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LUCAS HEIGHTS

REDUCTION OF CAPITAL COSTS IN REPROCESSING POWER REACTOR FUELS
A DESIGN STUDY

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ABSTRACT

The design and cost structures of conventional types of commercial reprocessing plant have been reviewed. Process functions offering the greater potential saving in capital cost have been identified. Unorthodox approaches to philosophies of plant layout, maintenance procedures, and radiation shielding have been formulated with the primary aim of reducing capital cost. Four conceptual designs have been made of a large scale reprocessing plant based on existing shear leach-solvent extraction technology, incorporating novel equipment types and unorthodox design philosophy. Preliminary capital cost estimates for the designs are in the range \$37-41 x 10^6, indicating a 30-37% reduction in comparison with a conventional plant of identical capacity and scope. Additional costs exceeding \$30 x 10^6 could be incurred for refinement to effluent treatment and storage systems envisaged for environmental protection for any type of reprocessing plant.

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BIOLOGICAL SHIELDING; CAPITAL; COMPARATIVE EVALUATIONS; COST; ECONOMICS; FUEL REPROCESSING PLANTS; LEACHING; MAINTENANCE; MANAGEMENT; OPERATION; OPTIMIZATION; PLANNING; PLUTONIUM; POWER REACTORS; RADIATION PROTECTION; RADIOACTIVE WASTE DISPOSAL; RECOMMENDATIONS; REPROCESSING; SOLVENT EXTRACTION; SPECIFICATIONS; SPENT FUELS; URANIUM

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1. SUMMARY

1.1 Residual fissile uranium and plutonium contained in irradiated fuel may be recovered for re-use in the fuel cycle by chemical reprocessing of the spent fuel. The problems of radiological control peculiar to fuel reprocessing result in high capital costs for reprocessing facilities, and in unit costs for reprocessing which are dominated by a capital cost component.

1.2 Economic reprocessing of spent fuels is at present achieved by construction of large capacity plants (3-5 t U/d). Operation of these plants at the required high utilisation demands a reactor programme of 30-50,000 MW(e), and countries with a modest installation programme for reactors cannot take advantage of economies of scale in reprocessing for their national power programme.

1.3 A more fundamental approach in the achievement of economic reprocessing is the reduction of capital costs through selection and engineering of the process employed. It seems unlikely that new processes will be introduced to effect substantial cost saving in relation to the well established commercial solvent extraction processes, and major prospects for cost reduction appear to be in improvement of the design and layout of the chemical plant.

1.4 A critical examination of design and cost structures of existing types of reprocessing complex has shown that the reprocessing plant to separate and purify uranium and plutonium from fission products accounts for 45-66% of the capital investment, the remainder being incurred in receipt and storage of spent fuel, in waste disposal, and in utilities, laboratories, health physics and administration. 70-80% of the reprocessing plant cost is incurred by costs of materials, fabrication and installation of the process building, process equipment and pipework, with the building accounting for 30-40% of the reprocessing plant cost.

1.5 Process functions offering the greatest potential saving in capital cost have been identified as those of Mechanical Pretreatment of Fuel, Leaching, and Solvent Extraction, which together account for about 75% of costs of materials and installation. Possible approaches to obtaining increased economy in building structures, equipment and pipework in these key areas have been reviewed, and include:

- Optimisation of mechanical operations in fuel pretreatment.
- Reduction of shielded volume in mechanical handling areas by elimination of general purpose handling equipment.

- Reduction of shielded volume and pipework costs in areas based on size and shape of vessel rather than segregation of areas on the level of activity handled.

- Investigation of alternative engineering designs of equipment for spatial economy.

- Examination of maintenance philosophy for routine and non-routine maintenance. Modification of procedures for replacement of failed equipment, using handling equipment external to active cells in conjunction with temporary radiation shields constructed when and if failure occurs.

- Examination of potential savings from modification of existing shielding philosophy.

1.6 The wide range of speculative ideas for reduction of capital cost suggested that a greater possibility of success would be derived from a broad based assessment covering a range of themes for cost reduction, incorporating selected combinations of the design variants. A complete plant has been designed around each theme, to make a preliminary comparison of overall costs, and to eliminate uncertainties of requirements and costs of supporting facilities which could result from a differential comparison of isolated plant functions.

Four conceptual designs have been assessed incorporating the following design themes:

1.7 Design 1

A fully shielded plant using adequate shielding for routine operations, with a new approach to maintenance procedures for removal and replacement of failed equipment by constructing temporary shielding as and when failure occurs, and utilising handling tools and equipment external to the cell. This design is based on whole assembly shearing, batch leaching, and solvent extraction in pulsed columns. Process operations have been simplified to reduce mechanical and chemical engineering costs, and building and pipework cost reductions have been achieved by compact design and layout in addition to the savings resulting from the maintenance philosophy adopted.

1.8 Design 2

A variant of Design 1 in which special emphasis has been placed on compaction of the shielded space required to house solvent extraction equipment, substituting compact centrifugal contactors of high capacity in place of pulsed columns.

1.9 Design 3

A design where the majority of active equipment for the high cost areas of mechanical prebreakdown, leaching, and solvent extraction is submerged in a water filled pond. This design involves prebreakdown of fuel assemblies into elements, shearing in a small scale disposable shear feeding a semi-continuous rotary drum leacher, with pulsed column solvent extraction equipment.

1.10 Design 4

A design based on application of protective radiation shielding to the plant personnel rather than to the plant equipment. The active plant is contained underground, with access for performance of active handling and maintenance operations by an operator in a mobile shielded cabin; all inactive functions are carried out above ground level in a radiation exclusion zone.

1.11 All designs have been based on a common plant specification - a plant capacity of 3 t U/d of fuel discharged from light water reactors, producing purified solutions of uranium and plutonium nitrates, and conforming to present basic requirements in effluent treatment in storage of high and low level liquid wastes as concentrated solutions. Published estimates of requirements and costs for more sophisticated processes such as fixation of high level wastes, removal of ^{85}Kr and $^{3}H_{1}$, and recovery of ^{237}Np, which are not included in the designs, have been reviewed.

1.12 Each design has been developed in sufficient detail for derivation of preliminary estimates of capital cost. Detailed information on plant technology and equipment requirements for the commercial reprocessing facilities of Nuclear Fuels Services, N.Y., Allied-Gulf Nuclear Services (Barnwell, Carolina) and General Electric Company (Morris, Illinois) were considered in the designs. The Nuclear Fuels Services plant was selected as a basis for flowsheet and equipment sizing considerations, as it relied on proven technology, its design, cost breakdown and operational history were well documented, and because its design could be adapted to the capacity requirements of the study with marginal resizing of equipment items. This approach eliminated the need to develop a plant design from first principles, and permitted the study to concentrate on alternatives in plant equipment and layout in high cost areas.

1.13 A chemical flowsheet common to the four plant designs has been drawn to define requirements for plant equipment sizing operations. Equipment flow diagrams have been detailed for plant functions including batch and semi-continuous shear-leaching, solvent extraction, treatment and disposal of plant solid, liquid and gaseous effluents including recovery processes, and concentration and final purification of uranium and plutonium product solutions.

The size and shape of process equipment items has been determined, and equipment layout studies conducted to define the area and height requirements for containment of active process functions. Building elevations, plans and sections have been drawn for each of the designs. A tentative site layout has been made for Design 1 to indicate requirements for land and site improvements.

1.14 A preliminary estimate of overall plant costs was regarded as an aim of the study; estimates have been made in as much detail as was permitted by the manpower and timescale allocated to the work, to minimise underestimation of overall cost through omission of plant items.

1.15 All costing has been carried out in mid-1972 Australian dollars with Sydney as a reference location. Initial studies indicated that a satisfactory site could be in Southern Central Australia close to the Main East-West Standard Gauge Railway and the geology of that area has been assumed for costing purposes. Costs for buildings and structures were estimated by civil engineering consultants, who confirmed the structural adequacy of the proposals.

1.16 The cost of active chemical engineering equipment was derived by detailed assessment of a number of representative vessels, to obtain relationshipsbetween equipment size and cost for a variety of basic types of vessel. Estimates for materials and fabrication were derived from standard costing reference works, by direct quotations from Australian suppliers and fabricators, and from published costs for some equipment in the Nuclear Fuels Services plant, converted and updated using the Australian implicit price level index. Process vessels designed for reprocessing service were found to incur significantly greater fabrication costs than conventional equipment fabricated to pressure vessel standards, possibly because of higher fabrication standards and more extensive inspection procedures demanded. Installation costs of process chemical plant was taken as 21% of the ex-works costs, conforming to the NFS average. Costs of process mechanical equipment were obtained where possible from published information on similar equipment in the Nuclear Fuels Services plant.

1.17 A detailed estimate of process pipework costs was made for the first uranium purification cycle. Equipment layouts in conjunction with engineering flow diagrams were used to determine piping layouts and fittings, and costs determined from a detailed assessment of material costs and fabrication time. The data wereextended to other plant functions by prorating estimates from the detailed calculations according to the number of lines indicated on the engineering flow diagrams.

1.18 Costs for process instrumentation and control were determined on a unit rate basis by extraction of requirements from the engineering flow diagrams. Analytical equipment costs were based on updated NFS data. Unit costs for utilities were obtained by direct quotation from suppliers or from the literature. Estimates for reticulation of services and for site improvements were derived from the conceptual site layout. Costs of auxiliary buildings were derived by application of unit rates supplied by civil engineering consultants to area requirements obtained from NFS data. The liquid waste storage facility was not designed; costs were updated from published information.

1.19 Overall capital costs for the four reprocessing plant designs have been estimated by determination of the direct construction cost (installed cost of individual components of plant and structure) together with an engineering, construction and contractors fee assessed at 27% of direct construction cost, and a 15% contingency allowance. Interest on construction funds has assumed a 3 year construction period with successive yearly investments of 20%, 30% and 50% of capital, based on a debt:equity ratio of unity. The costs also include working capital (updated from NFS estimates), pre-operational costs including operating costs and interest on capital during a nine month commissioning period, and a remote site loading of 5%.

1.20 Comparable costs for plants of 1 and 5 t U/d capacity have been estimated from a cost capacity exponent of 0.4 usually accepted for fuel reprocessing facilities.

2. CONCLUSIONS

2.1 Preliminary capital cost estimates in the range \$A36.5-40.5 x 10^6 have been derived for 4 designs of a chemical reprocessing plant, capacity 3 t U/d, in which an unorthodox approach to layout and maintenance philosophy has been linked to existing shear-leach-solvent extraction technology. The estimates indicate a cost reduction through design of up to 30-37% in comparison with an interpolated cost estimate of \$A58 x 10^6 for a conventional plant of identical capacity and scope, producing uranium and plutonium as nitrate solutions, and with 5 years storage capacity for fission product waste solutions. The estimates exclude development costs. Additional costs exceeding \$A30 x 10^6, dependent on philosophy, could be incurred for refinements to effluent treatment and storage systems envisaged for environmental protection for any type of reprocessing plant whether conventional or unconventional. The additional costs could be incurred mainly by a highly active waste storage vault (\$A11-18 x 10^6), facilities for removal of tritium (\$A10 x 10^6), and fixation of highly active wastes ($A5-7.5 x 10^6). Smaller costs could be incurred for removal and storage of noble gases ($A2.8 x 10^6) and radioiodine ($A0.8 x 10^6).

2.2 Lowest overall capital costs were indicated for a design featuring use of a water radiation shield for some highly active plant areas, while the greatest costs were obtained for a plant deriving radiation protection from ground and air exclusion zones, with active operations performed by an operator in a mobile shielded cabin. Intermediate estimates apply to a plant featuring a compact and unorthodox layout of equipment, avoidance of duplication of facilities, elimination of general purpose handling equipment in active cells, and with active maintenance performed by decontamination and erection of temporary radiation shields as and when necessary. The preliminary estimates are substantially below predictions for conventional plant. The apparent saving is sufficient to absorb a significant under-estimate and still leave a worthwhile capital saving from the radical approaches to design.

2.3 Comparison of equipment options for the head end shear-leach operation have indicated that fuel prebreakdown and semi-continuous leaching below a water shield could effect some economy over whole assembly shearing and batch leaching in a dry cell, maintained by decontamination and construction of temporary shielding facilities. The greatest potential for capital cost reduction in this function appears to be derived from savings in civil costs through elimination of fuel prebreakdown operations in shielded dry cells, and through the elimination of general purpose handling equipment and shielded cells for remote replacement of failed plant items.

2.4 A more compact plant design with some pipework economy resulted from employment of centrifugal contactors in place of pulsed column equipment, but an increase in overall cost for the solvent extraction function resulted through the relatively high cost of the centrifugal contactors.

2.5 Non-routine operations are concerned with removal and replacement of failed plant or clearance of plant blockages. The likelihood of failure is unknown; it is more probable in leachers and evaporators which operate with corrosive solutions at high temperatures than in the relatively low corrosive environment of solvent extraction. The attitude in the United Kingdom has been to construct plant to high standards with duplication in some areas, and accept a low probability of failure which would require repair by contact maintenance following a protracted decontamination and plant outage of several months. The USA plants adopt this approach for solvent extraction plant, but provide for remote replacement of leachers and evaporators without a prolonged shutdown of plant.

2.6 Design 1 incorporates adequate equipment for routine active handling procedures, which have been simplified through installation of handling equipment designed for the specific routine operations. A compromise between the UK and USA approaches has been adopted for removal and replacement of failed plant; equipment would be partially decontaminated to permit removal by handling equipment external to the cell operated behind a temporary biological shield. This scheme would require development of equipment modules designed for removal; in an extreme case the plant could be maintained by entry into the active cell after protracted decontamination. This design eliminates the capital cost inherent in the USA approach and is no less operable than the equivalent UK plant.

2.7 Design 2 is similar to Design 1 in maintenance provisions for the shear-leach and evaporator area; the use of centrifugal contactors imposes further requirements for maintenance. These contactors may be replaced after preliminary decontamination by shielded flasks carrying lifting equipment. The design offers no economic advantages over Design 1.

2.8 Design 3 appears to offer advantages both in economics and maintenance. Fuel prebreakdown conducted underwater has the advantage of visibility without incurring high costs inherent in a shielded dry cell. Similarly, removal and replacement of partially decontaminated plant under a water shield appears preferable to removal requiring operations conducted behind temporary remote shields, as in Designs 1 and 2. Design 3 involves risk of contamination of pond water; an allowance for equipment to remove activity from the water has been made but the design inevitably increases the risk of spread of contamination. The advantages of Design 3 in the head end shear-leach area require satisfactory development of a semi-continuous leacher to eliminate handling of active baskets containing leached cladding. The scheme appears to offer cost advantages, though in view of the novel equipment incorporated, the costing for scheme 3 is more susceptible to under-estimation than for Designs 1 and 2.

2.9 Design 4 appears to offer no cost advantage over the alternative designs considered. It is doubtful if one shielded mobile cabin is capable of conducting the routine active operations in the timescale required. If the head end shear-leach operation were modified to eliminate the need for active handling by the mobile cabin (e.g. by incorporation of the underwater head end facility of Design 3), then the cabin would be used for sampling and replacement of active instrumentation, changing of filters, and for replacement of plant in the event of failure. The residual routine functions could be performed by means alternative to the use of a mobile cabin, and the utility of the latter would be dependent on the maintenance capability and requirement for replacement of failed plant. For equipment constructed to high standards of containment, the probability of failure is low; the mobile cabin philosophy would hence imply that a reduction in standards in construction of active plant could be exploited to derive a reduction in capital costs. The design offers poor operating conditions for personnel in the mobile. It has disadvantages in guaranteeing adequate control of contamination and radiation, and results in capital costs similar to those estimated for alternative designs more in keeping with international standards for radiological handling procedures.

2.10 The narrow range of cost estimates derived from diverse approaches to economy through design indicates the high cost inherent in production scale reprocessing operations employing existing technology. It suggests that the design theme used for cost reduction should be examined against criteria other than capital cost. Thus the apparent advantage in reduction of capital charges in reprocessing should be assessed against the likelihood of major plant failure, and the penalty of an increased plant outage for maintenance by ad hoc methods. Similarly, any design featuring reduction in the degree of radiological shielding or containment to achieve capital cost economy should be assessed against the increasing stringency of world standards for the nuclear industry.

2.11 The most promising aspect for development work to reduce capital costs in fuel reprocessing by existing technology appears to be in the field of mechanical handling. This work could include development of special purpose tools for routine process operations (e.g. basket movements, or maintenance of shear packs), and evaluation of maintenance techniques for removal and replacement of active equipment liable to failure, using lifting devices external to the active cell in conjunction with temporary radiation shields. This approach would extend the trend towards contact maintenance of partially decontaminated plant, accepted for solvent extraction equipment.

2.12 The type of mechanical handling equipment and maintenance facilities suggested for capital cost reduction would best be developed in conjunction with detailed plant design work. It is not appropriate for the Research Establishment to contribute to this field in the absence of a commitment for an Australian reprocessing plant. More general development work which could be considered by the RE could be:

- Continuous moving bed ion exchange contacting equipment.

This will probably be required for reprocessing operations with plutonium, but could be of general interest in uranium milling and to the metallurgical industry.

- Equipment for electrolytic reduction and separation of uranium and plutonium. Experience gained could be of general use in alternative manufacturing routes for UF_4 in fuel manufacture.

- Rapid decontamination techniques, including electrolytic deplating procedures. This could be of general interest in decontamination of active equipment arising from HIFAR, in addition to its specific use in treatment of shear packs.

3. RECOMMENDATIONS

It is recommended that the Commission:

- Note the unconventional approaches to design and maintenance philosophy likely to reduce capital cost,

- Propose a more detailed examination of these principles in the event of a firm commitment for an Australian reprocessing plant.

- Consider background development work in the reprocessing field where the results are of general application to the uranium industry.

- Consider an examination of possible savings through design of the repository for highly active wastes.

1. INTRODUCTION

Most modern thermal reactors are fuelled with uranium oxide enriched in the fissile uranium isotope, ^{235}U. Energy is generated by fission of ^{235}U, and, to a lesser extent, by plutonium produced as a by-product. The reactivity of the fuel falls progressively through depletion in fissile content, and accumulation of fission products which absorb neutrons required to sustain the chain reaction. Fuel is discharged when its reactivity is no longer able to generate the required power output.

The discharged fuel contains appreciable quantities of fissile uranium and plutonium, which could be recovered for reintroduction into the fuel cycle.

Waste Management of the spent fuel involves either:

(a) permanent safe storage of the fuel assemblies, or

(b) reprocessing and safe storage of fission products with recovery of uranium and plutonium.

The value of the fissile content of the spent fuel for recycle is equal to the cost of an equivalent quantity of enriched uranium. Reprocessing is worthwhile economically only if the reprocessing cost is less than the value of the fissile content, plus the cost of storage of unreprocessed fuel.

Chemical reprocessing plants are usually regarded as expensive, although this reputation is largely undeserved in the overall context of a large, growing, nuclear industry (ORNL 1970a). The reprocessing operation is highly capital intensive, because spent fuel is intensely radioactive, and the chemical reprocessing plants pose special problems in radiological control. Capital charges may account for 50-70% of the total cost of reprocessing power reactor fuel (Detilleux 1972, Terjesen 1970, Schuller 1970, Rodger and Reece 1969), the proportion depending on the ground rules adopted in costing (i.e. amortisation period, rate of interest, debt equity ratio, utilisation factor, etc.). The capital investment is the largest single factor in the unit cost of reprocessing.

The optimum plant size for economic reprocessing appears to be in the range 3-5t U/d (Marcus and Seynaeve 1971) and the trend is towards centralised plants to take advantage of the economy of scale. However, a large capacity plant serving an escalating reactor programme may lose part of its economy through under utilisation in early years of operation before its capacity is matched by the discharge rate from the reactors.

Increased transport charges arise for large capacity plants serving a wide area of reactor sites. Countries such as Australia with a modest reactor programme cannot take full advantage of economies of scale in reprocessing.

Construction of large capacity plant is therefore only a partial solution to the high capital cost of reprocessing plant. A more fundamental approach is to achieve economy through selection and engineering of the process employed.

It seems unlikely that new processes will be introduced to bring substantial cost savings in relation to the well-established solvent extraction processes now used in almost all plants overseas, and which are capable of processing advanced fuels of short cooling time. The major prospects for cost reduction in chemical reprocessing appear to lie in improvement of the design and layout of chemical plant, along lines suggested by Baczewski (1958).

This document reports studies, based on existing solvent technology, of a single purpose plant to process LWR oxide fuel, to investigate the effect on capital cost of alternatives in equipment choice, in design and layout, and in philosophies of shielding and maintenance.

2. TRENDS IN REPROCESSING TECHNOLOGY, CAPACITY, AND ECONOMICS

Processes for treatment of irradiated fuels fall into three categories: dry processes, aqueous processes, and processes involving both dry and aqueous steps. The merits of these alternative categories have been reviewed by Costello (1972a).

Dry techniques such as pyrochemistry and pyrometallurgy have a low decontamination efficiency, and lead to an increase in capital costs in remote refabrication of fuel (Levensen et al. 1966). Fluoride volatility techniques offer high decontamination efficiency, but the capital cost of a conceptual production plant estimated by Breton et al. (1968) was similar to that of an aqueous process of identical throughput. Furthermore the technology of dry processes has not been developed to the stage where a wholly dry plant could be designed with confidence.

Single purpose reprocessing plants, designed to handle only one type of fuel (e.g. LWR fuels) are potentially a more economical choice for that specific fuel type than a plant with capability for processing a range of fuel types. The additional capital investment for the multipurpose plant is chiefly derived from the multiplicity of head end treatments required.

The following estimate (ORNL 1968) illustrates the additional costs incurred by multiple fuel reprocessing plant:

Plant Capability	Fuel Type	Capital Cost US$(1968)
Basic Capability	Light Water Reactor (LWR) oxide fuel - 1 t U/d (260 t U/a)	29.0×10^6
Additional Capability	1. Plutonium recycle LWR fuel - 1 t U/d (260 t U/a) 2. Heavy water organic cooled reactor (HWOCR) thorium fuel - 0.5 t Th/d	2.5×10^6
	3. Fast breeder reactor fuel (FBR) - 0.5 t (U + Pu)/d	5.0×10^6
	4. Sintered alumina clad fuel (Some HWOCR types) 5. Carbide fuels (some HWOCR and FBR types) 6. Sodium bonded fuels (FBR types)	1.0×10^6
	7. High temperature gas cooled reactor (HTGR) Thorium-uranium fuel	9.0×10^6

Estimates made by Terjesen (1970) of total capital costs of conventional reprocessing facilities, inclusive of ancillary costs and interest during construction, updated to equivalent Australian currency on a 1972 basis, indicate capital costs of \$A43 x 10^6 and \$A 80 x 10^6 for plant capacities of 300 and 1500 tonnes per year, respectively.

The effect of plant capacity and utilisation on the overall cost of reprocessing and fission product waste storage is shown in Table 1 and the value of uranium and fissile plutonium recovered from BWR and PWR assemblies at equilibrium irradiation is shown in Table 2. The tables were based on isotopic compositions of uranium and plutonium in discharged fuel given by Jackson and Moreland and Stoller Associates (1968) and values of uranium (as UF_6) from the USAEC schedule of base charges (USAEC 1969b) and an assumed value of US\$8-10/g Pu fissile.

Fissile plutonium accounts for the major proportion of recovery value in both fuel types; the value of uranium is significantly greater for PWR fuel.

The tables show that for the ground rules adopted a plant capacity of 300 t U/a even if operated at 100% utilisation is uneconomic for treatment of BWR fuels, whereas a plant capacity of 1500 t U/a can achieve economic recovery at load factors of about 35-40%.

TABLE 1

ILLUSTRATIVE OVERALL COSTS FOR REPROCESSING

Plant Capacity	1 t U/day (300 t U/a)	5 t U/day (1500 t U/a)		
% Plant Utilisation	100%	20%	50%	100%
	US$/t U	US$/t U	US$/t U	US$/t U
1. Fuel Assembly Transport	6,000	6,000	6,000	6,000
2. Reprocessing Inclusive of Aqueous Storage of Fission Product Wastes				
Capital Charge (at 30%)	40,000	75,000	30,000	15,000
Constant Operating Cost	10,000	14,000	6,000	3,000
Variable Operating Cost	5,000	4,000	4,000	4,000
3. Additional Charge for Storage of Fission Products as Solids	8,500	8,500	8,500	8,500
Total Cost of Reprocessing US$/t U	69,500	107,500	40,500	34,500

TABLE 2

VALUE OF URANIUM AND PLUTONIUM FROM REPROCESSING

	Unit	Reactor Type	
		BWR	PWR
^{235}U Content in Irradiated Fuel	%	0.62	0.84
Fissile Plutonium Content	kg/t U	5.54	6.47
Value, U as UF_6	US$/kg U	16.4	34.3
Estimated Conversion Cost for Reprocessed Uranium	US$/kg U	4-5	4-5
Estimated Value of Plutonium	US$/g $Pu_{(f)}$	8-10	8-10
Value of Recovered U	US$/t U	11,400-12,400	29,300-30,300
Value of Recovered Pu	US$/t U	44,300-55,400	54,800-64,700
Total Value of Fuel	US$/t U	55,700-67,800	80,100-105,000

The greater inherent value of PWR fuel could be recovered economically at either plant capacity.

The minimum plant capacity for commercial viability is dependent on the fissile value of the fuel to be processed (i.e. whether PWR or BWR fuel, which differ significantly in residual ^{235}U content), and on the mode of apportioning plant financial charges. Two plants of nominal capacity 1 t U/d have recently been constructed in the USA; the Nuclear Fuels Services Reprocessing Plant, West Valley, operational in 1966 at a total capitalisation (1966 price) of US\$32.5 x 10^6 inclusive of waste storage facilities (Abrams 1969, Runion and Lewis 1969, Ferguson and Blanco 1968), and the General Electric Mid West Fuel Recovery Plant (Smith and Bigge 1970, Ferguson and Blanco 1968) which was scheduled for active commissioning in 1972, and was reported (USAEC 1971) to cost US\$25 x 10^6. A third plant of nominal capacity 5 t U/d is under construction by Allied-Gulf Nuclear Services, Barnwell, S. Carolina, at a capital cost reported between US\$70 x 10^6 (ORNL 1970a), and US\$80 x 10^6 (USAEC 1971). A combination of aqueous and volatility processes is employed in the General Electric Mid West Fuel Recovery Plant, whereas both the Nuclear Fuels Services and Allied-Gulf Nuclear Services plants are based on aqueous (solvent extraction) technology, as are major reprocessing facilities in the UK and France.

The General Electric plant apparently shows a saving in capital cost over that for the NFS facility of equal throughput, and additionally effects conversion of uranium as purified uranium hexafluoride.

However, it is not clear whether the ground rules for cost comparison are identical, and the economics of the combination 'Aquafluor' process require validation by sustained operation at full capacity.

It has been concluded that the NFS plant operation will not yield a normal chemical-industry return on equity until such time as the potential load increases to permit expansion of plant capacity to a level approaching 500 t U/a at current unit reprocessing costs (Abrams 1969). Planned expansion of the NFS facility is to a capacity of 3 t U/d at an additional capital investment of US\$12 x 10^6 (1971 prices) (USAEC 1971). The annual operating year of the NFS plant has been quoted as 285 days, equivalent to a fuel capacity of 855 t U/a.

3. BACKGROUND TO THE REPROCESSING STUDY

3.1 Terms of Reference

The terms of reference were:

(1) To examine critically the design and cost structures of existing types of chemical reprocessing plant, and to identify the aspects

of the plant in which significant cost reductions might be achieved by different design approaches.

(2) To carry out conceptual design and layout studies applying new design principles to existing solvent extraction processes, in sufficient detail to enable preliminary cost estimates to be made of promising new design concepts, with the primary aim of achieving capital cost reduction through design.

(3) To keep a watching brief on overseas development of processes and equipment, and to suggest and incorporate into the design studies new equipment which appears to offer a reduction in capital cost.

(4) To identify and define areas requiring experimental study which are worthwhile in terms of potential cost savings in a large plant and which will enable the Commission to maintain expertise in fuel reprocessing by having staff working in this field.

The minimum objectives of the study were to produce a reference plant design incorporating new design concepts in which the major areas of cost were identified.

3.2 Specification of Requirements for the Reference Plant Design

The plant was to receive irradiated fuel from thermal oxide reactors, to process fuel after a cooling period of 180 days, and to produce decontaminated uranium and plutonium as nitrate solutions. Liquid effluents of high and medium activity would be stored as liquid concentrates; waste of low activity content would be discharged after treatment to reduce concentrations of active isotopes below levels recommended by the International Commission on Radiological Protection.

The plant was to conform to present basic requirements in effluent treatment; the design study would additionally state estimates of costs, where available, for more sophisticated processes of krypton-85 and tritium fixation, and recovery of neptunium-237.

A plant capacity of 3 t U/d (900 t U/a) has been adopted in the study on the following grounds:

(a) Present reprocessing plants are considered likely to be uneconomic below a capacity of 300 t U/a, and the world trend is towards capacities of 1500 t U/a. Present optimum capacity appears in the range 3-5 t U/d (900-1500 t U/a) (Marcus and Seynaeve 1971).

(b) The anticipated Australian Power Programme for thermal reactors (M. Silver, AAEC private communication 1972) could generate annual arisings of 300 t U/a by early to mid 1990's, rising to 500-700 t U/a by 2000. Subsequent increase in annual arisings of thermal oxide fuel will be dependent on the rate of installation and self sufficiency of Fast Reactors; a speculative maximum total Australian arising of 900 t U/a of thermal fuel does not seem unreasonable.

(c) The minimum economic capacity (300 t U/a) could not be justified on present estimates for Australian arisings alone before the 1990's. However, the capacity requirement could be radically altered if Australia were to act as a toll processor for fuel from other countries. For example, the anticipated combined arisings from Japan and Australia could require a capacity of about 1800 t U/a by 1988. The installed capacity and timing would require detailed optimisation to achieve economy by a high utilisation factor. The utilisation could probably be obtained by construction of a plant with capability for extension of capacity to match the increasing reprocessing load. On this basis a 3 t U/d plant installed in 1979 would require doubling in capacity in 1985, and reach full capacity in 1988.

3.2.1 Alternatives in selection of technology and equipment

Figure 1 shows a sequence of the more important alternatives on which decisions are required for specification of the main line of a Thermal Oxide Reprocessing Plant. A review by Costello (1972b), of the alternatives available in choice of technology and equipment showed that in many cases the decisions were interactive, and that later options were governed by, or arose out of, earlier decisions. The present study has been based on the route shown in heavy outline in Figure 1.

It was concluded that a design *ab initio* of a reprocessing plant was neither necessary nor desirable for the aims of the present study, which has concentrated on aspects of design and layout philosophy. A greater effort on generation of new concepts was possible by adaptation of process flowsheets and equipment size and shape utilised in existing operational plants, confining design effort to those items affected by throughput, and by injection of alternative equipment options (e.g. continuous leaching equipment and centrifugal solvent extraction contactors) into the design variants.

3.2.2 Selection of plant technology

Detailed information on plant flowsheets and equipment requirements for reprocessing thermal reactor fuels is available in docket form from the following sources:

(a) The Nuclear Fuels Services Plant, West Valley, N.Y., operational since 1966. It has a nominal capacity of 300 t U/a capable of extension to 900 t U/a (USAEC 1971) and employs a shear-leach head end process followed by separation of uranium-plutonium and fission products by solvent extraction. (Nuclear Fuels Services, Inc., USAEC Docket Series 50201.)

(b) The Allied-Gulf Nuclear Services Plant, Barnwell, S. Carolina, (in course of construction). It is designed for a nominal capacity of 1500 t U/a, and employs a shear-leach head end followed by solvent extraction. (Allied-Gulf Nuclear Services, USAEC Docket Series 50332.)

(c) The General Electric Company Plant, Morris, Illinois, which is at the commissioning stage. It is designed for a nominal capacity of 300 t U/a, and employs a shear-leach process with one cycle of solvent extraction, followed by fluoride volatility processing for uranium purification and ion exchange treatment for plutonium purification. (General Electric Company, USAEC Docket Series 50268.)

The Nuclear Fuels Services (NFS) plant was selected as a basis for equipment sizing and flowsheet consideration, as it relied on proven technology, its design, cost breakdown and operational history were well documented, and because its design could be adapted to the capacity requirements of the study with marginal resizing of equipment items.

This approach eliminated the need to develop a plant design from first principles, and permitted concentration of the study effort on alternatives in plant equipment and layout in high cost areas.

3.3 Design and Cost Breakdown of Solvent Extraction Reprocessing Plants

A critical review of the design and cost structure of existing types of chemical reprocessing plants by Costello (1972a) concluded that the Reprocessing Plant, which separates and purifies uranium and plutonium from fission products, accounted for 45-66% of the capital investment for materials and installation in existing reprocessing complexes. The remainder of the cost was incurred in receipt and storage of fuel, in waste disposal, and in utilities, laboratories, health physics monitoring services and administration.

Seventy to eighty per cent of the Reprocessing Plant cost was incurred in the Process Building, Process Equipment and Pipework. The Building contributed 30-40% of the Reprocessing Plant cost.

Mechanical pretreatment, leaching, and solvent extraction accounted for about three quarters of the Reprocessing Plant materials and installation cost.

These process functions offered the greatest potential saving in capital expenditure.

Reduction in capital cost involved optimisation of mechanical handling operations, and the use of high capacity compact process equipment to reduce physical size and complexity.

The most promising areas for investigation were:

(a) Examination of maintenance philosophies, modified remote maintenance with modular equipment installation for highly active equipment, with the option of direct maintenance for items handling lower activity.

(b) Optimisation of mechanical operations in fuel breakdown and leaching, and investigation of overall cost savings in continuous leaching of fuel.

(c) Reduction of shielded volume by elimination of general purpose handling equipment in cells in favour of single purpose simplified equipment.

(d) Possible reduction of the number of solvent extraction cycles by improvement of chemical technology.

(e) Investigation of alternative engineering designs of extraction equipment for economy in shielded volume.

(f) Selection of the optimum economic means for criticality control.

(g) Examination of the potential savings from modification of existing shielding philosophy.

(h) Reduction of costs incurred outside the Separation Plant with particular reference to Fuel Receipt and Storage, and Treatment of Radioactive Effluents.

3.4 Definition of Concepts for Assessment

Reprocessing design studies have been carried out by Cairns et al. (1967), Unger et al. (1967), Schneider et al. (1961) and Lloyd (1965) with the aim of reducing capital costs to attempt economic justification of small capacity reprocessing facilities. In general, these studies have each pursued a major theme considered at the outset to have significant cost saving potential (e.g. installation of equipment in modular racks, reshaping of equipment for close packing in a single shielded cell, or maintenance and operation of plant from a shielded mobile control station) and each study has led to one plant design embodying the theme.

The success or failure of this approach depends on the effectiveness of the theme selected. Its major disadvantage is that several schemes cannot be compared on the same ground-rules owing to differing capacity, flowsheet, and equipment type.

This study has assumed that a greater probability of success in developing schemes for cost reduction would be derived by a broad based assessment, covering a range of plant concepts. The extremes in concept range from a fairly conventional plant embodying new approaches in operating and maintenance philosophy, to a completely unconventional design with shielding requirements tailored to the operators instead of to the plant equipment. These design alternatives have been based on a standard flowsheet and plant capacity to facilitate ultimate comparison.

Four conceptual designs have been assessed as follows:

Design 1 : A fully shielded plant using adequate shielding concrete for routine operations, but adopting a new approach to maintenance procedures by constructing temporary shielding as and when equipment failure occurs. This design is based on whole assembly shearing, batch leaching, and solvent extraction in pulsed columns.

Design 2 : A variant of (1) utilising centrifugal contactors as solvent extraction equipment.

Design 3 : A design where the majority of highly active equipment is submerged in a water filled pond. This design involves prebreakdown of fuel assemblies into elements, shearing in a small scale disposable shear feeding a semi-continuous rotary drum leacher, with pulsed column solvent extraction equipment.

Design 4 : A scheme based on containment of plant underground, with access for operation and maintenance by a mobile shielded crane.

4. PROCESS TECHNOLOGY

4.1 Fuel Specification

The reprocessing plant is assumed to service a reactor programme of advanced LWR's consisting of 1000 MW(e) boiling and pressurised water reactors.

Fuel assemblies for these reactors are square in cross section, and of approximate overall dimensions 5 m (length) and up to 210 mm (across flats). The fuel consists of UO_2 pellets clad in Zircaloy, at an initial enrichment of up to 3.3% ^{235}U. After an exposure of 33,000 MWd/t U at an average power level of 34.8 MW/t U, the discharged fuel will contain up to 0.84% ^{235}U and 9.3 kg Pu (total)/t U at a fissile content ($^{239 + 241 + 243}Pu$ in total Pu) of 69.4%. Fuel will be cooled for a minimum of 180 days prior to reprocessing.

Physical characteristics of reference fuel assemblies (Jackson and Moreland and Stoller Associates (1968)) are given in Table 3. Approximate compositions

TABLE 3

DATA FOR REFERENCE REACTOR FUEL DESIGN

Mechanical Design Parameters	Unit	Reactor Type	
		BWR	PWR
Fuel Element			
Fuel Material		Sintered UO_2 Pellets	Sintered UO_2 Pellets
Fuel Pellet Diameter	mm	12.4	9.4
Fuel Pellet Length	mm	17.8	15.3
Fuel Pellet Density	$g\ cm^{-3}$	10.22	10.19-10.3
Cladding Material		Zircaloy-2	Zircaloy-4
Cladding Outside Diameter	mm	14.3	10.7
Cladding Thickness	mm	0.81	0.62
Active Fuel Length	mm	3,660	3,660
Length of Gas Plenum	mm	410	100
Fuel Assembly			
No. of Fuel Elements per Assembly		49	204
Weight of UO_2 per Assembly	kg	221.0	517.0
Fuel Element Array		7 x 7	15 x 15
Fuel Element Pitch	mm	18.75	14.3
Overall Cross Sectional Dimensions including Shroud	mm	138 x 138	214 x 214
Overall Assembly Length	m	4.9	4.9
Fuel Shroud Material		Zircaloy-2	None
No. of Spacer Grids per Assembly		7	9
Spacer Grid Material		Zircaloy-4	Inconel-718
Fixed Burnable Poisons		Gadolinium Oxide	None

of actinides and fission products contained in the irradiated fuel are detailed in Table 4.

TABLE 4

NUCLEAR DESIGN DATA (EQUILIBRIUM CORE AVERAGES)

Nuclear Design Parameters	Unit	Reactor Type	
		BWR	PWR
Fuel Rating	MW/t U	22.0	34.8
Fuel Irradiation	MWd/t U	27,500	33,000
Cooling Time -			
(a) on receipt	days	90.0	90.0
(b) before processing	days	180.0	180.0
Initial Enrichment	% ^{235}U	2.56	3.3
Discharge Parameters			
^{235}U	%	0.62	0.84
^{236}U	%	0.33	0.42
^{238}U	%	95.3	94.5
^{239}Pu	kg/t U	4.61	5.30
^{240}Pu	kg/t U	2.07	2.40
^{241}Pu	kg/t U	0.93	1.17
^{242}Pu	kg/t U	0.36	0.44
$Pu_{(f)}$/total Pu	%	69.5	69.4
Fission Products (180 days cooled)	Ci/t U		$\sim 4 \times 10^6$
Decay Heat (180 days cooled)	kW/t U		18.0

4.2 Chemical Flowsheet

The chemical flowsheet for the process has been based on the Nuclear Fuels Services Flowsheet for low enriched uranium dioxide fuel clad in Zircaloy (Nuclear Fuels Services, Inc., USAEC Docket 50201-4). The specific throughput of plutonium has been increased in accordance with the greater irradiation and rating parameters of the reference fuel specification.

A block diagram of process operations is shown in Figure 2. Irradiated fuel elements are received at the reprocessing centre after a period of 90 days' cooling at the reactor sites. The fuel is stored under water for a total cooling time of 180 days to decay the bulk of radioisotopes of iodine. The reprocessing operations consist of shearing through the fuel elements to expose the fuel pellets to nitric acid leachant. A Purex solvent extraction flowsheet is used to decontaminate uranium and plutonium from fission products, and to separate uranium and plutonium.

The uranium stream is further purified by two cycles of solvent extraction, concentrated by evaporation, and receives a final decontamination by adsorption of fission products on silica gel. Uranium is exported as uranyl nitrate solution.

Plutonium is purified by one cycle of solvent extraction followed by adsorption-desorption purification on ion exchange resin, evaporation, and exported as plutonium nitrate solution.

Fission product liquid wastes are concentrated to small volume and stored as neutralised solutions pending ultimate conversion to solid form. Nitric acid is recovered from waste treatment for recycle. Solid wastes are buried in sealed containers; gaseous wastes are scrubbed with sodium hydroxide, filtered and discharged to atmosphere.

Liquid wastes of large volume are decontaminated by evaporation and ion exchange to below acceptable concentrations of active isotopes recommended by the International Commission on Radiological Protection, and discharged to the environment.

Detailed chemical flowsheets for plant functions are given in Figures 3 - 8.

4.3 Fuel Shearing and Leaching (Figure 3)

An irradiated fuel assembly consists of a parallel array of fuel element tubes, each tube being fabricated from Zircaloy or stainless steel, sealed at both ends and containing pellets of oxides of uranium and transuranics (e.g. plutonium, neptunium) and fission products. The tube plenum contains helium introduced during manufacture; after irradiation the plenum gas contains activity due to krypton-85, tritium, and volatile fission products such as iodine-129 and iodine-131.

A shear blade is used to cut transversely through the containment tubes and fuel to expose the latter to boiling leachant (nitric acid). Active plenum gases released during shearing are ducted into the leacher off-gas treatment system.

The stoichiometry of oxide fuel leaching in nitric acid is dependent on acid concentration and temperature. Approximate relations appropriate to leaching conditions are as follows:

$$UO_2 + 3\ HNO_3 \rightarrow UO_2(NO_3)_2 + \frac{1}{2}\ NO + \frac{1}{2}\ NO_2 + \frac{3}{2}\ H_2O$$

$$UO_2 + \frac{8}{3}\ HNO_3 \rightarrow UO_2(NO_3)_2 + \frac{4}{3}\ H_2O + \frac{2}{3}\ NO \quad .$$

Uranium is oxidised to the hexavalent state and the plutonium to a mixture of tetra- and hexavalent ions.

Off-gases from the leacher consist mainly of water vapour, nitric oxide, nitrogen peroxide, together with activity from krypton-85, iodine-129, iodine-131, entrained droplets of leachant containing dissolved fission products, and tritiated water. Oxygen is fed into the off-gas stream, which is cooled to effect recovery of part of the nitrogen oxides according to the relations:

$$2NO + \frac{1}{2}\ O_2 \rightarrow NO_2$$

$$3NO_2 + H_2O \rightarrow 2\ HNO_3 + NO \quad .$$

Gases are scrubbed countercurrently with a solution of dilute nitric acid, and the reclaimed nitric acid is returned to the leaching operation.

After leaching is complete, the quantity of uranium and plutonium contained in solution is determined, and the acidity of the solution adjusted prior to solvent extraction.

The leached cladding solid waste arising is monitored for residual fuel content and discharged to burial in sealed containers.

4.4 Solvent Extraction

The solution of uranium, transuranic elements (notably plutonium, neptunium, and americium), and fission product nitrates resulting from the fuel leaching operation, is treated by a solvent extraction process to yield purified product solutions of uranium and plutonium nitrates, and a waste solution of fission product nitrates, which are concentrated and stored, and which may be used as a source for recovery of neptunium.

The aqueous nitrate solutions are contacted countercurrently in solvent extraction equipment with an organic solvent consisting of 30% tributylphosphate (TBP) in odourless kerosene diluent (OK). The functions of the solvent extraction process are:

(a) separation of uranium and plutonium from fission products,

(b) separation of uranium from plutonium,

(c) purification of the separated uranium and plutonium.

Separation of uranium and plutonium from fission products is achieved by selective extraction of uranium and plutonium into solvent.

In a solution of irradiated fuel in nitric acid, the actinide elements are principally present as tetravalent or hexavalent ions, which form strong metal nitrate-TBP complexes of the form $M(NO_3)_4.2TBP$ and $MO_2(NO_3)_2.2TBP$ respectively. These complexes are readily solvated and are preferentially extracted into the solvent phase. Most of the fission products form ions of a lower charge which give rise to weak metal nitrate-TBP complexes, and are less readily extracted into solvent. The relative extractability of the TBP complexes is exploited in the design of the solvent extraction contactor to effect the desired degree of separation of the actinides from fission products.

Separation of uranium from plutonium in the solvent extract is achieved by selective control of valency state.

Plutonium is reduced from the tetravalent or hexavalent state to the trivalent state by ferrous ion at a redox potential where uranium remains hexavalent. The partition coefficient of plutonium is reduced in the trivalent state; plutonium tends to transfer from the organic phase into the aqueous phase, while uranium remains in the organic phase.

The solvent extraction process consists of four cycles:

(a) a partition cycle where uranium and plutonium are extracted from the bulk of fission products and separated into two partially decontaminated product streams;

(b) two consecutive cycles of uranium purification from residual contamination by plutonium and fission products;

(c) one cycle of plutonium purification from uranium and fission products.

Gross contamination of activity between the partition cycle and the two uranium cycles is reduced by segregation of solvent into three separate solvent cycles, each with its own equipment for solvent treatment.

4.4.1 Partition cycle (Figure 4)

The solution of leached fuel is accounted for uranium and plutonium content, conditioned to 2M in nitric acid, and is fed countercurrent to a stream of 30% TBP/OK solvent. More than 99.5% of the uranium and plutonium are extracted into the solvent phase. More than 99% of fission products originally present in the irradiated fuel remain in the aqueous phase, designated the high level waste stream (HAW), which is discharged to waste evaporation and storage.

Solvent phase containing the uranium and plutonium is scrubbed countercurrently with a nitric acid stream which re-extracts into aqueous phase a proportion of fission products extracted into solvent (notably zirconium, niobium, and ruthenium) and additionally tends to wash back droplets of entrained aqueous phase containing fission product nitrates.

The fission product content of the extract solvent solution of uranium and plutonium is mainly due to residual zirconium, niobium, and ruthenium, which form labile complexes with TBP. Extraction of zirconium and niobium is reduced by operation at low acidity at which these ions tend to hydrolyse; however, low acid conditions favour extraction of ruthenium ion. A compromise is achieved in practice by control of acidity in the extraction and scrubbing operations.

Solvent extract is combined with plutonium bearing solvent raffinates from plutonium scrubbing, partition cycle and plutonium stripping, plutonium purification cycle (IIBW). The combined solvent stream is fed countercurrent to a reducing aqueous strip stream containing ferrous sulphamate. Plutonium is reduced to the trivalent state in which its distribution ratio favours the aqueous phase. The nitric acid concentration in the partition operation is maintained sufficient to preserve the bulk of the uranium in the solvent phase (HBU). The plutonium emerges in an aqueous stream, which is fed to plutonium scrubbing, where separation of uranium from plutonium is completed.

The aqueous product from plutonium partition is contacted countercurrently with fresh solvent to remove uranium, which is transferred back to the partition stage. The aqueous plutonium product stream (HBP) is transferred to the plutonium purification cycle.

Solvent product containing uranium from partition (HBU) is pre-heated and contacted countercurrently with an inactive heated solution of weak nitric acid. The low concentration of nitric acid allows transfer of uranium from the solvent into the aqueous phase. The product aqueous solution of uranium (HCU) is transferred to the first uranium purification cycle.

Solvent raffinate from the uranium stripping operation (HCW) is fed to the first cycle of solvent washing equipment, for ultimate recycle to the partition cycle.

4.4.2 <u>First uranium purification cycle</u> (Figure 5)

The first uranium purification cycle of extraction, scrubbing, and backwash effects further decontamination of uranium from residual plutonium and fission products.

The aqueous uranium stream from the partition cycle is acidified with nitric acid to favour re-extraction of uranium into the solvent phase. Uranium is re-extracted from the acidified aqueous phase into a countercurrent stream of solvent, which is scrubbed countercurrently by nitric acid to remove entrained aqueous phase, fission products, and plutonium which were co-extracted with the uranium. The solvent is scrubbed by nitric acid at two levels of acidity to achieve optimum conditions for decontamination from both zirconium and niobium, and ruthenium. The aqueous raffinate from extraction contains about 95% of the fission products and trace plutonium entering the cycle, and is routed to low active waste evaporation and storage.

The solvent extract (IAP) containing >99.5% of uranium fed to the cycle, is heated and contacted countercurrently with a heated stream of low acidity nitric acid. Uranium is stripped into aqueous phase which is transferred to the second uranium purification cycle.

Solvent raffinate (ICS) from the uranium stripping operation is washed with sodium carbonate and nitric acid, and recycled to the first uranium purification cycle.

4.4.3 Second uranium purification cycle (Figure 5)

The second uranium purification cycle is analogous to the first; it consists of extraction, scrubbing, and stripping operations.

The aqueous uranium product stream from the first uranium cycle (ICU) is acidified with nitric acid. Uranium is extracted with a countercurrent stream of solvent, which is scrubbed by nitric acid. More than 99.5% of the uranium fed to the cycle emerges in the solvent extract stream (IDP). The aqueous raffinate stream (IDW) contains about 90% of the fission products entering the second cycle and is routed to low active waste evaporation and storage.

Uranium is stripped from the solvent extract stream (IDP) into a countercurrent stream of weak nitric acid (IEU), and is routed to uranium evaporation.

The solvent raffinate from uranium stripping (IEW) is fed to the third cycle of solvent washing equipment.

4.4.4 Plutonium purification cycle (Figure 6)

The aqueous plutonium product (HBP) from the partition cycle is blended with raffinate from the plutonium ion exchange stage (XAW), conditioned with sodium nitrite solution to reoxidise plutonium to the tetravalent state, and acidified to favour conditions for re-extraction of plutonium into solvent.

A single cycle of forward extraction, scrubbing and stripping is employed. Plutonium is extracted from the conditioned feed solution into a countercurrent stream of solvent. The plutonium bearing solvent stream is scrubbed with a countercurrent stream of nitric acid, to remove contaminant uranium and fission products. The aqueous raffinate stream from extraction (IIAP) contains about 95% of the fission products entering the cycle, and is routed to low active effluent evaporation.

The solvent extract (IIAP) contains more than 99.5% of the plutonium entering the cycle. Plutonium is stripped from solvent into a countercurrent aqueous stream containing ferrous sulphamate, and is routed to plutonium ion exchange.

Solvent raffinate from plutonium stripping is recycled to the partition cycle.

4.4.5 Solvent treatment (Figure 7)

Both the TBP solvent and the kerosene diluent undergo radiolytic degradation in the reprocessing operation. TBP is decomposed progressively to dibutyl phosphate, monobutyl phosphate, and phosphoric acid. Kerosene is degraded to aliphatic hydrocarbons, which are converted to nitro-compounds in the presence of nitric acid.

The by-products formed by radiolysis may complex both with certain fission products and with plutonium; the overall effect of solvent degradation is a progressive reduction in decontamination efficiency and an increase in losses of fissile material in aqueous raffinates.

The solvent is treated to remove degradation products prior to recycle to the process. Solvent is washed with sodium carbonate solution to remove traces of uranium, plutonium, and fission products in addition to the degradation products of TBP and kerosene. A sodium carbonate wash is preferable to sodium hydroxide in that it avoids precipitation of uranium hydroxide. The nitric acid wash neutralises residual carbonate and assists disengagement of the solvent phase for recycle from the aqueous wash raffinate.

The partition cycle and the first and second uranium purification cycles employ segregated streams of solvent; each cycle possesses its own solvent washing/equipment. The three solvent wash cycles operate on similar flowsheets in identical equipment; they differ essentially in the level of activity processed.

Solvent raffinate from the partition or uranium purification cycles is fed into the appropriate wash cycle. The solvent is twice contacted countercurrently with sodium carbonate solution, the operations differing in the residence time of solvent being treated. Solvent is finally washed countercurrently with nitric acid before recycle to the process.

Aqueous carbonate raffinate from the second wash contactor is fed into the first contactor; aqueous raffinate from the first and third contactors are combined and ducted to low active waste evaporation, together with aqueous raffinate from the other two solvent wash cycles.

4.5 Product Treatment

4.5.1 Uranium product evaporation and silica gel purification (Figure 5)

The aqueous uranium product stream from the second uranium purification cycle IEU is concentrated to 300 g U/l. The weak acid overheads from evaporation are recycled as strip feed in the uranium purification cycles. The concentrated solution is finally treated with silica gel, which removes residual fission products Zr and Nb. Fission products are absorbed from the silica gel by a solution of oxalic acid, and routed to waste evaporation and disposal.

4.5.2 Plutonium ion exchange purification and concentration (Figure 6)

The aqueous plutonium product stream from solvent extraction (IIBP) is conditioned to 7.2 M acidity and contacted with anion ion exchange resin (Deacidite FF) at 60°C. Plutonium is absorbed on the resin to a concentration of 70 g Pu/l of bed. Fission products and uranium remain in solution and are recycled to the plutonium purification (solvent extraction) cycle to recover trace plutonium. Plutonium is elutriated from the resin with 0.6 M nitric acid, and concentrated to 300 g Pu/l.

4.6 Waste Treatment and Disposal

Gaseous, liquid, and solid wastes containing activity arise from the reprocessing operation.

4.6.1 Gaseous wastes (Figure 3)

The sources of gaseous activity are

(a) off-gases from the leaching operation

(b) extract air from the vessel ventilation system

(c) extract air from cells.

Treatment of leaching off-gases was described in Section 4.3. The active constituents are krypton-85, residual iodine-129 and iodine-131, entrained droplets of fission product bearing solution, and some tritiated water vapour. Of these, krypton-85 and tritium oxide are discharged to the environment.

Off-gases from the leacher condensers are scrubbed countercurrently with sodium hydroxide solution to remove elemental iodine, and most of the activity in entrained droplets. The residual gases are combined with the vessel ventilation extract air, and the combined stream scrubbed countercurrently with sodium hydroxide solution, preheated above the dew point, and filtered prior to discharge to atmosphere.

Extract air from cells is filtered and discharged to atmosphere.

4.6.2 Liquid wastes (Figure 8)

Liquid wastes arising from the process are divided into three categories according to the level of contained fission product activity: High Level, Low Level, and Trace Active Wastes, containing respectively activity concentrations of more than 500 Ci/m^3, about 1-10 Ci/m^3, and less than 10 mCi/m^3.

The fission product and actinide contents of liquid high and low level wastes are initially stored as solutions pending conversion to an insoluble solid form.

Liquid wastes are evaporated prior to neutralisation and storage; evaporation reduces the requirement for waste storage capacity, effects recovery of part of the nitric acid content of the wastes, reduces the requirement for sodium hydroxide neutralising reagent, and hence the salt content of the stored waste, which in turn increases the ultimate concentration factor achievable before solidification.

High Level Waste contains >99% of the fission products originally present in the irradiated fuel. This waste arises in aqueous raffinate from the partition cycle. It is evaporated to low bulk, neutralised with sodium hydroxide, and stored as liquid waste pending ultimate disposal by solidification. An arbitrary limitation of nitric acid concentration in the evaporator to $8\underline{M}\ HNO_3$ is imposed to limit corrosion due to acidity and temperature; this corresponds to a concentration factor of 20:1.

Low Level Waste contains <1% of the fission product content of the fuel. The main sources of this arising are aqueous raffinates from the first and second uranium purification cycles and the plutonium purification cycle, from the solvent washing cycles, overheads from the higher level waste evaporator, and from off-gas treatment operations. The waste is evaporated to low bulk, neutralised with sodium hydroxide and stored in a similar manner to high level waste. The limitation of concentrate acidity to $8\ \underline{M}\ HNO_3$ is equivalent to a concentration factor >60:1.

Trace Active Waste contains $<10^{-5}\%$ of the fission product originally present in the fuel. This waste arises in overheads in recovery of nitric acid by fractional distillation of evaporator condensates. This waste is further decontaminated by evaporation or ion exchange prior to discharge.

4.6.3 Solid wastes

Solid wastes arising in the reprocessing operation are:

(a) leached sections of fuel element cladding (hulls) discharged from the leacher,

(b) finely divided particles of cladding metal and elemental fission products (Ru, Rh, Mo, Tc, Pd) suspended in the solution of leached fuel,

(c) contaminated waste trash arising from maintenance operations, e.g. reject shear blades, contaminated protective clothing.

Fuel element cladding is washed from containment fuel solution. It contains activity through neutron activation of the parent metal and of impurities in the metal, together with traces of unleached fuel. It is buried below ground in concrete containers.

Finely divided suspended activity in the fuel solution will either be removed prior to solvent extraction (by filtration or centrifugation), or will enter solvent extraction and be removed either continuously in the high level raffinate, or batchwise at plant washout. In each case the solid waste will be directed into high level storage.

Contaminated waste arising from maintenance will be sealed and buried.

4.6.4 Nitric acid recovery (Figure 8)

Nitric acid is recovered by fractional distillation of the acid condensates from evaporation of the high and low level waste streams.

Condensate from high level waste evaporation is blended and evaporated with low level wastes. Condensate from evaporation of the low level wastes is evaporated and fractionally distilled at reduced pressure to produce products of 8 $\underline{M}$ and 12 $\underline{M}$ HNO_3 for recycle to the process. Overhead condensate from the fractionation contains <0.002 $\underline{M}$ HNO_3, and represents the trace active effluent. Dependent on its activity content, this stream may be further decontaminated by evaporation or ion exchange. This waste is ultimately neutralised and discharged to the environment when the level of active species is below the maximum permissible concentrations recommended by the ICRP.

4.6.5 Rework procedure

In-line monitors for uranium and plutonium have been incorporated on all raffinate streams where significant losses could occur. It is expected that these monitors will indicate the need for adjustment of plant controls in adequate time to maintain average losses at an acceptably low level. In the event of a waste stream containing more product than economically can be discarded, that stream may be directed to rework evaporation. The concentrated waste stream is returned to the partition cycle and the condensate from the evaporation is directed to low level waste evaporation.

5. DESIGN 1

5.1 Themes for Reduction of Capital Cost

The themes taken as a basis for cost reduction in Design 1 are:

(a) Simplification of process operations to reduce mechanical and chemical engineering costs.

(b) Compact design and layout of plant to reduce building and civil costs.

(c) A new approach to maintenance requirements for failed equipment, to reduce building costs.

The areas involving the largest proportion of costs in a reprocessing complex are:

(a) Mechanical pre-breakdown of fuel.

(b) Shear-leach.

(c) Solvent extraction.

5.1.1 Mechanical pre-breakdown of fuel

The Head End Treatment in Design 1 has incorporated the principles of whole assembly shearing into a basket contained in a batch leacher. These operations have been employed successfully in the BNFL Windscale reprocessing plant (Hughes & Kemp 1970) and have been incorporated in the Allied-Gulf reprocessing plant design (USAEC Docket 50332). This technique eliminates the complex and expensive equipment and shielded space employed in the NFS Plant (Nuclear Fuels Services, Inc., USAEC Docket, 50201-2) to disassemble fuel into elements prior to shearing, which involved removal of assembly sheath and end pieces by a remote saw, and breakdown of the lattice of elements into a geometry acceptable to the shear.

5.1.2 Shear-leach operations

The shielded volume of the shear-leach operational cell has been reduced to a minimum by two principles:

(1) Direct feeding of the assembly from a shielded flask into the shear.

(2) Substitution of general purpose handling equipment in the cell (cranes, power manipulators) by simple robust special purpose equipment designed for the routine movement functions necessary for a batch shear-leach operation.

The principles involved were derived from a critical examination of the elevation layout of the proposed Allied-Gulf shear-leach cell (Figure 9) which also contains some equipment for solvent extraction and highly active evaporation. The effective boundary enclosing the equipment for processing (i.e.

shearing and leaching fuel, and for disposal of leached cladding) is outlined. The processing area on an elevation basis is about 40% of the total shielded area in Figure 2; the additional shielded cell space is necessary for movement of traversing cranes and routine decontamination and maintenance of the active cranes and manipulators, for maintenance of the shear pack, and for receipt of the fuel assemblies. The latter are transferred from pond storage into the processing cell by a swinging arm, and then incrementally fed to a shear. The length of the cell is thus increased over the minimum required by a length approximately equal to that of a complete fuel assembly (about 5 m).

In the cell layouts for Design 1 (Figures 24, 25), fuel is transferred from the storage pond in a shielded flask, and is incrementally fed from the flask into the shear pack, to eliminate the need for an in-cell feed envelope and the concomitant penalty in shielded space. Sheared sections and end pieces are received into a basket located in the leacher; baskets containing leached cladding are lifted from the leacher and traversed over an arc to washing, monitoring and basket discharge stations by a rotating arm carrying a grab, mounted in a plug in a roof slab.

The shear pack is removed for maintenance by a similar special purpose lifting grab and rotating arm after being washed down in situ from contaminant fuel and cladding. The washed pack is transferred into a pond adjacent to the processing cell for underwater maintenance.

Routine handling operations within the cell are those of basket transfer from the leacher to leached hull monitoring, washing, and discharge stations, and of removal and replacement of failed shear packs. These transfer operations are effected by special purpose grabs suspended from rotating arms located in plugs in some roof slabs. The cell equipment is laid out so that successive transfer stations lie on the arc of traverse of the grabs. The lifting and rotating mechanisms are made sufficiently simple and robust to eliminate the maintenance required by gear drive and motors in the relatively complex general purpose handling tools (cranes and power manipulators), and thus eliminate the need for shielded decontamination and maintenance cells for the handling equipment.

The shear pack is designed to be washed from gross contamination in situ and transferred by special purpose grab into an underwater pond for remote replacement of failed parts. This approach eliminates the need for a shielded cell for remote dismantling and decontamination, prior to replacement of blades and bearings by contact maintenance.

Access to the cell for breakdown maintenance is via removal of the roof slabs, following a preliminary decontamination operation. It is assumed that equipment could be removed using an impact wrench attached to a device lowered into the cell by the external crane.

5.1.3 Solvent extraction and ancillary plant

Layouts for the active chemical engineering plant are shown in Figures 26-32 and for the cold chemical storage area in Figure 33.

To achieve a high degree of utilisation of shielded space the following unconventional principles have been followed:

- Equipment has been grouped in shielded cells on the criterion of shape, in contrast to the conventional approach of segregation of equipment areas on the basis of activity levels. Thus pulsed columns (and minor associated vessels) are contained in a single tall cell of small area, while large tankage associated with the columns (feed and waste tanks and conditioners) are contained in cells of large area but relatively low internal height.
- Internal divided walls between cells have where possible been eliminated. Thus there is no segregation of activity levels within the column cell (Figure 26).

The feed, conditioning and waste tankage associated with the leachers, solvent extraction columns and evaporators is essentially contained in one cell contiguous with the area housing equipment for solvent washing and off-gas treatment. Vessels containing a relatively high level of activity are grouped together and locally shielded within the cell to effect an overall reduction in external shielding requirements.

- Equipment has been close packed in cells on a minimum mutual spacing of about 0.6 m to permit access for contact maintenance. The majority of vessels appear superficially to have no problems of nuclear interaction which could restrict close spacing; it has been assumed that incorporation of fixed neutron absorbers in the design could eliminate any interactive problems appearing on detailed examination.

5.2 Equipment Description and Flow Diagrams

5.2.1 Fuel transport and receipt

A schematic layout of the pond receipt, storage, and export area is shown in Figure 35.

LWR fuel assemblies are transported in shielded flasks containing water as heat transfer medium. A transport flask is lifted from its transporter, and placed in a vertical position outside the receipt pond area. Excess pressure in the flask is vented to the ventilation system, the flask water content is drained and monitored for activity level. The flask is washed internally by hose connection, and the lid bolts loosened.

The flask is lifted and lowered into the low level fuel receipt section of the pond. The lid is removed, and the fuel assemblies lifted from the flask to storage racks in the main body of the pond. A minimum water submergence of at least 3.3 m will be maintained as radiation shielding for the fuel contained in the assemblies.

5.2.2 Shear-leaching operations

Figure 10 shows the equipment flow diagram for shear-leaching; details of shear-leach cell layout are shown in Figures 24 and 25.

Fuel transfer

An irradiated fuel assembly is transferred from an underwater storage rack, and loaded under water into the fuel transit flask. The flask is sealed, removed from the pond, decontaminated and transported to the processing plant. The flask is drained of water, and is positioned on the face of the cell. Interlocked doors on the flask and cell face are raised, and an incremental ram unit positioned at the rear door of the flask, which is then raised to permit incremental discharge of the fuel assembly from the flask into a guide tube leading to the assembly shear.

Assembly shearing

The whole assembly shear concept is similar to that adopted by B.N.F.L. and by the Allied-Gulf Reprocessing Plant. It is a robust unit containing profiled gags to grip the assembly transversely during the shearing stroke. The assemblies are sheared transversely by a moving and a fixed blade. Gags and the moving blade are actuated by mechanical linkages driven by hydraulic units. All hydraulics are external to the cell, the rams passing through seals to ensure containment of activity. Following B.N.F.L. design (Hughes et al. 1971), the shear operation is controlled by static switching units, interlocked with the incremental feed ram to shear the assembly into pre-determined lengths (50 mm). Sheared debris including end pieces fall from the shear pack into a radial feed chute which delivers the debris into a basket contained in one of two leachers.

The shear pack comprising blades, gags, and bearings is designed to be water washed in situ, and lifted by special purpose rotating grab from the pack frame into a pond for maintenance under water. While this operation is being carried out, a spare pack can be installed to resume continuity of processing.

Leacher design and operation

Two batch leachers (3D-2) have been incorporated in Design 1 to achieve a throughput of 3 t U/d. Each leacher has a liquor capacity of 9.3 m^3, and is fitted with a removeable basket with perforated walls and base, capable of containing 1 t of uranium as oxide fuel with associated cladding and end pieces in submergence under leachant acid. An ullage space of about 4 m^3 has been allowed for disengagement of froth. A cycle time between 14 and 18 hours is envisaged (Appendix B2).

To reduce problems of blockage in leacher baskets through interlocking of sections of assembly sheath, the approach of Allied Chemical Corporation has been followed; the baskets are about 0.76 m i.d., and criticality control of the leacher is obtained by incorporating a soluble neutron poison (cadmium nitrate) in the acid leachant feed.

Nitrous fume and steam discharged from a leacher at reflux flow to a down draft condenser, which reclaims and returns some nitric acid to the leacher and discharges incondensibles to off-gas treatment.

When leaching of uranium dioxide fuel in strong nitric acid is complete, the leacher is cooled and the product solution transferred by force ejector to the accountability and feed adjustment tank (3D-1). Several leacher batches are accumulated, homogenised, sampled and the total volume of solution deduced from hydrostatic pressure measurement to determine the quantity of uranium and plutonium fed to the solvent extraction plant.

Disposal of solid wastes

The basket containing leached cladding hulls and end sections is washed in the leacher in the successive batch of poisoned nitric acid. The basket is raised and drained into the leacher, and transferred by a rotating grab to a leached hull monitoring station (Figure 25), where residual uranium and plutonium are determined either by a γ-scanning device similar to B.N.F.L. (Hughes et al., 1971) or by a neutron irradiation technique similar to proposals by Allied-Gulf Nuclear Services.

Baskets containing appreciable fissile material are returned to the leachers; those in which the losses are acceptable are transferred by rotating grab to the basket tipping device (Figure 25), where the basket is inclined and the leached cladding contents discharged into a chute leading to the solid waste disposal system. The cladding is collected in a concrete flask sized to contain cladding from about 10 t U processed; the lid of the flask is placed in position and the flask and contents transported for burial in the solid waste storage facility.

Fines of cladding material generated in the shearing operation, and escaping through the basket perforations into the product solution, are allowed to settle and accumulate on the leacher floor.

The fines are periodically removed by a water jet positioned relative to the dished floor of the leacher to create a vortex swirl, which washes the fines from the floor to a central area below the suction pipe of a lift and force ejector. The fines and wash water are discharged into the highly active waste accountability and neutraliser tank, for ultimate storage and disposal with the highly active fission products.

Off-gas treatment (leachers and process vessels) (Figure 11)

Off-gases from the leacher condensers (3E-1, 3E-2) are scrubbed countercurrently with a recirculating aqueous solution in the leacher off-gas absorber (item 32001), which is a tower packed with stainless steel rings for absorption of nitrous fumes and entrained fission products (notably radio-iodine). The exit gases are combined with the extract from the vessel ventilation duct and scrubbed with sodium hydroxide solution in the iodine absorber (item 35002).

The aqueous solution is refluxed in a steam heated, air purged iodine stripper (item 35003) to remove volatile iodine. The residual off-gas stream is heated above the dew point to vaporise entrained liquid droplets and passed through a tank of absolute filters prior to discharge to atmosphere through a stack 65 m high.

Waste active sodium hydroxide solution is discharged to low active waste evaporation.

5.2.3 Solvent extraction

Continuous countercurrent solvent extraction in Design 1 is carried out in 10 pulsed columns, arranged in 4 cycles. Equipment diagrams for the solvent extraction cycles are shown in Figures 12-16.

The partition cycle (Figures 12, 13) contains 4 columns. Product from the shear-leach operation is fed into the extraction column (4C-1) where uranium and plutonium are extracted into solvent phase, and the bulk of fission products discharged as aqueous raffinate. In the partition column (4C-2), plutonium is separated from the solvent stream into aqueous phase. The plutonium product is purified from gross contaminant uranium in the plutonium scrub column (4C-3). The strip column (4C-4) transfers uranium from solvent into aqueous phase.

The aqueous solution of impure uranium is purified in two successive cycles of forward extraction, scrubbing and backwash. The first uranium cycle (Figure 14) contains an extraction and scrubbing column (4C-9) and a stripping column (4C-10); the second uranium cycle contains an extraction and scrubbing column (4C-11) and a stripping column (4C-12). (Figure 15).

The aqueous solution of impure plutonium is purified in a single plutonium purification cycle (Figure 16) containing an extraction and scrubbing column (4C-7) and a stripping column (4C-8).

Pulsed columns

A typical design of a pulsed column is shown in Figure 80. It consists of a stainless steel shell housing cartridges of perforated plates held on a central tie rod. The plates have a typical free area of 23% and are spaced about 50 mm apart. Two disengaging sections for phase separation are located above and below the perforated plate assembly. Typical values of HETS (Height Equivalent to a Theoretical Stage) quoted for pulsed columns in the Eurochemic plant (De Witte 1966, Joseph et al. 1971) are 0.7 m (extraction and strip columns) 0.7-0.9 m (scrubbing columns) and 1.1 m (U-Pu separation column, [corresponding to (4C-2) in function]).

Each column is pulsed by application of a varying air pressure to a line containing solvent, connected to the base of the column. The pulsed columns are controlled by maintaining the interface of solvent and aqueous phase at a preset position by controlling the rate of withdrawal of effluent from the base of the columns. The extraction columns are operated with organic solvent as the continuous phase; all other columns are operated with aqueous as the continuous phase.

The interface for the extraction column is in the bottom disengaging section (to reduce carryover of activity with the solvent product) and is located by sensing pots, filled with organic, connected to the interface control stubs of the column. These sensing pots are located near the top of the column to balance the hydrostatic head in the column. The position of the interface in the bottom disengagement section is known from the relative level of solvent in the sensing pots, monitored by pneumercator probes.

For the aqueous continuous columns the interface is located in the upper disengagement section by pneumercator pipes.

The aqueous stream from both aqueous and solvent continuous column is withdrawn from the bottom disengaging section to a level control pot located near the top of the column. The position of the interface in the columns is stabilised by a varying air pressure applied to the control vessel, the control air pressure being determined by the interface sensing equipment. The aqueous phase passes through a decanter to remove traces of entrained solvent phase.

The relative position of the solvent extraction columns and associated vessels carrying βγ-active solution have in the main been selected to allow transfer of active liquor by gravity flow, and eliminate the capital and maintenance costs of active pumps and shielded pump bulges. Air lifts have been employed in some areas to increase the head available for the gravity flow principle.

Solvent washing

There are three identical separate wash systems (Figures 13, 14, 15) for the partition cycle, the first uranium purification cycle and the second uranium purification cycle of solvent extraction. Waste solvent from the plutonium purification cycle is recycled into the contactor (4C-2) of the partition cycle.

Waste solvent is scrubbed countercurrently with sodium carbonate solution in a column packed with rings, to remove the bulk of lightly complexed or entrained activity, uranium and plutonium. Washed solvent flows by gravity into a long residence Holley-Mott type contactor, where it is washed in a single stage with recirculated sodium carbonate solution for an average residence time of about 45 minutes to remove strongly complexed fission products. It has been assumed that mixing in this contactor may be accomplished by pulse units rather than a stirrer, to simplify the mechanical section of the plant.

Solvent is finally washed with dilute nitric acid to neutralise and remove traces of entrained sodium carbonate solution. The solution is collected in stock tanks for recycle to the appropriate extraction cycles.

5.2.4 Product treatment and export

Uranium product concentration and tail end purification (Figure 20)

The uranium product solution from the second uranium purification cycle (IEU) is routed into the feed tank (5D-7) to the uranium product evaporator (5C-4). The solution is fed to the evaporator by air lift and heated by an external reboiler. The evaporator (5C-4) has been scaled up to 3 t U/d capacity from data provided by NFS on a 1 t U/d unit.

The feed rate to the evaporator is controlled by regulation of the air supply to the air lift according to the level of liquor in the feed tank. The evaporator steam supply is regulated by a density measurement on the boiling concentrate. Evaporator overheads are condensed (5E-4) and collected in a condensate tank (5D-8) from where they are cooled and recycled to the second uranium cycle strip make up tank. Continuous take-off rate for concentrate is controlled according to the level in the evaporator. The solution is concentrated to 300 g U/l.

Concentrate is removed from the base of the evaporator through a level control vessel (5D-4) which received an air pressure from a level controller on the evaporator. The concentrate is cooled (5C-10) and transferred to a uranium product surge tank (5D-9) feeding either of two silica gel bed columns (5C6A and B) for final uranium decontamination. The beds have been scaled up on diameter from NFS design pro rata to flowrate.

Uranium product solution from the silica gel beds is collected in product sample tanks (5D12A and B) to await analysis for specification. Satisfactory product uranium is pumped into uranium storage tanks (5D-15A and B) for export in tank cars.

Off-specification uranium is routed back to the partition cycle feed tank (4D-1).

Regeneration effluent liquor from the silica gel beds is routed into low active waste evaporation; waste rinse acid effluent from the beds is recycled to the uranium product evaporator.

Plutonium ion exchange purification (Figure 17)

The aqueous plutonium stream from the plutonium stripping column (4C-8) is collected in the ion exchange feed conditioner (5D-1), and acidified to 7.2 M in nitric acid. The plutonium is then loaded onto 6 anion exchange columns (5C-1A, B, C, D, E, F) at 60°C. Figure 82 shows a typical ion exchange column; it contains a resin bed 152 mm in diameter and 2.3 m tall, and is of identical design to those used in the NFS plant.

Six ion exchange columns were estimated to be adequate for the larger plutonium throughput (27.9 kg Pu/d) arising from a 3 t U/d flowsheet at increased burnup. This estimate was based on statements on the NFS plant in USAEC Docket 50201-3:

- the columns are loaded to a concentration of 60 g Pu/l.
- the maximum theoretical saturation is about 150 g Pu/l.

The cycle time and number of columns required is derived in the appendix. The columns are assumed to be loaded to about 70 g Pu/l.

Nicholson (1967) has estimated the maximum instantaneous capacity of the NFS 3-column anion exchange system to be 14.6 kg Pu/d; however, because of batch processing and column feed conditioning and waste handling facilities, an actual capacity of about 8 kg Pu/d is estimated for a 3-column system. Nicholson recommends a continuous system of ion exchange using a Higgins moving bed contactor.

The proportion of expense incurred by plutonium ion exchange is small in comparison with the total plant cost and the design and cost data have accordingly been based on a 6-column batch operated system.

The rates selected for washing, elution, and reconditioning washes for the ion exchange columns are identical to operating data for the NFS units (3.8, 0.28 and 3.8 $mm^3\ s^{-1}\ mm^{-2}$ respectively).

The rate of loading the column with plutonium is about 1.1 $mm^3\ s^{-1}\ mm^{-2}$ (compared with NFS data of 0.36 $mm^3\ s^{-1}\ mm^{-2}$). This is judged acceptable as

- loading rates of up to 100 (US) gal/hr.ft^2 (1.1 $mm^3\ s^{-1}\ mm^{-2}$) are referenced (Long 1967).
- raffinate from the loading operation is recycled to the plutonium purification cycle, and an excessive loading rate would not incur loss of plutonium.

Pressure gradients of 8.9 and 27.6 kPa have been estimated (see Appendix 2.6) for loading and washing conditions with a resin size of 20 mesh. These gradients are respectively increased to 68 and 234 kPa with a 40 mesh resin. Because of the potential resistance to flow the flow diagram (Figure 17) has incorporated pumped feed and wash to the ion exchange columns.

Plutonium evaporation (Figure 17)

Plutonium nitrate solution received from ion exchange in the product evaporator feed tank (5D-4) is metered by two level controlled air lifts into each of two steam jacketed, geometrically safe plutonium product evaporators (5C-2A, 2B), fabricated from titanium.

Duplicate plutonium product evaporator units of identical design to the original NFS units are provided to maintain dimensional guarantees for nuclear safety.

The evaporator overhead vapours are fed into a single condenser (5E-7) which routes condensate to two steel tanks (5D-6A, 6B), for discharge to low active waste evaporation via the plutonium waste catch tank (4D-8).

The condensate can also be recycled through the evaporation feed tank should it contain an excessive concentration of plutonium (e.g. through evaporator boil over or poor disentrainment).

The plutonium nitrate is concentrated to 200 g Pu/l and withdrawn from the base of the evaporators through control vessels (5Y 22 A & B, 5Y 23 A & B). The rate of discharge through these vessels is determined by the air pressure applied which in turn is controlled by a pre-set density recorder on the evaporator. Concentrated product solution is received in two plutonium storage tanks (5D-5A, 5B).

Plutonium packaging (Figure 17)

The plutonium solution is vacuum transferred from the product storage tanks (5D-5A, 5D-5B) into a calibrated vessel (5D-17), which delivers a measured volume into a product shipping container. The packaging equipment is contained in a ventilated glove box. The filled containers are stored on a safe lattice pending export.

5.2.5 Waste evaporation

Flow diagrams for the high and low level waste evaporator (7C-1 and 7C-2), and rework evaporator (7C-3) are given in Figures 18 and 19. These evaporators are replaced by remote operation; they are accordingly grouped in the same area, and located as modules in the cell floor. All linkages to external pipework are made through remote jumper connections. The evaporators are designed as thermosyphon units with external reboilers and condensers linked to a disentrainment vessel. The reboiler and condenser tube bundles are designed for removal and replacement in situ; the entire unit may be removed using lifting trunnions in the main shell. The evaporators are designed for batch operation, with feeding and removal of liquor by force lift ejectors.

The general purpose evaporator (7C-5) operates at a relatively low level of contamination; it is designed for contact maintenance and is located in the acid recovery area. Steam feed to the reboiler is held at a present level by a pressure controller (Figure 22). The dimensions of the evaporators have been derived from data (Nuclear Fuels Services, Inc., USAEC Docket 50201-2) scaled up to the increased capacity of 3 t U/d.

High level waste evaporation (Figure 18)

Aqueous raffinate from the partition cycle forward extractor (4C-1) is fed semi-continuously by force ejection from the feed tank (7D-1) to the high level waste evaporator (7C-1) which is heated by the external high level waste evaporator reboiler (7E-11). The rate of feed is controlled by regulation of the steam pressure to the force ejector by a level controller on the evaporator reboiler.

Overheads from the evaporator are condensed in the high level waste condenser (7E-1), and routed to the feed tank of the low level waste evaporator.

High level concentrate is withdrawn batchwise from the evaporator by cooling the evaporator reboiler, and transferring by force ejection into the high level waste accountability and neutralisation tank (7D-4). Concentrate is accounted for uranium and plutonium, neutralised, and discharged through a distributor into tanks for storage of high level liquid waste (Figure 23).

Low level waste evaporation (Figure 18)

Low level aqueous wastes are combined in the feed tank (7D-2) and fed by force ejector into the low level evaporator (7C-2) heated by the external reboilers (7E-1 and 7E-22). The feed rate to the evaporator is controlled in a similar manner to that in the high level waste evaporator.

Overheads are condensed in the low level waste evaporator condensers (7E-7) and routed to the acid fractionator feed vapouriser (item 31003).

Concentrate is withdrawn batchwise at a preset density by cooling the evaporator reboilers and transferring by force ejection into the low level accountability and neutralising tank (7D-10).

Following accounting for uranium and plutonium and subsequent neutralisation, the concentrate is discharged into waste storage tanks through a feed distributor.

5.2.6 Rework evaporation (Figure 19)

Effluents containing appreciable quantities of uranium or plutonium are concentrated and recycled.

The off specification effluent is routed into the feed tank (7D-8) for the rework evaporator (7C-4), and fed to the evaporator by force ejection. Evaporator overheads are condensed (7E-8) and routed to low active waste evaporation.

5.2.7 Recovery of nitric acid

Figure 21 shows equipment for nitric acid recovery. Effluents containing nitric acid (e.g. leacher off-gas scrubber raffinate, raffinate streams for forward extractors of the solvent extraction cycles, and condensate overhead from the high level and rework evaporators) are concentrated in the low level evaporator. The acid condensate from the evaporator contains the bulk of the nitric acid in the original effluents at a low concentration.

The condensate is fed to a vaporiser (item 31003). The underflow from the vaporiser is cooled and collected in 'hot' acid storage tanks, from where it is recycled to the leachers.

The acid vapour from the vaporiser, which was further decontaminated through evaporation is fed into a vacuum acid fractionator (Figure 81). The fractionation requires 13 trays having 13 mm dia. holes with a 10% free hole area per tray. The diameter of the concentration and stripping sections of the fractionator have been estimated by enlarging the dimensions quoted by NFS pro-rata to the increased throughput.

Vaporised feed is introduced to the column at trays 5, 6 and 7. The column is reboiled by an external heat exchanger (item 31005) at the base of the fractionator, from which recovered nitric acid is withdrawn for recycle.

Overhead vapour from the fractionator is condensed in the fractionation condenser (7E-3); part of the condensate is metered to the top of the fractionator stripping section to maintain the desired reflux ratio, and the excess is collected in a weak acid catch tank (7D-6) for effluent treatment and ultimate discharge.

The rate of discharge of overheads is controlled by heat input to the column, and the recovered acid withdrawal rate by a level recorder on the fractionator reboiler.

The reduced pressure in the fractionator and feed vaporiser is achieved by a two stage vacuum ejector system, with intermediate and after condensers. The preset degree of vacuum is automatically maintained by regulation of an air intake into the fractionator condenser by a pressure recorder controller on the fractionator.

5.2.8 Effluent treatment and discharge (Figure 22)

Condensed overheads from the acid fractionator are collected in the feed tank to the general purpose evaporator (7D-7). The liquor is pumped into the general purpose evaporator (7C-5), which is heated in an external reboiler (7E-51). The feed to the evaporator is controlled from the level in the evaporator.

Concentrate from the evaporator is withdrawn batchwise at a preset density after a cooling period and recycled by pump into the feed tank for the low level evaporator. Overheads from general purpose evaporation are condensed in the general purpose evaporator condenser (7E-13), and routed to the feed tank (7D-9) for the waste ion exchangers (7C-10 and 7C-11).

Liquor is fed through the ion exchange units into the storage lagoon for aqueous effluent. This liquor may either be recycled as make up water for the plant, or discharged as aqueous effluent. The ion exchange columns are regenerated by a flow recycled to the low level waste evaporator feed tank.

5.2.9 Storage of fission product wastes (Figure 23)

Concentrates from the high and low level accountability and neutralisation tanks (7D-4, 7D-10) are transferred batchwise by steam ejection through a storage distributor routing to the waste storage tanks, each having a liquor capacity of 2,850 m^3. Two tanks + 1 spare are required for about 5 years storage of mixed high activity and low activity waste; segregation of wastes would involve 4 tanks (1 high activity, 2 low activity, 1 spare).

5.3 Building Design and Layout

Sectional plans and elevations of the reprocessing plant are given in Figures 34-43. The plant consists of a fuel receipt, storage and transfer building linked to the building housing the reprocessing plant by an aisle carrying an overhead crane conveying shielded assemblies for reprocessing. Detailed layouts of process equipment in shielded cells are given in Figures 24-33.

The receipt and storage pond (Figure 35) contains areas for unloading transport flasks under a water shield, and for their subsequent decontamination, together with a storage area capable of accommodating about 1000 fuel assemblies single stacked in parallel rack storage, equivalent to about 3 and 11 months arising from a reactor programme of 20,000 MW(e) of BWR's and PWR's, respectively. The pond includes an area for underwater loading of assemblies into a transit flask and for flask decontamination prior to transfer to the shear-leach cell.

The reprocessing building (Figures 36-43) contains the shielded processing cells and three upper levels housing plant controls, ventilation systems, and inactive feed equipment.

The shear-leach and evaporator cells (Figures 24, 25, 31, 40), requiring remote maintenance techniques, are constructed of concrete 1.8 m thick and are housed in line together with a maintenance pond for disposal of failed equipment and a smaller pond for maintenance of shear packs. The cells are serviced by the 50 ton overhead crane used for fuel transfer.

Export of Zircaloy waste flasks is made at right angles to the shear-leach cell through a shielded cell provided with air locks.

The remainder of the processing cells, designed for contact maintenance, are constructed from concrete shielding ranging from 0.5 m to 1.8 m thick, and are located adjacent to the remote cells. Common walls for cells have been adopted where possible for economy. Six cells for analytical sample dispensing are provided adjacent to the main block, and contain shielded viewing windows and master-slave manipulators. Liquor transfer ejectors are located in external shielded bulges on the cell faces. Two aisles house pumping equipment for the acid recovery and uranium evaporation and the solvent washing cells; glove boxes contain sampling and pumping equipment for the plutonium evaporation cell. Plutonium nitrate solution is packed and stored in a shielded area adjacent to the processing cells, provided with an air lock for vehicular access. Uranyl nitrate product solution is pumped into storage tanks external to the plant.

The building is entered through an administrative area containing laboratories, offices and toilet facilities. Change rooms are provided in a separate building on the site.

A personnel and equipment lift has been provided to convey sample stillages from stations to the dispensing cells and to transport bulk solid chemicals to the process chemical make up area.

5.3.1 Site plan

A tentative layout of the reprocessing plant ancillary buildings, services and waste disposal facilities is shown in Figure 44, primarily for assessment of land requirements and costs and service reticulation charges.

An active area of 0.14 km^2 surrounded by a perimeter fence is adequate to house the reprocessing plant and high level liquid waste storage facilities and the solid waste burial ground, together with decontamination and active maintenance shops. An area within the fence has been allocated for possible future processes, e.g. waste fixation methods and a plant for recovery of neptunium from wastes.

Inactive workshops and service buildings, canteens, and administrative buildings are located outside the perimeter fence of the active area. The plant changerooms are located adjacent to the fence.

An overall exclusion area of 16 km^2 is allowed for the plant in accordance with USAEC standards.

5.4 Plant Maintenance Philosophy

The plant design requirements necessary to permit maintenance of equipment have been approached from three fundamental requirements:

- the frequency of maintenance - whether routine or non-routinely for repair or replacement of mechanical or chemical engineering plant items
- the $\beta\gamma$ activity level of the plant to be maintained
- the likelihood of plant failure with respect to its operational environment.

5.4.1 Routine maintenance

Typical routine maintenance operations are replacement of blades and bearing surfaces in the shear pack, maintenance of blocked force-lift ejectors, and replacement of instrument heads on active lines.

Shear packs (Figure 25)

There is no documented information on the frequency of maintenance for shear packs involved in transverse shearing of whole fuel assemblies. The life of blades and bearing surfaces in the pack will depend on pack design and choice of materials of construction, and on the physical size and shape of assemblies to be sheared and the physical characteristics of the irradiated metallic shroud and element cladding. The ultimate limitations on throughput of the BNFL Head End plant are likely to be associated with maintenance of the shear (Hughes et al. 1971) which is maintained by remote disassembly and prolonged decontamination to levels permitting replacement of parts by contact maintenance.

For purposes of the study, a pack life between 20 and 80 t U has been assumed, equivalent to operational intervals of about 1 and 4 weeks.

The shear pack is designed to be water washed in situ, the slurry of uranium dioxide being collected in one of the leachers. The necessary direction and force of wash jets is achieved by a rotating spray nozzle mounted on a wash pipe inserted through a sleeve in the cell roof through an aperture in the casing of the shear.

The washed shear pack is lifted from its actuating linkages by a special purpose rotating grab, which lowers the pack onto a trolley in a maintenance pond. A new pack is installed in the shear cell by the rotating grab to effect continuity of plant operation, and the failed pack is dismantled and rebuilt under a water radiation shield.

Force-lift ejectors

These ejectors are liable to failure through blockage or gradual falloff in performance through erosion of the venturi throat. They are located in shielded bulges external to the biological shield, and are maintained either by contact, or behind a temporary shield after decontamination to acceptable levels. Ejectors in continuous operation (e.g. feed ejectors to the waste evaporators) are duplicated to assure continuity of operation.

These items are mounted on shielded cubicles external to the biological shield, and are removed by preliminary washdown and transfer into shielded disposal flasks.

5.4.2 Philosophy of non-routine maintenance

Plant equipment requiring non-routine maintenance has been divided into two categories.

(a) Equipment which is highly contaminated by $\beta\gamma$ activity and which is vulnerable to infrequent mechanical breakdown or failure from corrosion. This category includes mechanical equipment in the shear-leach area, and chemical engineering equipment in the shear-leach and waste evaporation areas. This equipment is removed and replaced by semi-remote means. Access to the cell is gained by removal of roof slabs, and tools (e.g. impact wrenches) necessary to remove equipment fixtures are lowered into the cell from the external crane. Local shielding for the maintenance operation is erected as and when necessary.

(b) The remainder of the plant where $\beta\gamma$ activity is handled under conditions not favouring corrosion failure (e.g. the low acidity and low temperature prevailing for solvent extraction) or where liquors of relatively low $\beta\gamma$ activity are handled in a corrosive environment (e.g. product evaporation). These areas will be maintained by direct access after prior decontamination. Experience at the Dounreay and Windscale plants of BNFL (Boyle et al. 1971) has demonstrated that, given an adequate decontamination time, modifications are possible in even the highly active sections of a reprocessing plant.

Semi-remote maintenance

The conventional design for remote maintenance of plant involved provision of in-cell general purpose handling equipment (traversing power manipulators and cranes). The bulk of operations performed by this equipment are routine handling operations for fuel assemblies and leacher baskets within the cell. The general purpose handling equipment is employed to disconnect failed plant items from service and process lines, and to convey the items to shielded decontamination and maintenance facilities. This design approach results in a considerable increase in shielded volume requirements over the minimum necessary to house process equipment and perform routine operations (Figure 9). The additional capital investment is incurred during plant construction as an investment against a failure contingency which may not occur.

A system of semi-remote maintenance has been adopted for the shear leach and waste evaporation areas of Design 1. Routine transfer operations for baskets and shear pack maintenance are performed by special purpose handling equipment tailored to the movements required. Removal and replacement of failed plant equipment is treated as an emergency operation, for which temporary shielding and containment is constructed, and the external crane employed to lift roof slabs and insert tools for semi-remote maintenance.

The concept of provision of mobile radiation shields or temporary shielding panels for maintenance personnel has been developed by the French (Vertut et al. 1971). These shields consist of interlocked lead or steel bricks and are provided with viewing windows and simple facilities for remote manipulation. Broadley (1969) described the use of a movable personnel shield in UKAEA facilities. The shield, suspended from an external crane permitted maintenance operations in a radiation background of up to 1000 R/hr $\beta\gamma$. It consisted of a lead box, 2.4 m high, approximately one square metre internal cross section, constructed from 100 mm thick lead bricks weighing about 5 tonnes. One side of the box was open to permit the contained operator to work an an item and be shielded on three sides and from below from the general radiation flux in the cell.

The shear-leach and waste evaporation area have been sited adjacent to each other and to a receipt pond for failed equipment. The cells have been provided with removable roof slabs, and are located under the traverse of the 50 ton crane (Figure 40). Chemical engineering equipment is positioned in the cells on a modular basis using dowel fixtures. Connections to process and service lines are made through remote disconnects, each disconnect serving a bank of lines.

In the event of equipment failure, the following operations are envisaged:

(a) The general $\beta\gamma$ background radiation level in the cell would be reduced by removal of active solutions from the cell under maintenance. The equipment to be removed is decontaminated from loosely held activity.

(b) Roof slabs to be removed for maintenance are defined, and a PVC tent is erected above the proposed aperture to reduce spread of airborne contamination. The area of roof over which failed equipment will be transported between the roof aperture and the maintenance and disposal pond is covered with sisal to retain spillages of activity during transfer.

(c) A radiation shield for maintenance personnel is located inside the containment area on the cell roof adjacent to the proposed aperture.

(d) The external crane is used to lift the appropriate roof slabs and expose the equipment to be removed. The latter is disconnected from service and process lines at the multibank jumper connections which are dismantled by an impact wrench lowered either from the external crane or from a subsidiary manipulator attached to the crane. An alternative approach would be the introduction of a power manipulator attached to and controlled from the shielded cubicle housing the maintenance operators. The external crane is finally used to transfer the equipment from the cell into the maintenance pond

(Figure 40) where it can be dismantled under a water shield for disposal in shielded flasks. The replacement equipment is introduced by the overhead crane and coupled to the service and process line disconnects.

6. DESIGN 2

6.1 Theme for Reduction of Capital Cost

An assessment of cost structures in conventional reprocessing plants showed that the solvent extraction function was a major contributor to capital costs, representing about 18% of the total cost of a reprocessing complex (Costello 1972a).

Design 2 is a variant of Design 1 in which special emphasis has been placed on compaction of the shielded space required for the solvent extraction areas by the use of compact extraction equipment of high capacity. The extractors have been sited in the roof of cells housing process feed and waste tankage, eliminating the 70 foot well required to house pulsed columns in Design 1. Design 2 operates on the chemical flowsheet for Design 1, previously described in Section 4 and Figures 2-8.

6.2 Equipment Description

With the exception of the solvent extraction equipment, Design 2 employs chemical engineering equipment identical to that in Design 1, for which flow diagrams are given in Figures 10-23. The solvent extractors have been allocated plant equipment identification numbers corresponding to the pulsed columns which they replace, to facilitate use of flow diagrams Figures 12-16. Thus contactor 4CC-1E corresponds to the extraction section of pulsed column 4C-1.

Other differences in plant equipment between Designs 2 and 1 are the elimination of interface control vessels for the contactors, the use of mixer settlers as solvent washers, and the provision of clarification equipment prior to the first extraction unit.

6.2.1 Solid clarification

The product solution resulting from the fuel leachers will contain some finely divided solid particulate in suspension, which is slow to settle under gravity. The solids range from particles of Zircaloy cladding and ceramic fuel spacers, fragmented by the shear, to insoluble carbonaceous material resulting from the organic binder used in fuel manufacture, elemental fission products, and possibly undissolved plutonium oxide.

Pulsed column solvent extraction equipment accepts these solids to some degree without operational difficulty; however, the solids may settle out and block internal liquor routes under the high settling forces of a centrifugal contactor, and it is considered desirable to remove solids from the leacher product. Suitable equipment for this purpose includes filters or centrifuges.

A centrifugal separator with batch removal of collected solids into High Active Evaporation has been logically included in the design, as it involves similar problems in design, operation and maintenance as the contactors it serves and does not require the periodic removal and replacement of highly active filters.

6.2.2 Centrifugal contactors

Centrifugal solvent extraction contactors operate on similar principles to those of gravity mixer-settlers, where mass transfer of solute between two immiscible liquid phases is effected by successive mixing and separation of the phases in separation chambers. The main drawback of the mixer-settler is the relatively large area and hence equipment size required to effect phase disengagement; in centrifugal contactors, centrifugal fields in excess of 1000 g ($g = 9.81 \text{ m s}^{-2}$) achieve rapid phase separation in highly compact equipment.

The advantages of the centrifugal contactor in the context of fuel reprocessing derive primarily from its relatively small physical size when compared with alternative contactor types having similar throughput. This results in the following design features:

(a) a small shielded volume for contactor installation, and hence a reduction in the civil costs of the plant. Against this, however, must be set the increased cost of the contactor, which may be 4 times as expensive as alternative solvent extraction units;

(b) a low hold up of process solutions, resulting in reduction of inventory charges for uranium and plutonium. The small size of the contactors also facilitates design for nuclear safety against criticality;

(c) a short contact time, resulting in an increased solvent life in the highly active extractor of the partition cycle;

(d) ease of startup and rundown operations, and acceptance of a wide range of flowrates of fluids having a variety of physical properties.

High speed centrifugal contactors with holdup times of the order of 5 s per stage and capacities up to 100 gal/min have been developed at Savannah River (Long 1967). These are single stage units comprising a mixing chamber, a centrifugal separator, an air controlled weir for interface control, collection chambers for separated phases, and interstage piping. Each contactor has its own drive motor (typically 5 HP) and a cascade of eight units necessary for countercurrent extraction could occupy a shielded space of about 2 m x 6 m in area and 1.5 m in height for the active contactors alone. The motors would increase the overall height by about 1 m. This design of centrifugal contactor occupies shielded space similar to that of a mixer settler of equivalent capacity.

The 'Robatel' contactor

A major advance in compaction of the centrifugal contactor has been achieved in a design marketed by SGN-Robatel, where up to 10 actual stages of solvent extraction with a liquor throughput (solvent + aqueous) of 5.5 m^3h^{-1} are contained in a single unit of approximate dimensions 1 m (dia) x 1.4 m. Twelve stages has been considered a possible maximum for the design (Bernard 1971). The principles of the design are shown in Figure 45. A stage of extraction consists of an annular gap between a fixed central drum and a rotating bowl. The annulus is divided into a mixing chamber and a disengagement chamber by radial blades attached to the rotating bowl. A disc attached to the stationary drum is located between a pair of radial blades forming the mixing compartment. In operation, the bowl is rotated about a vertical axis. The relative velocities between moving and stationary surfaces give rise to an intense turbulent mixing of the two phases. The moving blades and fixed disc further act as a centrifugal two stage pump and transfer the mixed phase from the mixing compartment into the adjoining settling chamber, where phase disengagement occurs under centrifugal force.

Solvent is transferred from a settler to the successive mixer over a circular weir of precise radius. The aqueous phase is transferred through a series of axial and radial channels drilled in the periphery of the cylinder, ducting the liquor into the mixer of the preceding stage. The geometry of the weirs and ducts is determined by hydrostatic balance for the practical range of phase densities to be handled by the contactor. Ducts may be provided in the central cylindrical drum to introduce auxiliary process feeds into intermediate stages of the contactor.

In the nuclear application of the contactor (Figure 46), the bowl is attached to the lower end of a pendular shaft, which passes through a seal in the roof of the cell in which the contactor is located. The bowl is contained in an external cylindrical casing to which all pipe connections are made. The mechanical bearing support for the pendular shaft and the electrical driving unit are supported by a removable circular steel central shielding plate recessed into the shielding of the cell roof.

Drainage of contactors

Stoppage of a contactor drive (e.g. through a power failure, or preceding maintenance operations, or at plant shutdown) results in drainage of the combined volume of both phases from the contactor (about 100ℓ max.) into collection vessels from which they are recycled at a low flowrate to the first extraction contactor (Bernard 1971). Two collecting vessels have been incorporated in the design for nuclear safety to effect segregation of liquors containing high and low levels of plutonium. Thus a vessel of 0.20 m^3 capacity receives phases from the plutonium contactors 4C-3, 4C-7E, 4C-7S, and 4C-8, while a vessel of 2.5 m^3 capacity collects drainage from all other contactors. These collection vessels are assumed to contain fixed neutron absorbers in the form of rings.

Selection of contactors

Table 5 shows the type and location of centrifugal contactors proposed for the solvent extraction duty of Design 2. Fourteen contactors are employed, with separate units for extraction, scrubbing and stripping stages.

The contactors were selected by comparison of flowsheet duties (Figures 4-6) with data supplied by SGN-Robatel on the number of stages and total flowrate capacity of production models.

6.2.3 Solvent washing

As for Design 1, there are three separate identical wash systems, for solvent from the partition cycle, and the first and second uranium purification cycles. Waste solvent is scrubbed countercurrently with sodium carbonate solution in a 4-stage mixer settler, followed by a long residence wash with sodium carbonate in a Holly-Mott type contactor, and final washing with dilute nitric acid in a 2-stage mixer settler. Sizing calculations for the mixer settlers given in Appendix 2.5 have been based on mixing residence times of 60 s and specific settling areas of 1.5 $mm^3\ s^{-1}\ mm^{-2}$.

6.3 Building Design and Layout

Figures 47-50 show the layout in those cells affected by the inclusion of centrifugal solvent extraction contactors in Design 2. Figures 47 and 48 show a sectional elevation and plan view respectively of the layout of the solids centrifuge for leached liquor and the highly active contactors 4CC-1E and 4CC-1S, located in the roof of the highly active liquor cell, and of the location of all other contactors in the roof of the low active liquor cell.

The necessary pressure head to effect transfer of liquor between contactors is achieved in the contactor design. For each contactor, the aqueous phase is inducted by a centrifugal turbine set at the base of the rotating bowl,

TABLE 5

LOCATION AND DUTY OF CENTRIFUGAL CONTACTORS

Contactor Type Number	Location On Flow Diagrams	No. Of Stages		Maximum Total Phase Capacity		Overall Size Under Shielding	
		(a) Required	(b) Maximum Available In Production Models **	(a) Required	(b) Available *	Dia	Height
				(m^3 h^{-1})	(m^3 h^{-1})	(mm)	(mm)
240-N	4C-7S	8	8	0.2	0.24	450	900
	4C-8	12		0.2			
420-N	4C-3	9	10	0.5	1.2	780	1,200
	4C-7E	11		0.7			
530-N	4C-1E	9	10	2.0	2.7	940	1,350
	4C-1S	6		1.7			
	4C-2	10		2.0			
	4C-9S	5		1.8			
	4C-11S			1.8			
670-N	4C-4	9-15	10	3.4	5.5	1,100	1,500
	4C-9E	9		3.7			
	4C-10	10		3.1			
	4C-11E	9		3.7			
	4C-12	10		3.7			

* At Solvent Fraction = 0.5

** Up to 12 stages are considered possible (Bernard 1971) while the organic phase is discharged under pressure through the radial perforations of the fixed disc which dips into the annular groove in the rotating bowl. These devices allow installation of the contactors at the same level without need to provide intermediate transfer systems.

The size of the cells has been increased marginally to accommodate the drain tanks for the contactors. The solvent washing equipment for the three process cycles has been incorporated in a single cell (Figure 50).

Sectional plans and elevation of the reprocessing plant are given in Figures 51-55.

6.4 Plant Maintenance

Maintenance of failed equipment in the shear-leach and waste evaporation cells is conducted by the semi-remote technique proposed in Design 1; equipment in all other cells with the exception of the centrifugal contactors is maintained by contact, after all equipment in the cell to be entered has been decontaminated.

6.4.1 Maintenance of a contactor

The unit has no bearing surfaces inside the active cell, and routine maintenance is confined to the external shaft support bearings and motor drive.

In the event of contactor failure, the entire unit may be removed from its casing into a flask mounted externally on the cell roof. Such operations are preceded by emptying the bowl and performing a preliminary decontamination to reduce gross levels of contamination and radiation. The mechanical driving system and bearing support are removed; remote couplings are operated to withdraw pipe connections from the path of the bowl. A shielded flask carrying a winch is located by travelling crane over the plug housing the contactor, and the unit (comprising the shielding plug, bowl, shaft and fixed drum) is withdrawn into the flask, which is closed by a horizontal door, and transported to maintenance or disposal. A standby closure is used for the aperture in the cell roof; the disposal flask is subsequently used to install a new contactor and shielding plug.

7. DESIGN 3

7.1 Theme for the Reduction of Capital Cost

Earlier reviews by Baczewski (1958) and Costello (1972a) of the cost structures of reprocessing plants have suggested that a substantial contribution to capital costs is incurred through shielded structures housing process equipment, and the means for remote decontamination and maintenance of equipment, together with the constraints which radiation shielding places on economic layout of vessels and piping.

Design 3 presents the major cost areas accounting for 30 to 50% of conventional reprocessing fixed capital investment (i.e. machanical pretreatment, leaching, and solvent extraction) in the unorthodox situation of underwater location.

A small pilot reprocessing facility for operation on the fuel stream from an Aqueous Homogeneous Reactor featuring replacement of failed equipment below a water shield, has been described by Rom & Weerem (1959). Capital costs for the underwater facilities have not been disclosed.

Underwater location also allows introduction of neutron absorbers as dissolved salts in the pond water surrounding the equipment for additional nuclear safety.

7.2 Location of Facility Above or Below Ground Level

In general, complete location of a reprocessing plant below ground level by excavation into a rock face would not be expected to lead to significant economy in capital cost.

A report on costs of siting of the Halden reactor in Norway (Jamne 1971) indicated that the background siting selected increased the construction time by 10-12 months, and led to an increase of 4% in the total station construction costs, inclusive of interest during construction, as compared to a corresponding reactor constructed above ground.

A tentative comparison of excavation costs in hard and soft rock versus costs of material and construction in ordinary concrete was made by Ebeling (1957) for lined pits and above ground caves (Figure 56). The exercise excluded costs of a roof for the cave or pit, or of an external building. It indicated that for a shielding thickness of about 0.7 m in ordinary concrete, equivalent costs for a pit or above ground cave were attained at internal dimensions of 5 - 6 m for a cubic excavation or cell. Extrapolation suggested that equivalent costs for a shielding thickness of 1.8 m would be reached by an internal side dimension of about 9 m. The size of shielded cells housing the close packed active equipment of Design 3 generally approached or exceeded these dimensions.

A water filled pond of about 9 m depth constructed above ground level would require concrete walls of about 0.75 m thickness to withstand water pressure. Containment of highly active plant required up to 1.8 m of concrete shielding, equivalent to about 1.8 m of water shield in addition to the external wall. Positioning of plant equipment to develop a gradient of activity increasing towards the centre of the pond, to achieve the additional shielding without increasing the floor area of the pond around the equipment, led to many constraints in layout.

It appeared preferable to locate the equipment in a water filled pond submerged below ground level, with all access for operation and maintenance from above, through a water shield of a minimum 3.3 m in thickness.

7.3 Chemical Engineering Equipment

Design 3 operates on a chemical flowsheet similar to that described for design 1.

The type of equipment amenable to underwater operation was restricted. Whole assembly shear-leach operations, which required the opening of a leacher to remove baskets containing leached cladding for disposal, were precluded. Operation of a leacher under a diving bell or alternative means of batch removal of cladding seemed neither realistic nor attractive from the viewpoints of operation, cost and likely spread of contamination.

Separation of the cladding from the fuel prior to feeding into a dissolver was not amenable to underwater operation owing to spread of contamination, and the solution to the leaching stage appeared to lie in the development of a continuous or semi-continuous leacher, with automatic mechanisms for charging sheared fuel and discharge of leached cladding sections.

Operation of a continuous leacher in turn implied a mechanical prebreakdown stage in which end pieces and sheaths would be removed from the assembly for disposal, and individual elements fed to a small scale shear located at the leacher, which would be air jacketed for thermal insulation.

Pulsed column solvent extraction equipment with no moving parts was an automatic choice for underwater operation; the external surfaces of stripping columns operating at 60°C required external insulation.

Problems of insulation arose in areas such as evaporation of effluents and product streams, and fractional distillation of recovered nitric acid. Because of the increased cost involved in providing air jackets for the complex shape of reboilers, interconnecting piping, and disengagement sections of six evaporators and a fractionation column, a compromise was made; the heated units would be operated in a dry cell which could be flooded with pond water in the event of removal and replacement of equipment.

For the purposes of the study it has been assumed that any problem in nuclear safety occurring through operation under fully water reflected conditions can be overcome either by a controlled concentration of soluble neutron absorber (e.g. boron as sodium borate) in the pond water, or by incorporation of fixed neutron absorbers into the constructional material of the relevant items of plant.

7.3.1 Mechanical prebreakdown and shear-leaching

Figure 57 shows the alternative equipment flow diagram for the prebreakdown operations, element shearing and semi-continuous leaching equipment incorporated in Design 3.

The principles adopted for prebreakdown of assemblies underwater into fuel elements for charging to a small underwater shear have followed the approach of an earlier study by Unger et al. (1967).

End pieces are removed from the fuel assembly by underwater saw and elements are removed by a pusher ram. Assembly hardware (sheaths, end pieces) are cut into suitable sizes for burial.

The individual elements are loaded underwater in a rotatable magazine. The magazine is transported and lowered into a shear feed chamber. Elements are discharged from the magazine by ram into a small shear which discharges sheared sections into a semi-continuous leacher.

Fuel magazine

The magazine is a cylindrical device with peripheral grooves along cylinder radii, each groove capable of containing four elements in line. The magazine is contained in a close fitting cylindrical shroud to retain elements (Figure 59).

The basis of the magazine design, detailed in Appendix B3 was on the reference LWR Fuel parameters given in Table 3. A throughput of 3 t U/d was equivalent to 76 BWR elements 14.3 mm diameter x 3.66 m active fuel length, or 1340 PWR elements 10.7 mm diameter x 3.66 m active length. The design of magazine was based on the PWR case.

On a basis of operating schedules given in the Appendix (6 magazines/day operated at ram speeds of 80 mm/s, with handling times per magazine of 0.5 h), the magazine is about 1.4 m diameter and 3.7 m in length having 112 peripheral grooves each holding four elements.

Magazine chamber

The magazine chamber consists of a sealed air-lock provided with rotating and indexing gear for the magazine. The chamber has a lower shaped guide into which groups of elements fall from the magazine grooves during each indexing step and from which a discharge ram feeds the group of elements through a valved tube leading to the shear.

Fuel shear

The conceptual fuel shear consists of two close fitting concentric cylinders drilled transversely to provide a shearing action. The cylinders are driven hydraulically. The unit is mounted in a standpipe projecting above water level, the shear housing being connected to the feed tube from the magazine chamber and to a discharge tube leading to the continuous leacher. A worn shear unit is removed from the guide tube into a shielded discharge flask for disposal.

7.3.2 Continuous leaching equipment

The inherent problems in the design of a continuous leacher are

- Control of contact time of fuel segments with leachant to ensure low losses of unleached fuel.
- Maintenance of fresh leachant adjacent to fuel surfaces by recirculation.
- Adequate discharge for leacher off-gases which could prevent access of acid to the fuel.
- Prevention of accumulation of settled fines, leading to blockages.
- Limitation of leacher geometry, if nuclear safety is to be achieved without addition of soluble neutron poisons.
- Prevention of blockage through springs and spacers contained in many designs of fuel element.
- Compatibility of a leacher to a wide variety of fuel types.

Operation of a continuous leacher requires pre-breakdown of the fuel assembly, to remove end sections and assembly shrouds and present the leacher with a uniform feed of sheared fuel element sections. Any saving in capital investment from design of a continuous leaching unit and elimination of in-cell batch handling steps is offset to some degree by the increased capital requirement inherent in provision, housing, and maintenance of mechanical prebreakdown equipment.

Development work has been reported (Groenier 1971) on five basic types of continuous leacher in the fuel reprocessing context, the types differing in the means for conveying sheared fuel segments through the leacher. The five basic types are discussed below.

Vibrating tray

In this design, the sheared segments are vibrated countercurrently to a stream of leachant acid recirculated from a reservoir. This design has been employed in the General Electric Midwest Fuel Recovery Plant where a design capacity of about 1 t U/d is claimed for a totally enclosed horizontal - slab, vibrated trough leacher of critically safe geometry. The stainless steel trough is about 0.46 m wide by 3.2 m long and contains solution at a depth of

100 mm. The unit is designed to leach all UO_2 into solution on a once through basis. Leached cladding is rinsed and discharged by a vertical vibrating conveyor; insoluble fines are collected and discharged batch-wise.

The operational reliability of this unit has not yet been proven; there is no positive control over the rate at which fuel is transferred through the leacher (i.e. the fuel contact time for leaching).

The spiral vibrating tray design is an undeveloped variant of the above in which fuel is fed to the base of the unit, vibrated upwards along a spiral path flooded with leachant and discharged from the top of the spiral channel. Theoretically, this design had improved control over leaching residence time for segments, but was more susceptible to blockages, and had several problems in collection of off-gases, in recirculation of acid and in suspension of fines, in addition to mechanical failure through vibration fracture.

Spiral screw

The enclosed spiral design features positive feeding of the fuel through a tank of leachant using an Archimedean screw mounted on an inner drum, revolving in an outer steel drum. Fuel contained between adjacent flights of the spiral is advanced through leachant as the screw is rotated, and is eventually discharged from the leacher. Perforations in the screwed flights and the inner drum facilitated zoning of the drum into leaching and rinsing sections. The chief disadvantages of the design were the poor heat transfer characteristics from the drum, and lack of ullage for disengagement of froth within the enclosed screw.

Pulsed U-tube or air lift

A design developed by AERE (Carley-MacAuly and Moulton 1967) employed a U-tube with limbs about 75 mm diameter connected for recirculation of leachant. Sheared fuel sections are charged into one limb onto a stack of sections in contact with acid. Leached cladding is continuously removed from the base of the U-tube through the second limb either by an air lift or liquid pulse. The second limb contains internal projections which are climbed (pulsed ladder technique) by sections being discharged.

A problem with this type of unit is the avoidance of blockages from the variety of shapes and sizes of inert metal solids arising from element shearing (e.g. internal springs) which may not be amenable to discharge at the air lift rate satisfactory for the bulk of cladding sections.

Rotating drum leacher

The basic design of unit consists of a single compartment containing both longitudinal and radial baffles inside a rotating cylinder. Sheared sections fed in through an end port are transferred axially along the drum, countercurrent to a stream of leachant, facilitated by a slight inclination of the baffles in the desired direction of transfer.

A problem encountered with this design is of forward mixing of solids, causing some fuel to be passed rapidly through the leacher and be discharged prematurely, resulting in high losses of unleached fuel. Subsequent development has concentrated on reduction of forward mixing by increase in the length of the barrel, in the number of baffles, or in a multi-compartment design, the most successful being a multi-compartmented trough with conical shaped transfer ducts between compartments (Groenier 1971).

Rotating slab

Figure 58 shows the principles of a semi-continuous slab leacher under development by the French CEA (1970). It consists of an external casing containing a cylindrical slab capable of rotation about an axial mount. The slab is divided into compartments for containment of fuel sections. The slab is submerged for about 30% of its diameter in leachant contained in the external casing; it is rotated by a motorised shaft carrying a bevel gear which meshes with teeth located near the periphery of the slab.

Sheared segments are introduced into a compartment through a feed port in the side of the leacher.

Rotation of the slab results in submergence of the compartment and contents in boiling leachant. The rate of rotation is dependent upon the submergence time required to effect leaching; leaching times of 1.5-2.5 hours are reported for sheared sections varying from 10 mm to 50 mm in length.

The slab is rotated semi-continuously in angular increments equivalent to the spacing between successive compartments. Each port rotation results in discharge of cladding from a fully leached compartment into an exit port, and the presentation of an empty compartment below the inlet port of the leacher. Off-gases are discharged from the top of the containment shell, and product solution removed from a luted vessel attached to the base of the leacher.

It is claimed that a capacity of 3 t U/d can be obtained from a leacher unit of approximate dimensions 3 m x 3.7 m x 0.25 m thickness. The design eliminates problems of forward mixing as the contents of each compartment are leached in isolation. The bearing design and drive mechanisms are simplified.

No design of continuous leacher has been proven in commercial scale reprocessing operations. The rotating slab design of semi-continuous leacher has been incorporated into Design 3 for the following reasons:

- It has a relatively simple design, and apart from two robust slow moving bearings has no moving internal parts and is relatively free from vibration.

- Its mode of operation suggests that it is capable technically of achieving high leaching efficiencies comparable with those of batch leaching units; it is a compact item, requiring a smaller space than a vibrating tray model of similar capacity.

7.3.3 Operation of shear-leach equipment

A schematic layout of the equipment is given in Figure 59. A fuel assembly is lifted from a storage rack in the pond and transferred underwater into the disassembly area by the element hoist on the pond bridge crane. The assembly is lowered in a horizontal position on to a cutting table and the sheath is gripped by two clamps attached to a horizontal shaft capable of rotation through 180°. A movable saw mounted on rails alongside the table is positioned to cut transversely through the assembly adjacent to both top and bottom end plates, which are lifted to a waste flask for burial. The clamped assembly is swung into position over the loading tray, and elements discharged by a removal ram on to the tray.

The assembly sheath and grids remaining after completion of the discharge operations is swung back to the cutting tray and sectioned by saw into pieces for burial.

Elements are manoeuvred in groups of four to the side of the loading tray, which is profiled as a loading hopper for the shear feed magazine.

A magazine is positioned beneath the hopper, and the grooves successively loaded with fuel elements. A loaded magazine is removed from storage and transferred to the magazine chamber (Figure 59). The chamber lid is closed, and the water contents discharged by pressurisation with air. The magazine chamber is vented to atmospheric pressure, and the magazine indexed to allow the contents of one magazine groove to fall into the charging slot at the base of the chamber. The isolating valve to the shear is opened and the elements discharged through the feed tube to the shear. The feed ram propels the cluster of four rods against the shear stop, the shearblade is advanced to shear elements into 50 mm lengths, sections falling through the chute into the leacher. The cycle of feeding and shearing is sequentially repeated; finally the feed ram is retracted, the magazine indexed, and the discharge feeding and shearing operations repeated to empty the magazine.

7.3.4 Disposal of leached cladding sections

Waste cladding discharged intermittently from the semi-continuous leacher is received at the base of a vibrating vertical tower conveyor (Figure 59).

Cladding sections are vibrated upwards and washed with a countercurrent stream of water which is ducted to the leacher as part of the feed acid stream. The washed cladding emerges from the conveyor into a shielded waste flask similar to that in Design 1. The flask is transferred by trolley and hoist through an air lock to a transporter for ultimate transport to the solid waste burial area.

Solid fines settling in the lower section of the leacher are removed at intervals by lift and force ejector and transferred into the account and neutralisation tank for H.A. wastes (item 7D-4) for ultimate storage with fission product concentrates.

7.3.5 Solvent extraction and ancillary processes

The equipment previously described in Section 5.2.3 of Design 1 and given in equipment flow diagrams (Figures 11-24) has been incorporated in Design 3.

7.4 Equipment and Building Layout

A schematic layout of the major equipment items in the mechanical pre-breakdown and shear-leach is shown in Figure 59, and plans and sectional elevations of the processing and ancillary areas in Figures 60-69.

The receipt and storage pond and subsequent active processing areas have been contained in a single excavation, with internal subdivision into cells by walls or gates, to enable traverse of the main 50 ton crane over those areas maintained by semi-remote means in Design 1.

In some cases it has been necessary to make minor alterations to the shape of processing areas for their accommodation within the active excavation; the floor areas required by the detailed equipment layouts of Design 1 (Figures 26-33) have been preserved.

The cells for waste evaporation, uranium and plutonium product evaporation and export, and acid recovery are non-flooded. The waste evaporator cell is segregated from the main pond by gates which permit submergence of equipment at times of maintenance.

8. DESIGN 4

8.1 Theme for Reduction of Capital Cost

Previous design studies for reducing capital investment in small scale reprocessing plants have concentrated on minimising the capital investment in radiological shielding by compaction of the plant equipment, and this approach has been further developed in the preceding three studies. The complexity and size of equipment items in a large scale reprocessing plant in Design 1 shows that, even with compact layout and adoption of unorthodox maintenance concepts to reduce shielded structures to facilitate remote removal, decontamination and replacement of failed equipment, there remains a considerable investment in radiation shielding to protect plant operational personnel.

The design philosophy forming the key of Design 4 has arisen from a suggestion from Sir Philip Baxter that further reduction in shielded space requirements might be obtained by application of the protective shielding to the operational and maintenance personnel rather than to the plant equipment. This approach has been followed by Lloyd (1965) in a design concept for a small scale plant.

This concept presupposes

- that the reprocessing plant would be sited in a remote location, with ground and air exclusion zones to protect the public at large from external γ radiation;
- that the process equipment would be confined to existing world standards for containment to prevent spread of activity by air or water into the biosphere;
- that plant operators and maintenance personnel whose duties cannot be performed remotely are confined within shielding provided with visibility and adequate handling devices, capable of performing all routine and non-routine operations in plant control and maintenance, and with ability to access all items of plant;
- that operators whose duties are of a non-contact nature should obtain radiation protection either through radiological shielding of a non-costly nature (e.g. locating the plant behind a natural earth or rock formation, or the use of an earth bunker) built around the plant.

8.2 Reduction of Shielding Requirements by Separation from the Source of Radiation

One interpretation of the 'shield the active operator' principle could be location of the active plant above ground level in a building designed for containment of activity rather than for attenuation of γ radiation. All plant functions not involving direct contact with large scale active equipment (e.g. make-up and flow control of process feeds, supply of utilities, operation of control station and analytical laboratories) would be located remote from the main plant in a radiation field below that recommended by the ICRP for occupational exposure (2.5 mR/h). This level could be attained either by separation from the source or by a combination of separation and shielding.

Figure 70 puts into perspective the effect of separation and/or shielding requirements, based on the γ radiation from fission products contained in 1 tonne of uranium irradiated to 20,000 MWD/t U at a rating of 20 MW/t U, with preprocessing cooling times for fuel ranging from 180 days to five years. The effect of self absorption of γ within the uranium has been ignored; the effect of distance on attenuation of radiation has been estimated from the inverse square law and the specific absorption of air at NTP for an average γ energy of 0.8 MeV.

At a normal fuel cooling time of 180 days, an operator-source separation of nearly 1 km would be required to reduce the intensity of γ radiation in the open air to 2.5 mR/h. Location of personnel and equipment in a building equivalent to a concrete wall 159 mm thick would reduce the necessary separation distance to about 800 m. At 400 and 100 m from the source, the equivalent of about 0.5 and 0.9 m of concrete shielding would be required respectively; at 3 m separation (corresponding to radioactive plant contained in a shielded cell) a concrete thickness of 1.3 m is indicated.

Protracted cooling of the fuel for five years before processing would not have a marked effect on plant design; the required separation distance in the open air would be reduced to about 0.7 km and the shielded thickness at 3 m separation reduced to about 1 m of concrete.

Location of the non-active plant functions at a location remote from the active plant is likely to lead to a substantial penalty in capital cost through the increased requirement for process and service piping. It is undesirable operationally to locate in the high radiation field adjacent to the plant that equipment which could otherwise be operated and maintained by contact. Active and non-active plant could be located on alternate sides of a natural rock formation, but the capital cost of an adequate containment structure for activity would be incurred.

Location of active plant below ground level in a natural fissure or artificial excavation could offer some advantages. Containment of activity could be achieved by reinforcement of the walls of the excavation and provision of a roof structure. Utilities and inactive feed services could be located on ground level inside the shadow cast in the radiation field by the sides of the excavation. The likelihood of a natural fissure of desired dimensions existing in a suitable site is speculative; furthermore a ravine generated by water erosion would be liable to flooding.

For the purpose of this study this concept has been explored in the context of an artificial excavation in a hypothetical area with a low water table.

8.3 Chemical Engineering Equipment

Design 4 employs chemical engineering equipment identical with that in Design 2. The process receives and shears whole assemblies without prebreakdown, followed by batch leaching of fuel, and solvent extraction in centrifugal contactors.

Equipment Flow Diagrams (Figure 10-23) are applicable to Design 4 with the substitution of centrifugal contactors for the extraction, scrubbing, and stripping functions of the pulsed column units.

8.4 Plant Design Concept

In Design 4, all active shielded cells have been eliminated. The active chemical engineering processing equipment specified for Design 2 has been laid out in two parallel rows in a single chamber, excavated below ground level, and provided with a sealed roof for containment. The equipment is laid out either side of tracks carrying a heavily shielded mobile cabin containing personnel for plant operation and maintenance functions. A schematic layout showing the relative size and shape of containment canyon plant equipment, and mobile shielded cabin is given in Figure 72.

The equipment is grouped according to its location on the flow diagrams; this results in a gradient of activity level along the canyon. Pipework connections from each plant item are made to pipe racks running the length of the chamber on either side of the mobile cabin. Flanged pipework connections to this rack are made to permit replacement of equipment using the manipulators and crane possessed by the cabin. Instrument heads and sampling points are located at the pipe rack for access by the cabin.

The floor under the process equipment and pipe racks is lined in stainless steel, following normal plant practice, to contain spillages of liquor through plant failure or malfunction, or during replacement of pipe runs or failed equipment. Spillages will be contained in three zones: a high and a low γ active zone, and a plutonium active zone resulting from subdivision of the containment area along the activity gradient of the canyon by 1 m high walls clad in stainless steel.

Figure 73 shows a general view of a reprocessing site incorporating these concepts. The receipt and storage pond is in an excavation adjoining the highly active end of the canyon. Fuel assemblies are transferred into the processing area through a water lock. At the low active end of the canyon are areas for entry into and exit from the shielded cabin, for cabin maintenance and for removal of shielded active waste for burial. A turntable is located at the maintenance end of the canyon to facilitate cabin maintenance.

All other plant functions are conducted above ground level outside the direct radiation beam emitted by the canyon. These functions include active laboratories, administration offices, the main control station, services building, canyon ventilation equipment, and feed tanks for process chemicals.

8.5 Operational and Maintenance Cabin

All γ active operations within the plant are assumed to be accomplished by an operator contained in a shielded mobile cabin provided with windows, manipulators and carrying crane. Similar concepts for non-routine intervention in a field of γ radiation have been developed by the French (Vertut 1971). A schematic view of the cabin is shown in Figure 71. It contains an internal space of dimensions 2 m x 2 m x 2 m for an operator and control consoles. Windows are provided on the sides of the cabin, and an entry hatch with shielding door on one face. The cabin is propelled on wheels running on a central track between two rows of equipment in the plant area (Figure 72). It carries its own motorised drive and would additionally require failsafe retrieval gear.

Mechanical operations are performed by the cabin using a 15 ton crane and power manipulator. Shielding provided for the cabin design is equivalent to that for the highly active leaching cell of Design 1, i.e. equivalent to 1.8 m of ordinary concrete. The cabin is assumed to be shielded mainly by concrete containing recessed shielding windows. Alternative shielding materials (lead or steel) are likely to lead to a greater capital cost and could result in shielding windows standing proud of the external faces of the cabin.

8.5.1 Functions to be performed by the cabin

Table 6 lists the major operational and maintenance procedures for a shear-leach-solvent extraction reprocessing plant. Operations selected for the cabin operator are limited to those which would normally be conducted directly on active equipment behind shielding. These operations include all mechanical operations in shear leaching (fuel shearing, basket movement, washing, monitoring and disposal of leached cladding). Operations in the solvent extraction area are restricted to direct sampling of active process streams. Maintenance procedures include changing and disposal of blades and bearings in the shear pack, active filters, pumps and ejectors, and infrequent replacement of silica gel and ion exchange beds. The cabin will also be required to conduct non-routine maintenance (removal and replacement of failed plant items).

All functions which are normally performed on non-active streams are conducted remote from the active plant. These include make-up, sampling and

TABLE 6

PROCESS AND MAINTENANCE OPERATIONS IN FUEL REPROCESSING

Operation	Frequency	Operations Performed by Shielded Cabin
ROUTINE OPERATIONS		
Fuel Charging		
Receipt and storage of assemblies	Weekly	
Transfer of assemblies from store to plant	7-15/day	
Feeding of assemblies to shear	7-15/day	X
Control station - shearing	Intermittent Daily	X
Control station - leaching	Continuous	
Handling of fuel baskets Cladding - washing - monitoring - discharge	4/day	X
Removal of cladding waste	Daily	X
Solvent Extraction		
Make-up and analysis of process feeds	Each Shift	
Control station - solvent extraction	Continuous	
Operation of pumps and ejectors	Intermittent	
Sampling of active streams	20-30/day	X
Setting of pulse units (pulsed columns only)	Daily	
Packaging - Pu product	Continuous	
Export - U product	Daily	
ROUTINE MAINTENANCE		
Removal and repair of shear pack	Weekly/Monthly	X
Replacement of:		
. off-gas filters	Infrequent	
. solvent filters	Weekly	X
. silica gel bed	Infrequent	X
. ion exchange resin	Infrequent	X
Maintenance of active pumps and ejectors	Infrequent	X
Maintenance of inactive pumps and ejectors	Daily/Weekly	
NON ROUTINE MAINTENANCE		
Removal and replacement of failed plant equipment	Infrequent	X

flow control of process feeds, and flows of steam and cooling water, transfer of liquor by air lift ejector or pump, and read-out of plant instrumentation.

Examination of estimated handling times in Table 6 suggests that the cabin operator would be occupied with routine shearing and basket movements for at least 10 - 12 hours per day. Most of these operations require continuous attendance, and severe problems are envisaged of phasing the head end operations with daily sampling requirements, maintenance operations, and the effective outage of the cabin during changeover of shift personnel.

9. INSTRUMENTATION AND SAMPLING

Flow diagrams, Figures 10-23 show the locations of points for sampling and control instrumentation in all designs. These are divided into three classes, in-line radiometric equipment, in-tank radiometric and physical equipment, and inactive feed instrumentation.

9.1 In-line Instrumentation and Sampling of Process Stream

Radioactive instruments with pre-set alarm levels have been installed to monitor process streams for continuous detection of changes in concentration of uranium and plutonium in main and raffinate streams, to achieve early information of plant malfunctions leading to excessive losses of product or lowering of decontamination factors for process stages. Departure from flowsheets may thus be corrected at an early stage, reducing the need for frequent recycle of process solutions. A sample point is provided at the detection head of each instrument for calibration purposes.

Continuous in-line instrumentation on active streams was not incorporated in the NFS plant design, which relies on plant control through periodic sampling and analysis of process streams, with collection of raffinates in waste catch/hold tanks and analysis of the bulk prior to discharge to effluent treatment or to recycle evaporation.

The waste collection tanks for a capacity of 3 t U/d in Design 1 are of the same physical size as those installed in the 1 t U/d NFS plant. The reduction in residence time by a factor of three is judged to be acceptable in view of the greater reliance placed on early correction of departures from flowsheet.

The location of in-line radiological instrumentation are given in Table 7. They are of basically three types: γ monitors to detect fission product breakthrough between successive cycles of decontamination, uranium colorimeters and γ absorptiometers to monitor raffinates and main streams respectively for uranium, and alpha scintillators to monitor plutonium in raffinates and main streams.

9.2 In-tank Instrumentation and Sampling

All major vessels where process operations occur (leachers, conditioners, solvent extraction equipment, evaporators, fractionators) together with feed and waste tanks where liquor is collected are provided with systems to measure physical conditions (e.g. liquor level, temperature, density, interface position) as appropriate. Measurements of temperature are made by thermocouple, resistance thermometer or thermistor, and level, density, and interface position are sensed by pneumatic probes (pneumercators). Information is presented to a central control station by indicators or recorders with high or low level alarms as necessary.

All major process vessels in which liquor is collected (feed tanks, waste tanks, conditioners) are provided with sample points for periodic analysis of liquor content.

9.3 Instrumentation on Inactive Feed Streams

This instrumentation is required to record and control flow rates and concentrations of inactive feeds within pre-set limits, and to provide a positive warning on key streams where a departure from flowsheet could lead to high material losses or impaired safety of the process.

The detection heads and flow controllers are mounted on the outlet lines from process head tanks, and information is presented to a central control station.

The centrifugal contactor has an operational problem of short response time of concentration profiles within the contactor to changes in external flow rates or concentrations. So called 'in situ' instrumentation, applicable in mixer settlers for monitoring of internal stage profiles, is not possible in the centrifugal design. Flowrate controls on the main feed streams to these contactors have been assumed to be duplicated for additional confidence in plant operation, in accordance with a recent evaluation of centrifugal contactors in the reprocessing context (Bernard et al. 1971).

9.4 Stations for Sampling and Radiometric Instrumentation

(Designs 1, 2 and 3)

Each vessel to be sampled is connected by a sampling and a return line to centrally located multi-point shielded sample stations. The liquor to be sampled is air lifted from the vessel to the sampling station, and is received in a glass phial. Sample phials are transferred in shielded containers to the analytical cells. The design of the stations is similar to that of those in the NFS plant; samples of low $\beta\gamma$ activity ($<10^{-3}$ Ci/ml) are taken in an unshielded glove box, samples with medium $\beta\gamma$ levels (<2 Ci/ml) are taken in a glove box surrounded by lead shielding of thickness 50 mm, while highly active samples are taken by remote tongs in a station shielded by 380 mm of steel. The stations are designated types A, B, and C. respectively.

Active liquor is air lifted from process lines to shielded stations provided with sampling facilities where the liquor is presented to the detection heads of radiometric instruments. The shielding requirements of the stations have been assessed from NFS information; five highly active and seven low active detection heads have been grouped in line with external shielding of 380 mm and 50 mm of steel respectively with an overall station dimension of 3.4 m x 1.2 m x 0.6 m.

A detailed breakdown of the requirements of sampling and instrumentation is given in Table 7.

9.5 Sampling and Instrumentation (Design 4)

In this design it has been assumed that sampling and radiometric instrumentation heads will be located for access by the shielded mobile cabin, which will be responsible for sampling and maintenance of instrument detection heads. Read-out of physical and radiometric instruments will be displayed at a central control panel in the shielded zone external to the plant.

10. SAFETY ASSESSMENT

10.1 Criticality

Criticality control for individual plant items in the four design schemes has been achieved by a combination of dimensional, mass, and administrative restrictions, together with the introduction of fixed and soluble neutron absorbers, with the primary aim of achieving a compact plant installation.

The approach to the study has employed the Purex chemical flowsheet and major plant items specified for the operational Nuclear Fuels Services Plant. Equipment has been resized as necessary for a throughput of 3 t U/d, and alternative equipment types (e.g. batch leachers with soluble neutron absorbers, continuous leachers, and centrifugal solvent extraction contractors) have been introduced.

Detailed criticality assessments of equipment items for Shear-Leach Purex Solvent Extraction Processes have been docketed (Nuclear Fuels Services, Inc., USAEC Docket 50201, Allied-Gulf Nuclear Services, USAEC Docket 50332).

A detailed criticality analysis is beyond the scope of this report; the review has concentrated on the principles employed for nuclear safety in the major plant areas with particular reference to areas where the minimum critical infinite sea concentration is exceeded in normal operation (e.g. the fuel leacher basket, the plutonium purification columns, ion exchange columns,

TABLE 7

SUMMARY OF MAJOR REQUIREMENTS FOR SAMPLING AND RADIOMETRIC INSTRUMENTATION IN THE REPROCESSING PLANT

Sample Type	Ident. No.	Sampler Location	Radiometric Instrumentation	Order of Activity Ci dm^{-3}
C, Tank	3C-1	Leacher No. 1	--	1,300
C, Tank	3C-2	Leacher No. 2	--	1,300
C, Tank	3C-3	Semi-continuous leacher *	--	1,300
C, Tank	35001	Leacher off-gas scrubber recirculating tank	--	2
C, Tank	35003	Iodine stripper	--	2
C, Tank	35004	Vessel off-gas scrubber recirculating tank	--	0.3
C, Tank	3D-1	Accountability and feed adjust tank	--	1,300
C, Tank	3D-101	Continuous Leacher overflow tank *	--	1,300
C, Tank	4D-1	Partition cycle feed tank	--	1,300
C, Line	HAW	Partition cycle extraction contactor	UR	800
C, Line	HAP	Partition cycle scrub contactor	UR, βγR	3
C, Tank	4D-2	Partition cycle waste catch tank		800
C, Line	HBU	Partition cycle partition contactor	αR, UR	3
C, Line	HBP	Partition cycle, Pu scrub contactor	αR, UR	3
C, Line	HCU	Partition cycle, strip contactor	UR, βγR	3
C, Line	HCW	Partition cycle, strip contactor	UR	3
C, Tank	4D-6	Pu cycle feed conditioner	--	3
C, Line	IIAW	Pu cycle extraction contactor	αR	3
B, Line	IIAP	Pu cycle extraction contactor	αR, βγR	0.003
C, Tank	4D-8	Pu cycle waste catch tank	--	3
B, Line	IIBW	Pu cycle strip contactor	αR	0.003
B, Line	IIBP	Pu cycle strip contactor	αR	0.003
C, Tank	4D-9	1st U cycle feed conditioner		3
B, Line	1 AU	1st U cycle extraction contactor	UR, βγR	0.003
C, Line	1AW	1st U cycle extraction contactor	UR	3
C, Tank	4D-10	1st U cycle waste catch tank	--	3
B, Line	1BU	1st U cycle strip contactor	UR	0.003

TABLE 7 (cont'd.)

B, Line	1CW	1st U cycle strip contactor	UR	0.003
B, Tank	4D-12	2nd U cycle feed conditioner	--	0.003
B, Line	1DU	2nd U cycle, extraction contactor	UR, $\beta\gamma$R	0.003
B, Line	1DW	2nd U cycle, extraction contactor	UR	0.003
B, Tank	4D-13	2nd U cycle, waste catch tank	--	0.003
B, Line	1EU	2nd U cycle, strip contactor	UR	0.003
B, Line	1EW	2nd U cycle, strip contactor	UR	0.003
C, Tank	4D-101	Uranium centrifuge dump tank *	--	1,300
C, Tank	4D-103	Plutonium centrifuge dump tank *	--	3
C, Tank	13D-7	Solvent treatment waste catch tank	--	0.3
B, Tank	13D-4	No. 1 solvent storage tank	--	0.003
B, Line	13D-1	1st U cycle solvent washer	--	0.3
B, Tank	13D-5	No. 2 solvent storage tank	--	0.0003
B, Line	13D-2	2nd cycle solvent washer	--	0.3
B, Tank	13D-6	No. 3 solvent storage tank	--	0.0003
B, Line	13D-3	3rd cycle solvent washer	--	0.3
B, Line	5C-1A	No. 1 Pu ion exchange column		0.2
B, Line	5C-1B	No. 2 Pu ion exchange column		0.2
B, Line	5C-1C	No. 3 Pu ion exchange column		0.2
B, Line	5C-1D	No. 4 Pu ion exchange column		0.2
B, Line	5C-1E	No. 5 Pu ion exchange column		0.2
A, Tank	7D-9	Condensate ion exchanger feed tank	--	--
A, Line		Condensate ion exchanger	$\beta\gamma$R	--
A, Tank		Storage lagoon		--
C, Tank		Leacher cell sump		1,300
C, Tank		Column cell sump		1,300
C, Tank		Active liquor cell sump		1,300
C, Tank		Evaporator cell sump		1,500
C, Tank		Solvent wash cell sump		2
B, Tank		Pu evaporation cell sump		0.5
B, Tank		Acid recovery cell sump		0.003
B, Tank		Uranium evaporation cell sump		2.0

product evaporator and associated feed tanks) and where maloperation could lead to the minimum critical concentration being exceeded (e.g. waste tanks and evaporators, uranium product evaporator and silica gel purification columns). Nuclear safety has been assessed against the criterion of two simultaneous maloperations.

10.1.1 Criticality - Design 1

Up to 1000 fuel assemblies of fissile content 3.3% ^{235}U or equivalent fissile (U + Pu) are stored in a water pond on a safe fixed lattice spacing.

Assembly shear

The shear delivers cut sections of assembly through the delivery chute. Normally the holdup of fuel in the shear and chute is negligible; for criticality considerations a chute blockage is postulated, allowing sheared debris from one leacher charge to build up in the chute and shear pack. Unmoderated uranium of $\leqslant 5\%$ ^{235}U or equivalent is subcritical.

Leacher basket; batch fuel leacher

Up to 1 t U of fuel as sheared assembly sections is received into a basket 760 mm in diameter and submerged below acid level in a leacher.

The main problem in criticality control is in the high fissile concentration of unleached fuel in the basket. Criticality control is achieved by the guaranteed presence of a soluble neutron absorber, $Cd(NO_3)_2$, in the nitric acid contained in the leacher. A nominal 5-10 g Cd/litre has been shown on the flowsheet; the presence of the absorber in the feed acid is guaranteed by analysis and by instrumentation (BF_3 counters and an associated neutron source) for continuous monitoring of all liquid streams entering the leacher.

Leacher vessel, accountability and feed adjust tank, and partition cycle feed tank

The solution under normal operation contains 320 g U/l, at an enrichment <3.3% ^{235}U, i.e. <10.6 g ^{235}U/l, or equivalent fissile concentration (a mixture of uranium and plutonium isotopes equivalent in reactivity to a solution containing 10.6 g ^{235}U/l and 309.4 g ^{235}U/l). Thus maloperation equivalent to a 50% increase in fuel charged to a leacher would not exceed the critical concentration for an infinite system (16.5 g/l ^{235}U for fuels containing 4% ^{235}U). Additionally, the leacher product solution contains the soluble neutron absorber.

Pulsed column solvent extractors

Concentrations of uranium and plutonium considerably in excess of flowsheet conditions can occur in the column during conditions of maloperation. The maximum concentrations of fissile material for the range of maloperation procedures have been assessed in Table 8 and compared with the concentrations for criticality in the unmodified NFS column geometry. An increase in column diameter is judged necessary to accept the greater throughput (3 t U/d) on the original Purex flowsheet. Columns processing plutonium have been assumed to contain sufficient fixed neutron absorber to be equivalent in nuclear safety criteria to the unmodified columns. The enlarged disengagement sections of the columns also contain fixed neutron absorber plates.

It is concluded that -

- the maximum concentrations of ^{235}U are all below the critical concentration in an infinite system (16.5 g ^{235}U/l at 4% ^{235}U).
- The maximum concentrations of ^{239}Pu or mixture of ^{239}Pu and ^{235}U are considerably below the levels required for criticality in any one column. The Pu columns contain fixed neutron absorbers.

Feed and waste tanks associated with the solvent extraction cycles

The partition cycle feed tank has been considered previously with the fuel leacher vessel; criticality control is achieved through the presence of soluble neutron absorber.

The maximum fissile concentration possible for receipt by the waste tank is derived from a maloperated partition column (Table 8). The equivalent reactivity of ^{235}U + ^{239}U is in excess of infinite sea concentration on the approximate equivalent % total fissile content = % ^{235}U + 1.5 x % ^{239}Pu. The partition cycle waste tank primarily derives its safety from the presence of the soluble neutron absorber present in the leachant acid. Secondary controls are provided by the alarms on radiometric instruments and on inactive feed streams.

The feed and waste tanks, uranium purification cycle, derive their nuclear safety by concentration limitation; the maximum ^{235}U concentration possible in the uranium cycle is less than the infinite sea critical concentration (Table 8).

Control of the feed and waste tank plutonium purification cycle, is achieved by the presence of fixed neutron absorbers (borated steel or glass raschig rings) located in the tanks.

Plutonium ion exchange and evaporation

Criticality control of the ion exchange columns and evaporators has been achieved through geometry limitation.

The increase in plutonium throughput has been achieved by six anion exchange columns, and two product evaporators each of safe diameter. Control of the associated tankage is achieved by geometry limitation or through fixed

TABLE 8

NUCLEAR SAFETY OF PULSED COLUMN CONTACTORS

Column	Fault Condition	Maximum Concentrations of Total Pu and U in Column	Maximum Concentrations of Fissile Pu and U	Concentration for Criticality (Ref.3) in Modified NFS Columns	Background to Assessment
4C-1	1. Loss of acid in scrub feed 2. Interruption of solvent feed	U <226 g/ℓ^{-1} Pu ~ 25 g/ℓ^{-1}	^{235}U <7.5 g/ℓ^{-1} ^{239}Pu ~18 g/ℓ^{-1}	^{235}U = 25 g/ℓ^{-1}* to 40 g/ℓ^{-1}** ^{239}Pu = 16 g/ℓ^{-1}* to 28 g/ℓ^{-1}**	Maximum solubility of U in 30% TBP = 124 g/ (Ref.1); equilibrium maximum saturation of U in low acid nitric solution = 226 g/ (Ref.2); 30% aqueous dispersion (Ref.3). Also Ref. 5, 6, 7.
4C-2	1. Loss of reductant	U 69 g/ℓ^{-1} Pu <2.3 g/ℓ^{-1}	^{235}U 2.4 g/ℓ^{-1} $Pu_{(f)}$ 1.6 g/ℓ^{-1}	^{235}U = 29* to 45 g/ℓ^{-1}** or ^{239}Pu = 13 g/ℓ^{-1}* to 30 g/ℓ^{-1}**	Statements. Refs. 3, 4, 6, 7.
	2. Failure of scrub feed	U 69 g/ℓ^{-1} Pu <46 g/ℓ^{-1}	^{235}U 2.4 g/ℓ^{-1} $Pu_{(f)}$ <32 g/ℓ^{-1}		Column requires fixed neutron absorbers. Statements. Refs. 3 and 4, also Ref. 6, 7.
	3. Loss of acid in scrub feed	U <226 g/ℓ^{-1} Pu 8.4 g/ℓ^{-1}	^{235}U <7.5 g/ℓ^{-1} $Pu_{(f)}$ <5.9 g/ℓ^{-1}		As for Column 4C-1, with 30% solvent dispersion
4C-3	1. Loss of solvent feed	U 10 g/ℓ^{-1} Pu 8.4 g/ℓ^{-1}	^{235}U 0.3 g/ℓ^{-1} $Pu_{(f)}$ 5.9 g/ℓ^{-1}	infinitely safe (152 mm dia.)	
4C-4	1. Failure of strip feed	U <226 g/ℓ^{-1} Pu NIL	^{235}U <7.5 g/ℓ^{-1} $Pu_{(f)}$ -	^{235}U = 20 g/ℓ^{-1}* to 36 g/ℓ^{-1}**	Ref. 5, 6, 7.
4C-7	1. Loss of acid in scrub feed	U NIL Pu <96 g/ℓ^{-1}	^{235}U - $Pu_{(f)}$ 67 g/ℓ^{-1}	^{239}Pu = 38 g/ℓ^{-1} to 600 g/ℓ^{-1}**	Refs. 3, 4, 6, 7. Column requires fixed neutron absorbers
4C-8	1. Failure of strip feed	U NIL Pu <84 g/ℓ^{-1}	^{235}U - $Pu_{(f)}$ 59 g/ℓ^{-1}	Infinitely safe (152 mm dia.)	Refs. 3, 4.
4C-9 and 4C-11	1. Loss of acid in scrub feed	U <226 g/ℓ^{-1} Pu NIL	^{235}U <7.5 g/ℓ^{-1} $Pu_{(f)}$ -	^{235}U = 18 g/ℓ^{-1}* to 28 g/ℓ^{-1}**	As for Column 4C-1
4C-10 and 4C-12	1. Failure of strip feed	U <226 g/ℓ^{-1} Pu NIL	^{235}U 7.5 g/ℓ^{-1} $Pu_{(f)}$ -	^{235}U = 21 g/ℓ^{-1}* to 38 g/ℓ^{-1}**	As for Column 4C-4

Note: Assumed (a) ^{235}U 3.3% (^{235}U + ^{238}U)

(b) $Pu_{(f)}$ = 70% Pu (Total)

* Full water reflection assured (Ref. 7).

** Partial reflection assured (Ref. 6).

References:
1. Stoller, S.M. and Richards, R.T. (1961) p.152.
2. Long, J.T. (1967).
3. Nuclear Fuels Services Inc. (1962a), Table 6.143).
4. Nuclear Fuels Services Inc. (1962b), Appendix 6.143.
5. Allied Gulf Nuclear Services (1970), Table II.
6. Paxton, H.C. et al. (1964), Figures 10 and 29.
7. UKAEA (1965).

neutron poisons. Thus the feed vessels for the ion exchange columns and evaporators, and the product storage vessels are slab tanks of restricted thickness, while the waste tank for the ion exchange column raffinate contains borated steel raschig rings.

Uranium product evaporation and silica gel purification

Under normal operating conditions, the concentrate in the uranium evaporator contains 300 g U/l, or <9.9 g/l ^{235}U, which is below the minimum critical infinite sea concentration. Maloperation through simultaneous stoppage of the evaporator feed pump and failure of the density controller to isolate the steam supply (130°C) could result in concentration of the evaporator contents to 1390 g U/l.

Clearance by geometrical restriction would limit the diameter of reboiler and downcomer limbs to below the minimum critical dimension (about 13 in.). This could be achieved by an evaporator with two reboilers and downcomer in-line. The design has adhered to the prorated NFS unit, i.e. a single reboiler and downcomer, with maximum dimension of about 18 in. and criticality clearance by concentration limitation, with alarms on measurements of density and temperature of the evaporator contents, and on the flow controller regulating feed to the evaporator. Additional safeguard could be obtained by incorporating fixed neutron absorbers into the evaporator structure.

The silica gel purification columns operate with concentration limitation obtained at the evaporator stage above.

Waste evaporation

The high level waste evaporator receives raffinate from the partition column and overheads from the low level waste evaporator. It is controlled on criticality through the presence of the soluble neutron absorber (cadmium nitrate) introduced in the leacher feed acid.

The low level waste evaporator receives various raffinates including those from the plutonium and uranium purification cycles. It obtains criticality control from the radiometric instrumentation used to monitor cycle raffinates and by control of concentration factor through alarms on liquor density and temperature.

Further confidence may be achieved by incorporating fixed neutron absorbers into the design.

The rework evaporator is operated batchwise with criticality control by analysis of feed concentration.

10.1.2 Criticality - Design 2

This design differs from Design 1 in the substitution of centrifugal contactors for pulsed column solvent extraction units.

Centrifugal contactors

Contactors processing uranium in the absence of plutonium derive criticality safety from concentration limitation; the maximum concentration of ^{235}U generated under fault conditions is below the critical concentration for an infinite system (Table 8).

Contactors handling plutonium in the presence or absence of uranium require detailed examination of the volume and geometry of the contactor stages and the possibilities for build up of plutonium either in solution or as a precipitate. The critical concentration for large centrifugal contactors has been claimed (Bernard et al. 1971) to exceed 20 g/l of fissile plutonium, but this level is exceeded in routine flowsheet operation in the plutonium purification cycle. The annular shape of the stages is favourable to geometrical safety; a restricted shape may be acceptable for the external vessel and the internal cylindrical drum, and it is possible to fill the central cavity with a neutron absorbing material.

A major problem peculiar to centrifugal contactors is their short response time to a disturbance and the lack of in-situ control; Bernard et al. (1971) recommended duplication of the flow rate controls on the main (feed) streams to increase the safety.

For the purpose of this report it has been assumed that nuclear safety in contactors handling plutonium may be achieved by a combination of fixed neutron absorbers, geometrical restriction, and guarantees on flowrate and concentration of inactive feeds.

Drainage tanks for contactors

Stoppage of the contactor drive results in drainage of both phases from the contactor. The phases are collected in drainage tanks which derive nuclear safety from the presence of fixed neutron absorbers.

10.1.3 Criticality - Design 3

This design differs from Design 1 in that fuel is shear leached in a semi-continuous leacher without introduction of soluble neutron absorbers. Some equipment is contained submerged below a water shield. It has been assumed that any problems arising from full water reflection may be overcome by incorporation of a soluble neutron absorber (boron as boric acid) in the pond water.

Fuel magazine

The magazine may contain up to 253 kg of 3.3% (or equivalent) ^{235}U in an effective annulus 30 mm thick and 1200 mm dia. It derives nuclear safety from geometrical restriction; in addition, fixed neutron absorbers may be installed in the central space.

Assembly shear and delivery chute

This system derives safety from mass and volume limitation.

Semi-continuous leacher

Fuel is received into a cylindrical rotating slab about 150 mm thick, contained in a jacket of slab geometry and 250 mm thick.

Under normal operation, the bulk of fuel charged will dissolve in one hour i.e. the holdup within the cylindrical rotating slab should not exceed 125-150 kg U of 3% ^{235}U enrichment, and the product solution will contain 320 g U/l.

Conditions of maloperation considered are:

(a) Lack of acid in leacher feed

On the assumption of a total leaching time of four hours, the rotating slab could contain up to 500 kg U. The slab derives nuclear safety primarily from its restricted thickness (the minimum critical infinite slab thickness corresponding to 3% ^{235}U is about 170 mm).

(b) Stoppage of the acid feed to the leacher

The maximum concentration possible for an initial acid concentration of 6N is about 500 g U/l, compared with a minimum critical concentration of 1000 g U/l for 3% ^{235}U or fissile equivalent for an infinite slab 250 mm thick.

(c) Overconcentration of the solution in the leacher

This could occur through simultaneous maloperation of the steam supply to the leach jacket, or of cooling water to the leacher condenser, and failure of the acid feed to the leacher. It is prevented by administrative controls and by alarms on steam pressure and flows and on instrumentation monitoring the level and density of the leacher solution.

Leacher overflow vessel

Product solution overflowing from the leacher is collected and cooled in this vessel prior to accountancy and feeding to solvent extraction. It is a vessel of non-ever-safe shape, and derives its safety from:

(a) limitation on the concentration of solution in the leacher;

(b) limitation of volume (a volume of 1.6 m^3 is the minimum critical corresponding to a concentration of 500 g U/l of 3% ^{235}U or equivalent);

(c) fixed neutron absorbers may be introduced into the tank if desired to increase its capacity or as further guarantees for nuclear safety.

Accountancy and feed vessel to partition cycles

Nuclear safety of these items is assured by concentration limitation.

Partition extraction column 4C-1

Maximum concentration of ^{235}U and fissile plutonium is considerably below the level required for criticality in the colum (Table 8) even in the absence of soluble neutron absorber.

High level waste tank for the partition cycle

This vessel receives raffinate from the partition column. Under maloperation conditions (Table 8) the concentration of fissile material in the column could exceed the minimum critical; nuclear safety for this vessel is achieved by installation of fixed neutron absorbers.

High level waste evaporation system

This system may be controlled by installation of fixed neutron absorbers and administratively by provision of alarms for flow controllers on column inactive feeds and by alarms on the raffinate stream from the partition column to detect abnormal levels of fissile material.

10.1.4 Criticality - Design 4

This design uses equipment identical to that in Design 2.

10.1.5 Nuclear interaction

A principle adopted for cost reduction in design has been improvement in utilisation in shielded space by grouping of equipment by shape, and spacing of equipment on access requirements for installation and maintenance.

A detailed analysis of interaction has not been part of this study.

It was assumed that any nuclear interaction problems arising out of a detailed examination of the compact layouts could in principle be overcome without an appreciable increase in the cell volumes (e.g. by separation of vessels containing the fissile interactants by 200 mm of concrete).

The design principles adopted to avoid interaction are as follows:

The batch leachers located adjacent in the shear leach cell of Design 1 contain a soluble neutron absorber and operate normally at a fissile concentration below that required for criticality in an infinite system. The extraction columns containing plutonium at above 'infinite sea' concentration in normal operation (4C-7, 4C-8), and 4C-2, where the 'infinite sea' levels can be exceeded in maloperation are spaced in the cell at least 2 m apart; concentrations of uranium cannot exceed the 'infinite sea' concentration for 3% ^{235}U.

The centrifugal solvent extraction contactors in Design 2 have been located in the roof of the high and low liquor cells. They are mutually spaced by about 1 m of concrete.

The shear leach, and solvent extraction equipment and associated tankage in Design 3 are located below a water shield. It has been assumed that any problems in nuclear interaction could be resolved by incorporation of a cheap soluble neutron absorber (e.g. boron or boric acid) dissolved in the water surrounding the equipment.

Evaporation of uranium and plutonium product streams and of wastes from solvent extraction are carried out in dry cells; with an equipment layout similar to that for Design 1.

There is no subdivisions of equipment into cells in Design 4; it is located in two parallel arrays spaced about 8 m apart. It has been assumed that any interaction problems arising out of a detailed examination could be resolved by construction of local concrete shields.

10.2 Containment

10.2.1 Ventilation

Ventilation systems are segregated on an activity basis into three levels, (a) vessel off-gas extract, (b) cell air extract, and (c) extract from operating aisles and offices.

Vessel off-gases

Vessels other than the leachers are maintained at about 250 mm (w.g.) below cell pressure. The leachers are maintained at a reduced depression during the fuel charging operation.

Off-gases from the leacher condensers are scrubbed with alkali to remove nitrous fume and entrained fission products, and combined with the vessel ventilation extract. The joint stream is scrubbed with alkali, heated above the dew point, and passed through a bank of absolute filters prior to discharge to atmosphere through a stack at 65 m above ground level. All design schemes require a similar capacity for vessel ventilation requirements.

Cell ventilation

Table 9 details the ventilation extract requirements for the active cells in the four schemes. An extraction rate of 6 changes per hour (0.1 cell volumes per minute) has been adopted to accommodate fire or explosion without pressurising the cell and to deal with in-leakage of air and preserve a depression of 18 mm (w.g.) between the cell and the operating area. The exhaust air is discharged through the stack to atmosphere.

Designs 1 and 2 require cell extraction capacities of about 9 and 8 $m^3 s^{-1}$ respectively. This requirement is reduced to about 3.5 $m^3 s^{-1}$ in Design 3, where a substantial proportion of active equipment is maintained submerged under water. Heavy ventilation requirements of 50 $m^3 s^{-1}$ are estimated for the canyon of Design 4 taking the same criterion of 6 cell air changes per hour.

10.2.2 Gaseous effluent discharge

The ventilation systems for the various plant designs have been described in sections 4.6.1 and 10.2.1. There are separate systems for process vessels, leachers and the cells. These are combined and filtered before discharge through a 65 m stack. Under normal operating conditions negligible quantities of particulate fission products will escape the three stages of absolute filters.

TABLE 9

VENTILATION REQUIREMENTS FOR ACTIVE CELLS

Cell	Design 1		Design 2	Design 3	Design 4
	Extraction Capacity (1) (m^3 s^{-1})	Approx. dia of Duct (mm)	Extraction Capacity (m^3 s^{-1})	Extraction Capacity (m^3 s^{-1})	Extraction Capacity (m^3 s^{-1})
Leaching	2.2	18	2.2	-	
Evaporation	0.7	10	0.7	0.7	
Extraction Column	0.6	9	-	0.6	
H.A. Vessel	0.5	9	0.5	0.5	0.1 x Canyon Volume
L.A. Vessel	1.0	12	1.0	1.0	(Approx. 500 m^3)
Solvent Washing	0.9	12	0.4	0.9	
U Purification and Evaporation	0.4	9	0.4	-	
Pu Ionex and Evaporation	0.5	9	0.5	-	
Acid Recovery	1.5	16	1.5	-	
Ventilation Off-gas	0.3	6	0.3	-	
Pu Packaging	0.9	12	0.9	-	0.9
TOTAL	9.5		8.4	37	500

Notes (1) Extraction capacity = 6 changes h^{-1}

The main volatile fission products in the off gas system are krypton, xenon, iodine and tritium. Table 10 lists the specific activity of voltatile radionuclides in the reference PWR fuel (Table 1 and 2) after cooling times of 30 and 180 days.

Scrubbing of the off gas with nitric acid and caustic soda can be expected to remove at least 99.5% of the iodine (ORNL 1970b). There is strong evidence to suggest that most of the tritium either escapes from the fuel assembly during irradiation or is chemically bound with the cladding (ORNL 1971). The major part of the remaining tritium will find its way into liquid streams as HTO.

Gaseous activity discharged from the 65 m stack will be dispersed by wind currents. The variation of long-term average concentration (x,θ) with distance can be calculated for a given site location from a knowledge of wind velocity and atmospheric characteristics viz., (Turner 1970).

$$\chi(x,\theta) = \sum_{S}\sum_{N} \frac{2\,Q\,f\,(\theta,S,N)}{\sqrt{2\pi}\,\sigma(x,S)\,u_N\,(\frac{2\pi x}{n})} \exp\left[-\frac{1}{2}\left(\frac{H}{\sigma(x,S)}\right)^2\right] \qquad (1)$$

where f(θ,S,N) is the frequency that the wind is from direction θ for the stability condition S and wind speed class N,

Q is the release rate (Ci s^{-1}),

χ(x,θ) is the average concentration in direction θ at distance x from the stack (Ci m^{-3}),

σ(x,S) is the vertical dispersion parameter (m),

n is the number of sectors considered,

u_N is the average wind speed for class N (m s^{-1}),

and H is the stack height (m).

The maximum value of χ(x,θ) at the site boundary fence determines the largest airborne radiation dose that any member of the public could conceivably receive under normal operating conditions. In the USA the maximum permissible concentration at the site boundary is one-third of the ICRP or 10CFR20 regulations (USAEC 1969, ICRP 1959) which are equivalent to 170 m rem/year.

The value χ/Q is a measure of the meterorological suitability of the site. Values of χ/Q at the site boundaries of three existing plants in the USA and three other hypothetical plant locations are given in Table 11. The range of values in Table 11 reflects not only the differences in meterorological conditions from site to site but also differences in the assumptions and computational methods adopted.

The data for Lucas Heights were calculated from Equation 1 using available wind rose and stability characteristics (Charash and Bendun 1968). In applying this equation, three wind speed classes, four stability classes and sixteen

TABLE 10

VOLATILE RADIONUCLIDES IN REFERENCE PWR FUEL (ORNL 1970a)

Burn-up 33,000 MWd/t U, Rating 30 MW/t U

Isotope	Half-life	Activity in Fuel (Ci/tU)	
		30 days cooled	180 days cooled
^{3}H	12.5 yrs	700	690
^{85}Kr	9.4 yrs	11,400	11,200
^{131m}Xe	12 days	2,700	0.6
^{133}Xe	5.3 days	30,000	-
^{129}I	1.7×10^{7} yrs	0.037	.038
^{131}I	8.1 days	66,800	0.2

TABLE 11

ATMOSPHERIC DISPERSION CHARACTERISTICS FOR SELECTED SITES

Location	Distance to Site Boundary	Stack Height	Dilution Factor at Boundary ($\times 10^{8}$)	Reference
West Valley, N.Y., USA (NFS)	1.5 km	65 m	22	ORNL 1970a
Morris, Ill, USA (MFRP)	0.6-3 km	91 m	11	ORNL 1970a
Barnwell, S.C., USA (BNFP)	2 km	100 m	5.7	ORNL 1970a
Oak Ridge, Tenn, USA	2 km	100 m	56	ORNL 1970a
United Kingdom (typical)	2 km	120 m	9.3	Bryant 1964
Lucas Heights, Aust.	2 km	65 m	9.7	Charash and Bendun 1968

sectors were considered. Figure 75 shows the variation of χ/Q with distance in the critical SSE direction from a hypothetical 65 m stack. The maximum concentration is reached about 500 m from the stack.

In the absence of specific information from possible plant locations, the Lucas Heights data have been used to give an indication of the typical radiation dose at the boundaries of a 3 t U/d reprocessing plant (see Table 12). In calculating the effects of routine stack releases it has been assumed that 0.5% of the iodine, 25% of the tritium and 100% of the noble gases are discharged from the stack. For a site boundary fence 2 km from the plant, stack discharge rates fall within current international limits.

It should be noted that the maximum permissible concentration (MPC) for krypton in air as shown in Table 12 is now regarded as too conservative. Hendrickson (1970) has pointed out that, because ^{85}Kr is predominantly a beta emitter, the ICRP maximum concentration of 3×10^{-7} Ci m^{-3} which supposedly corresponds to a whole body dose of 500 millirems/year, in reality, corresponds to a dose of 300 millirems to the skin and only 7 millirems to the whole body.

TABLE 12

ESTIMATED STACK EFFLUENTS FROM A 3 t U/d PLANT

Cooling time - 180 days, Distance to site boundary - 2 km

Isotope	Release Rate (μCi s^{-1})	Concentration at Site Boundary (Ci m^{-3})	MPC (Ci m^{-3}) 1/3 (10 CFR 20 limits)	% MPC
^{3}H	6.0×10^{3}	5.8×10^{-10}	7×10^{-8}	0.8
^{85}Kr	3.9×10^{5}	3.8×10^{-8}	1×10^{-7}	38.0
^{131m}Xe	21	2.0×10^{-12}	1.3×10^{-7}	-
^{129}I	6.5×10^{-3}	6.3×10^{-16}	3×10^{-15} *	21.0
^{131}I	3.5×10^{-2}	3.4×10^{-15}	1.4×10^{-13} *	2.4

* The values for ^{129}I and ^{131}I have been reduced by factors of 7000 and 700 respectively to take account of the dose to the child thyroid arising from the grass-cow-milk pathway. These factors can be ignored if no milk is produced within a radius of 50 km.

10.2.3 Disposal of liquid and solid wastes

Liquid wastes (Section 4.6.2)will be concentrated and stored pending an ultimate fixation process. Trace active wastes of large volume will either be recycled to the process, or discharged after the level of active species has been decontaminated below the maximum permissible levels recommended by the ICRP.

Solid waste arisings (Section 4.6.3) will be buried in sealed containers or stored with highly active liquid wastes.

10.2.4 Failure of containment systems

Structures, systems and equipment are conventionally grouped into three classes:

Class 1 comprises areas where failure could result in an uncontrolled release of radioactivity or where functioning is required to effect and maintain a safe plant shutdown. Class 1 areas would therefore include all process areas containing plutonium and/or $\beta\gamma$ activity, the ventilation systems and the plant control stations. These systems are normally required to withstand conditions of loading through normal operation, accident condition, and through natural phenomena (seismic and meteorological) without loss of function.

Class 2 comprises areas where failure would not result in an uncontrolled release of radioactivity, and where functioning is not necessary for effecting and maintaining a safe shutdown. These could include the areas of low activity (such as uranium evaporation); these areas are required to withstand a design basis earthquake and a 100 mph wind load.

Class 3 refers to areas not essential for safe shutdown or maintenance of the plant.

Activity is primarily contained in process vessels and pipework. Subsequent shells of containment are provided by the shielded areas housing process equipment, and by the envelope of the reprocessing building.

Activity discharged would be received into the subsequent containment areas. The likelihood of breaching primary containment through vessel corrosion is minimised by high standards of construction and inspection. Activity egress through corrosion would be retained by subsequent containment shells. Breaching of primary and secondary containment may occur through

(a) Fires
(b) Explosions
(c) Nuclear excursions
(d) Natural phenomena (earthquake and tornadoes)
(e) External missiles.

Fires and explosions

These are possible in three areas: in the fuel leachers through accumulation of Zircaloy fines and subsequent reaction with nitric acid, in the solvent extraction system through ignition of solvent, and in the evaporation areas by reaction of entrained or degraded solvent with nitric acid or nitrates.

The leachers are provided with ejectors for removing gross accumulation of Zircaloy fines as settled sludge on the vessel floor. The Zircaloy is routed into active waste storage. Fires in the solvent extraction systems are prevented by maintaining the operating temperature in equipment below the flash point of the solvent diluent, typically about 70-80°C. Fire or explosion in evaporators through reaction between solvent or its degradation products and nitric acid or nitrate salts is avoided by maintaining temperatures below 135°C through limitation of the steam pressure applied to the evaporator jacket.

Nuclear excursions

Pressurisation or loss of containment through a nuclear excursion is primarily prevented by avoidance of criticality through design or administrative control, and has been dealt with in detail in Section 10.1.

Natural phenomena

It has been assumed that the plant would be constructed in a zone of low seismic activity, for which a 'design basis' earth tremor would be determined from past records and that plant structures would be examined to ensure stability.

The conceptual site for the reprocessing plant has been assumed to lie in a region of comparative calm, with extreme conditions not exceeding a maximum 70-90 m s^{-1}. Structures would require examination for adequacy against wind loadings of this magnitude.

External missiles

These fall into three types:

(a) Airborne missiles generated by a freak storm

A typical missile assumed in the safety analysis for the NFS. Reprocessing Plant Nuclear Fuels Services Inc., USAEC Docket Series 50201-42 was a wooden plank of dimensions 100 mm x 300 mm x 3 m travelling at 134 m s^{-1}. In view of the lower wind speeds cited for the conceptual location, a maximum of 90 m s^{-1} is proposed for the same missile. Detailed calculations would be necessary for all schemes to verify that this unlikely case would not be unsafe; in particular, the control station areas for all schemes could require added protection.

(b) Aircraft

It is assumed that there will be an air corridor above the plant site prohibited to aircraft, and that the plant would be located clear of existing flight tracking paths for missiles under test.

(c) Land vehicles

The relevant safety document for the NFS plant assumes impact at the plant external wall of a motor vehicle of mass 1360 kg travelling at 22 m s^{-1}. It has been assumed that roads around the process building could be constructed to eliminate this possibility.

10.3 Radiation Shielding

The thickness of concrete shielding for cell walls has been derived by inspection of data on the NFS and Allied Chemical Reprocessing plants. The thicknesses are:

Cell Function	Ordinary Concrete (m)
Shear leach	1.8
Evaporator	1.8
High level liquor tankage	1.8
Column wall	0.9 (High density)
Low level liquor tankage	0.9
Solvent wash and off-gas treatment	0.9
Plutonium ion exchange and evaporation	0.9
Uranium evaporation	0.3
Acid recovery	0.6
Storage pond	0.6

The column wall has been allocated 0.9 m of high density concrete (equivalent to 1.5 m of ordinary concrete). The plutonium ion exchange and evaporation area is at a relatively low level of γ radiation but has been allocated 0.9 m of concrete shielding from criticality considerations.

Thicknesses of alternative shielding materials have roughly been assessed on a density basis, using the above values:

Densities of Shielding Materials	Density kg m^{-3}
Water	1,000
Concrete (ordinary)	2,360
Concrete (high density)	3,770
Steel	7,850
Lead	11,400

10.4 Comparative Safety and Environmental Aspects of the Designs

10.4.1 Containment

All four plant designs operate on an identical flowsheet using the same design of equipment for treatment and disposal of gaseous, liquid and solid wastes; the activity contribution to the plant environment in normal operation is thus identical for all schemes. Isotopes which are discharged to the biosphere (tritium and krypton-85 in gaseous discharges, and trace fission products in liquid waste) are diluted to concentrations below those recommended by the ICRP; all other activity arisings are stored. Normal operation of any of the designs thus conforms to accepted reprocessing practice.

Leakage of activity from process vessels may be due either to conditions inside the vessels (vessel corrosion, fires, explosions, nuclear excursions) or to external causes (earthquakes, tornadoes, external missiles). The degree of hazard to the plant personnel and the public at large which could result from egress of activity from active equipment is dependent on the integrity of subsequent containment shells.

Designs 1 and 2 embody the greatest degree of conservatism in design for containment. Leakage from the process vessels would be received in the cell surrounding the vessel. Detailed calculations of pressure transients would be required to demonstrate the adequacy of cell containment against rupture in the event of a fire, explosion or criticality incident. Calculations would be required to ensure that all four designs were capable of withstanding seismological and meteorological stresses.

Designs 1 and 2 further enclose the active cells in the building envelope; in the event of structural failure of the cells, a leakage of gaseous activity into the operating area would be discharged through the ventilation stack, 60 m above ground level, and would be subject to considerable dilution on entry into the biosphere.

Failure of plant under water in Design 3 could result in both ingress of water into process vessels and escape of activity into the surrounding water. In the case of an explosion, activity could be discharged into the operating area, and radiation levels could be increased due to presence of activity in the water shield. This scheme offers a lower containment standard for plant personnel than Designs 1 and 2.

Activity arising as a result would be confined within the process building envelope, and gaseous activity would be discharged through the 60 m stack as in Designs 1 and 2.

Design 4 is weakest from the standpoint of containment in the event of an accident; there is no process building enclosing the canyon, and containment safety would depend on the integrity of the canyon construction.

10.4.2 Radiation shielding

All four designs contain thicknesses of shielding materials commensurate with personnel protection requirements in normal operation in the NFS and Allied-Gulf Nuclear Services reprocessing plants.

Maloperation of plant is unlikely to result in an appreciable increase in radiation level to the plant operators in any of the designs; the operator is shielded from all equipment processing plutonium by about 1 m of concrete (or equivalent) to cater for increased radiation arising in the event of maloperation resulting in a nuclear excursion.

Plant maintenance

Removal and replacement of defect equipment in Designs 1 and 2 involves removal of roof slabs from active cells, and transfer of partially decontaminated active equipment from the cell using lifting and manipulating devices external to the cell. These operations are envisaged as being controlled by maintenance staff protected by temporary radiation shields. There is inevitably a greater probability of increased radiation dosages to maintenance personnel during these non-routine operations.

Designs 1 and 2 economise in shielded space by direct feeding of a fuel assembly from a shielded transfer flask into the shear pack. This system has the inherent possibility of a shear pack failure while the fuel assembly is still partly contained within the shielded flask. In this event it would be necessary to withdraw the partially sheared assembly (contaminated with exposed fission products) into the flask prior to commencement of maintenance operations. Increased radiation dosages to personnel could result from spread of contamination, and a detailed study of maintenance operations would be required to ensure a satisfactory plant design.

In contrast, failed equipment in Design 3 is removed under a water shield, and superficially, the design appears more satisfactory from a radiation shielding aspect than Designs 1 and 2. However, the effect of activity leakage into the water at time of failure could result in absorption of activity on the pond walls at water surface level, and give rise to an increase in background radiation levels which could be difficult to remove by decontamination.

Failed equipment in Design 4 is removed by the shielded mobile cabin, which is designed for the maximum radiation flux existing in normal or maintenance operations. The operator(s) would be shielded to the same extent as in conventional plant. However, sustained working conditions inside the confined space of the cabin could be unacceptable to workers' unions. Design 4 also requires exclusion zones for aircraft and ground vehicles; the zones could be several kilometres in radius to reduce radiation to non-occupational levels.

10.4.3 General comparison

Under normal operating conditions, all designs have a similar degree of containment and biological shielding for operators, comparable with conventional reprocessing plants.

Containment under fault conditions is greatest in Designs 1 and 2, which have effectively 3 shells of containment, Design 3 is subject to spread of contamination in the water shield. Gaseous activity arising under fault conditions in Designs 1, 2 and 3 would be discharged with considerable dilution through a stack 60 m above ground level. Design 4 is poorest from a containment standpoint; it has only two shells of containment, and could lead to a discharge of gaseous activity at ground level in the event of failure of the plant canyon containment.

Designs 1 and 2 are more prone to increased radiation dosages to maintenance personnel; an increased background dose rate could apply to Design 3 consequent upon leakage of activity into the pond water. Design 4 presupposes an extensive exclusion area for the public at large, and the confined working conditions for plant operators with the mobile cabin could be opposed by unions. The wisdom of placing the key operational component (the mobile cabin) in the hands of one or two isolated personnel could also be criticised.

11. ECONOMIC EVALUATION

11.1 Scope of Costing

This study determines the total capital costs for four plant design variants of an Australian reprocessing facility treating thermal oxide reactor fuels at a nominal capacity of 3 tonnes of contained uranium/day (900 tonnes/year). The conceptual plants produce uranium and plutonium products as nitrate solutions and provide for storage of liquid wastes for a period of five years after which solidification is envisaged. Not included in the total cost estimate is the cost of waste solidification and other feasible plant refinements, krypton and tritium removal and neptunium recovery. Cost estimates for each of these additions are given but these can only be approximate since insufficient plant scale experience is available at present.

The conceptual plants are self sufficient except for water and power. The cost of these services as well as access roads and railways have been determined as far as the site boundary.

11.2 Effect of Site Location

Because reprocessing plants handle large quantities of highly active materials, they should be located as far as possible from centres of population. In practice, it is necessary to compromise this ideal to minimise infrastructure and transport costs. Other factors which affect site selection include availability of power and water, access to plant via rail, road or water, meteorology, seismology, hydrology, terrain and the local availability of labour and building materials.

Possible sites for the conceptual plants have not been considered in depth but initial scoping studies indicate that Southern Central Australia could provide a satisfactory site location (L. Keher - AAEC private communication). For costing purposes the geology of that area - limestone rock overlain by a thin layer of sandy soil - has been assumed.

All costing, however, has been carried out in mid 1972 dollars with Sydney as the reference location because cost data is more readily available for that area. It is expected that location of the plant on a suitable site in Southern Central Australia would increase the capital cost of the project by about 5% (L. Keher - AAEC private communication).

11.3 Sources of Costing Data

Since the economic evaluation presented in this study is regarded as of prime importance, every effort has been made to ensure that the figures produced are as accurate and complete as possible. Costs were obtained in the following ways:

1. Crooks, Michell, Peacock and Stewart, a firm of civil engineering consultants, were commissioned to cost buildings and structures.

2. Direct quotations were obtained for special equipment from Australian and overseas suppliers and fabricators.

3. Piping and some equipment costs were determined from a detailed breakdown of material costs and fabrication man-hours.

4. Costs for various nuclear plant facilities were obtained from reported overseas costs (Bechtel Corporation 1966, Robertson and Stockdale 1955, Barendregt et al. 1964), and from prior costing studies (Division of Civilian Application 1957, Runion 1962, Breton et al. 1968, Amdel 1970, Charlton and Alfredson 1971, Farrow 1961, Schneider et al. 1961, Cairns 1967, Unger 1967).

5. Costs for conventional plant were obtained from standard Australian (Buchanan and Sinclair 1964) and overseas costing references (Aries and Newton 1955, Bauman 1964, Vilbrandt and Dryden 1959, Popper 1970, Peters and

Timmerhaus 1968, Institution of Chemical Engineers, London 1969, Perry 1963).

6. Some costing data were obtained from internal AAEC sources. As much of the cost data available was out of date, it was often necessary to update to 1972 Australian dollars. Australia does not have chemical plant and equipment indices equivalent to the American Marshall and Stevens Equipment Index or the Chemical Engineering Plant Index. It was, therefore, not possible to use these specific indices to update published Australian cost figures. In the absence of specific indices it was decided to use the implicit price level index which measures the inflation in the cost of all goods and services relative to a base year. Table 13 and Figure 76 show a comparison between the implicit price levels in the USA and Australia and the specialised American chemical engineering indices. The closeness of agreement between the American indices suggests that updating using the Australian implicit price level index is unlikely to introduce a significant error in the costing of chemical plant.

It was also necessary at times to determine the cost of equipment fabrication in Australia from overseas equipment cost data (chiefly USA). Discussion with a representative of a consulting firm specialising in chemical engineering plant (A. Rintoul - Davy Ashmore, private communication) confirmed that the best and simplest method for relating plant and equipment costs between the USA and Australia was to convert currencies at parity. The procedure followed in updating overseas costs, then, was first to convert to Australian dollars at parity and then to update using the Australian implicit price level index.

11.4 Total Capital Cost

Table 14 sets out the overall capital costs for the four reprocessing plant designs. The direct construction cost was determined, in each case, by adding together the individual items whose cost breakdown is given in following sections. The total capital expenditure was then obtained by adding the following items to the direct construction cost:-

1. Engineering, construction and contractor's fee assessed at 27% of direct costs as recommended by Peters (1958). This item accounts for all costs associated with design including engineering and construction drawings and all construction overheads such as temporary buildings, field supervision, purchasing and administration.
2. Rolling stock estimated at $600,000 includes the cost of fire trucks, locomotive, rail-cars, mobile cranes, tractors, trucks and sundry other vehicles.

TABLE 13

COMPARISON OF COST INDICES

	Marshall & Stevens (1,2)	Chemical Engineering (1,2)	Implicit Price Level, USA (3,4)	Implicit Price Level Aust. (5)
BASE YEAR	1926	1958	1963	1965
1960	237.7	101.7	96.4	91.2
1961	237.2	100.2	97.6	94.0
1962	238.5	100.6	98.7	95.1
1963	239.2	100.5	100.0	96.3
1964	241.8	101.2	101.5	97.2
1965	244.9	102.1	103.5	100.0
1966	252.5	105.3	106.3	103.1
1967	262.9	107.7	109.7	105.9
1968	273.1	111.5	114.1	109.1
1969	285.0	119.0	119.5	113.3
1970	303.3	125.7	126.2	118.7
1971	321.3	132.3	132.0	125.6
1972 (est)	330.1	136.5	134.7	134.3
ANNUAL INCREASE				
12 year average 1960-1972	2.8%/a	2.5%/a	2.8%/a	3.3%/a
6 year average 1960-1966	1.0%/a	0.6%/a	1.7%/a	2.1%/a
6 year average 1966-1972	4.4%/a	4.0%/a	4.5%/a	4.5%/a

1. Norden, R.B. (1970)
2. Chemical Engineering (1972)
3. OECD (1970)
4. OECD (1972)
5. Richmond, M. (1972) - Private Communication

Timmerhaus 1968, Institution of Chemical Engineers, London 1969, Perry 1963).

6. Some costing data were obtained from internal AAEC sources. As much of the cost data available was out of date, it was often necessary to update to 1972 Australian dollars. Australia does not have chemical plant and equipment indices equivalent to the American Marshall and Stevens Equipment Index or the Chemical Engineering Plant Index. It was, therefore, not possible to use these specific indices to update published Australian cost figures. In the absence of specific indices it was decided to use the implicit price level index which measures the inflation in the cost of all goods and services relative to a base year. Table 13 and Figure 76 show a comparison between the implicit price levels in the USA and Australia and the specialised American chemical engineering indices. The closeness of agreement between the American indices suggests that updating using the Australian implicit price level index is unlikely to introduce a significant error in the costing of chemical plant.

It was also necessary at times to determine the cost of equipment fabrication in Australia from overseas equipment cost data (chiefly USA). Discussion with a representative of a consulting firm specialising in chemical engineering plant (A. Rintoul - Davy Ashmore, private communication) confirmed that the best and simplest method for relating plant and equipment costs between the USA and Australia was to convert currencies at parity. The procedure followed in updating overseas costs, then, was first to convert to Australian dollars at parity and then to update using the Australian implicit price level index.

11.4 Total Capital Cost

Table 14 sets out the overall capital costs for the four reprocessing plant designs. The direct construction cost was determined, in each case, by adding together the individual items whose cost breakdown is given in following sections. The total capital expenditure was then obtained by adding the following items to the direct construction cost:-

1. Engineering, construction and contractor's fee assessed at 27% of direct costs as recommended by Peters (1958). This item accounts for all costs associated with design including engineering and construction drawings and all construction overheads such as temporary buildings, field supervision, purchasing and administration.
2. Rolling stock estimated at $600,000 includes the cost of fire trucks, locomotive, rail-cars, mobile cranes, tractors, trucks and sundry other vehicles.

TABLE 13

COMPARISON OF COST INDICES

	Marshall & Stevens (1,2)	Chemical Engineering (1,2)	Implicit Price Level, USA (3,4)	Implicit Price Level Aust. (5)
BASE YEAR	1926	1958	1963	1965
1960	237.7	101.7	96.4	91.2
1961	237.2	100.2	97.6	94.0
1962	238.5	100.6	98.7	95.1
1963	239.2	100.5	100.0	96.3
1964	241.8	101.2	101.5	97.2
1965	244.9	102.1	103.5	100.0
1966	252.5	105.3	106.3	103.1
1967	262.9	107.7	109.7	105.9
1968	273.1	111.5	114.1	109.1
1969	285.0	119.0	119.5	113.3
1970	303.3	125.7	126.2	118.7
1971	321.3	132.3	132.0	125.6
1972 (est)	330.1	136.5	134.7	134.3
ANNUAL INCREASE				
12 year average 1960-1972	2.8%/a	2.5%/a	2.8%/a	3.3%/a
6 year average 1960-1966	1.0%/a	0.6%/a	1.7%/a	2.1%/a
6 year average 1966-1972	4.4%/a	4.0%/a	4.5%/a	4.5%/a

1. Norden, R.B. (1970)
2. Chemical Engineering (1972)
3. OECD (1970)
4. OECD (1972)
5. Richmond, M. (1972) - Private Communication

TABLE 14

ESTIMATED TOTAL CAPITAL COSTS FOR FUEL REPROCESSING PLANTS (1972 $A)

	Design 1 ($A)	Design 2 ($A)	Design 3 ($A)	Design 4 ($A)
Site Improvements	835,000	835,000	835,000	835,000
Process Building	4,932,000	4,881,000	4,380,700	5,313,500
Process Equipment	4,214,200	4,893,500	3,709,900	4,561,500
Process Piping	1,177,200	1,054,800	1,168,100	1,922,400
Process Instrumentation and Control	869,500	830,400	735,700	878,900
Sampling and Analytical	982,000	982,000	982,000	527,600
Utilities	1,465,000	1,465,000	1,465,000	1,465,000
Services Reticulation	727,000	727,000	727,000	727,000
Auxiliary Buildings	1,079,000	1,079,000	1,079,000	1,079,000
Liquid Waste Storage	3,328,000	3,328,000	3,328,000	3,328,000
Direct Construction Cost	19,608,900	20,075,700	18,410,400	20,637,900
Engineering Construction and Contractor's Fee (27% of direct construction cost)	5,294,400	5,420,400	4,970,800	5,572,200
Rolling Stock	600,000	600,000	600,000	600,000
Site Investigation and Survey	300,000	300,000	300,000	300,000
Estimated Plant Cost	25,803,300	26,396,100	24,281,200	27,110,100
Contingency (15%)	3,870,500	3,959,400	3,842,200	4,066,500
Interest on Construction Funds (average 7.80%)	2,314,600	2,367,700	2,178,000	2,431,800
Working Capital	1,575,000	1,575,000	1,575,000	1.575,000
Preoperational Costs (9 months)	3,290,600	3,364,100	3,109,000	3,446,400
Estimated Total Capital Expenditure Sydney-Basis (rounded)	36,800,000	37,700,000	34,800,000	38,600,000
Remote Site Allowance (5%)	1,800,000	1,800,000	1,700,000	1,900,000
Estimated Total Capital Expenditure Remote-Site (rounded)	38,600,000	39,500,000	36,500,000	40,500,000

3. Site investigation and survey. $300,000 has been allocated for a detailed study of the geology, hydrology, seismology and meteorology of the site.

4. A contingency allowance taken at 15% of the estimated plant cost.

5. Interest on construction funds calculated by assuming a three year construction period with 20% of the investment committed in the first year, 30% in the second and 50% in the third. Interest has been charged at 8.75% on debt capital only. For the purposes of this report, the debt/equity capital ratio has been taken as unity.

6. Working capital taken from the NFS figures (Nuclear Industry 1970) and updated.

7. Preoperational costs including operating costs and interest on capital during a commissioning period of nine months.

8. A remote site allowance excluding infrastructure costs of 5%.

The total capital expenditure for the four plants range from $36.5 million for design 3 to $40.5 million for design 4.

Site improvements

It is assumed that the conceptual plants are independent facilities located on a 16 km^2 site which is relatively flat and sparsely wooded. Spent fuel can be received by either rail or road. Only a small fraction of the total site, some 140,000 m^2, is required for the actual reprocessing complex, the remainder is an undeveloped but controlled exclusion area.

Table 15 shows a breakdown of the estimates for site improvements. Costs have been based on the site layout for design 1 (Figure 44). While minor alterations to the site plan would be required for the other designs, these would not significantly affect the overall plant area or the cost of site improvements.

Buildings and structures

The proposed building designs were submitted to a firm of civil engineering consultants, Crooks, Michell, Peacock and Stewart Pty. Ltd, for examination of structural feasibility, for design of crane supporting buildings and for costing. The consultants confirmed that the designs were structurally adequate, practicable and economic.

In designs 1 and 2, the relative merits of above-ground versus below-ground storage pond location were studied. The consultants breakdown of construction costs for those items affected by pond elevation is shown in Table 16. Above ground construction leads to lower excavation costs, but this is balanced by the higher cost of the pond building which needs to be both taller and wider. Overall, underground location is marginally cheaper. More

TABLE 15

COST OF SITE IMPROVEMENTS

Item	Requirement	Unit Cost $A	Total Cost $A
Roads - plant roads	2 km of 9 m width, heavy duty	10 per m^2	180,000
access road	2.1 km of 6.7 m width, heavy duty	10 per m^2	141,000
Railroads	3.5 km	55 per m	193,000
Kerb and Guttering	1.4 km	11 per m	15,000
Footpaths	1.5 km of 0.5 m width	5 per m^2	4,000
Site Clearing and Levelling	230,000 m^2	0.40 per m^2	92,000
Landscaping	140,000 m^2	0.20 per m^2	28,000
Parking Areas	4,000 m^2	5 per m^2	20,000
Security Fences	18 km	9 per m	162,000
TOTAL SITE IMPROVEMENTS			835,000

TABLE 16

BREAKDOWN OF COSTS FOR THOSE ITEMS IN DESIGNS 1 AND 2 AFFECTED BY STORAGE POND ELEVATION

	Design 1		Design 2	
	Above Ground ($A)	Below Ground ($A)	Above Ground ($A)	Below Ground ($A)
Foundation, Excavation and Backfill	84,900	128,400	83,700	127,200
Reinforced Concrete	694,000	691,100	694,900	692,000
Crane Supporting Buildings	361,100	319,000	411,200	369,100
Cranes and Hoists	236,600	231,600	236,600	231,600
	1,376,600	1,370,100	1,426,400	1,419,900

importantly, below ground construction leads to better utilisation of storage space within the pond. While fuel elements can be stored right up to the side walls in the underground pond, a water corridor, 2 metres in thickness, must be maintained around the above ground pond for radiation attenuation. For this reason, underground pond location was adopted in the four designs and all further costs are reported for this case.

The consultants recommended a steel framed building with metal cladding of roofs and walls as the most economical form of construction for the crane supporting building in designs 1, 2 and 3. No provision was made by the consultants for the cost of services in the crane supporting building. The requirement for ancillary areas such as laboratories, offices and control rooms was taken largely from the NFS plant and was costed on a unit area basis to include all services except air conditioning. Unit area rates for active laboratories were obtained from internal sources (J.D. Wilson 1972 - AAEC private communication), other rates were supplied by the consultants. Appendix A1 summarises the estimating rates used in costing the process buildings.

A breakdown of building costs on a functional area basis for designs 1, 2 and 3 is given in Table 17. The complete cost of building and structures was obtained by taking the consultant's estimate of the basic shell cost and adding to this the cost of services and equipment as necessary. Provision was made for air conditioning in ancillary areas such as laboratories, offices and cold chemical rooms etc., but it was found to be impractical and uneconomic to condition the air in the large crane supporting buildings. Costs for the ventilation system were based on six changes per hour in active cells and twelve changes per hour in working areas. Allowance has been made in the estimates of designs 1 and 2 for a closed circuit television system to be used for maintenance operations in partially decontaminated cells.

It was not possible to partition building costs on a functional basis in design 4 because all equipment is housed in a single large canyon. Instead, costs were divided on a materials and construction item basis and these are tabulated in Table 18. It was proposed that the canyon could most economically be constructed by excavating the limestone with explosives and then lining the exposed walls with 300 mm of concrete. Due to the depth of excavation it would be necessary to provide positive anchorage of the concrete lining to the rock face by means of bolts. The proposed roof construction to withstand damage by external missiles would consist of precast, prestressed concrete T beams with an in-situ concrete topping and bituminous waterproof membrane.

TABLE 17

COST OF PROCESS BUILDINGS AND STRUCTURES - DESIGNS 1, 2 & 3

1. Fuel Receipt and Storage	Designs 1 & 2		Design 3
	Above Ground $A	Below Ground $A	$A
Civil and Structural - CMPS	389,100	387,600	239,600
Fuel Storage Racks - 1000 Elements 50 Racks at $A5,500 ea.	275,000	275,000	275,000
Pool Clean-up			
. Filtration and Pumping Circuit 120 m^3 h^{-1} at $A130 $m^{-3}h$	15,600	15,600	15,600
. Demin. Water System 25 m^3 h^{-1} at $A3,500 $m^{-3}h$	87,500	87,500	87,500
. Fuel Pond Cooler, 175 kW	22,000	22,000	22,000
Ruptured Fuel Storage Canisters - 5 off	25,000	25,000	25,000
Cranes and Hoists	236,500	236,500	207,400
Underwater Tools etc.	30,000	30,000	30,000
Pool Gates	14,000	14,000	14,000
Storage Pond Building Services at $A50 m^{-2}	56,000	56,000	42,000
Sub Total	1,150,700	1,148,700	958,100
2. Shear and Leach	Design 1 $A	Design 2 $A	Design 3 $A
Fuel Disassembly and Shearing Pool - CMPS			119,600
Leaching Pool - CMPS			53,700
Shear and Leach Cell - CMPS	344,100	348,900	
Cell Windows,			
. 4 off (1.5 m x 1.2 m x 1.8 m) at $A20,000 m^{-3}	259,200	259,200	
. 1 off (0.9 m x 0.75 m x 1.8 m) at $A20,000 m^{-3}	24,300	24,300	
Special Lighting	10,000	10,000	20,000
Fixtures and Services at $A100 m^{-2}	7,700	7,700	11,800
Zircaloy Flask Removal Cell - CMPS	88,100	88,100	88,100
Sub Total	733,400	758,000	313,200

3. Waste Evaporation	Design 1 \$A	Design 2 \$A	Design 3 \$A
Civil and Structural - CMPS	55,600	56,400	68,600
Cell Fixtures and Services $\$A70\ m^{-2}$	3,300	3,300	3,500
Sub Total	58,900	59,700	72,100

4. Solvent Extraction and Off-gas Treatment	Design 1 \$A	Design 2 \$A	Design 3 \$A
High Activity Liquor Cell - CMPS	28,600	35,700	55,500
Low Activity Liquor Cell - CMPS	36,200	44,700	47,100
Solvent Wash and Off-gas Treatment Cell - CMPS	40,300	40,400	63,100
Column Well - CMPS	81,700		214,100
Fixtures & Services at $\$A70\ m^{-2}$	16,200	15,500	16,500
Special Lighting			15,000
Sub Total	203,000	136,300	411,300

5. Uranium Evaporation and Purification	Design 1 \$A	Design 2 \$A	Design 3 \$A
Civil and Structural - CMPS	22,500	26,400	29,100
Fixtures and Services at $\$A120\ m^{-2}$	3,600	3,500	3,100
Sub Total	26,100	29,900	32,200

6. Plutonium Purification, Evaporation and Packaging	Design 1 \$A	Design 2 \$A	Design 3 \$A
Plutonium Purification and Evaporation Cell - CMPS	35,300	31,400	44,000
Cell Fixtures and Services at $\$A120\ m^{-2}$	3,600	3,400	3,100
Plutonium Packaging Area - CMPS	31,800	33,200	31,200
Plutonium Packaging Equipment at $\$A180\ m^{-2}$	113,000	113,000	113,000
Sub Total	183,700	181,000	191,300

7. Acid Recovery and Effluent Treatment	Design 1 \$A	Design 2 \$A	Design 3 \$A
Civil and Structural - CMPS	36,100	42,000	26,200
Cell Fixtures and Services at $\$A70\ m^{-2}$	7,400	7,700	7,900
Sub Total	43,500	49,700	34,100

8. Hot Analytical Cells	Design 1 \$A	Design 2 \$A	Design 3 \$A
Civil and Structural - CMPS	18,300	20,000	20,000
Cell Linings at \$A80 m^{-2}	15,000	15,000	15,000
Services to Cells	60,000	60,000	60,000
Manipulators	30,000	30,000	30,000
Viewing Windows (6 off)	132,000	132,000	132,000
Sub Total	255,300	257,000	257,000

9. Other Process Buildings and Structures	Design 1 \$A	Design 2 \$A	Design 3 \$A
Crane-Supporting Building over Cells - CMPS	157,100	164,200	82,300
Operating and Pump Aisles CMPS	28,700	39,500	64,400
Stairs, Lift Wells etc. CMPS	19,800	19,800	25,600
Lift - CMPS	32,000	32,000	32,000
Fixtures and Services in			
Pump Aisles and Operating Areas at \$A40 m^{-2}	88,500	75,800	73,000
Ejector Bulges at \$A500 per tonne	120,000	120,000	120,000
Sub Total	445,500	451,300	397,300

10. Laboratories, Offices and Ancillary Areas in Process Building	Design 1 \$A	Design 2 \$A	Design 3 \$A
Control Room at \$A320 m^{-2}	54,600	54,600	54,600
Laboratories at \$A530 m^{-2}	172,500	172,500	172,500
Offices at \$A240 m^{-2}	24,200	24,200	24,200
Toilets at \$A480 m^{-2}	18,000	18,000	18,000
Aisles at \$A210 m^{-2}	32,000	32,000	32,000
Cold Chemicals Room at \$A210 m^{-2}	141,700	136,600	153,900
Sub Total	443,000	437,900	455,200

11. Ventilation Systems	Design 1 \$A	Design 2 \$A	Design 3 \$A
Ventilation Supply Room at \$A190 m^{-2}	8,600	8,600	8,600
Off Gas Filter Cell - CMPS	44,000	44,000	44,000
Off-gas Filters	20,000	20,000	12,000
Cell Ventilation System at \$A20,000 m^{-3} s	182,000	168,700	90,200
Air Conditioning System at \$A50 m^{-2}	122,300	121,400	127,600
Building Ventilation at \$A1,800 m^{-3} s	380,000	375,000	260,000
Ventilation Exhaust Room at \$A190 m^{-2}	21,200	21,200	21,200
Stack - Stainless Steel at \$A5,000 per tonne Erected	107,000	107,000	107,000
Remote Handling Flask	100,000	100,000	100,000
Sub Total	985,100	965,900	770,600

12. Laboratories, Offices and Ancillary Areas in Plant Office (adjoining Process Building)	Design 1 \$A	Design 2 \$A	Design 3 \$A
Offices at \$A240 m^{-2}	59,400	59,400	59,400
Laboratories at \$A530 m^{-2}	44,400	44,400	44,400
Change Rooms at \$A300 m^{-2}	61,600	61,600	61,600
Toilets at \$A480 m^{-2}	13,500	13,500	13,500
Canteen at \$A280 m^{-2}	10,400	10,400	10,400
Aisles and Halls at \$A210 m^{-2}	44,000	44,000	44,000
Sub Total	233,300	233,300	233,300

13. Maintenance Systems	Design 1 \$A	Design 2 \$A	Design 3 \$A
Failed Equipment Maintenance Pond - CMPS	20,800	20,800	45,100
Closed Circuit Television System	69,000	69,000	
Temporary Shielding	61,500	61,500	7,400
Special Tools	20,000	20,000	20,000
Sub Total	171,300	171,300	72,500

14. Process Pool Systems	Design 3 $A
Pool Clean-up	
. Filtration and Pumping Circuit 150 m^3h^{-1} at $A130 $m^{-3}h$	19,500
. Demin. Water System 30 m^3h^{-1} at $A3,500 $m^{-3}h$	105,000
Underwater Tools	20,000
Pool Gates	38,000
Sub Total	182,500

	Design 1 Below Ground $A	Design 2 Below Ground $A	Design 3 $A
TOTAL BUILDING AND STRUCTURES	4,932,000	4,881,000	4,380,700

TABLE 18

COST OF PROCESS BUILDINGS AND STRUCTURES - DESIGN 4

Buildings and Structures	\$A
Foundation, Excavation and Backfill - CMPS	465,600
Reinforced Concrete - CMPS	526,400
Stainless Steel Linings - CMPS	196,800
Crane Supporting Buildings - CMPS	167,800
Crane and Hoists - CMPS	168,600
Acid Resisting Paint - CMPS	10,400
Bituminous Roofing - CMPS	17,900
Excavation for Roads and Ancillary Areas - CMPS	116,400
Mobile Operational Cabin	750,000
Track and Turntable for Mobile Cabin - CMPS	90,000
Fuel Receipt and Storage Equipment (as per designs 1 & 2)	525,100
Laboratories, Offices and Ancillary Areas (as per designs 1 & 2)	676,300
Ventilation Systems	
. Ventilation supply room at \$A190 m^{-2}	17,600
. Off-gas filters	20,000
. Canyon ventilation system \$A10,000 $m^{-3}s$	574,000
. Air conditioning system at \$A50 m^{-2}	117,300
. Ventilation exhaust room at \$A190 m^{-2}	53,000
. Stack - stainless steel at \$A5,000 per tonne erected	107,000
Plutonium Packaging Area	113,000
Plutonium Packaging Equipment and Services at \$A 180 m^{-2}	21,400
Canyon Fixtures and Services	158,300
Special Canyon Lighting	100,000
Hot Analytical Cells	257,000
TOTAL PROCESS BUILDINGS & STRUCTURES	5,313,500

The cost for the mobile operational cabin in design 4 has been estimated at $750,000. This figure must be regarded as only tentative because development costs for such a specialised vehicle are likely to be only high. Lloyd (1964) estimated the cost of a similar cabin employing lead shielding as $345,000.

The total building estimates for the four designs excluding overheads ranged from approximately $4.5 million for design 3 to $5.3 million for design 4. The relatively lower cost of design 3 was due primarily to,

(1) savings in the cost of crane supporting buildings made possible because storage pond and processing areas are adjacent,
(2) elimination of heavy shielding in the shear and leach area,
(3) elimination of concrete shielding roofs in most areas.

Active ventilation requirements in design 3 are below those for the other designs but this is offset by the extra water treatment facilities for the processing ponds.

The pulsed column and centrifugal contactor schemes (designs 1 and 2 respectively) have similar costs but, because of the extra pulsed column well in design 1, its costs are about $50,000 higher. Design 4 is the most expensive scheme principally because of higher excavation costs and the added cost of the mobile cabin. Furthermore, the consultants reported "because design 4 involves a much larger quantity of excavation, approximately two and half times design 3 and seven times designs 1 and 2, it is much more susceptible than the others to an increase in cost should poor rock conditions occur. Design 4 would be considerably more expensive if constructed in a sandy soil even in best conditions of a dense sand with ground water below excavated depth".

Process equipment

Appendix A2 lists the 248 chemical process equipment items in the four plant designs, their dimensions, location and estimated cost. Most of the equipment is common to all designs. Except where indicated, all process vessels are fabricated in AISI 3041 stainless steel. A higher grade of stainless, AISI 309SCb, which is more resistant to boiling nitric acid solutions is used for the leachers, leacher off-gas condensers and waste evaporators and condensers.

The procedure adopted in pricing process equipment was to cost a number of representative vessels in detail to establish a relation between equipment size and cost which could then be applied to other vessels of the same basic type. The "materials plus labour" method outlined in Buchanan and Sinclair

(1964) was used for the detailed costing calculations. Fabrication man-hours were estimated from the data of Page and Nation (1958) and Buchanan and Sinclair (1964). Labour was costed at $6 per hour to include all overheads, but not profit, which was calculated separately at 15 per cent.

To confirm the general method of costing, quotations were obtained from CSR Engineering Workshop, Sydney for a representative cross-section of process equipment. Drawings of these vessels are shown in Figures 77-82 and a fabrication specification is given in Appendix A3. Table 19 shows a comparison between the detailed 'materials plus labour' method and the CSR quotations. The updated Nuclear Fuel Services data where obtainable (Bechtel Corporation 1966) are also listed. The agreement between these three sources was sufficiently close to give confidence in the general method of costing.

Figure 83 shows the ex-works cost of tanks fabricated to radio-chemical standards obtained by the detailed 'materials plus labour' method. Two curves are given depending on whether the tanks are fitted with or without internal coils for heat transfer. Figure 84 shows a similar curve for heat exchange (reboilers and condensers) plotted as a function of external heat transfer area.

Figures 83 and 84 were used to obtain cost estimates for the majority of the process vessels. Where applicable, allowance was made in the costing for special features such as external jackets or internal neutron absorbers for criticality control. Cold chemical storage vessels were costed using Figure 85 which is based on the updated figures of Peter and Timmerhaus (1968). The costs of special equipment such as pulsed columns and slab tanks were obtained by updating the NFS figures. Robatel SLP1, Genas, France, supplied quotations for their centrifugal contactors.

Compared with conventional chemical plant, process vessels designed for reprocessing service were found to be significantly more expensive to fabricate. For example, the reported costs (Institution of Chemical Engineers 1969) of conventional equipment fabricated to pressure vessel standards in 18/8 stainless steel are $3,500-$6,500 (1972 $A) depending on size and complexity while the corresponding figures for vessels costed in this study are $5,000-$18,000. Similarly, as shown in Figure 84, heat exchangers for reprocessing plants are 25% more expensive than conventional exchangers fabricated to TEMA standards. The additional cost of process equipment designed for radiochemical duties arises from the higher fabrication standards demanded and the more extensive testing and inspection procedures.

A list of process mechanical equipment used in the head-end breakdown

TABLE 19

COMPARISON OF CHEMICAL PROCESS EQUIPMENT COST ESTIMATES

Item	Drawing No.	Figure No.	NFS Plant Updated to 1972 $A (Bechtel 1966) ($A)	CSR Workshops ($A)	'Materials & Labour' Method ($A)
Pulsed Column - without plates	NDD 5015	80	-	11,600	11,700
- with plates			22,400	30,500	23,600
Nitric Acid Fractionator	NDD 5020	81	23,700	22,000	22,400
Partition Cycle Feed Tank	NDD 5021	79	-	21,000	20,300
Condensate Catch Tank	NDD 5022	77	-	14,400	11,200
Plutonium Ion Exchanger	NDD 5023	82	3,900	3,200	3,500
Decanter	NDD 5024	78	-	3,300	2,200

and zircaloy disposal is given in Appendix A4. Cost estimates were obtained wherever possible by updating figures from NFS (Bechtel 1966) or other existing facilities. The cost of some items which were of original design were estimated on a weight basis or from experience.

A breakdown of installed process equipment costs for the four designs is presented in Table 20. The installation costs of process chemical equipment was taken at 21% of the ex-works cost in agreement with the NFS average.

11.4.4 Process Piping

Process piping, as defined here, includes all in-cell piping directly associated with the process flowsheet and all external process piping either as far as the service mains or control room. It does not include service mains piping or piping associated with sampling and instrumentation.

To estimate the reduction in cost achieved by equipment compaction, detailed estimates of piping costs were prepared for the first uranium cycle in designs 1 and 2. Equipment layouts were used in conjunction with the engineering flow diagrams to determine piping layouts and these, in turn, were used to calculate pipe lengths, number of welds, bends, fittings, wall penetrations and testing man-hours. The time required for various pipe fabrication operations was calculated from the data of Popper (1970) and Page and Nation (1958).

The results of the piping cost estimation are summarised in Figure 86 which shows unit installed costs of process piping as a function of line size. Not included in this graph are allowances for wastage, contingencies and sub-contractor profit. Also excluded is the cost of other piping components such as valves, ejectors, steam traps, jumper connections and pumps. These were costed separately from information taken from the engineering flow diagram. Appendix A5 sets out the basis for the cost estimates and gives a detailed breakdown of individual cost components. The total piping bill was determined by combining the pipe and piping component costs and then adding 30% for wastage contingencies, rewelds etc., and a further 15% for sub-contractor profit.

A comparison of process piping estimates for the 1st uranium cycle for both pulsed column and centrifugal contactor designs is given in Table 21. Because of the greater equipment compaction possible with centrifugal contactors, piping costs for the solvent extraction cycles in design 2 are 34% less than the corresponding costs for design 1.

In extending the detailed piping estimates to other parts of the plant, two approaches were considered. One which is commonly recommended for

TABLE 20

INSTALLED PROCESS EQUIPMENT COSTS (1972 $A)

Equipment	Design 1	Design 2	Design 3	Design 4
	$A	$A	$A	$A
Fuel Leaching -				
chemical equipment	110,100	110,100	116,800	110,100
mechanical equipment	1,213,000	1,213,000	702,000	881,000
Partition Cycle Solvent Extraction	218,500	469,700	218,500	469,700
1st U. Cycle Solvent Extraction	154,400	330,500	154,400	330,500
2nd U. Cycle Solvent Extraction	154,400	330,500	154,400	330,500
Uranium Evaporation and Purification	197,400	197,400	197,400	197,400
Plutonium Cycle Solvent Extraction	41,900	117,800	41,900	117,800
Plutonium Purification and Evaporation	92,700	92,700	92,700	92,700
High Level Waste Evaporation	107,700	107,700	107,700	107,700
Low Level Waste Evaporation	389,600	389,600	389,600	389,600
Rework Evaporation	167,600	167,600	167,600	167,600
Nitric Acid Recovery	405,400	405,400	405,400	405,400
Effluent Treatment	218,400	218,400	218,400	218,400
Off-gas Treatment	81,700	81,700	81,700	81,700
Cold Chemicals	248,400	248,400	248,400	248,400
Pumps	55,000	55,000	55,000	55,000
Pu Packaging Equipment	130,000	130,000	130,000	130,000
Mobile Crane, Solid Waste Disposal	95,000	95,000	95,000	95,000
Area Monitoring Equipment	133,000	133,000	133,000	133,000
Total	4,214,000	4,893,000	3,709,900	4,561,500

TABLE 21

COMPARISON OF PIPING COSTS FOR 1ST URANIUM CYCLE

Item	Design 1 Pulsed Columns	Design 2 Centrifugal Contactors
	$A	$A
Stainless Steel Piping	39,600	23,800
Mild Steel Piping	1,900	1,500
	41,500	25,300
Valves, Ejectors, Pumps etc.	15,900	12,700
	57,400	38,000
Add 30% for Wastage, Rewelds, Contingency	17,200	11,400
	74,600	49,400
Add 15% Profit	11,200	7,400
DIRECT PIPING COST	85,800	56,800

preliminary cost estimates is to prorate piping costs as a fixed percentage of process equipment costs. This method has no fundamental basis and analysis of the available data from NFS (Bechtel 1966) did not support such an approach. The method adopted was to prorate estimates from the detailed calculations for the 1st uranium cycle according to the number of lines indicated on the engineering flow diagrams. This 'line-count' method assumes that the average cost of installing a pipe run is the same. Overall, it is considered that this method is the best available for this type of plant, short of preparing detailed pipe layouts for the whole plant. As before, estimates of the cost of valves, pumps etc. were taken from the engineering flow diagram.

Process piping costs for the four designs are compared in Table 22. The piping estimates for design 4 are higher than those for the other designs for two reasons. First, because the mobile shielded cabin must be capable of removing and replacing all failed equipment from the process canyon, flanges must be provided on vessels and piping at regular intervals. It was estimated that this provision would increase piping costs by about 40% (see Appendix A5).

TABLE 22

PROCESS PIPING COSTS

Process	Design 1 $A	Design 2 $A	Design 3 $A	Design 4 $A
Fuel Leaching	69,500	69,500	60,400	141,000
Partition Cycle Solvent Extraction	142,000	94,000	142,000	190,800
1st U. Cycle Solvent Extraction	85,800	56,800	85,800	115,300
2nd U. Cycle Solvent Extraction	85,800	56,800	85,800	115,300
Uranium Evaporation and Purification	99,600	99,600	99,600	202,200
Plutonium Cycle Solvent Extraction	48,400	32,000	48,400	65,000
Plutonium Purification and Evaporation	137,500	137,500	137,500	278,100
Plutonium Product Packaging	100,000	100,000	100,000	100,000
High Level Waste Evaporation	24,600	24,600	24,600	49,900
Low Level Waste Evaporation	80,300	80,300	80,300	163,000
Rework Evaporation	43,000	43,000	43,000	87,300
Nitric Acid Recovery	94,300	94,300	94,300	191,500
Effluent Treatment	48,700	48,700	48,700	48,700
Off-gas Treatment	54,900	54,900	54,900	111,500
Cold Chemicals	62,800	62,800	62,800	62,800
TOTAL	1,177,200	1,054,800	1,168,100	1,922,400

Also, since it was necessary to separate inactive services some distance from the canyon, service pipe runs were lengthened, thereby increasing piping costs by a further 65%.

11.4.5 Instrumentation and control process

Process instrumentation and control costs were determined by extracting the requirements from the engineering flow diagrams and applying a unit rate for each control function. Appendix A6 ummarises the unit rates used for estimating purposes. These rates were derived from information supplied by Leeds and Northrup Australia Pty. Ltd. and include an allowance for control piping where necessary.

Table 23 summarises process control costs on a functional area basis. In addition allowance has been made for radiation monitoring instruments and alarms throughout the plant. A 500 point data logger and a control panel with flowsheet display have also been included in the total cost.

Of the four designs, design 3 has the lowest instrumentation costs because of savings in the head end area. Design 2 requires less instrumentation than Design 1 because, although it is necessary to double controls in some feed lines to the centrifugal contactors, this is more than balanced by the extra cost of pulsed column instrumentation. The canyon scheme (design 4) has the highest costs of all because of longer instrumentation pipe runs.

11.4.6 Analytical and sampling

Table 24 sets out the cost estimates for analytical and sampling facilities. As the cost of analytical cells has been included in the building estimates it is excluded from this breakdown. Analytical equipment costs were taken from the NFS figures (Bechtel Corporation 1966) and updated to 1972 Australian dollars while sampling and glove box stations were estimated separately.

Designs 1, 2 and 3 have similar sampling requirements and the same overall costs. Samples in design 4 are taken from the mobile operational cabin and sampling costs are proportionately lower. The cost of sampling piping was estimated from the data in Appendix A5 with allowance made for ejectors, valves, air lifts and sample return lines. This gave a total cost for each sample point of $1,800 for designs 1, 2 and 3 and $200 for design 4.

11.4.7 Utilities

Utilities provided include steam, air, water, emergency power and nitrogen. Requirements were obtained from the anticipated future demand for the NFS plant (Nuclear Fuel Services 1962b). Estimated costs exclusive of buildings are given in Table 25. Unit costs for utilities were obtained from

TABLE 23

PROCESS INSTRUMENTATION AND CONTROL COSTS (1972 $A)

	Design 1 $A	Design 2 $A	Design 3 $A	Design 4 $A
Fuel Leaching				
Chemical Equipment	19,800	19,800	10,800	24,300
Mechanical Equipment	159,800	159,800	35,000	159,800
Partition Cycle Solvent Extraction	95,000	81,000	95,000	89,000
1st U Cycle Solvent Extraction	44,000	36,000	44,000	39,500
2nd U Cycle Solvent Extraction	44,000	36,000	44,000	39,500
Uranium Evaporation and Purification	39,800	39,800	39,800	44,300
Plutonium Cycle Solvent Extraction	37,700	28,600	37,700	33,300
Plutonium Purification and Evaporation	48,100	48,100	48,100	55,200
High Level Waste Evaporation	14,300	14,300	14,300	16,200
Low Level Waste Evaporation	19,600	19,600	19,600	22,100
Rework Evaporation	7,900	7,900	7,900	8,800
Nitric Acid Recovery	42,500	42,500	42,500	47,200
Effluent Treatment	18,800	18,800	18,800	20,500
Off-gas Treatment	21,200	21,200	21,200	23,200
Cold Chemicals	37,000	37,000	37,000	37,000
Miscellaneous Monitors etc.	60,000	60,000	60,000	60,000
Control Panel	50,000	50,000	50,000	50,000
Data Logger	110,000	110,000	110,000	110,000
	869,500	830,400	735,700	878,900

TABLE 24

ANALYTICAL INSTRUMENTATION AND SAMPLING COSTS

		Designs 1, 2 & 3 \$A	Design 4 \$A
Analytical Instrumentation and Sampling		510,000	510,000
Sampling Piping (90 sample points)		162,000	17,600
Sampling stations (with stillage and sample castles)	2 at \$90,000	180,000	
	2 at \$40,000	80,000	
Sampling Glove Boxes		50,000	
		982,000	527,600

TABLE 25

UTILITY REQUIREMENTS AND COSTS

	Unit	Requirement	Unit Cost (\$A)	Total Cost (\$A)
Steam Raising	$\mathrm{kg\ h^{-1}}$	55,000	10	550,000
Compressed Air Systems	$\mathrm{m^3\ h^{-1}}$ (STP)	4,800	15	72,000
Electrical Substation	kW	2,000	40	80,000
Pumping and Filtration Station	$\mathrm{m^3\ h^{-1}}$	110	130	14,000
Cooling Towers	$\mathrm{m^3\ h^{-1}}$	570	110	63,000
Demin. Water Plant	$\mathrm{m^3\ h^{-1}}$	10	4,500	45,000
Nitrogen Supply	$\mathrm{m^3\ h^{-1}}$ (STP)	9	5,000	45,000
Standby Power Supply	kW	400	100	40,000
Water Storage Tank	$\mathrm{m^3}$	400	30	12,000
Sewers and Sewerage Disposal	$\mathrm{m^3\ day^{-1}}$	1,600	100	160,000
Lagoons	$\mathrm{m^3}$	19,200	20	384,000
			TOTAL UTILITIES	1,465,000

direct quotations from Australian suppliers of packaged plants or from the literature (Buchanan and Sinclair 1964, Bowman 1964, Peters and Timmerhaus 1968). A water storage tank of 400 m^3 capacity is included for emergency use and for fire protection. Two large lagoons similar to those built at NFS are also provided to allow for precipitation and settling of particulates and for monitoring and controlled release of effluents.

11.4.8 Services reticulation

Table 26 lists the requirements and costs for distribution of services. The estimates costs are those required to provide service mains from the site boundary or from the utilities building to the points of use. It does not, however, include the cost of distribution within a building which is allowed for in the building costs themselves. Unit costs for piping were obtained from Popper (1970)

TABLE 26

COST OF SERVICES RETICULATION

Service	Requirement	Unit Cost per m ($A)	Total Cost ($A)
Power Supply to Site	2 km of overhead cable	50	100,000
Electrical Distribution from Substation	1.2 km of underground cable	110	132,000
Water Supply and Distribution	3.2 km of 200 mm pipe	65	208,000
Steam Distribution	250 m of 350 mm pipe	200	50,000
	850 m of 125 mm pipe	60	51,000
Cooling Water Reticulation	900 m of 300 mm pipe	100	90,000
Compressed Air Reticulation	450 m of 200 mm pipe	65	29,000
	500 m of 100 mm pipe	30	15,000
Nitrogen Supply	450 m of 50 mm pipe	16	7,000
Demin. Water Reticulation	600 m of 50 mm pipe	25	15,000
Effluent Discharge	1.2 km of 250 mm pipe	25	30,000
	TOTAL SERVICES RETICULATION		727,000

11.4.9 Auxiliary buildings

Table 27 shows the cost of non-process buildings. The area requirements were taken mainly from data on the NFS plant (Nuclear Fuels Services, Inc., USAEC Docket 50201) but were increased in certain cases to accommodate the increased throughput. Unit area rates for buildings were supplied by the civil engineering consultants, Crooks, Michell, Peacock and Stewart.

TABLE 27

COST OF AUXILIARY BUILDINGS

Building	Floor Area (m^2)	Unit Cost per m^2 ($A)	Total Cost ($A)
Administration Office	1,250	240	300,000
Warehouse	1,000	185	185,000
Active Workshop	250	280	70,000
Inactive Workshop	250	280	70,000
Canteen	210	280	59,000
Utilities Building	1,000	185	185,000
Laundry	110	250	28,000
Fire House	110	240	26,000
Gate House	130	240	31,000
Change Room	150	310	46,000
Pump House	90	200	18,000
Sewage Treatment Building	130	185	24,000
Waste Tank Farm Building	200	185	37,000
TOTAL AUXILIARY BUILDINGS			1,079,000

11.4.10 Liquid waste storage facilities

The four conceptual designs allow for storage of highly active fission product wastes as alkaline liquids prior to ultimate solidification. Three mild steel tanks, 24 m in diameter and 8 m high with a safe storage capacity of 3,000 m^3 are used for this purpose with one tank acting as a spare should a leak develop in either of the other tanks. Each tank has a storage capacity equivalent to 780 days operation at the rated plant capacity of 3 t U/day. For a plant load factor of 85% the proposed storage facilities will last 5 years.

No attempt has been made to design or cost the waste storage facilities in detail, instead costs have been updated from a prior ORNL study (Blomeke et al. 1965). The Oak Ridge designs, which are based on Savannah River practice, feature triple containment. Each tank is fabricated from mild steel plate varying in thickness from 9.5 mm to 17.5 mm and is housed in a steel lined reinforced concrete vault with walls 0.8 m to 1.2 m thick. The entire unit is buried under 3 metres of earth. Monitors in the space between the tank and the steel lined vault allow any leak to be detected quickly. The tanks are equipped with cooling coils as a primary means of removing decay heat but a water cooled condenser is available in case of primary system failure.

Costs for a single waste storage tank have been estimated as follows:

	Cost (1972 $A)
Mild steel tank	81,100
Liner	38,600
Coils	235,000
Concrete	340,000
Excavation and Backfill	42,300
Appurtenances	46,500
Instrumentation	12,600
	$796,100

Table 28 sets out the capital cost estimate for the proposed waste storage facility exclusive of outside services. Costs have been scaled down and updated from figures for a much larger facility which had 31 tanks each of 3,000 m^3 capacity (Blomeke et al. 1965). Tank costs have been simply prorated according to the number of tanks but to scale other costs in the same manner would underestimate the total cost because the large tank farm offers economies of scale in piping, heat exchanger and building costs which cannot be matched in a smaller facility. Estimates for these items were based on the assumption that their costs were proportional to the number of tanks to the power of 0.65.

12. OVERALL EVALUATION

12.1 Operational Assessment of the Designs

Operational problems in reprocessing plant are concerned mainly with active handling operations, which require precise movements to be conducted by an operator separated from the operation by a biological shield.

These operations fall into the categories of routine and non-routine

TABLE 28

CAPITAL COST OF LIQUID WASTE STORAGE FACILITIES

Storage Facilities	Cost (1972 \$A)
Waste Storage Tanks (3 off)	2,388,000
Condensate System	240,000
Waste Lines	124,000
Heat Exchanger and Cooling Coil Circuits	147,000
Operations Building	226,000
Valve Pits and Pit Piping	203,000
TOTAL LIQUID WASTE STORAGE	3,328,000

procedures. Routine active handling operations include fuel movement, basket transfer, cladding disposal, routine replacement of shear pack components, replacement of active plant filters, and maintenance of failed active pumps and ejectors. These operations must be carried out without extensive downtime in operation.

Non-routine operations are concerned with removal and replacement of failed plant or clearance of plant blockages. The likelihood of failure is unknown; it is more probable in leachers and evaporators which operate with corrosive solutions at high temperatures than in the relatively low corrosive environment of solvent extraction. The UK attitude has been to construct plant to high standards, and accept a low probability of failure which would require repair by contact maintenance following a protracted decontamination and plant outage of several months. The USA plants adopt this approach for solvent extraction plant, but provide for remote replacement of leachers and evaporators without increasing a prolonged shutdown of plant.

Design 1 incorporates adequate equipment for routine active handling procedures, which have been simplified through installation of handling equipment designed for the specific routine operations. A compromise between the UK and USA approaches has been adopted for removal and replacement of failed plant; equipment would be partially decontaminated to permit removal by handling equipment external to the cell operated behind a temporary biological shield. This scheme would require development of equipment modules designed for removal; in an extreme case the plant could be maintained by entry into the active cell after protracted decontamination. This design eliminates the capital cost inherent in the USA approach and is no less operable than the equivalent UK plant.

Design 2 is similar to Design 1 in maintenance provisions for the shear leach and evaporator area; the use of centrifugal contactors imposes further requirements for maintenance. These contactors may be replaced after preliminary decontamination by shielded flasks carrying lifting equipment. The design offers no economic advantages over Design 1.

Design 3 appears to offer advantages both in economics and maintenance. Fuel prebreakdown conducted under water has the advantage of visibility without incurring high costs inherent in a shielded dry cell. Similarly, removal and replacement of partially decontaminated plant under a water shield appears preferable to removal requiring operations conducted behind temporary remote shields, as in Designs 1 and 2. Design 3 involves risk of contamination of pond water; an allowance for equipment to remove activity from the water has been made but the design inevitably increases the risk of spread of contamination. The advantages of Design 3 in the head end shear-leach area require satisfactory development of a semi-continuous leacher to eliminate handling of active baskets containing leached cladding. The scheme appears to offer cost advantages, though in view of the novel equipment incorporated, the costing for Scheme 3 is more susceptible to under-estimation than for Designs 1 and 2.

Design 4 appears to offer no cost advantage over the alternative designs considered. It is doubtful if one shielded mobile cabin is capable of conducting the routine active operations in the time required. If the head end shear-leach operation were modified to eliminate the need for active handling by the mobile (e.g. by incorporation of the underwater head end facility of Design 3), then the mobile would be used for sampling and replacement of active instrumentation, changing of filters, and for replacement of plant in the event of failure. The residual routine functions could be performed by means other than a mobile cabin, and the utility of the latter would be dependent on the maintenance capability and requirement for replacement of failed plant. For equipment constructed to high standards of containment, the probability of failure is low; the mobile cabin philosophy would hence imply acceptance of a reduction in standards in construction of active plant to derive a reduction in capital costs.

The design offers poor operating conditions for personnel in the mobile.

It has disadvantages in demonstrating adequate control of contamination and radiation, and results in capital costs similar to those estimated for alternative designs more in keeping with international standards for radiological handling procedures.

12.2 Cost Effectiveness of Design Themes

The themes taken as a basis for cost reduction through design were:-

Design 1 : Simplification of process operations, with compact plant layout, and a new approach to plant maintenance.

Design 2 : As design 1, with centrifugal solvent extraction contactors in place of pulsed columns.

Design 3 : Use of a water shield for the high cost areas of assembly breakdown, shear leaching, and solvent extraction.

Design 4 : Application of protective shielding to the operational and maintenance personnel, rather than to the plant equipment.

The preliminary estimates of overall cost suggest that all the designs could be effective in reduction of capital investment. The similarity of the overall cost estimates indicate that the combination of themes in each design scheme has roughly the same potential for capital cost reductions.

12.2.1 Overall capital cost estimates

The preliminary estimates of capital cost presented in Table 14 for the four designs for a reprocessing plant of 3 t U/d capacity range from \$A 36.5 x 10^6 to \$A 40.5 x 10^6. These costs are inclusive of waste storage facilities (\$A 3.3 x 10^6) but exclude possible refinements for recovery of neptunium or removal of tritium and krypton-85 from plant effluents.

Data on capital cost and capacity of conventional reprocessing plant is presented in Table 29. The costs have been converted and updated to 1972 Australian dollars, and are shown graphically in Figure 87. The cost-capacity scaling exponent varies from 0.21 for derivation from NFS data (Nuclear Industry, 1970 and 1971) and 0.4 for derivation from data by Terjesen (1970). Exponents in the range 0.24 to 0.40 have been cited by Bosch (1968), Terjesen (1970) and Detilleux (1972).

It is notable that the capital cost of the GE plant is considerably below other estimates for a plant of identical capacity. However, as stated in Section 2, the economic ground rules and cost exclusions for the estimates have not been stated.

Figure 87 indicates a cost range of \$A 59.4-65.5 x 10^6 for a reprocessing plant capacity 3 t U/d (1972 prices), with a mean value of \$A 58 x 10^6. The

TABLE 29

CAPITAL COST ESTIMATES FOR REPROCESSING PLANTS

Plant	Capacity (tU/day)	Cost ($\$US \times 10^6$)	Date	Reference	Estimated Updated Cost (1972) ($\$A \times 10^6$)	Identification No. (Fig. 87)
Nuclear Fuels Services	1 ext. to 3	32.5 12	1964* 1971*	Nuclear Industry (1971) Nuclear Industry (1971)	40.6 10.8	1 1
General Electric (Mid West)	1	25.5	1968*	Nuclear Industry (1970)	28.3	2
Allied-Gulf Nuclear Services (Barnwell)	5	80 (approx.) 70 (approx.)	1971* 1970*	Nuclear Industry (1971)	72	3
British Nuclear Fuels Ltd. (Windscale)	5	£stg. 17	1964*	Hughes & Kemp (1970)	58.7	5
Estimated Costs	1	45 40 40-45	1970* 1969	Marcus & Seynaeve (1971) Terjesen (1970) Detilleux (1972)	45.8 42.7 40-45	6 7 8
	5	90 75 80-85	1970* 1969	Marcus & Seynaeve (1971) Terjesen (1970) Detilleux (1972)	<91 80 80-85	6 7 8

Note: 1. Costs updated by conversion at parity and application of Australian Implicit Price Level Index (Table 13).

2. Parity Correlations:
1964-1970 $US1.11 = $A1.0
1971 $US1.19 = $A1.0
1964 £stg.1 = £A1.25 = $A2.5

*Estimated date of construction.

overall cost figures in the present study show an apparent reduction of about 30-37% on the mean value derived for a conventional plant, or a reduction of about 20-28% in the overall expenditure to increase the NFS plant to 3 t U/d capacity.

Design 3, which explored the use of a water radiation shield for some parts of the plant, involved the least capital cost. Design 4, in which the plant was operated from a shielded mobile, led to the greater capital cost. The cost of Design 2, which employed centrifugal solvent extraction equipment was greater by \$1.9 x 10^6, than that of Design 1, where solvent extraction was carried out in pulsed columns.

The Mean Capital Cost of the four design concepts was \$A 38.7 x 10^6. Designs 1 and 3 were respectively 0.4% and 6% below the mean cost while Designs 2 and 4 were respectively 2% and 4% above the mean cost.

12.2.2 Process building costs

The capital cost of the process building structure for the four designs ranged from \$A 4.4 x 10^6 to \$A 5.3 x 10^6, with the greatest capital investment incurred by Design 4. These capital costs are remarkably similar in view of the diverse range of design principles employed. Building costs for Designs 1-3 might be expected to be of similar order, as these schemes all embody a compact plant layout. Design 4 has a relatively simple process building, but has an inherently poor utilisation of active contained volume.

12.2.3 Comparison of cost intensive process functions

The major cost areas in conventional reprocessing plants are those of fuel prebreakdown, shear-leaching and solvent extraction. Overall cost estimates for these functions in Designs 1, 2 and 3 have been made to assess the effect of equipment alternatives on function costs. Cost estimates for these areas in the NFS plant have been presented for comparison to indicate where economies may be achieved through the different approaches to plant layout, design, and maintenance philosophy. Costs for the process functions in Designs 1, 2 and 3 include a factor of 1.88, derived from Table 14, to convert direct construction costs to total costs; costs for NFS have been converted and updated to 1972 \$A. (Appendix C).

Fuel prebreakdown and shear-leaching

Table 30 compares capital investments incurred by the whole assembly shearing and batch leaching of Designs 1 and 2 with the alternative of underwater fuel prebreakdown, element shearing and semi-continuous leaching of Design 3, and the dry prebreakdown, element cluster shearing and batch leaching of the NFS plant.

TABLE 30

SUMMARY OF COSTS FOR THE FUEL PREBREAKDOWN, SHEAR AND LEACH FUNCTIONS

	NFS ($A)	Design 1 ($A)	Design 2 ($A)	Design 3 ($A)
Buildings	4,362,000	1,801,000	1,850,000	899,000
Chemical Process Equipment	620,000	207,000	207,000	220,000
Mechanical and General Equipment	3,455,000	2,410,000	2,410,000	1,360,000
Process Piping	630,000	132,000	132,000	113,000
Instrumentation and Electrical	853,000	337,000	337,000	86,000
	9,920,000	4,887,000	4,936,000	2,678,000

TABLE 31

SUMMARY OF COSTS FOR THE SOLVENT EXTRACTION FUNCTION

	NFS ($A)	Design 1 ($A)	Design 2 ($A)	Design 3 ($A)
Buildings	1,764,000	532,000	406,000	921,000
Chemical Process Equipment	677,000	1,070,000	2,350,000	1,070,000
Process Piping	1,037,000	681,000	451,000	681,000
Instrumentation and Electrical	388,000	415,000	342,000	415,000
	3,866,000	2,698,000	3,549,000	3,087,000

Note: Costs include indirects, $A 1972.

A comparison of data for these functions in Design 1 and in the NFS plant shows that building costs are 41% of those for NFS. This figure is consistent with the shielded volume requirements which are 43% of those in the NFS plant. Chemical process equipment costs are about 30% of NFS, arising probably from the elimination of two complex 6-basket leachers, 2.5 m dia., and substitution of two simple single basket leachers, 1.5 m dia., employing soluble neutron absorber for criticality control. Mechanical and general equipment is reduced by about 30% through elimination of in-cell general purpose handling equipment. The cost of process piping in Design 1 is only 20% of that for NFS. Over 80% of the NFS piping cost in this area is accounted for in provision for remote disconnects, used throughout for all equipment in the head end. Design 1 employs disconnects only for pipework attachments to the leachers, which are the vessels most susceptible to failure by corrosion. Instrumentation and electrical costs are about 40% of those in the NFS plant, probably due to reduction in the overall complexity and elimination of duplicate instrumentation.

Comparison between Designs 1, 2 and 3 indicates a substantial reduction in building costs in Design 3, probably from the elimination of cell roofs, windows, and 2 m thick shielding walls. Designs1 and 2 are substantially identical as is indicated by the cost data.

Elimination of the expensive whole assembly shear and transit flasks in favour of underwater breakdown and a disposable 4-element shear leads to a considerable reduction in mechanical and general equipment costs in Design 3. The continuous leacher is judged to be more expensive than the two batch leachers of Designs 1 and 2; accordingly, Design 3 has slightly greater process equipment costs, but shows a substantial reduction in piping and instrumentation.

Solvent extraction

A comparison of overall costs for the solvent extraction function in the NFS plant and in Designs 1, 2 and 3 is made in Table 31. Shielded volume requirements for this function in Design 1 are about 70% of those in the NFS plant, but building costs are estimated to be about 30% of NFS. The cost reduction is ascribed to the sharing of walls between cells as far as practicable, and in the elimination of walls serving only to segregate areas of differing activity.

Chemical process equipment in Design 1 costs are 58% above that of NFS, reflecting the greater nominal capacity of Design 1, in which the NFS design of pulsed column was increased in diameter to accept a 3 t U/d throughput.

Process piping is apparently 34% of that in the NFS plant, probably from the policy of compaction of layout and elimination of duplicate pipework connections. Instrumentation and electrical costs for the solvent extraction function are comparable for both Design 1 and the NFS plant. Overall costs for solvent extraction, inclusive of solvent treatment, have apparently been reduced by about 30% despite the increase in throughput.

A comparison between Designs 1 and 2 shows that use of centrifugal contactors results in a 25% reduction in shielded volume and a corresponding reduction in building costs. Total process piping length was reduced by about 50%, and overall piping costs by 34% in Design 2, employing centrifugal contactors, which is in reasonable agreement with Bernard (1971). Instrumentation costs were reduced by about 18%.

Chemical process equipment costs for Design 2 are over twice those for Design 1, employing pulsed columns, reflecting the high capital cost of the centrifugal extraction equipment, and the use of centrifugal contactors does not apparently result in an overall economic advantage for the design philosophies examined. (Bernard et al. 1971, predicted a minor (3.7%), overall economic advantage for centrifugal contactors versus pulsed columns, with conventional design philosophy.)

Design 3 has substantially greater building costs than Design 1, probably through the additional height of the water shield in the pond containing the equipment. The column well in Design 3 was designed to resist hydrostatic pressure and lacks the advantage of sharing common walls. It is accordingly more expensive than the equivalent well in Design 1.

The greatest saving in capital cost is therefore judged to be derived from a combination of pulsed column solvent extraction in dry cells (Design 1) with the underwater breakdown and shear-leaching of Design 3.

12.2.4 Effect of plant throughput on capital cost

The foregoing work has compared the cost effectiveness of alternative design concepts at a constant (3 t U/d) throughput. The effect of throughput on capital cost may be expressed by

$$\frac{I}{I_o} = \left(\frac{A}{A_o}\right)^n$$

where I and I_o are the capital investments for plants of throughput A and A_o respectively, and n is a scaling factor. The value of n for conventional chemical industry may vary between 0.5 and 0.9, depending on the degree of replication of plant items; a mean value of 0.7 is recommended (Buchanan and

Sinclair 1966) for a first order cost estimate.

In contrast, the investment cost of fuel reprocessing plant is less influenced by throughput, and cost exponents lie between 0.24 and 0.40 (Detilleux, 1972).

In the absence of other evidence, a value of n of 0.33 has been used to estimate the cost of reprocessing facilities of 1 t U/d and 5 t U/d as follows:

Design No.	Cost Estimates for a Capacity of 3 t U/d ($\$ \times 10^6$)	Extrapolation of Cost Estimates (a) 1 t U/d plant ($\$ \times 10^6$)	(b) 5 t U/d plant ($\$ \times 10^6$)
1	38.6	26.9	45.7
2	39.5	27.5	46.8
3	36.5	25.4	43.2
4	40.5	28.2	48.0

12.2.5 Comparison of plant shielding requirements

It is interesting to compare the active processing areas and volumes required for Designs 1, 2 and 3 with those for the NFS facility. Table 32 shows the shielding plant requirements for the NFS plant; the corresponding requirements for Designs 1, 2 and 3 are given in Table 33. All three designs show a significant saving in shielded space compared with the NFS plant, despite the fact that each is based on 3 U/d throughput whereas the NFS facility was rated at 1 t U/d.

Design 2 has the lowest processing plant requirements because compact centrifugal contactors are a feature of that design. As Design 3 requires a water shield, 2 to 3.5 metres in thickness, about the equipment it has the largest volume of the three schemes.

Table 34 compares the NFS plant and three of the designs proposed in this study on a process function basis. The major savings over the NFS facility occur in the head end, solvent extraction, waste evaporation and maintenance areas.

The largest saving arises in the head end region which is the most important in the overall economics since this area is the most heavily shielded (1.7 - 2.0 m of concrete). In Designs 1 and 2 savings in shielded volume have been achieved by:

(a) eliminating the fuel disassembly area where sheaths are split, end hardware cut away and pins removed in clusters prior to shearing,

TABLE 32

SHIELDED PLANT REQUIREMENTS FOR NUCLEAR FUEL SERVICES FACILITY

Nominal Capacity - 1 tonne U/day

Plant	Area (m^2)	Volume (m^3)
Chemical Process Cell	190.1	2,490
Process Mechanical Cell	58.0	442
Extraction Cell 1	24.5	411
Extraction Cell 2	40.5	710
Extraction Cell 3	29.6	524
Product Purification Cell	32.3	572
Equipment Decontamination Room	127.9	974
General Purpose Cell	46.7	277
Liquid Waste Cell	96.7	353
Chemical Crane Room	49.1	239
Hot Analytical Cells	20.1	61
Scrap Removal Area	39.6	133
Acid Recovery Cell	58.6	439
Mechanical Crane Room	30.1	133
Off Gas Cell	33.5	288
Crane Decontamination Room	13.3	93
Miniature Cell	11.2	72
Hot Acid Cell	28.3	129
TOTAL	930.1	8,340

TABLE 33

SHIELDED PLANT REQUIREMENTS FOR DESIGNS 1, 2 AND 3

Capacity 3 tonnes U/day

Function	Design 1		Design 2		Design 3	
	Area (m^2)	Volume (m^3)	Area (m^2)	Volume (m^3)	Area (m^2)	Volume (m^3)
Shear and Leach	77.2	750	77.2	750	118.6	897
Zircaloy Removal	74.3	249	74.3	249	34.8	169
Solvent Extraction - HA Liquor Storage	60.4	258	69.7	297	58.6	446
Solvent Extraction - LA Liquor Storage	107.5	557	113.5	588	112.9	723
Solvent Extraction - Pulsed Columns	15.7	335			15.7	373
Solvent and Off-gas Treatment	47.1	490	38.7	393	48.0	525
Uranium Evaporation and Purification	29.7	314	28.4	286	25.7	266
Plutonium Purification and Evaporation	29.8	300	29.5	208	25.8	267
Acid Recovery and Effluent Treatment	105.7	789	110.5	796	112.9	747
Waste Evaporation	47.4	347	47.6	349	49.4	362
Off-gas Filtration	27.6	46	32.5	54	41.5	70
Equipment Maintenance	47.4	347	47.4	347	51.1	375
Shear Maintenance	16.0	64	16.0	64		
Hot Analytical Cells	20.1	61	20.1	61	20.1	61
TOTAL	705.9	4,907	705.4	4,542	715.1	5,281
Percentage of NFS Requirements	76%	59%	76%	54%	77%	63%

TABLE 34

COMPARISON OF SHIELDED VOLUME REQUIREMENTS ON A PROCESS FUNCTION BASIS

Process Function	NFS (1964) (m^3)	Design 1 (m^3)	Design 2 (m^3)	Design 3 (m^3)
Head End*	2,346	999	999	1,066
Solvent Extraction	1,998	1,459	1,120	1,873
Off-gas Treatment	288	277	212	264
Uranium and Plutonium Purification	572	614	594	533
Acid Recovery and Effluent Treatment	568	789	796	747
Waste Evaporation	996	347	349	362
Maintenance Facilities	1,511	411	411	375
Hot Analytical Cells	61	61	61	61
	8,340	4,907	4,542	5,281

* includes fuel breakdown, leaching and cladding disposal facilities.

(b) feeding the whole assembly into the shear from a special transit flask outside the shear and leach cell,

(c) reducing cell height by eliminating in-cell cranes and power manipulators.

In Design 3, savings in the head end arise from the elimination of basket handling facilities and the use of a single continuous leacher.

The volume for waste evaporation in Designs 1, 2 and 3 has been reduced by the elimination of overhead in-cell cranes and power manipulators. As a consequence maintenance facilities are also reduced because shielded crane maintenance areas are not required. Savings in the solvent extraction area arise from compaction of equipment, by locating all pulsed columns in a single well and in the case of Design 2 by the substituting centrifugal contactors.

The canyon in Design 4 has a volume of 34,000 m^3 which is from 6-8 times the requirement for the other three designs. The poor utilisation of processing area arises because space must be provided for the travelling mobile. Also only about one half of the canyon depth can be efficiently used since a space must be provided above the tallest equipment to minimise the radiation shine through the roof.

13. POSSIBLE FUTURE REFINEMENTS

13.1 Waste Solidification

High level wastes can be stored more safely if they are converted from liquids to solids before ultimate disposal. Solid wastes are intrinsically safer being immobile, more rugged, less soluble and more compact. Four processes for waste solidification have been developed and demonstrated on an engineering scale in the USA.

Pot Calcination is a batch process in which liquid wastes are evaporated to dryness and calcined in a stainless steel cylinder which also serves as the final storage container.

Spray Solidification is a continuous process in which feed liquor is pneumatically atomised into the top of a heated cylindrical tower. The evaporated and calcined residue falls into a continuous melter from where it flows to the receiver-storage vessel.

Phosphate Glass Solidification is carried out by adding phosphoric acid to the waste and evaporating and denitrating to a thick, syrupy, aqueous slurry. The slurry is then fed to a continuous melter where water is completely volatilised and the glassy product forms. Finally, the molten glass is discharged into the final storage vessel.

Fluidised Bed Calcination is a high capacity continuous process in which feed waste is injected into a heated fluidised bed of granular solids through a pneumatic atomiser nozzle. The liquid deposits as a thin coating on the solid particles and is quickly evaporated and calcined. Product powder is continuously removed and pneumatically conveyed to storage bins.

Table 35 summarises the characteristics of each process; a full description is given elsewhere (Schneider 1970, ORNL 1970a).

The cost of waste solidification depends on the age of the wastes since the decay heat determines the size of the storage vessel and the extent to which wastes must be diluted with inert material. Capital costs for a pot calcination plant have been estimated by Perona et al. (1961) and recently updated (ORNL 1970a). Based on these figures, the capital cost of a solidification facility for a 3 U/d plant is estimated to be $A 7.5 million if wastes are immediately treated or $A 5 million if they are stored as liquids for five years. Immediate solidification is generally regarded as uneconomic because processing and perpetual storage costs are comparatively high. Costs for solidification process other than pot calcination are unknown but Schneider

TABLE 35

CHARACTERISTICS OF WASTE SOLIDIFICATION PROCESSES

	Pot Calcination	Spray Solidification	Phosphate Glass Solidification	Fluidised Bed Calcination
Form of Product	Oxide cake	Rocklike	Glass	Granular
Additives	Calcium, sulphate	Phosphate	Phosphate	None necessary
Typical Temperature of Formation	850°C	900°C	1,100°C	500°C
Typical mole % Fission Products	60	20	20	40
Typical Bulk Density (kg m^{-3})	1,300	3,000	2,800	1,500
Advantages	. Simple process . Very flexible	. Continuous process . Fairly flexible . High thermal conductivity . Not easily leached	. Continuous process . High thermal conductivity . Very resistant to leaching	. Continuous process . High capacity . Product easily handled
Disadvantages	. Low thermal conductivity . Easily leached . Requires stainless steel pot	. Moderately complex . Control of flow rates critical	. Moderately complex . Operates with slurries which are difficult to handle	. Moderately complex . Low thermal conductivity . Easily leached

(1970) estimates that the fluidised bed process should be comparable while conversion costs for spray and phosphate glass solidification are likely to be higher.

13.2 Further Off-Gas Treatment

As pointed out in Section 10.2.2, the proposed stack discharge rates fall within current international limits. However, with increasing emphasis now being given to reducing release rates from all nuclear facilities to 'near zero', some discussion of the technology and possible costs of more extensive off-gas treatment methods is warranted.

13.2.1 Voloxidation

Treatment of volatile off-gas streams is simplified if the fission products can be released in a concentrated form. This is best achieved before dissolution by heating fuel pellets which have been treated in some way to release occluded volatiles. In the voloxidation process (ORNL 1979b, ORNL 1972) sheared fuel pieces are heated in a stream of oxygen at temperatures from 450-700°C to oxidise the UO_2 to U_3O_8. In the resultant phase change the fuel pellets are broken down into a powder and fission product volatiles can escape.

At present, the main application envisaged for the voloxidation process is in LMFBR or short cooled LWR reprocessing where decontamination factors approaching 10^8 for ^{131}I are required. The technique also has special application to tritium recovery because, if tritiated water is not removed before dissolution, it becomes for practical purposes inseparable from the aqueous effluent stream. Research has not proceeded to a stage where the required process equipment has been identified and no reliable costs are available.

13.2.2 Iodine

In discussing the treatment of iodine, its various forms must be distinguished. Iodine may be present in the off-gas system as I_2, HI, HIO, iodates and as organic iodides and aerosols (ORNL 1970b). Elemental iodine and HI are most simply removed by caustic scrubbing but organic iodides pose greater problems. At the Hanford Purex plant, for example, 80% of the iodine is volatilised from the dissolver. Even so, about 80% of the iodine released from the stack originates from the vessel off-gas system (ORNL 1970b). The bulk of the effluent iodine is presumably present as organic iodides formed by reaction of iodine with the solvent.

A great many methods have been developed to improve the overall decontamination factor for iodine. Among the more promising are concentrated nitric acid or mercuric nitrate-nitric acid scrubbing (ORNL 1972), absorption on charcoal beds (Adams et al. 1968) and reaction with silver or silver zeolites (Maeck et al. 1968). If very high decontamination factors are to be obtained, a method of iodine treatment which is effective for organic iodides must be incorporated in the off-gas train or else the organics must be oxidised to a form which is more readily removed.

If an overall decontamination factor of about 10^5 is desired, the present off-gas system could be modified by adding further stages of treatment after the existing caustic scrubber (see Figure 11). Typical additions to the off-gas train could be as follows: a catalytic oxidiser bed of $CuO\text{-}MnO_2$, silver reactor, caustic scrubber and finally a charcoal bed absorber. The total cost for such a system including indirects is estimated to be about $800,000 if incorporated in a grass roots plant.

13.2.3 Noble gases

Many processes have been proposed for the isolation of noble gases from off-gas streams. These include adsorption on solids (Weller 1959, Stevenson 1958), cryogenic distillation (Stevenson 1958), absorption in liquids (Merriman et al. 1968, 1970), separation using permselective membranes (Rainey 1968), clathrate or fluoride precipitation (ORNL 1970b) and thermal diffusion (Slansky 1969). For 189 day cooled fuel, only ^{85}Kr represents a significant radiological hazard. Table 36 summarises the status and estimated costs for the more promising methods of noble gas treatment.

It is not clear if the capital costs in Table 35 include buildings and indirect costs. A more recent estimate of $US750,000-$US1,500,000 is given by General Electric (1971) based on industry quotations. The costs were based on a packaged plant for a 1 t U/d facility and does not include installation and other project costs. From these figures, it would appear that the 'all-up' cost of a noble gas removal plant designed for 3 t U/d throughput is probably about $2,000,000.

13.2.4 Tritium

Tritium occurs in irradiated fuel mainly as HTO. While it is technically feasible to recover the bulk of the tritium from the H.A. waste by fractional distillation, it appears more logical to isolate the tritium before the leaching stage. The voloxidation process allows tritium to be removed from the fuel with yields of over 99% (ORNL 1972). Once liberated, HTO can be trapped on any common dessicant.

General Electric Company, USAEC Docket 50268-21, estimated the cost of a tritium recovery facility for a 1 t U/d plant at about (US)$10 million. This estimate appears rather high; the complex voloxidation process is presumably

TABLE 36

METHODS OF NOBLE GAS TREATMENT

(Slansky et al. 1969, ORNL 1970b)

Process	Kr Recovered	Status	Advantages	Disadvantages	Capital* Cost ($US 1969)	Operating Cost ($US/tU)
Adsorption on Charcoal Beds at Ambient Temperature	99%	Pilot plant with reactors	.Simple operation .Accepts dilute feed gas	.Large bed volume .Charcoal can ignite	1,000,000	170
Adsorption on charcoal or Silica-gel Beds at Low Temperature	99%	Pilot plant at Idaho Processing Plant	.Small bed volumes .Accepts dilute feed gas	.High refrigeration costs .Charcoal can ignite	1,000,000	150
Cryogenic Distillation	98%	Pilot plant at Idaho Processing Plant	.Compact equipment	.Explosion hazard with ozone	800,000	100
Absorption in Freons	98%	Inactive pilot plant	.Compact equipment .No explosion hazard*	.High pressures required	800,000	100
Permselective Membranes	99%	Laboratory and bench scale studies	.Compact equipment .No safety hazards	.Moderate pressures required .High power costs	1,500,000	200

* Refers to reprocessing plant capacity of 3 tonnes U/day.

responsible for the major component of this cost.

13.3 Ultimate Waste Disposal

13.3.1 Solid wastes

Each year a 3 t U/d reprocessing plant generates about 55 m^3 of cladding hulls, 600-3500 m^3 of low-level solid wastes and if waste solidification facilities are installed, about 85 m^3 of high-level fission product wastes (USAEC 1972). It is proposed that cladding hulls and low-level solids will be buried on site. Cladding will be isolated from the environment by using the concrete-shielded, zircaloy waste flask (see Figure 25) as the final storage container. Each flask has a capacity equivalent to about three days operation.

Only two methods for ultimate disposal of high level solid wastes are currently feasible; storage in geological formations or in man-made structures.

(1) Storage in Geological Formations In this approach, sealed, solid waste canisters are stored in individual holes deep (typically 150 m) below ground at regular spacings determined by the capacity of the rock to dissipate the decay heat. The major research effort in the USA has been devoted to storage in salt which appears the most attractive economically (Perona et al. 1963). Project Salt Vault was inaugurated in the early sixties and a pilot facility was set up at Lyons, Kansas to study the feasibility of ultimate storage in salt beds. Recently the USAEC decided to abandon the Lyons site because of engineering problems and community opposition (Nucleonics Week 1972). Storage in other rock formations such as granite and limestone has been considered (Perona et al. 1963, ORNL 1970a). Limestone is of interest because it predominates in Southern Central Australia but its low thermal conductivity is a distinct disadvantage.

(2) Storage in Man-Made Structures This involves storing wastes close to the surface (about 4-10 m underground) in steel or concrete vaults. Most of the decay heat is dissipated to the surface through the earth overlay. Storage in vaults is initially a more positive form of containment though many experts consider it inferior for ultimate disposal since wastes are close to the earth's surface and man-made structures are unlikely to survive the geological time-scale over which fission product wastes must be isolated from the biosphere.

Insufficient information is available to give a reliable estimate of the capital cost of ultimate disposal facilities although some estimates of overall costs have been attempted (Perona et al. 1963, Bradshaw et al. 1969, ORNL 1970a). Capital costs are dependent on the age of wastes at burial, the method of disposal, the proximity to the reprocessing site and the manner in which new storage areas are developed. Since it is more economical to develop storage facilities as the need arises, the initial capital facilities can be expected to be small compared to total costs.

The initial capital cost of a national salt-vault facility in the USA catering for a 20,000 MW nuclear economy has recently been estimated at $US 11-18 million with total costs over the expected twenty year lifetime of from $US 90-106 million. (ORNL 1970a). These figures correspond to a waste storage cost of about $6 per million kWh generated. The cost of storage in concrete vaults has been estimated to be 5-7 times as much i.e. about $35 per million kWh generated.

13.3.2 Noble gases

Krypton and xenon are produced in one tonne of the reference PWR fuel to the extent of 0.37 kg and 5.4 kg respectively. Suggested methods of disposal include storage in underground geological formations (Tadmor and Cowser 1967), trapping as clathrates or fluorides and storing as solid wastes (ORNL 1970b) and compression and storage of the gas in cylinders at a pressure of 15 MPa (Blomeke and Perona 1969).

The capital cost of a noble gas bottling facility has been estimated to be about $A250,000 (Blomeke and Perona 1969). The noble gas output of a 3 t/d reprocessing plant would require that fifty cylinders per year be filled. If krypton and xenon were separated before bottling and only krypton was encapsulated, ten cylinders would be required. Xenon is essentially inactive one year after irradiation and could be released to the atmosphere or sold commercially if sufficient markets could be developed. At the current price of about $25 per litre (STP), the value of xenon in the spent fuel is about $24,000/t U. The possible output from a 3 t U/d reprocessing plant of over 800,000 litres/year could depress the price of xenon considerably. The entire USA consumption of xenon for the year 1964 was only 31,000 litres (Kirk-Othmer 1966).

Once bottled, the noble gases could be stored under water or in a cell shielded by about 0.3 m of concrete. Alternatively, the cylinders could be shipped to a fuel depository with the solid wastes. The storage area needed for krypton disposal would be insignificant compared to that required for solid wastes.

13.4 Additional Cost of Possible Facilities for Radioactive Wastes

Table 37 presents overall cost estimates for advanced facilities for ultimate disposal of radioactive wastes from a reprocessing plant of 3 t U/d

TABLE 37

ESTIMATED ADDITIONAL COST
OF POSSIBLE FACILITIES FOR WASTE TREATMENT AND DISPOSAL

Treatment of Wastes	$\$A \times 10^6$
1. Highly Active Waste Solidification	
1.1 Pot or Fluidised Bed Calcination (after 5 years decay storage)	5
1.2 Pot or Fluidised Bed Calcination (with no decay storage)	7.5
1.3 Glassification Processes	7.5
2. Krypton Removal	2
3. Iodine Removal	0.8
4. Tritium Removal	10
Subtotal, Treatment	17.8-20.3
Storage of Wastes	$\$A \times 10^6$
1. Highly Active Waste Storage Vault (20,000 MWe programme)	11 - 18
2. Noble Gas Bottling	0.25
Total Additional Cost (rounded)	29 - 39

capacity. Treatment processes could cost in the range $\$A18\text{-}20 \times 10^6$ with major costs incurred by facilities for removal of tritium ($\$A10 \times 10^6$) and fixation of highly active wastes ($\$A5\text{-}7.5 \times 10^6$); smaller costs ($\$A2.5 \times 10^6$) could be incurred for treatments to remove krypton and radio-iodine.

Estimated costs of ultimate storage facilities are in the range $\$A11\text{-}18 \times 10^6$, incurred mainly by a Highly Active Waste Storage Vault.

Total additional costs imposed by waste disposal from environmental considerations could be in the range $\$A29\text{-}39 \times 10^6$.

13.5 Neptunium Recovery

Neptunium-237 is produced during nuclear fuel irradiation by neutron reactions with both uranium-235 and 238. It is a valuable intermediate in the production of plutonium 238 which is used as an isotopic heat source in applications such as space probes and cardiac pacemakers.

One tonne of reference PWR and BWR fuel (Table 4) contain respectively approximately 0.44 kg and 0.35 kg of ^{237}Np (McKee and Nunn 1969). The value of ^{237}Np has been variously estimated at \$40,000-\$150,000 per kg (Ramdohr 1970, Smith and Bigge 1970). Assuming 80% recovery, an average content of 0.4 kg/t U and a worth of \$40,000/kg the value of neptunium recovered is \$12,800/t U processed.

Neptunium can be extracted from process raffinate wastes by solvent extraction and/or ion exchange. Solvents that have been studied include tributyl phosphate, tri-n-octylamine and trilaurylamine(Sheppard 1957, Gourisse 1971, Berger et al. 1971) but most production experience has been with tributyl phosphate (TBP). The order of extraction into TBP from nitrate solutions is Np(VI)$\gg$Pu(IV)>Np(IV)>Pu(VI)>>Np(V). Tests at NFS showed that, without flowsheet modifications, 50-70% of the neptunium appeared in the partition cycle aqueous waste. Savannah River experience indicates that at least 95% of the neptunium can be directed to the HA raffinate stream by addition of sodium nitrite near the raffinate discharge point to form Np(V) (NFS 1968).

Nuclear Fuel Services have outlined a possible flowsheet for recovery of neptunium from concentrated raffinate streams. Neptunium is conditioned to the Np(IV) form by addition of ferrous sulphamate and hydrazine and then extracted into TBP in a pulsed column. The neptunium is stripped back to the aqueous phase with dilute nitric acid, concentrated by evaporation and acidified to 6 M HNO_3. The conditioned neptunium solution is loaded onto an anion exchange bed, washed to remove fission products and plutonium, and eluted with 0.3 M HNO_3 to recover neptunium.

No detailed costing of a neptunium recovery facility has been attempted in this report, but a rough estimate of capital cost can be obtained from the NFS preliminary design which is based on a throughput equivalent to 2 t U/d (Nuclear Fuels Services, Inc., USAEC Docket 50201-1). The proposed NFS plant required a building volume of 16,000 m^3 and an active cell volume of 1,050 m^3. If it is assumed that building costs are proportional to volume while equipment, piping, instrumentation and sampling costs are proportional to cell volume, it is possible to estimate capital costs for a neptunium recovery plant from those determined for the reprocessing plant proper. Table 38 shows a breakdown derived in this way but modified to allow for processing of 1 kg

TABLE 38

ESTIMATED CAPITAL COST OF 1 kg/d NEPTUNIUM RECOVERY PLANT

	Cost (1972 $A)
Building Cost	1,440,000
Equipment, Piping and Instrumentation	1,300,000
Sampling and Analytical	230,000
Services Reticulation	130,000
Direct Construction Cost	3,100,000
Engineering, Construction and Contractor's Fee (27%)	840,000
Estimated Plant Cost	3,940,000
Contingency (20%)	790,000
Interest on Construction Funds	230,000
Working Capital	250,000
Preoperational Costs (9 months)	420,000
Estimated Total Capital Expenditure (rounded)	5,600,000

Np/d (approximately equivalent to 3 t U/d) using the equation (Nuclear Industry 1970).

$$\text{Capital Cost} \propto (\text{Plant Capacity})^{1/3}$$

The contingency allowance has been set at 20% for the neptunium recovery facility (compared with 15% for the reprocessing plant) because of greater uncertainties in the technology. Interest on funds has been calculated assuming a construction period of two years with 30% of the capital investment in the first year and the remaining 70% in the second.

The estimated capital expenditure for a 1 kg/d neptunium recovery plant is $5.6 million. At a neptunium price of $12,800/t U processed, about $9 million could be derived from neptunium sales per year. Even after allowing for operating costs and taxes, the payback time is likely to be less than two years.

Superficially then, neptunium recovery appears most attractive. But there is a danger that production will exceed demand resulting in a drastic reduction in Np price. In the USA, General Electric plan to recover ^{237}Np in their Midwest Fuel Recovery Plant and both Allied-Gulf and NFS have seriously examined the technology and market (Sinclair et al. 1971). Because the demand for neptunium in the future is difficult to predict due to the limited and special uses for ^{238}Pu, it would be essential for a reprocessor to negotiate contracts for sale of neptunium before committing capital.

14. CONCLUSIONS

14.1 Preliminary capital cost estimates in the range $A 36.5-40.5 x 10^6 have been derived for 4 designs of a chemical reprocessing plant of 3 t U/d capacity, in which an unorthodox approach to layout and maintenance philosophy has been linked to existing shear-leach-solvent extraction technology. The estimates indicate a cost reduction through design of up to 30-37% in comparison with an interpolated cost estimate of $A 58 x 10^6 for a conventional plant of identical capacity and scope, producing uranium and plutonium as nitrate solutions, and with 5 years storage capacity for fission product waste solutions. The estimates exclude development costs. Additional costs exceeding $A 30 x 10^6 could be incurred for refinements to effluent treatment and storage systems envisaged for environmental protection for any type of reprocessing plant whether conventional or unconventional. The additional costs could be incurred mainly by a highly active waste storage vault ($A 11-18 x 10^6), facilities for removal of tritium ($A 10 x 10^6), and fixation of highly active wastes ($A 5-7.5 x 10^6). Smaller costs could be incurred for removal and storage of noble gases ($A 2.8 x 10^6) and radioiodine ($A 0.8 x 10^6).

14.2 Lowest overall capital costs were indicated for a design featuring use of a water radiation shield for some highly active plant areas, while the greatest costs were obtained for a plant deriving radiation protection from ground and air exclusion zones, with active operations performed by an operator in a mobile shielded cabin. Intermediate estimates apply to a plant featuring a compact and unorthodox layout of equipment, avoidance of duplication of facilities, elimination of general purpose handling equipment in active cells, and with active maintenance performed by decontamination and erection of temporary radiation shields as and when necessary. The preliminary estimates are substantially below predictions for conventional plant. The apparent saving is sufficient to absorb a significant under-estimate and still leave a worthwhile capital saving from the radical approaches to design.

14.3 Comparison of equipment options for the head end shear-leach operations have indicated that fuel prebreakdown and semi-continuous leaching below a water shield could effect some economy over whole assembly shearing and batch leaching in a dry cell, maintained by decontamination and construction of temporary shielding facilities. The greatest potential for capital cost reduction in this function appears to be derived from savings in civil costs through elimination of fuel prebreakdown operations in shielded dry cells, and through the elimination of general purpose handling equipment and shielded cells for remote replacement of failed plant items.

14.4 A more compact plant design with some pipework economy resulted from employment of centrifugal contactors in place of pulsed column equipment, but an increase in overall cost for the solvent extraction function resulted through the relatively high cost of the centrifugal contactors.

14.5 Non-routine operations are concerned with removal and replacement of failed plant or clearance of plant blockages. The likelihood of failure is unknown; it is more probable in leachers and evaporators which operate with corrosive solutions at high temperatures than in the relatively low corrosive environment of solvent extraction. The UK attitude has been to construct plant to high standards with duplication in some areas, and accept a low probability of failure which would require repair by contact maintenance following a protracted decontamination and plant outage of several months. The USA plants adopt this approach for solvent extraction plant, but provide for remote replacement of leachers and evaporators without involving a prolonged shutdown of plant.

14.6 Design 1 incorporates adequate equipment for routine active handling procedures, which have been simplified through installation of handling equipment designed for the specific routine operations. A compromise between the UK and USA approaches has been adopted for removal and replacement of failed plant; equipment would be partially decontaminated to permit removal by handling equipment external to the cell operated behind a temporary biological shield. This scheme would require development of equipment modules designed for removal; in an extreme case the plant could be maintained by entry into the active cell after protracted decontamination. This design eliminates the capital cost inherent in the USA approach and is no less operable than the equivalent UK plant.

14.7 Design 2 is similar to Design 1 in maintenance provisions for the shear leach and evaporator area; the use of centrifugal contactors imposes further requirements for maintenance. These contactors may be replaced after preliminary decontamination by shielded flasks carrying lifting equipment. The design offers no economic advantages over Design 1.

14.8 Design 3 appears to offer advantages both in economics and maintenance. Fuel prebreakdown conducted underwater has the advantage of visibility without incurring high costs inherent in a shielded dry cell. Similarly, removal and replacement of partially decontaminated plant under a water shield appears preferable to removal requiring operations conducted behind temporary remote shields, as in Designs 1 and 2. Design 3 involves risk of contamination of pond water; an allowance for equipment to remove activity from the water has been made but the design inevitably increases the risk of spread of contamination. The advantages of Design 3 in the head end shear-leach area require satisfactory development of a semi-continuous leacher to eliminate handling of active baskets containing leached cladding. The scheme appears to offer cost advantages, though in view of the novel equipment incorporated, the costing for scheme 3 is more susceptible to under-estimation than for Designs 1 and 2.

14.9 Design 4 appears to offer no cost advantage over the alternative designs considered. It is doubtful if one shielded mobile cabin is capable of conducting the routine active operations in the time required. If the head end shear-leach operation were modified to eliminate the need for active handling by the mobile cabin (e.g. by incorporation of the underwater head end facility of Design 3), then the cabin would be used for sampling and replacement of active instrumentation, changing of filters, and for replacement of plant in the event of failure. The residual routine functions could be performed by means alternative to the use of a mobile cabin, and the utility of the latter would be dependent on the maintenance capability and requirement for replacement of failed plant. For equipment constructed to high standards of containment, the probability of failure is low; the mobile philosophy would hence imply that a reduction in standards in construction of active plant could be exploited to derive a reduction in capital costs. The design offers poor operating conditions for personnel in the mobile. It has disadvantages in guaranteeing adequate control of contamination and radiation, and results in capital costs similar to those estimated for alternative designs more in keeping with international standards for radiological handling procedures.

14.10 The narrow range of cost estimates derived from diverse approaches to economy through design indicates the high cost inherent in production scale reprocessing operations employing existing technology. It suggests that the design theme used for cost reduction should be examined against criteria other than capital cost. Thus the apparent advantage in reduction of capital charges in reprocessing should be assessed against the likelihood of major plant failure, and the penalty of an increased plant outage for maintenance by ad hoc methods. Similarly, any design featuring reduction in the degree of radiological shielding or containment to achieve capital cost economy should be assessed against the increasing stringency of world standards for the nuclear industry.

14.11 The most promising aspect for development work to reduce capital costs in fuel reprocessing by existing technology appears to be in the field of mechanical handling. This work could include development of special purpose tools for routine process operations (e.g. basket movements, or maintenance of shear packs), and evaluation of maintenance techniques for removal and replacement of active equipment liable to failure, using lifting devices external to the active cell in conjunction with temporary radiation shields. This approach would extend the trend towards contact maintenance of partially decontaminated plant, accepted for solvent extraction equipment.

14.12 The type of mechanical handling equipment and maintenance facilities suggested for capital cost reduction would best be developed in conjunction with detailed plant design work. It is not appropriate for the Research Establishment to contribute to this field in the absence of a commitment for an Australian Reprocessing Plant. More general development work which could be considered by the RE could be:

- Continuous moving bed ion exchange contacting equipment. This will probably be required for reprocessing operations with plutonium, but could be of interest in uranium milling and to the metallurgical industry, in general.

- Equipment for electrolytic reduction and separation of uranium and plutonium. Experience gained could be of use in alternative manufacturing routes for UF_4 in fuel manufacture.

- Rapid decontamination techniques, including electrolytic deplating procedures. This could be of interest in decontamination of active equipment arising from HIFAR, in addition to its specific use in treatment of shear packs.

15. RECOMMENDATIONS

It is recommended that the Commission:

- Note the unconventional approaches to design and maintenance philosophy likely to reduce capital cost;

- Propose a more detailed examination of these principles in the event of a firm commitment to an Australian Reprocessing

Plant;

- Consider background development work in the reprocessing field where the results are of general application to the uranium industry;

- Consider an examination of possible savings through design of the repository for highly active wastes.

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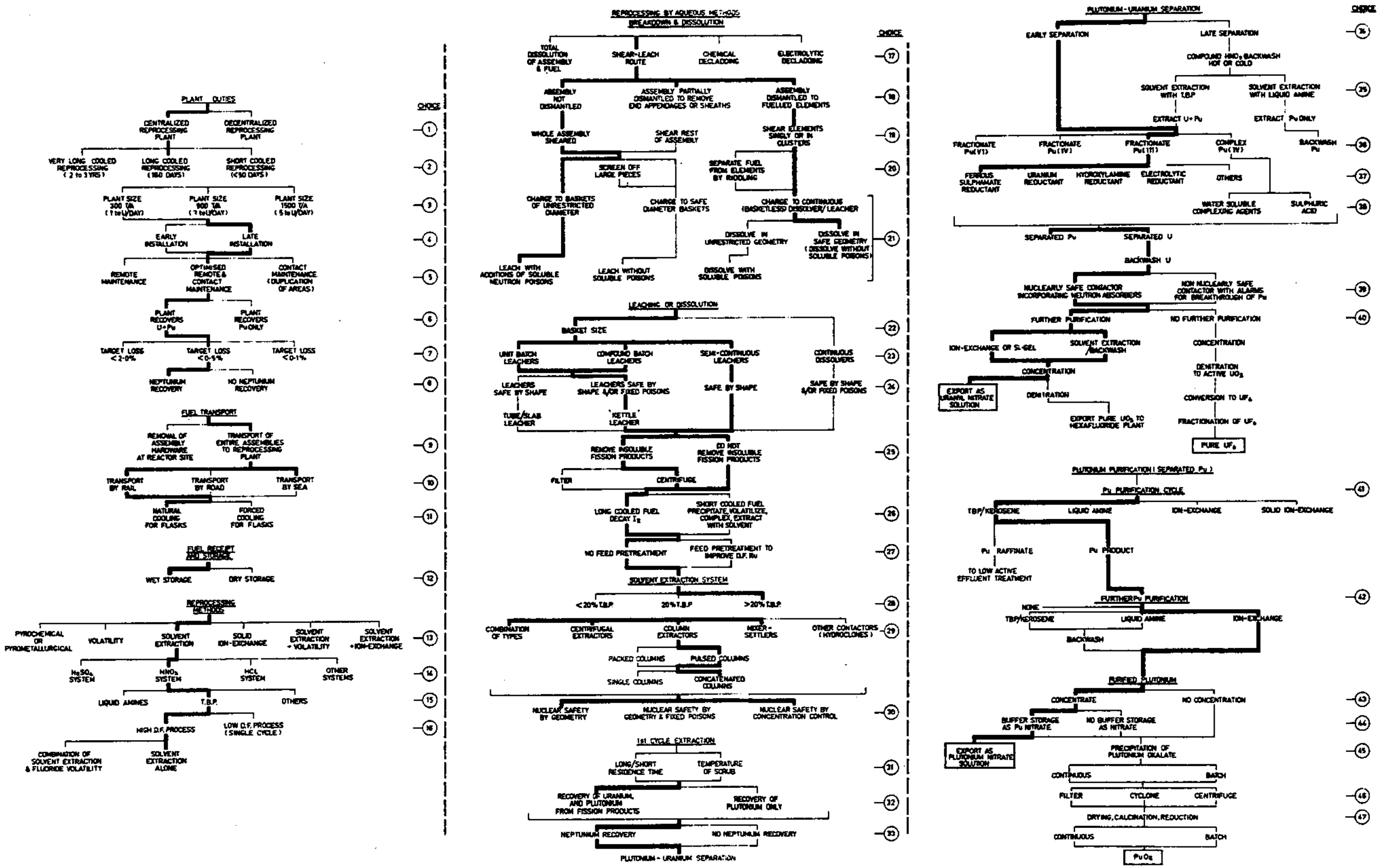


FIGURE 1. THERMAL OXIDE FUEL REPROCESSING PLANT AND PROCESS ALTERNATIVES

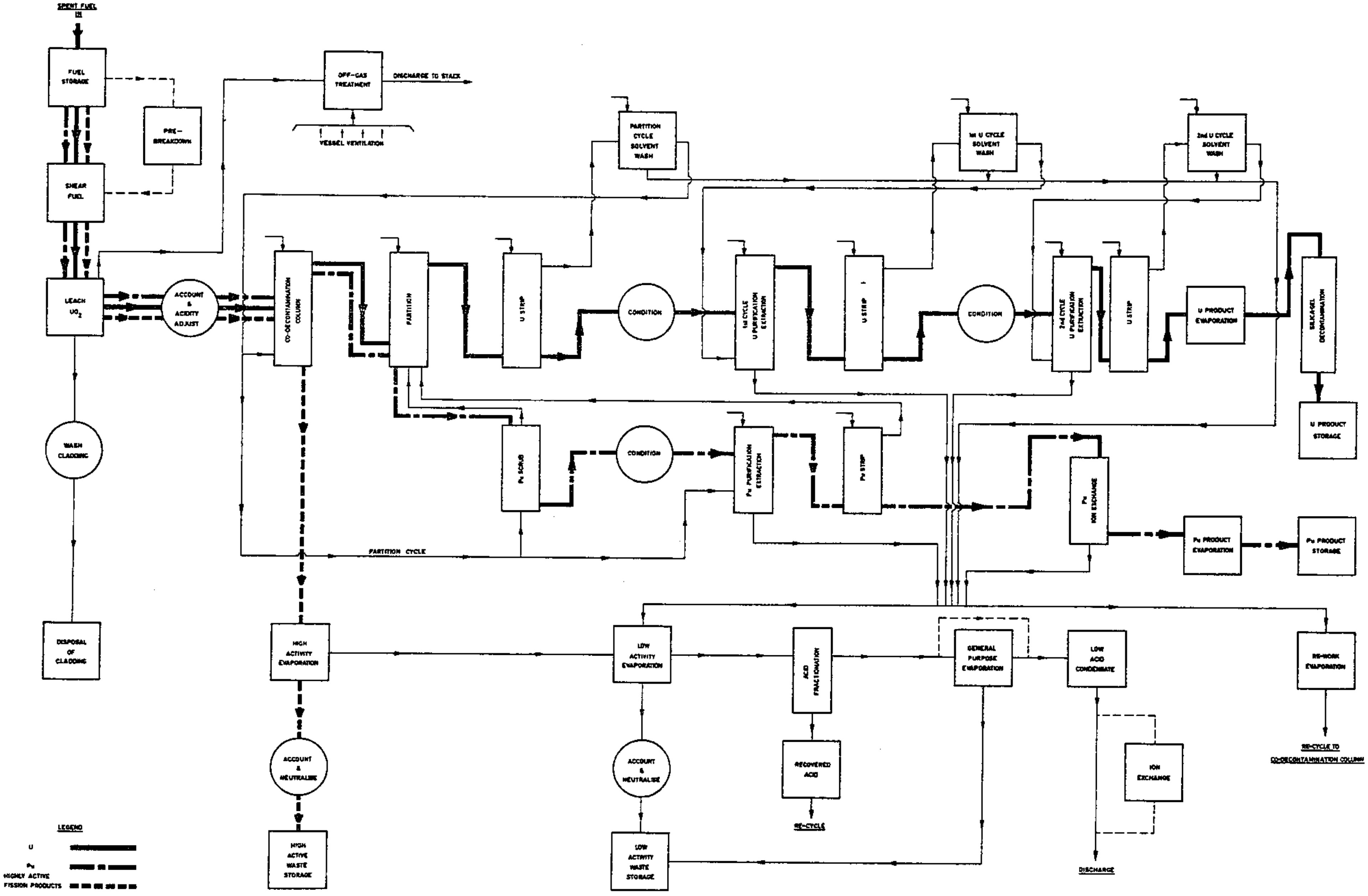
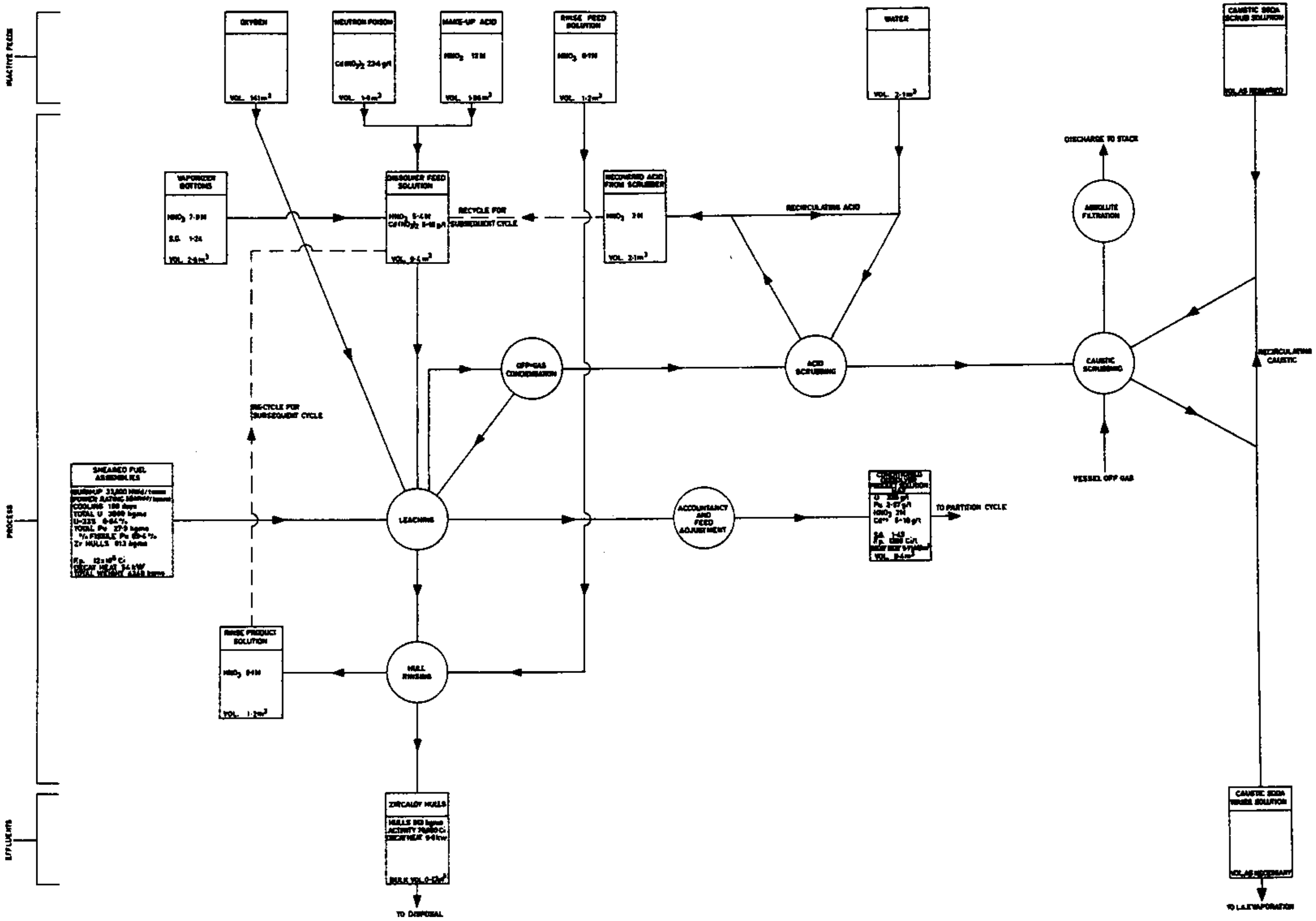
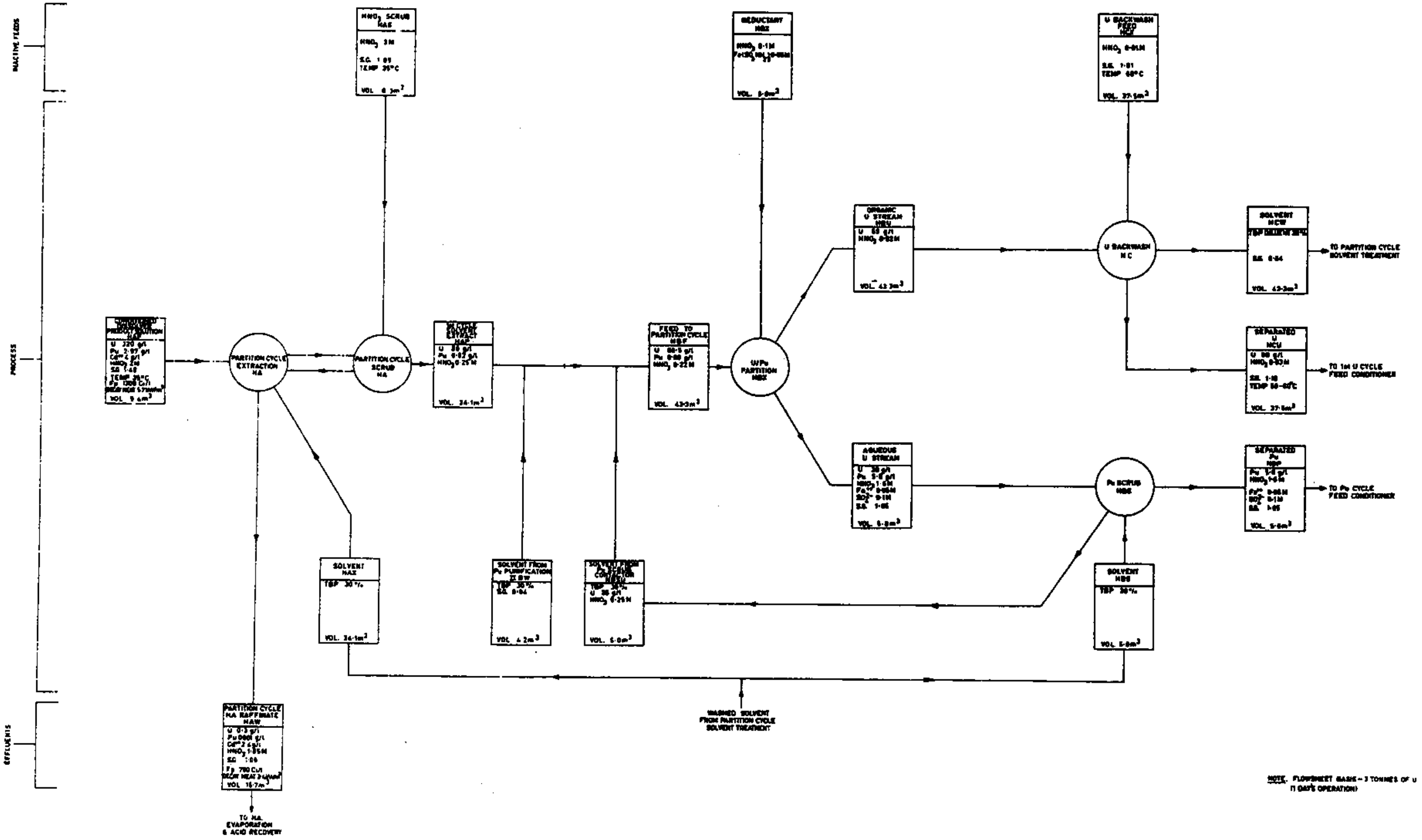


FIGURE 2. OVERALL PROCESS BLOCK DIAGRAM



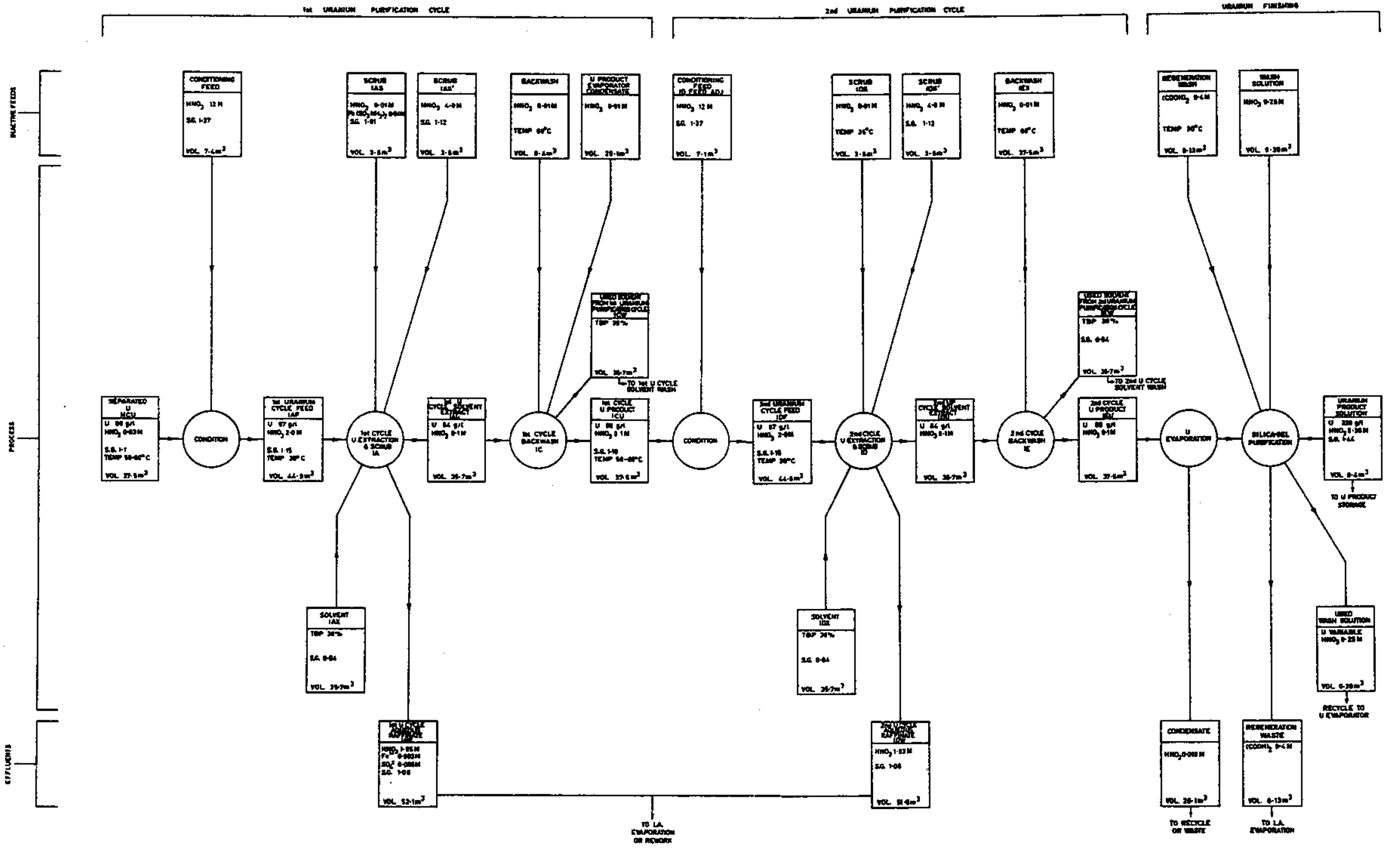
NOTE. FLOWSHEET BASIS – 3 TONNES OF U (1 DAY'S OPERATION)

FIGURE 3. CHEMICAL FLOWSHEET - LEACHING AND OFF-GAS TREATMENT



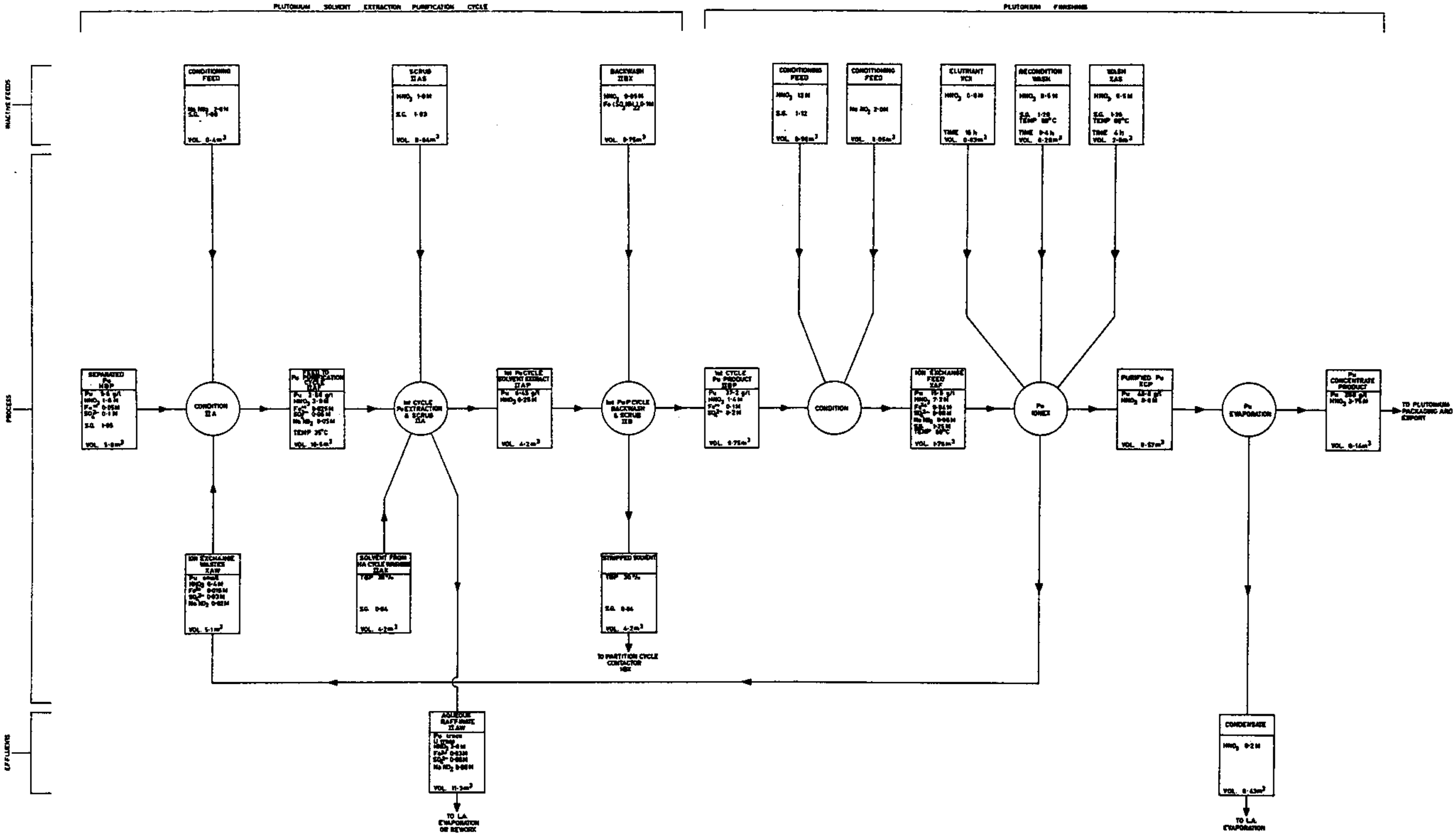
NOTE. FLOWSHEET BASIS – 3 TONNES OF U (1 DAY'S OPERATION)

FIGURE 4. CHEMICAL FLOWSHEET - PARTITION CYCLE



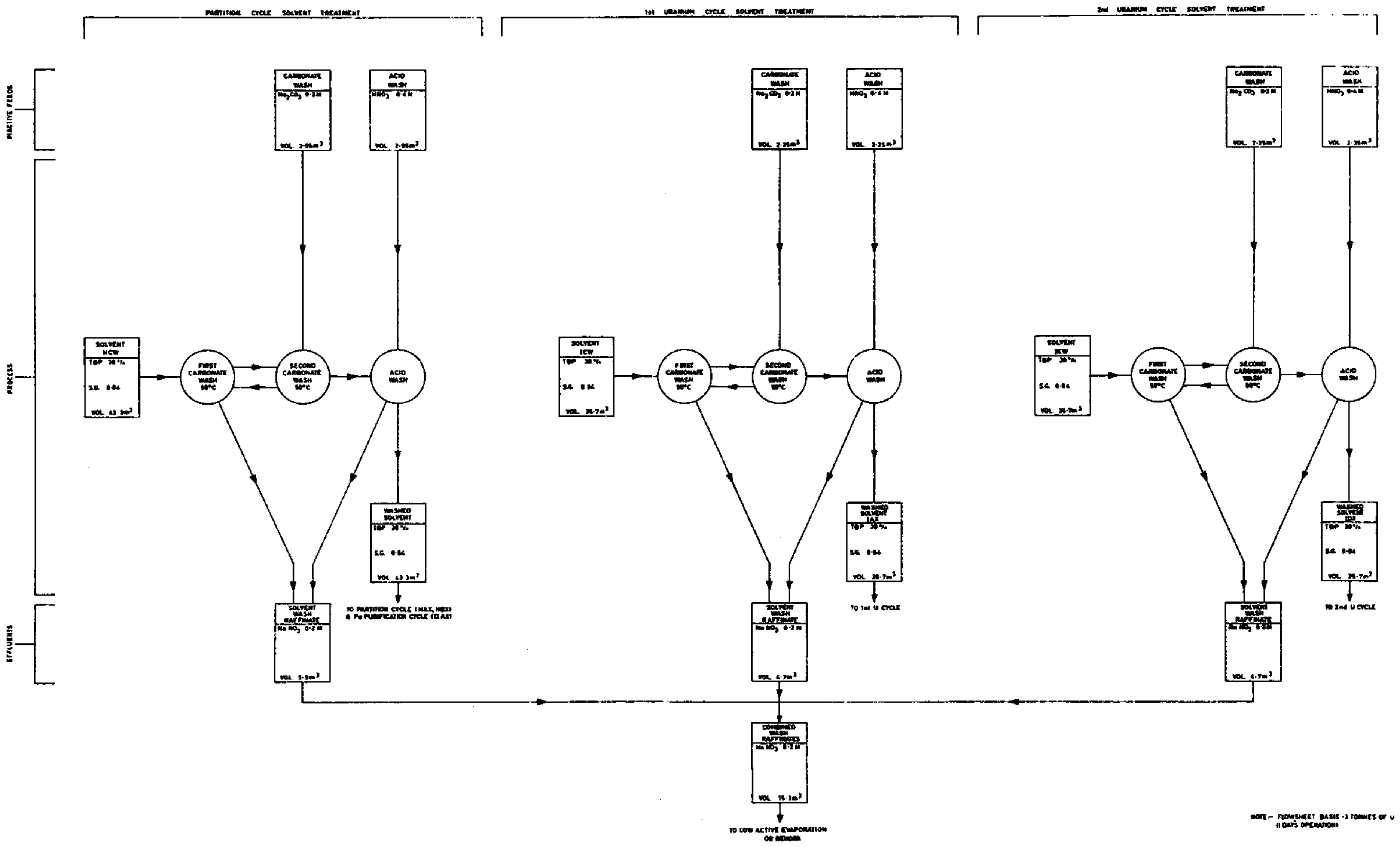
NOTE. FLOWSHEET BASIS – 3 TONNES OF U (1 DAY'S OPERATION)

FIGURE 5. CHEMICAL FLOWSHEET - URANIUM PURIFICATION



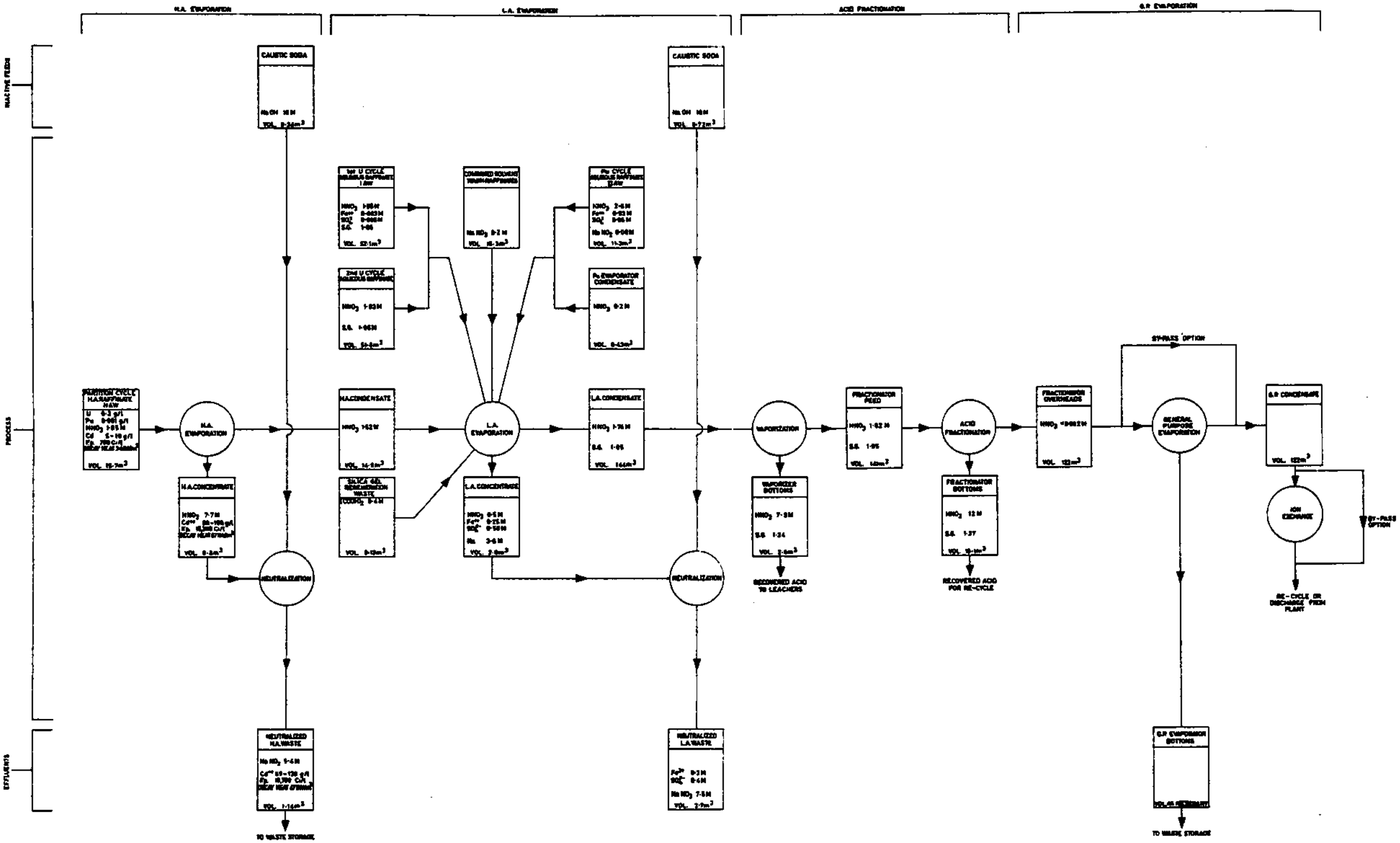
NOTE. FLOWSHEET BASIS—3 TONNES OF U
(1 DAYS OPERATION)

FIGURE 6. CHEMICAL FLOWSHEET - PLUTONIUM PURIFICATION



NOTE — FLOWSHEET BASIS - 3 TONNES OF U
(1 DAYS OPERATION)

FIGURE 7. CHEMICAL FLOWSHEET - SOLVENT TREATMENT



NOTE. FLOWSHEET BASIS — 3 TONNES OF U (1 DAY'S OPERATION)

FIGURE 8. CHEMICAL FLOWSHEET - WASTE EVAPORATION AND NITRIC ACID RECOVERY

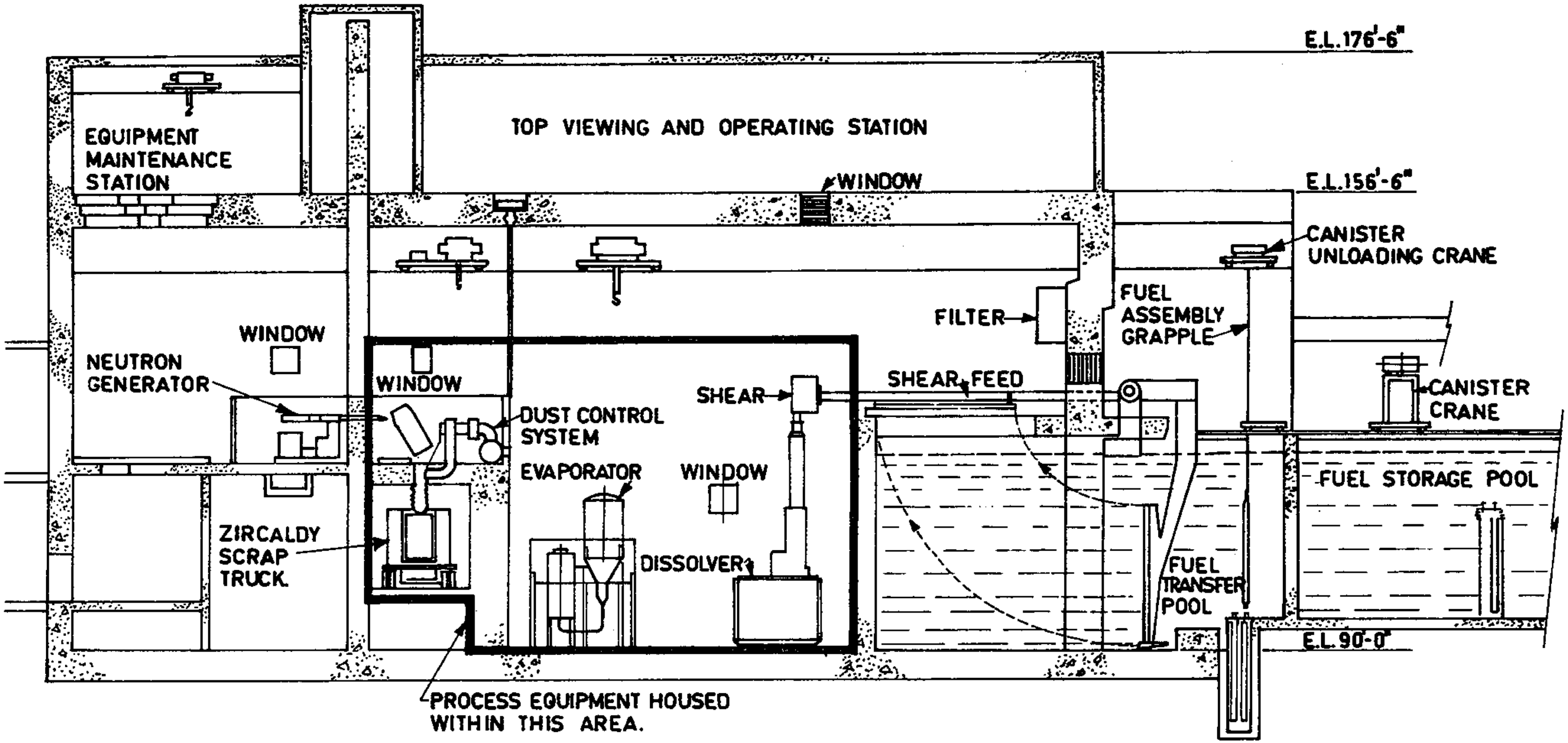


FIGURE 9. REMOTE PROCESSING CELL (ALLIED/GULF NUCLEAR SERVICES REPROCESSING PLANT)

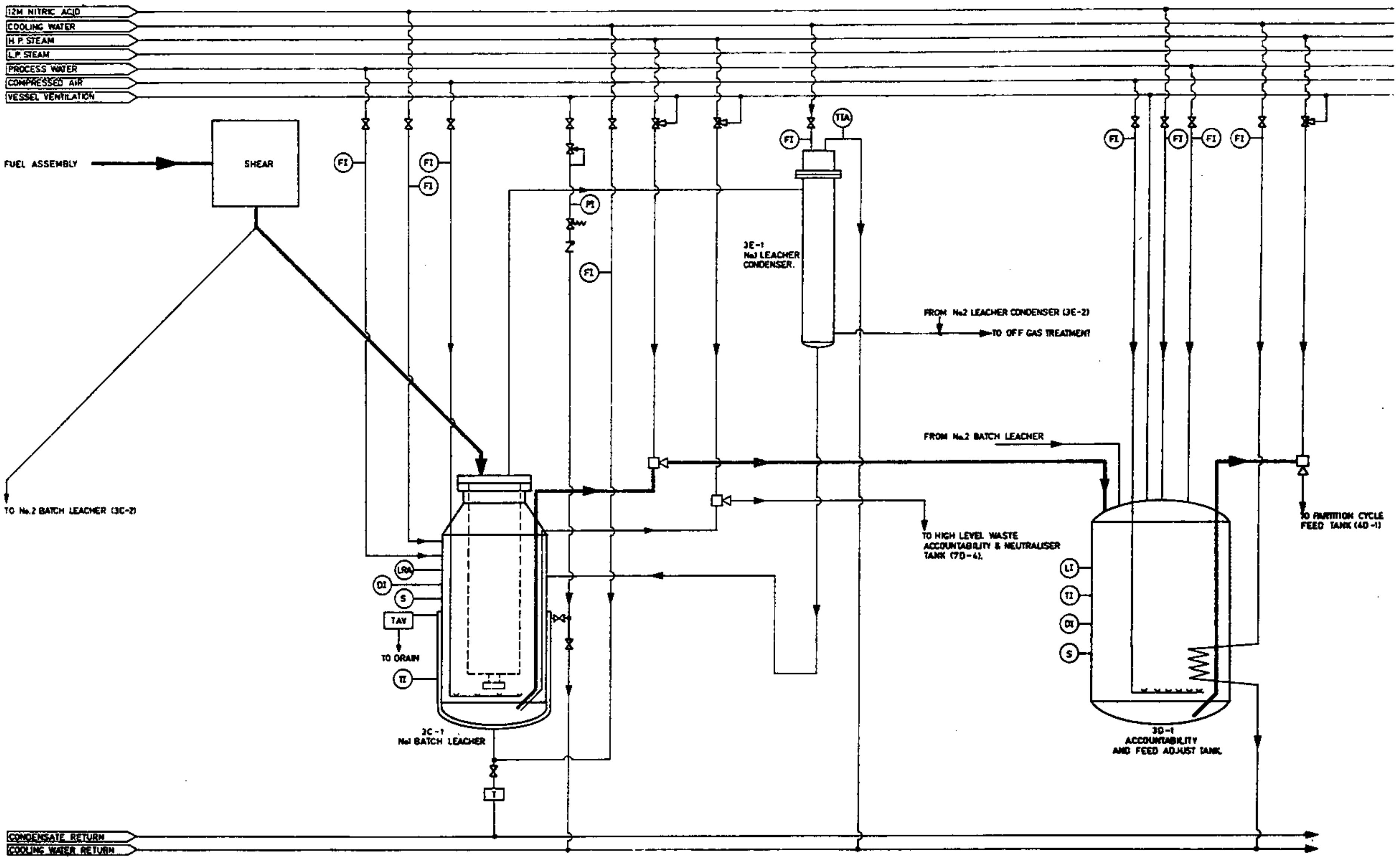


FIGURE 10. ENGINEERING FLOW DIAGRAM - BATCH LEACHING

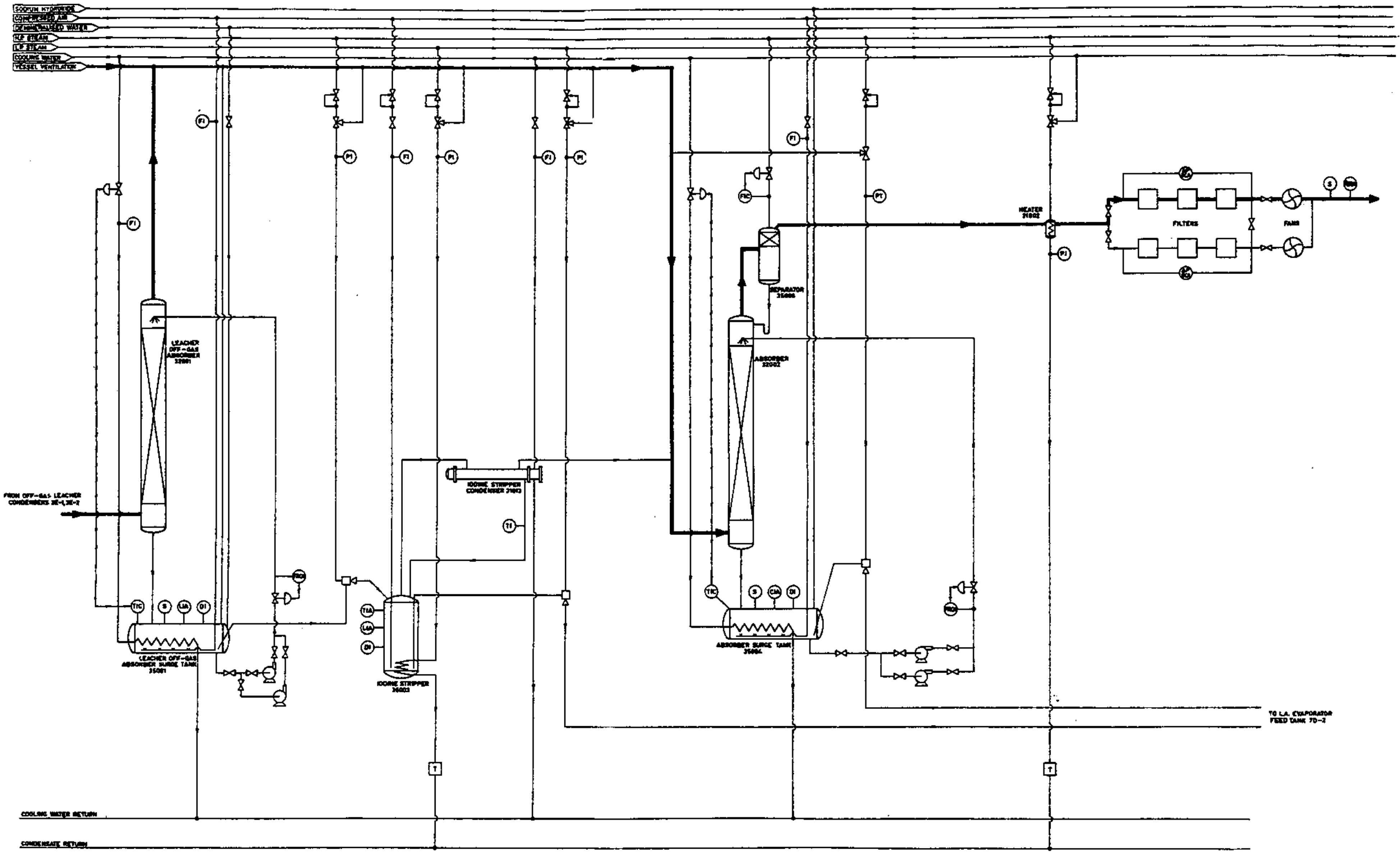


FIGURE 11. ENGINEERING FLOW DIAGRAM - OFF-GAS TREATMENT

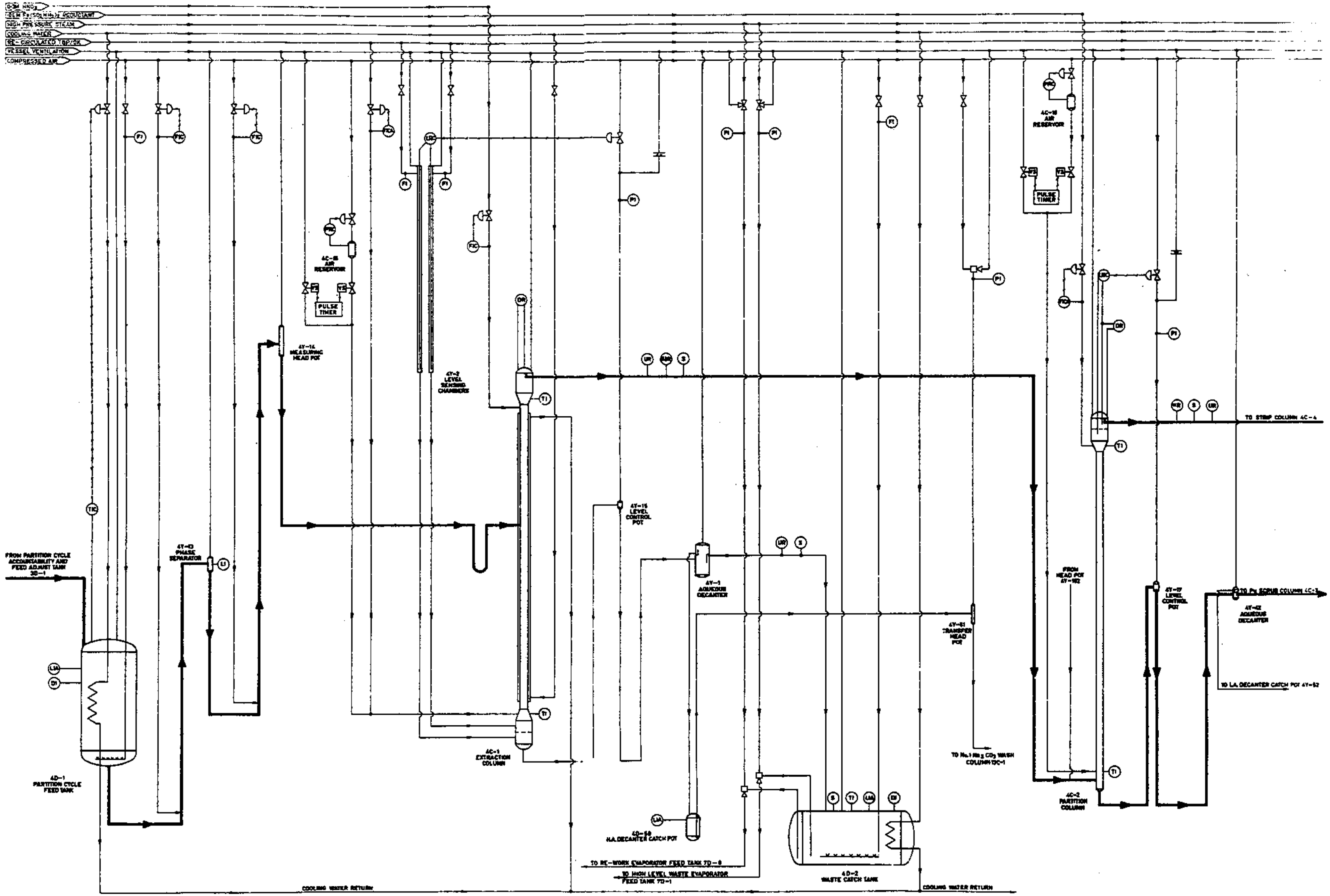


FIGURE 12. ENGINEERING FLOW DIAGRAM - PARTITION CYCLE PART 1 - EXTRACTION AND PARTITION

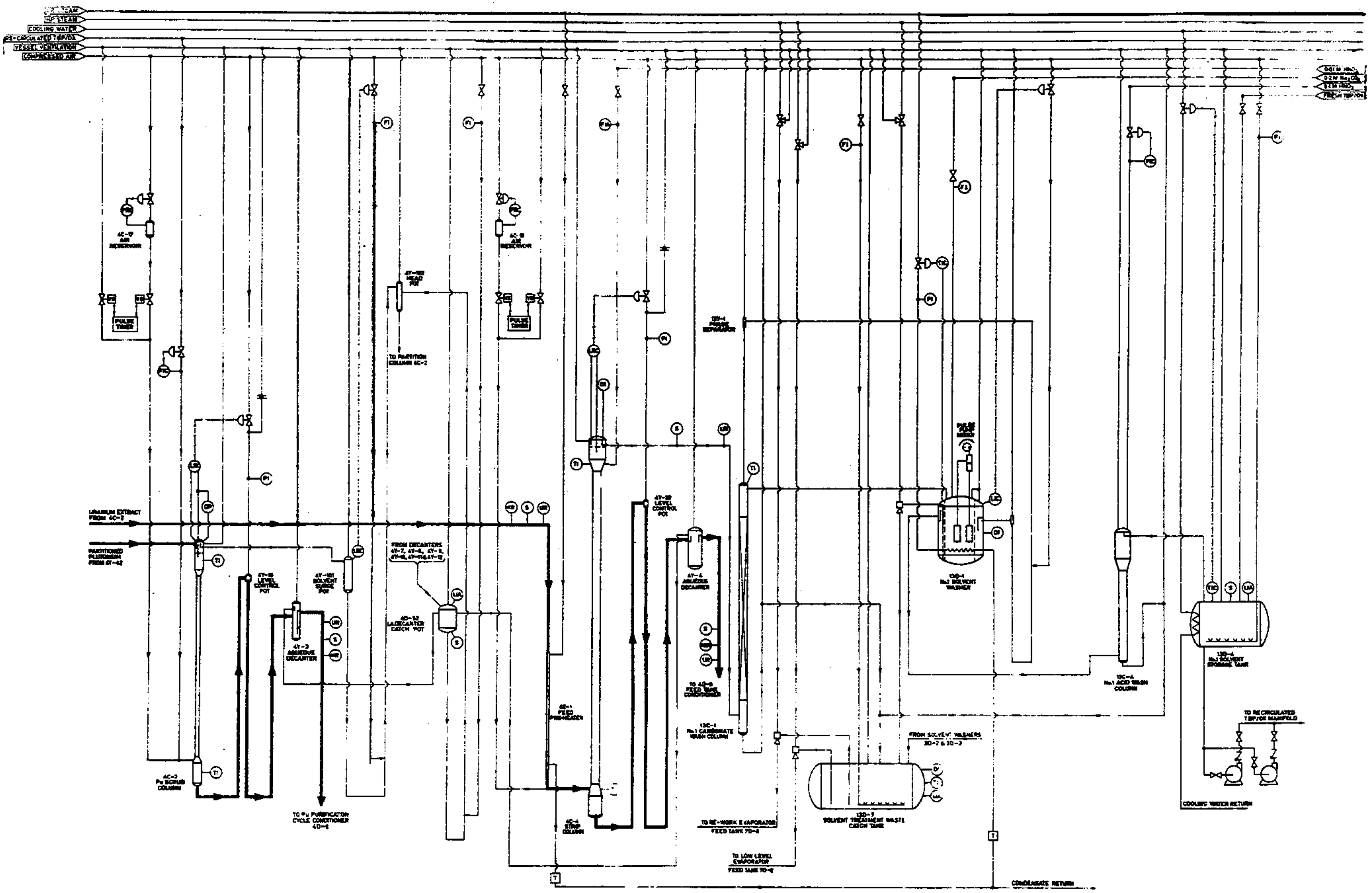


FIGURE 13. ENGINEERING FLOW DIAGRAM - PARTITION CYCLE PART 1
- SCRUBBING, STRIPPING AND SOLVENT TREATMENT

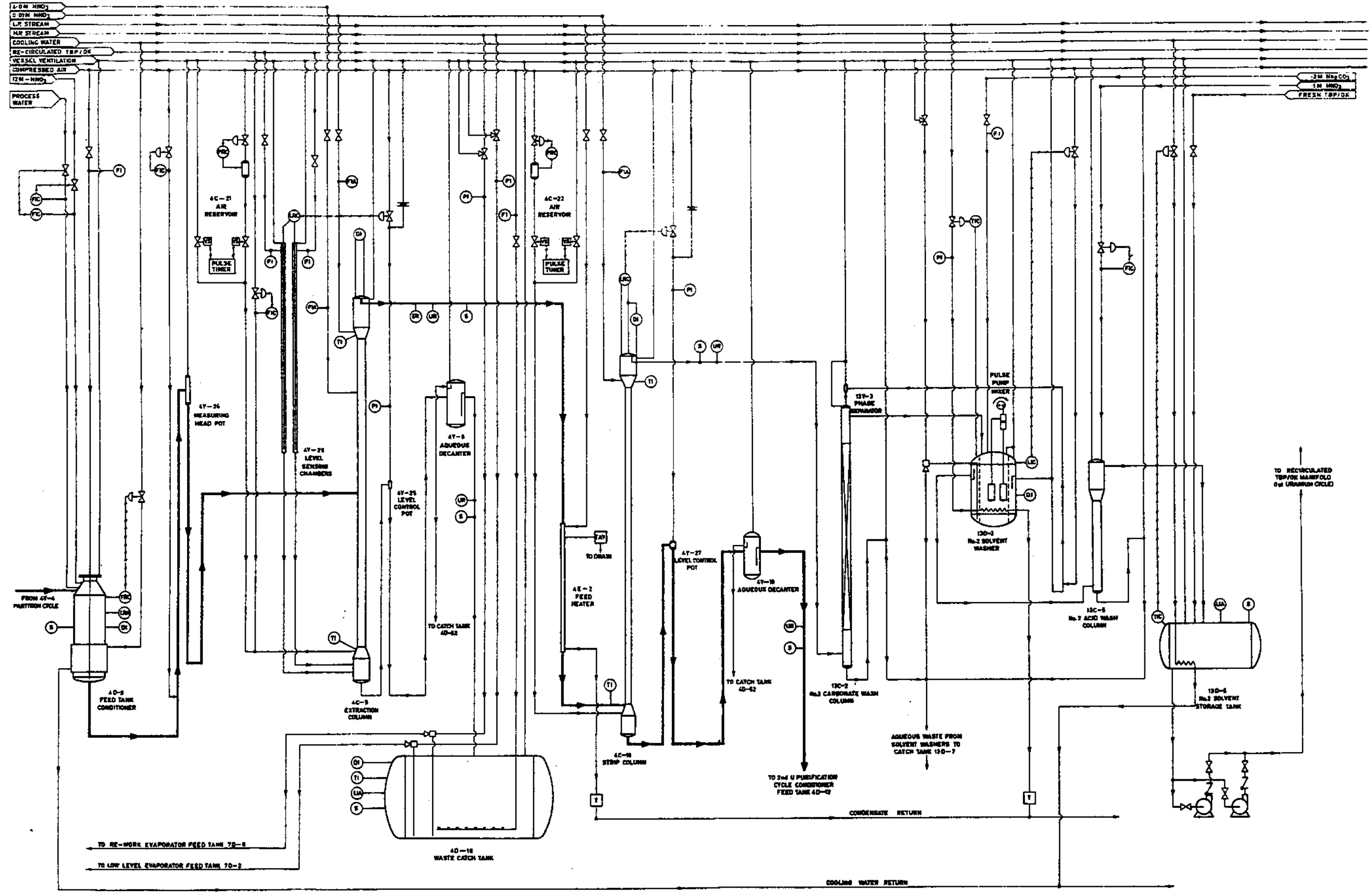


FIGURE 14. ENGINEERING FLOW DIAGRAM - 1st URANIUM CYCLE

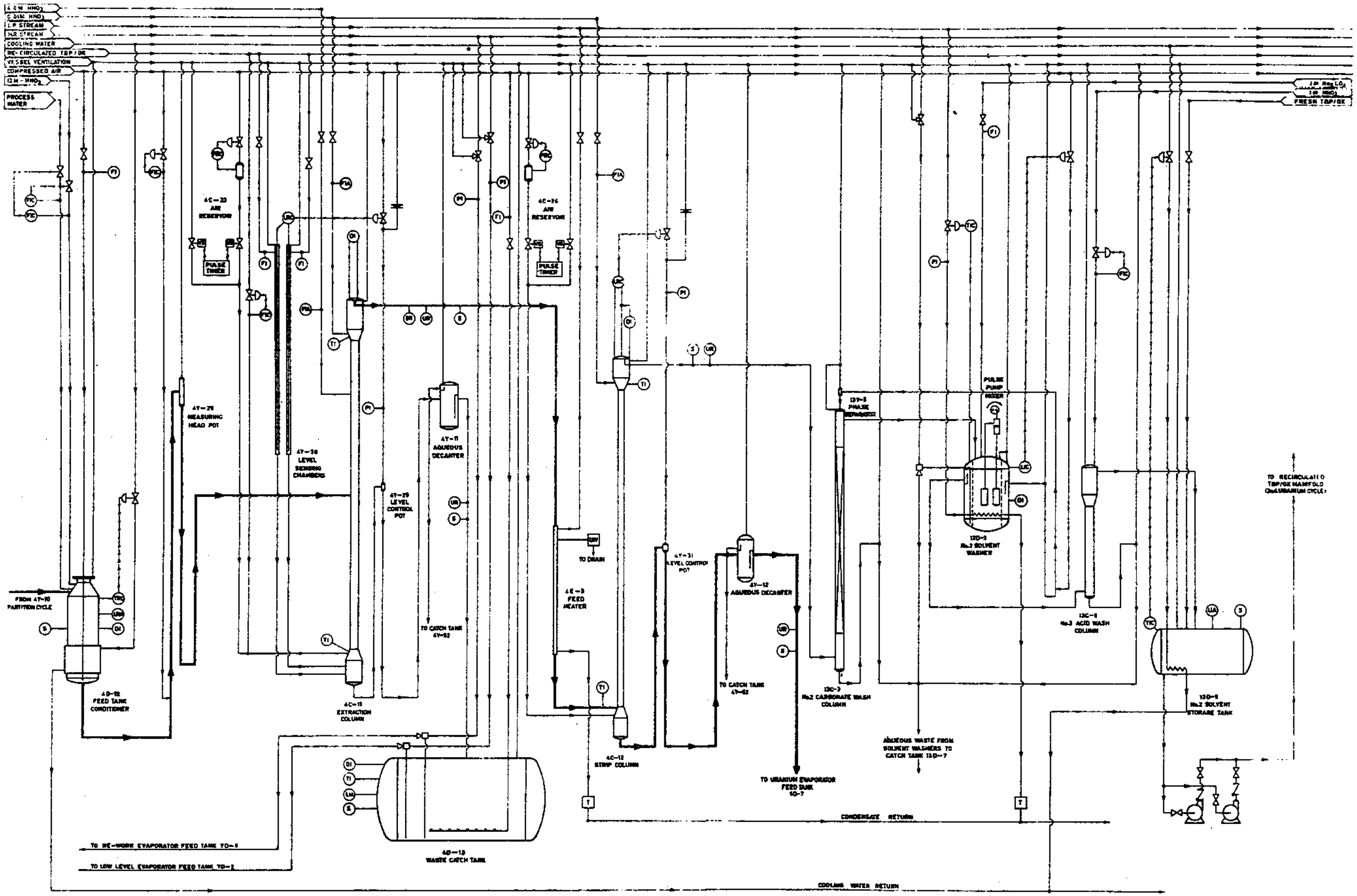


FIGURE 15. ENGINEERING FLOW DIAGRAM - 2nd URANIUM CYCLE

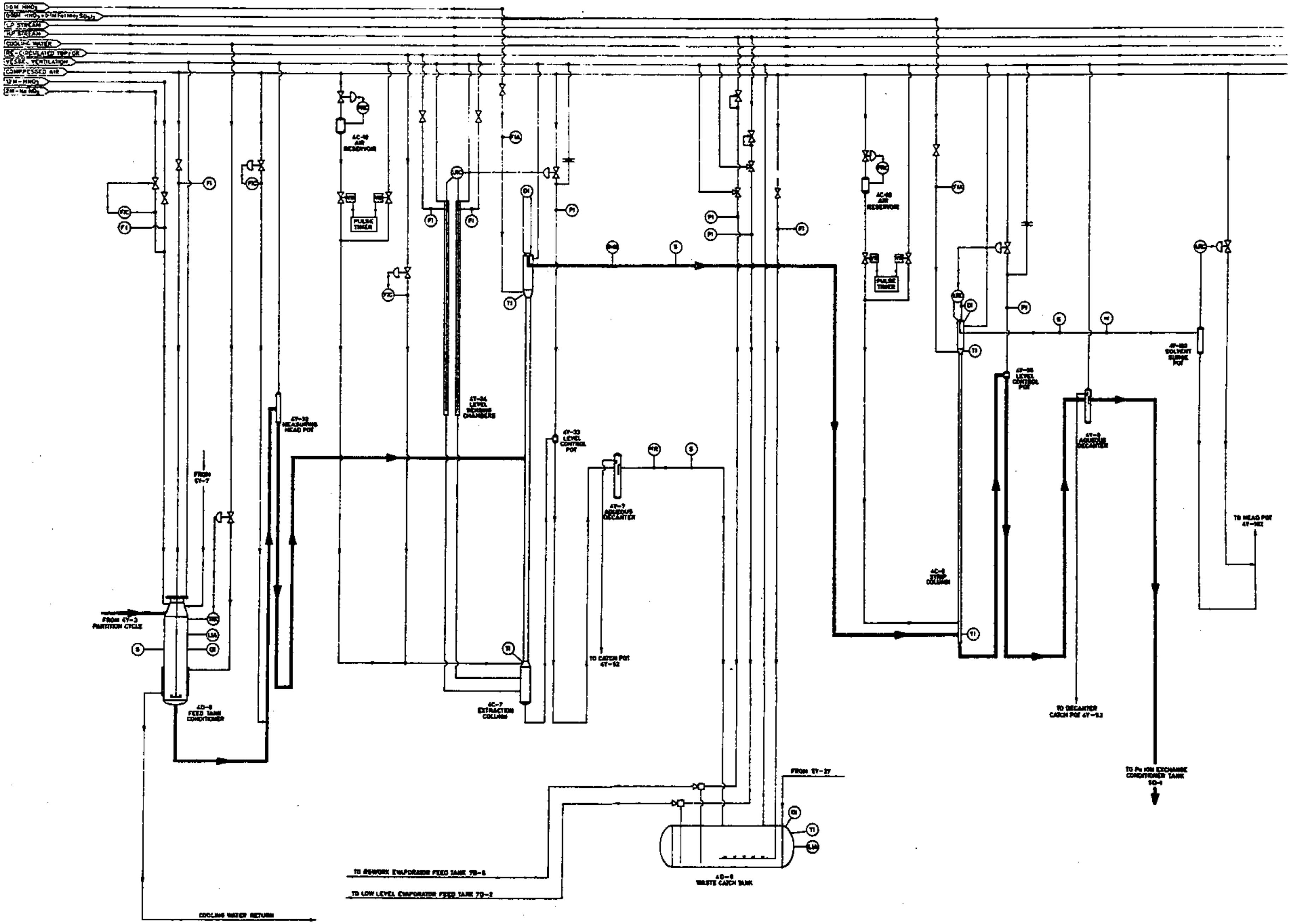


FIGURE 16. ENGINEERING FLOW DIAGRAM - PLUTONIUM PURIFICATION CYCLE

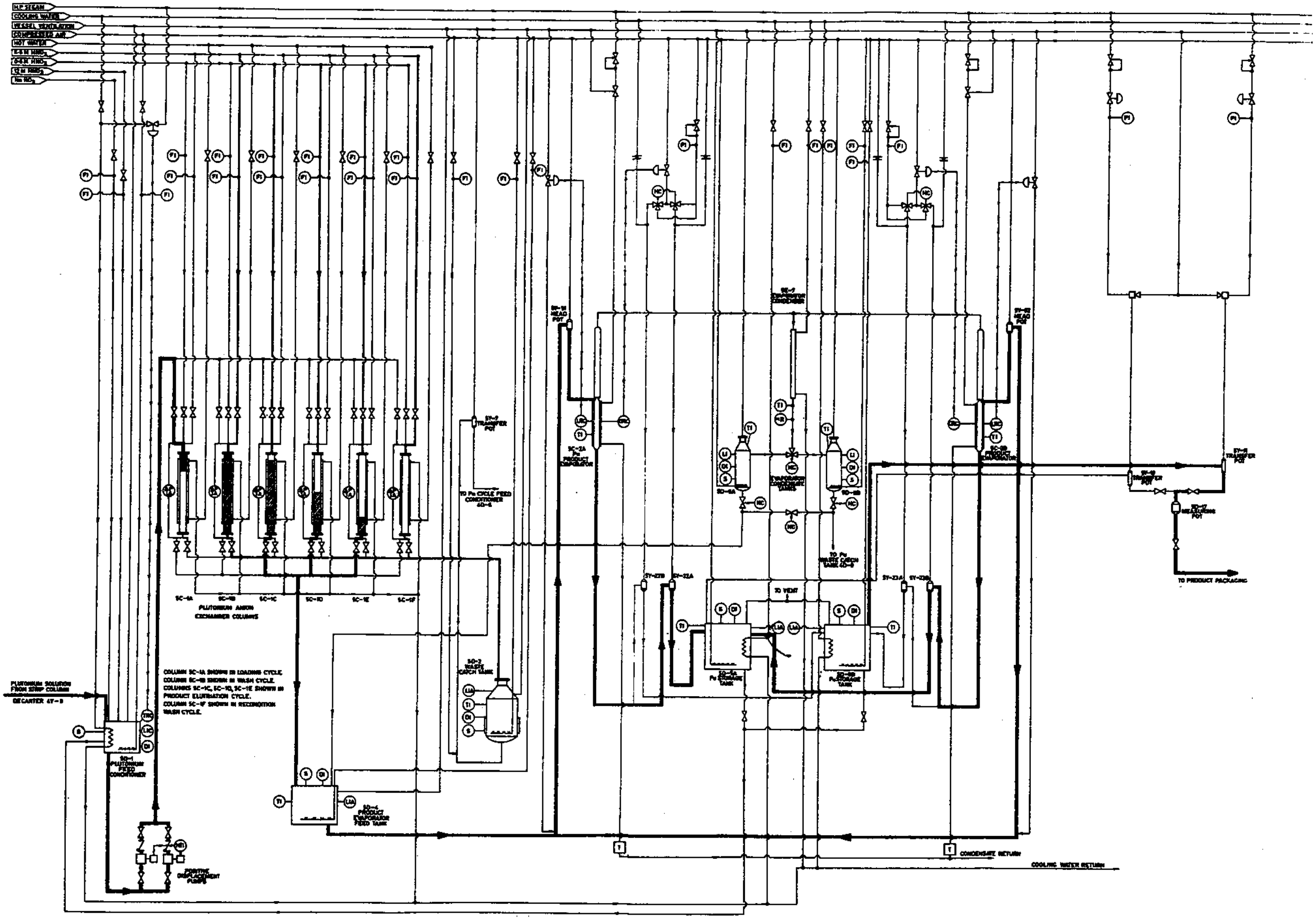


FIGURE 17. ENGINEERING FLOW DIAGRAM - PLUTONIUM ION EXCHANGE AND EVAPORATION

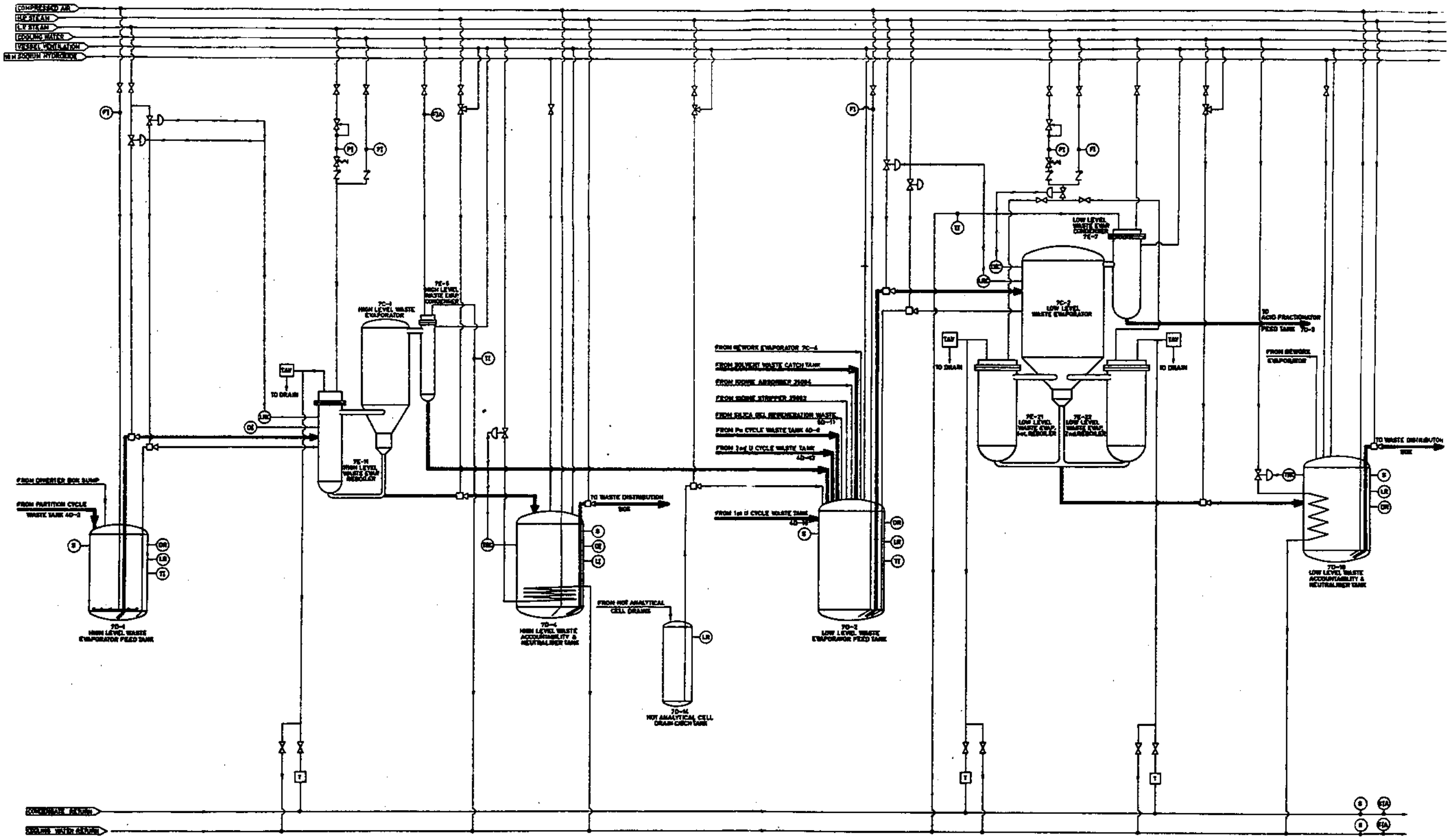


FIGURE 18. ENGINEERING FLOW DIAGRAM - WASTE EVAPORATION

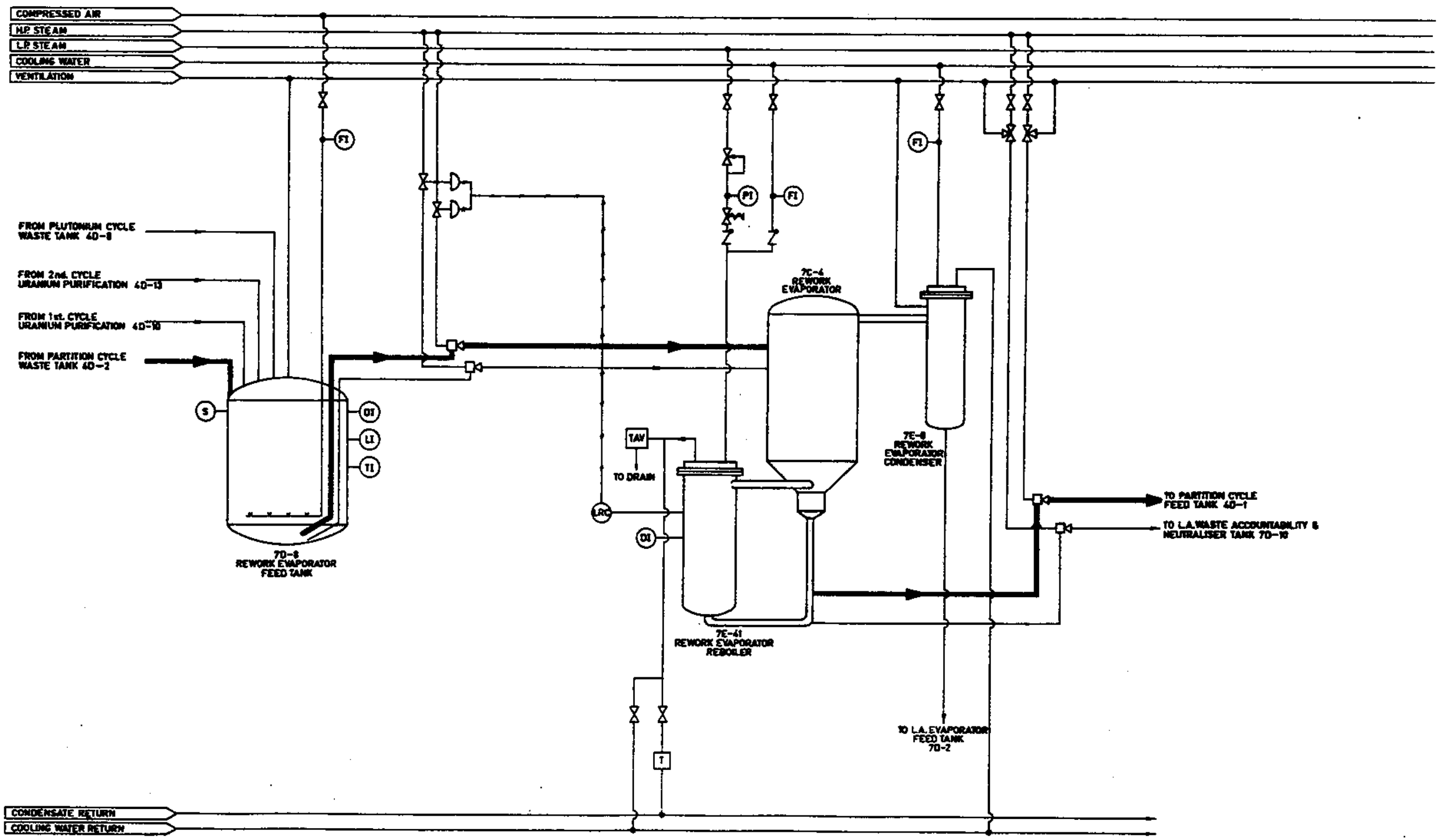


FIGURE 19. ENGINEERING FLOW DIAGRAM - REWORK EVAPORATION

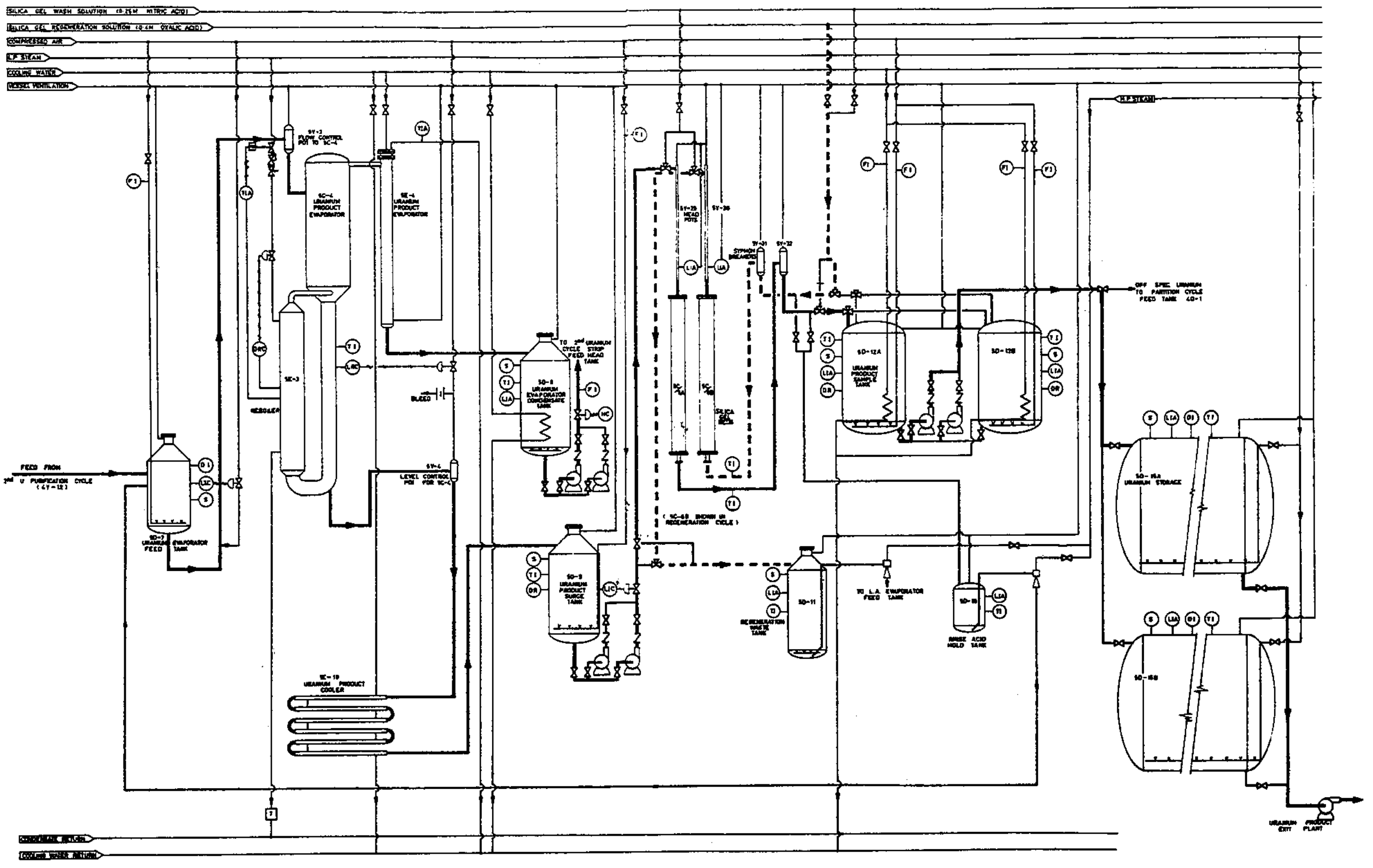


FIGURE 20. ENGINEERING FLOW DIAGRAM - URANIUM EVAPORATION AND SILICA GEL PURIFICATION
]

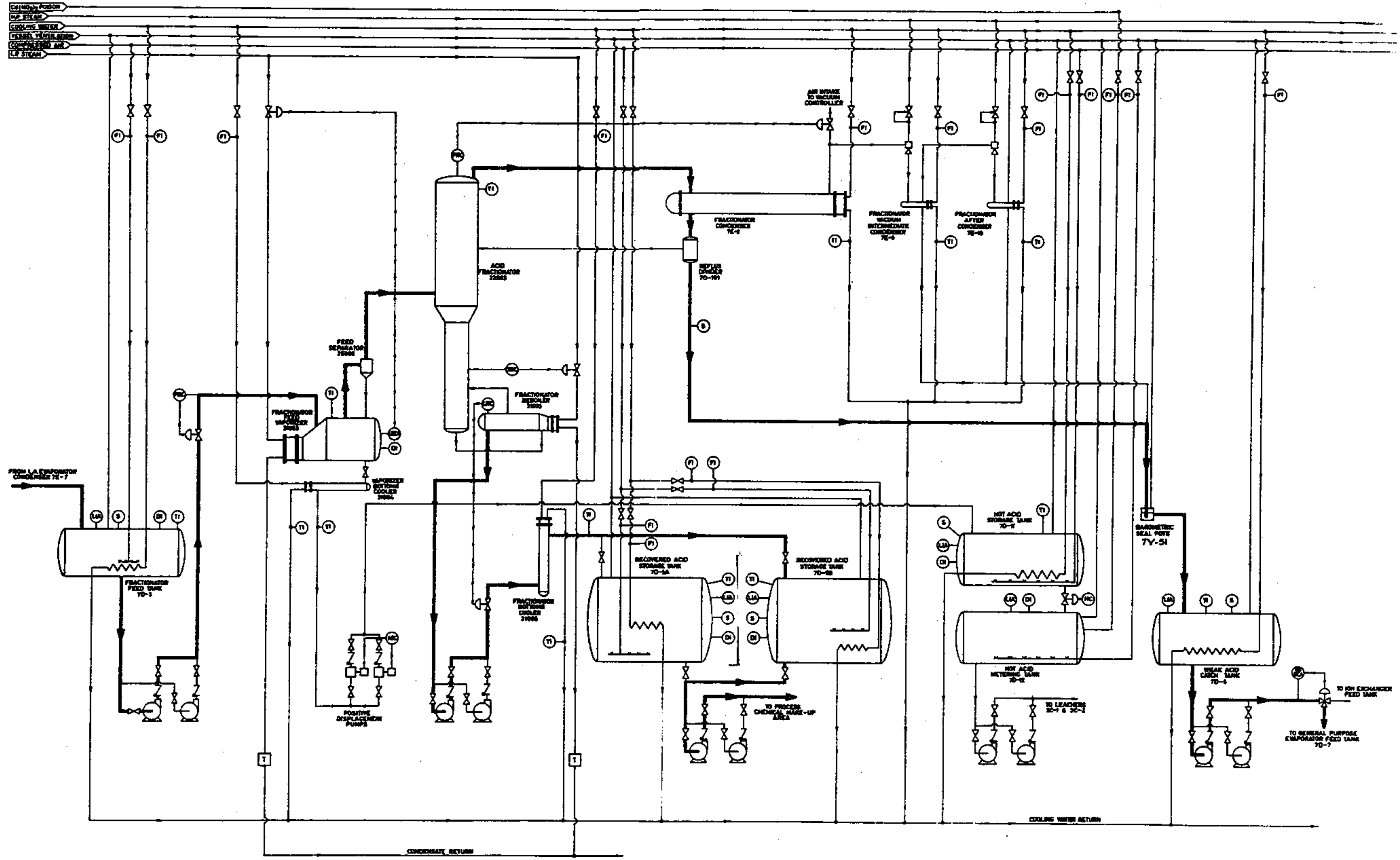


FIGURE 21. ENGINEERING FLOW DIAGRAM - NITRIC ACID RECOVERY

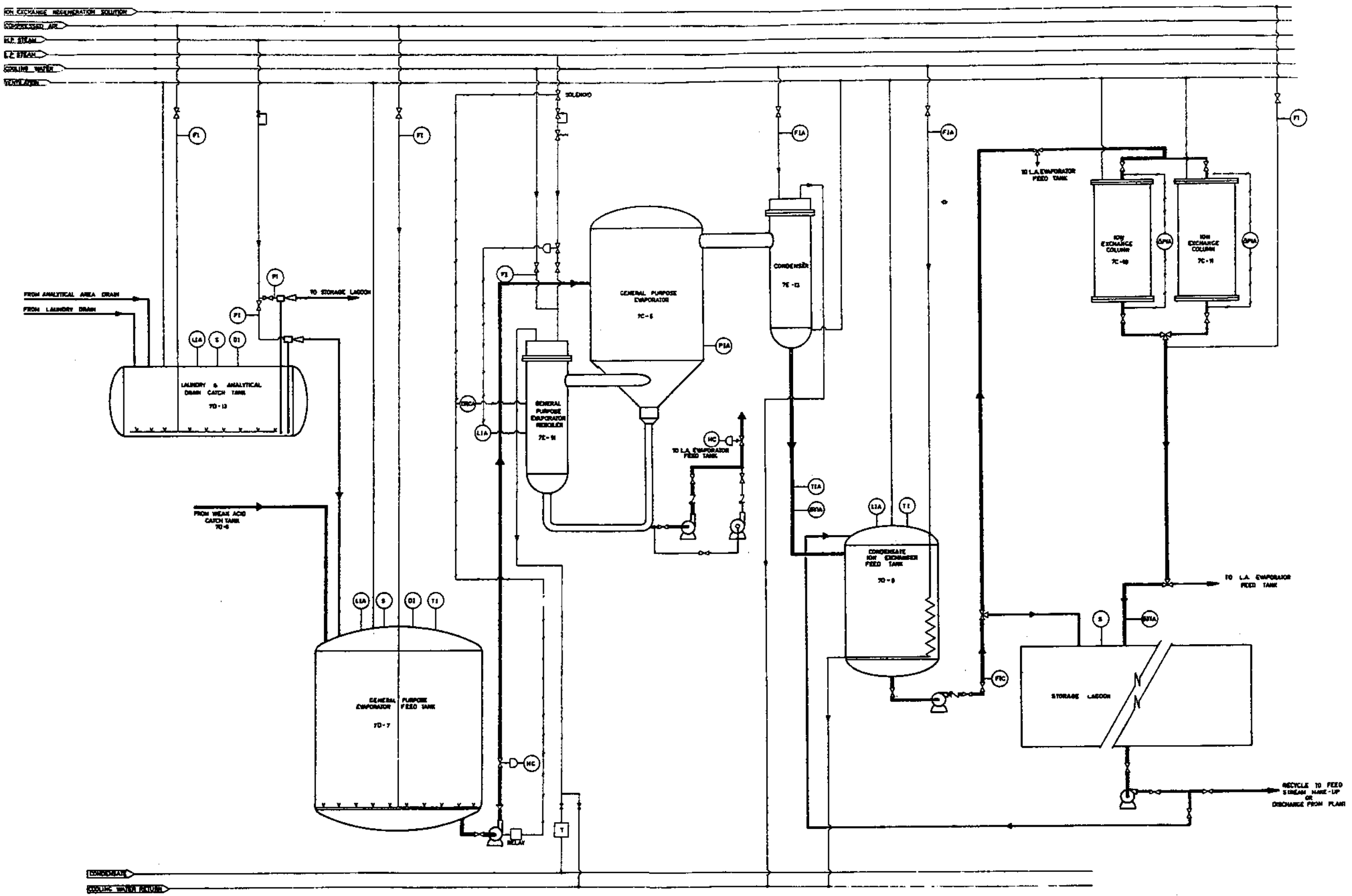


FIGURE 22. ENGINEERING FLOW DIAGRAM - EFFLUENT TREATMENT AND DISCHARGE

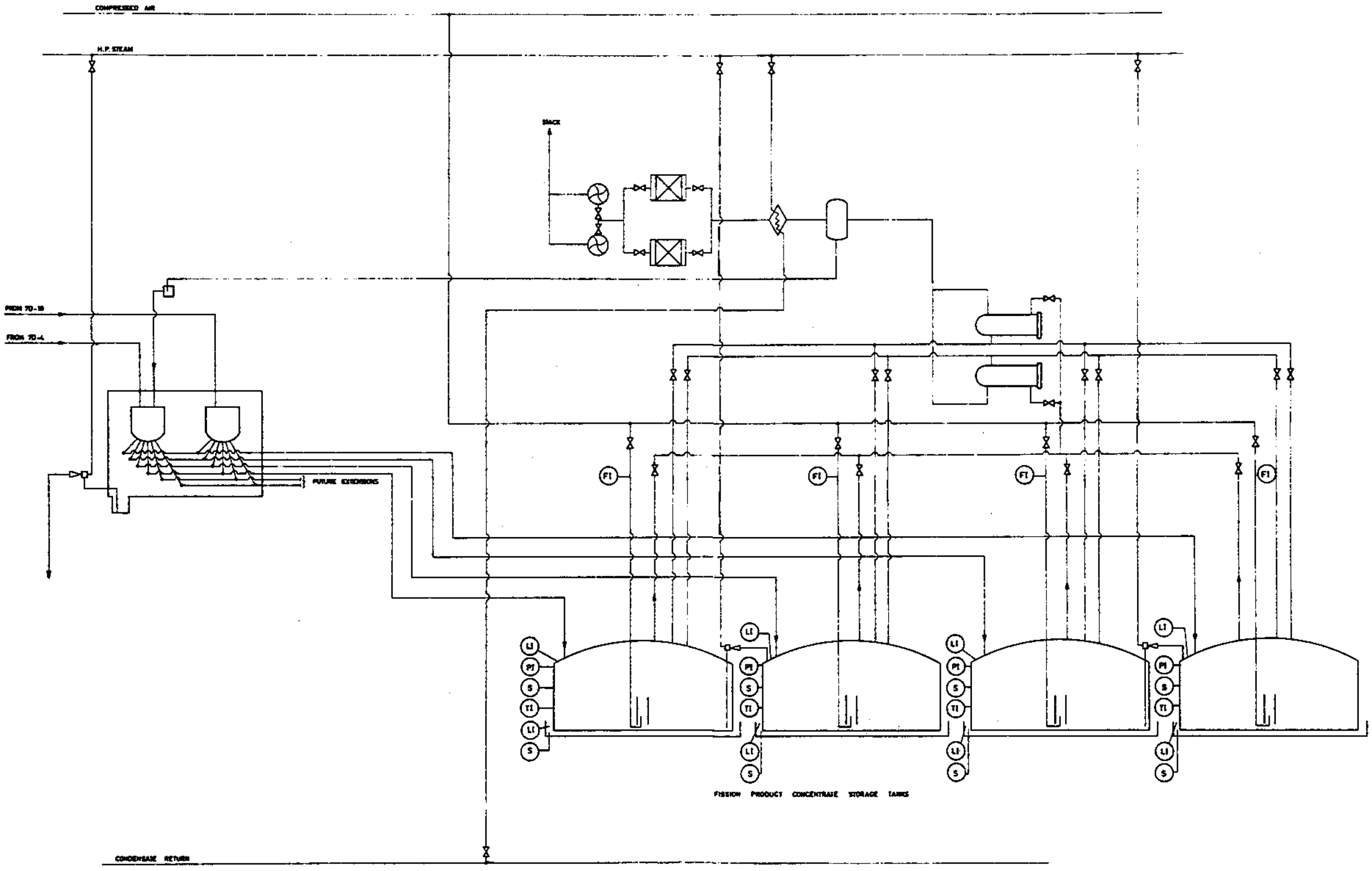


FIGURE 23. ENGINEERING FLOW DIAGRAM - STORAGE OF FISSION PRODUCT WASTES

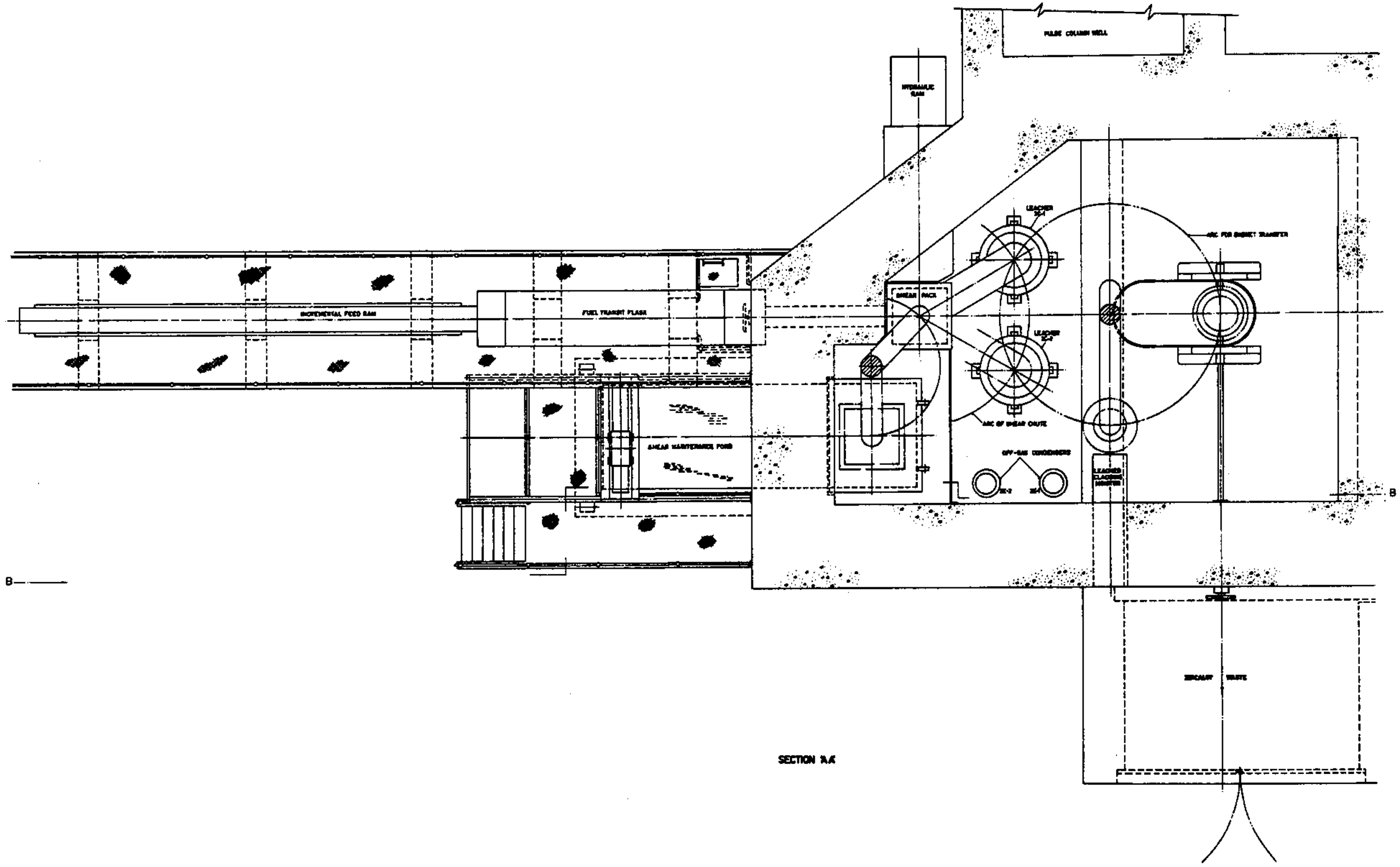
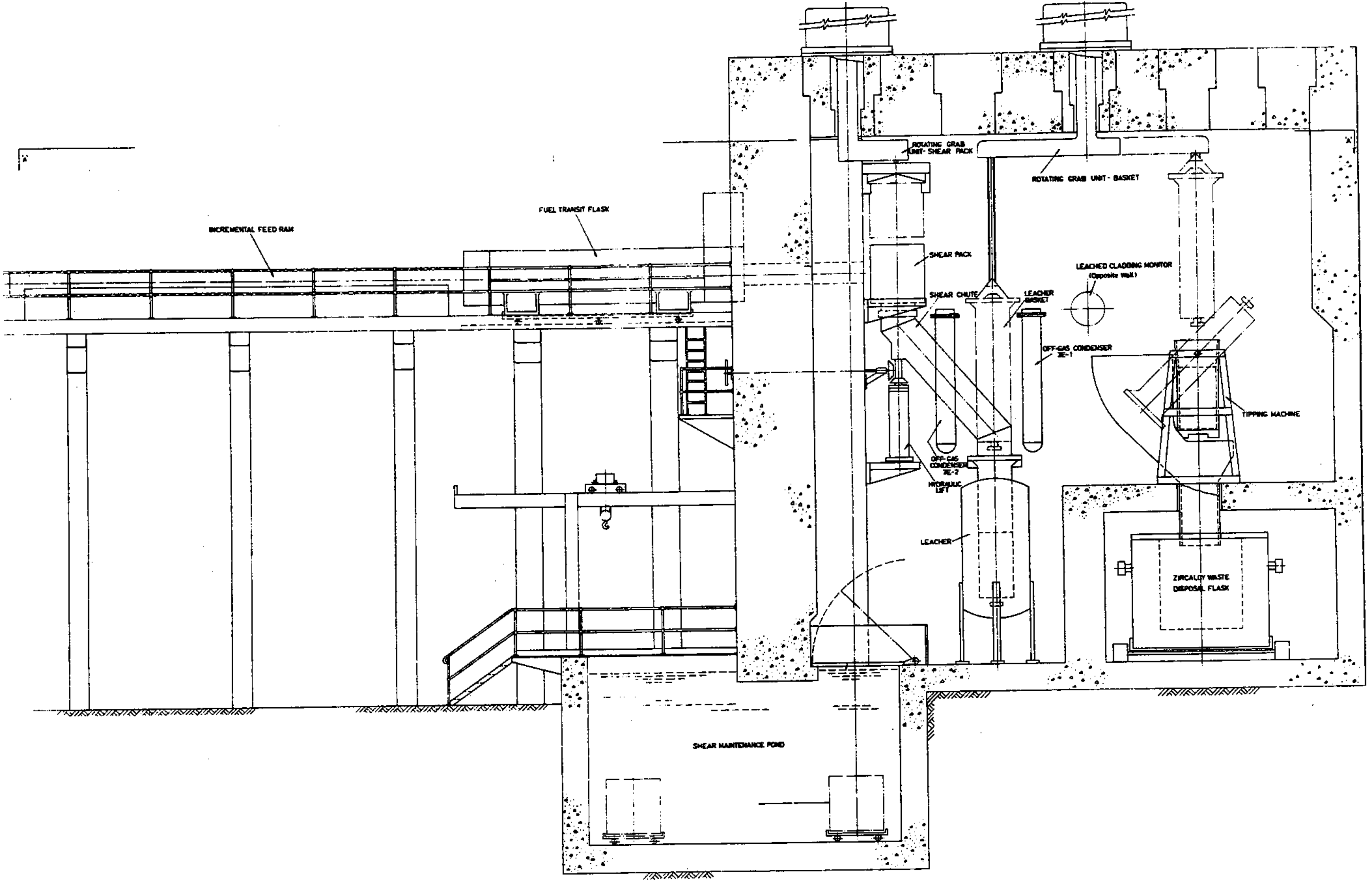


FIGURE 24. EQUIPMENT LAYOUT - SHEAR-LEACH CELL PLAN A-A



SECTION 'B B'

FIGURE 25. EQUIPMENT LAYOUT - SHEAR-LEACH CELL ELEVATION

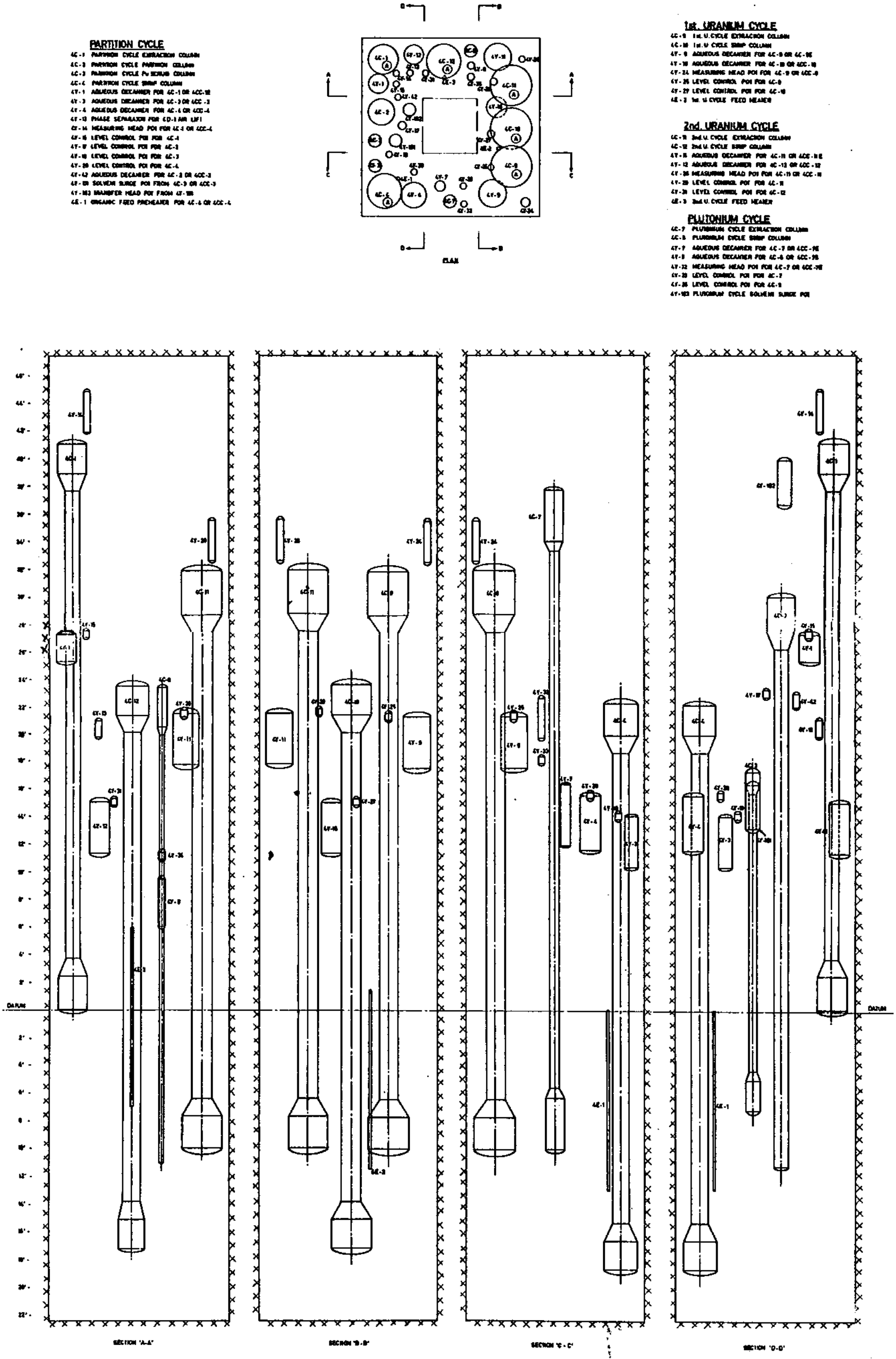


FIGURE 26. EQUIPMENT LAYOUT - PULSED COLUMN WELL

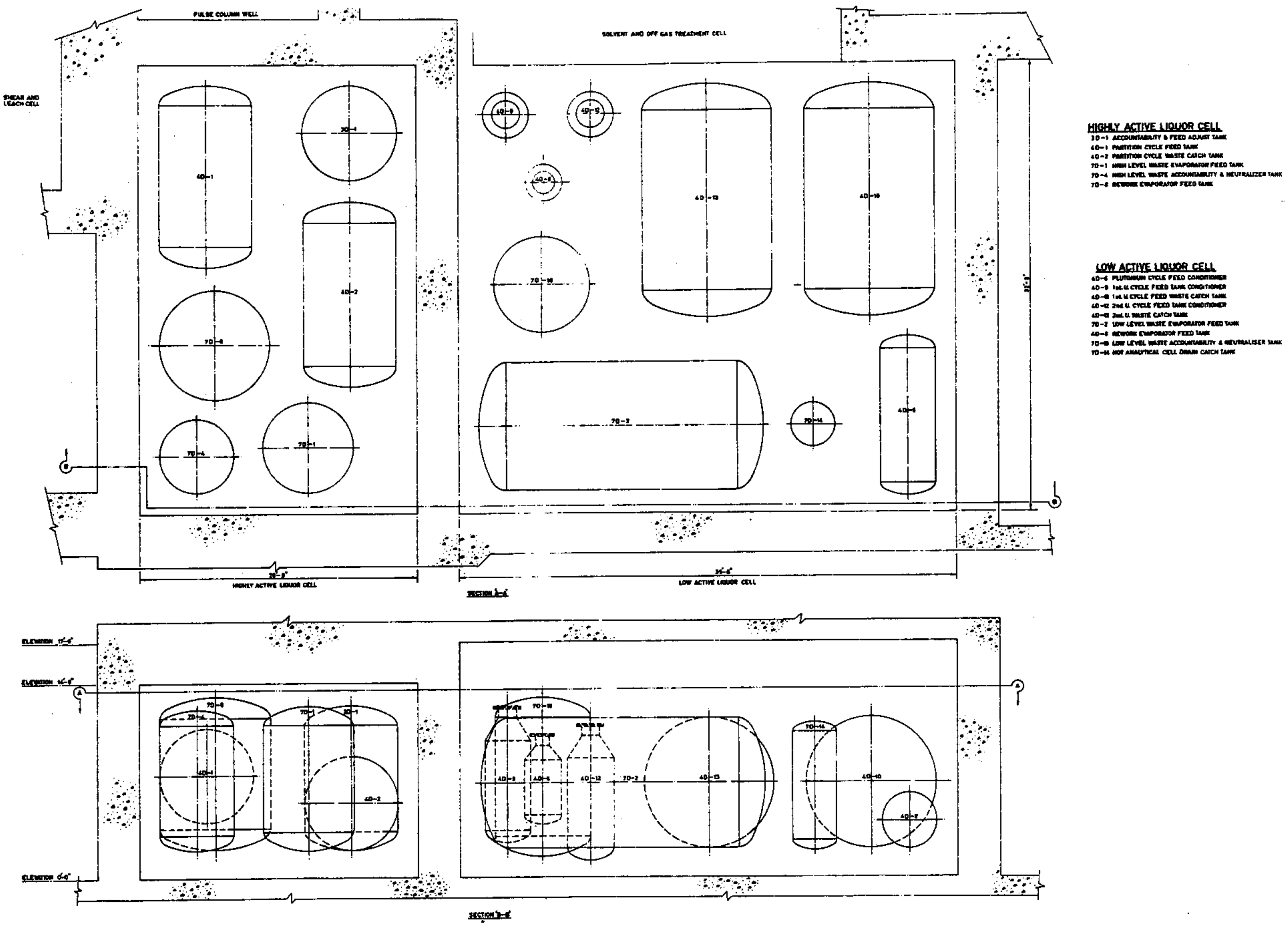


FIGURE 27. EQUIPMENT LAYOUT - ACTIVE LIQUOR CELLS

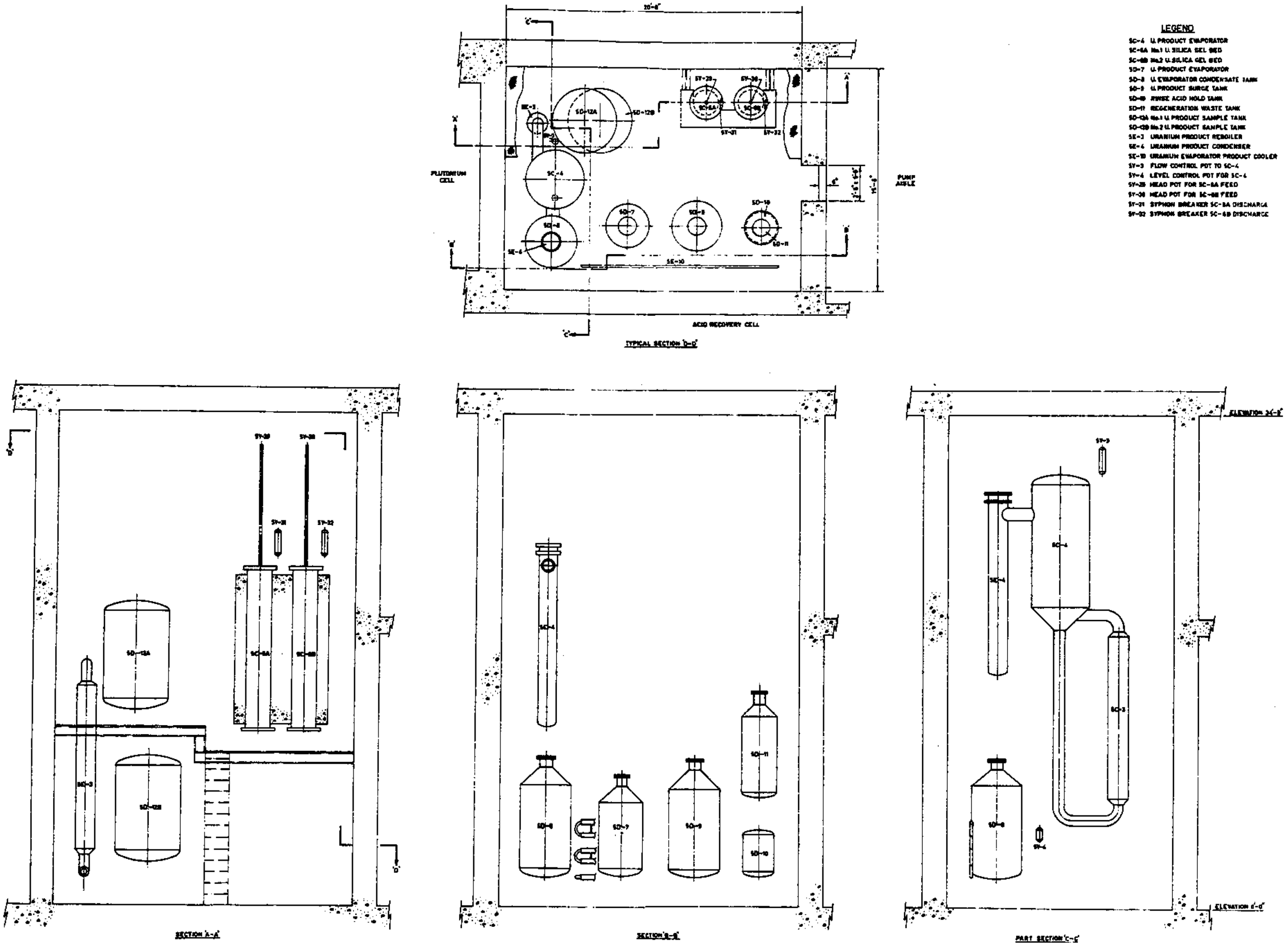
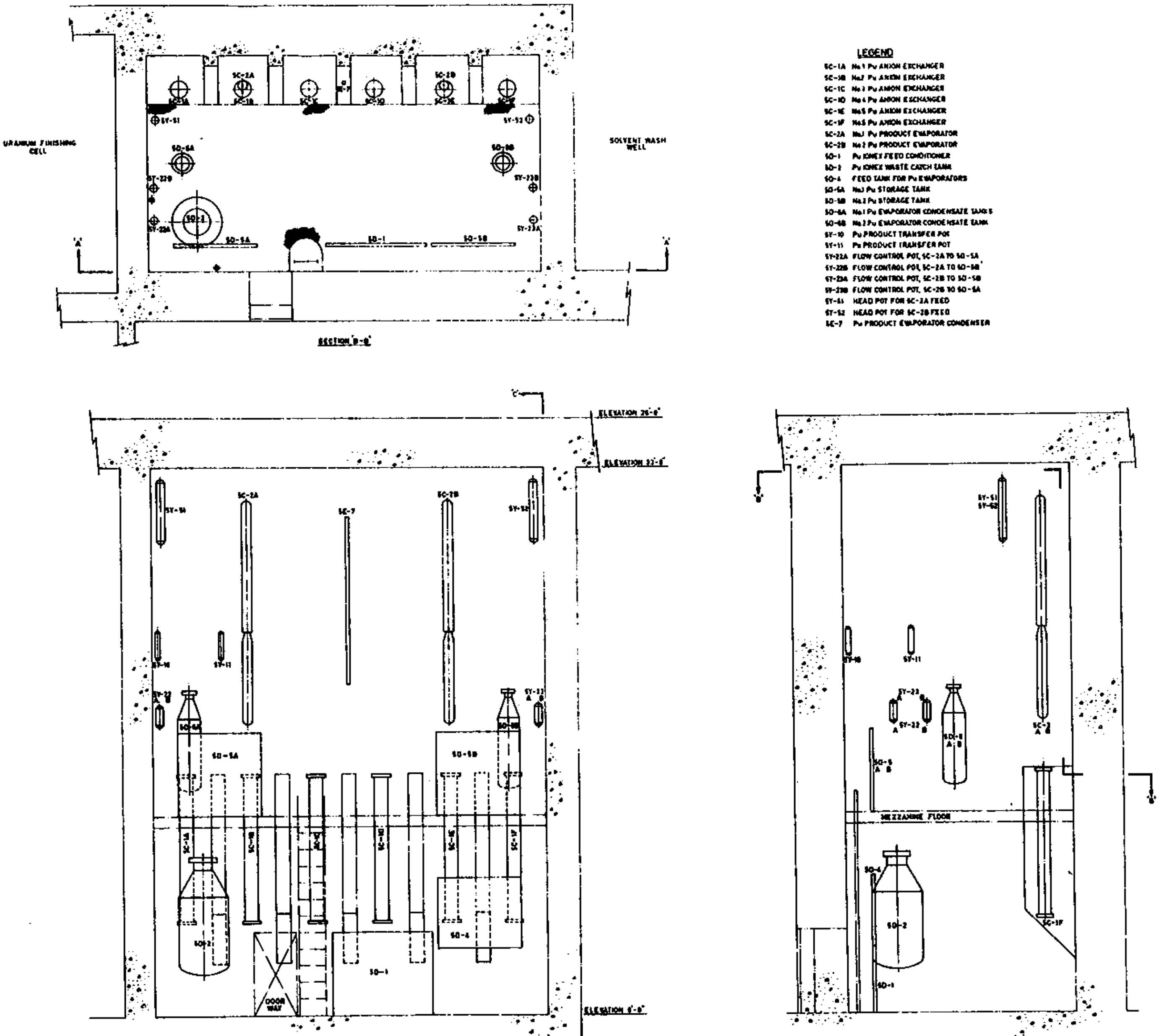


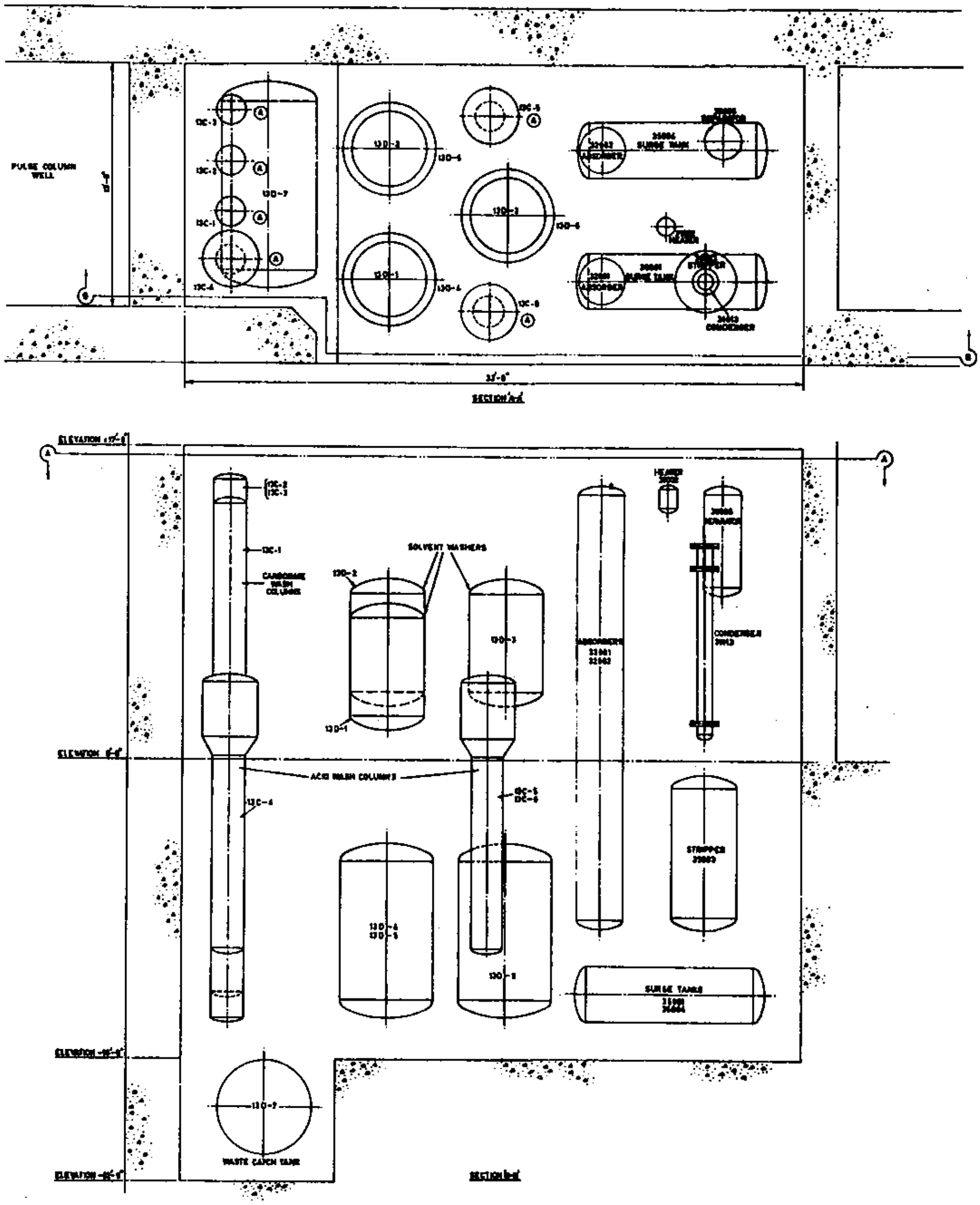
FIGURE 28. EQUIPMENT LAYOUT - URANIUM EVAPORATION AND SILICA GEL PURIFICATION



LEGEND

SC-1A	No.1 Pu ANION EXCHANGER
SC-1B	No.2 Pu ANION EXCHANGER
SC-1C	No.3 Pu ANION EXCHANGER
SC-1D	No.4 Pu ANION EXCHANGER
SC-1E	No.5 Pu ANION EXCHANGER
SC-1F	No.6 Pu ANION EXCHANGER
SC-2A	No.1 Pu PRODUCT EVAPORATOR
SC-2B	No.2 Pu PRODUCT EVAPORATOR
SO-1	Pu IONEX FEED CONDITIONER
SO-2	Pu IONEX WASTE CATCH TANK
SO-4	FEED TANK FOR Pu EVAPORATORS
SO-5A	No.1 Pu STORAGE TANK
SO-5B	No.2 Pu STORAGE TANK
SO-6A	No.1 Pu EVAPORATOR CONDENSATE TANKS
SO-6B	No.2 Pu EVAPORATOR CONDENSATE TANK
SY-10	Pu PRODUCT TRANSFER POT
SY-11	Pu PRODUCT TRANSFER POT
SY-22A	FLOW CONTROL POT, SC-2A TO SO-5A
SY-22B	FLOW CONTROL POT, SC-2A TO SO-5B
SY-23A	FLOW CONTROL POT, SC-2B TO SO-5B
SY-23B	FLOW CONTROL POT, SC-2B TO SO-5A
SY-51	HEAD POT FOR SC-2A FEED
SY-52	HEAD POT FOR SC-2B FEED
SC-7	Pu PRODUCT EVAPORATOR CONDENSER

FIGURE 29. EQUIPMENT LAYOUT - PLUTONIUM ION EXCHANGE AND EVAPORATION



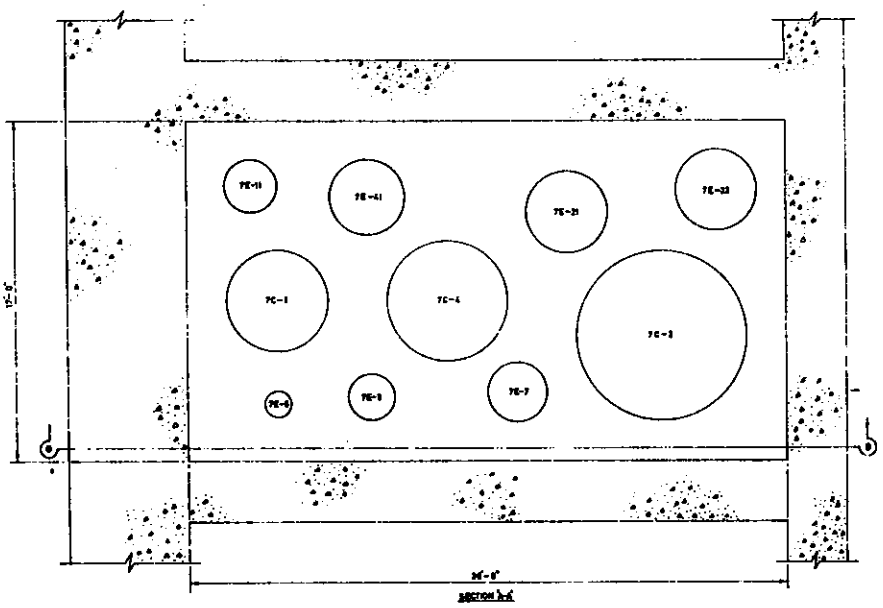
SOLVENT TREATMENT

13C-1	No.1 CARBONATE WASH COLUMN
13C-4	No.1 CARBONATE WASH COLUMN
13D-1	No.1 SOLVENT WASHER
13D-4	No.1 SOLVENT STORAGE TANK
13C-2	No.2 CARBONATE WASH COLUMN
13C-5	No.2 ACID WASH COLUMN
13D-2	No.2 SOLVENT WASHER
13D-5	No.2 SOLVENT STORAGE TANK
13C-3	No.3 CARBONATE WASH COLUMN
13C-6	No.3 ACID WASH COLUMN
13D-3	No.3 SOLVENT WASHER
13D-6	No.3 SOLVENT STORAGE TANK
13D-7	SOLVENT TREATMENT WASTE CATCH TANK

OFF-GAS TREATMENT

32001	DISSOLVER OFF-GAS ABSORBER
32002	IODINE ABSORBER
36001	DISSOLVER OFF-GAS ABSORBER SURGE TANK
32003	IODINE STRIPPER
36004	IODINE ABSORBER SURGE TANK
36006	IODINE ABSORBER SEPARATOR
31002	OFF-GAS HEATER
31013	IODINE STRIPPER CONDENSER

FIGURE 30. EQUIPMENT LAYOUT - SOLVENT WASHING AND OFF-GAS TREATMENT



HIGH LEVEL WASTES

7C-1 HIGH LEVEL WASTE EVAPORATOR
7E-11 HIGH LEVEL WASTE EVAPORATOR REBOILER
7E-5 HIGH LEVEL WASTE EVAPORATOR CONDENSER

LOW LEVEL WASTES

7C-2 LOW LEVEL WASTE EVAPORATOR
7E-21 LOW LEVEL WASTE EVAPORATOR FIRST REBOILER
7E-22 LOW LEVEL WASTE EVAPORATOR SECOND REBOILER
7E-7 LOW LEVEL WASTE EVAPORATOR CONDENSER

REWORK

7C-4 REWORK EVAPORATOR
7E-41 REWORK EVAPORATOR REBOILER
7E-3 REWORK EVAPORATOR CONDENSER

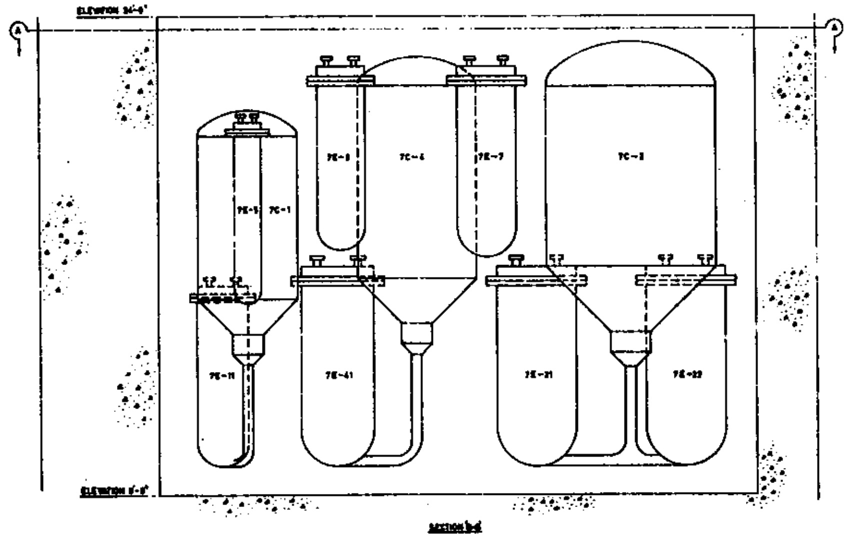


FIGURE 31. EQUIPMENT LAYOUT - WASTE EVAPORATION

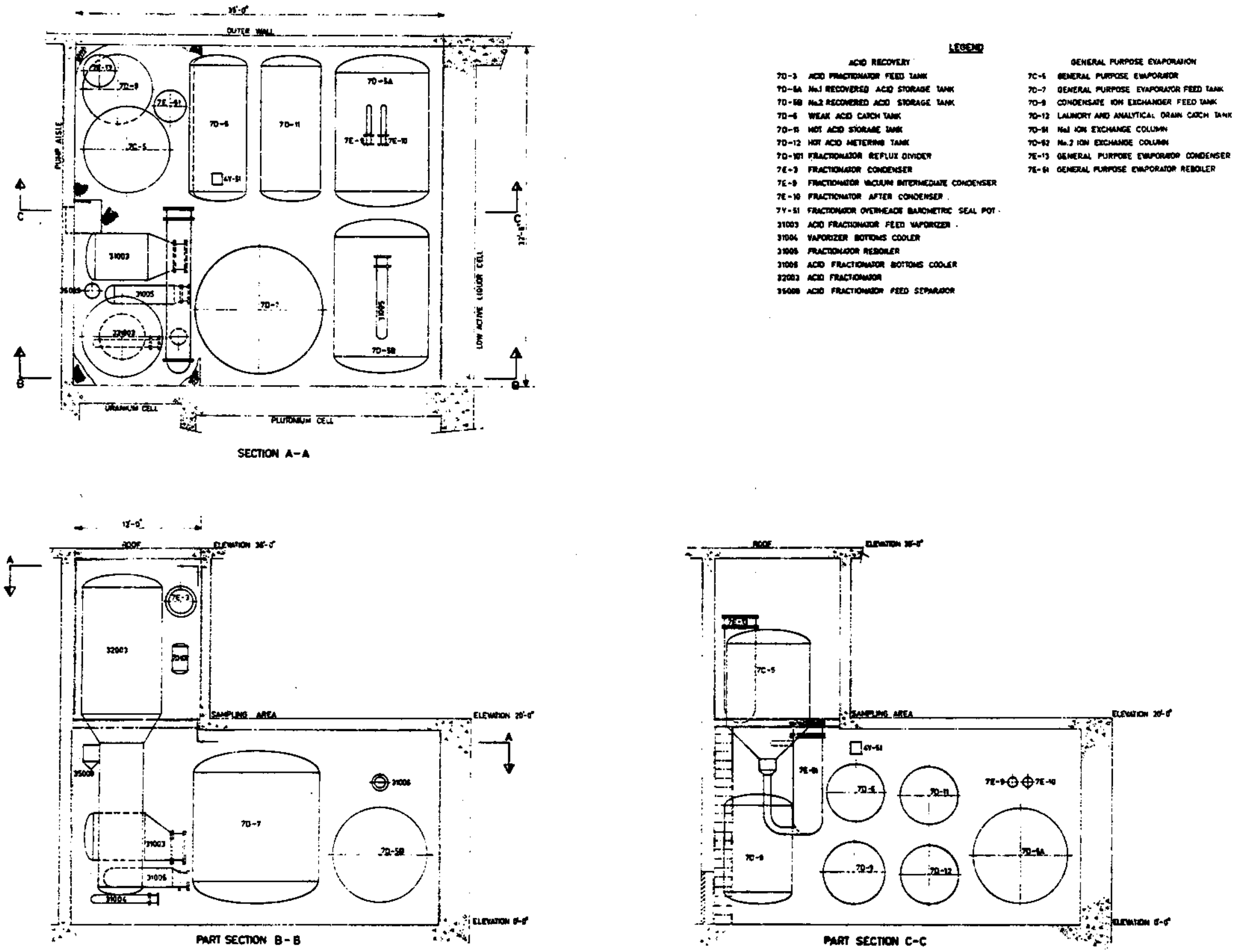
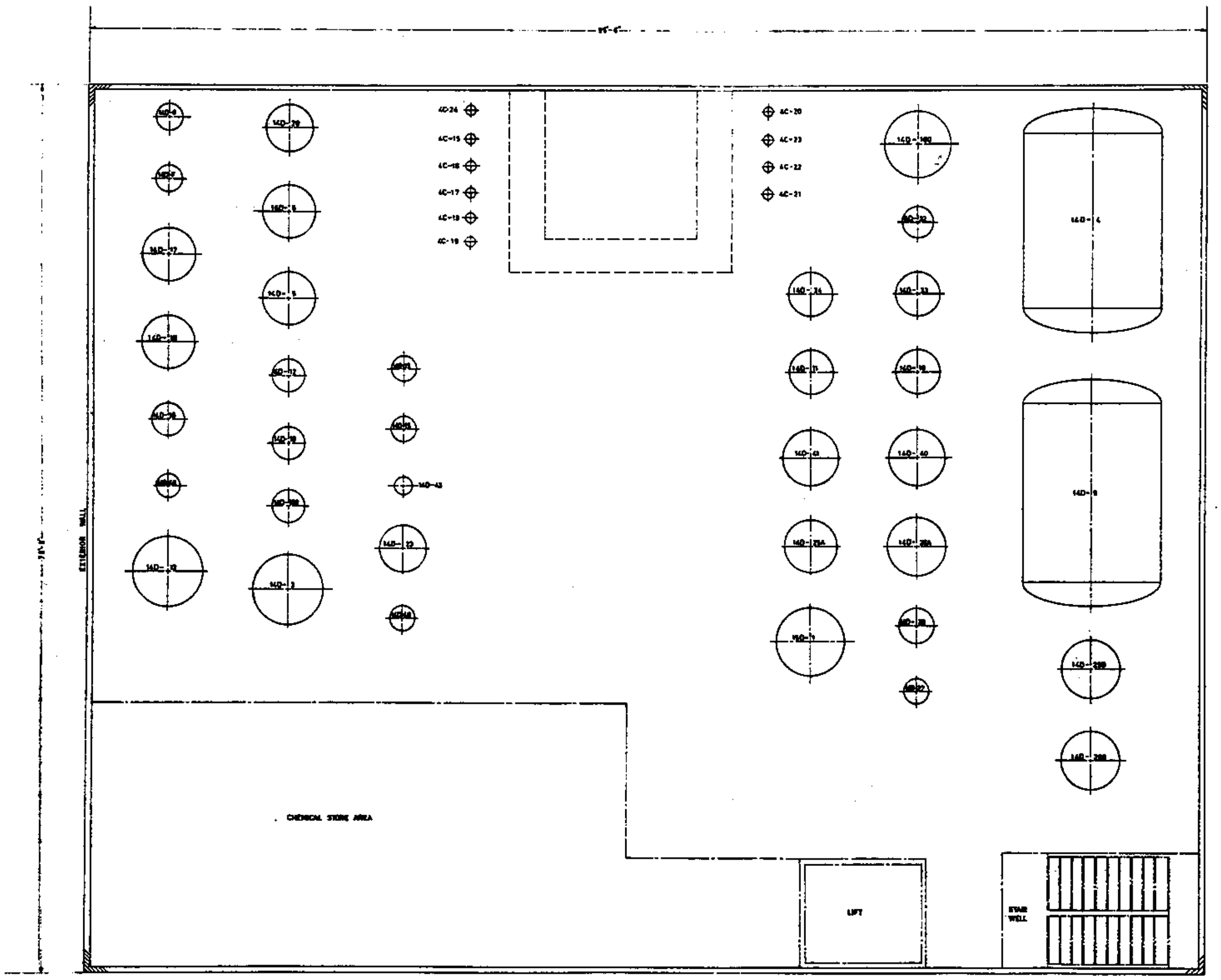


FIGURE 32. EQUIPMENT LAYOUT - ACID RECOVERY CELL



LEGEND	
4C-15	AIR RESERVOIR FOR 4C-1
4C-16	AIR RESERVOIR FOR 4C-2
4C-17	AIR RESERVOIR FOR 4C-3
4C-18	AIR RESERVOIR FOR 4C-4
4C-19	AIR RESERVOIR FOR 4C-7
4C-20	AIR RESERVOIR FOR 4C-8
4C-21	AIR RESERVOIR FOR 4C-9
4C-22	AIR RESERVOIR FOR 4C-10
4C-23	AIR RESERVOIR FOR 4C-11
4C-24	AIR RESERVOIR FOR 4C-12
14D-3	CONC. NITRIC ACID DAY TANK
14D-4	STRIP ACID DAY TANK
14D-5	PARTITION SOLUTION MIX TANK
14D-6	PARTITION SOLUTION RUN TANK
14D-7	PLUTONIUM STRIP MIX TANK
14D-8	PLUTONIUM STRIP RUN TANK
14D-9	STRIP ACID MIX TANK
14D-10	1st. U. CYCLE SCRUB MIX TANK
14D-11	1st. U. CYCLE SCRUB RUN TANK
14D-12	FERROUS SULFAMATE MIX TANK
14D-13	U. CYCLES FEED ACID TANK
14D-15	SODIUM NITRITE RUN TANK
14D-16	PARTITION CYCLE MIX TANK
14D-17	PARTITION CYCLE SCRUB RUN TANK
14D-18	PLUTONIUM CYCLE SCRUB MIX TANK
14D-19	PLUTONIUM CYCLE SCRUB RUN TANK
14D-20	PARTITION CYCLE FEED ACID TANK
14D-22	PU IONEX WASH MIX & RUN TANK
14D-25A	No.1 CAUSTIC DILUTION TANK
14D-25B	No.2 CAUSTIC DILUTION TANK
14D-26A	No.1 CARBONATE ACID TANK
14D-26B	No.2 CARBONATE ACID TANK
14D-27	SILICA-GEL REGENERATION WASH TANK
14D-28	TBP SOLVENT MIX & RUN TANK
14D-32	POISON MIX TANK
14D-33	2nd U. CYCLE SCRUB MIX TANK
14D-34	2nd U. CYCLE SCRUB RUN TANK
14D-40	U. CYCLES INTERMEDIATE SCRUB MIX TANK
14D-41	U. CYCLES INTERMEDIATE SCRUB RUN TANK
14D-43	PU IONEX SODIUM NITRITE RUN TANK
14D-46	PU IONEX ELUTRIANT MIX TANK
14D-100	LEACHER FEED ACID MIX TANK
14D-102	SILICA-GEL WASH TANK
15D-1	DECONTAMINATION SOLUTION MIX TANK

FIGURE 33. EQUIPMENT LAYOUT - COLD CHEMICAL ROOM

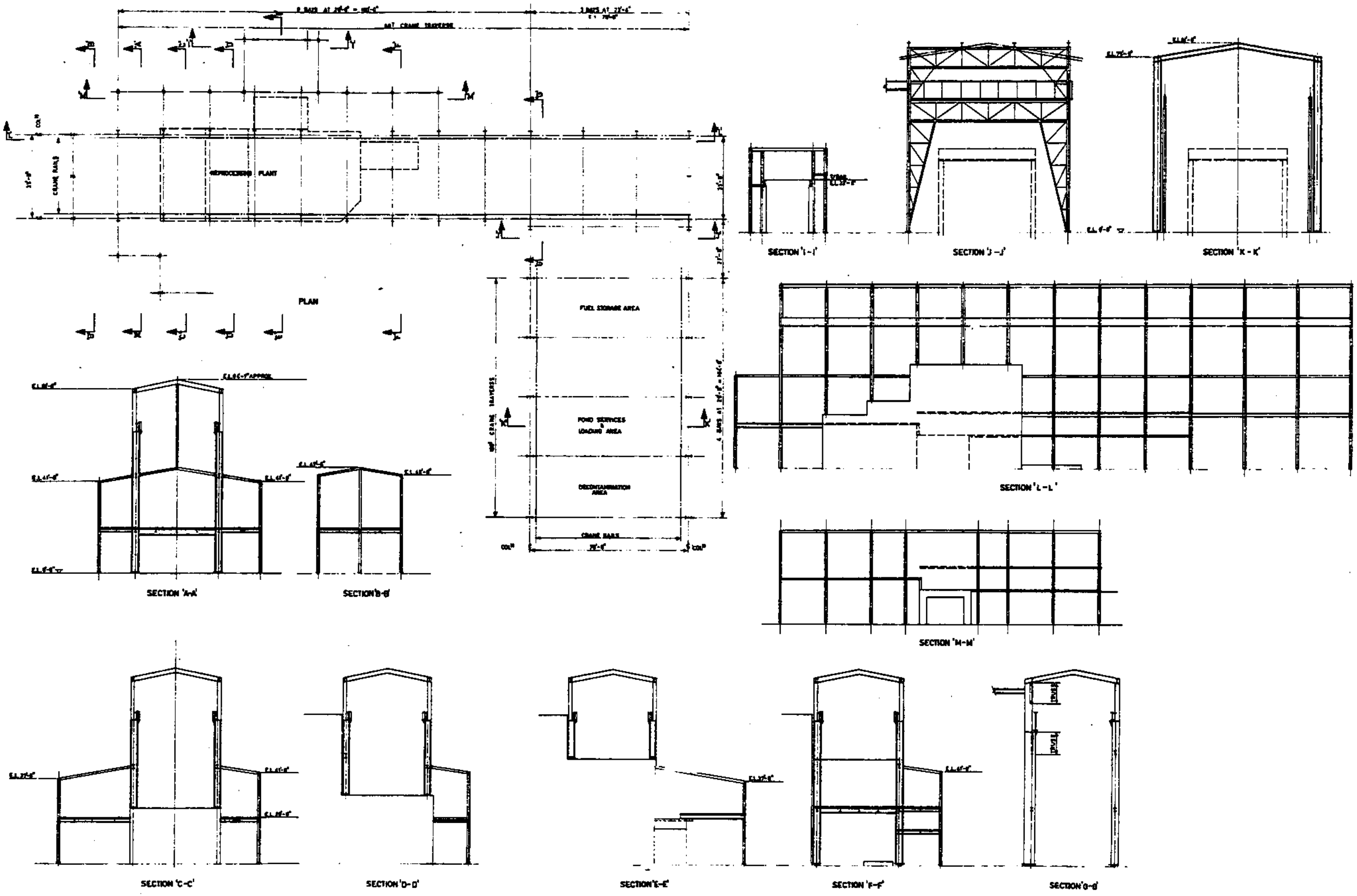


FIGURE 34. BUILDING DESIGN - CRANE SUPPORTING BUILDING

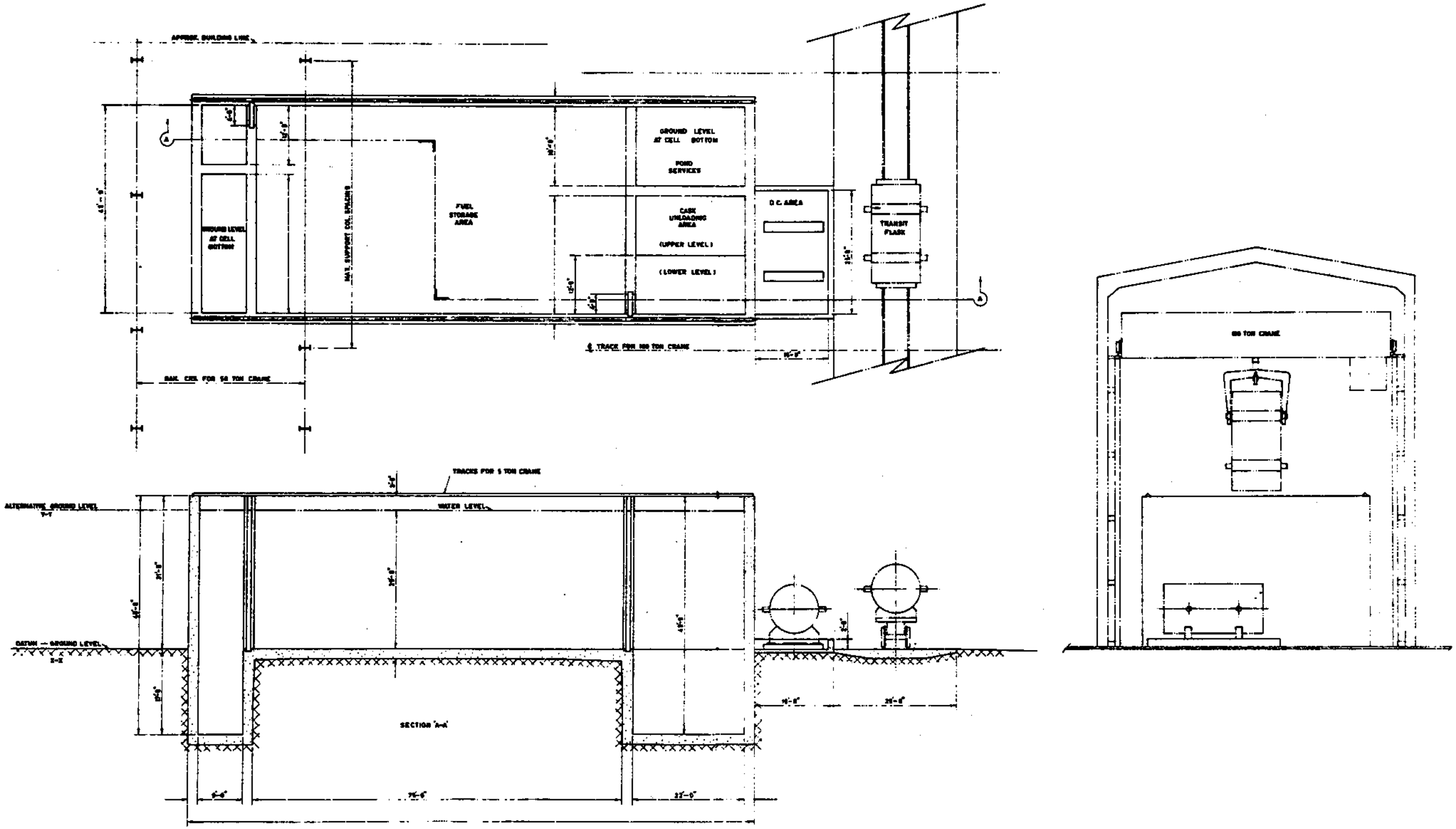
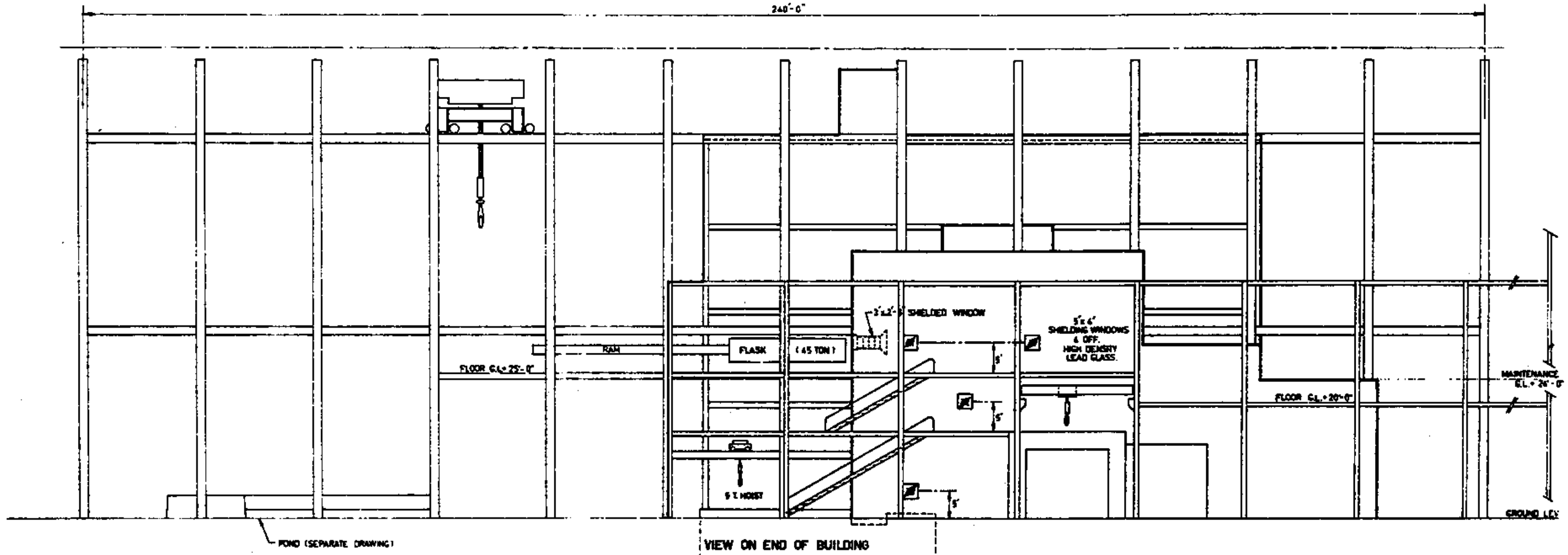
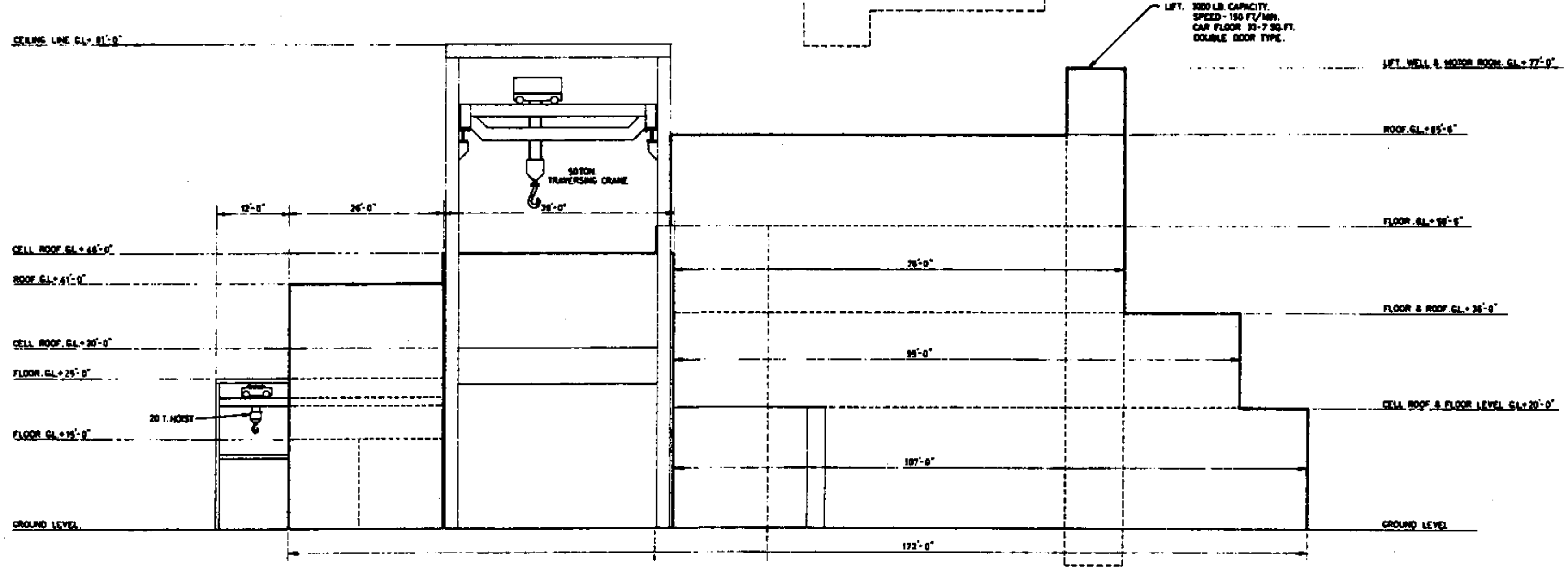


FIGURE 35. BUILDING DESIGN - FUEL STORAGE AREA

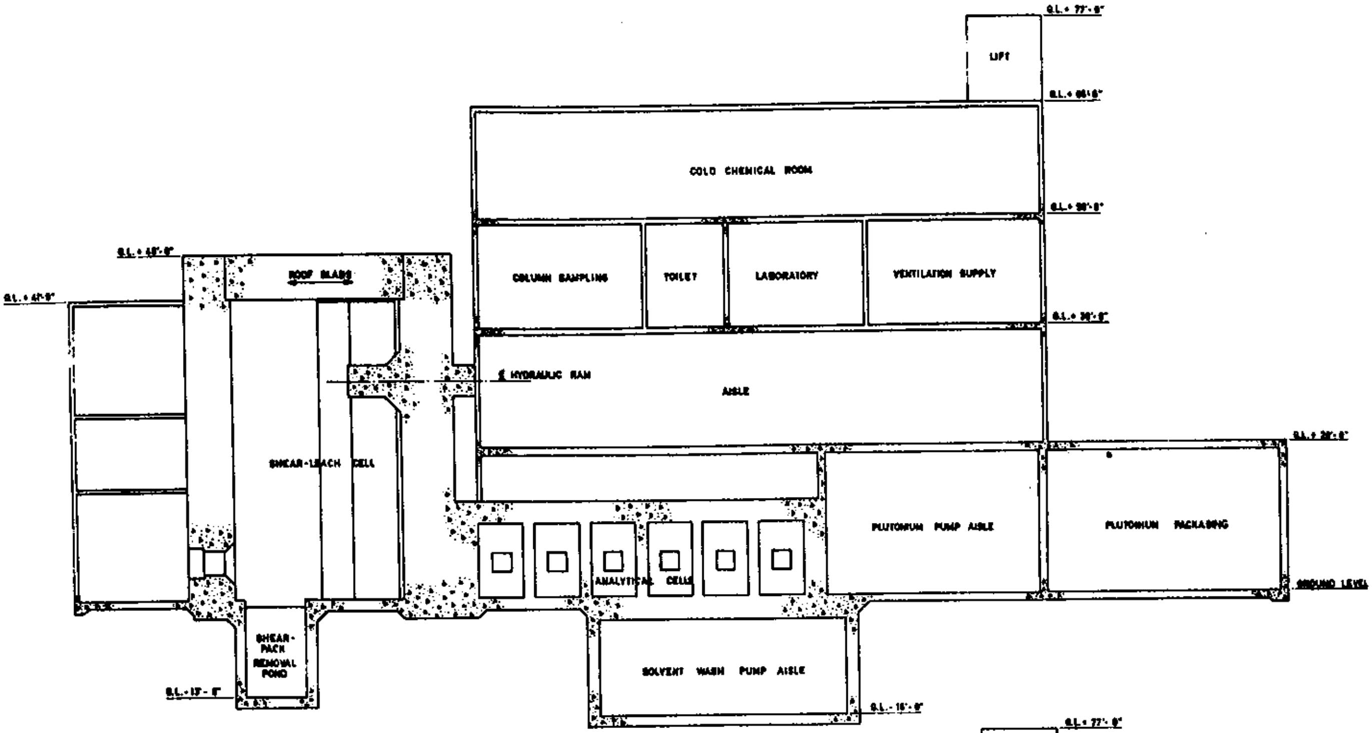


VIEW ON END OF BUILDING
(CLADDING REMOVED)

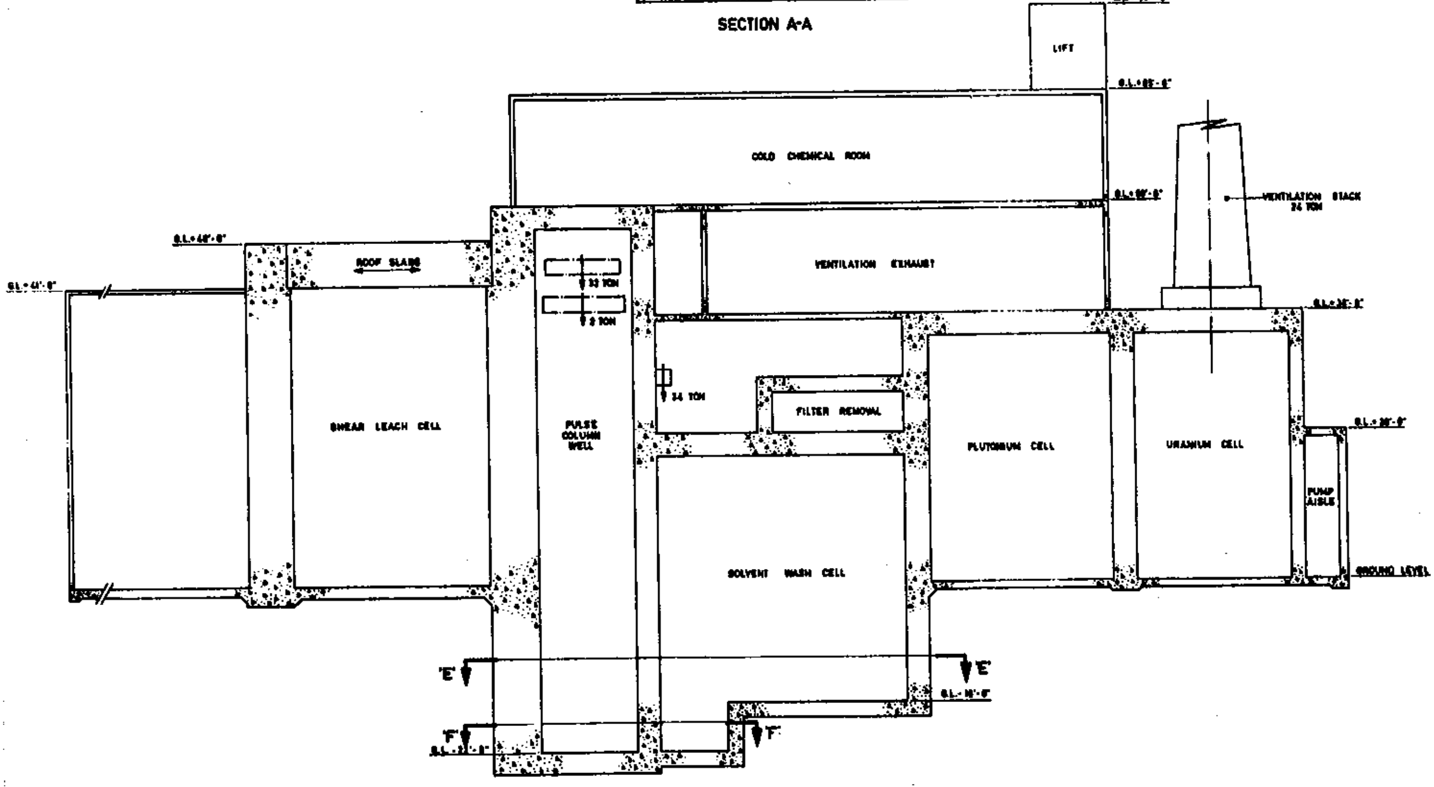


SIDE VIEW OF BUILDING

FIGURE 36. BUILDING DESIGN - BUILDING ELEVATIONS

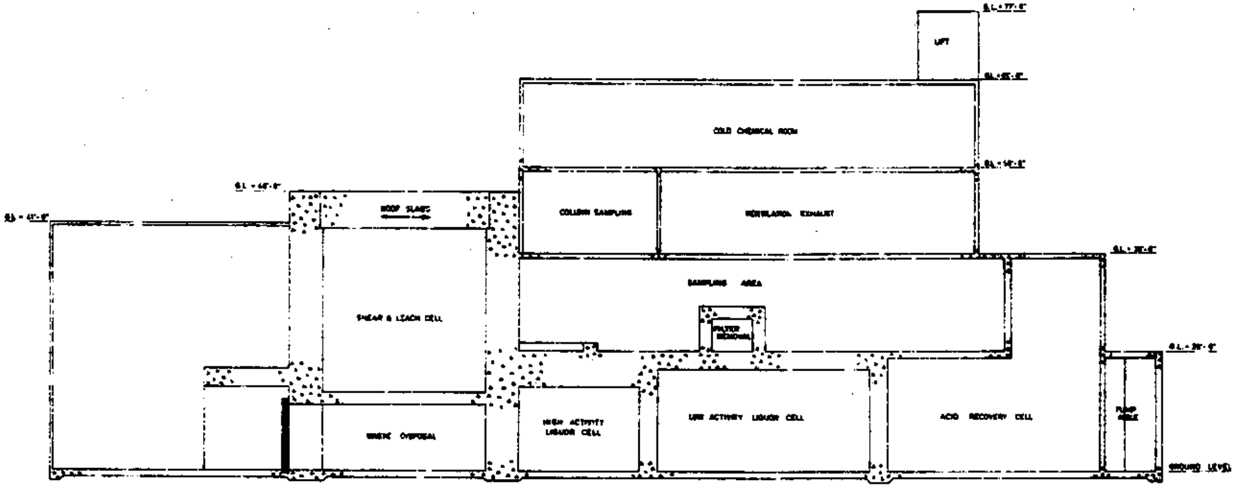


SECTION A-A

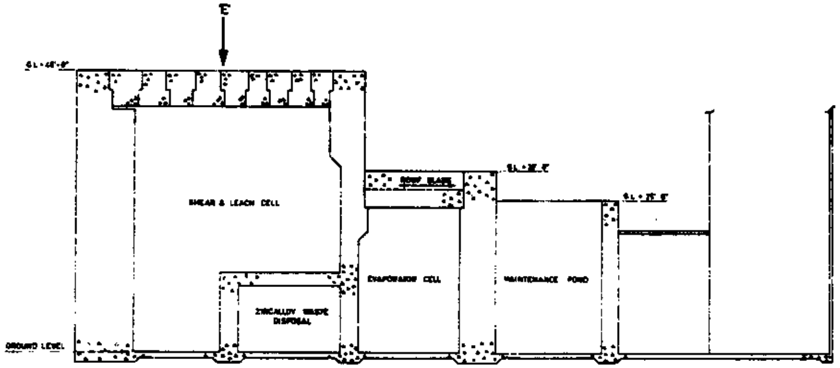


SECTION B-B

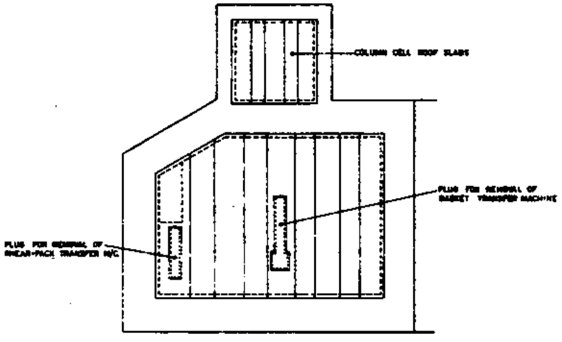
FIGURE 37. BUILDING DESIGN - BUILDING SECTIONS A-A, B-B



SECTION 'C-C'

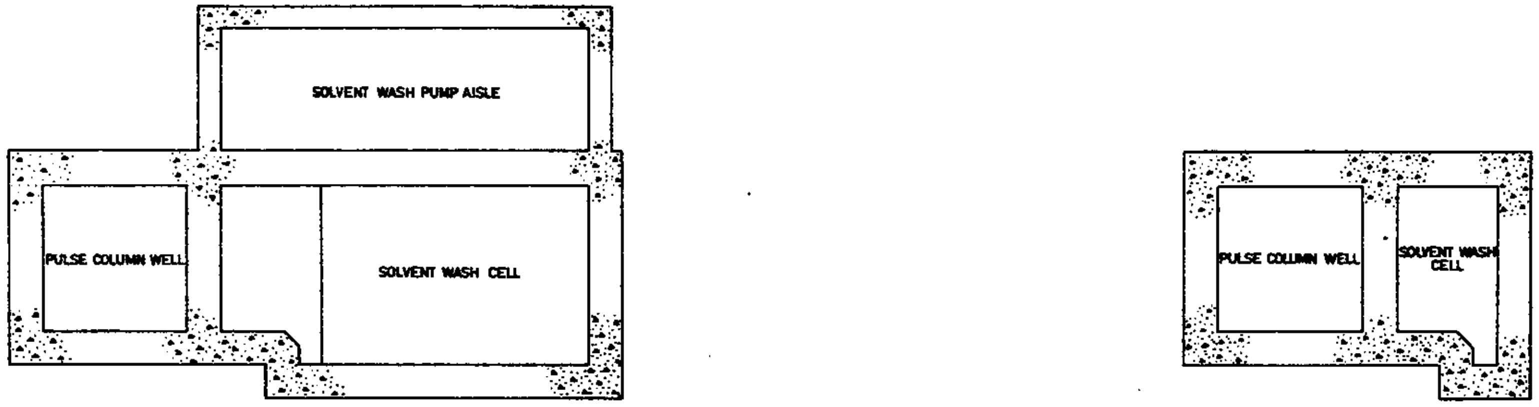


SECTION 'D-D'



VIEW IN DIRECTION OF ARROW 'E'

FIGURE 38. BUILDING DESIGN - BUILDING SECTIONS C-C, D-D



SECTION 'EE'

SECTION 'FF'

FIGURE 39. BUILDING DESIGN - BUILDING SECTIONS E-E,F-F

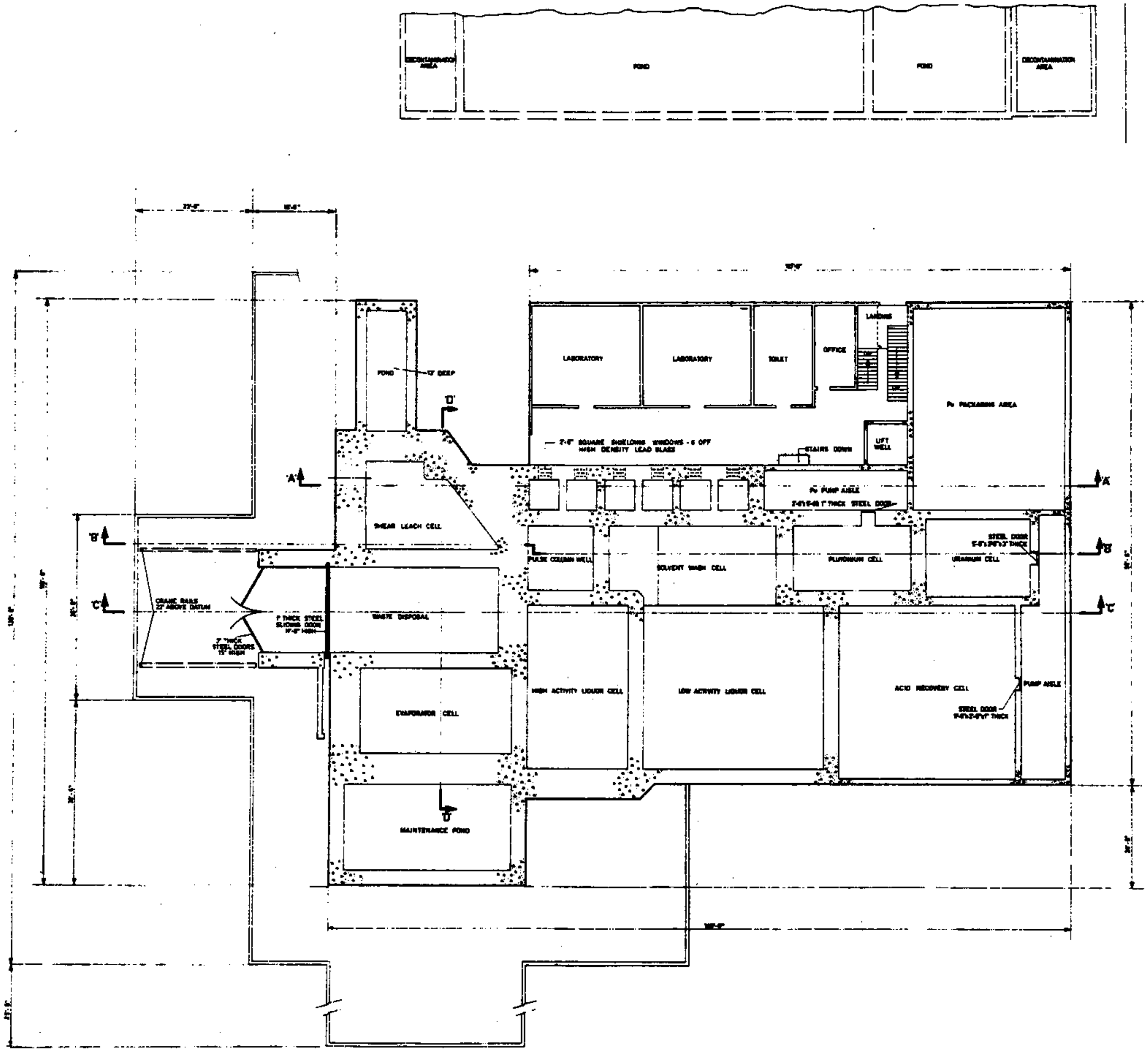


FIGURE 40. BUILDING DESIGN - PLAN 0 ft

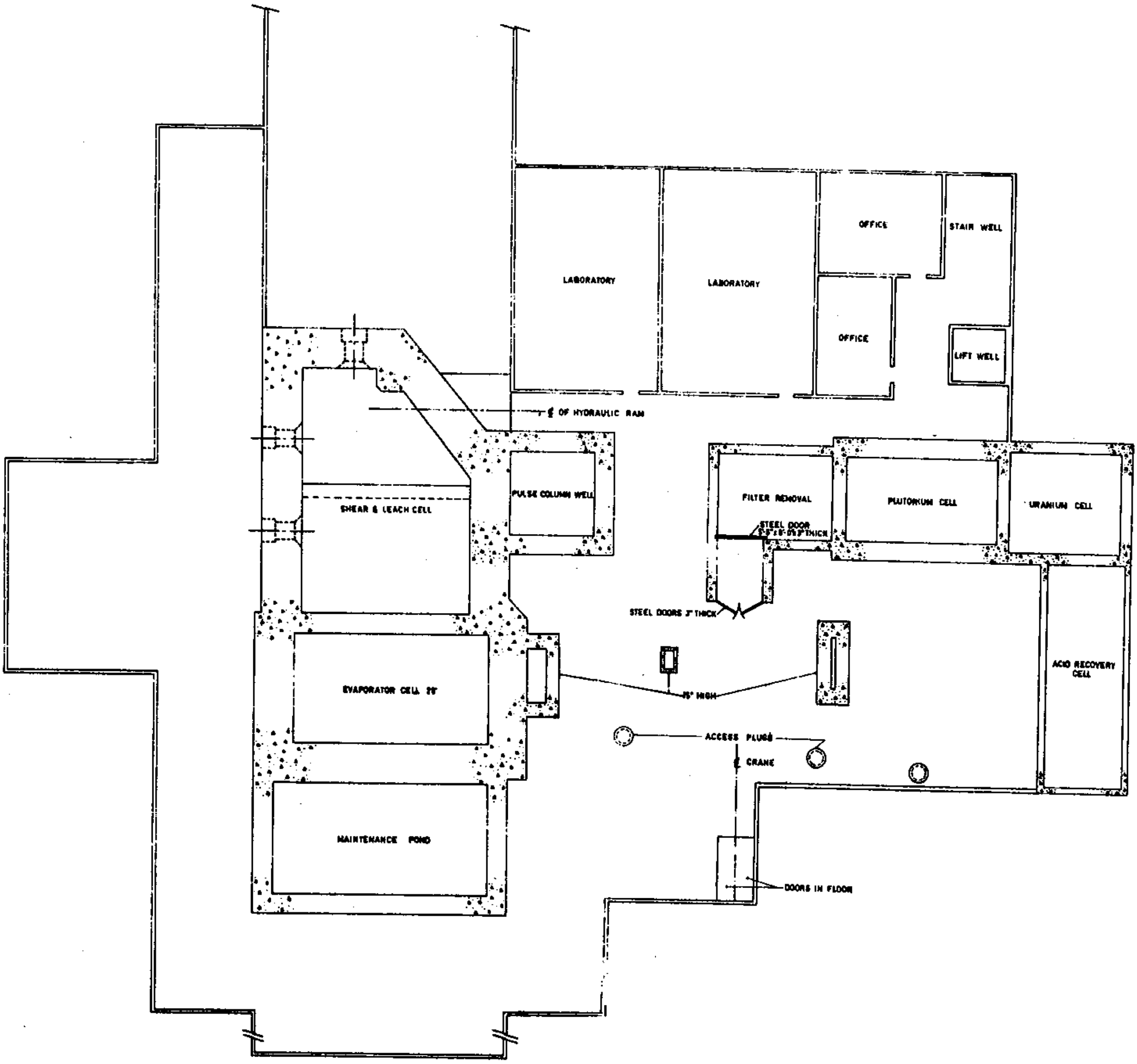


FIGURE 41. BUILDING DESIGN - PLAN 20 ft TO 36 ft

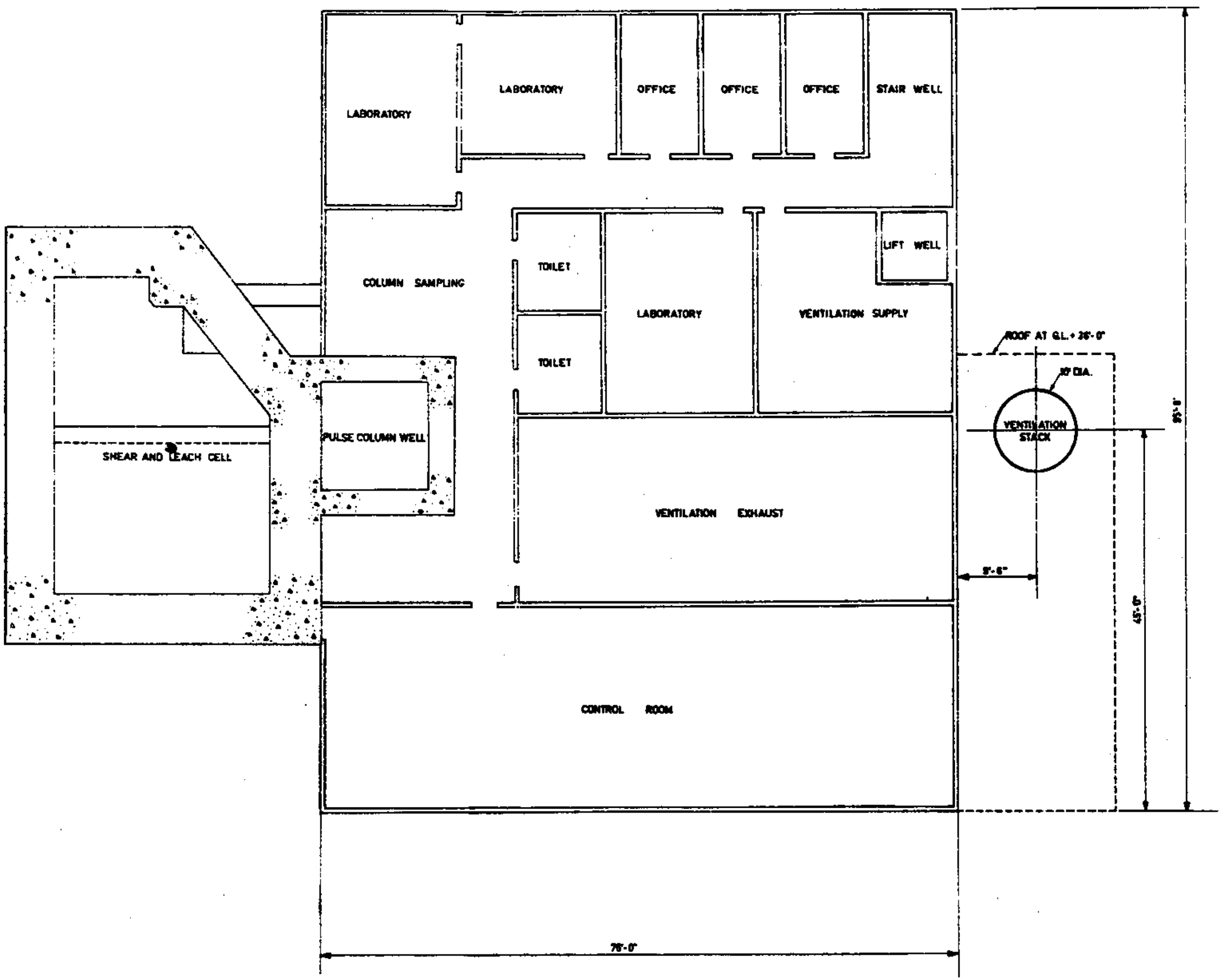


FIGURE 42. BUILDING DESIGN - PLAN 36 ft TO 50.5 ft

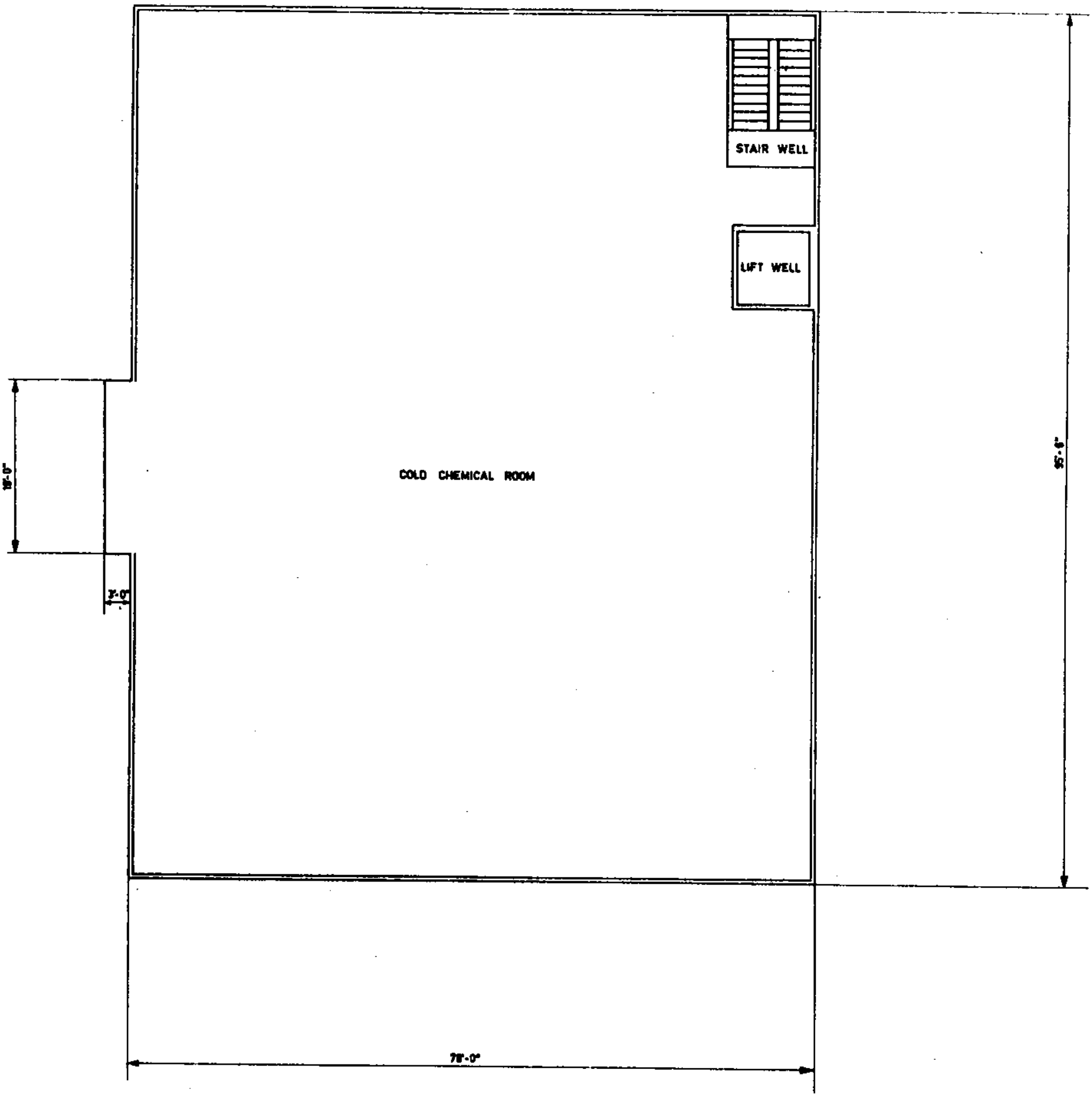


FIGURE 43. BUILDING DESIGN - PLAN 50.5 ft TO 65.5 ft

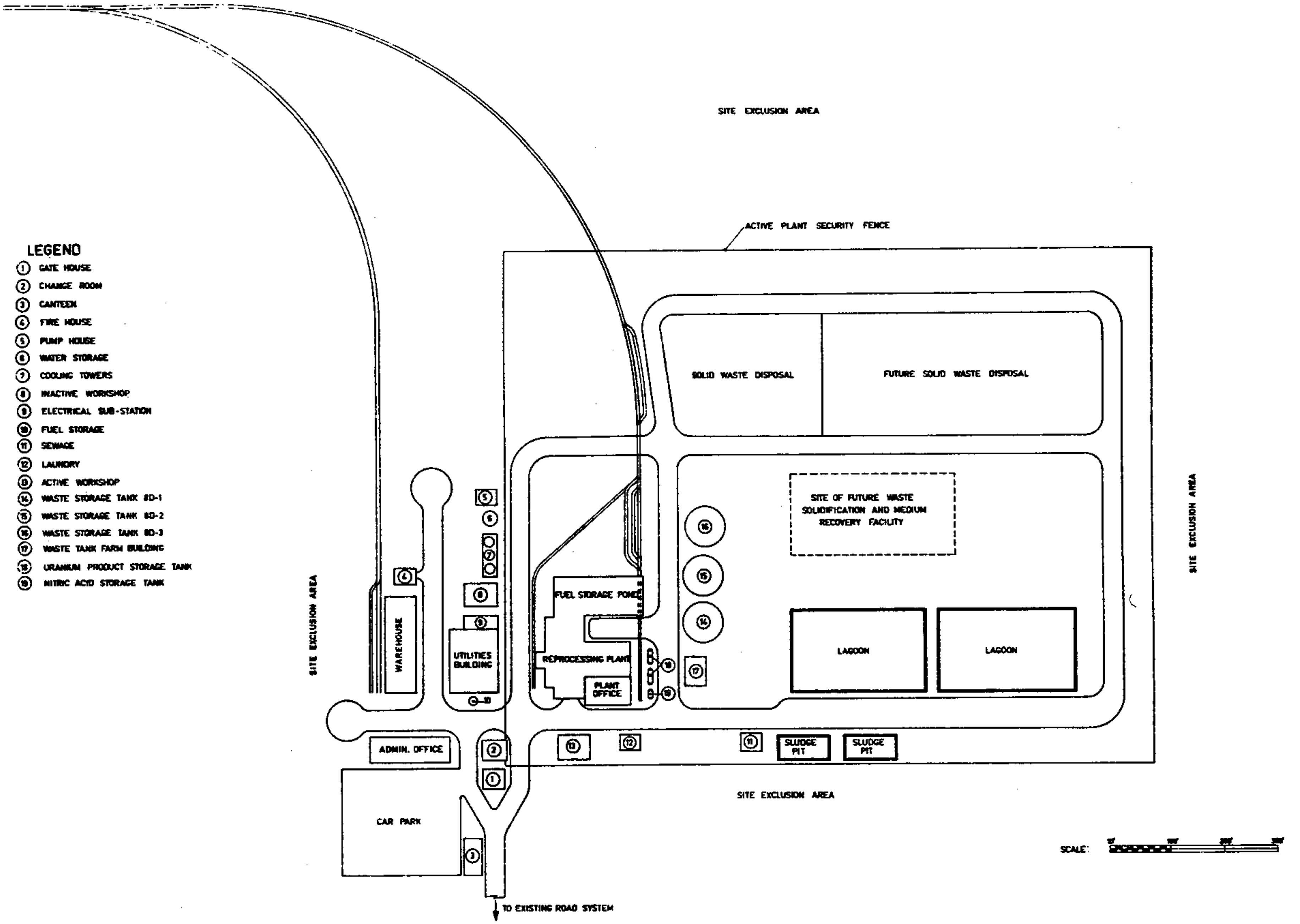


FIGURE 44. SITE LAYOUT

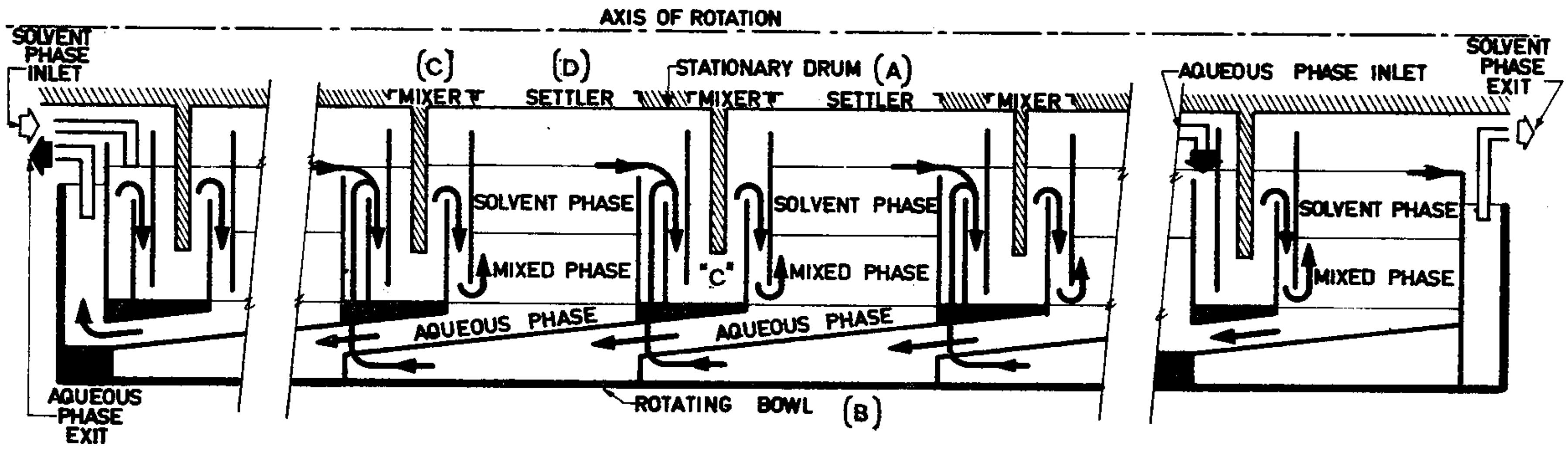


FIGURE 45. PRINCIPLES OF A CENTRIFUGAL STAGE

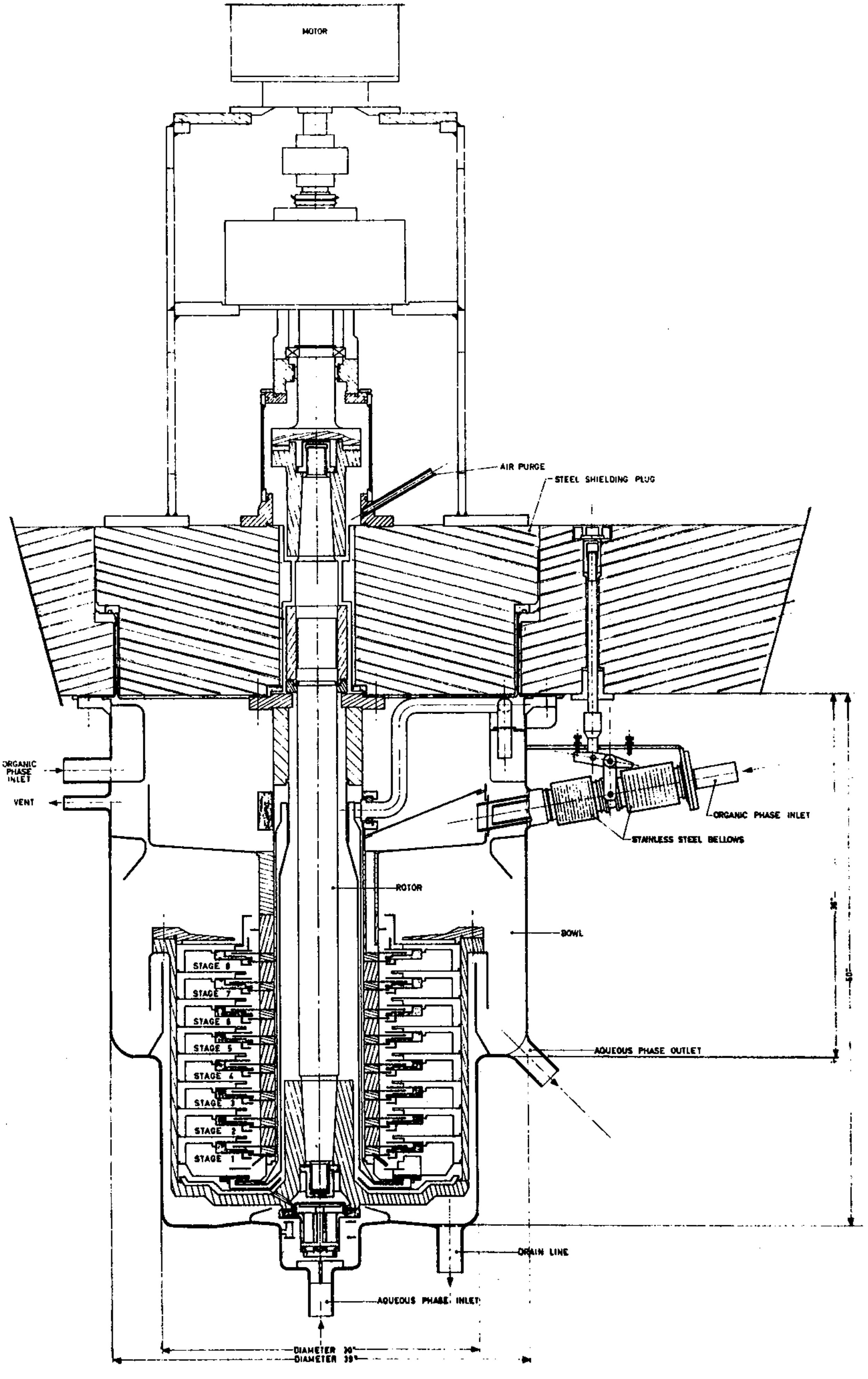
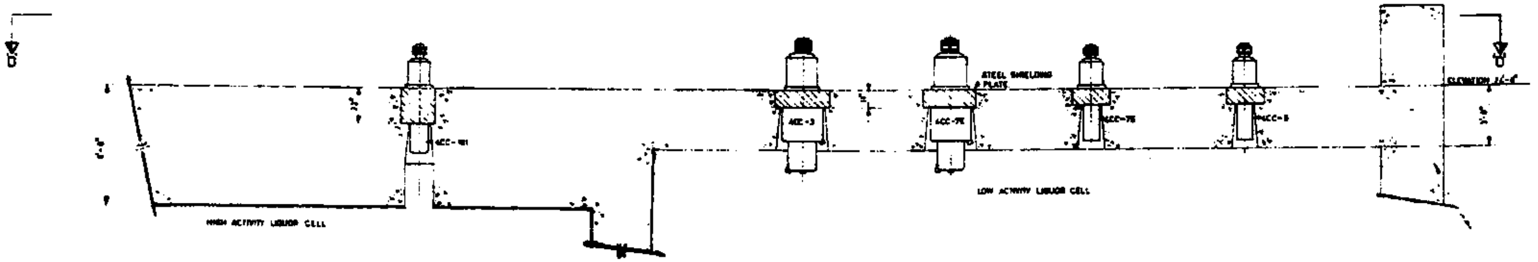
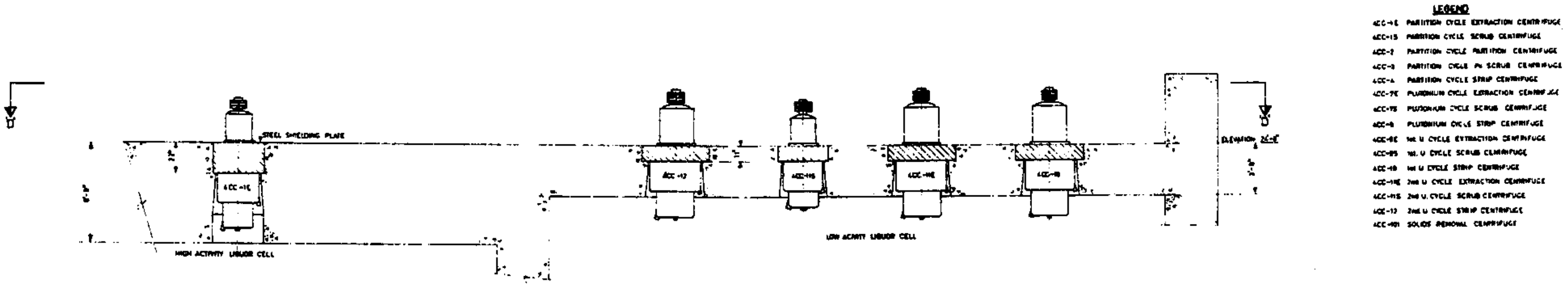


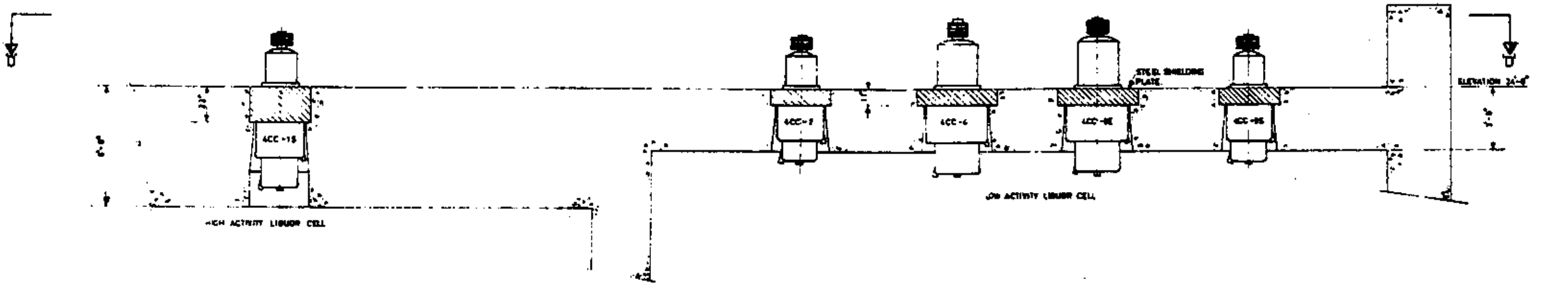
FIGURE 46. ROBATEL CENTRIFUGAL CONTACTOR



PART SECTION 'A-A'

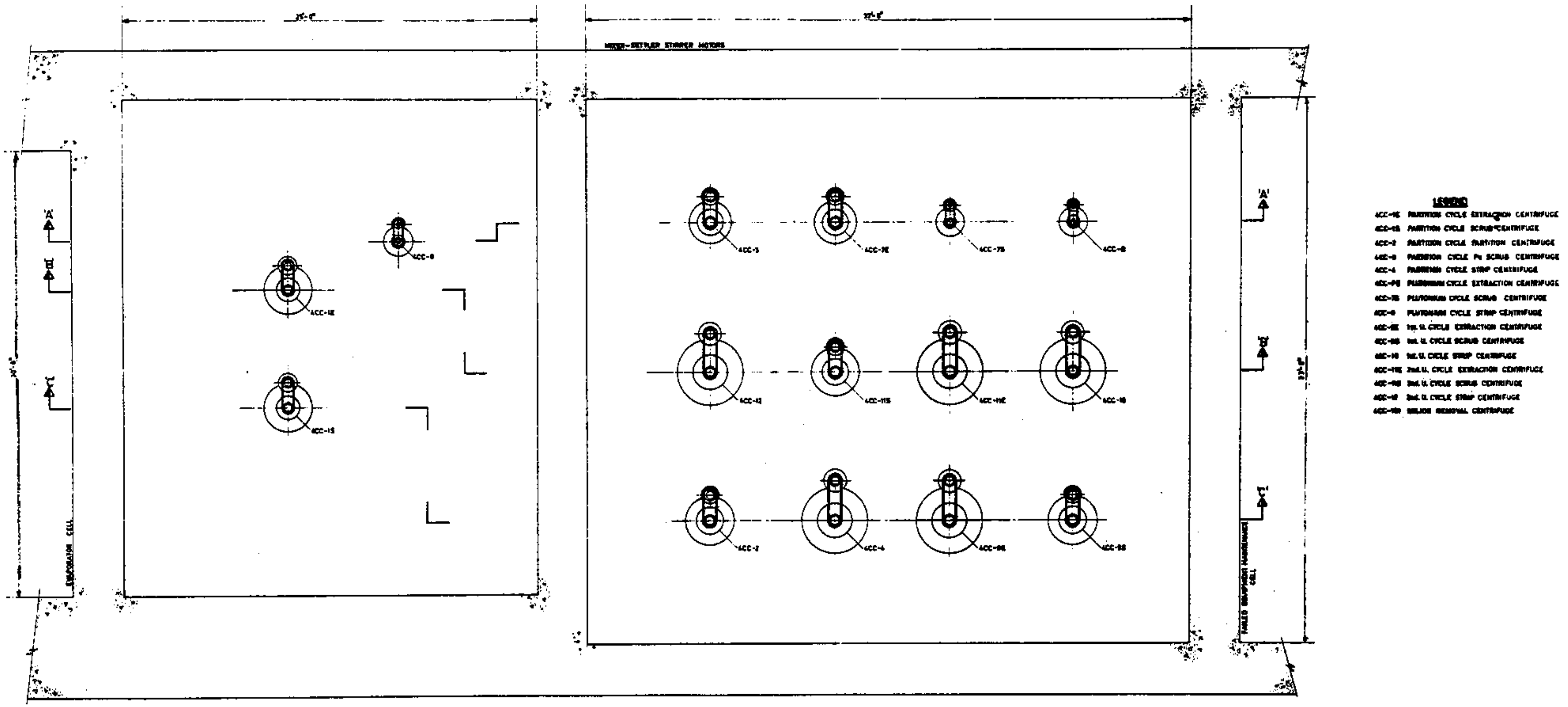


PART SECTION 'B-B'



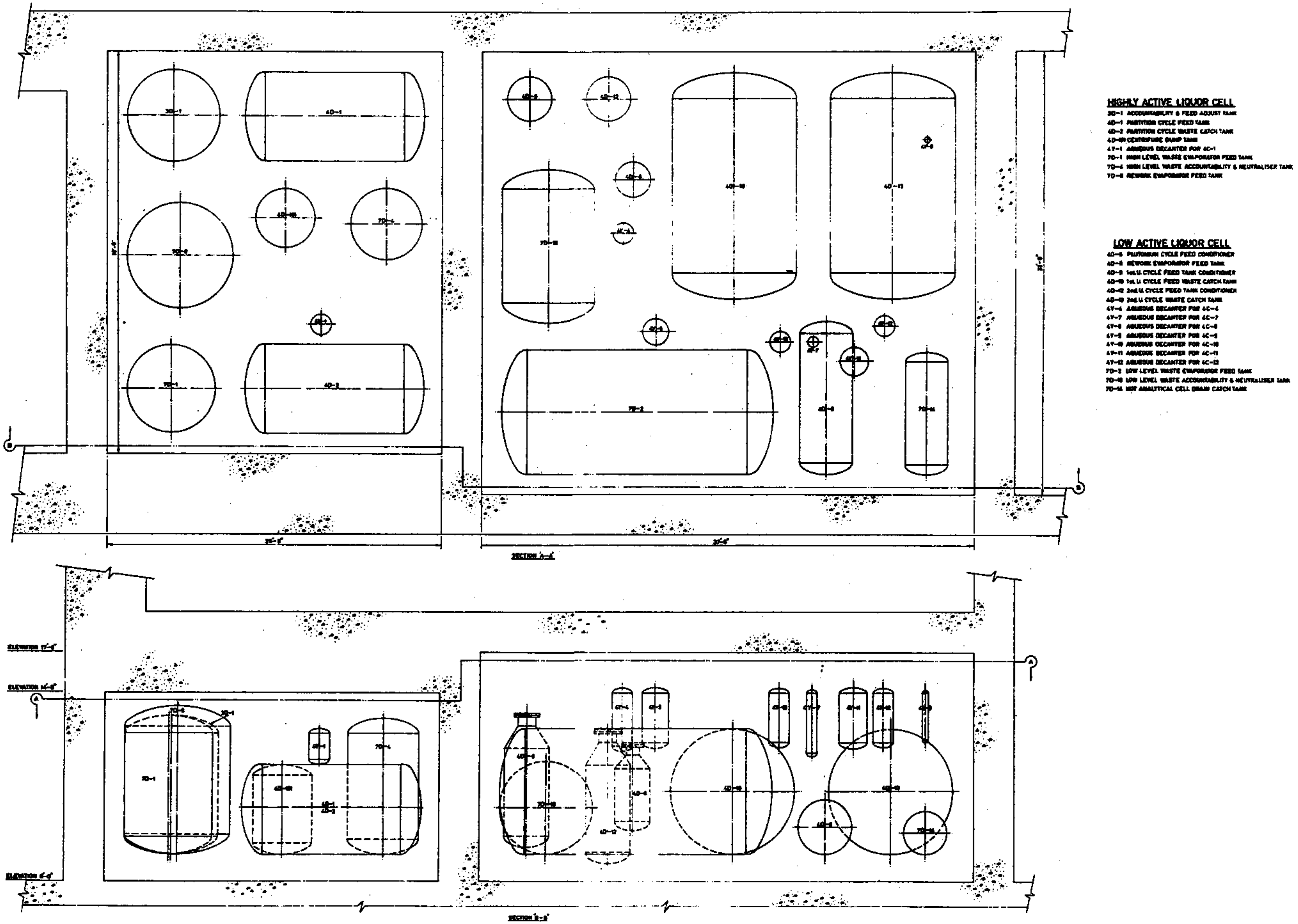
PART SECTION 'C-C'

FIGURE 47. EQUIPMENT LAYOUT - CONTACTORS, SECTION ELEVATION



SECTION 'D-D'

FIGURE 48. EQUIPMENT LAYOUT - PLAN VIEW OF CONTACTORS



HIGHLY ACTIVE LIQUOR CELL

- 3D-1 ACCOUNTABILITY & FEED ADJUST TANK
- 4D-1 PARTITION CYCLE FEED TANK
- 4D-2 PARTITION CYCLE WASTE CATCH TANK
- 4D-18 CENTRIFUGE DUMP TANK
- 4Y-1 AQUEOUS DECANTER FOR 4C-1
- 7D-1 HIGH LEVEL WASTE EVAPORATOR FEED TANK
- 7D-4 HIGH LEVEL WASTE ACCOUNTABILITY & NEUTRALISER TANK
- 7D-8 REWORK EVAPORATOR FEED TANK

LOW ACTIVE LIQUOR CELL

- 4D-6 PLUTONIUM CYCLE FEED CONDITIONER
- 4D-8 REWORK EVAPORATOR FEED TANK
- 4D-9 1st U CYCLE FEED TANK CONDITIONER
- 4D-10 1st U CYCLE FEED WASTE CATCH TANK
- 4D-12 2nd U CYCLE FEED TANK CONDITIONER
- 4D-13 2nd U CYCLE WASTE CATCH TANK
- 4Y-6 AQUEOUS DECANTER FOR 4C-6
- 4Y-7 AQUEOUS DECANTER FOR 4C-7
- 4Y-8 AQUEOUS DECANTER FOR 4C-8
- 4Y-9 AQUEOUS DECANTER FOR 4C-9
- 4Y-10 AQUEOUS DECANTER FOR 4C-10
- 4Y-11 AQUEOUS DECANTER FOR 4C-11
- 4Y-12 AQUEOUS DECANTER FOR 4C-12
- 7D-2 LOW LEVEL WASTE EVAPORATOR FEED TANK
- 7D-10 LOW LEVEL WASTE ACCOUNTABILITY & NEUTRALISER TANK
- 7D-14 HOT ANALYTICAL CELL DRAIN CATCH TANK

FIGURE 49. EQUIPMENT LAYOUT - ACTIVE LIQUOR CELLS

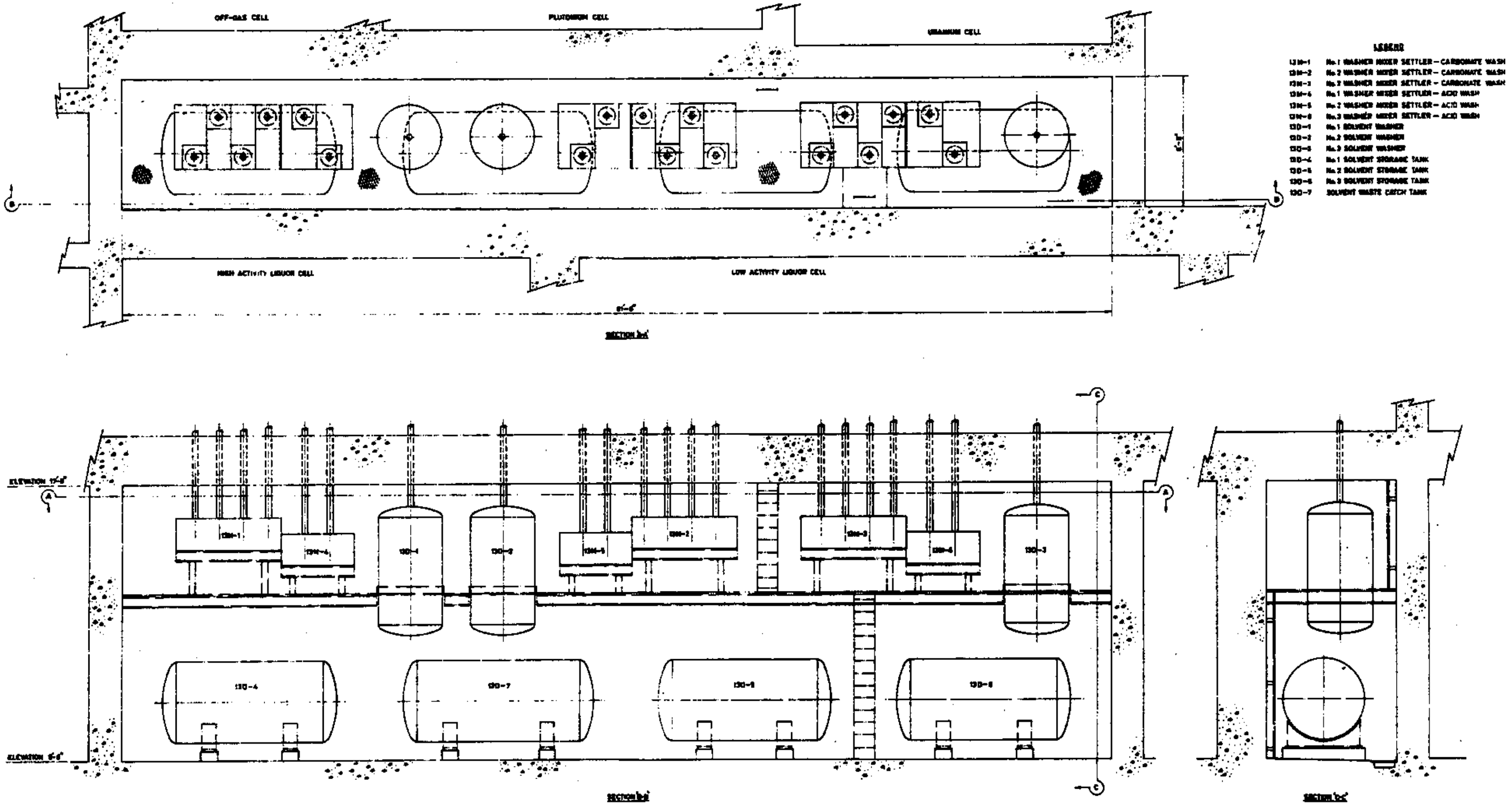
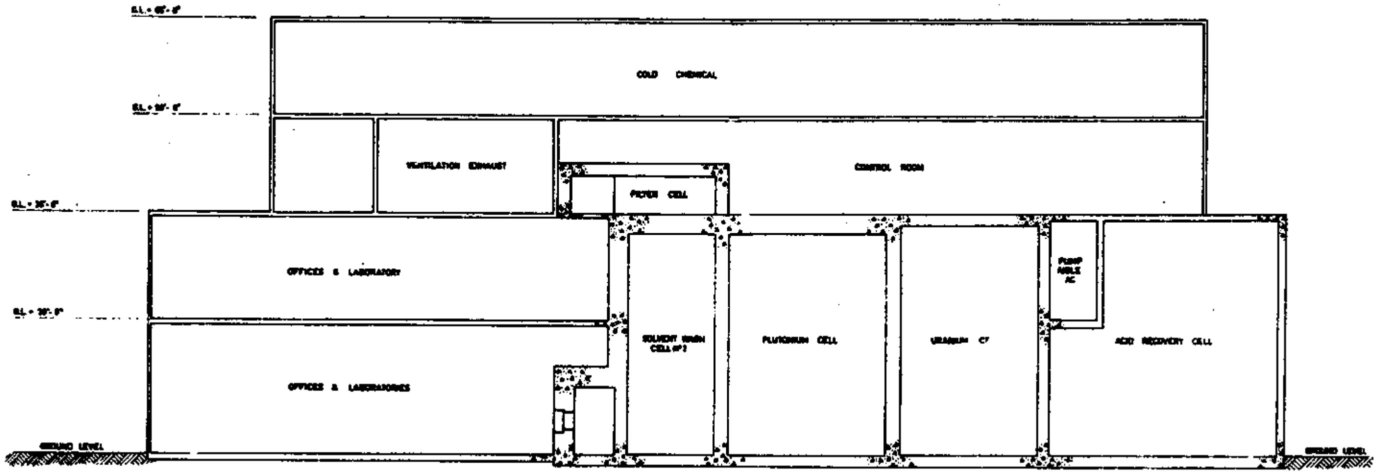
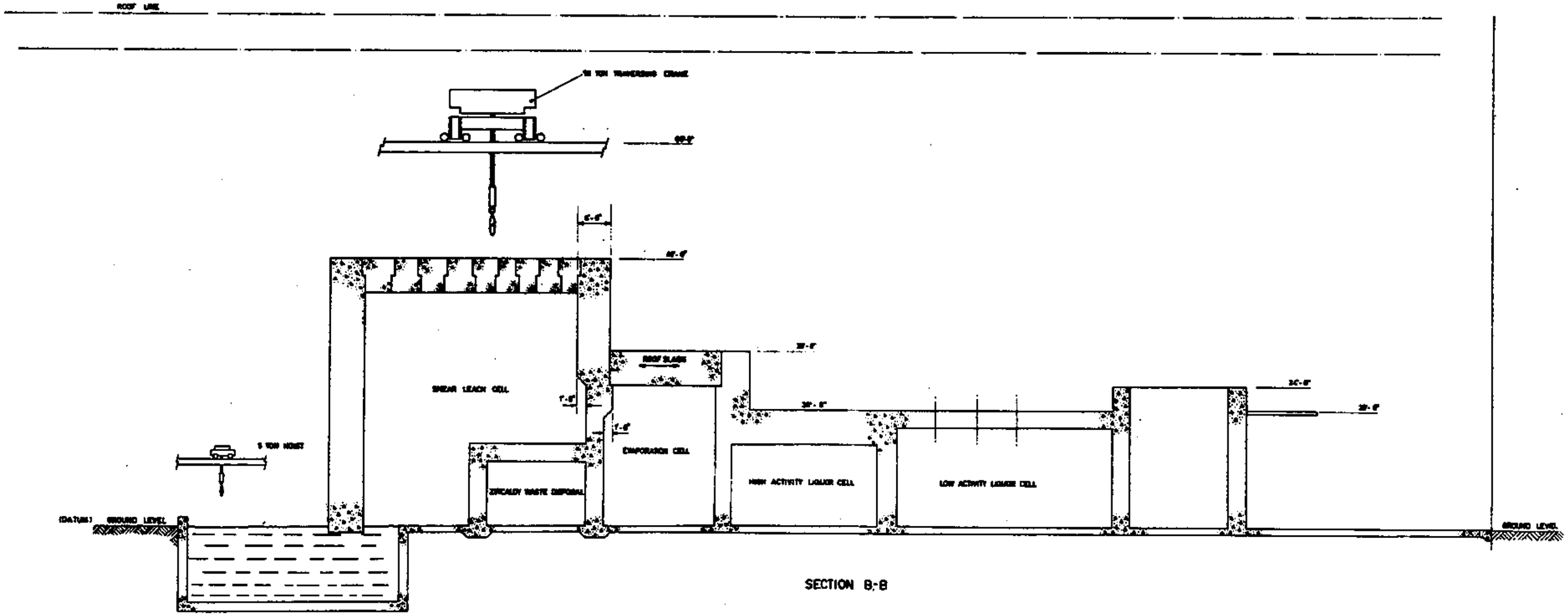


FIGURE 50. EQUIPMENT LAYOUT - SOLVENT WASH CELL



SECTION C-C



SECTION B-B

FIGURE 51. BUILDING DESIGN - BUILDING SECTIONS

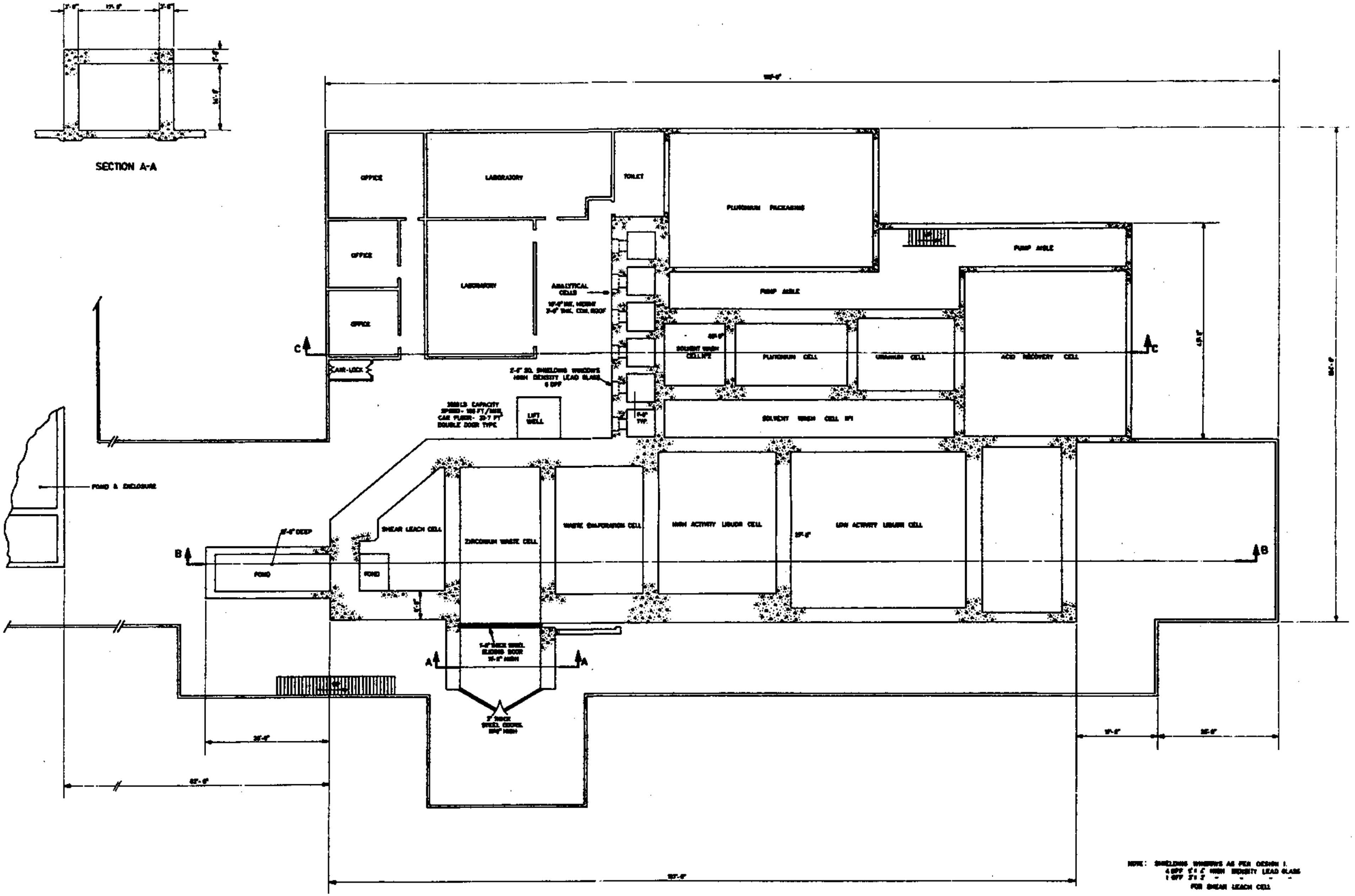


FIGURE 52. BUILDING DESIGN - PLAN 0 ft TO 20 ft

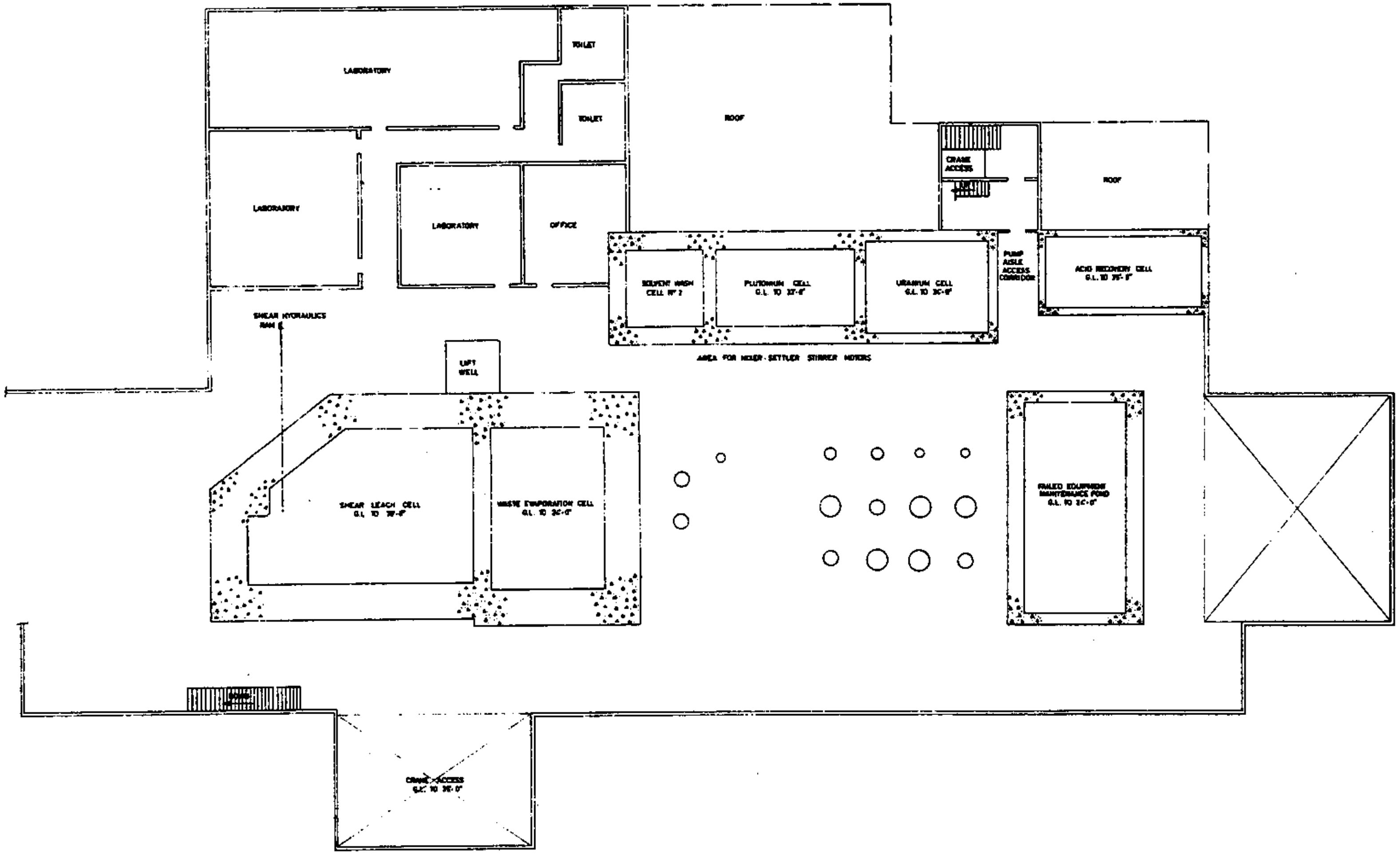


FIGURE 53. BUILDING DESIGN - PLAN 20 ft TO 36 ft

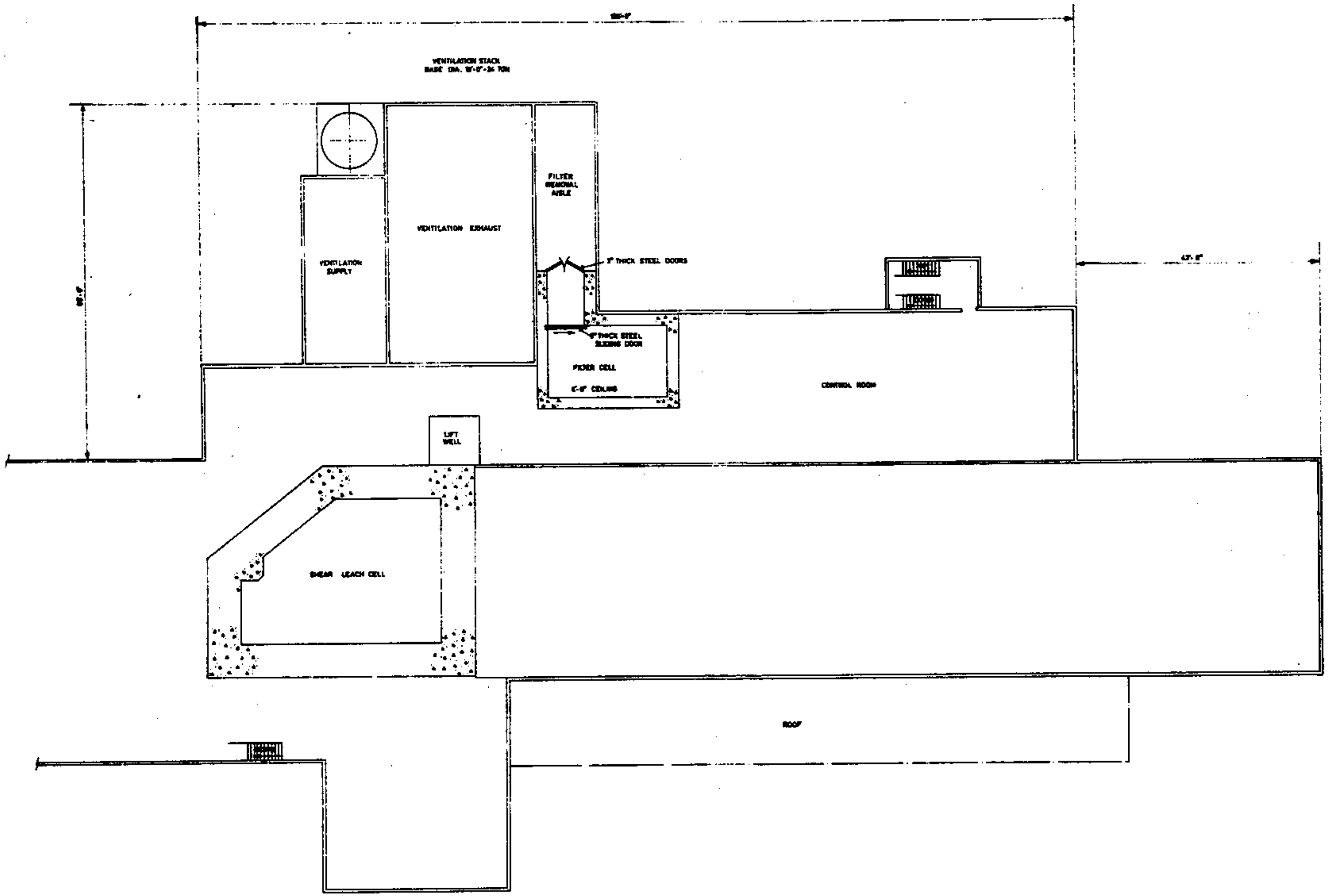


FIGURE 54. BUILDING DESIGN - PLAN 36 ft TO 50.5 ft

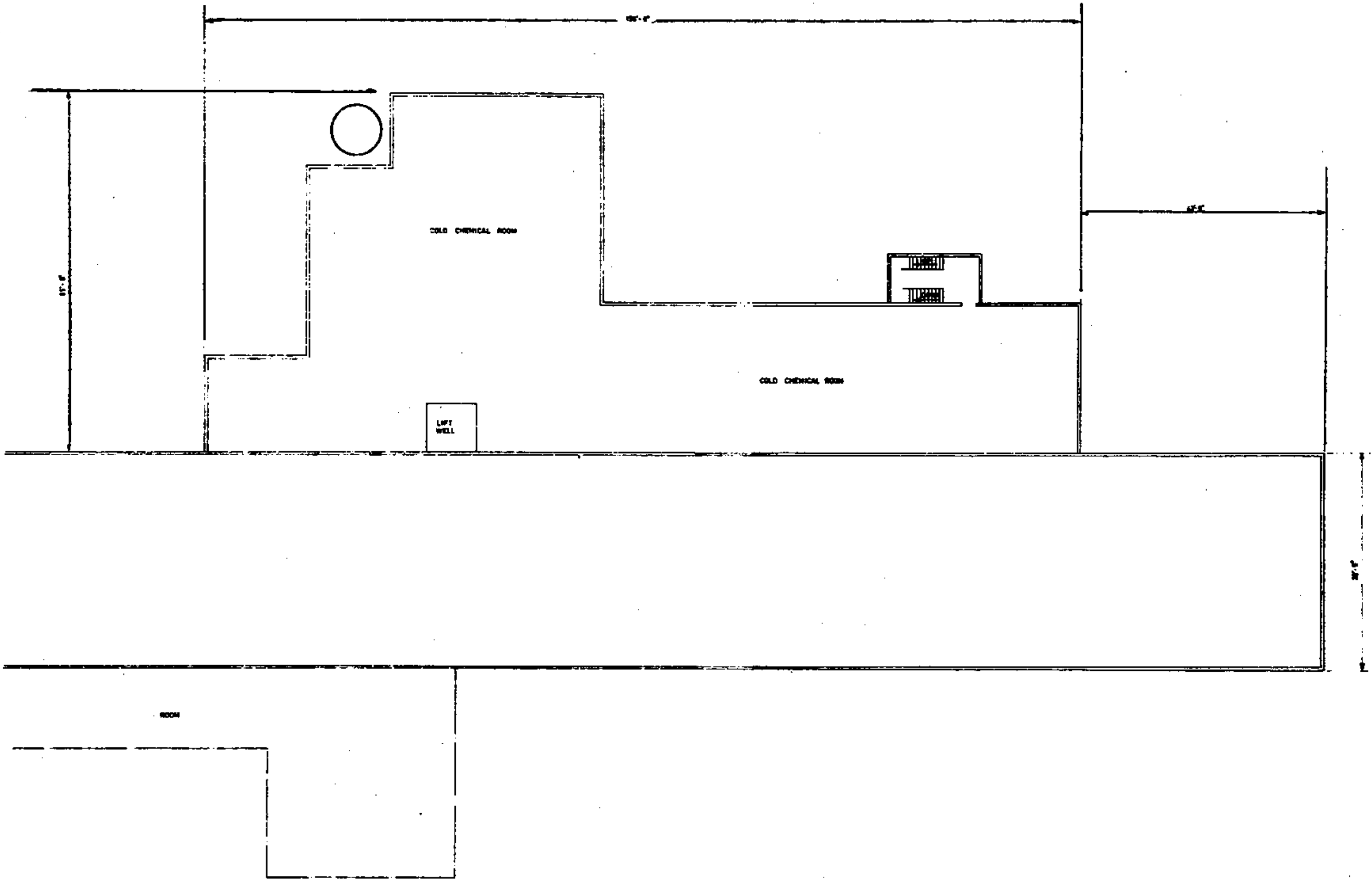


FIGURE 55. BUILDING DESIGN - 50.5 ft TO 65 ft

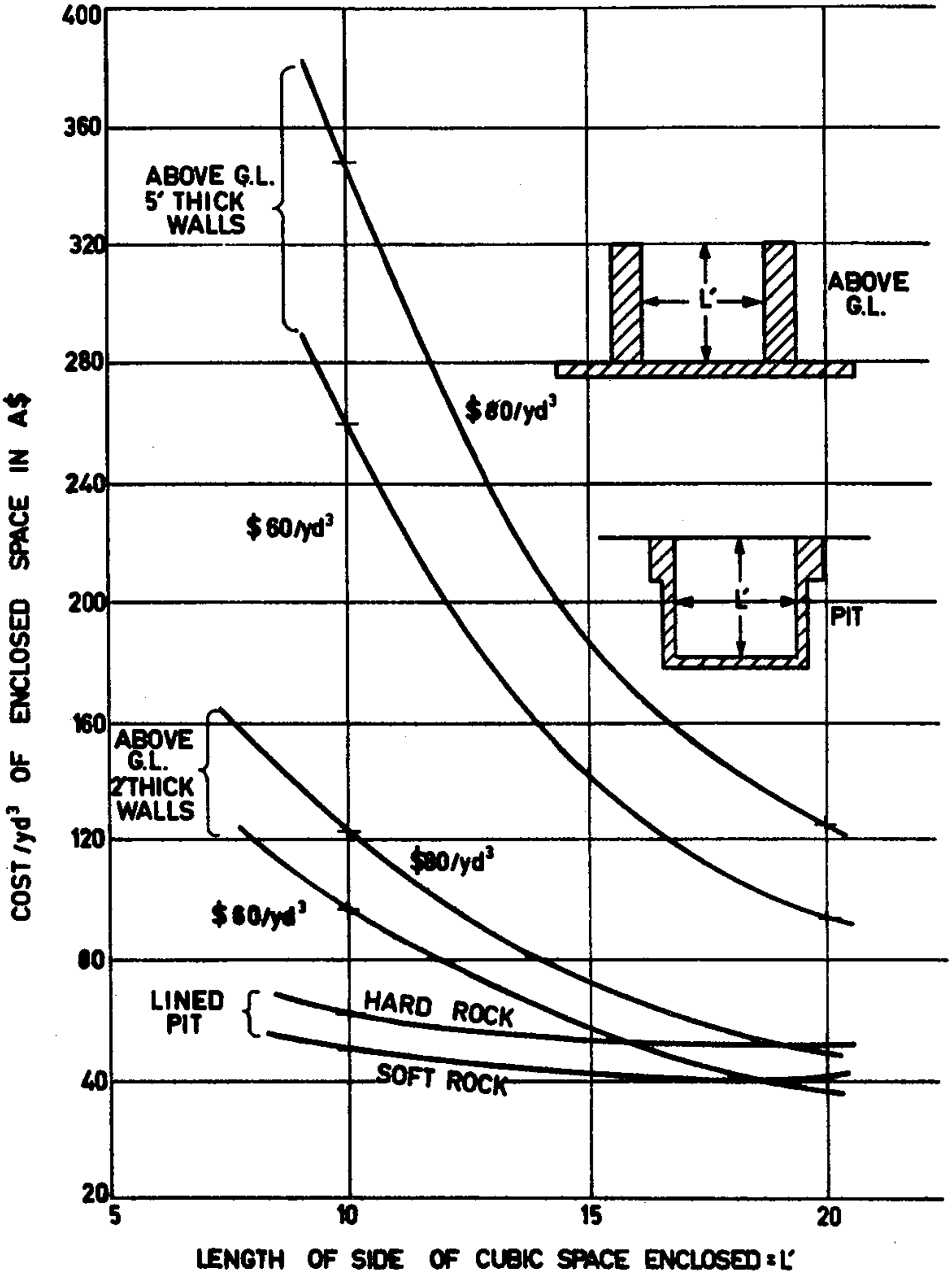


FIGURE 56. COMPARATIVE COSTS OF CONSTRUCTING PITS AND ABOVE GROUND CAVES FROM ORDINARY CONCRETE

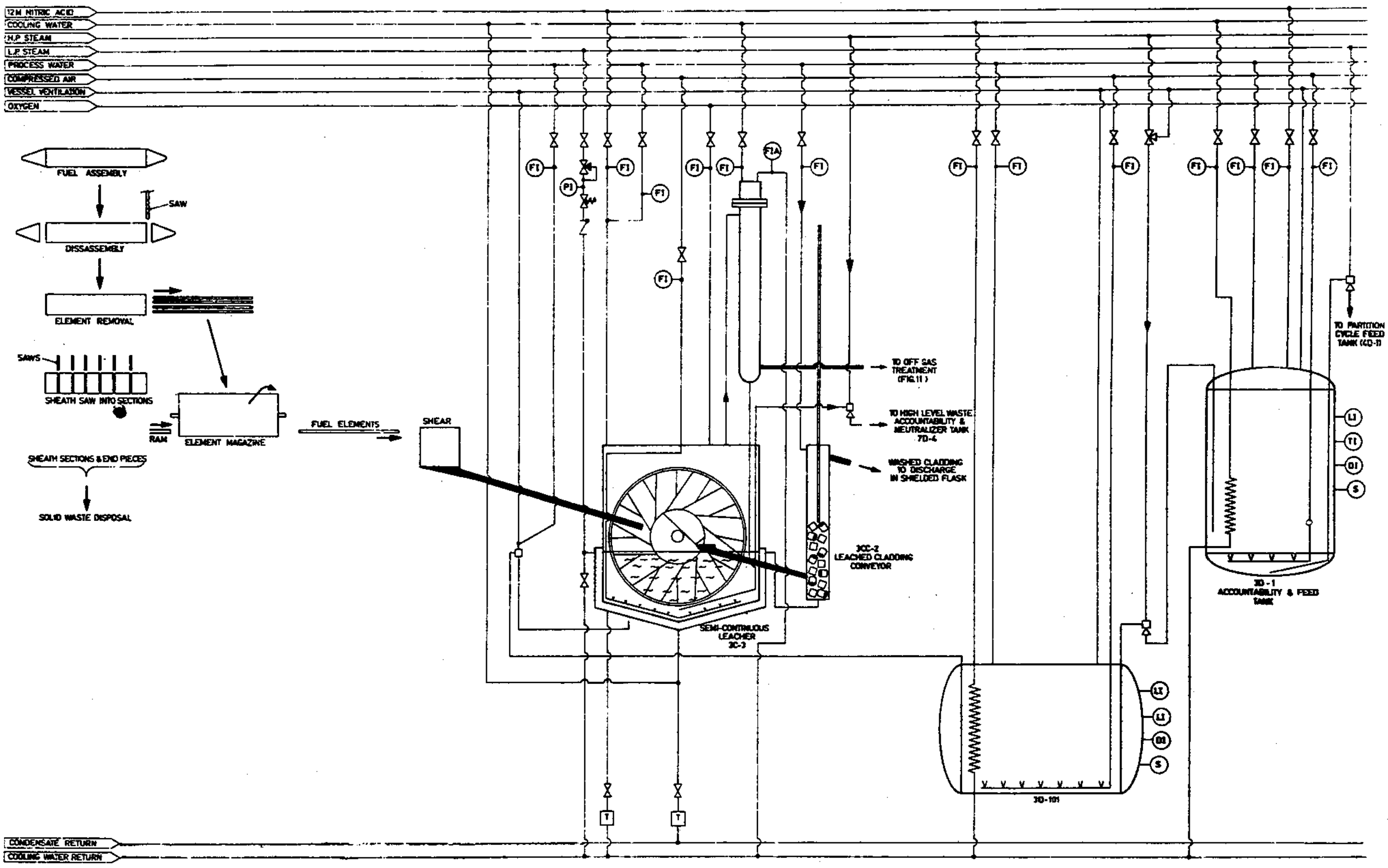


FIGURE 57 SEMI-CONTINUOUS LEACHING

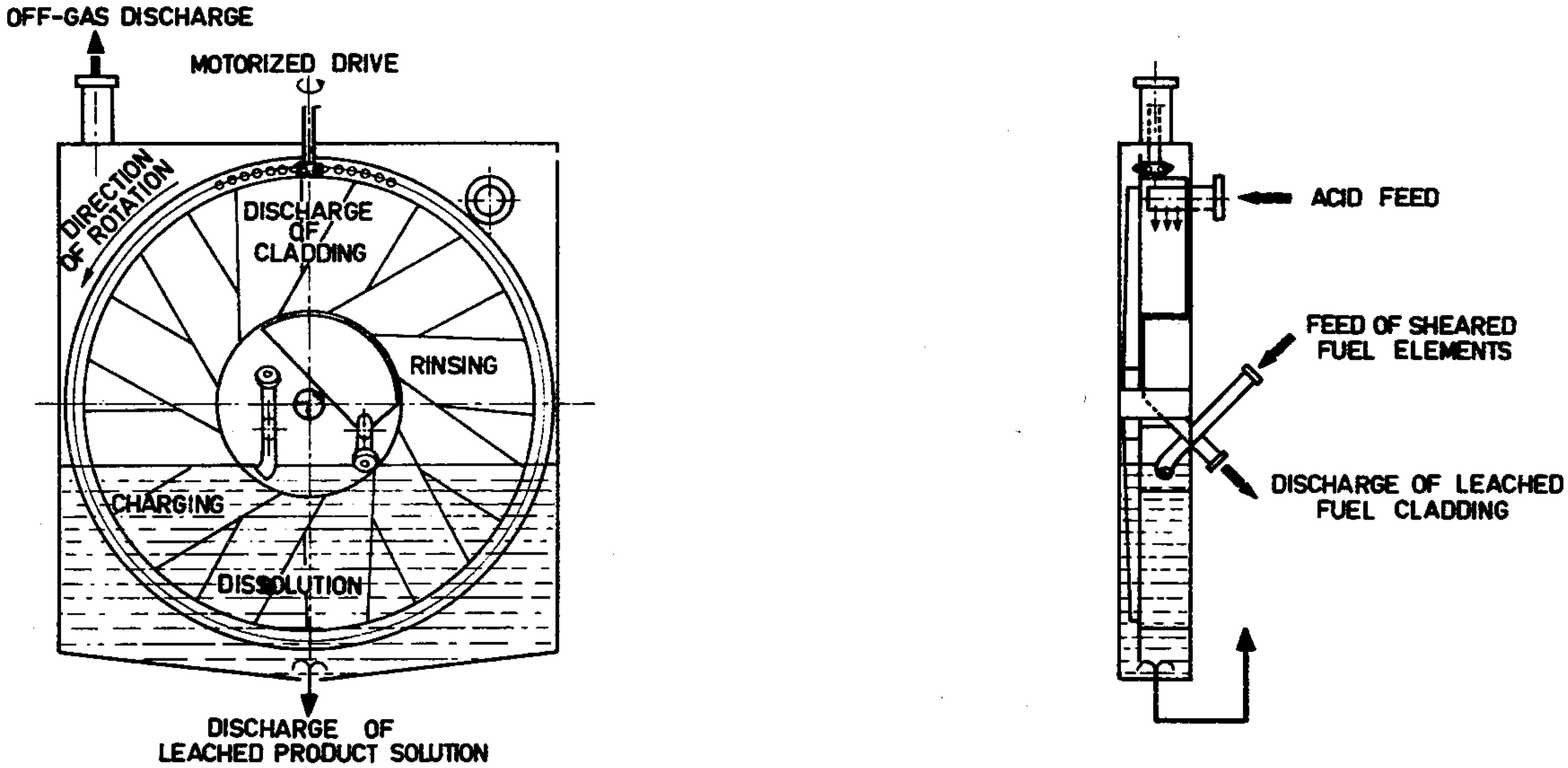


FIGURE 58. PRINCIPLES OF A ROTATING SEMI-CONTINUOUS LEACHER FOR RE-PROCESSING IRRADIATED OXIDE FUELS

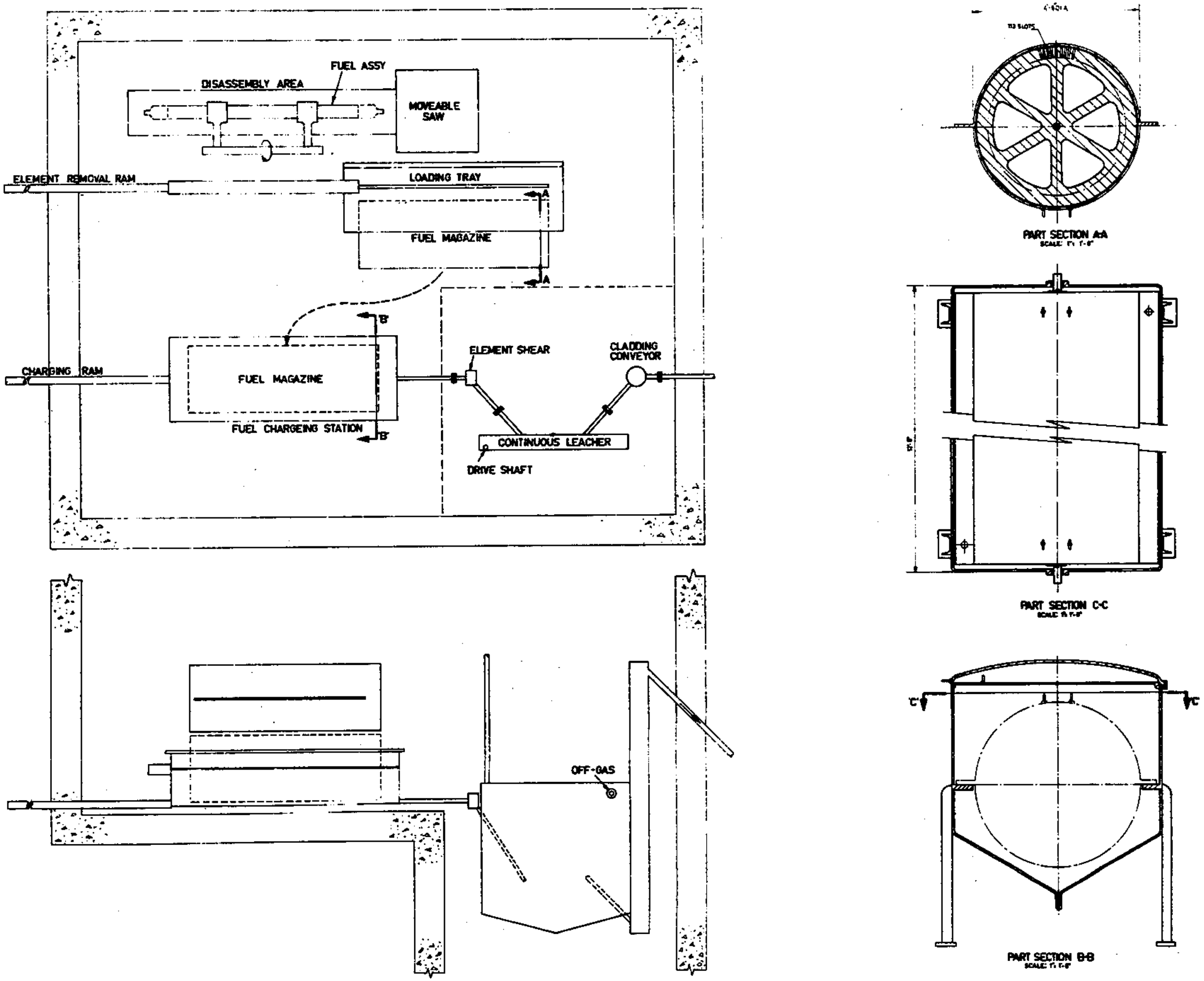


FIGURE 59. PREBREAKDOWN, SHEAR AND LEACHING CELL LAYOUT

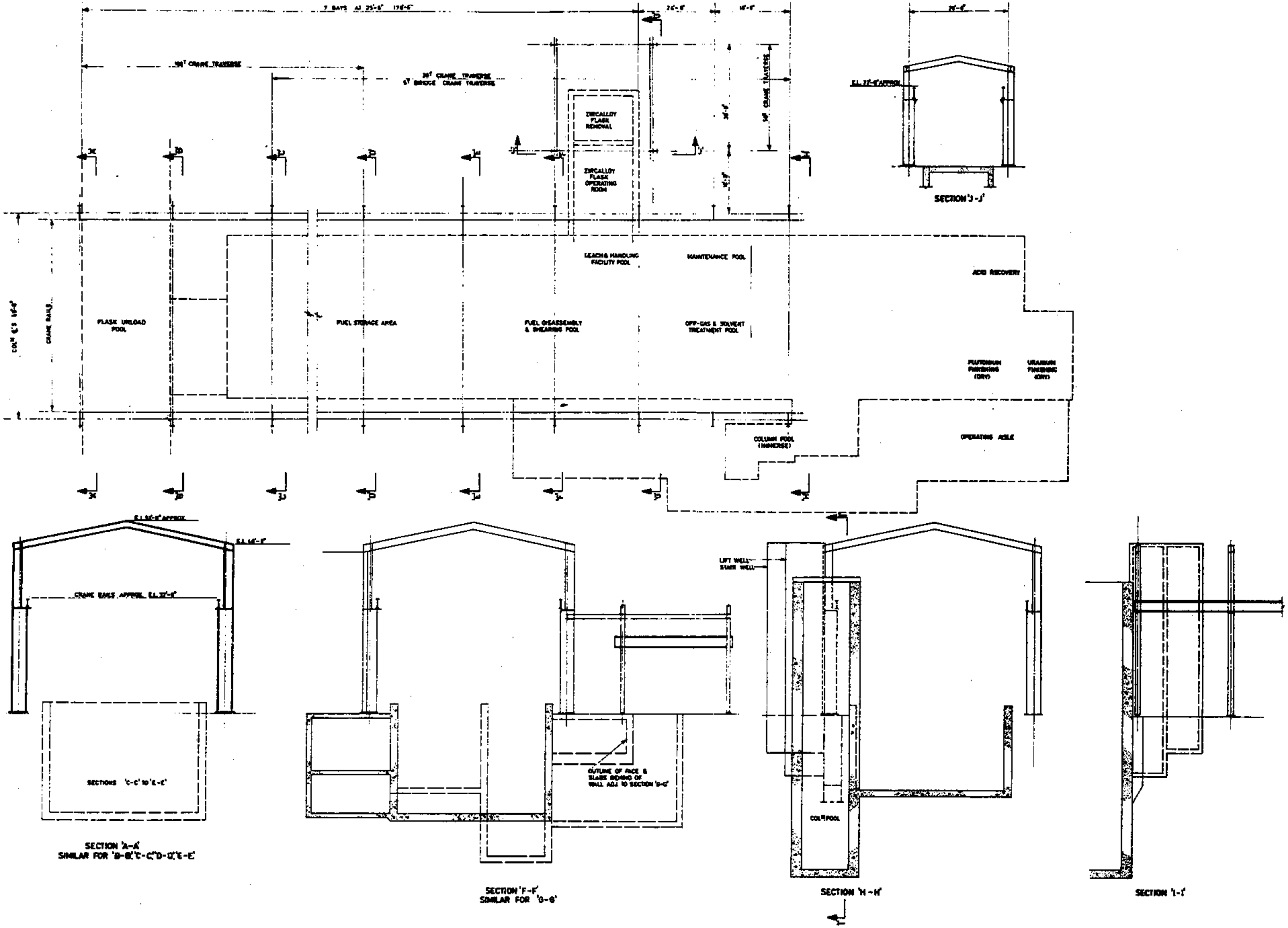


FIGURE 60. BUILDING LAYOUT - CRANE SUPPORTING BUILDING - SECTIONAL ELEVATION

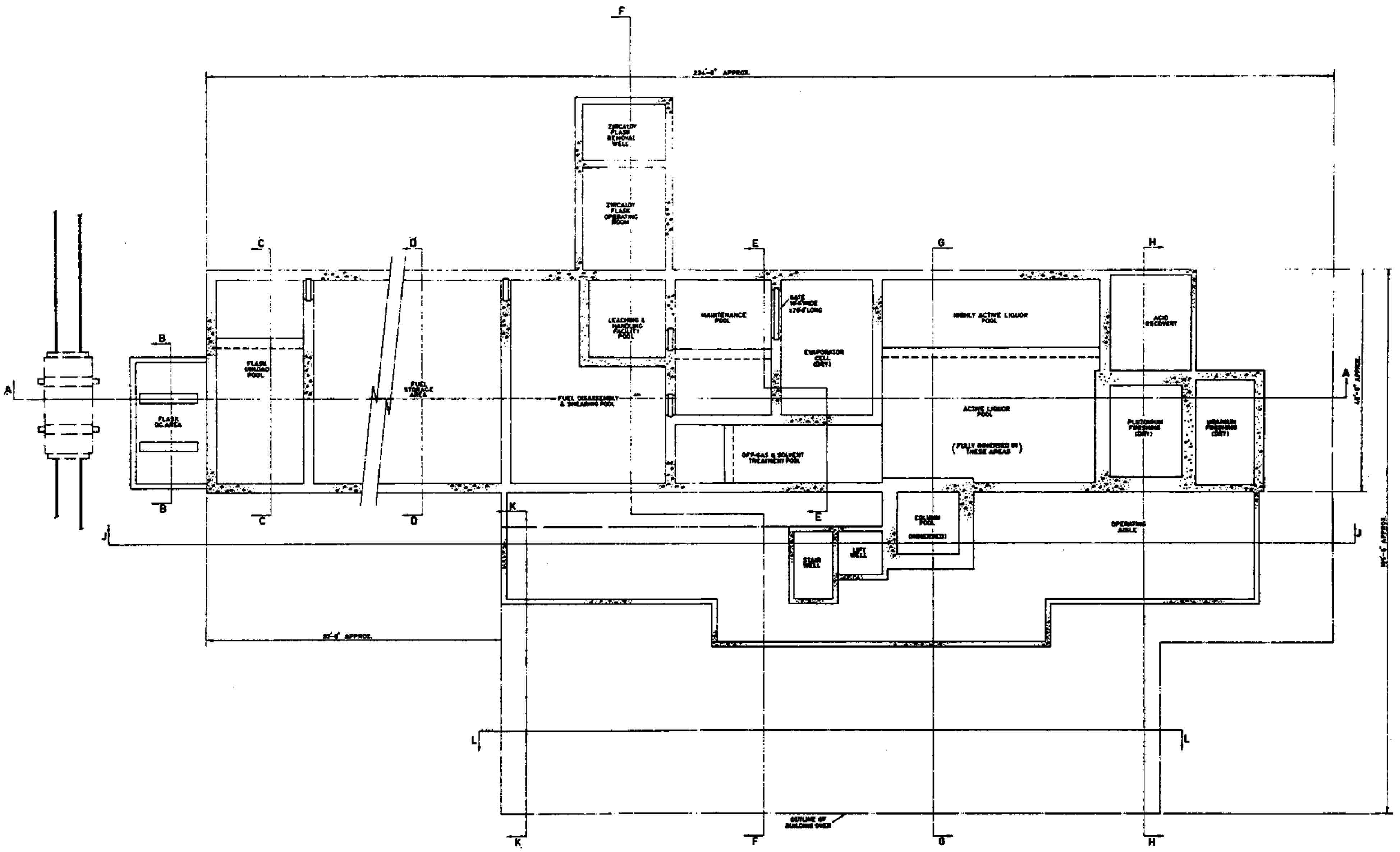


FIGURE 61. BUILDING LAYOUT - PLAN AT -12 ft

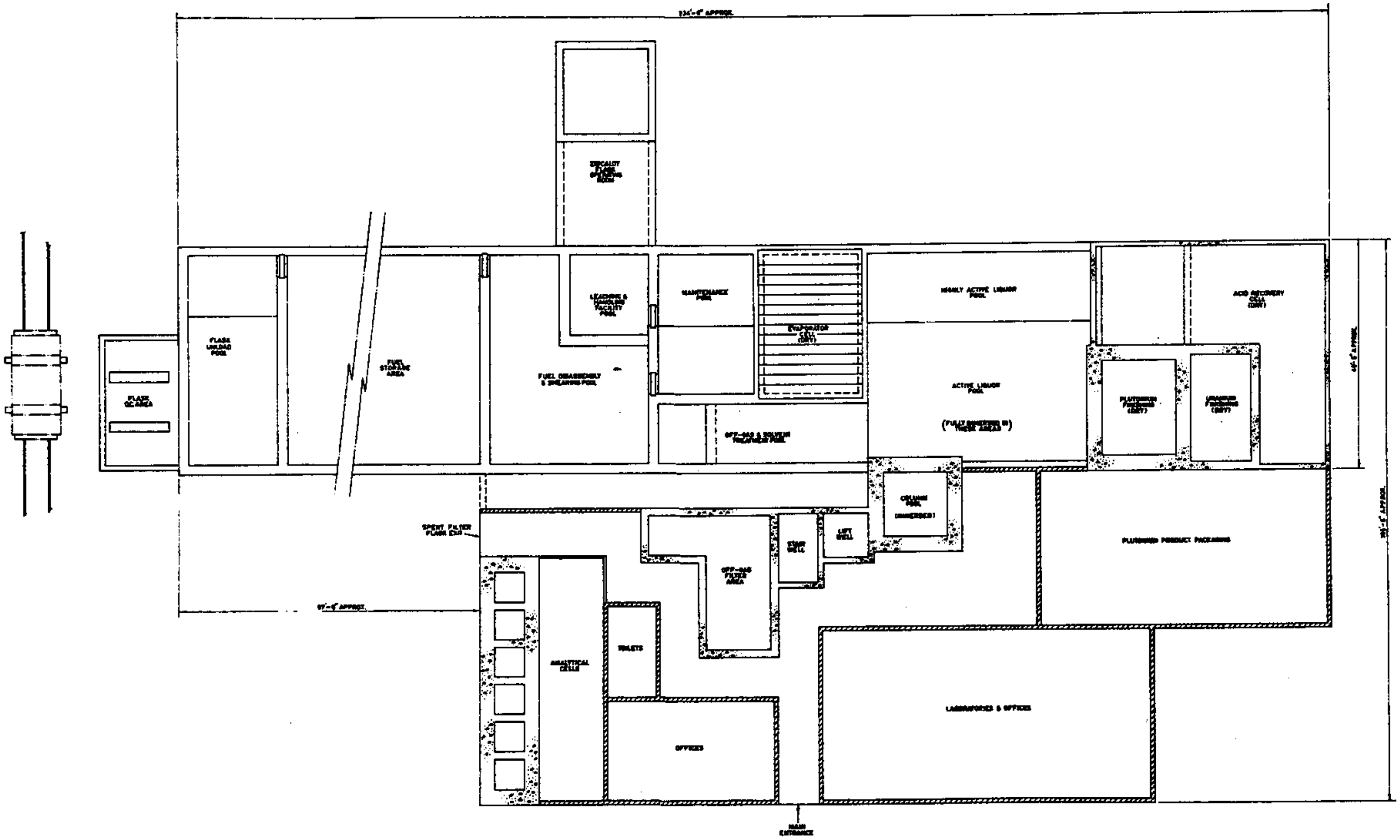


FIGURE 62. BUILDING LAYOUT - PLAN AT 0 ft

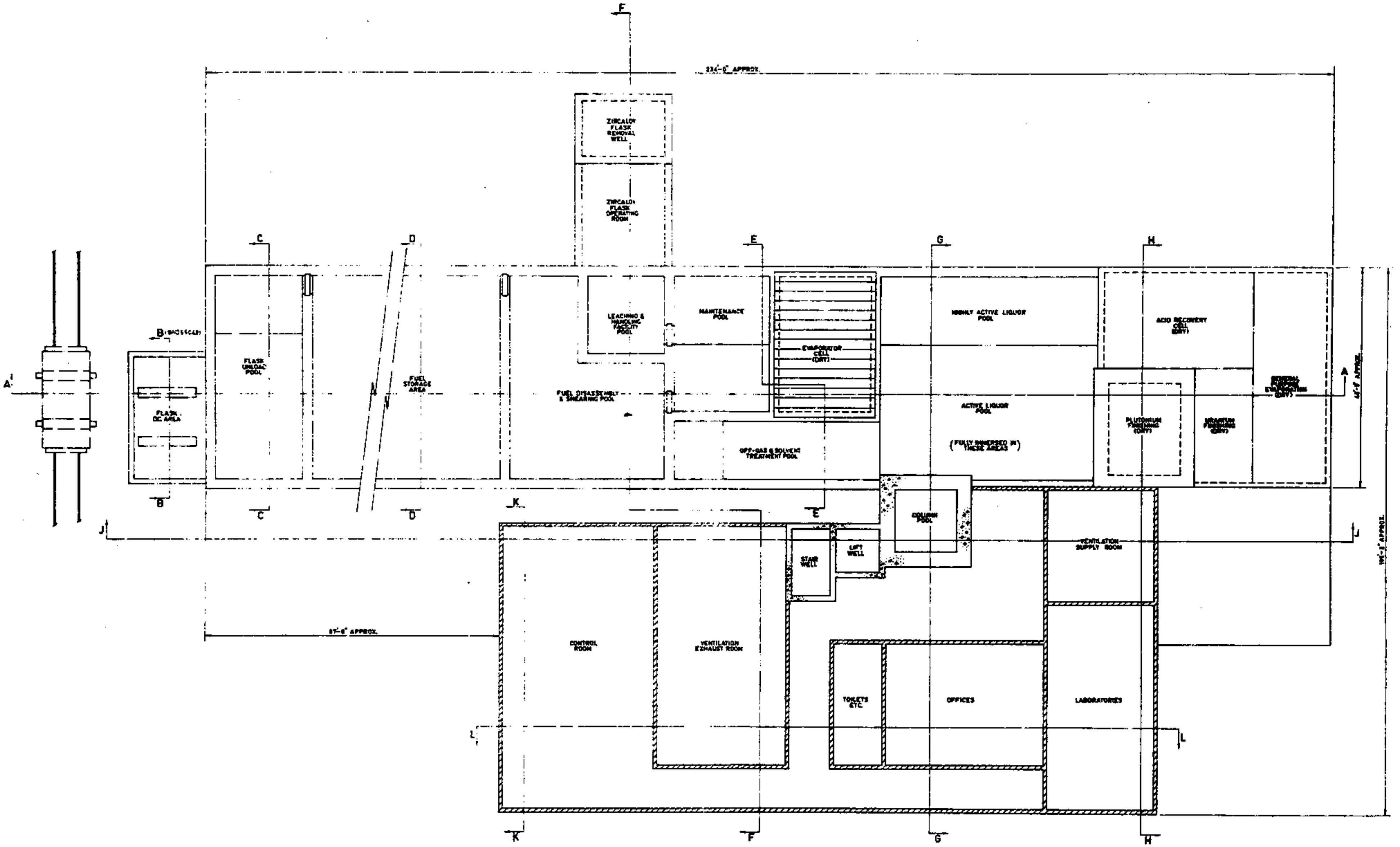


FIGURE 63. BUILDING LAYOUT - PLAN AT 16 ft

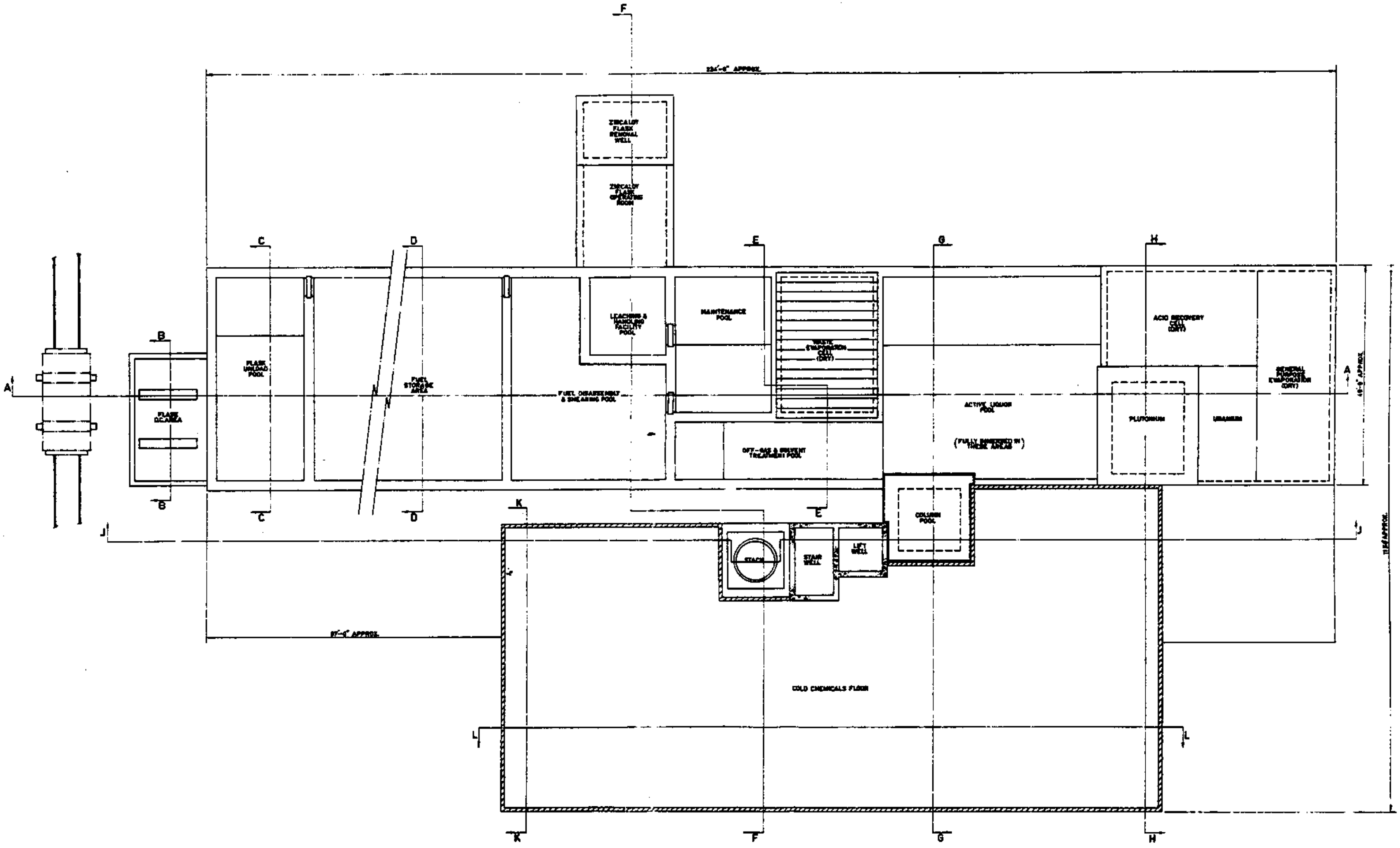


FIGURE 64. BUILDING LAYOUT - PLAN AT 32 ft

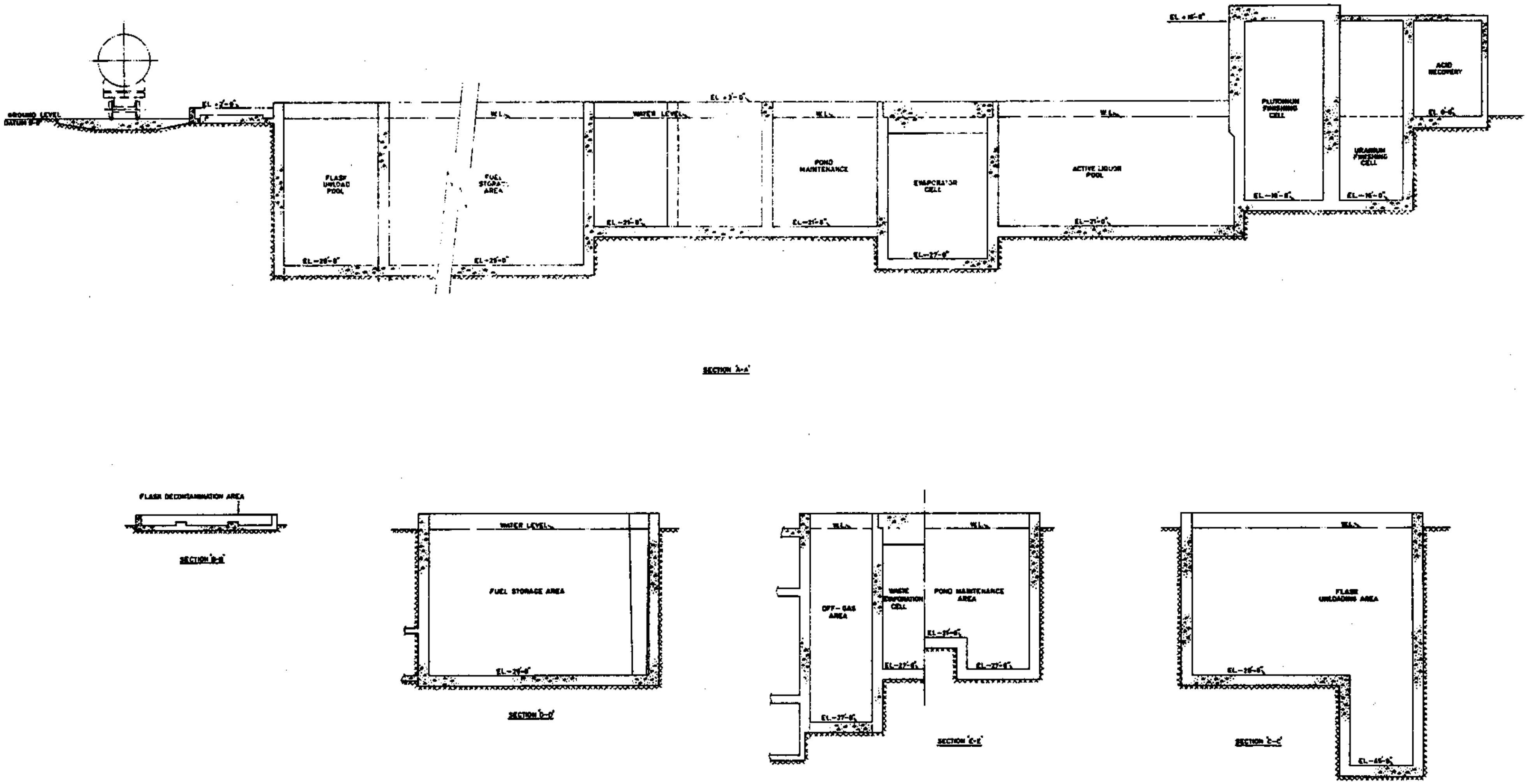


FIGURE 65. BUILDING LAYOUT - SECTIONS A-A, B-B, C-C, D-D AND E-E

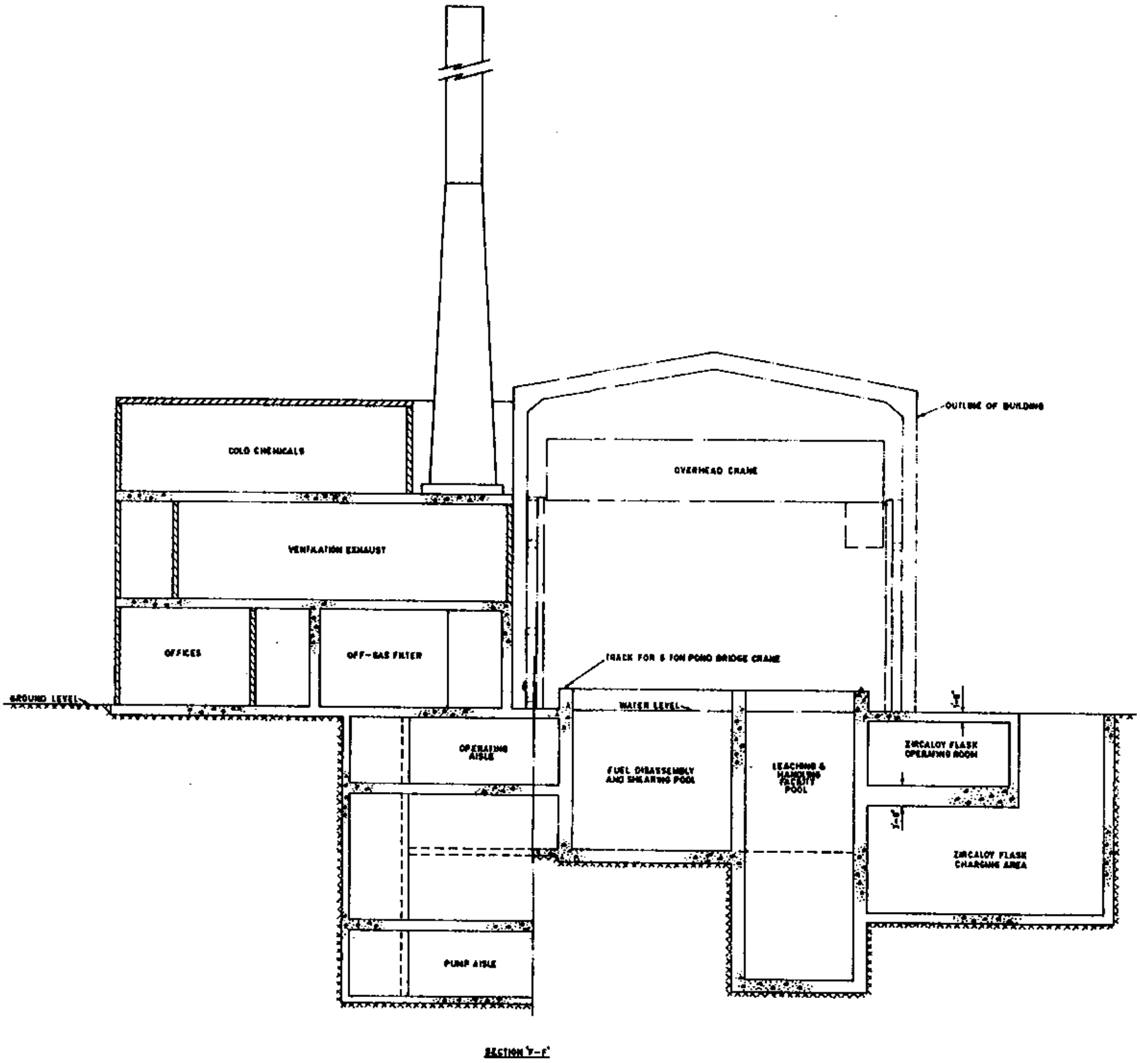
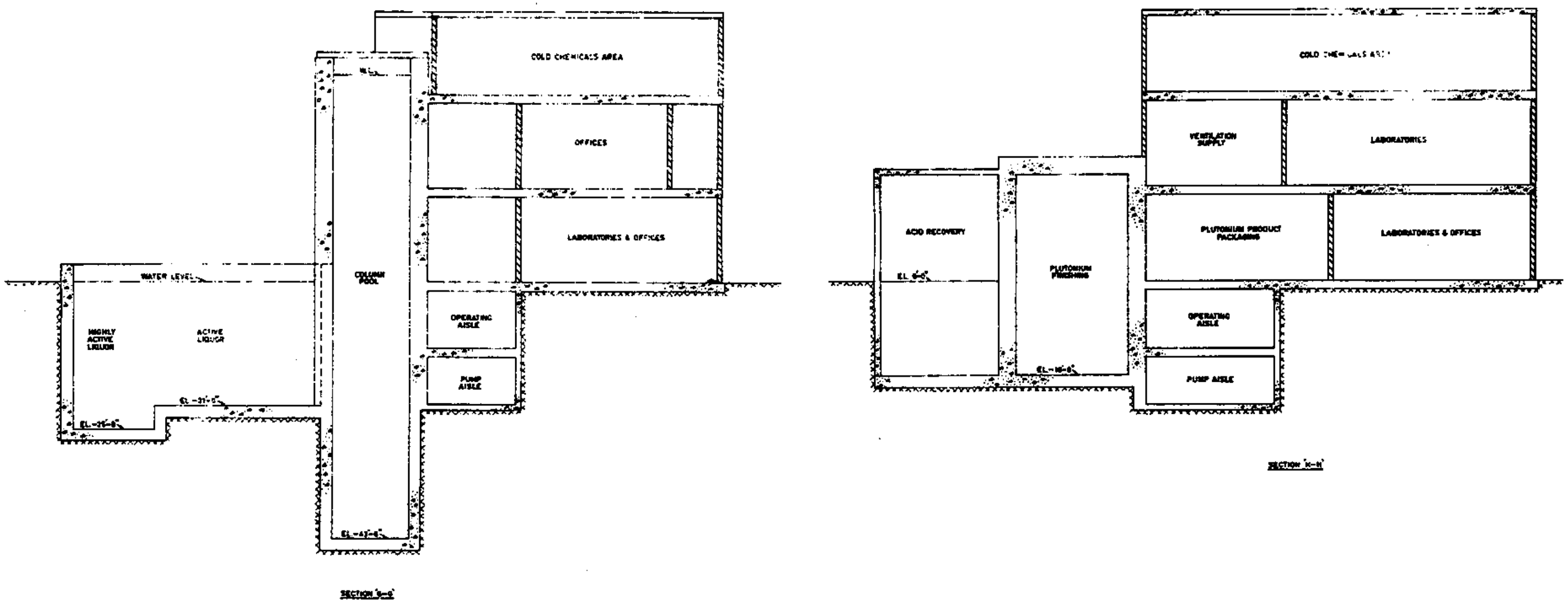


FIGURE 66. BUILDING LAYOUT - SECTION F-F



SECTION "G-G"

SECTION "H-H"

FIGURE 67. BUILDING LAYOUT - SECTIONS G-G AND H-H

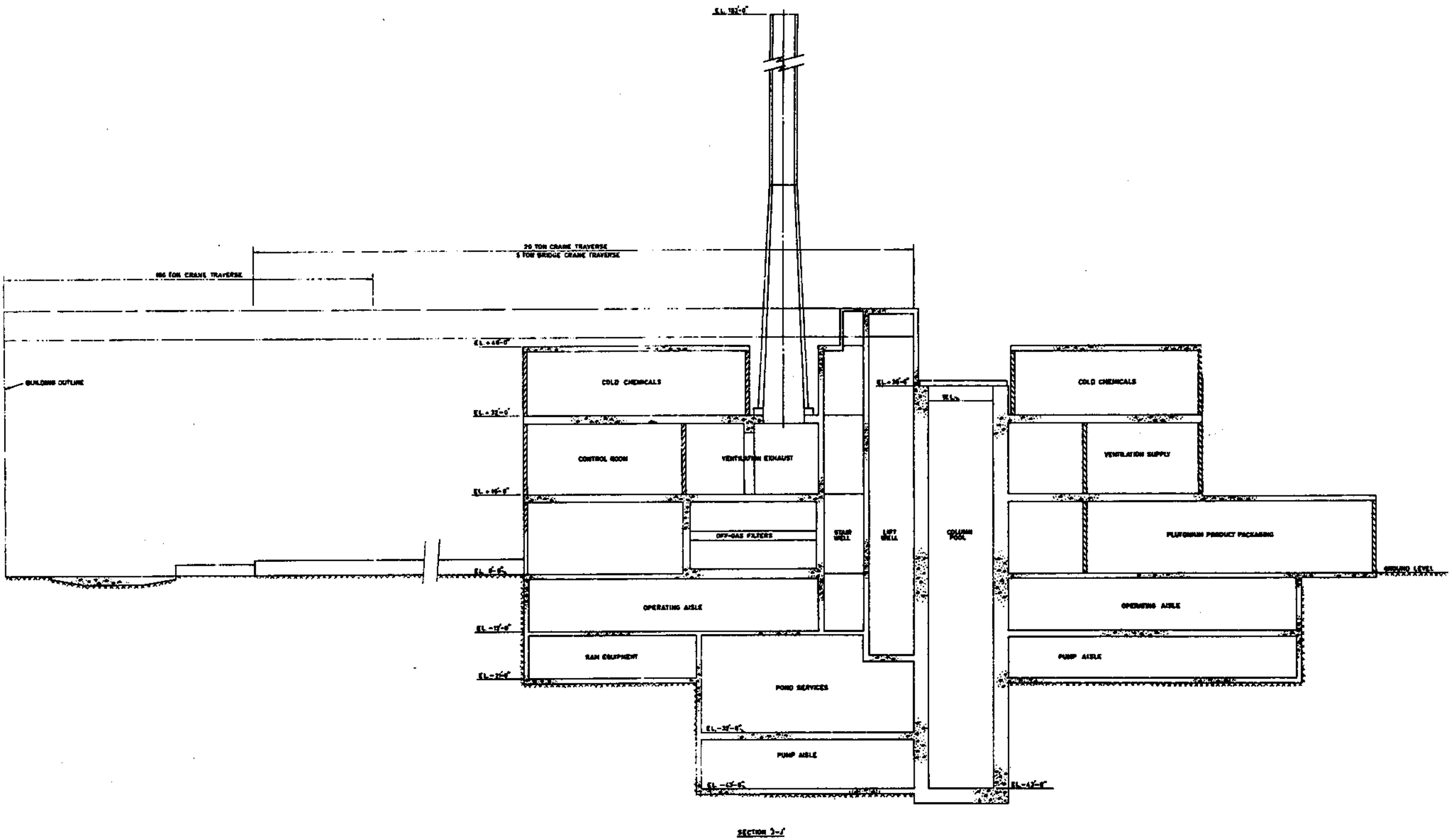
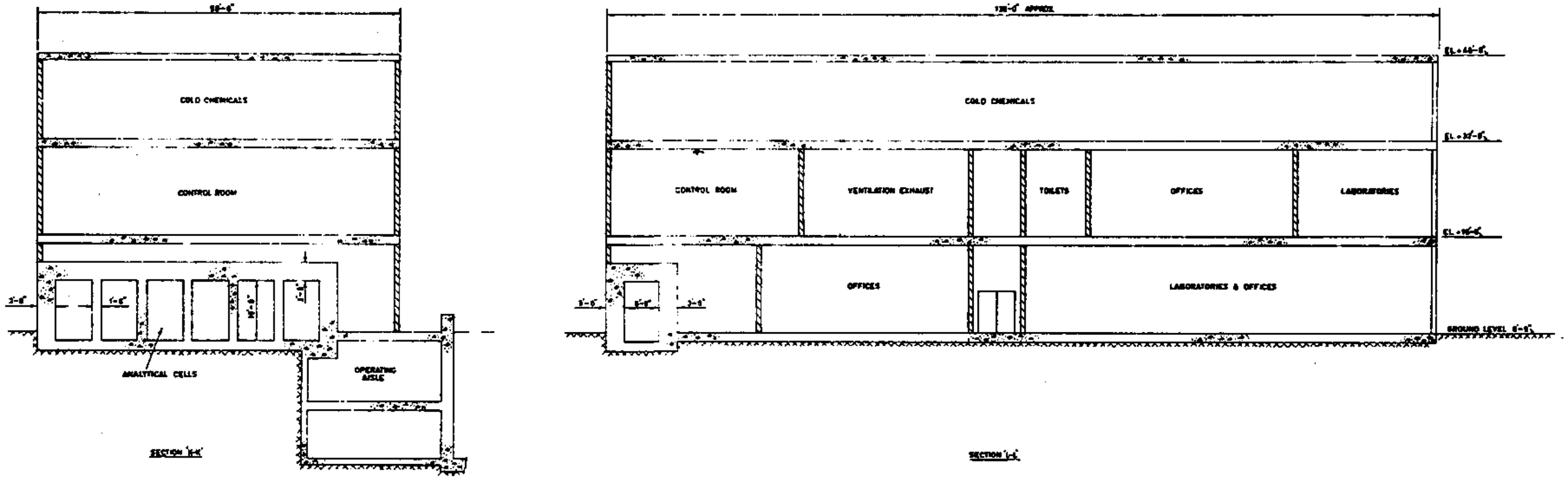


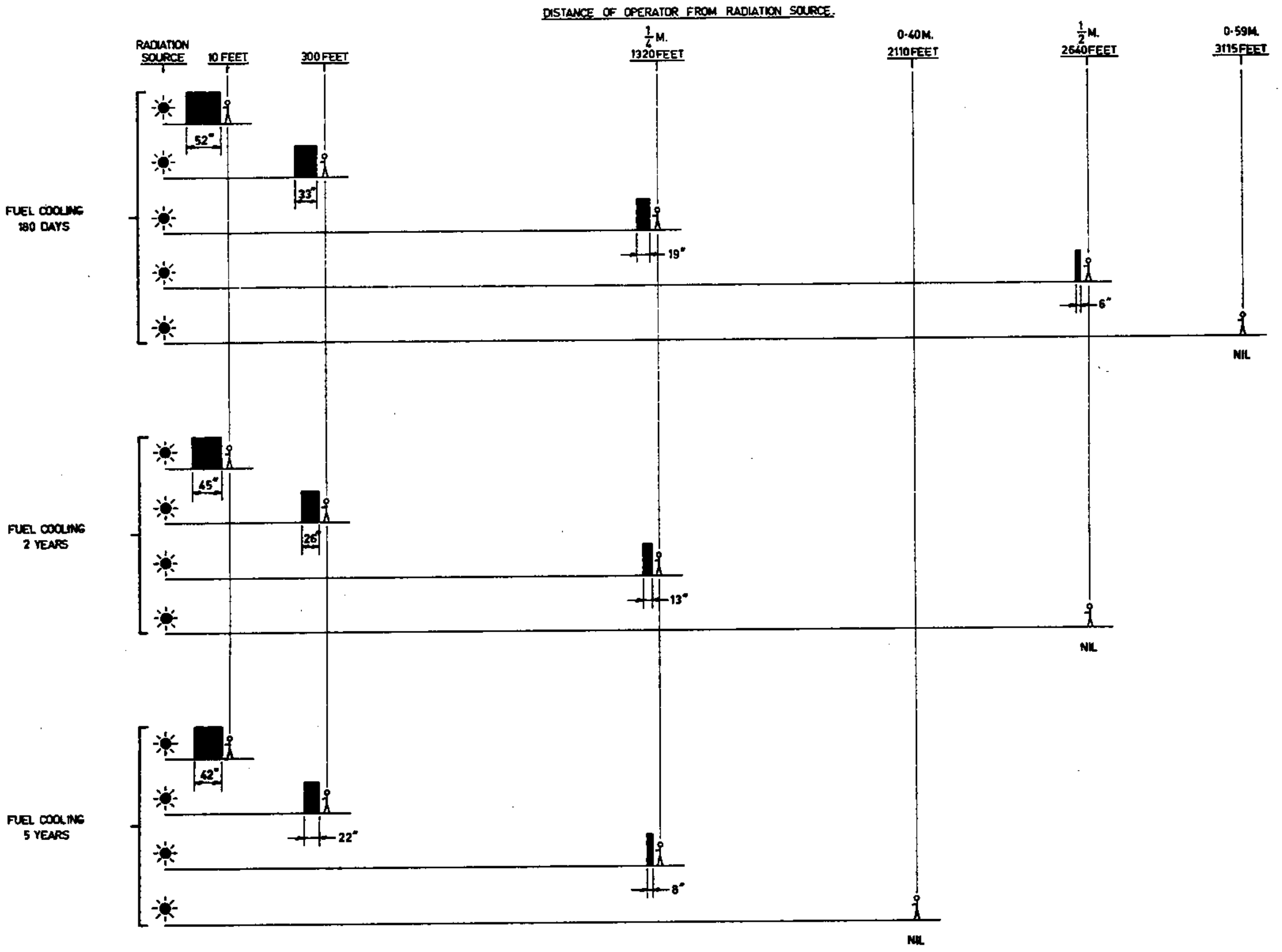
FIGURE 68. BUILDING LAYOUT - SECTION J-J



SECTION 'K-K'

SECTION 'L-L'

FIGURE 69. BUILDING LAYOUT - SECTIONS K-K AND L-L



EFFECT OF COOLING TIME, INVERSE SQUARE LAW, AIR ABSORBTION & CONCRETE THICKNESS ON REDUCTION OF GAMMA RADIATION FROM 1 te of U, 20,000 MWD/teU, 20 MW/teU TO 2·5mR/HR.

FIGURE 70. SHIELDING REQUIREMENTS

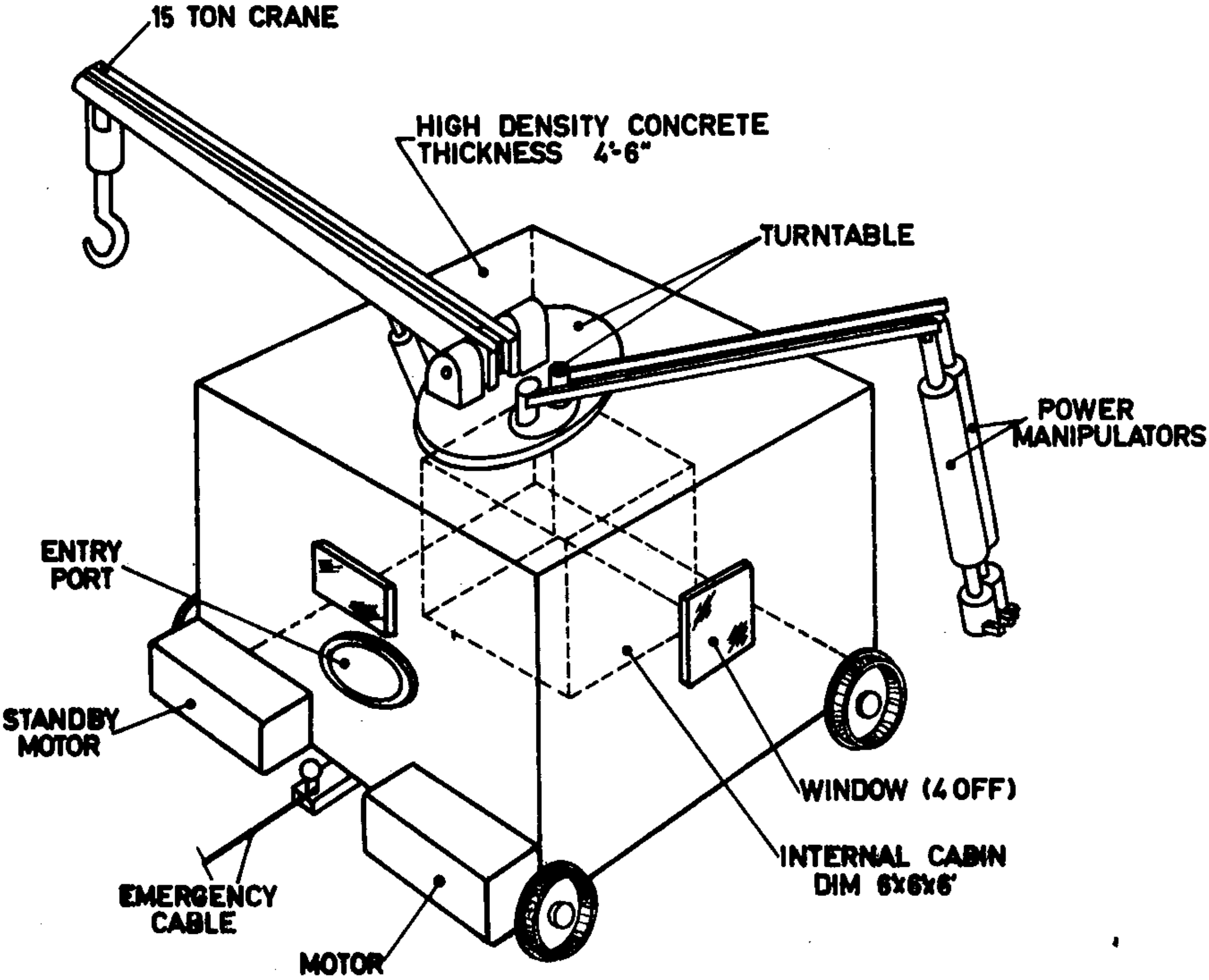
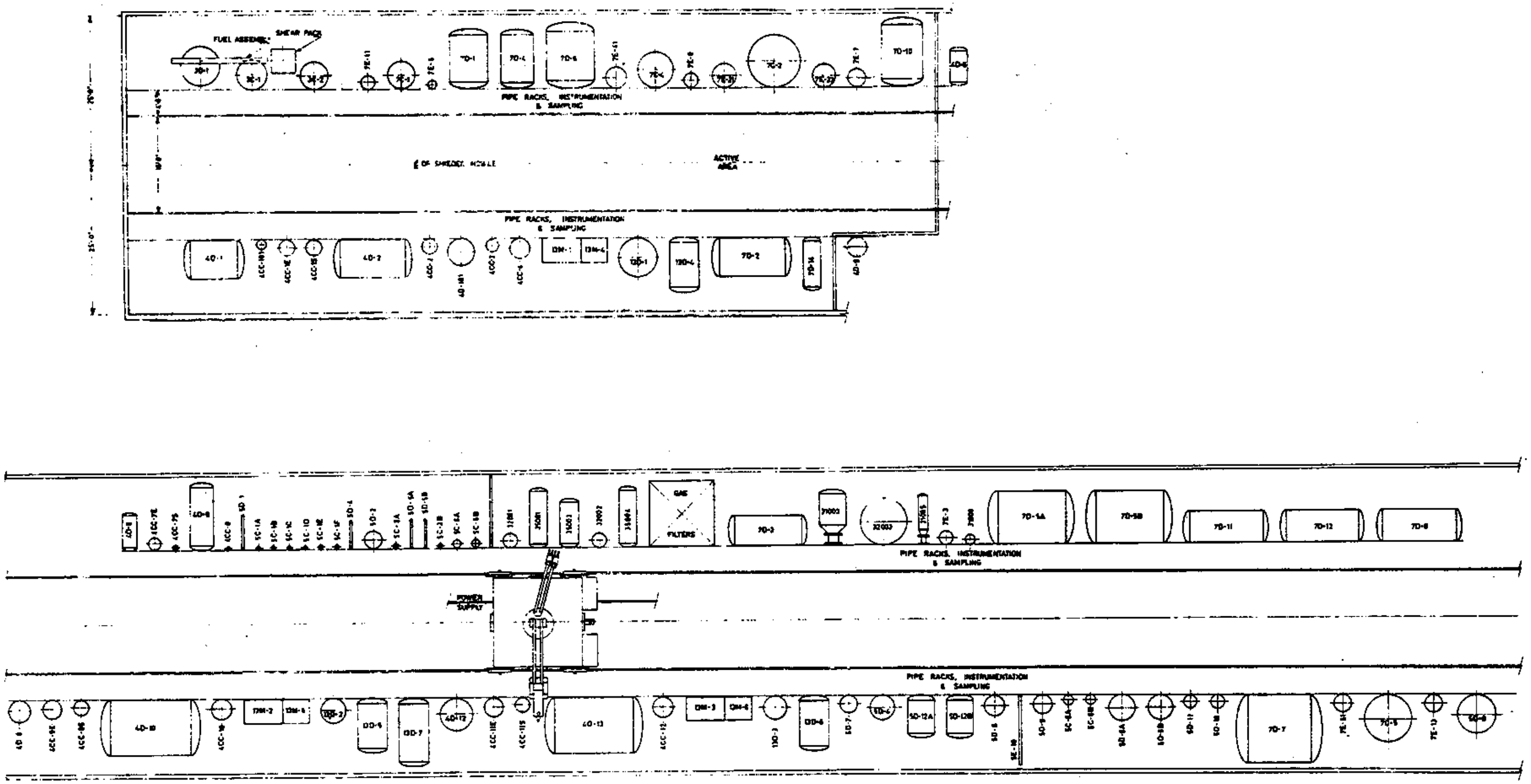


FIGURE 71. MOBILE SHIELDED CABIN



GENERAL PURPOSE EVAPORATION

7C-5	GENERAL PURPOSE EVAPORATOR
7D-7	GENERAL PURPOSE EVAPORATOR FEED TANK
7D-9	CONDENSATE ION EXCHANGER FEED TANK
7E-13	GENERAL PURPOSE EVAPORATOR CONDENSER
7E-18	GENERAL PURPOSE EVAPORATOR REBOILER

FUEL LEACHING

3C-1	No.1 BATCH DISSOLVER
3C-2	No.2 BATCH DISSOLVER
3D-1	ACCOUNTABILITY & FEED ADJUST TANK

PARTITION CYCLE SOLVENT EXTRACTION

4CC-1E	PARTITION CYCLE EXTRACTION CENTRIFUGE
4CC-1S	PARTITION CYCLE SCRUB CENTRIFUGE
4CC-2	PARTITION CYCLE PARTITION CENTRIFUGE
4CC-3	PARTITION CYCLE Pu SCRUB CENTRIFUGE
4CC-4	PARTITION CYCLE STRIP CENTRIFUGE
4CC-101	SOLIDS REMOVAL CENTRIFUGE
13M-1	No.1 CARBONATE WASHER
13M-2	No.1 ACID WASHER
4D-1	PARTITION CYCLE FEED TANK
4D-2	PARTITION CYCLE WASTE CATCH TANK
4D-101	URANIUM CENTRIFUGE DUMP TANK
13D-1	No.1 SOLVENT WASHER
13D-4	No.1 SOLVENT STORAGE TANK

WASTE EVAPORATION

7D-1	HIGH LEVEL WASTE EVAPORATOR FEED TANK
7C-1	HIGH LEVEL WASTE EVAPORATION
7D-4	HIGH LEVEL WASTE ACCOUNTABILITY & NEUTRALIZER TANK
7E-5	HIGH LEVEL WASTE EVAPORATOR CONDENSER
7E-11	HIGH LEVEL WASTE EVAPORATOR REBOILER
7D-2	LOW LEVEL WASTE EVAPORATOR FEED TANK
7D-14	HOT ANALYTICAL CELL DRAIN CATCH TANK
7C-2	LOW LEVEL WASTE EVAPORATOR
7D-10	LOW LEVEL WASTE ACCOUNTABILITY & NEUTRALIZER TANK
7E-7	LOW LEVEL WASTE EVAPORATOR CONDENSER
7E-21	LOW LEVEL WASTE EVAPORATOR 1st REBOILER
7E-22	LOW LEVEL WASTE EVAPORATOR 2nd REBOILER
13D-7	SOLVENT WASTE CATCH TANK
7D-6	REWORK EVAPORATOR FEED TANK
7C-4	REWORK EVAPORATOR
7E-41	REWORK EVAPORATOR REBOILER
7E-6	REWORK EVAPORATOR CONDENSER

1st URANIUM CYCLE SOLVENT EXTRACTION

4CC-8E	1st U. CYCLE EXTRACTION CENTRIFUGE
4CC-8S	1st U. CYCLE SCRUB CENTRIFUGE
4CC-10	1st U. CYCLE STRIP CENTRIFUGE
13M-3	No.2 CARBONATE WASHER
13M-4	No.2 ACID WASHER
4D-9	1st U. CYCLE FEED TANK CONDITIONER
4D-16	1st U. CYCLE WASTE CATCH TANK
13D-2	No.2 SOLVENT WASHER
13D-5	No.2 SOLVENT STORAGE TANK

2nd URANIUM CYCLE SOLVENT EXTRACTION

4CC-11E	2nd U. CYCLE EXTRACTION CENTRIFUGE
4CC-11S	2nd U. CYCLE SCRUB CENTRIFUGE
4CC-12	2nd U. CYCLE STRIP CENTRIFUGE
13M-5	No.3 CARBONATE WASHER
13M-6	No.3 ACID WASHER
4D-12	2nd U. CYCLE FEED TANK CONDITIONER
4D-13	2nd U. CYCLE WASTE CATCH TANK
13D-3	No.3 SOLVENT WASHER
13D-6	No.3 SOLVENT STORAGE TANK

PLUTONIUM CYCLE SOLVENT EXTRACTION

4CC-7E	PLUTONIUM CYCLE EXTRACTION CENTRIFUGE
4CC-7S	PLUTONIUM CYCLE SCRUB CENTRIFUGE
4CC-6	PLUTONIUM CYCLE STRIP CENTRIFUGE
4D-6	PLUTONIUM CYCLE FEED CONDITIONER
4C-9	PLUTONIUM CYCLE STRIP COLUMN
4D-8	PLUTONIUM CYCLE WASTE CATCH TANK

OFF GAS TREATMENT

32001	DISSOLVER OFF-GAS (DOG) ABSORBER
32002	IODINE ABSORBER
35001	DOG ABSORBER SURGE TANK
32003	IODINE STRIPPER
35004	IODINE ABSORBER SURGE TANK

URANIUM FINISHING

5C-6A	No.1 U. SILICA GEL BED
5C-6B	No.2 U. SILICA GEL BED
5C-4	U. PRODUCT EVAPORATOR
5D-7	U. EVAPORATOR FEED TANK
5D-8	U. EVAPORATOR CONDENSATE TANK
5D-9	U. PRODUCT SURGE TANK
5D-10	RINSE ACID HOLD TANK
5D-11	REGENERATION WASTE TANK
5E-16	URANIUM EVAPORATOR PRODUCT COOLER

PLUTONIUM FINISHING

5C-1A	No.1 Pu ANION EXCHANGER
5C-1B	No.2 Pu ANION EXCHANGER
5C-1C	No.3 Pu ANION EXCHANGER
5C-1D	No.4 Pu ANION EXCHANGER
5C-1E	No.5 Pu ANION EXCHANGER
5C-1F	No.6 Pu ANION EXCHANGER
5D-1	Pu IONEX FEED CONDITIONER
5D-2	Pu IONEX WASTE CATCH TANK
5D-4	FEED TANK FOR Pu EVAPORATORS
5D-5A	No.1 Pu STORAGE TANK
5D-5B	No.2 Pu STORAGE TANK
5D-6A	No.1 Pu EVAPORATOR CONDENSATE TANK
5D-6B	No.2 Pu EVAPORATOR CONDENSATE TANK
5C-2A	No.1 Pu PRODUCT EVAPORATOR
5C-2B	No.2 Pu PRODUCT EVAPORATOR

NITRIC ACID RECOVERY

33001	ACID FRACTIONATOR
7D-3	ACID FRACTIONATOR FEED TANK
7D-5A	No.1 RECOVERED ACID STORAGE TANK
7D-5B	No.2 RECOVERED ACID STORAGE TANK
7D-8	WEAK ACID CATCH TANK
7D-11	HOT ACID STORAGE TANK
7D-12	HOT ACID METERING TANK
7E-3	FRACTIONATOR CONDENSER
31003	ACID FRACTIONATOR FEED VAPORIZER
31005	FRACTIONATOR REBOILER
31006	ACID FRACTIONATOR BOTTOMS COOLER

FIGURE 72. PROCESS EQUIPMENT LAYOUT

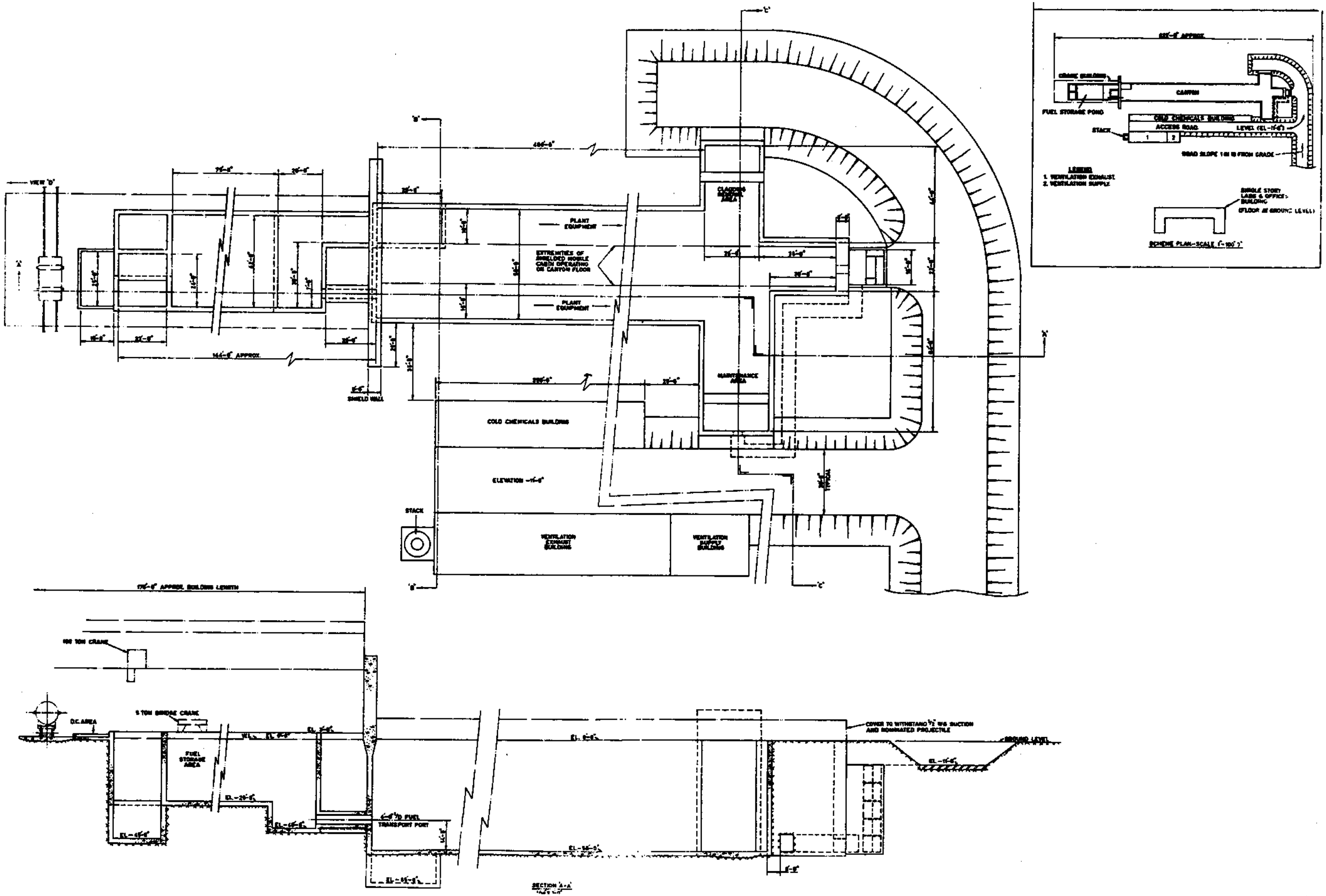


FIGURE 73. BUILDING EXCAVATION - PLAN AND ELEVATION

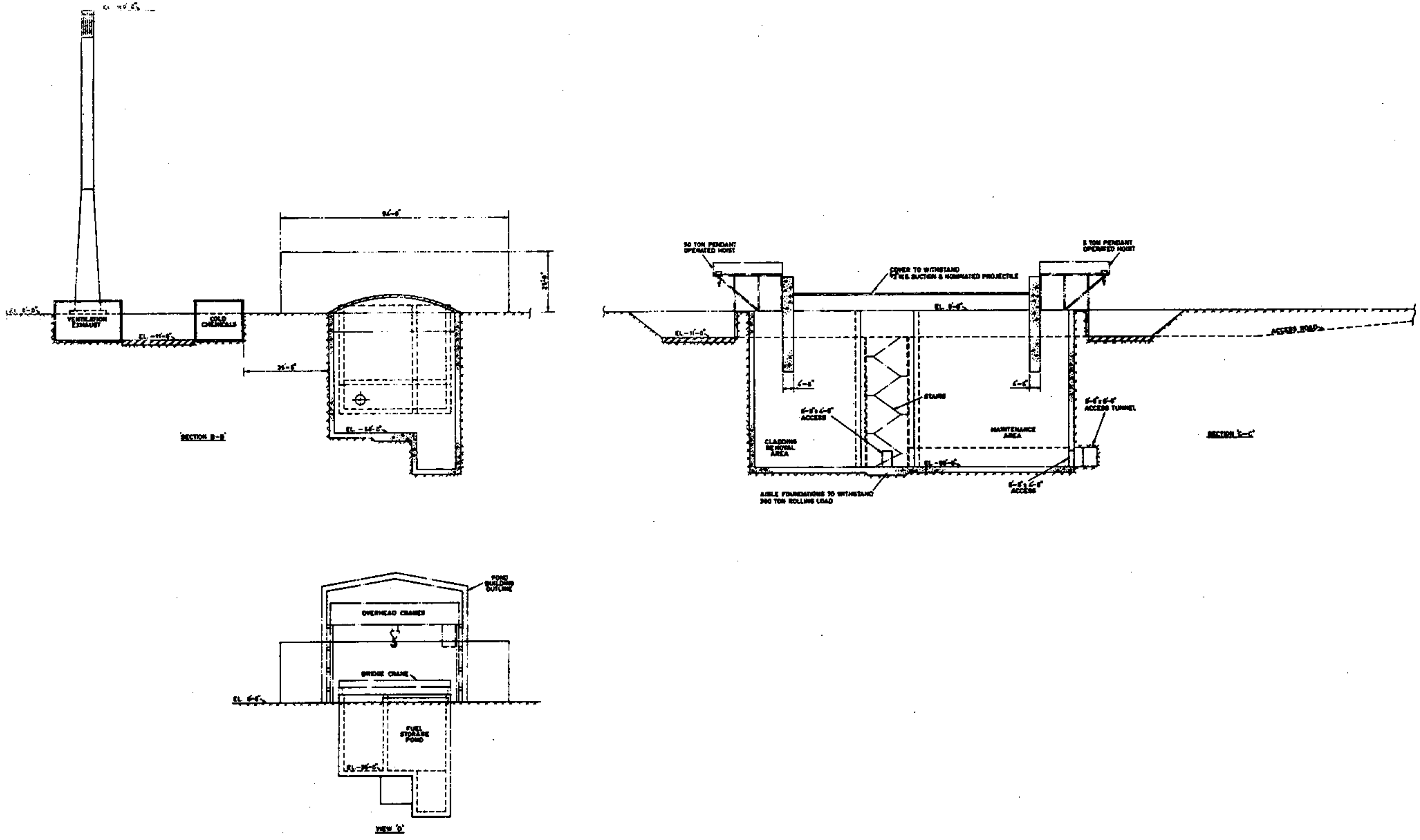


FIGURE 74. BUILDING EXCAVATION - SECTIONS B-B, C-C AND VIEW D

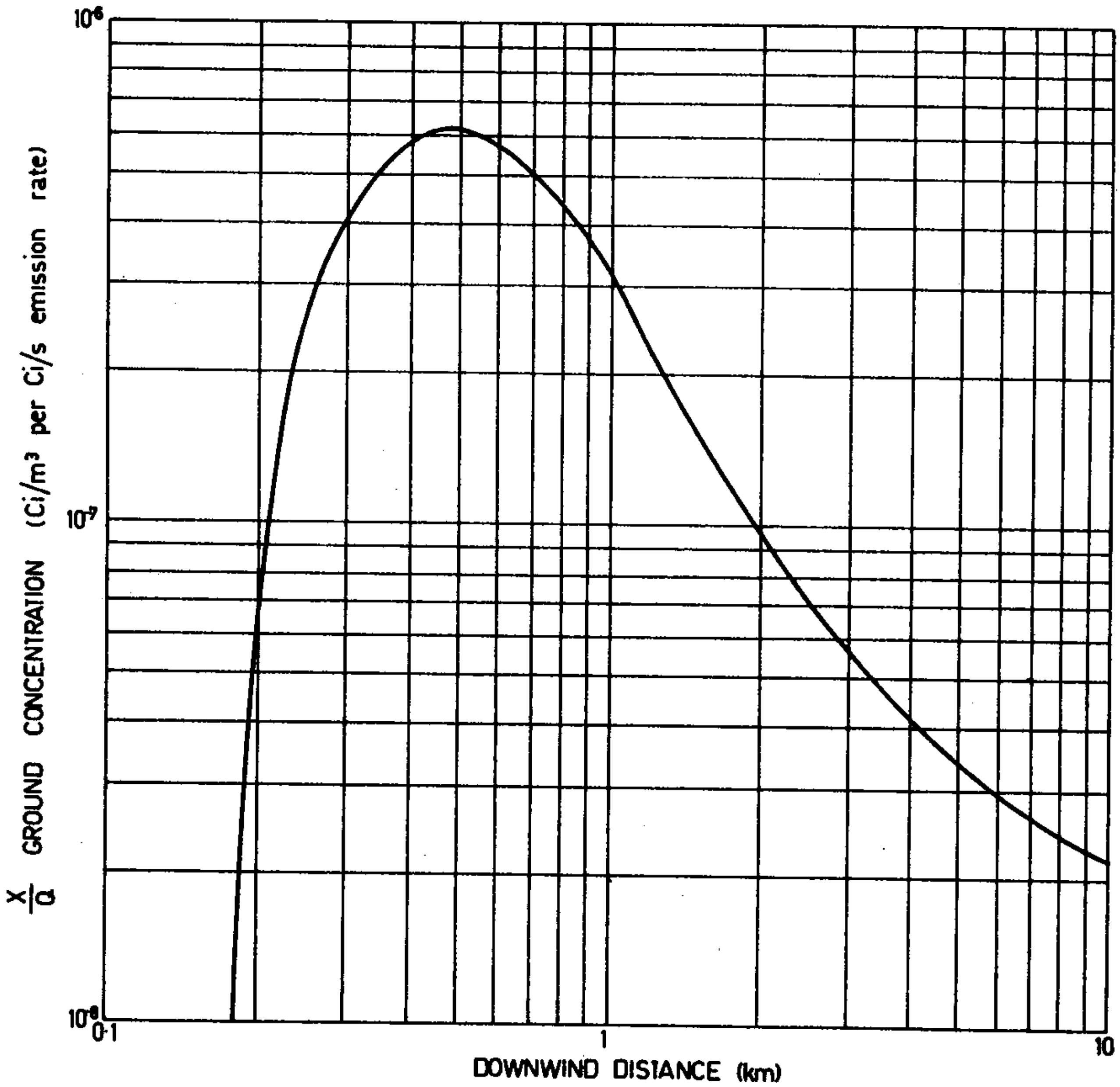


FIGURE 75. EFFECT OF DOWN WIND DISTANCE ON THE AVERAGE GROUND CONCENTRATION

LOCATION : LUCAS HEIGHTS, STACK HEIGHT = 65m, WIND DIRECTION: SSE.

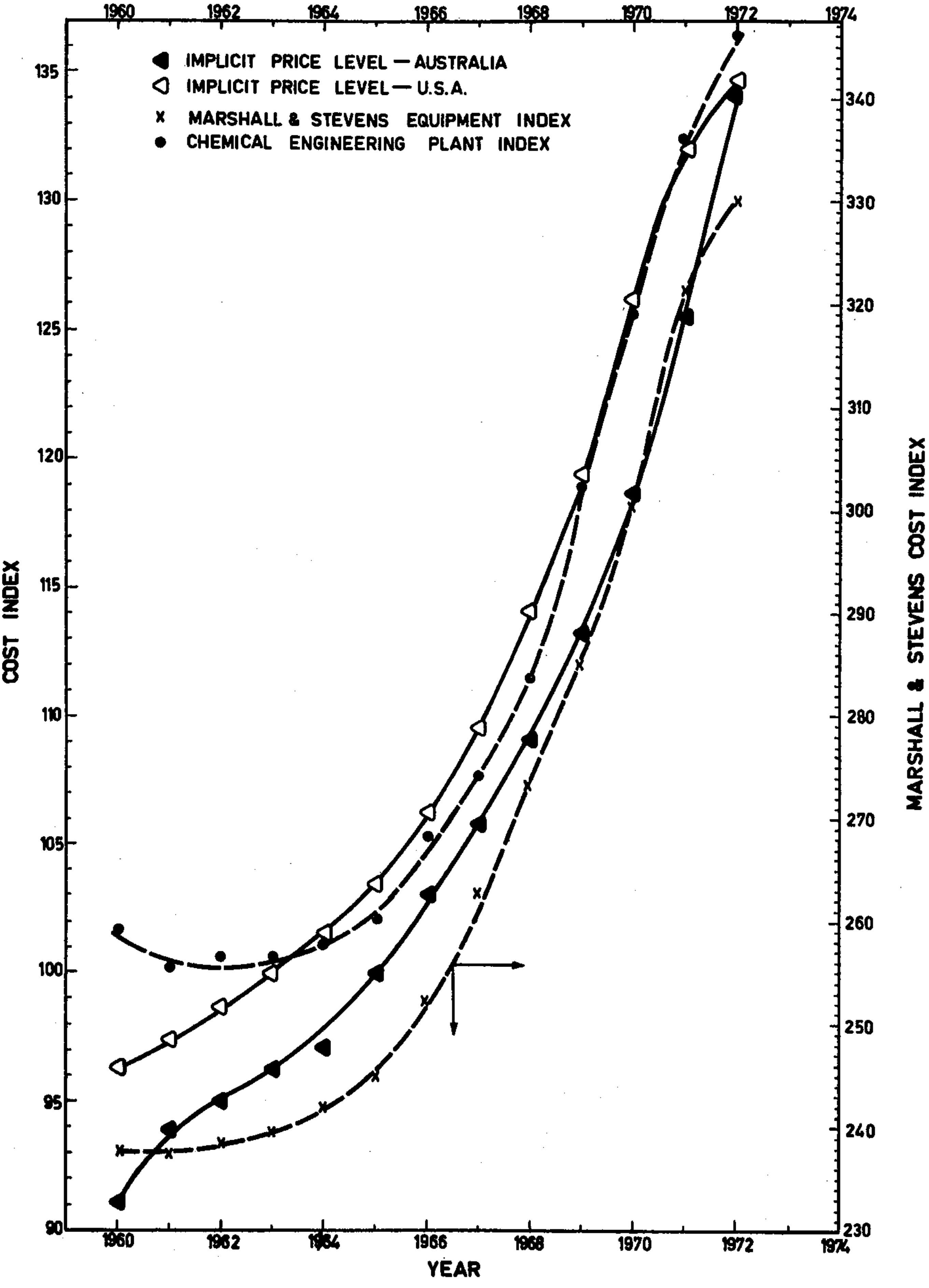
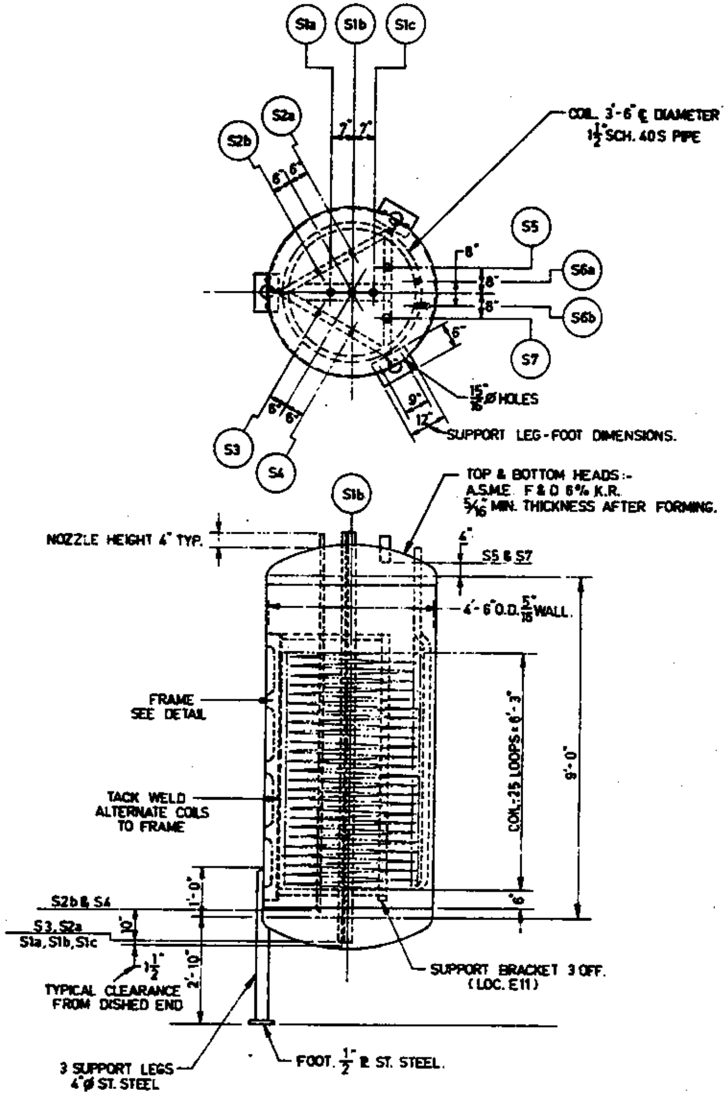
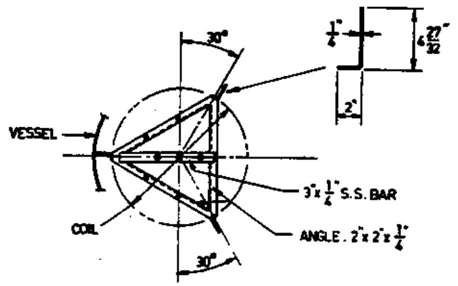
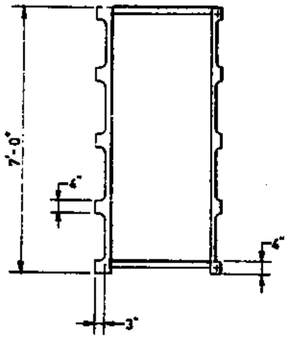


FIGURE 76. COMPARISON OF COST INDICES

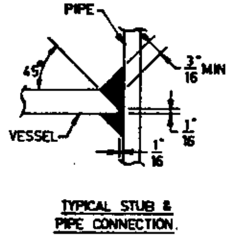


SCHEMATIC ELEVATION.

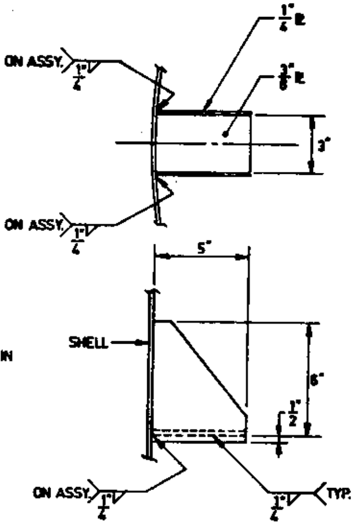




FRAME ASSEMBLY.



TYPICAL STUB & PIPE CONNECTION.



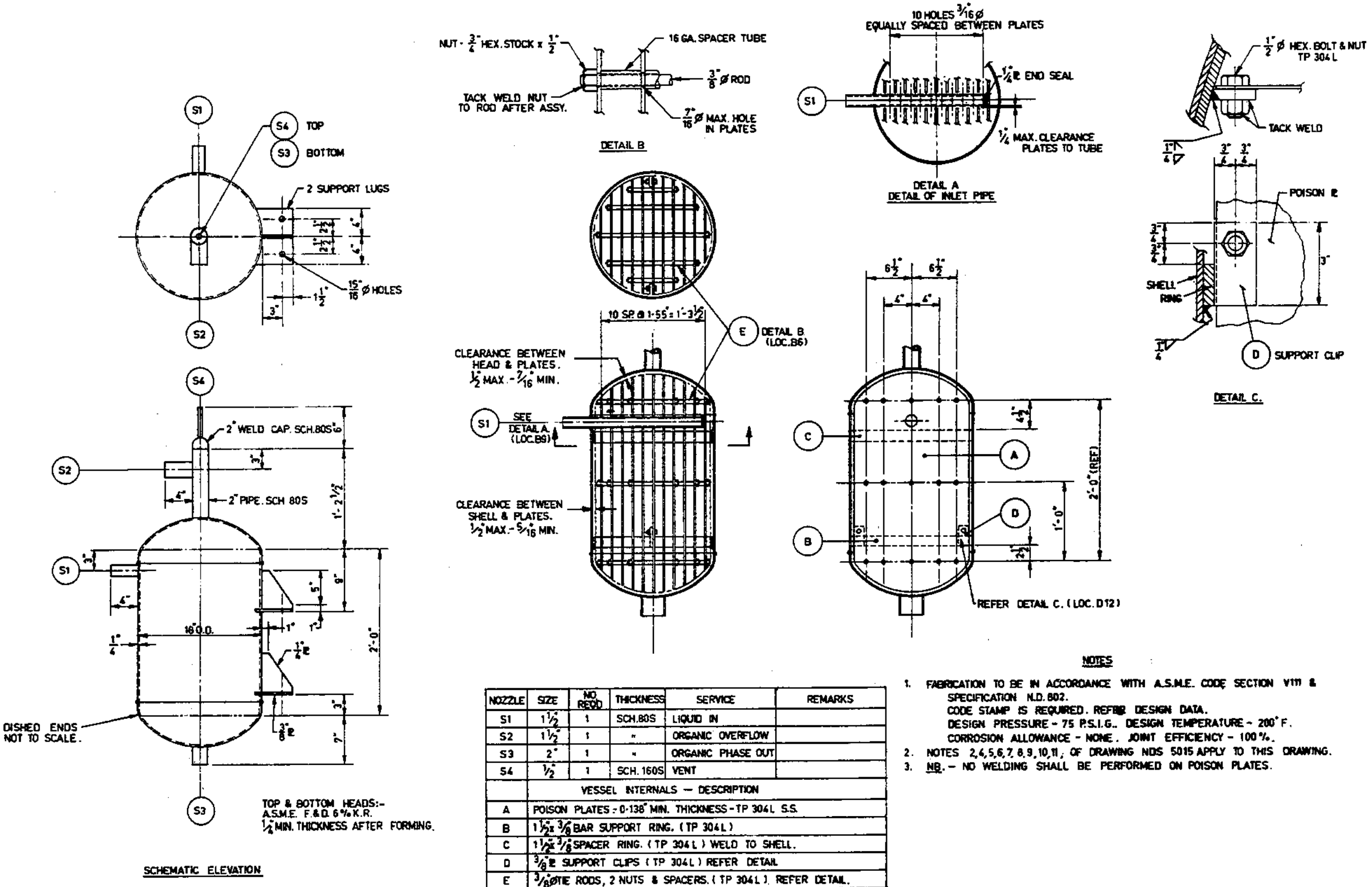
SUPPORT BRACKET.

NOZZLE	SIZE	NO. REQD	CODE	FACING INT.	FACING EXT.	SERVICE.
S1a	2"	1	SCH. 80S.	SQUARE	REF. NOTE 11	OUTLET
S1b	2"	1	DO.	DO.	DO.	DO
S1c	2"	1	DO.	DO.	DO.	DO. DRAIN
S2a	½"	1	SCH. 160S.	45° C/OFF	DO.	LEVEL RECORDER
S2b	½"	1	DO.	DO.	DO.	DO.
S3	½"	1	DO.	SQUARE	DO.	SAMPLE
S4	¾"	1	DO.	WELD CAP.	DO.	TEMP. RECORDER.
S5	2"	1	SCH. 80S.	SQUARE	DO.	INLET
S6a	1½"	1	SCH. 40S.	—	DO.	COIL - OUTLET
S6b	1½"	1	DO.	—	DO.	COIL - INLET
S7	3"	1	SCH. 40S	SQUARE	DO.	VENT.

NOTES.

1. FABRICATION TO BE IN ACCORDANCE WITH A.S.M.E. CODE SECTION VIII & SPECIFICATION NO 802.
   CODE STAMP IS REQUIRED. REFER DESIGN DATA.
   DESIGN PRESSURE - 75 P.S.I.G. DESIGN TEMPERATURE - 200° F.
   CORROSION ALLOWANCE - NONE. JOINT. EFFICIENCY - 100%
2. NOTES 2,4,5,6,7,8,9,11, OF DRAWING NDS 5015 APPLY TO THIS DRAWING.

FIGURE 77. CONDENSATE CATCH TANK 3D-2 - GENERAL DETAILS

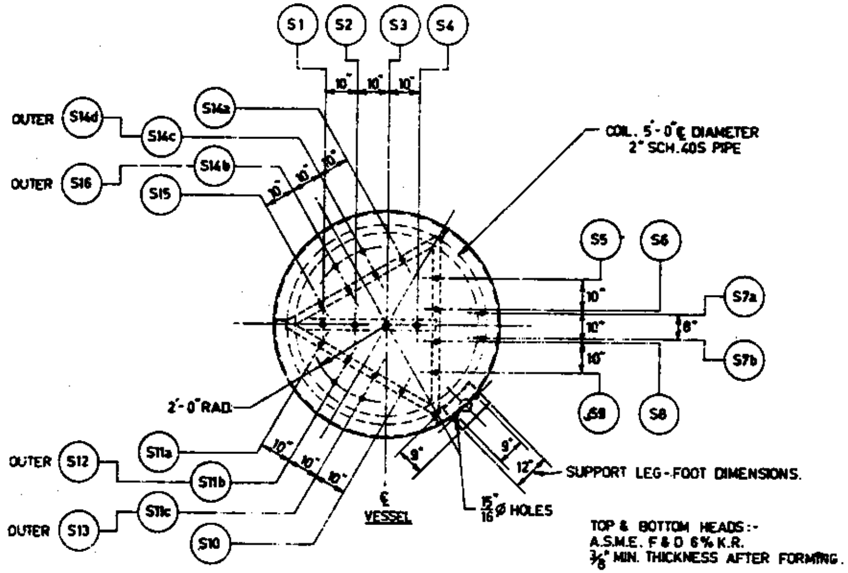


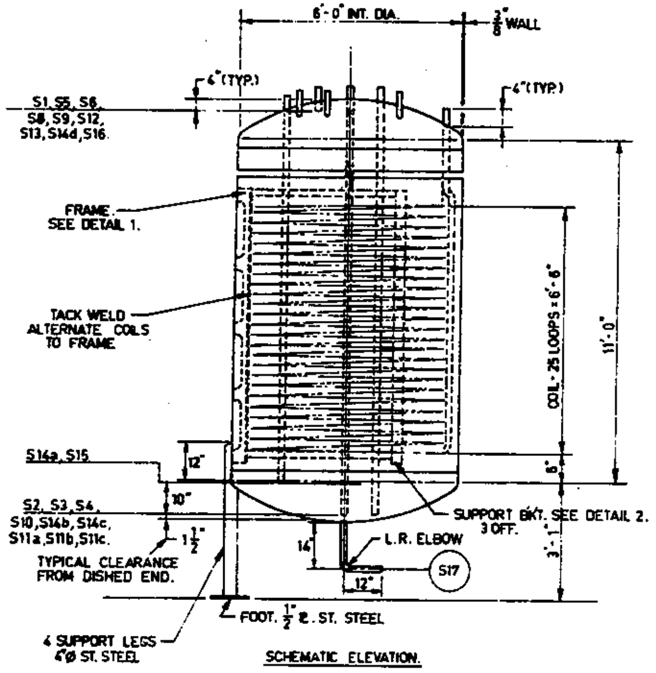
NOZZLE	SIZE	NO REQD	THICKNESS	SERVICE	REMARKS
S1	1½"	1	SCH.80S	LIQUID IN	
S2	1½"	1	"	ORGANIC OVERFLOW	
S3	2"	1	"	ORGANIC PHASE OUT	
S4	½"	1	SCH. 160S	VENT	
	VESSEL INTERNALS — DESCRIPTION				
A	POISON PLATES - 0·138" MIN. THICKNESS - TP 304L S.S.				
B	1½" x 3/8" BAR SUPPORT RING. (TP 304L)				
C	1½" x 3/8" SPACER RING. (TP 304L) WELD TO SHELL.				
D	3/8" Ⓡ SUPPORT CLIPS (TP 304L) REFER DETAIL				
E	3/8" Ø TIE RODS, 2 NUTS & SPACERS. (TP 304L). REFER DETAIL.				

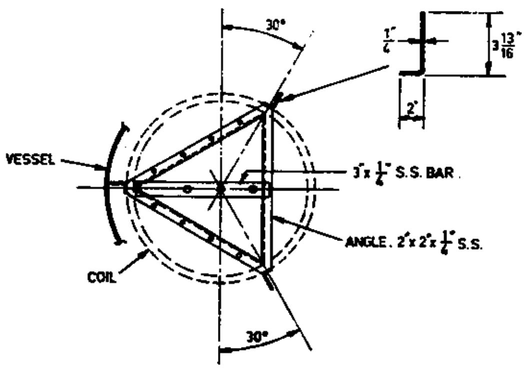
NOTES

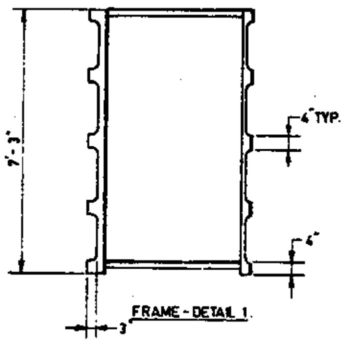
1. FABRICATION TO BE IN ACCORDANCE WITH A.S.M.E. CODE SECTION VIII & SPECIFICATION N.D. 802.
   CODE STAMP IS REQUIRED. REFER DESIGN DATA.
   DESIGN PRESSURE - 75 P.S.I.G.. DESIGN TEMPERATURE - 200° F.
   CORROSION ALLOWANCE - NONE. JOINT EFFICIENCY - 100%.
2. NOTES 2,4,5,6,7,8,9,10,11, OF DRAWING NDS 5015 APPLY TO THIS DRAWING.
3. NB. - NO WELDING SHALL BE PERFORMED ON POISON PLATES.

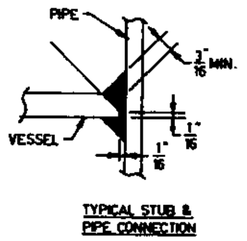
FIGURE 78. DECANTER 4Y-1 - GENERAL DETAILS

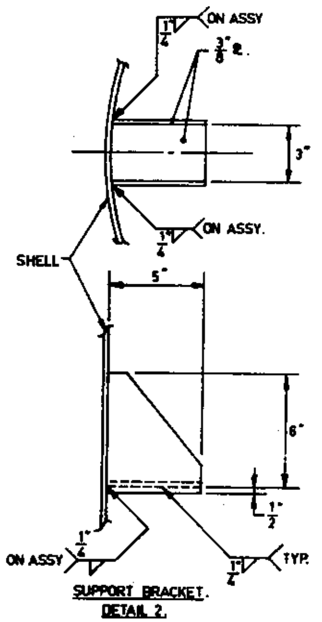










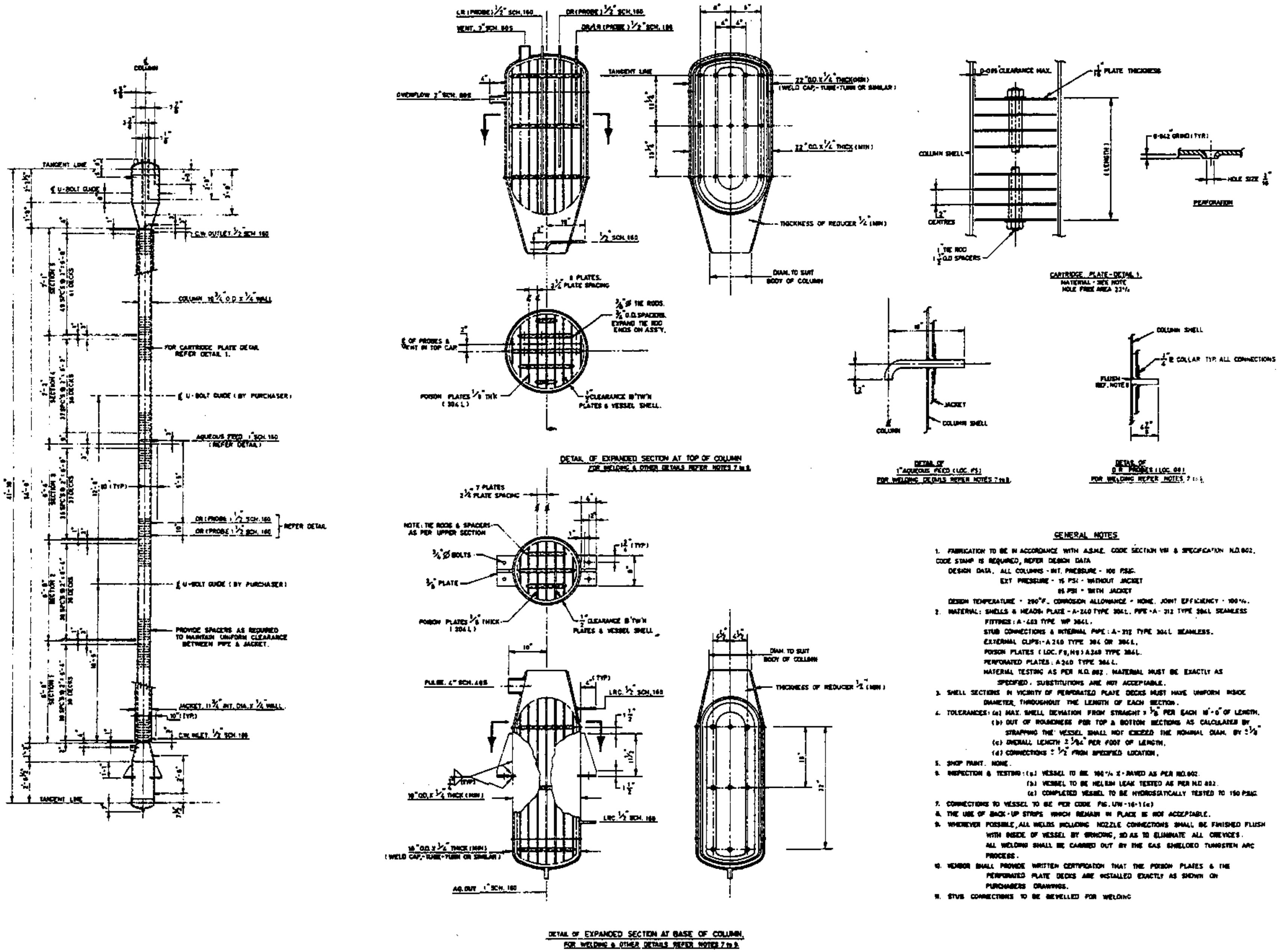


NOZZLE	SIZE	NO REQD	CODE	FACING INT.	FACING EXT.	SERVICE.
S1	3"	1	SCH. 40S	SQUARE	REF. NOTE 11.	VENT
S2	1"	1	SCH. 160S	DO.	DO.	AIR SPARGE
S3	1 1/2"	1	SCH. 80S	DO.	DO.	INLET.-
S4	1 1/2"	1	DO.	DO.	DO.	INLET.-
S5	2"	1	DO.	DO.	DO.	INLET.-
S6	2"	1	DO.	DO.	DO.	INLET.-
S7a	2"	—	SCH. 40S	—	DO.	COOLING COIL.
S7b	2"	—	DO.	—	DO.	DO.
S8	2"	1	SCH. 80S	SQUARE	DO.	INLET.-
S9	1 1/2"	1	DO.	DO.	DO.	INLET.- MISC. ADD'.
S10	1"	1	SCH. 160S	DO.	DO.	
S11a	1/2"	1	DO.	DO.	DO.	SAMPLE.
S11b	1/2"	1	DO.	DO.	DO.	DO.
S11c	1/2"	1	DO.	DO.	DO.	DO.
S12	1"	1	DO.	DO.	DO.	INLET.- PH. SEP.
S13	1 1/2"	1	SCH. 80S	DO.	DO.	INLET.-
S14a	1/2"	1	SCH. 160S	45° C/OFF	DO.	INSTRUMENT.
S14b	1/2"	1	DO.	DO.	DO.	DO.
S14c	1/2"	1	DO.	DO.	DO.	DO.
S14d	1/2"	1	DO.	SQUARE	DO.	DO.
S15	3/4"	1	DO.	WELD CAP.	DO.	TEMP. RECORDER.
S16	1/2"	1	DO.	SQUARE	DO.	INLET.-
S17	1"	1	DO.	DO	DO.	OUTLET.

NOTES

1. FABRICATION TO BE IN ACCORDANCE WITH A.S.M.E CODE SECTION VIII & ND 802.
   CODE STAMP IS REQUIRED, REFER DESIGN DATA.
   DESIGN PRESSURE - 75 P.S.I.G. DESIGN TEMPERATURE - 200° F.
   CORROSION ALLOWANCE - NONE. JOINT EFFICIENCY - 100%.
2. NOTES 2,4,5,6,7,8,9,11, OF DRAWING NDS 5015 APPLY TO THIS DRAWING

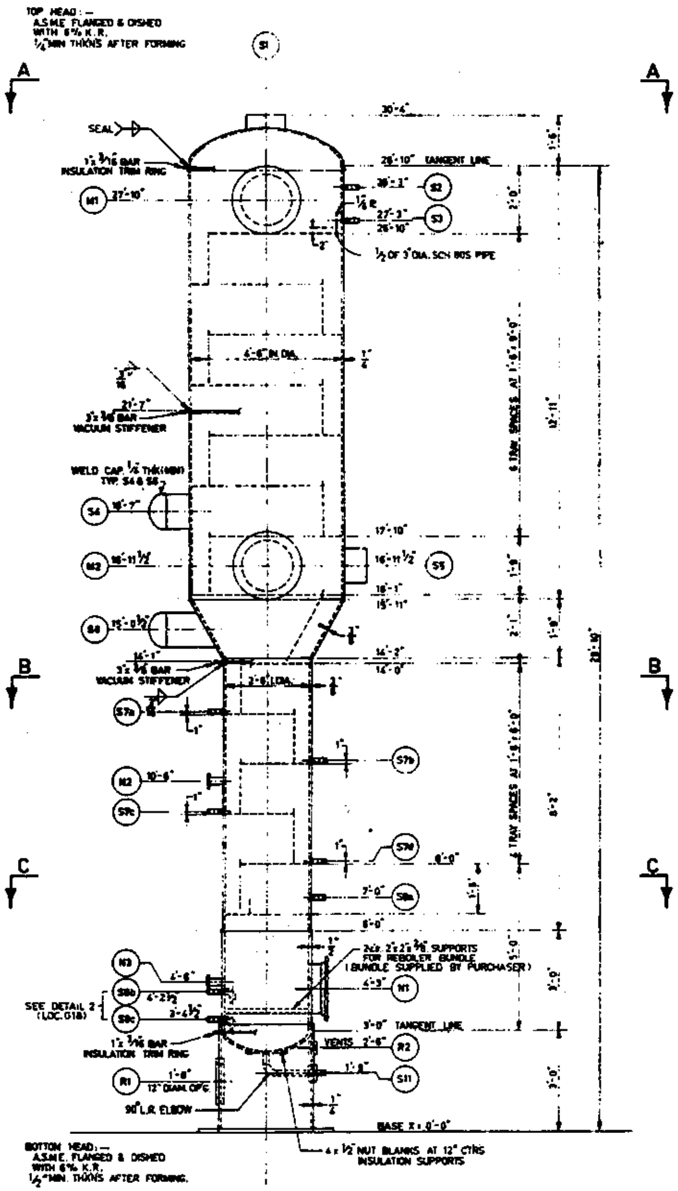
FIGURE 79. PARTITION CYCLE FEED TANK 4D-1 - GENERAL DETAILS



GENERAL NOTES

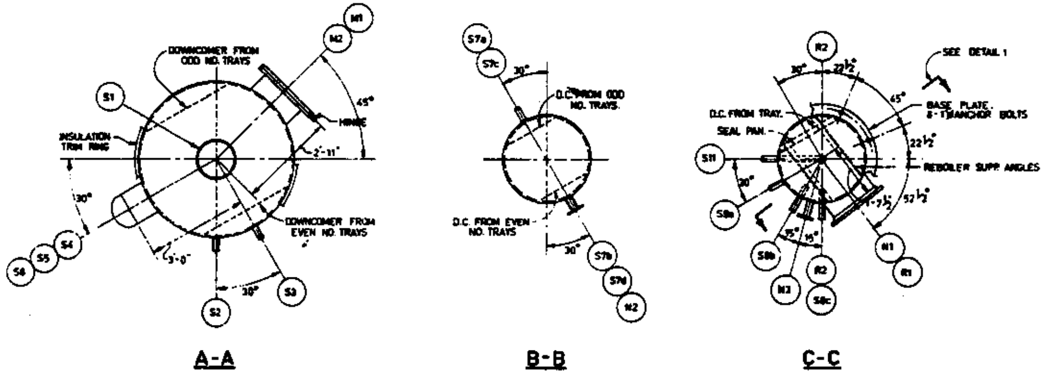
1. FABRICATION TO BE IN ACCORDANCE WITH A.S.M.E. CODE SECTION VIII & SPECIFICATION N.O. 802. CODE STAMP IS REQUIRED, REFER DESIGN DATA
DESIGN DATA. ALL COLUMNS - INT. PRESSURE - 100 PSIG.
EXT PRESSURE - 15 PSI - WITHOUT JACKET
85 PSI - WITH JACKET
DESIGN TEMPERATURE - 250°F. CORROSION ALLOWANCE - NONE. JOINT EFFICIENCY - 100%.
2. MATERIAL: SHELLS & HEADS: PLATE - A-240 TYPE 304L. PIPE - A-312 TYPE 304L SEAMLESS
FITTINGS: A-403 TYPE WP 304L.
STUB CONNECTIONS & INTERNAL PIPE: A-312 TYPE 304L SEAMLESS.
EXTERNAL CLIPS: - A 240 TYPE 304 OR 304L.
POISON PLATES (LOC. F8, H8) A240 TYPE 304L.
PERFORATED PLATES: A240 TYPE 304L.
MATERIAL TESTING AS PER N.O. 802. MATERIAL MUST BE EXACTLY AS SPECIFIED. SUBSTITUTIONS ARE NOT ACCEPTABLE.
3. SHELL SECTIONS IN VICINITY OF PERFORATED PLATE DECKS MUST HAVE UNIFORM INSIDE DIAMETER THROUGHOUT THE LENGTH OF EACH SECTION.
4. TOLERANCES: (a) MAX. SHELL DEVIATION FROM STRAIGHT ± 1/8" PER EACH 10'-0" OF LENGTH.
(b) OUT OF ROUNDNESS FOR TOP & BOTTOM SECTIONS AS CALCULATED BY STRAPPING THE VESSEL SHALL NOT EXCEED THE NOMINAL DIAM. BY ± 1/8"
(c) OVERALL LENGTH ± 1/64" PER FOOT OF LENGTH.
(d) CONNECTIONS ± 1/2" FROM SPECIFIED LOCATION.
5. SHOP PAINT. NONE.
6. INSPECTION & TESTING: (a) VESSEL TO BE 100% X-RAYED AS PER NO.802.
(b) VESSEL TO BE HELIUM LEAK TESTED AS PER N.O 802.
(c) COMPLETED VESSEL TO BE HYDROSTATICALLY TESTED TO 150 PSIG.
7. CONNECTIONS TO VESSEL TO BE PER CODE FIG. UW-16-1(c)
8. THE USE OF BACK-UP STRIPS WHICH REMAIN IN PLACE IS NOT ACCEPTABLE.
9. WHENEVER POSSIBLE, ALL WELDS INCLUDING NOZZLE CONNECTIONS SHALL BE FINISHED FLUSH WITH INSIDE OF VESSEL BY GRINDING, SO AS TO ELIMINATE ALL CREVICES.
ALL WELDING SHALL BE CARRIED OUT BY THE GAS SHIELDED TUNGSTEN ARC PROCESS.
10. VENDOR SHALL PROVIDE WRITTEN CERTIFICATION THAT THE POISON PLATES & THE PERFORATED PLATE DECKS ARE INSTALLED EXACTLY AS SHOWN ON PURCHASERS DRAWINGS.
11. STUB CONNECTIONS TO BE BEVELLED FOR WELDING

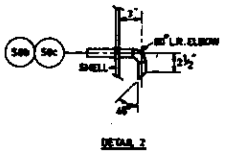
FIGURE 80. PULSED COLUMN 4C-1 - ELEVATION & GENERAL DETAILS



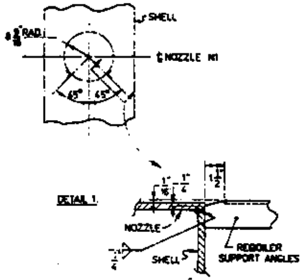
SCHEMATIC ELEVATION.

TRUE ORIENTATIONS SHOWN IN SECTIONS
ALL ELEVATIONS ARE TO BASE 'X'.
TRAY ELEVATIONS ARE TO TOP OF TRAY SUPPORT RING.





DETAIL 2



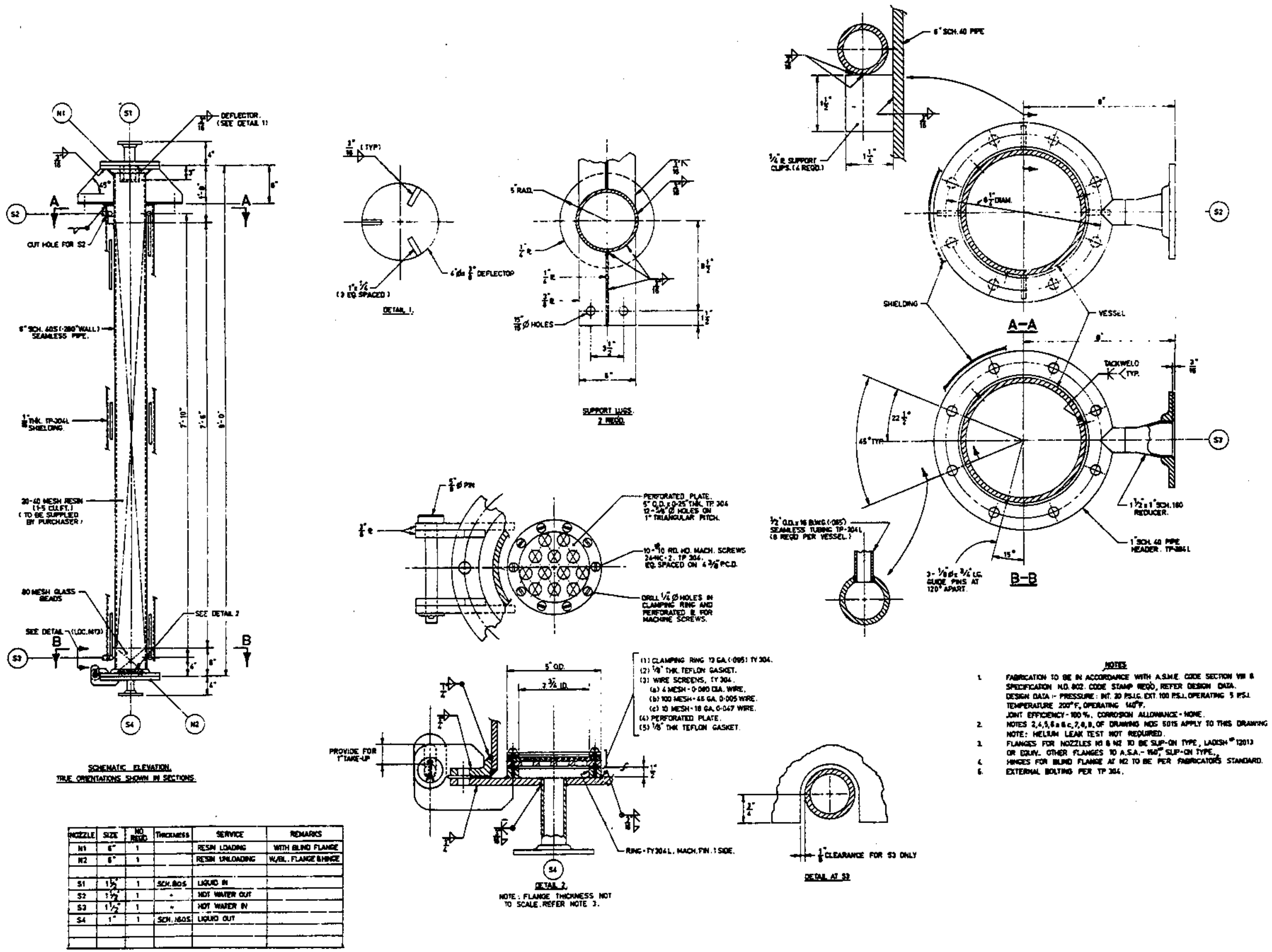
DETAIL 1.

NOZZLE	SIZE	NO. REQD.	WALL THICKNESS	SERVICE	REMARKS
M1	16"	1	1/4"	MANHOLE	W/HINGES & HANDLE
M2	18"	1	-	MANHOLE	DO.
N1	18"	1	1/2"	REBOILER	WITH FLANGE
N2	1 1/2"	1	1/4"	T.R.C.	W/FLANGE
N3	1 1/2"	1	-	T.R.	DO.
S1	14"	1	0·250"	VAPOR OUTLET	
S2	1"	1	SCH.180	P.R.C.	
S3	1"	1	SCH.180	REFLUX INLET	
S4	12"	1	1/4"	VAPOR FEED - ALT.	
S5	12"	1	1/4"	VAPOR FEED	
S6	12"	1	3/8"	VAPOR FEED - ALT.	
S7	1"	4	0·250"	CHLORIDE DRAIN	
S9	1"	3	0·250"	L.R. & D.R.	
S11	1 1/2"	1	0·261"	BOTTOM OUTLET.	

NOTES.

1. FABRICATION TO BE IN ACCORDANCE WITH A.S.M.E. CODE SECTION VIII & SPECIFICATION NO 802.
   CODE STAMP IS REQUIRED. REFER DESIGN DATA.
   DESIGN PRESSURE - 50 P.S.I.G. DESIGN TEMPERATURE - 280° F.
   CORROSION ALLOWANCE - NONE. JOINT EFFICIENCY - 100%.
2. NOTES 2,4,5,6a & c,7,8,9,10,11 OF DRAWING NOS 5815 APPLY TO THIS DRAWING
3. VALVE TRAYS TO BE 14 S.W.G., 304L STAINLESS STEEL ('KOCH' FLEXITRAY, 'HYDRONYL' TYPE NR-1 CONE VALVE OR SIMILAR).
   FLOW AREA 10% OF TRAY AREA. 1/2" HOLES ON TRIANGULAR PITCH.
4. SKIRT & BASE MATERIAL TP-304 OR 304L.
5 FLANGES FOR MANHOLES M1 & M2 TO BE SLIP-ON, LADISH PART NO 12411 OR EQUIVALENT.
   BLIND FLANGES TO BE SOLID ALLOY WITH THICKNESS EQUAL TO THAT OF THE MATING FLANGE.

FIGURE 81. NITRIC ACID FRACTIONATOR - GENERAL DETAILS



NOZZLE	SIZE	NO. REQD	THICKNESS	SERVICE	REMARKS
N1	6"	1		RESIN LOADING	WITH BLIND FLANGE
N2	6"	1		RESIN UNLOADING	W/BL. FLANGE & HINGE
S1	1 1/2"	1	SCH. 80S	LIQUID IN	
S2	1 1/2"	1	"	HOT WATER OUT	
S3	1 1/2"	1	"	HOT WATER IN	
S4	1"	1	SCH. 160S	LIQUID OUT	

NOTES

1. FABRICATION TO BE IN ACCORDANCE WITH A.S.M.E. CODE SECTION VIII & SPECIFICATION N.D. 802. CODE STAMP REQD., REFER DESIGN DATA.
   DESIGN DATA :- PRESSURE: INT. 20 P.S.I.G. EXT. 100 P.S.I., OPERATING 5 P.S.I.
   TEMPERATURE 200°F, OPERATING 140°F.
   JOINT EFFICIENCY - 100%. CORROSION ALLOWANCE - NONE.
2. NOTES 2,4,5,6a & 6c, 7, 8, 9, OF DRAWING NOS 5015 APPLY TO THIS DRAWING
   NOTE: HELIUM LEAK TEST NOT REQUIRED.
3. FLANGES FOR NOZZLES N1 & N2 TO BE SLIP-ON TYPE, LADISH # 12013 OR EQUIV. OTHER FLANGES TO A.S.A. - 150# SLIP-ON TYPE.
4. HINGES FOR BLIND FLANGE AT N2 TO BE PER FABRICATOR'S STANDARD.
6. EXTERNAL BOLTING PER TP 304.

FIGURE 82. PLUTONIUM ION EXCHANGER - GENERAL DETAILS

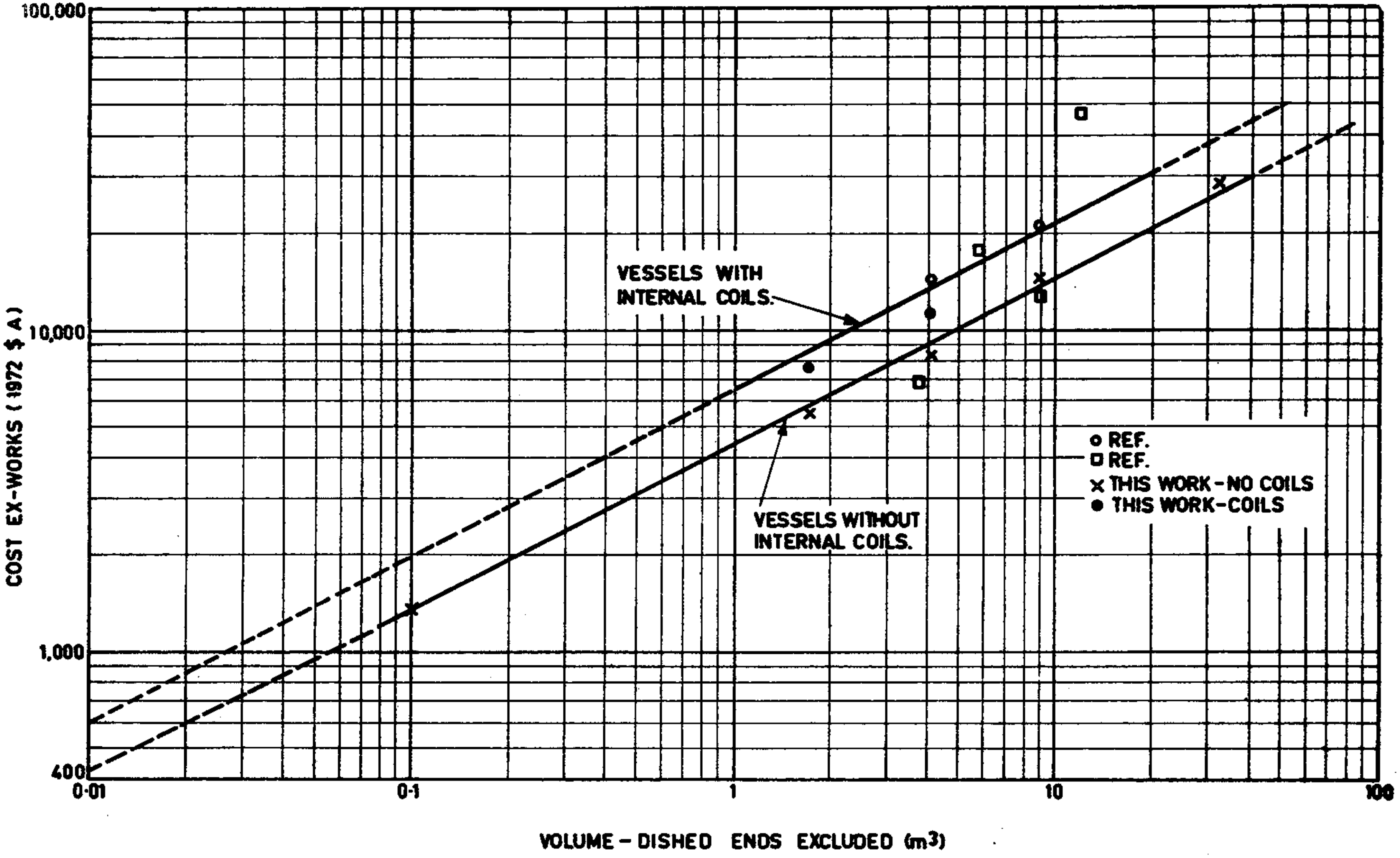


FIGURE 83. COST OF VESSELS FABRICATED TO RADIOCHEMICAL STANDARDS IN 304L STAINLESS STEEL

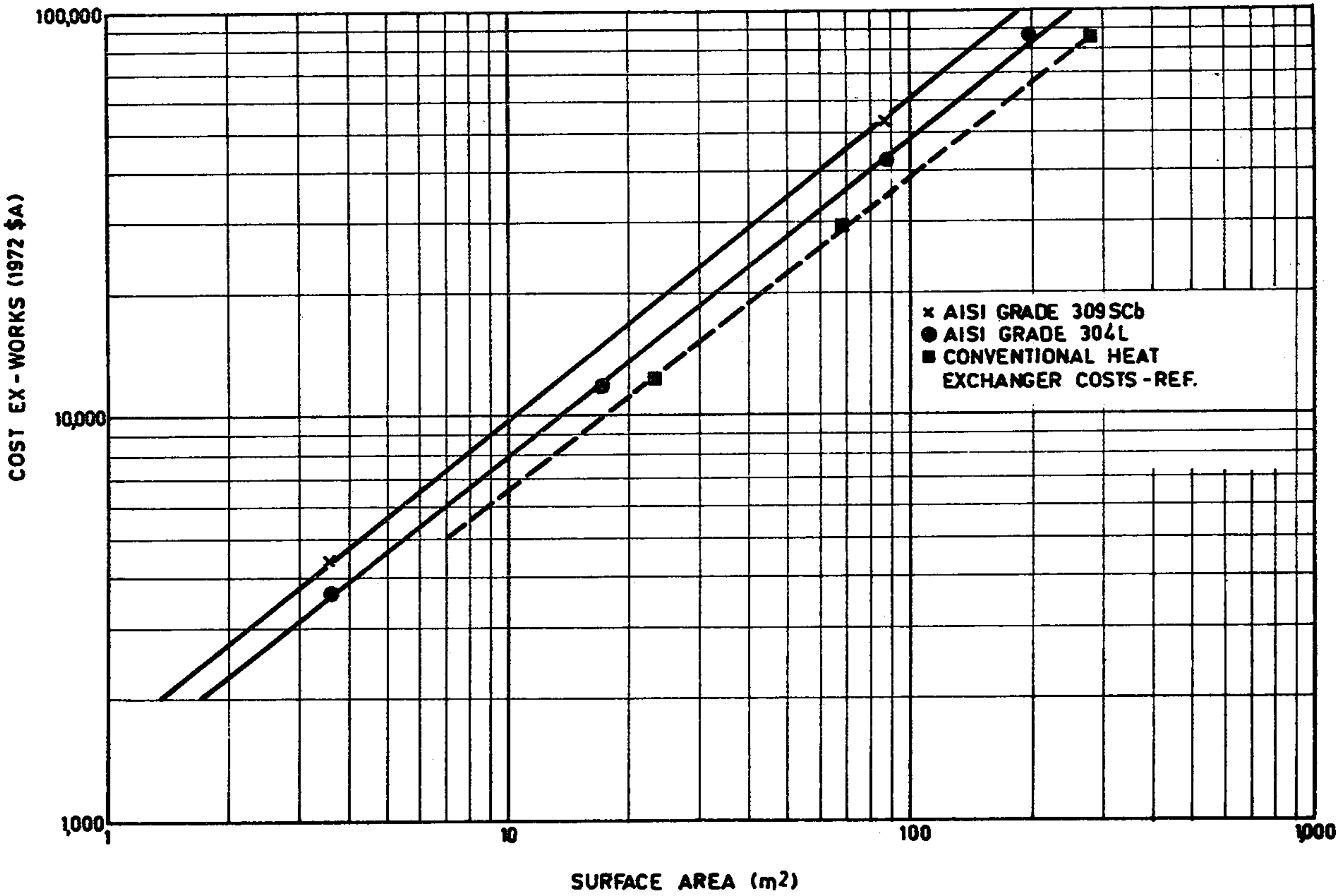


FIGURE 84. COST OF STAINLESS STEEL HEAT EXCHANGER FABRICATED TO RADIO-CHEMICAL STANDARDS

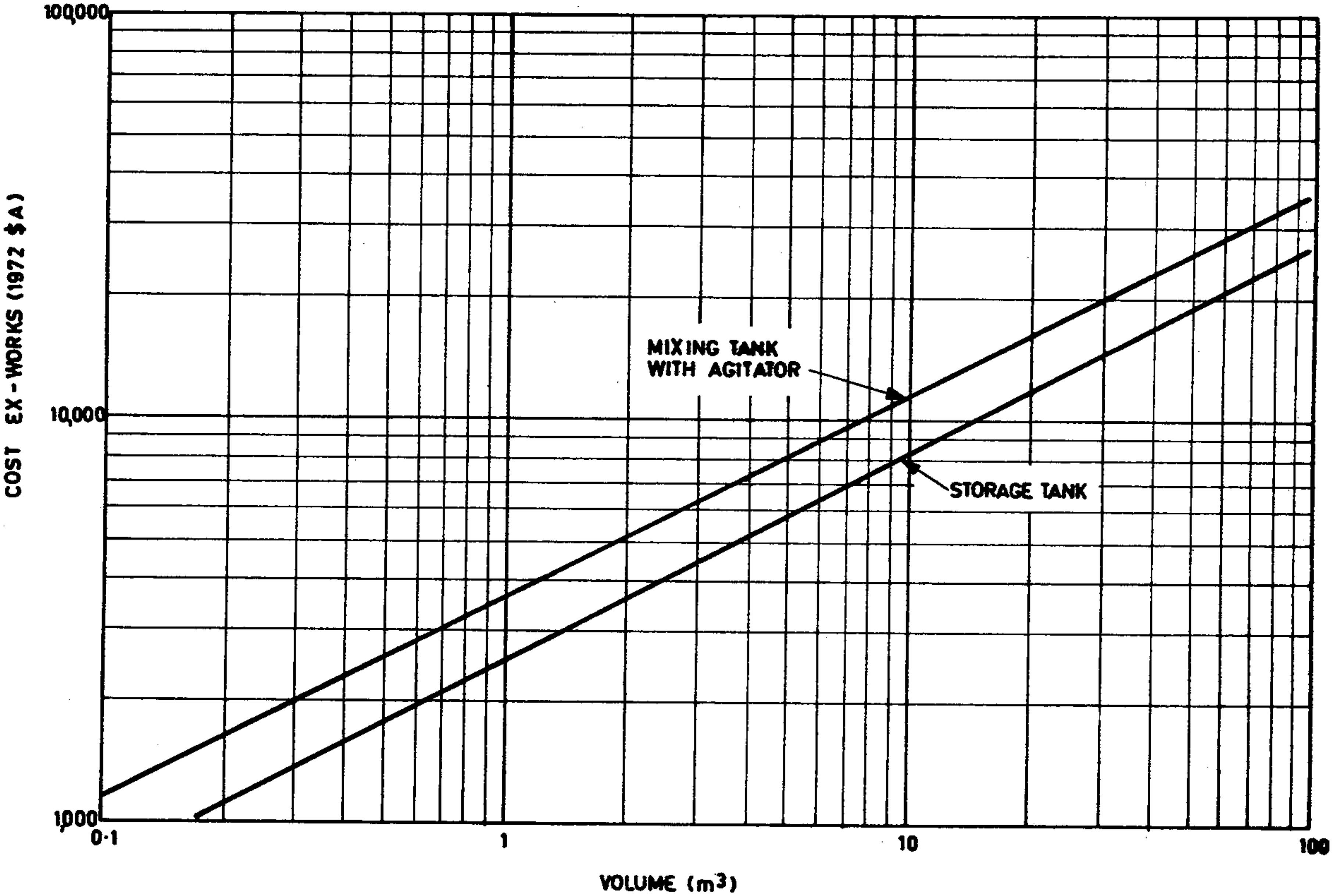


FIGURE 85. COST OF STORAGE AND MIXING TANKS FABRICATED FROM 304L STAINLESS STEEL PIPING

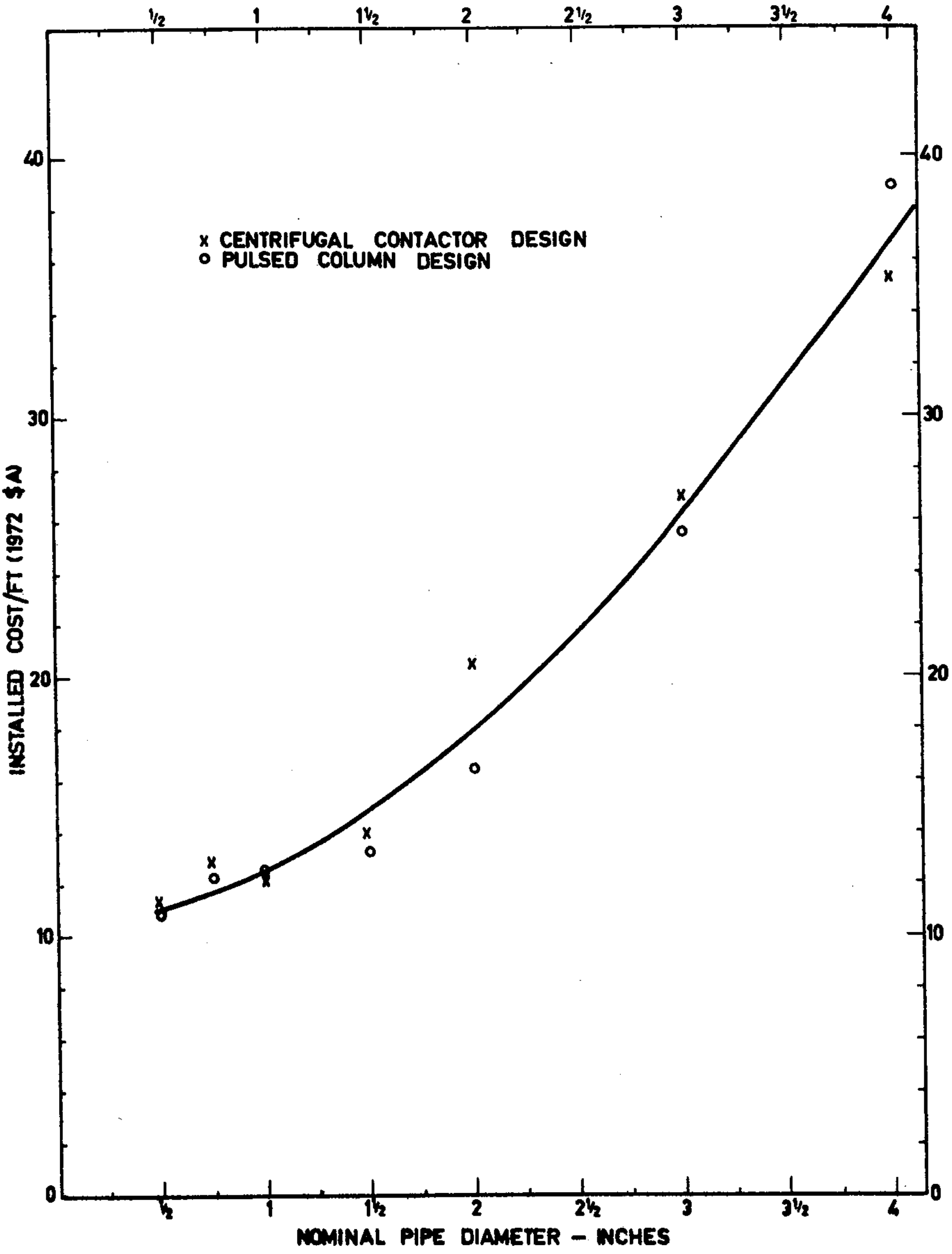


FIGURE 86. INSTALLED COST OF 304L STAINLESS STEEL PIPING

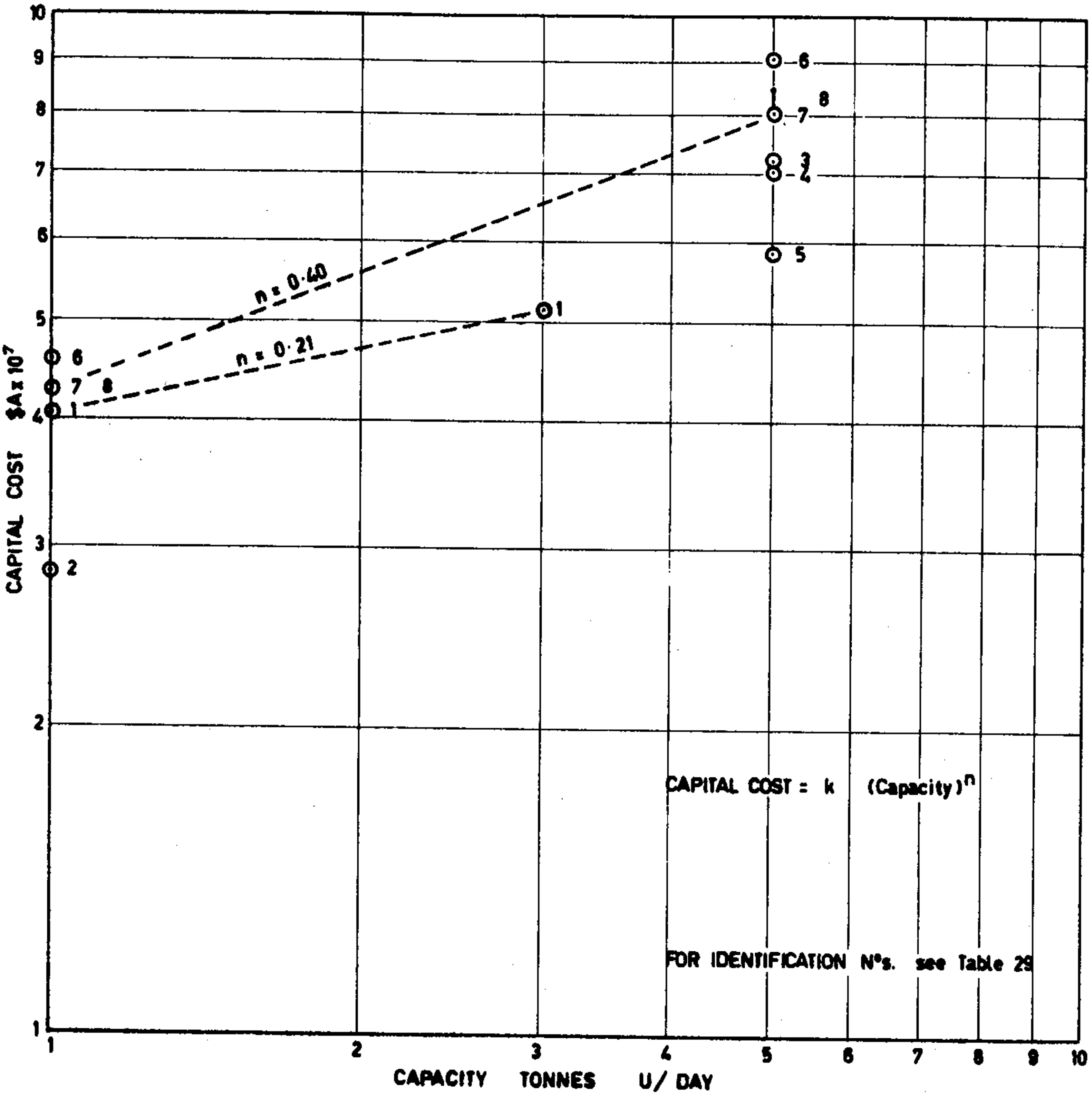


FIGURE 87. COMPARISON OF UPDATED CAPITAL COST ESTIMATES FOR REPROCESSING PLANTS

APPENDIX A

PLANT COSTING DATA

A1 CIVIL AND STRUCTURAL ESTIMATING RATES

PLANT AREAS

Item	Unit	Unit Rate ($A)
Rock Excavation		
• Bulk	m^3	8.00
• Small quantities	m^3	13.00
Concrete (placed)	m^3	42.00
Reinforcing (fixed)	tonne	320.00
Formwork		
• Walls	m^2	13.00
• Saffits	m^2	10.00
Structural Steelwork (fabricated, erected, painted)	tonne	500.00
Metal Purlins and Girts (erected)	m	3.60
Protected Metal Cladding (fixed)		
• Roof	m^2	8.10
• Walls	m^2	7.70
Paintwork (3 coats)	m^2	1.40
Acid Resisting Paint on Walls	m^2	3.80
Stainless Steel Linings	m^2	81.00

ANCILLARY AREAS

Item	Unit Rate ($A)
Rates per m^2 of floor area	
Cold Chemical Rooms	210
Control Rooms	320
Ventilation Rooms	190
Laboratories	530
Offices	240
Toilets	480
Aisles	210
Change Rooms	300
Canteen	280

A2 CHEMICAL PROCESS EQUIPMENT LIST

LOCATION LEGEND

Process Area	Symbol	DESIGN 1	2	3	4
Acid Recovery Cell	ARC	x	x	x	
Cold Chemical Room	CCR	x	x	x	x
Waste Evaporator Cell	WEC	x	x	x	
High Activity Liquor Cell	HALC	x	x		
Low Activity Liquor Cell	LALC	x	x		
Off-gas Cell	OGC		x		
Plutonium Purification and Evaporation Cell	PPEC	x	x	x	
Pulse Column Well	PCW	x		x	
Plutonium Packaging Area	PPA	x	x	x	x
Shear Leach Cell	SLC	x	x		
Shear Leach Pond	SLP			x	
Solvent and Off-gas Treatment Cell	SOGT	x			
Solvent Wash Cell	SWC		x		
Uranium Evaporation and Purification Cell	UEPC	x	x	x	

x indicates designs in which area is utilised

Unless indicated, all vessels are fabricated in AISI 304L stainless steel.

FUEL LEACHING

No.	Name	Location	Dimensions: Diameter	Overall Length	Cylinder Length	Capacity (m^3)	Ex-works Cost (1972 $A)	Design 1	2	3	4
VESSELS											
3C-1*	No. 1 Batch Dissolver	SLC	5'0"	11'9"	10'0"	5.7	23,000	x	x		x
3C-2*	No. 2 Batch Dissolver	SLC	5'0"	11'9"	10'0"	5.7	23,000	x	x		x
3C-3*	Continuous Dissolver**	SLP	1'0"	10'0"	10'0"	-	55,000			x	
3D-1	Accountability and Feed Adjust Tank	HALC	6'6"	10'6"	8'0"	7.3	18,000	x	x	x	x
3D-101	Continuous Dissolver Overflow Tank	SLP	6'6"	10'6"	8'0"	7.3	18,000			x	
HEAT EXCHANGERS											
3E-1*	No. 1 Batch Dissolver Condenser	SLC	1'6"	10'6"	8'0"	-	13,500	x	x		x
3E-2*	No. 2 Batch Dissolver Condenser	SLC	1'6"	10'6"	8'0"	-	13,500	x	x		x
3E-3*	Continuous Dissolver Condenser	SLP	1'0"	10'0"	8'0"	-	5,500			x	

* AISI 309SCb

** Non-cylindrical vessel. Dimensions quoted are thickness, width and height in that order

PARTITION CYCLE SOLVENT EXTRACTION

No.	Name	Location	Dimensions			Capacity (m^3)	Ex-works Costs (1972 $A)	Design			
			Diameter	Overall Length	Cylin. Length			1	2	3	4
COLUMNS											
4C-1	Partition cycle extraction column	PCW	13"	42'6"	34'0"	-	31,000	x		x	
4C-2	Partition cycle partition column	PCW	12"	41'10"	34'0"	-	19,000	x		x	
4C-3	Partition cycle Pu scrub column	PCW	6"	27'0"	20'0"	-	6,500	x		x	
4C-4	Partition cycle strip column	PCW	15"	42'2"	34'0"	-	24,500	x		x	
13C-1	No. 1 carbonate wash column	SOGT	1'0"	28'2"	27'6"	-	8,500	x		x	
13C-4	No. 1 acid wash column	SOGT	1'0"	15'0"	10'3"	-	6,000	x		x	
CENTRIFUGAL CONTACTORS *											
4CC-1E	Partition cycle extraction centrifuge	HALC	1'11"-2'6"	3'3"	2'8"	-	55,000		x		x
4CC-1S	Partition cycle scrub centrifuge	HALC	1'11"-2'6"	3'3"	2'8"	-	55,000		x		x
4CC-2	Partition cycle partition centrifuge	LALC	1'11"-2'6"	3'3"	2'8"	-	55,000		x		x
4CC-3	Partition cycle Pu scrub centrifuge	LALC	1'4"-2'0"	3'3"	2'8"	-	38,800		x		x
4CC-4	Partition cycle strip centrifuge	LALC	2'6"-3'3"	3'3"	2'8"	-	71,000		x		x
4CC-101	Solids removal centrifuge	HALC	8½"	1'9"	1'5"	-	7,000		x		x
MIXER-SETTLERS **											
13M-1	No. 1 carbonate washer	SWC	4'0"	2'0"	6'6"	1.1	8,500		x		x
13M-2	No. 1 acid washer	SWC	4'0"	2'0"	4'6"	0.8	5,500		x		x
MAJOR VESSELS											
4D-1	Partition cycle feed tank	HALC	6'6"	13'0"	10'6"	8.9	21,000	x	x	x	x
4D-2	Partition cycle waste catch tank	HALC	6'6"	13'0"	10'6"	8.9	21,000	x	x	x	x
4D-50	HA decanter catch pot	HALC	1'6"	3'0"	2'0"	0.1	1,400	x	x	x	x
4D-52	LA decanter catch pot	LALC	2'0"	2'10"	2'0"	0.2	1,900	x	x	x	x
4D-101	Uranium centrifuge dump tank	HALC	4'6"	6'9"	5'0"	2.2	9,500		x		x
4Y-1	Aqueous decanter for 4C-1 or 4CC-1E	PCW/HALC	1'6"	2'8"	2'0"	0.1	2,200	x	x	x	x
13D-1	No. 1 solvent washer	SOGT/SWC	4'0"	7'0"	5'6"	1.8	12,000	x	x	x	x
13D-4	No. 1 solvent storage tank	SOGT/SWC	5'0"	9'3"	7'6"	4.1	14,000	x	x	x	x
MINOR VESSELS											
4C-15*	Air reservoir for 4C-1	CCR	9"	2'0"	1'6"	-	200	x		x	
4C-16*	Air reservoir for 4C-2	CCR	9"	2'0"	1'6"	-	200	x		x	
4C-17*	Air reservoir for 4C-3	CCR	9"	2'0"	1'6"	-	200	x		x	
4C-18*	Air reservoir for 4C-4	CCR	9"	2'0"	1'6"	-	200	x		x	
4Y-2	Level sensing chambers for 4C-1	CCR	2"	23'0"	23'0"	-	800	x		x	
4Y-3	Aqueous decanter for 4C-3 or 4CC-3	PCW/LALC	1'0"	4'3"	3'9"	-	1,900	x	x	x	x
4Y-4	Aqueous decanter for 4C-4 or 4CC-4	PCW/LALC	1'6"	4'9"	4'0"	-	2,800	x	x	x	x
4Y-13	Phase separator for 4D-1 air lift	PCW/WEC	6"	1'6"	1'3"	-	250	x	x	x	x
4Y-14	Measuring head pot for 4C-1 or 4CC-1E	PCW/WEC	6"	3'6"	3'0"	-	450	x	x	x	x
4Y-15	Level control pot for 4C-1	PCW	6"	9"	6"	-	120	x		x	
4Y-17	Level control pot for 4C-2	PCW	6"	9"	6"	-	120	x		x	
4Y-19	Level control pot for 4C-3	PCW	6"	9"	6"	-	120	x		x	
4Y-20	Level control pot for 4C-4	PCW	6"	9"	6"	-	120	x		x	
4Y-42	Aqueous decanter for 4C-2 or 4CC-2	PCW/LALC	6"	1'6"	1'0"	-	300	x	x	x	x
4Y-51	Transfer head pot from 4D-50	SOGT	6"	3'6"	3'0"	-	500	x	x	x	x
4Y-101	Solvent surge pot from 4C-3 or 4CC-3	PCW/LALC	1'0"	3'9"	3'3"	-	1,100	x	x	x	x
4Y-102	Transfer head pot from 4Y-101	PCW/LALC	1'0"	3'9"	3'3"	-	1,100	x	x	x	x
13Y-1	Phase separator for 13C-1	SOGT	2"	8"	6"	-	50	x		x	
HEAT EXCHANGER											
4E-1	Organic feed preheater for 4C-4 or 4CC-4	PCW	2"	13'0"	13'0"	-	950	x	x	x	x

* Dimensions of centrifugal contactors quoted as follows:
- Diameter - both bowls
- Cylindrical length - inner bowl
- Overall length - excludes thickness of shielding

** Non-cylindrical vessel. Dimensions quoted are width, depth and length in that order.

1ST URANIUM CYCLE SOLVENT EXTRACTION

No.	Name	Location	Dimensions			Capacity (m^3)	Ex-works Costs (1972 $A)	Design			
			Diameter	Overall Length	Cylin. Length			1	2	3	4
COLUMNS											
4C-9	1st U cycle extraction column	PCW	18"	42'6"	34'0"		26,000	x		x	
4C-10	1st U cycle strip column	PCW	15"	42'2"	34'0"		24,500	x		x	
13C-2	No. 2 carbonate wash column	SOGT	1'0"	28'2"	27'6"		8,500	x		x	
13C-5	No. 2 acid wash column	SOGT	1'0"	15'0"	10'3"		6,000	x		x	
CENTRIFUGAL CONTACTORS *											
4CC-9E	1st U cycle extraction centrifuge	LALC	2'6"-3'3"	3'3"	2'8"		71,000		x		x
4CC-9S	1st U cycle scrub centrifuge	LALC	1'11"2'6"	3'3"	2'8"		55,000		x		x
4CC-10	1st U cycle strip centrifuge	LALC	2'6"-3'3"	3'3"	2'8"		71,000		x		x
MIXER SETTLERS **											
13M-3	No. 2 Carbonate washer	SWC	4'0"	2'0"	6'6"	1.1	8,500		x		x
13M-4	No. 2 Acid washer	SWC	4'0"	2'0"	4'6"	0.8	5,500		x		x
MAJOR VESSELS											
4D-9	1st U cycle feed tank conditioner	LALC	3'6"	10'0"	6'0"	1.6	7,000	x	x	x	x
4D-10	1st U cycle waste catch tank	LALC	9'6"	16'6"	13'0"	25.0	23,500	x	x	x	x
13D-2	No. 2 solvent washer	SOGT/SWC	4'0"	7'0"	5'6"	1.8	12,000	x	x	x	x
13D-5	No. 2 solvent storage tank	SOGT/SWC	5'0"	9'3"	7'6"	4.0	14,000	x	x	x	x
MINOR VESSELS											
4C-21 ***	Air reservoir for 4C-9	CCR	9"	2'0"	1'6"		200	x		x	
4C-22 ***	Air reservoir for 4C-10	CCR	9"	2'0"	1'6"		200	x		x	
4Y-9	Aqueous decanter for 4C-9 or 4CC-93	PCW/LALC	2'0"	4'6"	3'9"		2,400	x	x	x	x
4Y-10	Aqueous decanter for 4C-10or 4CC-10	PCW/LALC	1'6"	4'3"	3'9"		1,800	x	x	x	x
4Y-24	Measuring head pot for 4C-9 or 4CC-9	PCW/PPEC	6"	3'6"	3'0"		450	x	x	x	x
4Y-25	Level control pot for 4C-9	PCW	6"	1'0"	6"		120	x		x	
4Y-26	Level sensing chambers for 4C-9	CCR	2"	24'0"	24'0"		800	x		x	
4Y-27	Level control pot for 4C-10	PCW	6"	1'0"	6"		120	x		x	
13Y-3	Phase separator for 13C-2	SOGT	2"	8"	6"		50	x		x	
HEAT EXCHANGER											
4E-2	1st U cycle feed heater	PCW/LALC	2"	13'6"	13'0"		950	x	x	x	x

* Dimensions of centrifugal contactors quoted as follows:

- Diameter - both bowls
- Cylindrical length - inner bowl
- Overall length - excludes thickness of shielding

** Non-cylindrical vessel. Dimensions quoted are width, depth and length in that order.

*** Carbon steel to AS.B250

2ND URANIUM CYCLE SOLVENT EXTRACTION

No.	Name	Location	Dimensions			Capacity (m^3)	Ex-works Costs (1972 $A)	Design			
			Diameter	Overall Length	Cylin. Length			1	2	3	4
COLUMNS											
4C-11	2nd U cycle extraction column	PCW	18"	42'6"	34'0"		26,000	x		x	
4C-12	2nd U cycle strip column	PCW	15"	42'2"	34'0"		24,500	x		x	
13C-3	No. 3 carbonate wash column	SOGT	1'0"	28'2"	27'6"		8,500	x		x	
13C-6	No. 3 acid wash column	SOGT	1'0"	15'0"	10'3"		6,000	x		x	
CENTRIFUGAL CONTACTORS *											
4CC-11E	2nd U cycle extraction centrifuge	LALC	2'6"-3'3"	3'3"	2'8"		71,000		x		x
4CC-11S	2nd U cycle scrub centrifuge	LALC	1'11"2'6"	3'3"	2'8"		55,000		x		x
4CC-12	2nd U cycle strip centrifuge	LALC	2'6"-3'3"	3'3"	2'8"		71,000		x		x
MIXER SETTLERS **											
13M-5	No. 3 carbonate washer	SWC	4'0"	2'0"	6'6"	1.1	8,500		x		x
13M-6	No. 3 acid washer	SWC	4'0"	2'0"	4'6"	0.8	5,500		x		x
MAJOR VESSELS											
4D-12	2nd U cycle feed tank conditioner	LALC	3'6"	10'0"	6'0"	1.6	7,000	x	x	x	x
4D-13	2nd U cycle waste catch tank	LALC	9'6"	16'6"	13'0"	25.0	23,500	x	x	x	x
13D-3	No. 3 solvent washer	SOGT/SWC	4'0"	7'0"	5'6"	1.8	12,000	x	x	x	x
13D-6	No. 3 solvent storage tank	SOGT/SWC	5'0"	9'3"	7'6"	4.1	14,000	x	x	x	x
MINOR VESSELS											
4C-23 ***	Air reservoir for 4C-11	CCR	9"	2'0"	1'6"		200	x		x	
4C-24 ***	Air reservoir for 4C-12	CCR	9"	2'0"	1'6"		200	x		x	
4Y-11	Aqueous decanter for 4C-11 or 4CC-11E	PCW/LALC	2'0"	4'6"	3'9"		2,400	x	x	x	x
4Y-12	Aqueous decanter for 4C-12 or 4CC-12	PCW/LALC	1'6"	4'3"	3'9"		1,800	x	x	x	x
4Y-28	Measuring head pot for 4C-11 or 4CC-11	PCW/PPEC	6"	3'6"	3'0"		450	x	x	x	x
4Y-29	Level control pot for 4C-11	PCW	6"	1'0"	6"		120	x		x	
4Y-30	Level sensing chambers for 4C-11	CCR	2"	24'0"	24'0"		800	x		x	
4Y-31	Level control pot for 4C-12	PCW	6"	1'0"	6"		120	x		x	
13Y-5	Phase separator for 13C-3	SOGT	2"	8"	6"		50	x		x	
HEAT EXCHANGER											
4E-3	2nd U cycle feed heater	PCW/LALC	2"	13'6"	13'0"		950	x	x	x	x

* Dimensions of centrifugal contactors quoted as follows:

- Diameter - both bowls
- Cylindrical length - inner bowl
- Overall length - excludes thickness of shielding

** Non-cylindrical vessel. Dimensions quoted are width, depth and length in that order.

*** Carbon steel to AS.B250

URANIUM EVAPORATION AND PURIFICATION

No.	Name	Location	Dimensions			Capacity (m³)	Ex-works Costs (1972 $A)	Design			
			Diameter	Overall Length	Cylin. Length			1	2	3	4
COLUMNS											
5C-6A	No. 1 U silica gel bed	UEPC	1'9"	11'6"	10'6"	0.7	5,000	x	x	x	x
5C-6B	No. 2 U silica gel bed	UEPC	1'9"	11'6"	10'6"	0.7	5,000	x	x	x	x
MAJOR VESSELS											
5C-4	U product evaporator	UEPC	4'0"	22'6"	8'6"	-	10,200	x	x	x	x
5D-7	U evaporator feed tank	UEPC	3'0"	7'0"	4'6"	0.9	4,500	x	x	x	x
5D-8	U evaporator condensate tank	UEPC	3'6"	8'4"	6'0"	1.6	8,000	x	x	x	x
5D-9	U product surge tank	UEPC	3'6"	8'0"	5'6"	1.4	5,200	x	x	x	x
5D-10	Rinse acid hold tank	UEPC	2'3"	3'3"	2'6"	0.2	2,100	x	x	x	x
5D-11	Regeneration waste tank	UEPC	2'6"	7'6"	5'6"	0.85	4,200	x	x	x	x
5D-12A	No. 1 U product sample tank	UEPC	4'6"	7'4"	6'0"	2.6	10,500	x	x	x	x
5D-12B	No. 2 U product sample tank	UEPC	4'6"	7'4"	6'0"	2.6	10,500	x	x	x	x
5D-15A	No. 1 uranium storage tank	YARD	9'6"	30'2"	27'6"	56	38,000	x	x	x	x
5D-15B	No. 2 uranium storage tank	YARD	9'6"	30'2"	27'6"	56	38,000	x	x	x	x
MINOR VESSELS											
5Y-3	Flow control pot to 5C-4	UEPC	6"	2'0"	1'6"		250	x	x	x	x
5Y-4	Level control pot for 5C-4	UEPC	6"	1'0"	6"		250	x	x	x	x
5Y-29	Head pot for 5C-6A feed	UEPC	1½"	8'4"	8'0"		150	x	x	x	x
5Y-30	Head pot for 5C-6B feed	UEPC	1½"	8'4"	8'0"		150	x	x	x	x
5Y-31	Syphon breaker 5C-6A discharge	UEPC	6"	2'0"	1'6"		350	x	x	x	x
5Y-32	Syphon breaker 5C-6B discharge	UEPC	6"	2'0"	1'6"		350	x	x	x	x
HEAT TRANSFER EQUIPMENT											
5E-3	Uranium product reboiler	UEPC	1'4"	-	12'0"		10,200	x	x	x	x
5E-4	Uranium product condenser	UEPC	1'2"	12'9"	12'0"		8,600	x	x	x	x
5E 10	Uranium evaporator product cooler	UEPC	2"	70'0"	14'0" (5 passes)		1,800	x	x	x	x

PLUTONIUM CYCLE SOLVENT EXTRACTION

No.	Name	Location	Dimensions			Capacity (m^3)	Ex-works Costs (1972 $A)	Design			
			Diameter	Overall Length	Cylin. Length			1	2	3	4
COLUMNS											
4C-7	Plutonium cycle extraction column	PCW	8"	48'0"	39'0"		13,500	x	x	x	
4C-8	Plutonium cycle strip column	PCW	4"	35'0"	31'0"		5,000	x		x	
CENTRIFUGAL CONTACTORS *											
4CC-7E	Plutonium cycle extraction centrifuge	LALC	1'4"-2'0"	3'3"	2'6"		38,800		x		x
4CC-7S	Plutonium cycle scrub centrifuge	LALC	8½"	1'9"	1'5"		21,300		x		x
4CC-8	Plutonium cycle strip centrifuge	LALC	8½"	1'9"	1'5"		21,300		x		x
MAJOR VESSELS											
4D-6	Plutonium cycle feed conditioner	LALC	2'6"	6'6"	4'0"	0.6	3,500	x	x	x	x
4D-8	Plutonium cycle waste catch tank	LALC	4'0"	11'6"	9'6"	3.2	8,000	x	x	x	x
4D-103	Plutonium centrifuge dump tank	LALC	2'0"	2'9"	2'0"		1,300		x		x
MINOR VESSELS											
4C-19 **	Air reservoir for 4C-7	CCR	9"	2'0"	1'6"		200	x		x	
4C-20 **	Air reservoir for 4C-8	CCR	9"	2'0"	1'6"		200	x		x	
4Y-7	Aqueous decanter for 4C-7 or 4CC-7E	PCW/LALC	9"	5'0"	4'6"		1,400	x	x	x	x
4Y-8	Aqueous decanter for 4C-8 or 4CC-7S	PCW/LALC	5"	4'0"	3'7"		900	x	x	x	x
4Y-32	Measuring head pot for 4C-7 or 4CC-7E	PCW/PPEC	6"	3'6"	3'0"		500	x	x	x	x
4Y-33	Level control pot for 4C-7	PCW	6"	1'0"	6"		120	x		x	
4Y-34	Level sensing chambers for 4C-7	CCR	2"	24'0"	24'0"		800	x		x	
4Y-35	Level control pot for 4C-8	PCW	6"	1'0"	6"		120	x		x	
4Y-103	Plutonium cycle solvent surge pot	PCW/LALC	6"	2'6"	2'6"		400	x	x	x	x

* Dimensions of centrifugal contactors quoted as follows:

- Diameter - both bowls
- Cylindrical length - inner bowl
- Overall length - excludes thickness of shielding

** Carbon steel to AS.B250

PLUTONIUM PURIFICATION AND EVAPORATION

No.	Name	Location	Dimensions			Capacity (m^3)	Ex-Works Costs (1972 $A)	Design			
			Diameter	Overall Length	Cylin. Length			1	2	3	4
COLUMNS											
5C-1A	No. 1 Pu anion exchanger	PPEC	7"	9'0"	9'0"	0.05	3,400	x	x	x	x
5C-1B	No. 2 Pu anion exchanger	PPEC	7"	9'0"	9'0"	0.05	3,400	x	x	x	x
5C-1C	No. 3 Pu anion exchanger	PPEC	7"	9'0"	9'0"	0.05	3,400	x	x	x	x
5C-1D	No. 4 Pu anion exchanger	PPEC	7"	9'0"	9'0"	0.05	3,400	x	x	x	x
5C-1E	No. 5 Pu anion exchanger	PPEC	7"	9'0"	9'0"	0.05	3,400	x	x	x	x
5C-1F	No. 6 Pu anion exchanger	PPEC	7"	9'0"	9'0"	0.05	3,400	x	x	x	x
MAJOR VESSELS											
5C-2A *	No. 1 Pu product evaporator	PPEC	7"	13'6"	5'0" (reboiler) 7'4" (disengagement)	-	10,000	x	x	x	x
5C-2B *	No. 2 Pu product evaporator	PPEC	7"	13'6"	"	-	10,000	x	x	x	x
5D-1 **	Pu ionex feed conditioner	PPEC	2½"	6'0"	5'0"	0.18	6,900	x	x	x	x
5D-2	Pu ionex waste catch tank	PPEC	3'0"	7'0"	4'0"	0.82	3,800	x	x	x	x
5D-4 **	Feed tank for Pu evaporators	PPEC	2½"	5'0"	4'3"	0.12	5,000	x	x	x	x
5D-5A**	No. 1 Pu storage tank	PPEC	2½"	5'0"	5'0"	0.14	5,500	x	x	x	x
5D-5B**	No. 2 Pu storage tank	PPEC	2½"	5'0"	5'0"	0.14	5,500	x	x	x	x
5D-6A	No. 1 Pu evaporator condensate tanks	PPEC	1'4"	6'0"	4'0"	0.16	2,500	x	x	x	x
5D-6B	No. 2 Pu evaporator condensate tank	PPEC	1'4"	6'0"	4'0"	0.16	2,500	x	x	x	x
5D-20	Ionex hot water tank	PPA	2'3"	2'6"	2'6"	0.23	1,200	x	x	x	x
MINOR VESSELS											
5D-17	Pu product measuring tank	PPA	10"	1'3"	10"	-	500	x	x	x	x
5Y-7	Transfer pot for waste transfer to 4D-6	SWC	6"	1'6"	1'3"	-	250	x	x	x	x
5Y-10	Pu product transfer pot	PPEC	4"	1'8"	1'4"	-	200	x	x	x	x
5Y-11	Pu product transfer pot	PPEC	4"	1'8"	1'4"	-	200	x	x	x	x
5Y-22A	Flow control pot, 5C-2A to 5D-5A	PPEC	6"	1'0"	6"	-	120	x	x	x	x
5Y-22B	Flow control pot, 5C-2A to 5D-5B	PPEC	6"	1'0"	6"	-	120	x	x	x	x
5Y-23A	Flow control pot, 5C-2B to 5D-5B	PPEC	6"	1'0"	6"	-	120	x	x	x	x
5Y-51	Head pot for 5C-2A feed	PPEC	6"	3'6"	3'0"	-	450	x	x	x	x
5Y-52	Head pot for 5C-2B feed	PPEC	6"	3'6"	3'0"	-	450	x	x	x	x
HEAT EXCHANGER											
5E-7	Pu product evaporator condenser	PPEC	2"	10'0"	10'0"	-	800	x	x	x	x

* Titanium

** Non-cylindrical vessel. Dimensions quoted are thickness, length and height in that order.

OFF-GAS TREATMENT

No.	Name	Location	Dimensions			Capacity (m^3)	Ex-Works Costs (1972 \$A)	Design			
			Diameter	Overall Length	Cylin. Length			1	2	3	4
COLUMNS											
32001	Dissolver off-gas (DOG) absorber	SOGT/OGC	2'6"	24'0"	23'0"	3.2	16,500	x	x	x	x
32002	Iodine absorber	SOGT/OGC	2'6"	24'0"	23'0"	3.2	16,500	x	x	x	x
VESSELS											
35001	DOG absorber surge tank	SOGT/OGC	3'0"	10'3"	9'0"	1.7	8,600	x	x	x	x
35003	Iodine stripper	SOGT/OGC	3'6"	8'4"	7'0"	1.7	8,600	x	x	x	x
35004	Iodine absorber surge tank	SOGT/OGC	3'0"	10'3"	9'0"	1.7	8,600	x	x	x	x
35006	Iodine absorber separator	SOGT/OGC	2'0"	6'0"	5'0"	0.4	3,600	x	x	x	x
HEAT EXCHANGERS											
31002	Off gas heater	SOGT/OGC	1'6"	3'0"	1'6"	-	2,300	x	x	x	x
31013	Iodine stripper condenser	SOGT/OGC	9"	10'6"	10'0"	-	2,800	x	x	x	x

WASTE EVAPORATION

No.	Name	Location	Dimensions			Capacity (m^3)	Ex-Works Costs (1972 \$A)	Design			
			Diameter	Overall Length	Cylin. Length			1	2	3	4
HIGH LEVEL											
7D-1	High level waste evaporator feed tank	HALC	6'6"	10'3"	7'6"	6.8	13,000	x	x	x	x
7C-1 *	High level waste evaporator	WEC	5'0"	18'0"	8'0"	-	16,500	x	x	x	x
7D-4	High level waste accountability and neutraliser tank	HALC	5'6"	10'0"	7'9"	5.2	15,500	x	x	x	x
7E-5 *	High level waste evaporator condenser	WEC	1'4"	9'6"	8'0"	-	9,500	x	x	x	x
7E-11*	High level waste evaporator reboiler	WEC	2'6"	9'0"	7'0"	-	34,500	x	x	x	x
LOW LEVEL											
7D-2	Low level waste evaporator feed tank	LALC	9'6"	20'3"	16'6"	32	27,000	x	x	x	x
7D-14	Hot analytical cell drain catch tank	LALC	3'0"	9'0"	7'6"	1.5	5,200	x	x	x	x
7C-2	Low level waste evaporator	WEC	8'6"	21'6"	9'6"	-	31,200	x	x	x	x
7D-10	Low level waste accountability and neutraliser tank	LALC	7'0"	11'6"	8'6"	9.2	21,500	x	x	x	x
7E-7 *	Low level waste evaporator condenser	WEC	3'0"	10'0"	7'6"	-	53,000	x	x	x	x
7E-21*	Low level waste evaporator 1st reboiler	WEC	4'0"	10'0"	7'6"	-	87,000	x	x	x	x
7E-22*	Low level waste evaporator 2nd reboiler	WEC	4'0"	10'0"	7'6"	-	87,000	x	x	x	x
13D-7	Solvent waste catch tank	SOGT/SWC	5'0"	11'9"	10'0"	5.5	11,000	x	x	x	x
REWORK											
7D-8	Rework evaporator feed tank	HALC	8'0"	11'0"	7'6"	10.5	14,500	x	x	x	x
7C-4 *	Rework evaporator	WEC	6'0"	20'6"	10'0"	-	20,500	x	x	x	x
7E-41*	Rework evaporator reboiler	WEC	3'6"	10'0"	7'6"	-	67,000	x	x	x	x
7E-8 *	Rework evaporator condenser	WEC	2'6"	10'0"	7'6"	-	36,500	x	x	x	x

* AISI 309SCb

NITRIC ACID RECOVERY

No.	Name	Location	Dimensions			Capacity (m^3)	Ex-Works Costs (1972 $A)	Design			
			Diameter	Overall Length	Cylin. Length			1	2	3	4
32003	Acid fractionator	ARC	7'9"	30'0"	13'0"		41,500	x	x	x	x
MAJOR VESSELS											
7D-3	Acid fractionator feed tank	ARC	5'0"	13'9"	12'0"	6.7	18,000	x	x	x	x
7D-5A	No. 1 recovered acid storage tank	ARC	9'0"	14'6"	11'0"	19.0	29,400	x	x	x	x
7D-5B	No. 2 recovered acid storage tank	ARC	9'0"	14'6"	11'0"	19.0	29,500	x	x	x	x
7D-6	Weak acid catch tank	ARC	5'6"	14'0"	12'0"	8.2	19,500	x	x	x	x
7D-11	Hot acid storage tank	ARC	5'6"	14'0"	12'0"	8.2	19,500	x	x	x	x
7D-12	Hot acid metering tank	ARC	5'6"	14'0"	12'0"	8.2	13,500	x	x	x	x
MINOR VESSELS											
7D-51	Fractionator overheads barometric seal pot	ARC	1'0"	1'6"	1'0"		700	x	x	x	x
7D-101	Fractionator reflux divider	ARC	1'6"	3'0"	2'0"		1,200	x	x	x	x
35009	Acid fractionator feed separator	ARC	1'0"	2'9"	2'0"		900	x	x	x	x
HEAT EXCHANGERS											
7E-3	Fractionator condenser	ARC	2'6"	19'0"	17'0"		52,000	x	x	x	x
7E-9	Fractionator vacuum intermediate condenser	ARC	6"	3'0"	2'6"		500	x	x	x	x
7E-10	Fractionator after condenser	ARC	6"	3'6"	2'6"		500	x	x	x	x
31003	Acid fractionator feed vaporiser	ARC	4'6"	10'6"	5'0"		83,000	x	x	x	x
31004	Vaporiser bottoms cooler	ARC	1'0"	7'0"	5'9"		3,600	x	x	x	x
31005	Fractionator reboiler	ARC	1'9"	8'6"	5'9"		12,000	x	x	x	x
31006	Acid fractionator bottoms cooler	ARC	1'6"	8'6"	7'0"		9,600	x	x	x	x

EFFLUENT TREATMENT

No.	Name	Location	Dimensions			Capacity (m^3)	Ex-Works Costs (1972 $A)	Design			
			Diameter	Overall Length	Cylin. Length			1	2	3	4
MAJOR VESSELS											
7C-5	General purpose evaporator	ARC	8'0"	20'0"	8'0"	-	21,500	x	x	x	x
7D-7	General purpose evaporator feed tank	ARC	12'0"	14'6"	11'0"	34	27,000	x	x	x	x
7D-13	Laundry analytical drain catch tank	YARD	5'0"	13'9"	12'0"	6.8	9,500	x	x	x	x
7D-9	Condensate ion exchanger feed tank	ARC	6'6"	8'3"	10'6"	9.9	22,500	x	x	x	x
7D-51	No. 1 ion exchange column	YARD	4'0"	8'6"	8'6"	3.0	6,500	x	x	x	x
7D-52	No. 2 ion exchange column	YARD	4'0"	8'6"	8'6"	3.0	6,500	x	x	x	x
7E-13	General purpose evaporator condenser	ARC	3'0"	10'6"	8'0"	-	43,500	x	x	x	x
7E-51	General purpose evaporator reboiler	ARC	3'0"	10'6"	8'0"	-	43,500	x	x	x	x

COLD CHEMICALS

No.	Name	Location	Dimensions			Capacity (m^3)	Ex-Works Costs (1972 $A)	Design			
			Diameter	Overall Length	Cylin. Length			1	2	3	4
VESSELS											
14D-1	Conc. nitric acid storage tank	YARD	12'0"	19'6"	15'0"	49	18,500	x	x	x	x
14D-2	Conc. caustic storage tank	YARD	5'0"	8'6"	6'6"	3.3	4,600	x	x	x	x
14D-3	Conc. nitric acid day tank	CCR	6'0"	7'6"	7'6"	5.9	6,400	x	x	x	x
14D-4	Strip acid day tank	CCR	12'0"	19'6"	15'0"	49	18,500	x	x	x	x
14D-5	Partition solution mix tank	CCR	4'3"	6'0"	5'6"	2.1	5,300	x	x	x	x
14D-6	Partition solution run tank	CCR	4'3"	6'0"	5'6"	2.1	3,700	x	x	x	x
14D-7	Plutonium strip mix tank	CCR	2'3"	2'9"	2'9"	0.3	2,000	x	x	x	x
14D-8	Plutonium strip run tank	CCR	2'3"	2'9"	2'9"	0.3	1,400	x	x	x	x
14D-9	Strip acid mix tank	CCR	12'0"	19'6"	15'0"	49	26,500	x	x	x	x
14D-10	1st U cycle scrub mix tank	CCR	3'9"	4'9"	4'9"	1.5	4,500	x	x	x	x
14D-11	1st U cycle scrub run tank	CCR	3'9"	4'9"	4'9"	1.5	3,100	x	x	x	x
14D-12	Ferrous sulphamate mix tank	CCR	2'9"	3'6"	3'6"	0.6	2,800	x	x	x	x
14D-13	Uranium cycles feed add tank	CCR	6'0"	7'6"	7'6"	5.9	6,400	x	x	x	x
14D-14	Sodium nitrite mix tank	CCR	2'0"	2'6"	2'6"	0.2	1,600	x	x	x	x
14D-15	Sodium nitrite run tank	CCR	2'0"	2'6"	2'6"	0.2	1,000	x	x	x	x
14D-16	Partition cycle mix tank	CCR	4'6"	6'0"	6'0"	2.6	5,900	x	x	x	x
14D-17	Partition cycle run tank	CCR	4'6"	6'0"	6'0"	2.6	4,200	x	x	x	x
14D-18	Plutonium cycle scrub mix tank	CCR	2'9"	3'3"	3'3"	0.5	2,700	x	x	x	x
14D-19	Plutonium cycle scrub run tank	CCR	2'9"	3'3"	3'3"	0.5	1,800	x	x	x	x
14D-20	Plutonium cycle scrub add tank	CCR	4'0"	5'0"	5'0"	1.7	4,800	x	x	x	x
14D-23	Plutonium ionex waste mix and run tank	CCR	4'0"	5'0"	5'0"	1.7	4,800	x	x	x	x
14D-25A	No. 1 caustic dilution tank	CCR	4'6"	6'0"	6'0"	2.6	5,700	x	x	x	x
14D-25B	No. 2 caustic dilution tank	CCR	4'6"	6'0"	6'0"	2.6	5,700	x	x	x	x
14D-26A	No. 1 carbonate add tank	CCR	5'0"	6'6"	6'6"	3.3	6,500	x	x	x	x
14D-26B	No. 2 carbonate add tank	CCR	5'0"	6'6"	6'6"	3.3	6,500	x	x	x	x
14D-27	Silica gel regeneration wash tank	CCR	2'0"	2'6"	2'6"	0.2	1,100	x	x	x	x
14D-28	TBP solvent mix and run tank	CCR	3'0"	4'0"	4'0"	0.8	3,200	x	x	x	x
14D-32	Poison mix tank	CCR	3'6"	4'6"	4'6"	1.2	4,100	x	x	x	x
14D-33	2nd U cycle scrub mix tank	CCR	3'9"	4'9"	4'9"	1.5	4,500	x	x	x	x
14D-34	2nd U cycle scrub run tank	CCR	3'9"	4'9"	4'9"	1.5	3,100	x	x	x	x
14D-40	Uranium cycles intermediate scrub mix tank	CCR	4'9"	6'0"	6'0"	3.1	6,400	x	x	x	x
14D-41	Uranium cycles intermediate scrub run tank	CCR	4'9"	6'0"	6'0"	3.1	4,600	x	x	x	x
14D-43	Plutonium ionex sodium nitrate run tank	CCR	1'6"	2'0"	2'0"	0.1	800	x	x	x	x
14D-46	Plutonium ionex elutriant mix tank	CCR	2'0"	2'6"	2'6"	0.2	1,500	x	x	x	x
14D-47	Hutonium ionex elutriant run tank	CCR	2'0"	2'6"	2'6"	0.2	1,000	x	x	x	x
14D-100	Leacher feed acid mix tank	CCR	6'0"	7'6"	7'6"	6.6	9,500	x	x	x	x
14D-102	Silica gel wash tank	CCR	2'9"	3'3"	3'3"	0.	1,800	x	x	x	x
15D-1	Decontamination solution mix tank	CCR	6'0"	7'6"	7'6"	5.9	8,800	x	x	x	x

A3 SPECIFICATION FOR CHROMIUM-NICKEL STAINLESS STEEL EQUIPMENT (ND-802)

A3.1 Scope

This specification covers requirements for the materials, fabrication, inspection and testing of chromium-nickel stainless steel equipment for radiochemical service.

A3.2 Codes and Standards

Except where specified otherwise, the code governing the fabrication of equipment shall be the ASME Unfired Pressure Vessel Code.

A3.3 Materials

- Material Specifications

Plate, pipe, fittings and forgings shall be furnished in accordance with the following ASTM specifications:

Plate : ASTM A240
Pipe : ASTM A312
Fittings : ASTM A403
Forgings : ASTM A182.

Grade shall be that specified on the drawings. No substitutions will be allowed unless approved by purchaser.

- Additional Material Testing

In addition to the testing requirements of the material specifications, two samples of each heat shall be subject to a boiling nitric acid test per ASTM A262. All samples shall be sensitized at 1250°F for 1 hour prior to testing. No portion of a heat of material shall be used if intergranular attack (corrosion rate exceeding 0.02 inches per year) occurs on the samples subjected to the boiling nitric acid test.

A3.4 Welding

- Preparation for Welding

Surfaces to be welded shall be bevelled by machining or grinding. Solvents containing chlorine shall not be used for cleaning chromium-nickel stainless steel.

- Weld Metal Composition

The deposited weld metal shall have a composition within the limits specified for the base metal.

- Appearance of Weld

Welds shall be completely fused, smooth, uniform and free of cracks, pinholes, undercut, excess high crown, ridges or deep grooves between bends. Butt welds shall have full penetration.

- Weld Defects

Any defects that appear on the surface of any deposited bead shall be removed by grinding before depositing the next bead. Repair of welds shall not groove or notch the base metal or reduce the wall thickness.

- Surface Finish and Cleanliness

There are no requirements for surface finish or cleanliness, e.g. RMS finish, wipe test, etc., other than those for vessels built to the ASME Code for commercial use.

A3.5 Inspection

- Radiographic and Dye-penetrant Examination

All welds shall be 100% radiographed. Exceptions to this rule will only be made with the permission of the purchaser in cases where the welds cannot be designed to allow radiography. In such cases weld quality shall be examined by a liquid penetrant as per ASTM E165.

- Pressure Testing

Vessels shall be hydrostatically tested to 1½ times the design pressure unless otherwise specified. Coils shall be air tested with 100 psig and submerged in water to determine the location of leaks.

- Helium Leak Test

A helium leak test is not required unless specified on the drawings. When specified an internal pressure of 15 psig shall be used in conjunction with a mass spectrometer which is calibrated for full-scale deflection of a leak rate of 1×10^{-7} atm cm^3 s^{-1}. Any deflection of the leak rate meter shall be cause for rejection.

A4 MECHANICAL ENGINEERING EQUIPMENT

SHEAR AND LEACH CELL - DESIGNS 1 AND 2

Equipment	Estimated Installed Cost (1972$A)
Incremental Feed Ram	55,000
Shear Pack and Hydraulics	625,000
Spare Shear Pack	100,000
Transit Flask (2) - 35 tons at $1.54/lb.	238,000
Shear Chute Assembly	12,000
Rotating Grab Unit - Shear Pack	18,000
Rotating Grab Unit - Basket	12,000
Leacher Baskets - 2 + 1 Spare at $6,000 ea.	18,000
Basket Tipping Station	25,000
Zircaloy Waste Disposal Trolley	15,000
Transit Flask Transfer Port	8,000
Hull Monitoring Equipment	10,000
Maintenance Trolley	5,000
Closed Circuit Television System	72,000
DIRECT COST MECHANICAL ENGINEERING EQUIPMENT	$1,213,000

SHEAR AND LEACH CELL - DESIGN 3

Equipment	Estimated Installed Cost (1972$A)
Fuel Disassembly, Sawing & Pushout Table	420,000
Element Removal Ram	40,000
Element Charging Ram	40,000
Fuel Magazine Loading Tray	12,000
Fuel Charging Station	20,000
Fuel Magazine	50,000
Element Shear	23,000
Spare Shear Pack Inserts (2)	20,000
Fuel Assembly	5,000
Zircaloy Waste Disposal Trolley	15,000
Hull Monitoring Equipment	6,000
Cladding Conveyor	35,000
DIRECT COST MECHANICAL ENGINEERING EQUIPMENT	$702,000

A4 MECHANICAL ENGINEERING EQUIPMENT (Cont'd)

SHEAR AND LEACH CELL DESIGN 4

Equipment	Estimated Installed Cost (1972$A)
Incremental Feed Ram	55,000
Shear Pack and Hydraulics	625,000
Spare Shear Pack	100,000
Shear Chute Assembly	12,000
Leacher Baskets	18,000
Basket Tipping Station	25,000
Zircaloy Waste Disposal Trolley	15,000
Fuel Assembly Transfer Port	18,000
Hull Monitoring Equipment	10,000
Zircaloy Waste Transfer Bin	3,000
DIRECT COST MECHANICAL ENGINEERING EQUIPMENT	$881,000

A5 COST OF PROCESS PIPING

MAN-HOURS REQUIRED FOR PIPE FABRICATION AND ERECTION

(Popper 1970, Page and Nation 1958)

Handling and erecting pipe including hangers and supports
(includes allowance for working in confined area)

N.B.	304L Stainless	Carbon Steel
½"	0.50	0.35
¾"	0.52	0.37
1"	0.55	0.40
1½"	0.60	0.45
2"	0.65	0.50
2½"	0.70	0.55
3"	0.75	0.60
4"	0.80	0.65
5"	0.85	0.70
6"	0.90	0.75

Pipe bends

N.B.	Stainless Steel			Carbon Steel	
	90°	180°	Offset	90°	180°
½"	1.8	2.3	2.8	1.6	2.0
¾"	1.8	2.3	2.8	1.6	2.0
1"	1.8	2.3	2.8	1.6	2.0
1½"	2.3	3.0	3.6	2.0	2.6
2"	2.6	3.5	4.1	2.3	3.0

Butt Welds

N.B.	Stainless Steel	Carbon Steel
½"	1.5	0.8
¾"	1.6	0.9
1"	1.7	1.0
1½"	1.8	1.2
2"	2.2	1.4
2½"	2.5	1.6
3"	2.9	1.8
4"	3.5	2.0
5"	4.3	2.5
6"	5.1	3.0

Radiographs

N.B.	Stainless Steel
4" or less	1.0
5"	1.3
6"	1.5

Total testing man-hours estimated to 1.5 times radiography man-hours.

PRICE OF 304L STAINLESS STEEL PIPE TO ASTM A312 (June, 1972)

Prices for pipe up to and including 1" pipe are based on Schedule 80 pipe. Above this size, pri¢es quoted are for Schedule 40 pipe.

N.B.	Seamless		Welded	
	$A/ft	$A/lb	$A/ft	$A/lb
½"	2.81	2.57		
¾"	3.21	2.18		
1"	4.20	1.93		
1½"	4.60	1.69	3.70	1.36
2"	6.17	1.69	5.03	1.39
2½"	9.75	1.60	9.10	1.57
3"	13.40	1.77	12.80	1.69
4"	22.60	2.10	18.90	1.75
6"	45.50	2.40	34.80	1.84

PRICE OF CARBON STEEL PIPE TO AMERICAN STANDARDS (June, 1972)

Prices for pipe up to and including 1½" pipe are based on Schedule 80 pipe. Prices for pipe from 2" to 6" inclusive are based on Schedule 40 pipe. Above this size, prices quoted are for Schedule 20 pipe.

N.B.	Seamless API 5L Grade A or ASTM A53 Grade B		Seamless ASTM A106 Grade B		Welded ASTM A53 Grade B	
	$A/ft	$A/lb	$A/ft	$A/lb	$A/ft	$A/lb
½"	0.33	0.30	0.36	0.33		
¾"	0.41	0.28	0.44	0.30		
1"	0.57	0.26	0.62	0.29		
1½"	0.65	0.18	0.80	0.22		
2"	0.52	0.14	0.62	0.17	0.47	0.13
2½"	0.77	0.13	0.90	0.16	0.72	0.12
3"	0.98	0.13	0.18	0.16	0.90	0.12
4"	1.34	0.12	1.60	0.15	1.28	0.12
6"	2.34	0.12	2.74	0.14	2.21	0.12
8"					2.71	0.12
10"					3.32	0.12

The above prices do not allow for special discounts which may be negotiated for large quantities.

PRICE OF BUTT WELD PIPE FITTINGS (June, 1972)

Price for fittings up to and including 1" pipe are based on Schedule 80 pipe. Above this size, prices quoted are for Schedule 40 pipe.

Elbows		
N.B.	Carbon Steel $A	Stainless Steel (304L) $A
2½"	1.20	11.60
3"	1.70	16.10
4"	2.70	26.30
5"	6.60	49.40
6"	6.60	69.00

Flanges (Welding neck - 150 lb rating)		
N.B.	Carbon Steel	Stainless Steel (304L)
½"	2.30	9.80
¾"	2.30	12.00
1"	2.30	16.00
1½"	2.80	21.50
2"	3.30	29.90
3"	5.10	52.50
4"	7.00	75.20
6"	11.00	124.60

Reducers (Prices vary slightly according to smaller size)		
Larger Size N.B.	Carbon Steel	Stainless Steel (304L)
¾"	1.30	3.60
1"	1.50	4.20
1½"	1.60	4.70
2"	2.20	6.30
3"	2.90	11.00
4"	3.60	18.00
6"	4.70	46.00

Tees (Prices vary slightly according to smaller size)		
Equal or Larger Size N.B.	Carbon Steel $A	Stainless Steel $A
½"	1.40	3.20
¾"	1.40	5.00
1"	1.40	5.60
1½"	2.00	9.40
2"	2.90	13.20
3"	5.30	28.10
4"	7.60	44.90
6"	16.00	94.60

A5.1 Detailed Breakdown of Components of Piping Cost

1ST URANIUM CYCLE-PULSED COLUMN DESIGN 1

Stainless Steel Piping

Pipe Size	Length (ft.)	Materials Cost ($A)			Labour Man-hours						M + L Cost ($A)	Cost/ft ($A)
		Pipe	Fittings	Total	Handling	Bends	Welds	Penetr'ns	Tests	Total		
½"	664	1,866	209	2,075	332	79	150	140	150	851	7,181	$10.81
¾"	122	392	82	474	63	22	35	18	33	171	1,500	$12.30
1"	325	1,365	48	1,413	179	36	87	62	77	441	4,059	$12.49
1½"	951	4,375	406	4,781	570	153	230	154	192	1,299	12,575	$13.22
2"	110	679	204	883	72	13	35	10	24	154	1,807	$16.43
3"	212	2,841	249	3,090	159	-	118	55	61	393	5,448	$25.70
4"	180	4,068	342	4,410	144	-	168	50	72	434	7,014	$38.97
	2,564	$15,586	$1,540	$17,126	1,519	303	823	489	609	3,743	$39,584	

Carbon Steel Piping

Pipe Size	Length (ft)	Cost ($A)			Labour Man-hours				M + L Cost ($A)	Cost/ft ($A)
		Pipe	Fittings	Total	Handling	Bends	Welds	Total		
½"	124	41	10	51	43	10	14	67	453	$3.65
¾"	48	20	7	27	18	3	6	27	189	$3.94
1"	84	48	14	62	34	10	12	56	398	$4.74
1½"	120	78	20	98	54	4	22	80	578	$4.81
2"	56	29	10	39	28	9	6	43	297	$5.30
	432	$216	$61	$277	177	36	60	273	$1915	

1ST URANIUM CYCLE - CENTRIFUGAL CONTACTOR, DESIGN 2

Stainless Steel Piping

Pipe Size	Length (ft.)	Cost($A)			Labour Man-hours						M + L Cost ($A)	Cost/ft ($A)
		Pipe	Fittings	Total	Handling	Bends	Welds	Penetr'ns	Tests	Total		
½"	433	1,217	120	1,337	217	47	101	122	101	588	4,865	$11.24
¾"	25	80	16	96	13	4	7	7	7	38	325	$12.96
1"	134	563	25	588	74	18	31	25	27	115	1,638	$12.22
1½"	603	2,773	190	2,963	362	122	146	153	122	905	8,393	$13.92
2"	44	271	85	356	29	18	20	10	14	91	902	$20.50
3"	197	2,640	240	2,880	148	-	81	133	42	404	5,304	$26.92
4"	68	1,537	110	1,647	54	-	39	16	17	126	2,403	$35.30
	1,058	$9,081	$786	$9,867	897	209	425	466	330	2,327	$23,829	

Carbon Steel Piping

Pipe Size	Length (ft)	Cost($A)			Labour Man-hours				M + L Cost ($A)	Cost/ft ($A)
		Pipe	Fittings	Total	Handling	Bends	Welds	Total		
½"	78	26	18	44	27	3	10	40	284	$3.64
¾"	60	25	14	39	22	3	10	35	249	$4.15
1"	20	11	5	16	8	3	4	15	106	$5.30
1½"	120	78	34	112	54	4	22	80	592	$4.93
2"	38	20	13	33	19	9	88	36	249	$6.55
	316	$160	$84	$244	130	22	54	206	$1,480	

A5.2 Piping Costs for Design 4

Piping costs in Design 4 are increased relative to the other designs because of the increased length of external lines and the need to provide flanges at regular intervals for easy maintenance.

- Increased Length of External Lines

In 1st Uranium Cycle there are 21 external lines. Average increase in length per line - 40 ft. Hence, increase in total length = 21 x 40 - 840 ft. Cost of stainless steel pipe

$$= \frac{1058 + 840}{1058} \times 23{,}829 = \$42{,}700$$ (prorated from this appendix, section A5.1

- Addition of Flanges

Approximate average line size in plant is 1½" N.B. Assume average distance between flanges is 10 ft.

Cost of two 1½" welding neck, 150 lb stainless steel flanges	=	$44
Cost of additional butt weld, radiograph and 'bolt-up'		$21
Increase in cost per 10 ft		$65

An increase of $6.50 per ft in installed cost is equivalent to a further increase of 47% in the total stainless steel cost.

Hence stainless steel piping cost = $42,700 x 1.47 = $62,800

Revised breakdown of total piping bill for 1st Uranium Cycle is as follows :

Piping	DESIGN 2	DESIGN 4
	$A	$A
Stainless Steel Piping	23,800	62,800
Mild Steel Piping	1,500	1,500
	25,300	64,300
Valves, Ejectors, Pump etc.	12,700	12,700
	38,000	77,000
Add 30% for Wastage, Rewelds Contingency	11,400	23,100
	49,400	100,100
Profit at 15%	7,400	15,000
TOTAL DIRECT COST	56,800	115,100

This is equivalent to a 103% increase in piping cost. Of this increase, addition of flanges contributes 39% and the additional external pipe runs 64%.

A5.3 Assumed Unit Costs and Installation Man-hours for Piping Components

Components	Average Unit Cost ($A)	Average Installation Man-hours
Hand Valves	100	12
Check Valves	100	12
Control Valves	250	24
Steam Traps	300	30
Ejectors	200	20
Thermostatic Air Valves	100	20
Three Way Valves	150	24
Pumps	1300	40
Solenoid Valves	100	24
Jumpers	250	12

A6 UNIT RATES USED FOR ESTIMATING PROCESS INSTRUMENTATION COSTS

Temperature	Installed Cost ($A)
Indicator - TI	350
Indicator - alarm - TIA	400
Indicator-controller - TIC	500
Indicator-recorder-controller - TRC	1600
Level (Density)	Installed Cost ($A)
Indicator - LI (DI)	1600
Indicator-alarm-LIA (DIA)	1650
Recorder - LR(DR)	1800
Recorder - alarm - LRA (DRA)	1850
Indicator - controller - LIC (DIC)	2000
Recorder - controller - LRS (DRC)	2350
Recorder - controller - alarm - LRCA (DRCA)	2400
Flow	Installed Cost ($A)
Indicator - FI	200
Indicator - controller - FIC	900
Indicator - controller - alarm - FICA	1000
Indicator - alarm - FIA	300
Recorder - alarm - FRA	850
Recorder - controller - FRC	1100
Recorder - controller - alarm - FRCA	1200
Pressure	Installed Cost ($A)
Indicator - PI	50
Indicator - alarm - PIA	100
Differential indicator alarm - ΔPIA	900
Recorder controller - PRC	1500
Recorder controller alarm - PRCA	1600

APPENDIX B

PLANT DESIGN DATA

B1 FUEL ASSEMBLY TRANSIT FLASK. DESIGNS 1 AND 2

The flask is required to contain one BWR or PWR fuel assembly, square in cross section, and of maximum dimensions 305 mm (across flats) x 4880 mm.

Radiation shielding requirement (equivalent to that of the highly active cells) is about 250 mm of lead or 521 mm of steel.

The body of the flask consists of a lead cylinder of 250 mm wall thickness, lined internally and externally with steel 25 mm thick. The internal and external diameters of the flask are 432 and 1032 mm respectively.

The flask is closed at either end by lead filled steel doors 350 mm thick. One slab is square in cross section and contains a door sliding transversely to the flask body, for loading and discharge of a fuel assembly. The second slab is circular in section and contains a plug which is withdrawn to permit entry of the ram to feed an assembly into the shear pack.

The flask weighs about 35 tonnes.

B2 WHOLE ASSEMBLY SHEARING AND BATCH LEACHING

B2.1 Fuel Basket

Basis

- Internal diameter : 762 mm.
- Wall thickness : 25 mm.
- Base thickness : 38 mm, with pintle 150 mm deep.
- Walls to be drilled to retain cladding sections and surrounded by a perforated sheath to retain UO_2 fines, while permitting ingress of acid and egress of nitrous fume.
- Basket Capacity : 1 t uranium as sheared fuel assemblies.
- Estimated packed depth for fuel sections : 1.4 m.
- Submergence of cladding on completion of cycle : 150 mm.
- Space between pintle and leacher floor 300 mm.

B2.2 Calculation of Leacher Dimensions

- Volume of basket walls and base submerged in acid (neglecting drillings) = $0.132m^3$

- Volume of fuel cladding from 1 tU 0.05 m^3
- Total volume of basket + leached cladding 0.18 m^3
- Volume of product solution 3.1 m^3/tU
- Height of acid above floor of leacher = 2038 mm
- Approximate dia. of leacher = 1440 mm
- Allowance for disengagement of froth = 700 mm
- Approximate height of leacher (to centre line of off-gas pipe) = 2880 mm.

B2.3 Operational Cycle for Shearing and Batch Leaching of Fuel Assemblies

Assumptions

- Two transfer flasks for conveying assemblies from storage to the shear leach cell.
- Fractional shearing of assemblies to produce a leacher charge of 1 tU. (2.2 PWR assemblies).
- Turn round time of flask not more than 95 minutes.

Operation on PWR Fuel

No.	Operation	Time (min)
1	Receive Flask No. 1. Locate on face of shear leach cell, operate interlocked doors of flask and wall and attach incremental feed ram to flask.	30
2	Shear fuel assembly into 5 cm sections into a charge of nitric acid.	45
3	Remove Flask No.1 from cell face. (Return Flask No. 1 to pond and reload with assembly No. 3).	20
4	Repeat 1-3 with Flask No. 2 containing assembly No. 2	95
5	Repeat 1-2 with Flask No.1 containing assembly No. 3, shearing 0.2 of assembly	39

Cont'd.

No.	Operation	Time (min)
6	Remove shear chute and close lid on leacher	10
7	Heat acid to reflux	30
8	Leach fuel sections	300
9	Cool leacher	90
10	Transfer product liquor from leacher to accounting tank	30
11	Introduce fresh batch of acid into leacher	30
12	Wash basket and contents in fresh acid	10
13	Drain basket above leacher	30
14	Transfer basket to monitor for unleached UO_2	15
15	Monitor basket contents	30
16	Transfer basket to tipping machine	15
17	Discharge leached cladding from basket	15
18	Return basket to leacher	15
19	Position shear chute above leacher	10
	Total cycle time	= 14.3hrs.

Uranium charge fed = 1.0 t

Throughput of one leacher = 1.68 t/d

No. of leachers required = 2

Similar cycle time on BWR assemblies (5 assemblies for charge ≅ 974kgU) = 18.4 hr.

Capacity of 2 leachers for PWR assemblies = 3.36 tU/d

Capacity of 2 leachers for BWR assemblies = 2.54 tU/d

Mean capacity of cell on equal quantities of BWR and PWR assemblies = 2.95 tU/d.

Shearing of integral numbers of assemblies per charge would require an increase in leacher capacity to 1.37 t U (3 PWR fuel assemblies).

In that event,

Capacity of 2 leachers for PWR fuel = 4.3 tU/d.

and mean capacity for BWR/PWR fuel = 3.4 tU/d.

B3 Semi-Continuous Leaching

B3.1 Magazine for Fuel Elements (Design 3)

Basic Assumptions

Fuel

Throughput required	3 tU/d
No. of elements (PWR) equivalent to 3 tU	1340
Total length of element	3760 mm

Shearing

Transfer distance, element from magazine to shear	1220 mm
Length of sheared sections	50 mm
Distance of shear blade movement	50 mm
No. of cuts per element	74

Magazine

No. of magazines discharged per day	N
No. of grooves containing elements per magazine	n
No. of elements per groove	r
Time to insert a charged magazine or remove an empty magazine	T(h)

$$\therefore 1340 = N \times n \times r$$

No. of rods per magazine $n \times r = \dfrac{1340}{N}$

B3.2 Estimation of cycle time

Shearing Time

Time for ram advance (50 mm)	0.63 sec
Time for shear blade advance	0.63 sec
Time for shear blade retraction	0.63 sec

Cycle time per magazine

No.		Time (sec)
1	Load magazine into container	$T \times 3600$
2	Index magazine	2
3	Ram 1st element to shear (1220 mm)	16
4	Feed and shear element	140
5	Retract ram (5300 mm)	66
	Total time to index and shear an element	244
6	Repeat stages 2 - 5 (n-1) times	$\left(\frac{1340}{Nr} - 1\right) 224$
7	Remove magazine	$T \times 3600$
	Total time per magazine	$2T \times 3600 + \frac{1340}{Nr} \times 224$

Hence total time for 1 day's throughput (N magazines)

$$= 2\,NT + \frac{1340 \times 224}{r \times 3600} \ngtr 24$$

i.e. $r \geqslant \frac{1340 \times 224}{(24-2NT) \times 3600}$

i.e. $r \geqslant \frac{83.3}{24-2NT}$

which leads to the following design of mazagine (Figure 59).

No. of Magazines Per Day	Loading/Offloading Time T	No. of Elements Per Groove	No. of Grooves Per Magazine
N	hr.	r	(n + 1)
3	0.5	4	113

B4 PULSED COLUMNS

Typical comparative data (Groenier, 1966, de Witte, 1966) on capacities of pulsed columns operating under extraction and stripping conditions of a Purex (TBP) flowsheet are shown in this Appendix. In general there is reasonable agreement between the data sources; sizing calculations have been based on data for the pulsed columns in the Eurochemic plant (de Witte, 1966) Equivalent capacities of the existing design of NFS pulsed column calculated from these data indicate that the unmodified columns would be operating at about 70% of flooding on a 30% TBP flowsheet at a uranium throughput between 1 and 2 tonnes per day.

A capacity of 3 tonnes uranium per day may be achieved by three approaches.

- By operating the existing design of columns on a flowsheet containing a reduced % of TBP. Nicholson (1967) indicates a throughput of 3 t(U+Pu) per day at 75% of flooding for the highly active extraction column (4C-1) operating at a 15% TBP flowsheet. However, data by Groenier (1966) does not indicate a large capacity increase for a stripping column operating at a reduced TBP concentration.
- Operation of multiple lines of extraction equipment.
- By an increase in the column diameters, adhering to a 30% TBP flowsheet.

Inadequate data exist to predict the flooding characteristics of all columns on a reduced TBP flowsheets, (e.g. for stripping or partition columns); the larger aqueous and solvent flow rates resulting from this approach would result in additional costs for the waste and rework evaporators and the acid recovery equipment.

Operation of multiple solvent extraction lines is not desirable from capital cost considerations.

The diameters of pulsed columns used in this study have been increased to permit operation at 3 tU/d on a 30% TBP flowsheet. It has been assumed that fixed neutron absorbers have been introduced into the columns (e.g. by construction of perforated pulse plates of boron steel) to achieve a level of nuclear safety equivalent to that of the unmodified columns. The required dimensions are listed in this Appendix. The diameters of disengagement sections have been assumed to be twice that of the body of the columns; these sections also contain fixed absorbers.

COMPARISONS OF DATA ON FLOODING CHARACTERISTICS OF PULSED COLUMNS

Column Duty	Continuous Phase	% TBP Solvent	Pulsing Conditions			Column capacity at flooding	
			Amplitude (mm)	Frequency (cycles s^{-1})	Amp. x Freq. (mm s^{-1})	Groenier (1966) ($dm^3h^{-1}dm^{-2}$)	De Witte (1966)
Extraction	Organic	30	25	0.58	14.5	353	
	Organic	30	25	0.83	20.7	216-274	
	Organic	30	25	1.0	25.		320 (extract) 200 (scrub)
	Organic	20	23	1.0	23.	1218	
Stripping	Aqueous	30	25	0.58	14.5	337-453	
	Aqueous	30	25	0.83	14.	222-240	
	Aqueous	30	12.5	1.0	12.5		400
	Aqueous	20	23.	1.0	23.	378-463	

CALCULATION OF DIAMETERS OF PULSED COLUMNS

Column	Diameter of Existing NFS Column (mm)	Continuous Phase	Pulse Conditions		Flooding Capacity ($dm^3h^{-1}dm^{-2}$)	Flowrate		
			Amplitude (mm)	Frequency (cycles s^{-1})		Solvent (dm^3h^{-1})	Aqueous (dm^3h^{-1})	Total (dm^3h^{-1})
4C-1 (EXT)	260	Organic	25	1	320	1420	654	2074
4C-1 (SCRUB)	260	Organic	25	1	200	1420	263	1683
4C-2	260	Aqueous	20	1	400	1804	208	2012
4C-3	108	Aqueous	20	1	320	208	208	416
4C-4	260	Aqueous	12.5	1	400	1804	1562	3366
4C-7 (EXT)	133	Organic	25	1	320	175	471	646
4C-7 (SCRUB)	133	Organic	25	1	200	175	35	210
4C-8	133	Aqueous	12.5	1	400	175	31	206
4C-9/11 (EXT)	260	Organic	25	1	320	1488	2170	3658
4C-9/11 (SCRUB)	260	Organic	25	1	300	1488	300	1788
4C-10/12	260	Aqueous	12.5	1	400	1488	1563	3051

CALCULATION OF DIAMETERS OF PULSED COLUMNS

Column	Existing NFS Columns		Modified Columns	
	Diameters Of Unmodified NFS Columns (mm)	Equivalent Capacity Of Unmodified Columns (30% TBP Flowsheet) (tU/day)	Column Diameter At 70% Flooding (mm)	Diameter Of Disengagement Sections (mm)
4C-1 (EXT)	260	1.7	340	680
4C-1 (SCRUB)	260	1.3	330	780
4C-2	260	2.1	300	600
4C-3	108	1.5	150	300
4C-4	260	1.3	390	780
4C-7 (EXT)	133	1.4	190	380
4C-7 (SCRUB)	133	2.8	140	280
4C-8	133	>3.0	130 (Unmodified)	200
4C-9/11 (EXT)	260	1.0	460	920
4C-9/11 (SCRUB)	260	1.2	400	800
4C-10/12	260	1.5	370	740

B5 SOLVENT WASHING EQUIPMENT (DESIGNS 2 & 4)

This section presents calculations for design of the two and four stage mixer settlers used for washing solvent from the designs incorporating centrifugal contactors.

The designs have been based on a mixer contact time of 60 sec., and a specific settling rate of 1.5 mm^{3}s^{-1}mm^{-2}.

SOLVENT WASHING EQUIPMENT

Mixer-Settler	Flow Rates m^3h^{-1}			Aqueous Fraction	Temperature ^oC	Aqueous Phase	No. of Stages	Emulsion Depth mm	Settling Rate $mm^3s^1mm^{-2}$	Liquor Depth mm	Area of Settler m^2	Mixer Contact Time s	Mixer Side m	Settler Length m	Mixer + Settler Length m	Allowance For End Stages m	Unit Dimension m x m
	Organic	Aqueous	Total														
13C-1	1.82	0.12	1.94	0.06	50	0.2$\underline{M}$ Na_2CO_3	4	50-80	1.5	15	0.33	60	0.5	0.7	1.2	0.25	1.2 x 2.25
13C-4		0.12	1.94	0.06		0.4$\underline{M}$ HNO_3	2	50-80	1.5	15	0.33	60	0.5	0.7	1.2	0.5	1.2 x 1.5
13C-2	1.48	0.1	1.58	0.06	50	0.2$\underline{M}$ Na_2CO_3	4	50-80	1.5	15	0.27	60	0.4	0.7	1.1	0.2	1.1 x 1.8
13C-5		0.1	1.58	0.06		0.4$\underline{M}$ HNO_3	2	50-80	1.5	15	0.27	60	0.4	0.7	1.1	0.4	1.1 x 1.2
13C-3	1.48	0.1	1.58	0.06	50	0.2$\underline{M}$ Na_2CO_3	4	50-80	1.5	15	0.27	60	0.4	0.7	1.1	0.2	1.1 x 1.8
13C-6	1.48	0.1	1.58	0.06		0.4$\underline{M}$ HNO_3	2	50-80	1.5	15	0.27	60	0.4	0.7	1.1	0.4	1.1 x 1.2

B6 PLUTONIUM ION EXCHANGE COLUMNS

B6.1 Ion exchange columns

The NFS design of ion exchange column contains 41.8 litres of an ion exchange resin of maximum loading capacity 150 g Pu/dm^3 of resin. In operation the resin is loaded to 50 g Pu/dm^3.

This section details the operational information on the NFS anion exchange columns. It is inferred that specific flowrates for loading, elution, and washing are 13.3, 10.1 and 12.6 dm^3 h^{-1} dm^{-2} respectively. On this basis, a daily throughput of 3.9 kg Pu could be processed by 2 columns.

The ion exchange requirements for the greater plutonium throughput (27.9 kg/d) of a 3 tU/d flowsheet have been estimated on the following basis:

(a) An increased plutonium loading of 60 g Pu/litre.

(b) A rate of loading with plutonium solution increased to 41.1 dm^{3}hr^{-1}dm^{-2}, corresponding to maximum loading data reported (Long, 1967) for columns at Hanford.

(c) Rates for elution and washing as for the NFS flowsheet.

The requirements for ion exchange columns for a 3 tU/day flowsheet are shown. A minimum of 6 columns are indicated, each operating at 50% of maximum saturation of resin.

DATA ON NFS ION EXCHANGE COLUMN OPERATIONS

OPERATION	Pu Throughput (kg/d)	Volume (m^3/d)	Flowsheet Delivery Period (h)	Implied Flowrate (dm^3h^{-1})	Pu Loading Per Column (kg)	Volume Per Column Cycle (dm^3)	Specific Flowrate ($dm^3h^{-1}dm^{-2}$)	Operational Time (h)
Loading	3.89	0.585	24	24.4	2.1	314	13.3	12.9
Washing	--	0.927	4	232	-	498	126.	2.2
Elution	3.89	0.277	15	18.5	-	149	10.1	8.1
Reconditioning Wash	--	0.093	0.4	233	-	50	126.	0.2
					Total Time per Cycle h			23.4

No. of operational columns required for daily throughput $= \frac{3.89}{2.1} \times \frac{23.4}{24} = 1.8$

Basis of Calculation

Daily Pu throughput = 3.89 kg

Pu loading of resin = 50g dm^{-3}

Resin bed dimensions = 152.5 mm dia x 2290 mm

Vol. of bed per column = 41.8 dm^3

Pu loading per column = 2.09 kg

Cross sectional area of bed = 1.82 dm^2

Volumes per cycle pro rated to Pu throughput

REQUIREMENTS FOR ION EXCHANGE COLUMNS FOR A 3tU/day FLOWSHEET

OPERATION	Pu Throughput (kg)	Volume (m^3d^{-1})	Flowrate (dm^3h^{-1})	Specific Flowrate ($dm^3h^{-1}dm^{-2}$)	Pu Per Column (kg)	Volume Per Column Cycle (dm^3)	Time Per Cycle (h)
Loading	27.9	1.76	75.1	41.1	2.51	158	2.1
Washing	--	2.80	232	126.	--	498	2.2
Elution	27.9	0.830	18.5	10.1	--	149	8.1
Reconditioning Wash	--	0.280	233	126.	--	50	0.2
				Total time per cycle h			12.6

No. of columns required for daily throughput $= \frac{27.9}{2.51} \times \frac{12.6}{24} = \underline{5.8}$

Basis

Pu loading of resin = 60 g dm^{-3}

Loading rate increased to 41.1 $dm^3h^{-1}dm^{-2}$

B6.2 Pumping Head for the ion exchange columns

The pressure drop across a packed bed is given (Perry, 1963) by

$$p = \frac{2\ f_m\ G^2\ L\ (1\ -\ E)^{3-n}}{D_p\ g_c\ \rho \cdot \emptyset_s^{\ 3-n}\ E^3} \qquad \text{lb. force/ft}^2$$

Where L = length of bed (7.5 ft)

g_c = gravitational force (32.2 lb. ft/lb. force sec^2)

D_p = particle diameter (20-40 mesh = 0.0012-0.003 ft)

E = packing fraction (0.32)

$\emptyset_s$ = shape factor (0.9 for fused spherical aggregates)

G = superficial mass velocity of liquor, lb/sec.ft^2

ρ = liquor density, (62.4 lb/ft^3)

f_m = friction factor dependent on the modified Reynolds' number (Fig. 5-64, Perry, 1963)

n = exponent, dependent on the modified Reynolds' number.

Operation	Column Washing	Column Loading
G, lb/sec.ft^2	0.68	0.22
μ poise	0.01	0.01
D_p, ft		
∴ $N_{Re} = \frac{GD_p}{\mu}$	1.0-3.0	0.4-1.0
∴ f_m = (dimensionless)	100-35	300-100
and n =	1	1
$\frac{(1-E)^{3-n}}{E^3}$	13	13
Hence p = lb/ft^2	5550-620	1440-190
	≡ 38-42 psig.	≡ 10-1.3 psig
Equivalent water head, metres wg.	27-2.9	7.1-0.9

Feeding the columns by gravity flow would require a head of 3-27 metres. The ion exchange columns have accordingly been fed by a valved metering pump system, to reduce the cell height.

B7 PLUTONIUM EVAPORATION

B7.1 Plutonium Evaporator

Feed: 23.8 lh^{-1}, 49.8 g Pu l^{-1}, 0.57 m, 60°C.

Product: 5.8 lh^{-1}, 200 g Pu l^{-1}, 3.75m, ~ 107°C.

Maximum steam temperature 135°C.

Sensible heat load = 23.8 x 47 = 1119 kg cal h^{-1}.

Latent heat load ~ 18 x 641 = 11538 kg cal h^{-1}.

Total heat load ~ 12,657 kg cal h^{-1}.

The NFS plutonium evaporator is designed for a maximum heat load of 50,000 BTU h^{-1} $\equiv$ 12,600 kg cal h^{-1}.

The required capacity is therefore slightly in excess of the design load for an existing evaporator. To maintain nuclear safety through geometrical restriction, two evaporators of NFS design have been incorporated in an in-line arrangement, utilising a common overhead condenser.

B7.2 Plutonium Evaporator Condenser

Heat duty: 11538 kg cal h^{-1} (no super cooling assumed).

OHTC ~ 250 BTU h^{-1} ~ 1200 kg cal h^{-1} °C.

Cooling water temperatures assumed 25°C (inlet) and 35°C (exit).

$$\text{Required area for heat transfer} = \frac{11538}{1200 \times (107-30)}$$

$$= 0.125 \text{ m}^2$$

This corresponds to a jacketted pipe, 12 cm dia., about 0.4m in length.

B8 WASTE EVAPORATORS

Waste evaporators have been designed as recirculatory units having disengagement sections with external reboilers. Both reboilers and condensers have been designed with single pass U-tubes as heating and cooling tube bundles, removable vertically from the unit shell. Overall heat transfer coefficients of 200 and 250 BTU h^{-1} ft^{-2} $^oF^{-1}$ were adopted for the reboilers and condensers respectively. The low OHTC in the reboiler reflects the uncertainty in the efficiency of a vertical U-tube design, where partial blanketting of heat transfer area by condensate may occur.

Data on operating conditions for evaporators have been derived from USAEC Docket 50201, and are :

Evaporator	Highly Active	Low Active	Rework	General Purpose
Steam temp. ^oC	130	130	127	170
Feed temp.	30	49	30	38
Bottoms temp.	117	110	104	104
Feed, m^3/d	15.7	146	70	122

A typical design for the HA evaporator has been detailed.

B8.1 Design : Highly Active Evaporator

Basis

Liquor density = 1.06

Liquor specific heat = 0.9

Latent heat of vaporisation = 970 BTU lb^{-1}

Overall heat transfer coefficient = 200 BTU hr^{-1} ft^{-2} $^oF.^{-1}$

Flowsheet feed rate	15.6 m^3 d^{-1}
Dilution in steam force ejection (13%)	2.0 m^3 d^{-1}
Total design feed rate	17.6 m^3 d^{-1}

Evaporation load = $\frac{17.6 \times 10^3 \times 1.06 \times 2.2 \times 970}{24} = 1.66 \times 10^6$ BTU hr^{-1}

Sensible load = $\frac{17.6 \times 10^3 \times 1.06 \times 2.2 \times 0.9\ (117-30) \times 1.8}{24} = 2.41 \times 10^5$ BTU hr^{-1}

Total heat load = 1.90×10^6 BTU hr^{-1}

$\Delta T = 130 - 117 = 13^{o}C$

$\therefore$ heat transfer area required $= \dfrac{1.9 \times 10^{6}}{200 \times 13 \times 1.8} = 406\ ft^{2}$

Kern (1950) gives data on heat exchanger tubing.

1 inch N.B. tube : heat transfer area = 0.192 ft^{2}/ft.run

Total length of tubing required $= \dfrac{406}{0.192} = 2114$ ft.

Assuming a length of tube bank - 7 ft

No. of tubes $= \dfrac{2114}{7} = 302$

Diameter of tube bank (single pass) = 28 inches.

Allowance of 4 in. for U-tube design gives an overall dimension of 7 ft. x 30 in. dia. for the exchanger.

B8.2 Disengagement

Disengagement rate given by NFS = 60 lb h^{-1} $ft.^{-2}$

Overhead rate = 17.6 m^{3} d^{-1}

Area required for disengagement $= \dfrac{17.6 \times 10^{3} \times 2.2}{24 \times 60} = 26.8\ ft^{2}$

i.e. Diameter of disengagement section of evaporator = 5.8 ft

B8.3 Evaporator Condenser

Duty = 17.6 m^{3} d^{-1}

Heat load in condensation $= \dfrac{17.6 \times 10^{3} \times 2.2 \times 970}{24} = 1.56 \times 10^{6}$ BTU h^{-1}

Overall heat transfer coefficient (Kern) - 250 BTU hr^{-1} $ft.^{-2}$ $^{o}F^{-1}$

Assume cooling water inlet temperature = 25^{o}C

outlet temperature = 35^{o}C

$$LMTD = \frac{(117-25)-(117-35)}{2.303_{\log 10} \frac{162}{152}} = 95^{o}C$$

Take LMTD = 70^{o}C to allow for some cooling of the condensate below the boiling point -

Heat transfer area required $= \dfrac{1.56 \times 10^{6}}{70 \times 1.8 \times 250} = 49.4\ ft^{2}$

Use 10 gauge tubing, 1 in. OD. Internal area = 0.192 ft^{2} ft^{-1}

Length of tubing required $= \dfrac{49.4}{0.192} = 257$ ft.

Tube length of tube bank = 7 ft.

$\therefore$ No. of tubes = 37

Over design factor of 15% gives 44 tubes required.

A 12 in. shell diameter would be adequate for 48 tubes (Kern).
Add 4 in. to compensate for U-tube construction
Dimensions of tube bank of condenser are 7 ft. height x 16 in. dia.

Test for cooling water velocity :

$$\text{Mass flowrate} = \frac{1.56 \times 10^6}{10 \times 1.8 \times 62.5 \times 3600} = 0.38 \text{ ft}^3 \text{ s}^{-1}.$$

$$\text{Flow area} = \frac{44}{2} \times 0.421 \text{ in}^2 = 9.3 \text{ in}^2$$

$$\text{Flow velocity} = \frac{0.38}{9.3} \times 144 = 5.9 \text{ ft s}^{-1},$$ which is within recommended limits.

APPENDIX C

COMPARISON OF COSTS FOR THE NFS PLANT WITH DESIGNS 1, 2 & 3

Scale-up factors

NFS $\dfrac{32.5 \times 134.3 \times 1}{24.2 \quad 97.2 \quad 1.11} = 1.67$

AAEC $\dfrac{36.8}{19.6} = 1.88$

SHEAR AND LEACH AREA : - NFS

	$A
Buildings	
Process Mechanical cell and aisles	731,000
General Purpose cell and aisles	246,000
+0.60 x Chemical Process Cell and aisles	
+ Equipment covered under building allocation in AAEC system	904,000
Mechanical Pretreatment	356,000
Dissolution and Feed Preparation	375,000
	$2,612,000
Chemical Process Equipment	$371,000
Mechanical & General Equipment	
Mechanical Pretreatment	1,812,000
Dissolution & Feed Preparation	257,000
	$2,069,000
Process Piping	$377,000
Instrumentation and Electrical	$511,000

SHEAR AND LEACH AREA : - AAEC

	DESIGN 1	DESIGN 2	DESIGN 3
Buildings	$A	$A	$A
Shear and leach as per Table 17	733,000	758,000	313,000
60% Crane supporting building over cells	94,000	98,000	49,000
40% Cranes and hoists	95,000	95,000	83,000
20% Fixtures in operating areas	18,000	15,000	15,000
15% of ejector bulges	18,000	18,000	18,000
	$958,000	$984,000	$478,000

.../Cont.

	DESIGN 1	DESIGN 2	DESIGN 3
	$A	$A	$A
Chemical Process Equipment	$110,000	$110,000	$117,000
Mechanical Equipment			
As per Table 20	1,213,000	1,213,000	702,000
80% of temporary shielding	49,000	49,000	6,000
80% of special tools	16,000	16,000	16,000
	$1,278,000	$1,278,000	$724,000
Process Piping	$69,500	$69,500	$60,400
Instrumentation and Control	$179,000	$179,000	$46,000

SOLVENT EXTRACTION : - NFS

	$A
Buildings	
"Solvent extraction" cells and aisles	$1,415,000
Less area associated with product purification	365,000
(pro rated on volume $\frac{572}{2217}$ x 1,415,000	
	$1,050,000
Chemical Process Equipment	
Partition cycle	$110,000
Uranium	112,000
Plutonium	53,000
Solvent treatment	128,000
	$403,000
Process Piping	
Partition cycle	$138,000
Uranium 61% of	233,000
Plutonium 26% of	35,000
Solvent treatment	211,000
	$617,000
Instrumentation and Electrical	
Partition cycle	52,000
Uranium 69% of	58,000
Plutonium 38% of	9,000
Solvent treatment	112,000
	$231,000

SOLVENT EXTRACTION : - AAEC

	DESIGN 1	DESIGN 2	DESIGN 3
	$A	$A	$A
Buildings as per table 17	$203,000	$136,000	$411,000
Less 50% Solvent washing	20,000	20,000	32,000
	$183,000	$116,000	$379,000
+ 50% operating and pump aisles	14,000	20,000	32,000
+ 50% fixtures in operating areas	44,000	38,000	37,000

.../cont.

	DESIGN 1	DESIGN 2	DESIGN 3
	$A	$A	$A
35% ejector bulges	42,000	42,000	42,000
	$283,000	$216,000	$490,000
Chemical Process Equipment	$569,000	$1,250,000	$569,000
Process Piping	$362,000	$240,000	$362,000
Instrumentation and Control	$221,000	$182,000	$221,000

BREAKDOWN OF EQUIPMENT, PIPING AND CONTROL COSTS FOR SOLVENT EXTRACTION

Partition cycle	41%	
1st U cycle	24%	48%
2nd U cycle	24%	
Pu cycle	11%	