instead of a late split. The NFS and Eurochemic plants have 5 or 6 cycles but they are designed to take the more highly irradiated enriched uranium fuels.

It is suggested that the Argentina plant, for enriched fuels, should have a simple pre-treatment to modify the ruthenium, and then 5 cycles, which would thus be the equivalent of the maximum number of cycles of these plants listed. By comparison with the table, therefore, for Argentina is suggested 30% TBP, a late split, 3 uranium TBP cycles, 3 plutonium TBP cycles and one plutonium ion-exchange cycle. This should reduce the uranium fission product content down to about 0.1 millicuries/Kg and the plutonium fission product content to about 1 millicurie/Kg, both of which are acceptable. Furthermore, this is

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achieved without operating to a very high degree of solvent saturation in the forward extractors. The advantage of this is that flows do not have to be controlled with extreme accuracy, and since flow errors can be avoided, there is no need to have facilities for "re-work" cycles. It might be considered that the last uranium cycle, with its fission product decontamination factor of only about 10, is not essential. However, again, the writer would prefer to see this extractor included, rather than have to recycle unsatisfactory batches of uranium product. Neither would the writer replace this extractor by a silica gel column, which would be even less effective and the savings of space and operating materials would be very small on this size of plant. Furthermore, silica gel columns require more operational and maintenance attention, in general, than mixer-settlers.

It should be noted that the proposed flowsheet also does not include inter-cycle evaporation. These evaporators do not achieve great savings in the size of plant, at least where uranium is being extracted, since the volumetric flow of solvent, to an extractor is dependent mainly upon the weight of uranium in the feed solution to the extractor, and not upon the volume of the feed solution. Inter-cycle evaporators do, however, save nitric acid, but the writer has considered that the size of the plant (and therefore its cost) and the ease of operation, and absence of maintenance, are more important considerations. Another important reason to avoid inter-cycle evaporators is the fact that they cause greatly enhanced degradation of the TBP and diluent which is dissolved and entrained in the solutions being evaporated. These products can form precipitates and have a marked effect upon subsequent cycles. This effect is not a uniform one, since the quantity of entrained solvent varies, as do the evaporator heating conditions, and the quantity of the precipitated degradation product which leaves the evaporator with the concentrated solution. Consequently, this whole effect is one which varies from batch to batch. At its worst, it is responsible for a high proportion of the "re-work" which is necessary in those plants which include inter-cycle evaporation (e.g. in the U.S.). By contrast, the Windscale plant, with no inter-cycle evaporation, and no critical flow-rates for the achievement of very high solvent saturation, also has no "re-work" requirements.

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An attempt has been made to draft a possible flowsheet based upon the above principles, in Figure 2. It resembles the Windscale flowsheet in the details discussed above, but resembles the other flowsheets tabulated in that 30% TBP is used instead of 20% TBP, and a small but flexible ion-exchange cycle is proposed in place of the last plutonium solvent cycle. Many of the figures shown in this flowsheet are approximate, since they are not based upon direct experiment, and development work would be necessary.

The solvent wash systems are not shown in Figure 2. The writer at this stage, sees no reason to deviate from the standard practice of washing the solvent from the first highly active cycle separately, using first sodium carbonate. If found necessary during development, the sodium carbonate solution could be replaced by sodium hydroxide in one or both solvent treatment cycles. The Windscale system of washing is highly effective but much more complicated and it adds substantially to the whole cost of reprocessing. Much of the complication derives from caution, which is now not necessary, and also from the fact that, with a very large plant, it is important on criticality grounds to make sure that plutonium cannot deposit in the wash systems. These factors would not apply to the Argentina plant.

It is suggested that the extraction plant is divided into three separate cells. The first, highly active cell should contain the two "simple type" mixer-settler extractors, each of about 18 stages. With stage sizes of approximately 10 cm x 10 cm x 20 cm working height, for the mixers, and 10 cm x 30 cm x 20 cm working height, for the settlers, the overall dimensions of each extractor would be about 1.8M x 0.4M x 0.3M high. They could be located quite close together, with gravity flow of the solvent between them and gravity flow of the highly active fission product raffinate from the first extractor to a 2 M^3 tank underneath. The main active feed pump would be located in a bulge in the shield wall and the non-active feeds would come from a separate room, as discussed in Section 5.3. The recycled solvent could be kept to a small volume and might be located in the non-active feed room but with a little local shielding, if necessary. Other general mechanical features, such as the installation of wash-out facilities and neutron counters, would also be as in Section 5.3.

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The second extraction cell should contain the splitter cycle and the uranium purification cycle, together with a heat treatment tank for ruthenium and the two solvent wash cycles. It should be adequately shielded from the highly active cell in case access is ever required, since decontamination would be relatively easy in this case.

The principles of construction should otherwise be the same as for the highly active extraction cell, the first two mixer-settlers should be about the same size as the highly active ones, i.e. 1.8M x 0.4M x 0.3M high, with 18 stages. The third extractor, for uranium stripping, would only have 11 stages, although each stage would be of about the same size, so its overall dimensions would be about 1.1M x 0.4M x 0.3M high. The fourth extractor for the final extraction of uranium would be of 18 stages and the fifth of 11 stages, size 1.8 and 1.1 M in length respectively, by 0.4 M and 0.3 M high.

Two small mixer-settlers, of four stages each, would be used for solvent washing, one mixer-settler for the highly active solvent and one for low active solvent. In both cases, the first two stages treat with sodium carbonate solution (or possibly sodium hydroxide) and the second two with nitric acid.

The third extraction cell would be for plutonium purification only. Since the total fission product activity in the feed to this cell is expected to be only about 200 millicuries per day, in a total volume of less than 400 litres, this cell could perhaps be quite lightly built, with a little local shielding of the mixer-settlers and other vessels, as required, together with thick perspex panels. For example, the first extractor should, at any time, only have a fission product activity of about 10 millicuries, although the low energy gamma activity of plutonium isotopes would also be present.

The two extractors in this cell would be quite small, with mixers of about 6 cm x 6 cm x 10 cm working height, and settlers of 18 cm x 6 cm x 10 cm working height. The first, with 20 stages, would have overall dimensions of about 1.2M x 0.25M x 0.15M in height, and the second, with 12 stages, would be about 0.75M x 0.25M x 0.15M in height.

Installed wash-out facilities are not needed in the mixer-settlers since there is relatively easy access at all times. A portable neutron counter could be used, outside the box, underneath the extractors, to make sure that excessive quantities of plutonium were not building-up.

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The plutonium product solution from the last mixer-settler would pass through a small stirred vessel into which was fed nitric acid and sodium nitrite solution, to give a solution 5N in acid and with all the plutonium in the tetravalent condition. The residence time would be about two hours, and the solution would then flow continuously into a glass ion-exchange column containing about 8 Kg of De-acidite FF resin. The column would be about 12 cm internal diameter and 1.2 metres high. Two columns would be provided, so that one could be used for adsorption whilst the other was used for washing with 5N nitric acid and elution with 0.2N acid. The cycle time would be 24 hours. The wash volume would depend upon the required decontamination factor from fission product zirconium, and from iron, but about 20 litres should be satisfactory. Elution following this stage would have ample capacity.

6.5 Plutonium Evaporation

The same type of evaporator would be used as described in Section 5.5 for the comparable plutonium stream arising from the reprocessing of irradiated natural uranium fuel elements. In the present case, the plutonium throughput would be only about 0.6 Kg per day and the volume for evaporation would be perhaps only 10 L/day. The required evaporation rate, therefore, is below 0.5 l/hr.

The tubular evaporator would operate with a constant volume of concentrate, about 2.4 litres, and this would be discharged at the end of each day into 1 litre bottles, when the concentration had reached about 250 g/l.

6.6 Active Effluent Storage System

Since the flowsheet, for reprocessing, is fairly conventional in this case, and the quantity of uranium per year is low, the highly active effluent system would be very small. The 40 tons/year would yield highly active liquor of volume less than 200 M^3, which might be evaporated to 20 M^3 if stored alkaline, or 1 M^3 if stored acid, so the whole 10 year programme would only involve the storage of either 200 M^3 or 10 M^3 of highly active effluent, against 15,000 M^3 for the special natural uranium programme. On this small scale, the more complicated acid system is probably not competitive, and a small alkaline storage system is recommended.

In this case, with irradiation to a postulated 20,000 MWD/T, and evaporation, water cooling would be required.

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Medium active effluents would be evaporated and sent to the highly active storage system. Ground disposal would be used for low active effluents.

With this reprocessing plant, special provision will have to be made for the disposal of low active effluents. If the location is not near the coast, then ground disposal will presumably be used, although the site should be selected carefully to ensure that the ground had good ion-exchange properties for retention of the fission products, and the water table should be low.

Those low or medium active effluents which were free from salts could be evaporated and stored with the highly active effluent. With a plant of the size envisaged, it might, however, be difficult to justify the cost of such an evaporator, with its thick shielding and force-lift steam ejectors for discharging the concentrate. Nitric acid recovery, by re-evaporation of the distillate, is even less likely to be economically justified. This report, therefore, tentatively recommends ground disposal for all effluents except the highly active liquor from the first extractor.

6.7 Plant Cells and Main Plant Items

As in Section 5.7, the cells are listed, in the table below, with their suggested dimensions and the main plant items. Again, a very approximate cost estimate is included in the table, for the main plant items.

Cells	Internal Dimensions	Wall Thickness	Main Plant Items	Rough First Estimate of Cost of Plant Items - U$S
1. Chopping cell (2 cells, one only complete with equipment	2M x 1M x 4M high	1.5 M	Guillotine About 6 baskets and containers Cell inlet from shielded container Stainless-steel cell lining	75,000

- // -

2. Dissolver Cell (2 cells, one only complete with equipment)	6M x 2M x 8M high	1.5 M	Geometrically safe dissolver Geometrically safe slab tank Primary condenser Secondary condenser 2 Stainless-steel 1M^3 tanks Ejectors Spray traps Wash-out facilities Stainless-steel cell lining 2 Manipulators	50,000
3. Solid Waste Silo	3M x 4M x 12M deep (10M sunk below ground)	0.5 M (below ground) 1.5 M (above)	Crane (Concrete mixer)	
4. Highly Active Extraction Cell	4.5M x 3M x 3M high	1.5 M	2 mixer-settlers Raffinate tank (1.5 M^3) Uranic solvent stock tank Wash-out facilities Ejectors 2 Main active feed pumps (in cell wall) 2 Active solvent feed pumps (in cell wall) Stainless-steel cell lining 4 neutron counters	45,000
5. Medium Active Cell	6M x 4M x 3M high	1 M	5 mixer-settlers 2 solvent wash mixer-settlers 2 raffinate tanks (1.5 M^3) Uranium product tank (1.5 M^3) Plutonium product tank (1 M^3) Solvent stock tank (1 M^3) 2 Solvent wash raffinate tanks 4 Active feed pumps (in cell wall)	60,000

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			2 Active solvent pumps (in cell wall) Heat treatment vessel Ejectors Stainless-steel cell lining 4 neutron counters	
6. Low Active Cell	5M x 2M x 3M high	0.2 M	2 Mixer-settlers Raffinate tank 2 Active feed pumps 2 Active solvent pumps Conditioning vessel 2 Ion-exchange columns Ejectors Stainless-steel cell lining Portable neutron counter Continuous alpha counter Cell windows	10,000
7. Evaporation Cell	2M x 2M x 3M high	0.2 M	Plutonium evaporator Condenser Alpha counter Bottle filling system Bottle transfer system Product pump Stainless-steel cell lining Portable neutron counter Cell windows	15,000
8. Inactive Reagent Room	12M x 5M x 3M high (elevated 2.5M high)	0.1 M	About 14 tanks, each 2M^3, with pneumercator type level gauges 16 feed pumps 16 rotameters 16 filters Nitric acid tank waggon (out-side plant) Oxygen tank (out-side plant)	70,000

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			Nitrogen or carbon dioxide tank (outside plant)	

A simple sketch of the proposed layout is given in Figure 7.

6.8 Total Cost

The roughly estimated cost of the main plant items totals U$S 325,000
Doubling this figure to include piping and minor items gives U$S 650,000
Doubling again for the high standard of welding and inspection gives U$S 1,300,000.

The volume of concrete, for this plant is estimated at about 1,200 M^3, or 3,000 tons, which, as in the other plant, should cost about U$S 200,000.

(The two estimates for concrete are similar since although this one has more shielded cells, it does not have a pond).

The building volume shown on the sketch is about 2,800 M^3. At U$S 200/$M^3$, the cost is U$S 560,000. Doubling this for ventilation gives U$S 1,120,000. The total cost therefore is estimated at about U$S 2,620,000, or with design charges about U$S 3 M.

In the present case, the cost of the effluent storage plant can almost be neglected. The volume of effluent, after evaporation, is 200 M^3 in 10 years, agains 15,000 M^3 unevaporated from the natural uranium fuel element plant. On a pro rata basis, the full permanent storage cost would only be U$S 70,000, but it would be more reasonable to treble this figure for the very small scale of operation. Adding say U$S 90,000 for the highly active effluent evaporator gives U$S 300,000.

The additional cost of ground disposal of low active effluents is not known, but it is assumed that, on this small scale, it would not exceed U$S 100,000.

The total cost of the separation plant and effluent plants, therefore, is estimated as very approximately U$S 3.4 M.

By comparison, a normal plant at a throughput of 0.1 tons/day, costs about U$S 9 M.

The annual equivalent of a capital cost of U$S 3.4 M, with a plant life of 10 years, and on 8% interest rate per annum, is 15%, or U$S 510,000. At a throughput of 36 tons per year, the capital charges

per ton are about U$S 14,000. In this case, it is doubtful if operational costs could be reduced below about 100% of capital costs, so it would be more realistic to take the full reprocessing cost as about U$S 28,000 per ton.
Since, in this case, a ton of fuel should contain about 5 Kg of plutonium, after 20,000 MWD/T irradiation, the cost of extracting plutonium would be about U$S 5.6 per gram.

7.- CONCLUSIONS

1. It might be possible to build a very cheap, unorthodox reprocessing plant for 200 tons/YR of irradiated natural uranium fuel elements, based upon extracting the plutonium only, and leaving the uranium (which has no value) with the fission products.

2. This small, inexpensive reprocessing plant would have very high fission product storage charges. However, the combination of cheap reprocessing and expensive fission product storage appears to lead to an overall cost which would be competitive per ton of uranium, or per gram of plutonium, with the world's largest plant.

3. The calculated cost of reprocessing irradiated natural uranium fuel elements in the above plant is about U$S 9,000 per ton, or about U$S 4.5 per gram of plutonium extracted.

4. Large savings can be made in the building of a reprocessing plant for only 40 tons of enriched uranium oxide fuel elements, if the plant is built for a specific purpose and takes full advantage of previous technical and engineering experience.

5. Such a plant might be almost competitive with much larger reprocessing plants which exist.

6. The calculated cost of reprocessing irradiated enriched uranium oxide fuel elements in the above plant is about U$S 28,000 per ton, or about U$S 5.6 per gram of plutonium extracted.

8.- RECOMMENDATION

It should be noted that the reliability of the data used in this very rapid assessment, and of the cost figures obtained, is very low, since, in the nature of the assessment, it was not based upon specific

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development work. It is believed, however, that the two systems show enough promise to justify such development work and a more thorough assessment. This is recommended if Argentina wishes to reprocess its own irradiated fuel elements on this relatively small scale but on a competitive basis.

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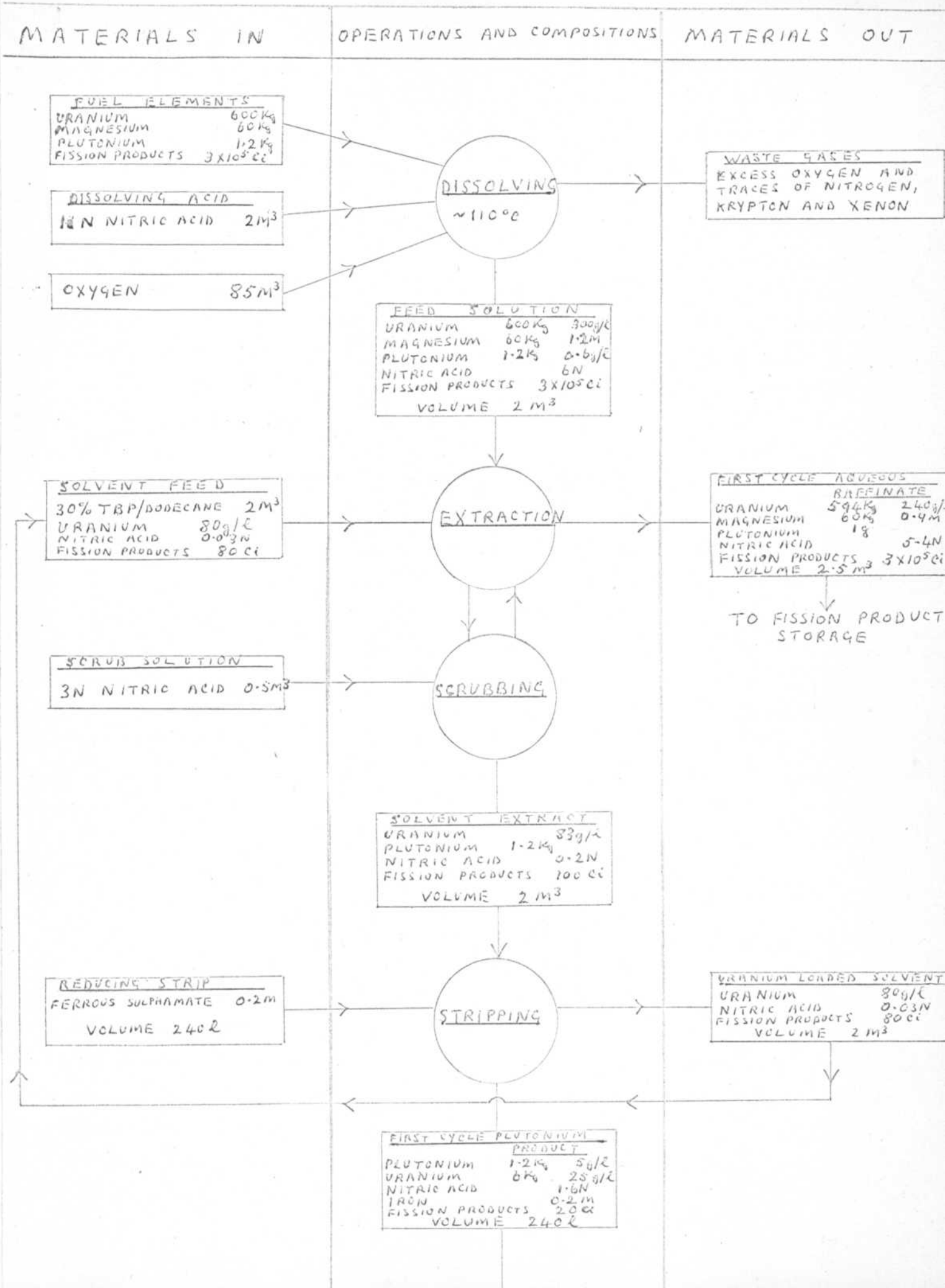
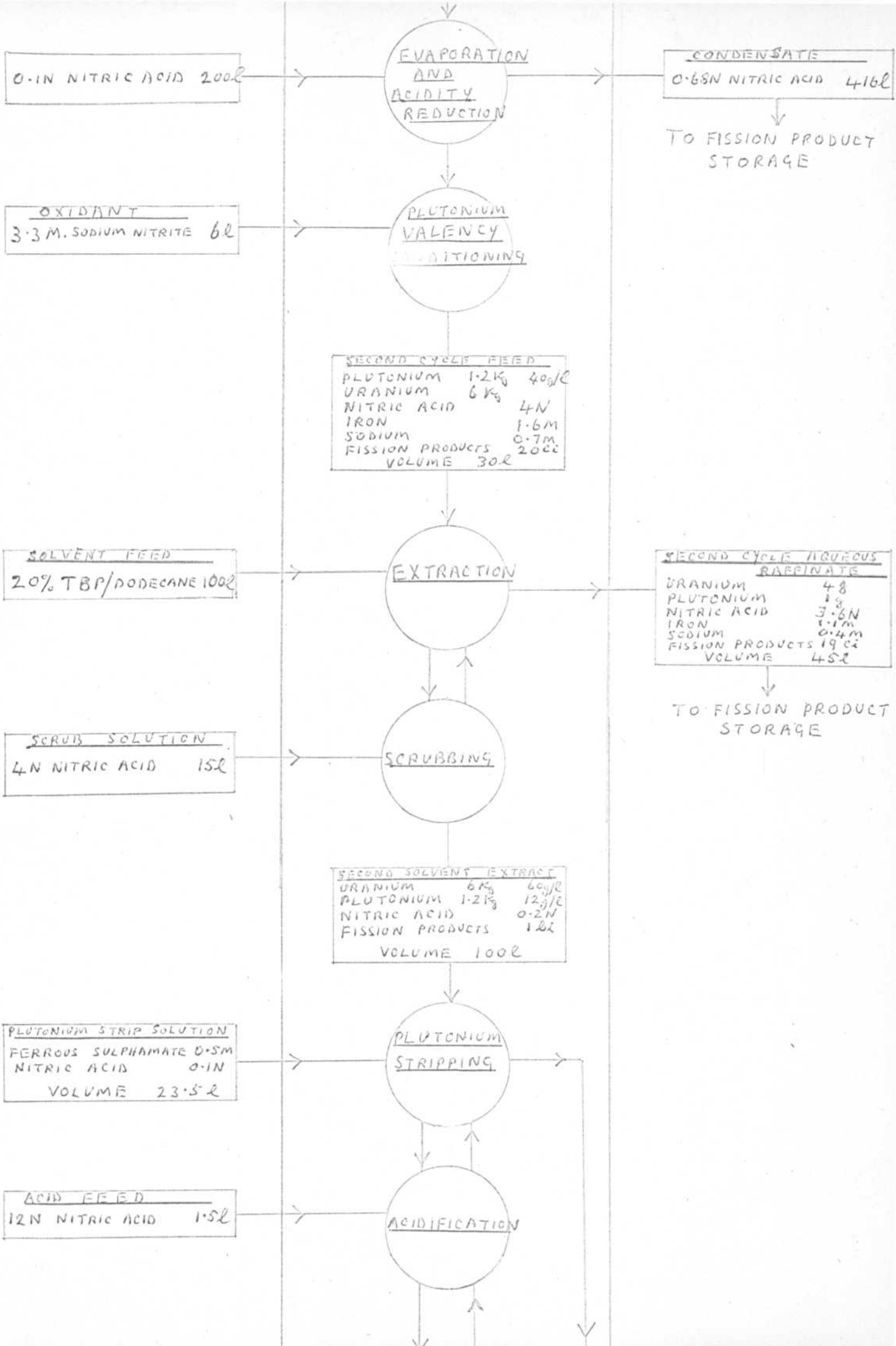
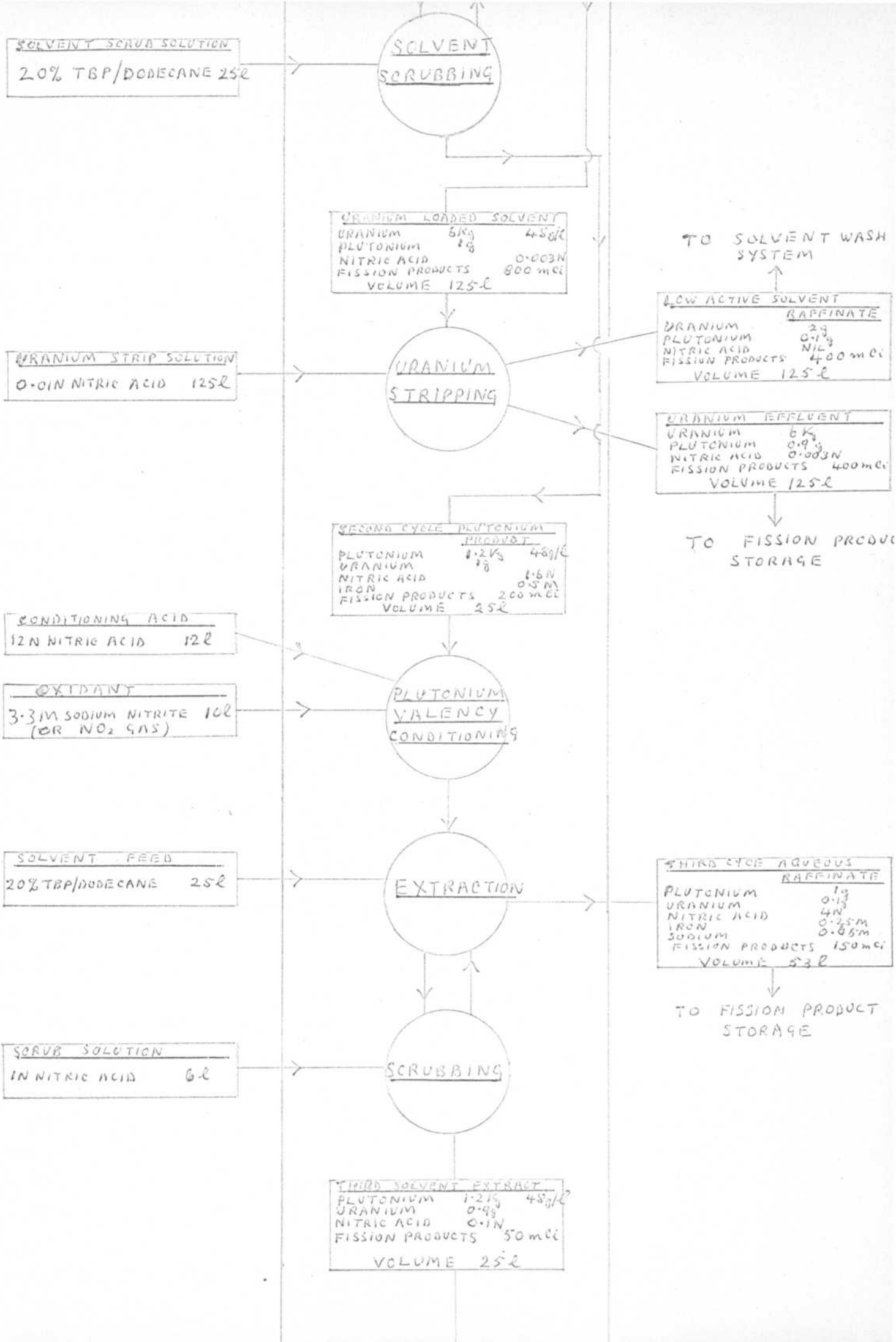
FIGURE 1. FLOWSHEET FOR THE REPROCESSING OF IRRADIATED NATURAL URANIUM FUEL ELEMENTS ON A SCALE OF 0·6 TONS U/DAY




0·1N NITRIC ACID 200ℓ
EVAPORATION AND ACIDITY REDUCTION
CONDENSATE
0·65N NITRIC ACID 416ℓ
TO FISSION PRODUCT STORAGE
OXIDANT
3·3 M. SODIUM NITRITE 6ℓ
PLUTONIUM VALENCY CONDITIONING
SECOND CYCLE FEED
PLUTONIUM 1·2Kg 40g/ℓ
URANIUM 6Kg
NITRIC ACID 4N
IRON 1·6M
SODIUM 0·7M
FISSION PRODUCTS 20ci
VOLUME 30ℓ
SOLVENT FEED
20% TBP/DODECANE 100ℓ
EXTRACTION
SECOND CYCLE AQUEOUS RAFFINATE
URANIUM 48
PLUTONIUM 1g
NITRIC ACID 3·6N
IRON 1·1M
SODIUM 0·4M
FISSION PRODUCTS 19ci
VOLUME 45ℓ
TO FISSION PRODUCT STORAGE
SCRUB SOLUTION
4N NITRIC ACID 15ℓ
SCRUBBING
SECOND SOLVENT EXTRACT
URANIUM 6Kg 60g/ℓ
PLUTONIUM 1·2Kg 12g/ℓ
NITRIC ACID 0·2N
FISSION PRODUCTS 1ℓ
VOLUME 100ℓ
PLUTONIUM STRIP SOLUTION
FERROUS SULPHAMATE 0·5M
NITRIC ACID 0·1N
VOLUME 23·5ℓ
PLUTONIUM STRIPPING
ACID FEED
12N NITRIC ACID 1·5ℓ
ACIDIFICATION


SOLVENT SCRUB SOLUTION
20% TBP/DODECANE 25ℓ
SOLVENT SCRUBBING
URANIUM LOADED SOLVENT
URANIUM 6Kg 48g/ℓ
PLUTONIUM 1g
NITRIC ACID 0·003N
FISSION PRODUCTS 800 mCi
VOLUME 125ℓ
TO SOLVENT WASH SYSTEM
LOW ACTIVE SOLVENT RAFFINATE
URANIUM 2g
PLUTONIUM 0·1g
NITRIC ACID Nil
FISSION PRODUCTS 400 mCi
VOLUME 125ℓ
URANIUM STRIP SOLUTION
0·01N NITRIC ACID 125ℓ
URANIUM STRIPPING
URANIUM EFFLUENT
URANIUM 6Kg
PLUTONIUM 0·9g
NITRIC ACID 0·003N
FISSION PRODUCTS 400mCi
VOLUME 125ℓ
TO FISSION PRODUCT STORAGE
SECOND CYCLE PLUTONIUM PRODUCT
PLUTONIUM 1·2Kg 48g/ℓ
URANIUM 1g
NITRIC ACID 1·6N
IRON 0·5N
FISSION PRODUCTS 200 mCi
VOLUME 25ℓ
CONDITIONING ACID
12N NITRIC ACID 12ℓ
OXIDANT
3·3M SODIUM NITRITE 10ℓ
(OR NO_2 GAS)
PLUTONIUM VALENCY CONDITIONING
SOLVENT FEED
20% TBP/DODECANE 25ℓ
EXTRACTION
THIRD CYCLE AQUEOUS RAFFINATE
PLUTONIUM 1g
URANIUM 0·1g
NITRIC ACID 4N
IRON 0·25M
SODIUM 0·65M
FISSION PRODUCTS 150mCi
VOLUME 53ℓ
TO FISSION PRODUCT STORAGE
SCRUB SOLUTION
1N NITRIC ACID 6ℓ
SCRUBBING
THIRD SOLVENT EXTRACT
PLUTONIUM 1·2Kg 48g/ℓ
URANIUM 0·9g
NITRIC ACID 0·1N
FISSION PRODUCTS 50 mCi
VOLUME 25ℓ

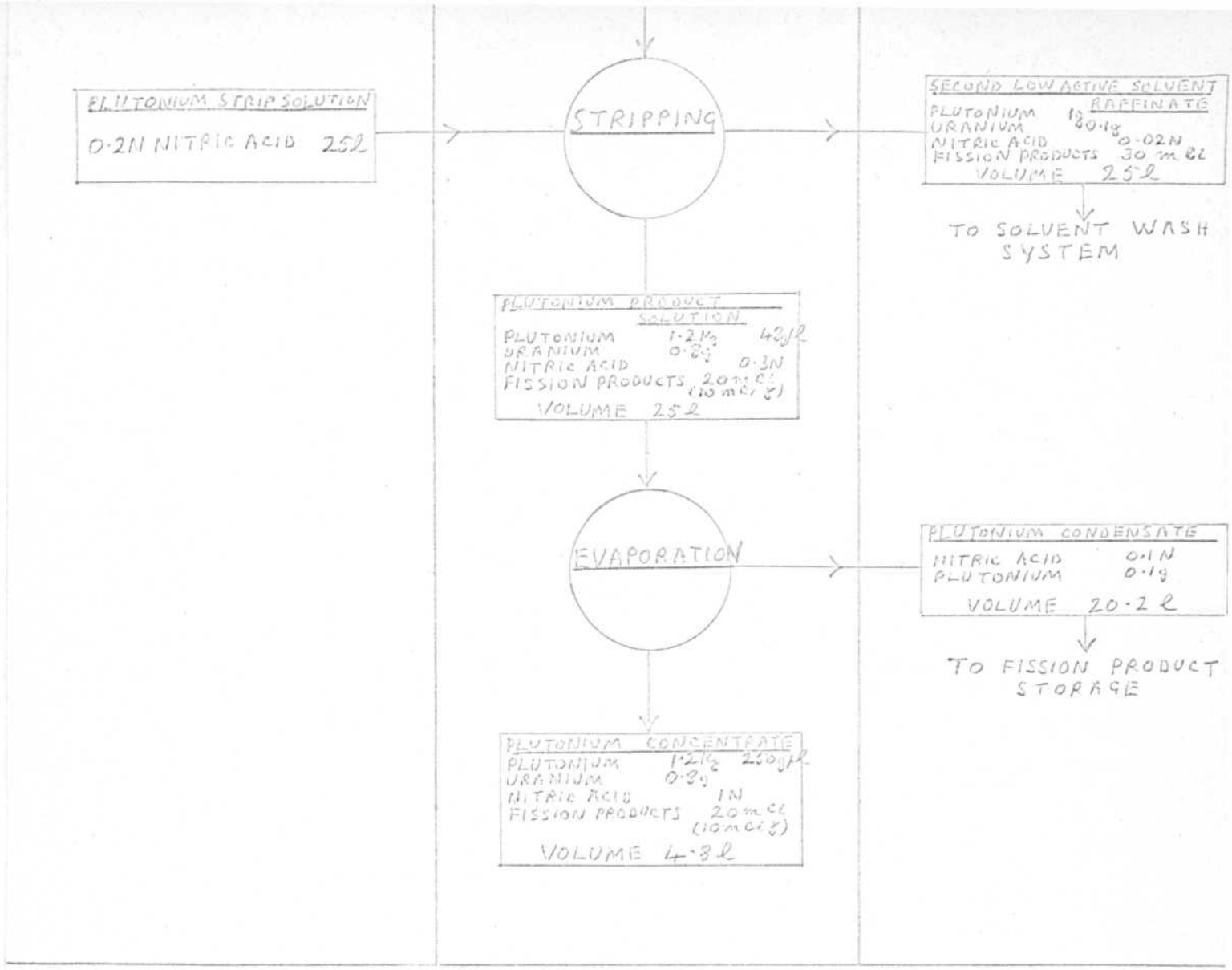
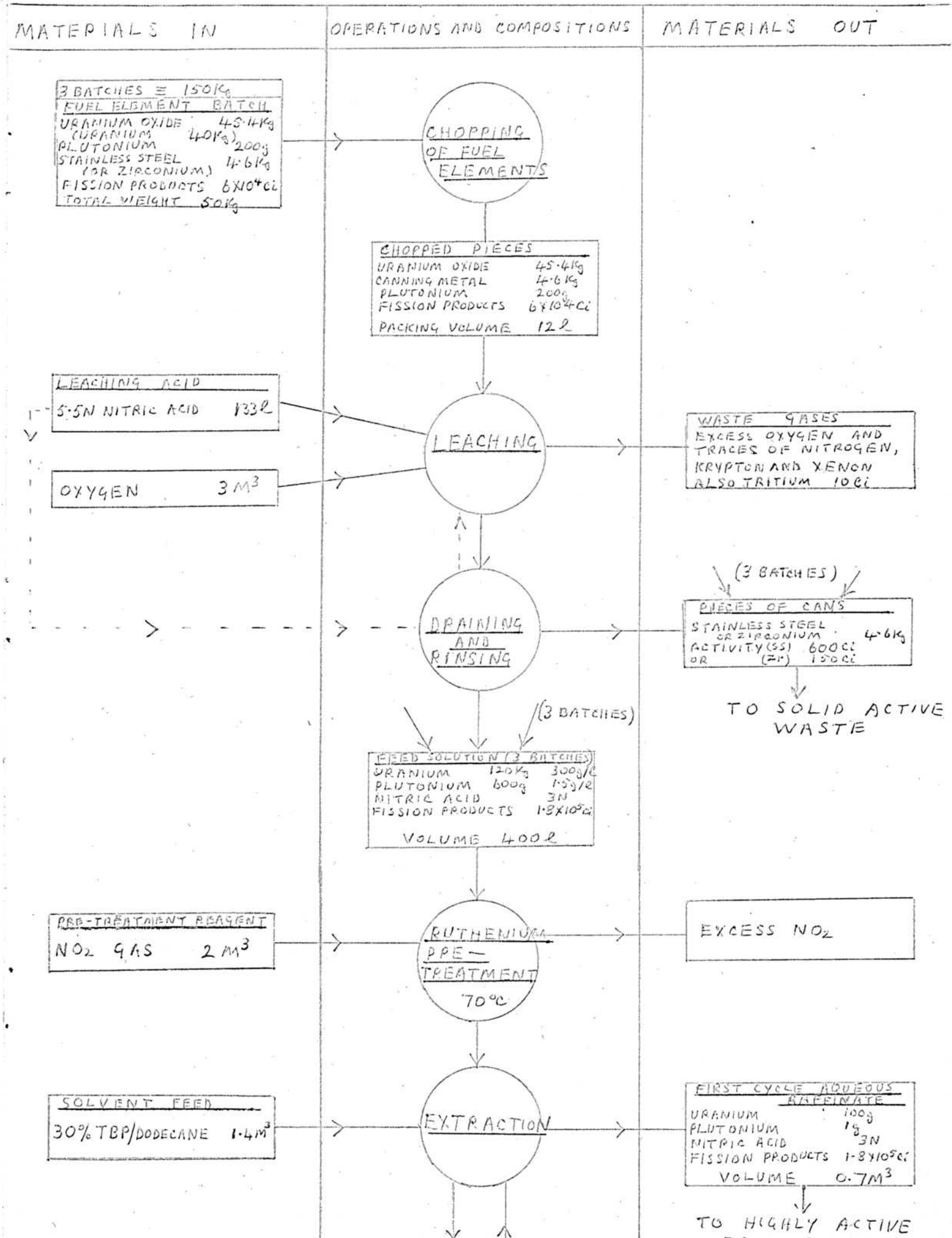
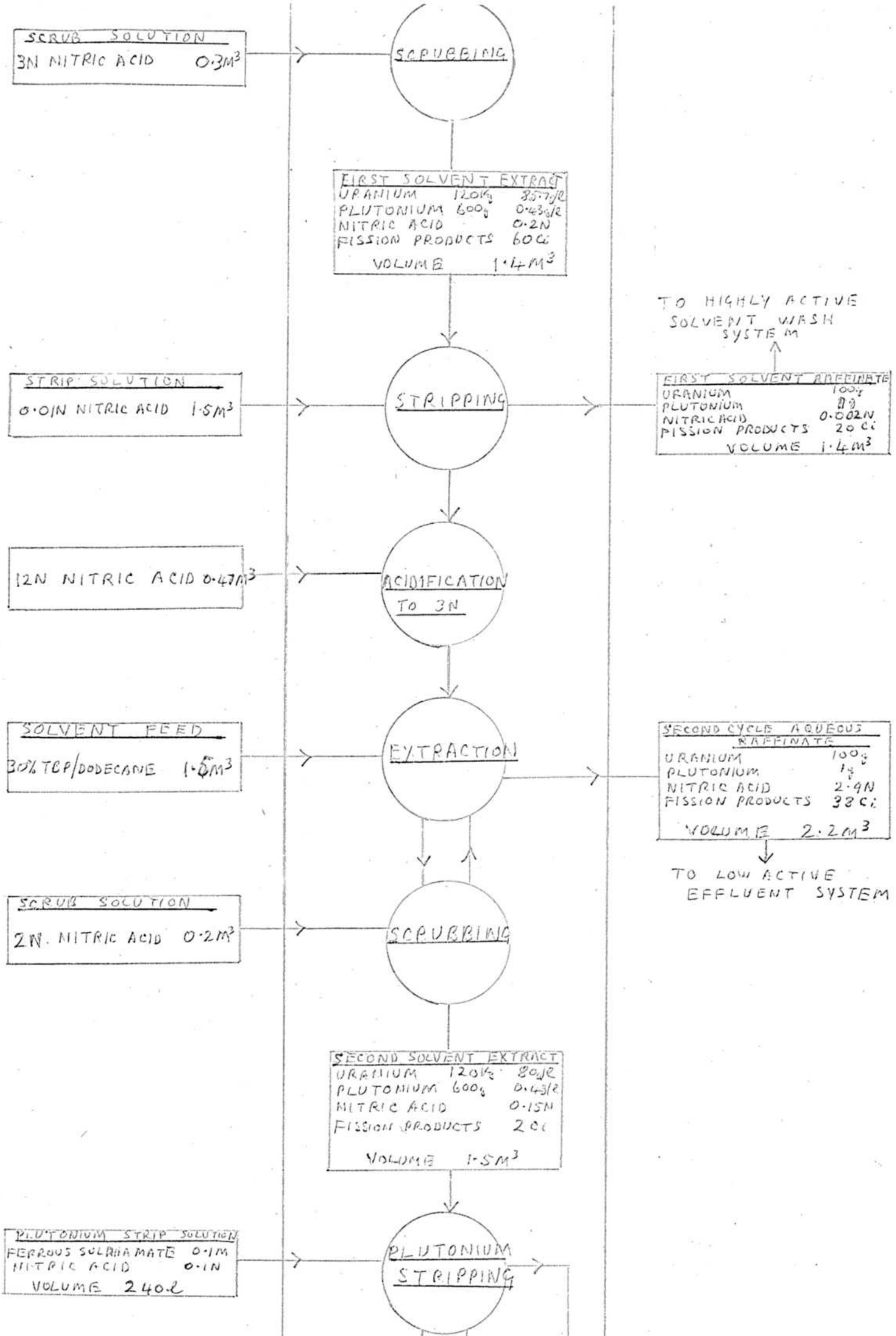
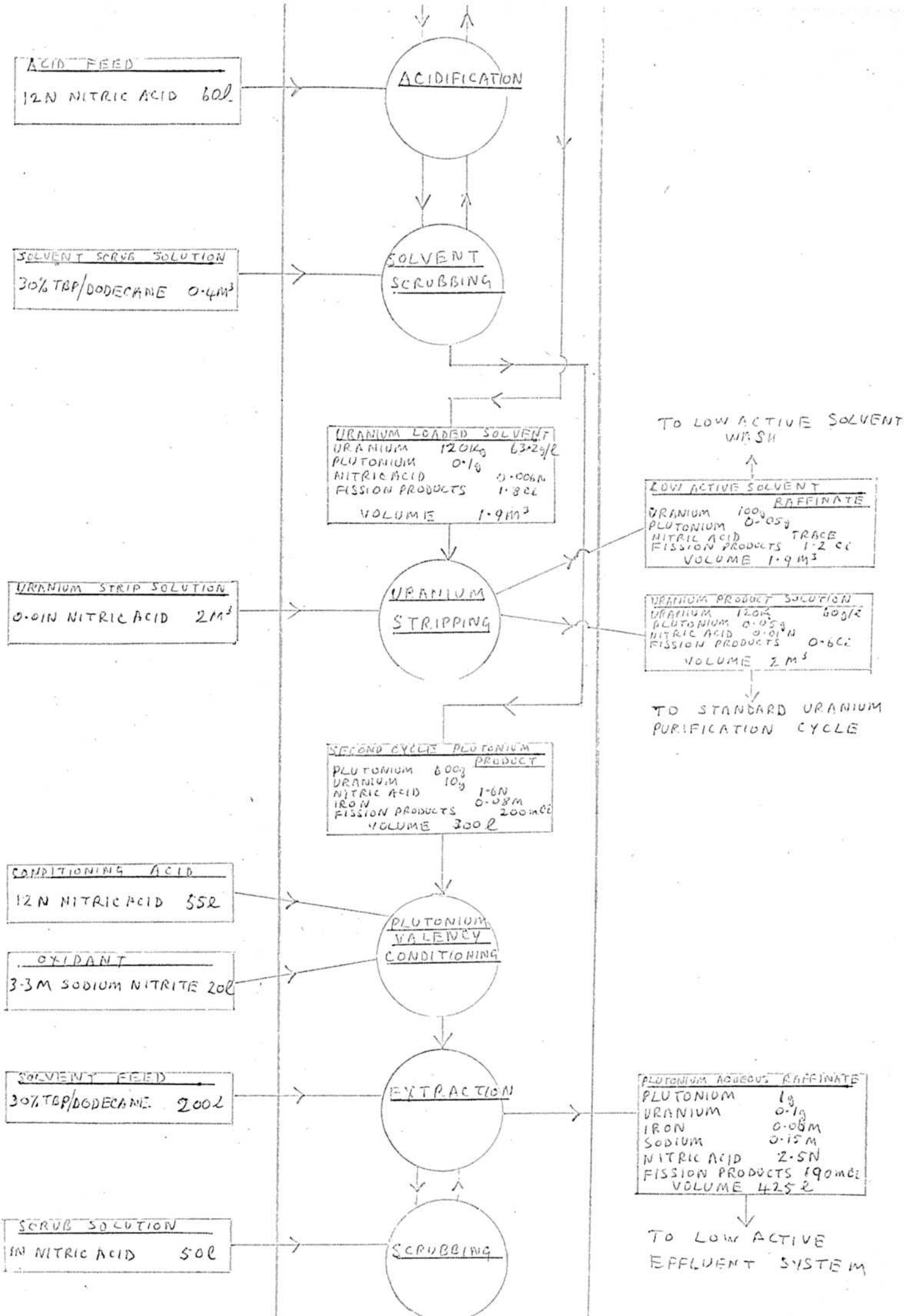


FIGURE 1. FLOWSHEET FOR THE REPROCESSING OF IRRADIATED NATURAL URANIUM FUEL ELEMENTS ON A SCALE OF 0·6 TONS U/DAY

FIGURE 2. FLOWSHEET FOR THE REPROCESSING OF IRRADIATED ENRICHED URANIUM OXIDE FUEL ELEMENTS ON A SCALE OF 120Kg URANIUM PER DAY




SCRUB SOLUTION
3N NITRIC ACID 0.3M³
SCRUBBING
FIRST SOLVENT EXTRACT
URANIUM 120Kg 85.7g/l
PLUTONIUM 600g 0.43g/l
NITRIC ACID 0.2N
FISSION PRODUCTS 60Ci
VOLUME 1.4M³
TO HIGHLY ACTIVE SOLVENT WASH SYSTEM
STRIP SOLUTION
0.01N NITRIC ACID 1.5M³
STRIPPING
FIRST SOLVENT RAFFINATE
URANIUM 100g
PLUTONIUM Nil
NITRIC ACID 0.002N
FISSION PRODUCTS 20 Ci
VOLUME 1.4M³
12N NITRIC ACID 0.47M³
ACIDIFICATION TO 3N
SOLVENT FEED
30% TCP/DODECANE 1.5M³
EXTRACTION
SECOND CYCLE AQUEOUS RAFFINATE
URANIUM 100g
PLUTONIUM 1g
NITRIC ACID 2.9N
FISSION PRODUCTS 38Ci
VOLUME 2.2M³
TO LOW ACTIVE EFFLUENT SYSTEM
SCRUB SOLUTION
2N. NITRIC ACID 0.2M³
SCRUBBING
SECOND SOLVENT EXTRACT
URANIUM 120Kg 80g/l
PLUTONIUM 600g 0.43/l
NITRIC ACID 0.15N
FISSION PRODUCTS 2Ci
VOLUME 1.5M³
PLUTONIUM STRIP SOLUTION
FERROUS SULPHAMATE 0.1M
NITRIC ACID 0.1N
VOLUME 240.l
PLUTONIUM STRIPPING


ACID FEED
12N NITRIC ACID 60l
ACIDIFICATION
SOLVENT SCRUB SOLUTION
30% TBP/DODECANE 0·4m³
SOLVENT SCRUBBING
URANIUM LOADED SOLVENT
URANIUM 120Kg 63·2g/l
PLUTONIUM 0·1g
NITRIC ACID 0·006M
FISSION PRODUCTS 1·8cc
VOLUME 1·9m³
TO LOW ACTIVE SOLVENT WASH
LOW ACTIVE SOLVENT RAFFINATE
URANIUM 100g 0·05g
PLUTONIUM TRACE
NITRIC ACID
FISSION PRODUCTS 1·2cc
VOLUME 1·9m³
URANIUM STRIP SOLUTION
0·01N NITRIC ACID 2m³
URANIUM STRIPPING
URANIUM PRODUCT SOLUTION
URANIUM 120Kg 60g/l
PLUTONIUM 0·05g
NITRIC ACID 0·01N
FISSION PRODUCTS 0·6cc
VOLUME 2m³
TO STANDARD URANIUM PURIFICATION CYCLE
SECOND CYCLE PLUTONIUM PRODUCT
PLUTONIUM 600g
URANIUM 10g
NITRIC ACID 1·6N
IRON 0·03M
FISSION PRODUCTS 200mcl
VOLUME 300l
CONDITIONING ACID
12N NITRIC ACID 55l
OXIDANT
3·3M SODIUM NITRITE 20l
PLUTONIUM VALENCY CONDITIONING
SOLVENT FEED
30% TBP/DODECANE 200l
EXTRACTION
PLUTONIUM AQUEOUS RAFFINATE
PLUTONIUM 1g
URANIUM 0·1g
IRON 0·06M
SODIUM 0·15M
NITRIC ACID 2·5N
FISSION PRODUCTS 190mcl
VOLUME 425l
TO LOW ACTIVE EFFLUENT SYSTEM
SCRUB SOLUTION
1N NITRIC ACID 50l
SCRUBBING

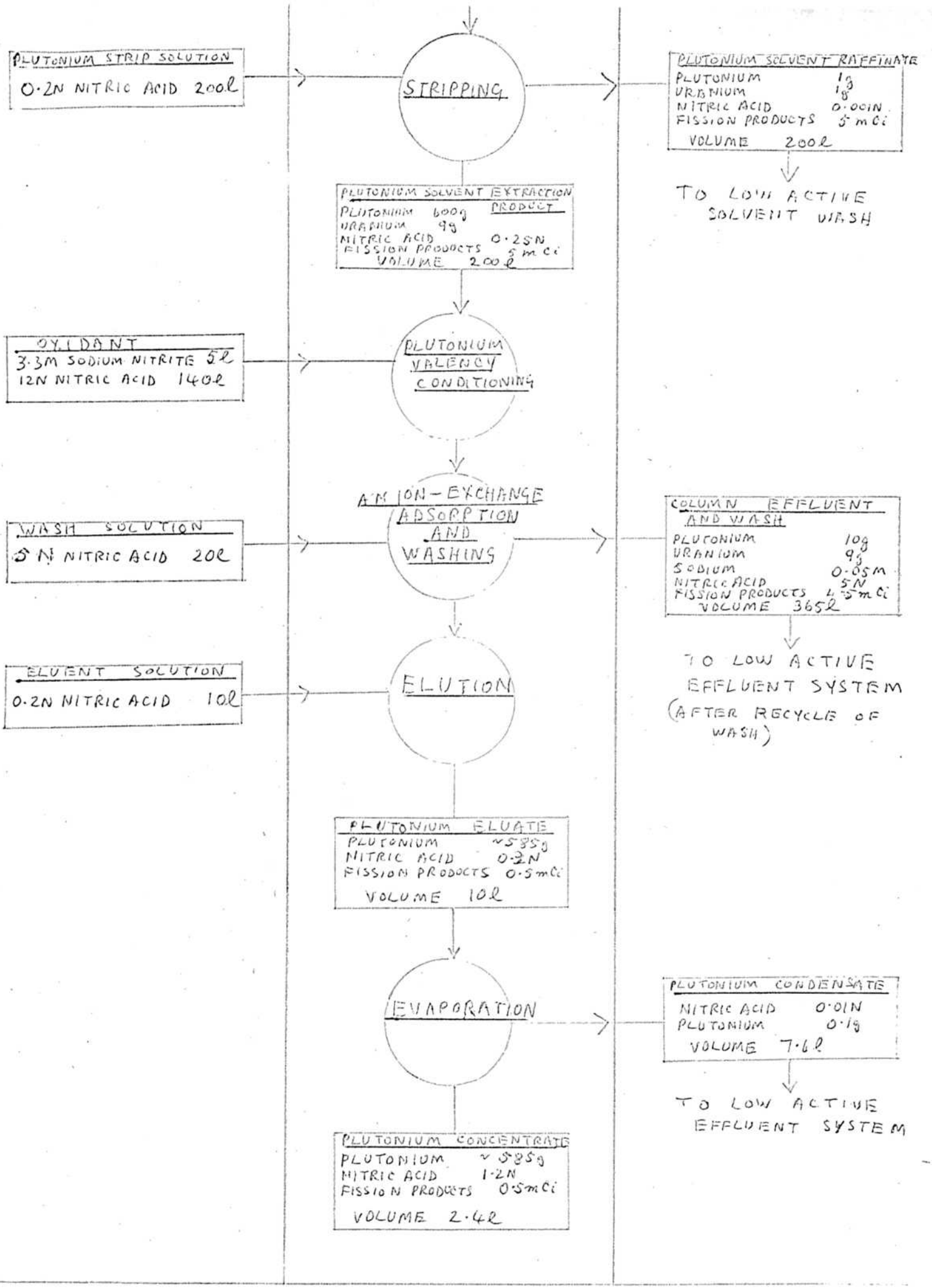


FIGURE 2. FLOWSHEET FOR THE REPROCESSING OF IRRADIATED ENRICHED URANIUM OXIDE FUEL ELEMENTS ON A SCALE OF 120Kg URANIUM PER DAY

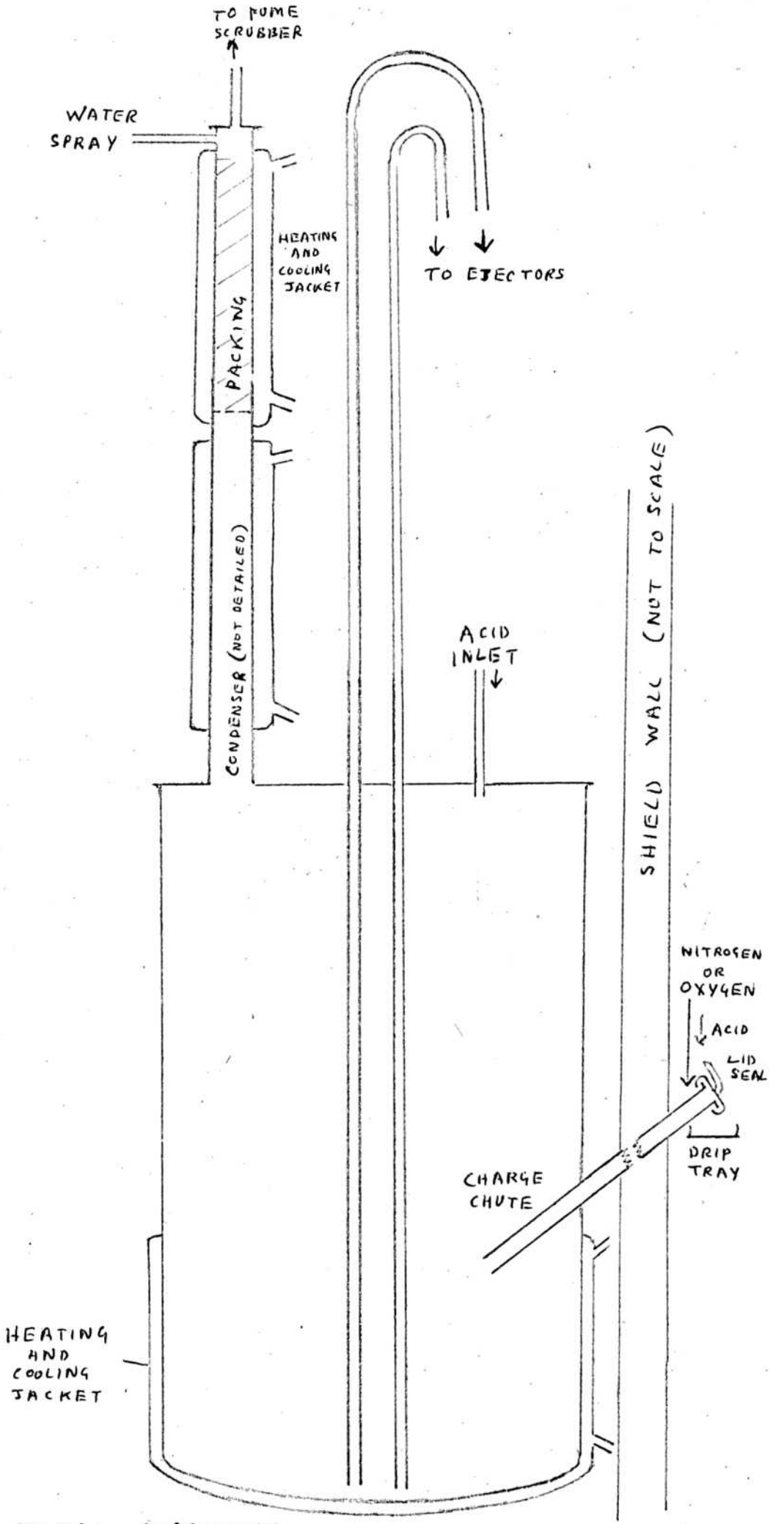


FIGURE 3. URANIUM METAL DISSOLVER

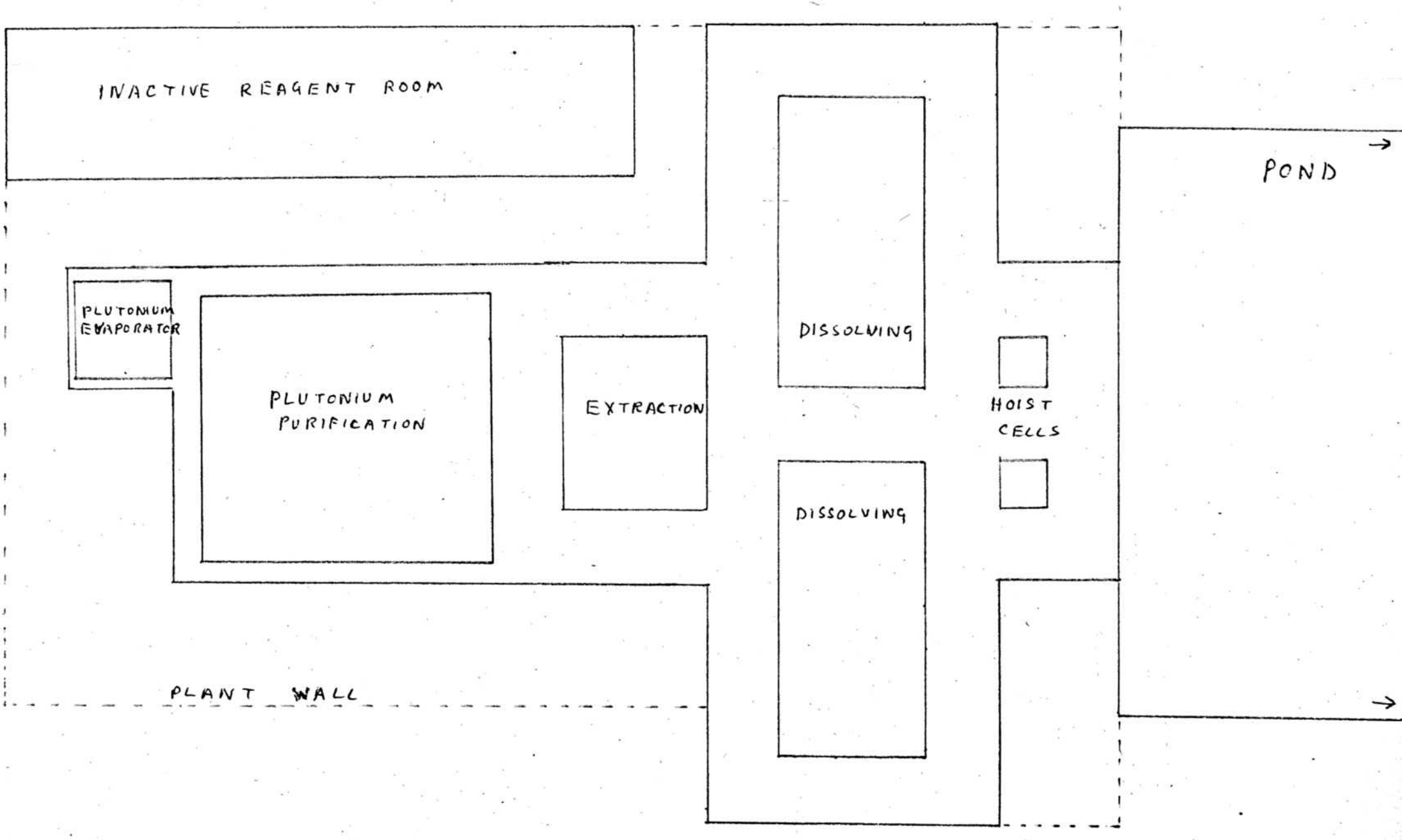


FIGURE 4. ARRANGEMENT OF CELLS FOR 200 T/YR METAL REPROCESSING PLANT

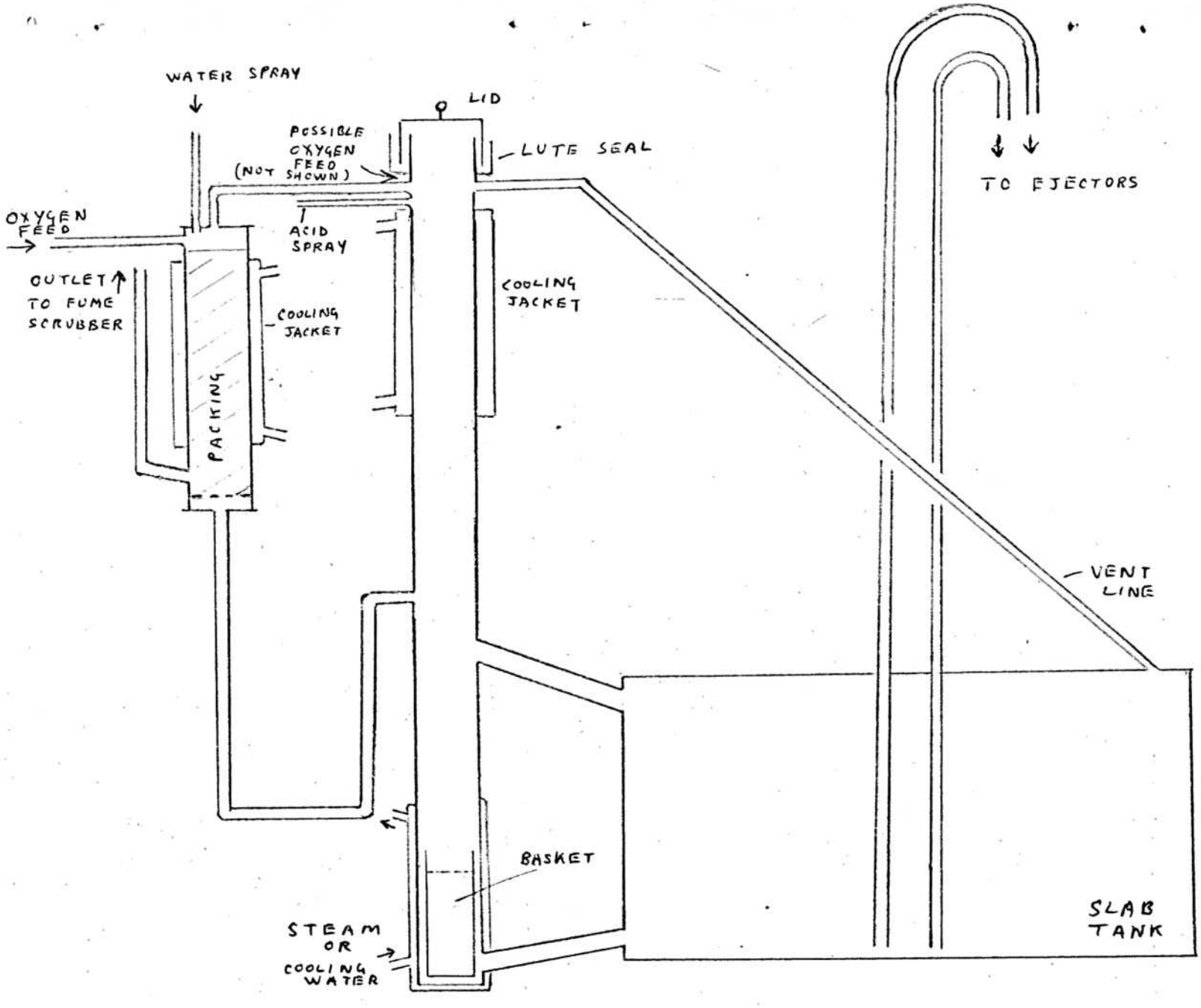


FIGURE 5. URANIUM OXIDE FUEL DISSOLVER

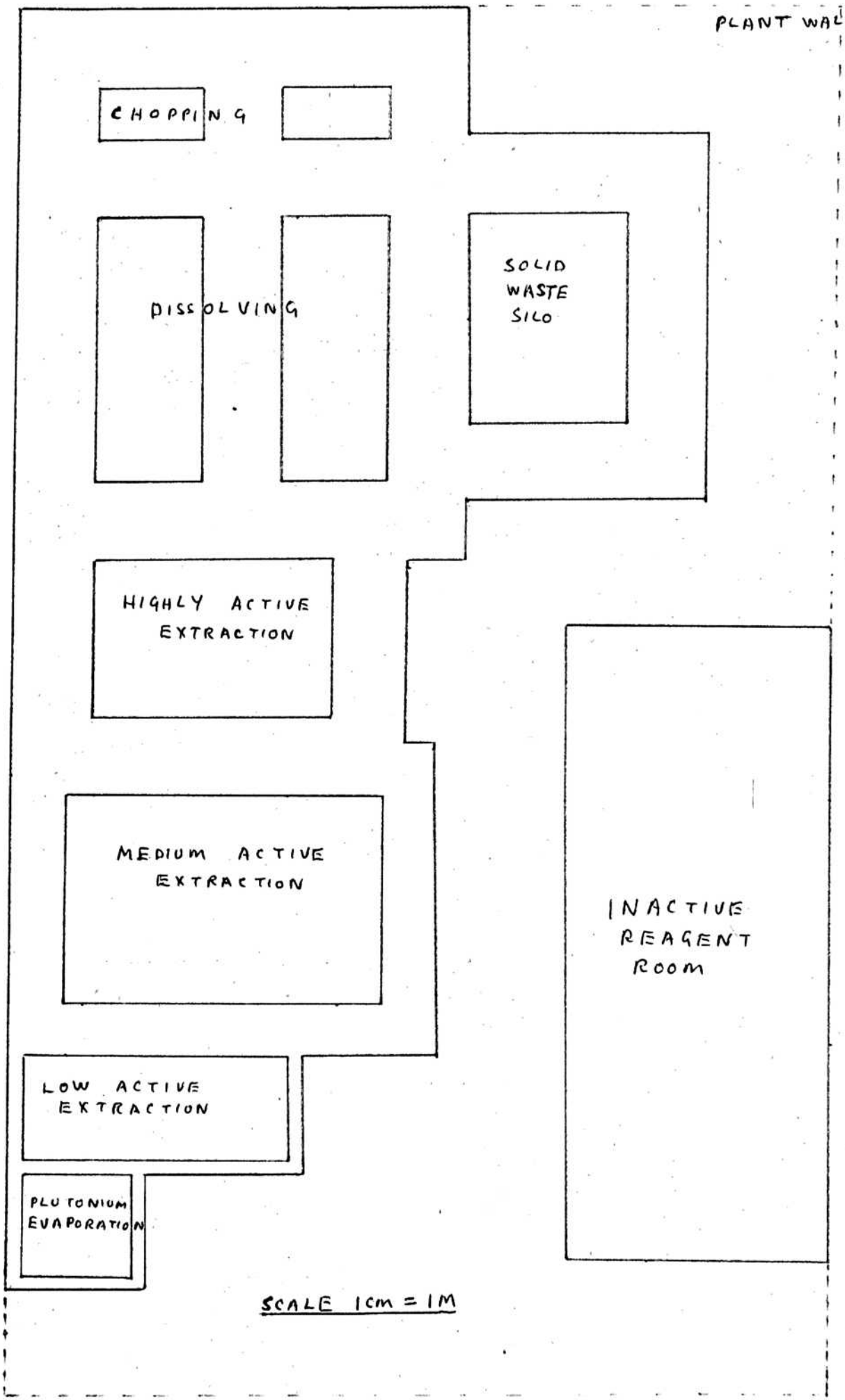


FIGURE 6. ARRANGEMENT OF CELLS FOR 40T/YR OXIDE REPROCESSING PLANT

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