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**FEASIBILITY OF
COMMERCIAL FUEL REPROCESSING
IN THE REDOX PLANT**

OCTOBER 1968

**AEC RESEARCH &
DEVELOPMENT REPORT**

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PACIFIC NORTHWEST LABORATORY

RICHLAND, WASHINGTON

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FEASIBILITY OF COMMERCIAL FUEL REPROCESSING
IN THE REDOX PLANT

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PREFACE

This report on the feasibility of commercial fuel reprocessing in the Redox plant was prepared as the result of an independent review and evaluation study carried out by Battelle-Northwest under the authority of Atomic Energy Commission Contract AT (45-1)-1830. The Commission offers the study to provide potential operators of the facility with further information regarding potential capability of the Redox plant and does not intend to advocate a method of operation or level of revenue to be anticipated from the plant as a functioning fuels reprocessing facility. This study provides a comparison between the Redox plant operation under modest extensions of the existing Hexone flowsheet with an alternate comprising a complete conversion to a Purex-TBP type process. It should be emphasized that conversion cost estimates are based on very conceptual designs and that market forecasts are the judgments of one group of investigators.

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1. INTRODUCTION

This report presents the highlight results of an independent study reviewing the feasibility of converting the Hanford Redox plant to commercial reprocessing of spent power-reactor fuels. The objectives were as follows:

1. Explore the feasibility of Redox conversion, reviewing previous studies and other readily available information.
2. Determine if it might be possible to enhance the conversion feasibility by more fully utilizing existing facilities than had been previously proposed.
3. Define major areas of technical uncertainty in schemes for using the Redox plant.
4. Define areas where further study or development show promise of enhancing the feasibility.

The primary incentive for the study was our interest in furthering the Hanford diversification program.

Two rather extensive previous studies provided the basis for the review. The earlier one, RL-SEP-595,⁽¹⁾ was directed toward adapting the present Redox or Hexone (methyl isobutyl ketone) process to power fuel processing while the more recent one, ISO-800⁽²⁾ was directed toward converting to a Purex or TBP (tri-butyl phosphate) process with special emphasis on multiproduct capabilities. Some of the findings in the later study superseded some findings of the earlier study as a result of the more extensive investigation.

After reviewing these previous studies, two areas in particular appeared worthy of further investigation and were examined in greater detail within the limits of the time allocated for this study.

These were:

1. The prospect for enhancing the feasibility of Redox conversion by more fully utilizing its high throughput capability than had been previously proposed.
2. The prospect for enhancing the feasibility of Redox conversion through flowsheet development that would permit use of the present Hexone process by reducing the large waste volumes associated with it and thus avoid the more costly revisions necessary for conversion to a TBP process.

The principal contribution of this study is in the considerations involving these two items and for this reason, more weight is given to these areas in the discussion that follows. This should not be construed in any sense as depreciating the value of the previous studies. The details of these prior studies are documented in the reports cited and we did not attempt to recite them all here. We have, however, endeavored to include enough information from the previous studies to give our readers an overall picture of the conversion feasibility.

2. SUMMARY AND CONCLUSIONS

The Redox facility, regardless of process, offers an opportunity for a reprocessing venture during the period of the 1970's and early 1980's. The unique feature of the Redox plant is its potential large capacity (6 MT/D or more) for power reactor fuels, available incrementally at modest incremental costs.

As a result of the intensified rate of nuclear power commitments starting in 1965, current projections indicate at least 140,000 MWe installed nuclear power generating capacity by 1980. The effect on fuel reprocessing requirements is profound. The annual U. S. reprocessing requirements will be expanding at a very rapid rate from less than 100 MT in 1970 to over 4000 MT in 1980. This provides an opportunity for relatively large reprocessing plants to begin operation in the mid-1970's.

The growth rate of nuclear power in the western U. S. is also expected to be rapid but lagging several years behind the overall U. S. pattern. The annual fuel reprocessing requirements for the western U. S. will reach 100 MT around 1973 or 1974 and increase to 500 to 700 MT in 1980. Ordinarily this would indicate that a western fuel reprocessing plant would not be feasible till the late 1970's but conversion of the Redox plant could offer an opportunity to enter the market at an earlier date.

There is a recognized economic advantage for high capacity (5 to 10 MT/D) reprocessing plants because the ratio of capital

investment and operating costs to plant design capacity drops sharply with increased design capacity. Our calculations have shown, for example, that for the same return on investment a 10 MT/D plant could price its services at \$12/kg fuel compared to \$27.50/kg fuel for a 2 MT/D plant. Because of this factor, reprocessing charges are expected to begin declining sometime around 1975 when larger plants are feasible. Using an idealized schedule of capacity additions, a projection of reprocessing prices was developed indicating a reduction starting in 1975 from the present \$31.3/kg down to \$12/kg in 1984. This is believed possible but optimistic from a buyer's viewpoint, and was used as the basis for a conservative estimate of revenue for a Redox reprocessing project.

Several conclusions reached as the result of this market examination were:

- Because of the rapid expansion of the reprocessing market, anyone entering the business with a small plant in the early 1970's will run a high risk of early obsolescence due to competition from more efficient large plants in the late 1970's.
- An expandable plant would be highly desirable.
- Three or four separate reprocessing plants will likely be in operation by 1980 and at least one may be as large as 10 MT/D.
- By about 1985, large plants will probably take over completely.

An extensive review and fitness examination of the Redox plant facilities was conducted in connection with the ISO-800⁽²⁾ study and resulted in a conclusion that the plant is sound and fit for additional

service. The fixed-in-concrete piping, the point of greatest concern, is either adequately spared or can be replaced by core drilling techniques.

Two process concepts have been the subject of extensive previous studies for application to power-reactor-fuel processing in the Redox plant -- the Purex or TBP process and the Redox or Hexone process.

Use of the TBP process requires extensive process modifications, including 16 completely new air-pulsed solvent extraction columns. It has the major advantages of low high-level waste volumes (220 gal/MT, alkaline) and better compatibility with developed byproduct recovery processes. Disadvantages include the large concentration factors for the aqueous waste and uranium concentrators and the associated large low-level waste volumes, limited column flow rangeability and possibly some solvent cleanup problems.

Use of the Hexone process would require fewer process changes; primarily four new plutonium columns for higher capacity, two new concentrators and some relatively minor process and instrumentation revisions for criticality control. The major advantages are lower conversion costs and the fact that this is the present process and fewer unforeseen design and startup problems would be encountered. It has the disadvantage of larger high-level waste volumes but low-level waste volumes are much lower.

This study was successful in defining a new Hexone flowsheet that reduces the neutralized high-level waste volumes from

2500 gal/MT to 660 gal/MT. Byproduct recovery appears feasible if desired, but more development work would be required with the Hexone process since it has not received the same attention as the Purex process. Further capital additions would be required for complete byproduct recovery in either process and this aspect requires further study to detail the relative costs.

Some features of the new Hexone flowsheet will require modest R&D expenditures to confirm the anticipated feasibility. These items include demonstration that the greatly increased recycle of aqueous phase raffinates will not seriously affect the limiting ruthenium decontamination and that high-level waste can be fractionated, if desired, by sulfate and phosphotungstate precipitation processes into low volume byproduct solutions amenable to conventional purification processes. The precipitation processes represent a fairly simple extension of present technology and should require only a modest R&D effort to refine. The ruthenium decontamination problem, however, may require an intensive review of ruthenium chemistry and the development of compatible auxiliary decontamination process -- perhaps involving the use of scavenging or complexing reagents.

To offset the effect of declining reprocessing prices, the objective of a Redox conversion can realistically be aimed at an ultimate capacity of at least 6 MT/D. This requires penetration of markets outside the western U. S. by offering prices that compensate for the shipping cost disadvantage. It was assumed that the Redox plant converted to power-reactor-fuel reprocessing could capture one-quarter of the

total U. S. market, with incremental capacity additions up to the time when an expanded capacity of 6 MT/D was reached.

Regardless of which process is utilized, certain basic facilities are required to convert the Redox plant for power reactor fuels. These include facilities for receiving, storing, mechanically preparing the fuels for processing, one or more new critically safe dissolvers, and new waste handling and storage facilities. The total cost of these basic facilities will amount to \$12 to \$13 million depending on options chosen.

The total requirement for initial capital facilities was estimated at \$18.5 million for the Purex-TBP process and \$13.3 million for the Hexone process. Additional capital is required for working capital and for incremental capacity expansions. Capital which may be required for aquisition of the Redox plant from the Atomic Energy Commission is not included. It is left to each reader to deduct his own estimate of aquisition costs from our calculated present worth values; however, each million dollars of additional capital expenditure reduces the net present worth for a 10-year (or longer) operation by only approximately \$680,000 with a 10% interest rate of return or \$750,000 with a 15% interest rate of return. This is because of the depreciation allowance for an additional capital expenditure and the resulting reduced federal income tax costs.

The following table summarizes the results of a cash flow analysis comparing the two process alternates using the reprocessing price projection discussed on page 4 as the basis for a conservative estimate of reprocessing revenue.

	6 MT/D TBP Process, <u>\$ millions</u>	6 MT/D Hexone Process, <u>\$ millions</u>
Initial Capital for Plant Modifications	18.5	13.3
Incremental Capital Additions	5.5	4.8
Project Costs	0.5	0.5
Working Capital	<u>1.0</u>	<u>1.0</u>
Total Capital Investment	25.5 plus aquisi- tion cost	19.6 plus aquisi- tion cost
Net present worth at startup date (1973) with:		
a. 10% interest rate of return after taxes:		
10 yr operation	11.0	12.5
15 yr operation	21.4	21.2
	} less aquisi- tion cost	} less aquisi- tion cost
b. 15% interest rate of return after taxes:		
10 yr operation	2.9	5.4
15 yr operation	8.9	10.4
	} less aquisi- tion cost	} less aquisi- tion cost

These results show that the high waste storage costs associated with the Hexone process can be offset by the lower initial capital required.

The Redox plant with either process offers an attractive opportunity for a reprocessing venture if the initial acquisition cost is reasonable. The lower capital required for the Hexone process favors its selection here but we recommend a more detailed analysis before a final selection is made.

3. FUEL PROCESSING MARKET

a. The U. S. Market

Starting in 1965 the nuclear power industry suddenly caught hold with the growth rate exceeding even the most optimistic estimates. Up through 1964, less than 1000 MWe of nuclear generating capacity had been placed in operation and none was added in 1965. However during 1965, 4900 MWe were contracted for and in 1966 the total was 16,580 MWe,⁽³⁾ representing 50% of the steam-electric capacity ordered that year. One business forecaster⁽⁴⁾ has predicted that "nuclear energy may well be the fastest growing industry in the United States for the remainder of the twentieth century."

The USAEC has revised their official forecast of nuclear generating capacity three times since their "1962 Report to the President" when 40,000 MWe by 1980 was predicted. Their most recent forecast in the spring of 1967 projected 120,000 to 170,000 MWe by 1980 and with a best estimate of 150,000 MWe.

The forecast of nuclear power growth used for the projections in this study is shown in Figure 1 and results in 140,000 MWe installed by 1980. This forecast is based on announced plans and contracts for nuclear stations through 1971, allowing for a small slump in the growth rate at that time due to the probable inability of reactor vendors to keep up with the demand, then starting in

Figure 1. Projected U. S. Electric Generating Capacity

1972 it was assumed that 60% of all additions larger than 400 MWe would be nuclear, increasing gradually to 70% in 1990. This forecast could very easily turn out to be too conservative.

For comparison, other forecasts of installed nuclear generating capacity in 1980 have been: an Edison Electric Institute projection⁽³⁾ of 109,000 to 117,000 MWe for investor-owned utilities alone; a General Electric Company estimate⁽³⁾ of 125,000 MWe; a Westinghouse estimate⁽³⁾ of 150,000 MWe; a prediction⁽⁵⁾ by Atomics International President John Flaherty of 150,000 to 200,000 MWe; and a projection⁽⁶⁾ published in March 1967 issue of Nucleonics of 140,000 MWe.

The projection of spent fuel discharges and quantities available for reprocessing based on our nuclear power growth projection is shown in Figure 2. The quantity of spent fuel available for reprocessing is based on allowing six months between reactor discharge and delivery to a reprocessor for decay-cooling and shipping time. The requirements for reprocessing increase spectacularly after 1970.

Except for a few large prototype developments, the fuel reprocessing requirements between now and the mid-1980's are expected to be almost entirely for the BWR and PWR type light water reactors. Fast breeder reactors are not likely to be economically competitive with light water reactors before about 1985 and will have virtually no

Figure 2. Total U. S. Spent Fuel Availability

impact on the fuels reprocessing market during the period through 1987 considered here for Redox plant operations.

b. The Western U. S. Market

The Federal Power Commission's 1964 "National Power Survey"⁽⁷⁾ which has provided the basis, in terms of electric generating requirements, for most of the recent nuclear power growth projections, divided the U. S. into four sectors for load growth evaluations. The FPC study sector boundaries are shown in Figure 3 along with another illustration from the FPC report showing their projected electrical load densities for the U. S. in 1980.

G. L. Wilfert has developed projections⁽⁸⁾ of nuclear power growth rates in each of these four sectors based on the previously mentioned U. S. nuclear power growth projection and a uniform scaleup of nuclear power growth in each sector. The spent fuel availability by sector based on these projections is shown in Figure 4.

Although the Western Sector reprocessing market will be growing rapidly after 1971, the implications of Figure 4 and the load pattern shown in Figure 3 are clearly that, other things being equal, the western U. S. would ordinarily be the last choice for locating a fuels reprocessing plant. With respect to markets, almost any place in the eastern U. S. would appear more attractive. However, as will be shown later, the unique potential of the Redox plant can offset its disadvantages with respect to markets.

Wilfert also developed a more detailed analysis of nuclear power growth potential in just the Western Sector which was used as the

COORDINATION STUDY AREA AND SECTOR BOUNDARIES

ESTIMATED U.S. LOAD PATTERN 1980

Figure 3. FPC Study Sectors and Projected Electrical Load Densities⁽⁷⁾

Figure 4. Spent Fuel Availability by Sector

basis for the fuel reprocessing load projections in the ISO-800 study. Information from the FPC's 1964 "National Power Survey"⁽⁷⁾ was combined with Bonneville Power Administration estimates, along with various assumptions regarding nuclear power penetration into steam electric capacity additions. Three separate projections were developed: a basic case, an optimistic case and a pessimistic case. The results in terms of nuclear power generating capacity are shown in Figure 5 and in terms of spent fuel availability for reprocessing in Figure 6.

The base case projection shown in Figure 6 was selected for the economic evaluations discussed later in Section 9.

c. Reprocessing Price Projections

In terms of current dollar values, reprocessing prices are expected to decline as the load builds up and especially when larger more efficient reprocessing plants are built. Even the initial NFS contracts had provisions for reduced charges when the total throughput exceeds 300 revenue units/yr.⁽⁹⁾ When the NFS load reaches 385 revenue units/yr their charges, for campaigns of 24 MT and larger, will have been reduced from \$31.3/kg to \$28.5/kg for 3% enrichment fuels, discounting the provisions for escalation to compensate for normal inflationary trends.

An evaluation of the overall price trend is essential to a realistic evaluation of the feasibility of converting the Redox plant. It is conceivable that competitive pricing could reduce the revenue for a fixed reprocessing load by 50% or more during the life of a 15-year Redox reprocessing project. We have therefore developed estimates of possible pricing trends to gain further insight on this question.

Figure 5. U. S. Western Sector Nuclear Power Growth Projections (excluding NPR)

Figure 6. Spent Fuel Available for Reprocessing in Western Sector (excluding NPP)

Using the forecast of fuel reprocessing requirements shown in Figure 2, a plausible sequence of reprocessing plant additions was developed to meet these requirements. The result is shown in Figure 7. The sequence is idealized and simplified and is only one of a number that could be hypothesized. A 10 MT/D plant was the maximum size considered. Regardless of the sequence hypothesized it seems probable that large plants will completely take over the market by sometime around 1985.

It has long been recognized that reprocessing costs would be highly sensitive to plant size and throughput. Some recent work to define these effects has been done in connection with a major study being conducted by the AEC-DRDT to update the program presented in the 1962 Report to the President on Civilian Nuclear Power. The major AEC laboratories as well as representatives of private industry are participating in this study. One of the study groups, the Fuels Recycle Task Force, has the responsibility to develop cost projections for the various steps in the fuel cycle. Although this task force has not reached agreement on a "best" projection for reprocessing charges, some of their findings have been helpful in developing a basis for this study.

Based on Purex type separations plant cost estimates from a number of sources, J. T. Roberts, ORNL, developed the following cost scaling equations for the Fuels Recycle Task Force.

$$C_c = 29 P_c^{0.35}$$

$$C_c = \text{Capital cost, } \$10^6$$

Figure 7. Possible Sequence of Reprocessing Plant Additions
(Total U. S. Requirement)

P_c = Plant capacity, MT/C

$C_o = 0.084 C_c + 0.13 P_o$

C_o = Operating cost, $\$10^6$

P_o = Actual throughput rate, MT/D

Fuel reprocessing charges for plant sizes up to 10 MT/D were calculated using these equations and the following assumptions:

1. Maximum overall operating efficiency = 72%
(80% on-line efficiency and 90% rate efficiency)
2. No allowance for plant clean-out between fuel batches. (This is discussed in Section 9).
3. A 25%/yr fixed-charge-rate on capital for return on investment, recovery of capital, federal and local taxes, insurance and interim replacements. (A 25%/yr charge rate is equivalent to approximately a 10%/yr return on investment after taxes using a sum-of-the-years digits depreciation basis for taxes and a 10-year depreciation period.)
4. Allowing \$3.5/kg for perpetual waste storage costs and for contingencies.

Results of these calculations are shown in Figure 8 and dramatically demonstrate how sharply reprocessing costs can decline with high throughput and large plants.

Figure 8. Effect of Throughput and Plant Capacity on Reprocessing Charges

Even if investors require much higher returns or shorter write-offs than a 25%/yr capital charge represents, the prices for reprocessing services will still drop sharply with large plants. This is shown by the 10 MT/D case with a 40%/yr fixed charge where the fully-loaded-plant reprocessing charge would be \$16/kg versus \$12/kg with 25%/yr fixed charges.

Using the sequence of capacity additions in Figure 7, break-even charges (with 25% capital charges) were tabulated for each plant in the sequence during its lifetime. From this, a smoothed out projection of reprocessing prices was developed assuming that the newest plants would tend to control the going rate. This projection is shown in Figure 9 and is intended to represent the fastest reasonably possible rate of decline; pessimistic from the seller's viewpoint regarding competition and optimistic from the buyer's viewpoint. Some of the variables that contribute uncertainties to this type of price projection include: the degree of stockpiling that will be practical, the strategy and motivation of a reprocessor entering the market, and any limitations that licensing might place on plant capacities.

Figure 9. Estimated Fuel Reprocessing Charges (Prices)

Several conclusions were drawn during development of these price projections:

1. Because of the rapid expansion of the reprocessing market, anyone entering the business with a small plant in the early 1970's will run a high risk of early obsolescence due to competition from more efficient large plants in the late 1970's.
2. An expandable plant would be highly desirable.
3. Three or four separate reprocessing plants will likely be in operation by 1980 and at least one may be as large as 10 MT/D.
4. By about 1985, large plants will probably take over completely.

d. Spent Fuel Compositions

To develop an appropriate design basis for converting the Redox plant to power-reactor-fuels processing, it is important to look at probable fuel compositions for the middle 1970's and beyond. Two of the most advanced design power reactors for which fuel composition data were available were selected for this purpose. These are the Diablo Canyon and TVA - Brown's Ferry reactors.

The Diablo Canyon reactor is 1050 MWe PWR designed by Westinghouse and scheduled for operation in 1972. The TVA Brown's Ferry reactors are 1100 MWe BWRs designed by General Electric with Brown's Ferry #1 scheduled for operation in 1970 and Brown's Ferry #2 scheduled for operation in 1971. The compositions of these fuels are summarized in Table I. The goal exposures and initial enrichments are designers data and the spent fuel compositions were calculated with the ALTHAEA computer program.

Table I

SPENT FUEL COMPOSITIONS

Exposure MWD/MT	Initial Enrichment, %		Final Enrichment, %			Pu Content		Neptunium g/MT
	U-235	Total	U-235	Pu(fiss)	Total	% Fissile	Total g./MT	
Diablo Canyon (PWR)								
Startup Discharges								
15 500	2.20	2.20	0.95	0.94	1.39	76.7	5 700	163
22 400	2.30	2.30	0.69	0.53	1.22	71.7	7 450	275
27 500	2.70	2.70	0.73	0.59	1.32	71.0	8 300	371
31 000	3.19	3.19	0.88	0.62	1.50	71.2	8 770	436
Equilibrium Discharges								
32 000	3.26	3.26	0.88	0.64	1.52	71.1	8 995	460
Pu Recycle Discharges								
32 000	2.63	3.25	0.82	0.77	1.59	64.2	12 000	399
32 000	2.55	3.29	0.84	0.81	1.65	61.0	13 900	389
32 000	2.54	3.32	0.85	0.83	1.68	59.1	14 100	386
Brown's Ferry (BWR)								
Startup Discharges								
20 900	2.19	2.19	0.73	0.58	1.31	74.4	7 700	286
18 000	2.19	2.19	0.86	0.54	1.40	76.5	7 100	236
23 000	2.19	2.19	0.65	0.60	1.25	73.0	8 200	322
19 000	2.19	2.19	0.82	0.56	1.38	76.8	7 400	253
36 000	2.19	2.19	0.32	0.74	1.06	67.9	10 900	569
Equilibrium Discharge								
27 500	2.56	2.56	0.71	0.67	1.38	72.5	9 140	417
Pu Recycle Discharges								
27 500	1.95	2.59	0.64	0.81	1.45	66.0	12 200	352
27 500	1.87	2.65	0.66	0.87	1.53	63.1	13 700	341
27 500	1.88	2.72	0.69	0.90	1.59	61.5	14 600	339

Using these compositions as a basis, one concludes that the enrichment criteria for the ISO-800 study could be relaxed somewhat. The basis for the ISO-800 design was 5% U-235 maximum unirradiated fuel capability (dissolvers) and 2.5% U-235 spent fuel (after burn-out) capability (solvent extraction section).

The dissolvers must be designed to be critically safe in the event that only slightly irradiated fuel must be processed due to premature failure or inadvertent removal from a reactor. The data in Table I show that a 4% enrichment limit for dissolvers should be adequate, at least for the first one. Some of the recent Yankee reactor fuel has been enriched as high as 4.9% but this is a small (185 MWe) and relatively inefficient machine by today's standards and the Diablo Canyon Reactor should be more representative of the long-term trend. Relaxing the dissolver enrichment criteria is probably not as important, however, as the criteria for solvent extraction.

After dissolution the dissolver solution can be isotopically analyzed, as proposed in ISO-800, to provide more positive control for enrichments admitted to the rest of the plant. The data in Table I shows that the total enrichment in the spent fuel does not exceed 1.7% and after partition of the U + Pu, the uranium enrichment does not exceed 1.0%. As already mentioned, some partially irradiated fuel may have to be processed but this should be only a small portion of the total and could be diluted with fully irradiated fuel. The fuel cycle economics are so dependent on high burnup

that it is inconceivable that gross quantities of partially burned fuel would have to be processed. A 2% enrichment limit for the uranium solvent extraction section should be a safe basis. This is important if utilization of the present Hexone process is considered since the existing uranium columns are safe for up to 2% enrichments.

Assuming plutonium recycle is initiated in 1972, there would be no appreciable impact on reprocessing requirements until around 1978 because of a 4 to 6 year lag between reactor charging and reprocessing and because when the initial recycle fuel is discharged it will represent only a small portion of the total fuel discharged at the same time.

For these reasons, the following is suggested as a design basis:

Maximum dissolver enrichment (U + Pu)	4.0%
Maximum co-decontamination column enrichment (U + Pu)	2.0%
Maximum uranium column enrichment	1.5%
Maximum fissile content of Pu	85%
Maximum capacity of Pu columns	
(a) Prior to 1978 - fissile Pu	7,500 g/MT
total Pu	11,000 g/MT
(b) After 1978 - fissile Pu	9,000 g/MT
total Pu	15,000 g/MT

For the infrequent occasions, if ever, when these limits might be exceeded it should be possible to handle them with special procedures.

4. DESCRIPTION OF PRESENT REDOX FACILITIES

a. Plant Facilities

The Hanford Redox plant was the first continuous-solvent-extraction-process plant ever built for remote processing of irradiated uranium fuels. It was constructed to process uranium metal fuels irradiated in the Hanford production reactors and was designed for completely remote operation and maintenance. The plant was started up in 1952 and operated until late 1966. It was shut down due to a general reduction in production levels at the Hanford complex, and because of the capability of the newer and larger Purex plant to handle the entire production requirements. A well managed program of routine maintenance and systems improvements kept the plant in good conditions and a number of technological improvements were incorporated during its period of operation.

The Redox building, 202-S, is a massive concrete structure 475 ft long x 162 ft wide with a tower section, known as the Silo, at one end that rises 120 feet above grade to accommodate the solvent extraction columns. The plant contains nine process cells in two rows with a pipe trench and air tunnel between. The cells are 27-1/2 ft deep x 12 ft wide and range from 25 to 115 ft in length. The column enclosure in the Silo is 84 ft high x 67 ft long x 11 ft wide and contains positions for 13 solvent extraction columns.

A recent improvement provided for consolidation of process instrumentation to centralize feed preparation and column control instrumentation at one location in the Silo operating gallery.

An extensive ventilation system insures a flow of air from the operating and service areas to the process areas. The ventilation system exhausts through an 85 ft square x 8 ft deep sand filter (99.8% removal of particulate contamination) and a 200 ft high stack. A separate vessel vent system is provided and an inert gas blanket system is provided for the hexone-containing tanks. A fire detection and water-fog system is provided for the areas where hexone is used.

Supporting facilities include: UNH storage facilities, 203-S and 204-S, with 210,000 gal total capacity; a silica-gel facility for final cleanup of off-standard UNH, 205-S; a chemical tank farm, 211-S; a plutonium concentration and load-out facility, 233-S; and a hexone treatment and storage facility, 276-S.

b. Physical Condition and Fitness

An extensive review and fitness examination of the Redox plant facilities was conducted in connection with the ISO-800 study. The examination included visual inspections and a review of operating and maintenance records as well as current engineering studies. The basic conclusion was that the plant is sound and fit for additional service.

The permanently fixed-in-concrete process piping was the greatest concern. Of 41 lines serving the solvent extraction system between the silo and the canyon cells, 24 are required for process service, 2 have failed, leaving 15 spares which is believed adequate. Sufficient routings also exist between process cells and in the event new lines are required they can be installed by core drilling, a technique successfully used both at Redox and Purex.

The outside canyon equipment and service systems including steam, air, water and electrical utilities plus chemical makeup and product storage facilities as well as operating instrumentation were all found to be in good condition.

Areas where major renovation efforts could be required include: the carbon steel portions of 3 support pads in "C" cells are corroded but techniques are available to re-establish these; 8 electrical nozzles and 12 pipe nozzles are out of service in the cells but can be replaced by core drilling; in the pipe trench, two lines are known to leak and spares are being used, one is suspect and one is plugged so some replacement by core drilling may be required; a major decontamination of the canyon crane would be required before it could be placed in first-class condition; three viewing windows and a number of light fixtures are in need of replacement in the silo; decontamination for the column carrier system (used for column replacement) is needed; the canyon walls and top sides of the cell cover blocks require decontamination before personnel entry can be permitted (included in deactivation program); and some minor modifications in the ventilation system are desirable.

c. Redox (Hexone) Process

The Redox process employs hexone (methyl isobutyl ketone) solvent extraction to recover and purify uranium and plutonium from irradiated reactor fuels. The process is based on the relative extractability of uranium and plutonium into hexone under certain conditions and the relative inextractability of fission products

and most inert contaminants. The transfer of uranium and plutonium to the hexone phase is controlled by the aqueous phase salting strength [aluminum nitrate (ANN) concentration] and the plutonium valence: high ANN concentrations favor extraction into hexone while reduction of plutonium to Pu(III) makes it almost inextractable. A solvent extraction cycle consists of an extraction contactor (packed column) in which the uranium and plutonium are quantitatively extracted into hexone out of a highly salted aqueous phase and a stripping contactor in which the uranium and plutonium are quantitatively removed from the solvent with water or a dilute acid solution. Several such cycles are required to obtain the high decontamination from fission products required to avoid high radiation dose rates on subsequent product handling. The plutonium decontamination requirements for power reactor fuels may be relieved somewhat due to the fact that higher fission product concentrations will be allowable in what is in any event high activity plutonium.

Several refinements of the basic cycle are employed. For example, the extraction contactors contain scrub sections wherein the solvent phase containing the extracted U and Pu is scrubbed with an ANN-salted stream to remove those contaminants with low distribution ratios. A partition column (the 1B column) is employed to selectively strip the plutonium from the solvent phase by reducing it to inextractable Pu(III) leaving the uranium in the solvent phase.

The basic flow scheme used in the Redox plant at shutdown was similar to that illustrated in Section 7 (Figures 10 thru 15). The uranium

and plutonium were extracted and stripped together in the head-end or codecontamination cycle (HA and HC columns) after prior oxidation of the Pu to Pu(VI) with sodium dichromate. The product from this cycle served as feed to the second, or partition, cycle (the 1A, 1B, and 1C columns). The 1B column, with a center feed point, served to partition the plutonium using ferrous sulfamate as the reductant; a solvent scrub stream re-extracted the small amount of uranium stripped along with the plutonium. The final uranium cycle (2D and 2E columns) was much like the first except for the absence of plutonium. The plutonium product from the 1B column was reoxidized with dichromate to a mixture of Pu(IV) and Pu(VI), both extractable under acidic flowsheet conditions, and processed through two additional extraction cycles (the 2A, 2B, 3A, and 3B columns).

Experience has shown that ruthenium is the most difficult fission product to remove in the Redox process. Decontamination from both ruthenium and the other fission products can be improved during solvent extraction by extracting the uranium and plutonium from acid-deficient ANN solutions (acid deficiency is obtained when some of the nitrate is replaced by hydroxide -- because of hydrolysis, the system still remains acidic). However, acid-deficiency must be used sparingly in the HA and 1A columns and not at all in the 2A and 3A columns because of the low extractability of Pu(IV) and the tendency of it to polymerize and deposit in the column at low hydrogen ion concentration. [Although plutonium is maintained in the plus six valence in the HA and 1A columns by dichromate ion, plutonium(IV) deposition

from the solvent phase has occurred at low-acidities because of hexone reduction of the Pu(VI)]. Recent practice has been to treat the uranium and plutonium product streams with ozone to oxidize residual ruthenium to the volatile ruthenium tetroxide. The procedure, however, is limited by the capability for containing the volatilized ruthenium within the off-gas treatment facilities.

During steady-state operation, the use of fresh ANN for salting agent is kept at a minimum by recycling the highly salted aqueous raffinates* from extraction columns where fission product concentrations are low back to that part of the process where fission product concentrations are high. Minimal flows of fresh salting agent solution, sufficient to provide adequate fission product decontamination, are added as scrub streams to the tops of all extraction columns downstream from the HA column. Generally, the aqueous raffinate flows from the extraction columns are proportioned to return some of the salting agent to the parent column (recycle), and to return the rest to the prior cycle feed (backcycle). In exceptions to this rule, the 3A raffinate is returned to the 1A feed and the 2A raffinate is used as the HA column scrub.

Neptunium is retained in the acid-deficient raffinate from the HA column. It is accumulated in the system for byproduct recovery by recycling a portion of the raffinate to the feed and acidifying

*Raffinate may refer to either aqueous or organic streams and refers to the feed stream after U or Pu has been extracted.

the remainder to allow dichromate to oxidize the Np(V) in the raffinate to the extractable Np(VI). The neptunium in the acidified raffinate fraction is then extracted in the 1S column with hexone, stripped back to the aqueous phase in the 1N column, and returned to the HA column raffinate concentrator for recycle. Periodically, the neptunium is recovered on a plant-campaign basis.

The raffinate from the 1S column represents the only true waste outlet for fission products and salting agent from the process during steady-state backcycle operation. This waste is concentrated and neutralized to pH ~ 13 for storage. Neutralization conditions are chosen to maintain the aluminum in the supernate as the soluble sodium aluminate (approximately 1.2 M NaAlO_2).

Solvent raffinates from the various extraction cycles are combined, scrubbed in the 1-0 column with a small amount of water to remove uranium and plutonium, and distilled. Hexone is appreciably soluble in aqueous streams, but it can be almost quantitatively recovered by steam-stripping the aqueous raffinates and the condensates from the raffinate stills. The recovered hexone is combined with the washed solvent raffinate in the hexone still, and the distilled solvent is stored in the lightly shielded 276-S hexone storage facility for reuse.

5. BASIC FACILITIES REQUIRED FOR REDOX CONVERSION

The Redox plant was designed to receive and process Hanford production reactor fuels. These are small aluminum-clad uranium metal elements around 1-1/2 inches in diameter and five to nine inches long. In contrast, the power reactor fuels are complex bundles of approximately 1/2 inch diameter zircaloy or stainless-steel clad elements packed with high density uranium oxide and the assemblies may range up to 15 feet in length with 49 to 204 elements or rods in an assembly. Zircaloy is presently preferred over stainless steel as the cladding material. Regardless of the chemical process selected for recovery of the spent-fuel components, certain basic facilities will be required to receive, store and prepare these power reactor fuels for processing. In addition, new waste handling and storage facilities will be needed and, for high capacities, an imbalance in the uranium and plutonium processing capability will have to be corrected. The facilities required to accomplish these things are described briefly in the following paragraphs.

a. Fuel Receiving, Storage and Mechanical Treatment

The ISO-800 study proposed a new annex adjoining the east end of the Redox plant for the functions of receiving, storing and mechanically preparing the power fuels for dissolution. The facility would be divided into a fuel receiving and storage area and two large "hot" cells for mechanical treatment.

The fuel receiving and storage area would provide for unloading of fuel shipping casks from either rail or truck shipment with a

150-ton bridge crane; a 25 ft x 25 ft x 50 ft deep unloading pool; a cask decontamination area; a storage pool for up to 650 fuel assemblies (receipts from a single reactor discharge will probably not exceed 200 assemblies but receipts from several reactors simultaneously would need to be provided for); and an additional crane for handling the fuel assemblies which would be stored in individual containers.

The mechanical cell would provide for sawing off the end hardware of the fuel assembly and for chopping the fuel elements into small pieces with a hydraulic shear in preparation for a leach dissolution of the core material.

The chopped fuel would be collected in stainless steel dissolver baskets containing thin expendable carbon-steel liners. A critically-safe storage cell facility would provide interim storage for the loaded dissolver baskets prior to transfer to the dissolver cell.

The total cost for these facilities was estimated at \$5.6 million. The earlier RL-SEP-595 study⁽¹⁾ had proposed modifying the existing J-Cell in the Redox canyon for a fuel storage facility. The estimated cost for the receiving, storage and mechanical treatment facility in that study was only \$2.5 million, but the more thorough ISO-800 investigations determined the earlier concept to be inadequate.

The ISO-800 concept for these facilities is thorough and reasonable in all respects, and opportunities for cost reduction are not apparent.

b. Low-Level Waste

The criteria provided by the AEC for the ISO-800 study included a prohibition against cribbing any low activity water. A design concept was developed for processing all low-level wastes consistent with the discharge limitations in 10 CFR part 20 and 10 CFR part 100. The facilities proposed provided for decontamination of various low activity wastes by evaporation and fractionation after neutralization of any acidity with discharge to the ground at less than MPC limits via retention basins. The estimated cost for these facilities was \$4.0 million for an instantaneous capacity of 2 MTU/D. The major cost for expanding the capacity to 3 MTU/D would be in providing larger retention basins.

We have estimated that the cost for these low level waste facilities could be reduced to \$3.0 million if the existing Hexone process is utilized instead of converting to the Purex-TBP process. This reduction results because the volume of process condensates produced in the Hexone process is a factor of four lower than with the TBP process and also, because a simple evaporator can be used instead of an acid fractionator for high-level waste condensate processing.

We have not attempted to investigate the possibilities or probabilities of escaping from the requirement for low-level waste processing, but have assumed that effluent discharge limitations will be strict and that these facilities, or their equivalent will be required.

c. High-Level Waste

Some type of storage facilities will be required for high-level aqueous wastes regardless of the process employed for power-fuel processing. Because of differences in compositions and quantity, however, there can be substantial differences between different processes in the cost of providing these storage facilities. Provision must also be made for perpetual care of any stored high-level waste.

For the TBP process detailed in the ISO-800 study, both acid and alkaline waste storage was discussed. The limitations and costs were as follows:

Acid Waste Storage

Initial minimum volume limited by heat rate	220 gal/MT
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Ultimate minimum volume after 2 to 4 years storage, limited by solubility (recycled for reconcentration)	55 gal/MT
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Cost of storage vault, transfer facilities and first 2-100,000 gal SS tanks	\$2.8 million
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Cost for additional SS tanks as required	approx. \$6.0/gal
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Tank life expectancy	50 years
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Alkaline Waste Storage

Minimum volume limited by solubility	220 gal/MT
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Cost of transfer facilities and first two carbon steel tanks	
2-300,000 gal tanks	\$2.5 million
2-600,000 gal tanks	\$3.4 million
2-1,000,000 gal tanks	\$4.5 million

Cost for additional carbon steel tanks	approx. \$1.50/gal
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Tank life expectancy	20 years
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The ISO-800 study placed heavy emphasis on multiproduct capability and favored acid storage because of better compatibility with the D2-EHPA waste fractionization process and potentially profitable recovery of fission products. We have taken the view here that profitability of fission product recovery remains to be proven and should be justified separately on its own merits.

Because the initial cost is somewhat lower for alkaline waste storage (\$2.5 million compared to \$2.8 million) and the cost of replacement tanks appears identical ($220 \text{ gal/MT} \times \$1.50/\text{gal} = \$330/\text{MT}$ versus $55 \text{ gal/MT} \times \$6.00/\text{gal} = \$330/\text{MT}$), alkaline waste storage was chosen for the economics developed in this evaluation. A more detailed analysis in an ORNL study⁽¹⁰⁾ resulted in a virtual standoff between cost of alkaline versus acid storage.

The cost of initial storage plus perpetual care and storage has been indicated⁽¹⁰⁾ to be on the order of \$3.00/gal for alkaline wastes. As a basis for the economic evaluation discussed in Section 9 we provided \$2.5 million for initial high-level waste in facilities for the TBP process and \$3.4 million in the case of the Hexone process because of a higher rate of waste accumulation. In addition, \$3.00/gal was assessed for each gallon of alkaline waste produced for replacements and perpetual storage costs. This may be more than is necessary since the investment in the initial facilities could reduce the assessment per gallon somewhat.

The \$3.00/gal would be placed in an escrow fund and used to provide new tanks as the initial storage capacity filled up (one

tank would always remain as a spare in the event of a failure). The fund would also provide, through accumulated interest, for replacement of failed tanks and for tank farm surveillance in perpetuity.

d. Minimum Process Requirements

The present Redox dissolvers are, without question, not suitable for power-reactor-fuel processing. A new critically-safe chop-leach-process dissolver with a capacity of 2 to 3 MT/D for power fuels can be provided for an estimated cost of \$250,000. Three dissolver cells are available so an ultimate dissolver capacity of 6 to 9 MT/D could be provided. A Zirflex type dissolver would be another possibility and would avoid the fuel chopping requirements but the mechanical cell would be traded for the problems of Zirflex waste solution processing to recover dissolved core material and a more difficult dissolver design problem. It is doubtful if a net gain would be made and flexibility would be reduced as well.

The solvent extraction section of the present Redox is out of balance for power fuel processing with respect to uranium and plutonium processing capability. The uranium section has a capacity for more than 6 MT/D (less than 2% enrichment) but the plutonium section is limited to about 10 kg/D which in effect limits it to around 2 MT/D on power-reactor-fuel processing. Our flowsheet study (see Section 7) has shown that the plutonium section capacity could be increased to process 50 kg/D or the equivalent of 6 MT/D of power fuel by replacing four columns and adding 2 concentrators. The ISO-800 study estimate for replacement of four uranium columns (not

needed if enrichment <2%) for Hexone processing was \$350,000 and we have assumed that new Pu columns and concentrators plus instrumentation changes could be provided for within \$500,000.

In any consideration of converting the Redox plant for power-fuels processing, prime consideration must be given to critical safety hazards which might develop. The enrichments and quantities of fissile material handled will be higher than the Redox plant was designed for. A system of basic, operative and administrative controls has been employed at Redox and was proposed in the ISO-800 study. The basic or primary controls include concentration, mass and geometry controls with geometry controls preferred wherever possible; the operative controls include instrumentation and sampling-analytical functions; and administrative controls include review and approvals of designs, procedures and in some cases continuing supervisory approval prior to certain routine procedures.

Some minimum modifications will undoubtedly be required for critical safety requirements. For example, modification of some of the cell sumps, duplicate instrumentation, instrument interlocks and possibly some vessel revisions. The scope of this review has not included a detailed examination of these requirements but we have endeavored to comply with accepted critical safety practice in options we considered. The ISO-800 study allowed \$150,000 for miscellaneous instrumentation and routing changes for critical safety in connection with replacing the four uranium columns previously mentioned. It would seem prudent to allow a minimum of \$150,000 for critical mass considerations in any concept for power-fuels processing.

e. Total Cost of Minimum Required Facilities

The total cost of the minimum facilities required for converting Redox to power-fuel processing based on the considerations just discussed is summed up in the following table.

Estimated Minimum Conversion Requirements

<u>Capacity</u>	<u>Less than 2 MT/D, \$ millions</u>	<u>Greater than 2 MT/D,* \$ millions</u>
Fuel receiving, storage and mechanical treatment	5.6	5.9
Low-level waste facilities	3.0	3.0
High-level waste facilities	3.4	3.4
Critically safe dissolver	0.25	0.25
Pu columns and concentrators		0.5
Critical Safety Considerations	0.15	0.15
Miscellaneous contingencies		<u>0.15</u>
TOTAL	12.40	13.35

*Other incremental capital additions would be required later as the load increased beyond 2 MT/D as outlined in Section 9.

It is apparent that substantial costs are involved even if only minimum revisions are made in the basic Redox process.

6. CONVERSION TO TBP PROCESS

a. Process Description

The TBP process proposed in the ISO-800 study is described in summary form in this section. The solvent is 30 vol% tributyl phosphate in a normal paraffin hydrocarbon diluent. The flow scheme employs an initial partition cycle in which uranium, plutonium and neptunium are coextracted in the 1A column from a nitric acid solution; the plutonium is selectively stripped from the solvent in the 1B column with a ferrous-hydrazine reductant added to reduce the extractable Pu(IV) to the inextractable Pu(III); and the uranium and neptunium are stripped with dilute nitric acid in the 1C column. The uranium is purified in two more solvent extraction cycles (2D, 2E, 3D, and 3E columns) with the flowsheet adjusted to divert the neptunium to the 2D column raffinate stream. The plutonium in the aqueous stream from the 1B column is oxidized to Pu(IV) with nitrite ion and purified in two additional solvent extraction cycles (2A, 2B, 3A, and 3B columns).

The aqueous raffinates from the 2D, 2A, and 3A columns are combined and processed in three columns (2S, 2F, and 2G) to recover neptunium. The 2S column serves to extract the uranium and plutonium values away from the neptunium which is then reduced to the extractable Np(IV) with ferrous ion and purified in the 2F/2G extraction cycle. A fixed bed anion exchange column is added for final neptunium purification and concentration.

The process uses extensive solvent backcycle: the 3E solvent raffinate is reused as 2D column extractant, the 3B solvent raffinate as 2A extractant, etc. Only the 1C, 2E, and 2G solvent raffinates are given the conventional sodium carbonate wash -- the latter two streams in the 2-0 column and the former in the 1-0 column. The only significant aqueous stream backcycle is the 3D column raffinate to the 2D column feed via the 1C uranium concentrator. Any uranium and plutonium that follow the other aqueous raffinate streams to the neptunium recovery columns are returned to the 1A column in the solvent raffinate from the 2S column.

The process uses a total of 16 extraction columns, 7 aqueous concentrators, and one ion exchange column. Most of the present concentrators and tanks can be reused, but the lower extraction efficiency obtained with the TBP process will require that the present packed columns be replaced with air-pulsed columns. Critically-safe diameters of 5 to 7 inches are proposed for all columns (except for the 1C column which would be 10 inches). Existing plutonium feed tanks would also be replaced with critically-safe-geometry tanks.

b. Advantages and Disadvantages

The major advantage of the TBP process is the production of a low volume waste. The waste volume will be limited initially by the ability to dissipate heat safely rather than by the solubility of the salts as is the case with the Redox process. The waste volume leaving the TBP process is limited to about 200 gal/MT by cooling requirements and later to 55 gal/MT (acidic) or 220 gal/MT (basic)

by solution stability. Redox waste volumes range upward from about 600 gal/MT, depending on the amount of plutonium present and the extent of backcycle.

Secondary advantages are the recovery and reuse of the nitric acid salting agent and the compatibility of the low-salt waste with fission product fractionization and recovery processes. (The fission product recovery aspect was heavily emphasized in the ISO-800 study.) The TBP solvent recovery system is somewhat simpler than the hexone recovery system because of the lower aqueous-phase solubility of the TBP and the ability to remove process-affecting impurities effectively with a simple alkaline wash. On the other hand, hexone after distillation is radioactively pure enough to be used in all cycles and only one solvent recovery system is required compared with two for the Purex process.

There are several areas of uncertainty in the proposed TBP process--some relatively minor. Among these are the undemonstrated use of commercially unobtainable ferrous nitrate stabilized with hydrazine as reductant and the use of extensive solvent backcycle. These innovations were made, in part, to eliminate sulfate from the process to make the waste more compatible with fission product fractionization. This objective is probably of relatively minor value.

Of more concern is the potential for increasing the amount of solvent degradation products in the systems both through the solvent backcycle and the limited provisions for solvent washing. Usual practice requires two or more wash contactors per solvent waste stream to assure adequate removal of dibutyl phosphate (DBP). Failure to remove all of

the DBP can lead to the formation of insoluble DBP compounds, high waste losses, lower column capacities, and poorer decontamination from zirconium and niobium.

A major drawback is the large concentration factor required in the aqueous waste and the uranium concentrators. Approximately 2000 to 4000 gal of water must be boiled off in each of the large concentrators per ton of uranium processed. This is at least double and in some instances more than five-fold greater than the boiloff required by the present Hexone process -- even considering waste back-cycle added to the uranium concentrators. These concentration requirements appear to limit the plant capacity to something like 3 tons of uranium per day by the heat duty obtainable from several of the existing concentrators.

The flooding capacity of TBP pulse columns is generally only 50 to 75% of the capacity of Hexone-process packed columns. This may prove a serious restriction in the capacity of the plant, particularly if the recommended critically-safe geometries are used. (The indicated column capacity was given as 4 MTU/day but our calculations indicate the capacity of several columns may be somewhat less.)

The ISO-800 process study also proposes using a single column for the 1B column. Usual Purex practice is to use separate columns for the 1B column functions of plutonium extraction and uranium scrubbing because of the large disparity of flows between the extraction and scrub sections and the difficulty of matching optimum pulsing

conditions in both sections. Hexone packed columns will accept a wider rangeability of flows; and a single, dual diameter 1B column has been used satisfactorily in the present Hexone process.

c. Capital Cost of Required New Facilities

The chemical processing facilities proposed in the ISO-800 study for converting the Redox plant to a TBP process with an instantaneous capacity of 3 MT/D or 700 MT/yr actual capacity were as follows:

1. New geometrically-safe annular dissolver in A-Cell.
2. New insert in present Mark V dissolver in B-Cell for processing metallic uranium fuels.
3. 16 new air-pulsed solvent extraction columns with 4 MT/D instantaneous capacity. Three of these columns for Np recovery would be located in E-Cell and the two solvent-wash columns would be located in G-Cell. The main 11-column solvent-extraction battery would be located in the Silo.
4. A new hot-cell annex for Np purification by anion exchange.

The cost of modifying the plant with these new facilities was estimated at \$6.2 million which includes a number of new tanks, instrumentation, etc., as well as the primary process facilities listed above.

In addition, facilities for Fuel Receiving, Storage and Mechanical treatment as well as low-level and high-level waste handling as previously discussed in Section 5 would be required. The two base estimates for total capital costs for the TBP process conversion in the ISO-800 study were as follows:

<u>Facility</u>	<u>Acid Waste Storage</u>	<u>Alkaline Waste Storage</u>
Fuel receiving, Storage and Mechanical treatment	\$ 5,600,000	\$ 5,600,000
Chemical Processing	6,200,000	6,200,000
High-Level Waste	2,800,000	2,500,000
Low-Level Waste	<u>4,000,000</u>	<u>4,000,000</u>
TOTAL	\$18,600,000	\$18,300,000

The economic evaluations in Section 9 are based on an ultimate capacity expansion to 6 MT/D. Because of the limitations of the aqueous concentrators and pulse column capacities pointed out in the previous discussion (6.b) as well as other uncertainties, we allowed an additional \$1,300,000 for the initial facilities (\$300,000 for head-end facilities, \$500,000 for chemical processing facilities and \$500,000 for low-level waste facilities) for the TBP conversion expandable to 6 MT/D in the Section 9 evaluation. The higher capacity would require new high-capacity aqueous concentrators and higher capacity pulse columns. The total initial capital cost for the TBP process expandable to 6 MT/D and including neptunium recovery would then be \$19.9 million or \$19.6 million for acid and alkaline waste storage respectively. Other incremental capital additions were also allowed as the load requirements increased and these are detailed in Section 9.

The estimated costs for several additional options related to waste processing for byproduct recovery were also provided in the ISO-800 study. The costs are in addition to the above basic costs:

Cost of additional facilities for processing up to 25 kg/D of Np target elements through the TBP process. (Dissolver capacity would limit processing to 8 kg/D if only one dissolver)

\$ 40,000

Cost for facilities to recover Am-Cm using combination D2-EHPA and TBP solvents (PAMEX Process)

a. Chemical Processing

\$ 2,400,000

b. Waste Facilities

3,500,000

Total

\$ 5,900,000

Cost for recovery of one or more of Cs, Sr, Ce, Pm and other rare earths and Am-Cm. A combination phosphotungstate acid precipitation process for Cs plus D2-EHPA solvent extraction for the others is proposed.

a. Chemical Processing

\$ 6,800,000

b. Waste Facilities

3,600,000

Total

\$10,400,000

7. CONVERSION USING PRESENT HEXONE PROCESS

a. RL-SEP-595 Study

An extensive study of using the present Hexone process for power fuel processing in the Redox plant was issued as RL-SEP-595 in 1965. A flowsheet for power reactor fuel processing proposed in this study was similar to the one described in Section 4 except that the decontamination cycle (the HA and HC columns) would be replaced by the partition cycle (the 1A, 1B, and 1C columns) to separate the uranium and plutonium as quickly as possible for criticality control. The additional decontamination required by the uranium stream would be obtained by adding a third uranium cycle -- the 3D and 3E columns. This flowsheet made maximum usage of the existing equipment and required only about \$250,000 for solvent extraction equipment changes and additions for criticality control for equivalent enrichments up to 5% U-235.

The RL-SEP-595 hexone process had two major drawbacks: it was limited by equipment capacity to about 10 kg plutonium/day, and it produced copious quantities of high-salt waste. The capacity bottleneck can be removed by replacing the four plutonium purification cycle columns, but the salt-waste volume problem can be alleviated only by major changes in the flowsheet bases, as described in the next subsection. To illustrate the magnitude of the waste volume handicap in the flowsheet specified in the RL-SEP-595 report, the neutralized waste volume derived from a typical fuel containing 7.2 kg of fissile plutonium per ton and 1% U-235 enrichment after

burnup would be 2520 gallons/MT, or about 11 times the volume of neutralized TBP process waste.

The flowsheet had two other weak points: elimination of the head-end codecontamination cycle removed one cycle of plutonium decontamination, and the neptunium product from the 1N column was back-cycled to the 1A column waste concentrator for accumulation. These processing modes would almost certainly increase the ruthenium activity in the plutonium product (or require more extensive ruthenium volatilization, which is undesirable) and increase the neptunium loss.

b. New Flowsheet Proposal

A conceptual Hexone process flowsheet has been designed to overcome some of the above deficiencies in the RL-SEP-595 hexone flowsheet and is illustrated in Figures 10 through 15. The major benefit gained from this flowsheet is the reduction of the neutralized waste volume from 2520 gallons/MT to 660 gallons/MT using the same typical fuel described above. This also reduces the chemical cost per ton by about the same factor -- from about \$770 to \$220. The saving is gained primarily by use of much more extensive recycle and backcycle of aqueous raffinate from the extraction columns in the new flowsheet. The following are some of the major flowsheet features:

- The flowsheet reverts back to the current practice of having a head-end codecontamination cycle. This in effect provides four purification cycles for plutonium and three for uranium. The two codecontamination cycles and the final uranium cycle are operated acid-deficient for maximum ruthenium decontamination.

Bases:

U = 1000 kg = 4200 moles

Fissile Pu = 7.2 kg = 30 moles

Np = 0.35 kg = 1.5 moles

Flow given in liters per 1000 kgs uranium

Figure 10. Conceptual Hexone Process for Power Reactor Fuels Codecontamination Cycle

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Figure 12. Conceptual Hexone Process for Power Reactor Fuels Final Uranium Cycle

Figure 13. Conceptual Hexone Process for Power Reactor Fuels
Second Plutonium Cycle

Figure 14. Conceptual Hexone Process for Power Reactor Fuels Final Plutonium Cycle

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- From 55 to 75% of the aqueous raffinates from the 3A, 2D, 1A and HA columns are recycled to the respective feed streams. This represents a several-fold increase in the 1A and 2D recycle and is an innovation in the case of the 3A recycle. The ruthenium decontamination will be decreased by these changes, but perhaps not as much would be predicted from the increased ruthenium holdup in the system. (Ruthenium is present in several extractable and inextractable forms. Presumably the inextractable forms would greatly predominate in the recycled waste, and equilibrium between extractable and inextractable species is achieved slowly.)
- The remainder of the 3A column raffinate stream is backcycled to the 1B column where, after ferrous sulfamate addition, it is used as the IBX stream. This innovation displaces a major use of fresh aluminum nitrate solution and is the single most important step in reducing the overall waste volume. The 2A column raffinate is backcycled to the HA column as HAS according to standard practice when operations ceased.
- Product uranyl nitrate solution is used as the scrub stream in the 1A and 2D columns and aluminum nitrate is omitted completely from the 2D column. (Water could be used in place of the UNH in the 2D scrub with a minor change in the flowsheet design.)
- Diban (dibasic aluminum nitrate) is used extensively for acid neutralization and to achieve acid deficiency where desired. This does double duty in providing the necessary aluminum salting agent and minimizing the criticality hazard of accidentally precipitating

plutonium hydroxide. Sodium hydroxide is used in only two places-- to adjust the acid-deficiency in the plutonium-free 2D column feed and to neutralize the final high-level waste.

- The concentration of plutonium in plutonium cycle streams has been increased about 30% above the values specified in the RL-SEP-595 report by reducing the flow of the plutonium streams. The limit in the new flowsheet was set at 67% (75% is an allowable safe concentration) of the critical concentration of 7.8 g fissile plutonium/liter. This volume reduction results in a significant saving in salting agent, since the quantity of ANN required is proportional to the volume of raffinate and not to the concentration of plutonium present.
- Neptunium is recovered from the HA column raffinate by a flowsheet similar to that in current use. The major difference is that the neptunium product from the 1N column is concentrated and stored separately rather than being returned to the HA column raffinate concentrator for accumulation in the regular backcycle system. This minimizes the neptunium holdup in the system and, hence, the loss to the overall waste stream, which was an objection to the Hexone process in the ISO-800 study. The higher concentrations of Np in power fuels contributes to the feasibility of this procedure.

This flowsheet still has the drawback that the capacity of the plant is only about 10 kg of fissile Pu/day with the present columns, however, up to 50 kg of fissile Pu/day (6 MT U/day) can be processed by replacing the plutonium columns with 6 to 7 inch diameter columns.

These columns would be critically safe for accidental accumulation of up to 70 g Pu/liter or about ten times the flowsheet concentrations. If this should turn out not to provide adequate safety margin, 5-inch diameter columns, safe for any concentration, could be used with the flowsheet concentrations increased to gain the required capacity. Critically-safe geometries would also be required on auxiliary tanks and equipment if the latter alternative is chosen.

Other possible drawbacks pertain mostly to questions of criticality safety. For example, recycle of the 3AW to the IBX tank may conceivably introduce enough oxidant to destroy the ferrous ion reductant. This could lead to plutonium reflux in the non-geometrically safe extraction section of the IB column or to gross carryover of plutonium into the 2D-2E column system -- again possibly leading to plutonium accumulation by reflux. Addition of an in-cell IBX feed tank and administrative control of the reducing normality of the IBX stream may provide solutions to that problem. Addition of a new 3AW concentrator and high salting agent strength in the 3AF tank can also lead to exceeding the critical plutonium concentration. Use of a critically-safe geometry in the concentrator and replacement of the present 3AF tank with one of safe geometry should solve that problem. (Replacement of the 2AF tank should also be considered for nuclear safety.) The addition of soluble poisons, such as boric acid, could also be considered as a secondary safeguard. A detailed criticality safety analysis would be necessary to define all of the required process equipment changes; however, no insurmountable problems are foreseen.

Product decontamination may be a problem, as discussed earlier -- particularly if the ruthenium decontamination provided by the present product ozonators is limited by restrictions on the amount of ruthenium that can be volatilized. Purification of the neptunium product may also prove to be a particularly troublesome problem. The stored neptunium product shown in Figure 15 will contain about 30 to 50% of the ruthenium in the feed. Reprocessing accumulated neptunium through additional hexone extraction on a campaign basis results in ruthenium decontamination factors of only about 3 per cycle because of the necessity of using an acidic flowsheet. Anion exchange purification in a new facility, such as that proposed for the TBP process, will likely prove the best alternative to solvent extraction purification. One or more cycles of solvent extraction may be required, however, to prevent excessive radiation damage to the anion resin. (The incremental waste volume due to neptunium processing has not been determined, but it should be less than 10% of the previously quoted value.)

A final, and obvious, drawback to the new flowsheet is the complexity of the system and the difficulties that may be encountered on startup and shutdown in establishing and maintaining the recycle stream compositions. Since the flowsheet is mainly just an extension of the present flowsheet, this disadvantage should be relatively easy to overcome. An R&D expenditure of about \$150,000 is anticipated.

c. New Facilities Required

The discussion in the preceeding section indicated that several new facilities are needed to make optimum high-capacity use of the

existing plant. These include:

1. New 2A, 2B, 3A and 3B columns to increase the plant capacity from about 10 kg to 50 kg Pu/day.
2. A new IBX tank and feed system for safe reuse of 3AW as IBX..
3. A new 3AW waste concentrator (preferably of geometrically safe design) and routings from the concentrator to the 3AF tank and the new IBX tank.
4. A new neptunium product concentrator.
5. Optional additions for criticality control, including new geometrically-safe 2AF and 3AF feed tanks.
6. Routings and appropriate flow instrumentation for use of uranium product as 1AS and 2DS feed streams.
7. An optional ion exchange facility for neptunium purification (consideration should be given to "selling" the impure neptunium product to the Purex plant or other processor for final purification).

Several additional items were listed in the RL-SEP-595 study.

A major alteration specified in that study was the replacement of the 1A, 1B, 2D and 3D columns for critical safety. This change does not appear warranted unless the equivalent U-235 enrichment exceeds 2%. From data presented earlier, this appears unlikely except in unusual cases. Isotopic analysis of the dissolver solution will be used to assure safe enrichment and administrative (concentration) control can be employed for the rare case when high enrichments must be processed. Other additions suggested by this study were primarily back-up

instrumentation to assure safe concentration control in the uranium concentrators. Again, this may not be necessary for the actual expected enrichments.

d. Capital Cost of Required New Facilities

The cost of the chemical processing facilities required for converting the Redox plant for power-fuel processing using the present Hexone process was previously discussed in Section 5c. For an ultimate plant capacity of 6 MT/D our guesstimate for the cost of a new dissolver, new Pu columns, two new concentrators, instrumentation, etc., was \$0.9 million. Since we did not make detailed estimates and since some requirement may have been overlooked, we increased the estimate to \$1.0 million and added \$0.3 million for head-end facilities to provide capability for later expansion. The ISO-800 study provided an estimate of \$1.1 million for an Np anion exchange annex. Including the other basic facilities required, the total initial capital cost for a plant expandable to 6 MT/D using the Hexone process and including Np recovery would be:

Fuel Receiving Storage and Mechanical Treatment	\$ 5,900,000
Chemical Processing	1,000,000
Anion Exchange for Np	1,100,000
High-Level Waste	3,400,000
Low-Level Waste	<u>3,000,000</u>
TOTAL	\$14,400,000

Other incremental capital additions were also provided later as the load increased and these are detailed in Section 9.

8. BYPRODUCT RECOVERY -- COMPARISON OF PROCESSES

The ability to recover byproducts from the high-level waste will be a secondary but perhaps significant consideration in the choice of the solvent extraction process. Byproducts of interest, in approximate order of value, might include neptunium (recovery of Np, at least as a crude concentrate, has been assumed and discussed in previous sections), americium-curium, and the radioisotopes of promethium, cerium, strontium, and cesium. Technology for recovering and purifying these constituents from Purex waste exists and forms the basis of the B-Plant waste management program. The ISO-800 study, however, assumes that the B-Plant facility will not be available for processing the future Redox waste and that facilities for fractionating the TBP process waste into the various byproduct fractions could be incorporated in the Redox plant if the recovery option were chosen.

The complete ISO-800 Am-Cm-fission product recovery process is quite complex. It utilizes phosphotungstic acid precipitation to remove cesium and a multi-contactor solvent extraction process using D2EHPA as solvent to extract and partition strontium, Am-Cm, cerium, and Pm-rare earth product fractions. The necessary columns, batch contactors, etc., would be installed in H-Cell and would be almost completely isolated from the rest of the process. This feature is essential because of the incompatibility of D2EHPA with the TBP process. The capacity limit of the proposed process is about 2 tons U/day because of space limitations. The volume of stored waste would be increased by about 85% by chemical additions.

The incorporation of the above byproduct facilities into the Redox plant appears to have several disadvantages. First, there is the definite problem of keeping the two dissimilar solvents separated -- TBP does not interfere with D2EHPA processing but D2EHPA contamination in the TBP process would lead to high product losses and poor decontamination performance. In addition, complete removal of fission products for waste management purposes would limit the plant capacity to 2 tons/day. The process equipment is crowded with 5 pulse columns, 4 batch solvent contactors, and several feed and product tanks and concentrators proposed for space presently occupied by only 9 (or 10) equipment sites. Prolonged product storage is required for several of the fission products to meet isotopic ratio requirements in certain applications; this would require a separate shielded storage facility. The basic processes have received only limited testing thus far in "cold" and "hot" semiworks facilities. (Process reliability, however, will soon be demonstrated in the B-Plant facility for most of the proposed separations.)

The Hexone process is less adaptable to fission product recovery from concentrated waste than the TBP process because of the greater waste volume and the large amount of aluminum present. However, the hexone process waste is uniquely suited to the sulfate precipitation process developed in our laboratories for concentrating the lanthanides, Am-Cm, and strontium into a low volume nitric acid solution amenable to D2EHPA processing. The process is similar to the lead sulfate scavenging procedure used routinely on Purex wastes to provide

crude strontium and rare earth fractions for further purification in the Strontium Semiworks. The basic process requires sulfate addition to 0.75 M (minimum), hydrogen ion adjustment to $\text{pH } 1 \pm 0.5$, and digestion at 60 to 80°C for an hour. This treatment produces a crystalline, easily settled precipitate consisting of the sodium-rare earth double sulfates. The precipitate also carried Am-Cm and strontium almost quantitatively if the original waste solution contains over 0.01 M rare earths. (Lead carrier can be added if the rare earth concentration is below this.) Digestion of the precipitate with sodium carbonate converts the sulfate precipitate to insoluble carbonates, which can then be dissolved readily in dilute nitric acid after the sulfate supernate has been removed. The volume of the acidified product is limited primarily by the ability to handle the heat load.

Several properties of the waste produced by the new Hexone process seem particularly favorable for the sulfate precipitation process. First, the pH of the concentrated waste is nearly ideal and no expensive complexing agents are required to prevent aluminum and iron precipitation on neutralization. Second, the unneutralized waste volume of 400 gal MT provides an adequate rare earth concentration for strontium and Am-Cm scavenging. Third, none of the gross impurities (aluminum, iron, chromium) and few of the unwanted fission products form insoluble sulfates; and fourth, up to 20% of the sulfate should already be present from destruction of the sulfamate ion. Further adjustment of sulfate ion and pH can be obtained by adding

either sodium sulfate or sodium hydrogen sulfate. Application of the sulfate process to present Redox-type waste was recently demonstrated successfully by Bray and Roberts (BNWL-CC-907).

The volume of the salt waste derived from this process, after neutralization, will be about 1000 gallons/ton without further evaporation, and the only long-lived fission product present in large quantities will be Cs-137. This isotope can probably be removed, either before or after the sulfate precipitation step, by the same phosphotungstic acid process proposed for the TBP waste. (In fact, it should be possible to precipitate cesium phosphotungstate along with the sulfate precipitation.) If adequate fission product decontamination is achieved, it may prove feasible to solidify the high-salt waste after a few years of storage to allow for decay of the short-lived fission products, as planned with B-Plant waste.

Further fractionization of the Hexone process waste in the Redox plant does not appear to be feasible without a major capital addition due to lack of space. (A major factor in the ability to fit both the TBP and D2EHPA processes in the present Redox plant was the replacement of the complex hexone recovery and purification equipment with the simpler TBP solvent treatment system.)

9. ECONOMICS

a. ISO-800 Study

It was our conclusion that selection of the TBP process in the ISO-800 study was based largely on qualitative considerations. A great deal of weight was given to the compatibility of the TBP process with developed waste fractionization processes. One of the most serious objections to the Hexone process was the high waste volume.

b. Revenue Estimates

Although there are promising prospects for developing reprocessing plant revenues from byproduct recovery operations and especially for Np recovery for Pu-238 production, there are as yet no firm large-scale markets or commitments for these byproducts (Np, Am, Cm, Sr, Cs, Ce, and Pm). For this reason our view is that the economic feasibility of converting the Redox plant for power-reactor-fuel processing should be based on revenues derived from fuel processing charges alone. The possibility of obtaining revenues from the recovery of byproducts cannot be ignored but it is highly speculative and providing for byproduct recovery at this point in time appears a high risk proposition and should be considered separately from the main business of fuel reprocessing.

One advantage that a power-reactor-fuel reprocessor has over almost any other type of manufacturing or process industry is the

almost absolute certainty of the size of the market for at least six or seven years into the future. It requires four years (possibly as much as seven years at the present time) from the time of contract award to commercial operation of a power reactor. It takes another two to three years before the first fuel discharge is ready for reprocessing. There is of course no assurance as to what portion of this fuel reprocessing market any particular reprocessor will serve -- this is a matter of aggressive marketing and competitive pricing. It does provide the opportunity for a reprocessor to develop a strategy to capture a certain portion of the market and to implement that strategy with actual sales commitments covering a significant portion of his capital recovery period, well in advance of the time he must actually commit his capital. A reprocessing plant with the capability of incremental capital additions to expand its capacity would seem to be particularly desirable in this situation. Converting the Redox plant to power-reactor-fuel processing provides an attractive opportunity to capitalize on this unique feature of the fuel reprocessing market.

Referring to Figure 7, an entirely reasonable objective for any single reprocessor during the period from the mid-1970's through the 1980's would be to capture one-quarter of the total reprocessing market. A previous analysis⁽¹¹⁾ by A. J. Scott, Isochem Inc., of spent-fuel shipping costs for a reprocessing plant located at Hanford concluded that a Hanford plant would have roughly a \$2.00/kg shipping cost disadvantage in competing with midwest or eastern plants serving reactors in those regions. (Total shipping costs would be on the order of \$6.00/kg from eastern reactors.) As a basis for projecting reprocessing

revenues we have assumed that the Redox plant would process one-quarter of the total spent fuels available up to 1600 MT/yr (6 MT/D), taking first the west-coast fuels and making up the balance at a \$2.00/kg price disadvantage. The base reprocessing charges assumed were those shown in the Figure 9 projection and are considered a conservative basis for revenue estimates. The resulting projection of fuel load and revenue by year for a 15-year project starting up in 1973 is shown in Table II. 1973 was selected as a reasonable startup date since it would be the first year that revenue could exceed costs with the above assumptions.

c. Capacity Requirements

We have assumed the same relationship between instantaneous and actual capacity as the ISO-800 study. A 5-day week (3-shift) operation results in a 4-day processing week and an 85% overall efficiency results in 177 full capacity days per year. With a 7-day week (3-shift) operation, maintenance considerations reduce the overall efficiency to 72% (80% on-line and 90% rate efficiency) and results in 263 full capacity days/yr.

It has been common practice in reprocessing evaluations to assume that each batch of reactor fuel processed would be followed by a complete plant cleanout to assure accurate accountability of all valuable fissile and fertile materials. This practice was assumed in the original AEC conceptual plant study and is an important part of the basis for reprocessing charges at the NFS plant. However, when larger plants are operating it is unlikely that this

Table IICONVERTED REDOX ESTIMATED FUEL LOAD AND REVENUE

<u>Year</u>	<u>Redox Reprocessing Load, MT</u>			<u>Revenue, \$⁽¹⁾</u>		
	<u>Total Assumed as 1/4 Total US</u>	<u>Western Sector Portion</u>	<u>Outside Western Sector</u>	<u>Western Sector Load</u>	<u>Other</u>	<u>Total</u>
1970	20	3				
1971	41	18				
1972	84	30				
1973	175	75	100	2,325,000	2,900,000	5,225,000
1974	286	104	182	3,224,000	5,278,000	8,502,000
1975	416	140	276	4,200,000	7,728,000	11,928,000
1976	538	250	288	7,000,000	7,488,000	14,488,000
1977	655	375	280	9,750,000	6,720,000	16,470,000
1978	783	470	313	11,280,000	6,886,000	18,166,000
1979	925	530	395	11,660,000	7,900,000	19,560,000
1980	1063	630	433	12,600,000	7,794,000	20,394,000
1981	1238	760	478	13,680,000	7,648,000	21,328,000
1982	1400	910	490	14,560,000	6,860,000	21,420,000
1983	1600	1100	500	15,400,000	6,000,000	21,400,000
1984	1600	1230	370	14,760,000	3,700,000	18,460,000
1985	1600	1350	250	16,200,000	2,500,000	18,700,000
1986	1600	1480	120	17,760,000	1,200,000	18,960,000
1987	1600	1620	0	19,200,000	0	19,200,000

(1) Based on estimated reprocessing charges shown in Figure 9, except fuel load from outside western sector discounted \$2/kg to offset shipping cost disadvantage.

practice will be continued because of unfavorable economics. The concern with losses involves two items: (1) a direct material loss, and (2) a loss of separative work or blending loss that results when two separate uranium enrichments are mixed. Assuming a spent fuel containing 1% U-235 and 8 kg Pu/MT and a 40 MT fuel batch, the value of a 1% loss is \$51,000. Based on daily charges developed from the price projections in Section 3c, the following table shows what cleanout costs mean in terms of material losses.

<u>Plant Capacity,</u> <u>MT/D</u>	<u>3-day Cleanout</u> <u>Cost, \$</u>	<u>Loss Equivalent to</u> <u>Cleanout Cost, %</u>
1	82,000	1.6
2	143,000	2.8
4	184,000	3.6
6	213,000	4.2
10	258,000	5.0

A 1% actual total loss is a reasonable value so it is unreasonable to spend several times the equivalent value of this loss to insure its accuracy. With respect to separative work losses, the spent fuel enrichments are so similar that even a batch mix of 10% fuel A with 10% fuel B would generally result in a blending loss of less than \$4,000 -- not enough to warrant cleanout either. Thus, it should be possible to develop mutually satisfactory contractual agreements between the reactor operator and the reprocessor to eliminate the cleanout requirement. Adjustments in credits for recovered fuel material could be based on periodic cleanouts after several fuel batches, which would be required anyway for critical

safety reasons. We have therefore assumed that plant cleanouts between fuel processing campaigns are not necessary.

The capacity requirements for the converted Redox plant based on these considerations and the previous load projection are shown in Table III.

d. Economic Comparison of Purex-TBP vs Hexone Processes

To provide additional insight into the proper choice of processes for converting the Redox plant for commercial fuel processing, a cash flow economic comparison was developed for both the TBP and the Hexone process. The input cost data were based on extrapolations of the cost data developed in the ISO-800 study. The TBP process is the ISO-800 Purex process expanded to an ultimate 6 MT/D capacity and the Hexone process is the revised Redox process flowsheet discussed in Section 7b expanded to an ultimate capacity of 6 MT/D.

A major uncertainty in an analysis of commercial use for the Redox plant is the question of aquisition cost (purchase or lease). The original plant cost approximately \$40 million. Because of uncertainties with respect to government policies for disposition of such facilities, the analysis does not include an estimate or allowance for this important unknown and it is left to each reader to deduct his own estimate of aquisition costs from our calculated present worth values.

The cost estimates of initial capital requirements for both process alternates are shown in Table IV as well as the incremental capital cost additions required to increase the capacity as the load

Table IIICONVERTED REDOX ESTIMATED CAPACITY REQUIREMENTS

<u>Year</u>	<u>Assumed Processing Load MT/Yr</u>	<u>Minimum Requirements</u>			<u>Basis for this Analysis</u>		
		<u>Nominal Instan- taneous Capacity MT/D</u>	<u>Days per Week</u>	<u>Annual Capacity MT</u>	<u>Nominal Instan- taneous Capacity MT/D</u>	<u>Days per Week</u>	<u>Annual Capacity MT</u>
1973	175	1	5	177	2	5	354
1974	286	2	5	354	2	5	354
1975	416	2	7	524	2	7	524
1976	538	3	5	531	3	7	789
1977	655	3	7	789	3	7	789
1978	783	3	7	789	3	7	789
1979	925	4	7	1052	4	7	1052
1980	1063	4	7	1052	4	7	1052
1981	1238	5	7	1315	6	7	1578
1982	1400	6	7	1578	6	7	1578
1983	1600	6	7	1578	6	7	1578
1984	1600	6	7	1578	6	7	1578
1985	1600	6	7	1578	6	7	1578
1986	1600	6	7	1578	6	7	1578
1987	1600	6	7	1578	6	7	1578
1988	1600	6	7	1578	6	7	1578

Table IVREDOX CONVERSION CAPITAL COST ESTIMATES, \$1000's

<u>TBP (Purex) Process</u>				
<u>Capacity, MT/D</u>	<u>2</u>	<u>3</u>	<u>4</u>	<u>6</u>
Fuel receiving, storage and mechanical treatment	5 900 ⁽¹⁾		1 000	2 800
Chemical processing	5 600 ⁽²⁾	250 ⁽³⁾		250 ⁽³⁾
High-level waste (alk.)	2 500 ⁽⁴⁾			
Low-level waste	<u>4 500 ⁽⁶⁾</u>	<u>50</u>	<u>1 000</u>	
Cumulative Total	18 500	18 800	20 900	23 950
Np Anion Exchange Option	<u>1 100</u>			
Cumulative Total	19 600	19 900	22 000	25 050

<u>Hexone (Redox) Process</u>				
Fuel receiving, storage and mechanical treatment	5 900 ⁽¹⁾		1 000	2 800
Chemical processing	1 000	250 ⁽³⁾		250 ⁽³⁾
High-level waste (alk.)	3 400 ⁽⁵⁾			
Low-level waste	<u>3 000 ⁽⁷⁾</u>		<u>500</u>	
Cumulative total	13 300	13 550	15 060	18 100
Np Anion Exchange Option	<u>1 100</u>			
Cumulative Total	14 400	14 650	16 150	19 200

(1) Increased from ISO-800 estimate of \$5.6 million to provide facility expandable to 6 MT/D.

(2) Increased from ISO-800 estimate of \$5.1 million to provide facility expandable to 6 MT/D.

(3) New 2 MT/D dissolver.

(4) Two 300,000 gal. alkaline waste tanks plus transfer facilities.

(5) Two 600,000 gal. alkaline waste tanks plus transfer facilities.

(6) Increased from ISO-800 estimate of \$4.0 million to provide facility expandable to 6 MT/D.

(7) Reduced from ISO-800 estimate of \$4.0 million because of lower volume of low-level wastes (4-fold reduction in volume relative to TBP process).

requirements increased. Some of the modifications for the high capacity that were assumed in the initial cost might be deferred till later but probably would involve higher costs and might or might not be more economical. An allowance for working capital of \$1 million was provided in each case in addition to the facilities costs.

One reservation regarding the estimated costs for conversion to a TBP process should be mentioned. The very large volumes of process condensates produced in the TBP process may result in process bottlenecks (pumps, piping, etc.) at an expanded 6 MT/D capacity that won't show up without a detailed process design. This is not a concern with a Hexone process conversion.

The most significant difference in operating costs for the two processes relates to the cost of perpetual storage for high-level wastes. In addition to the initial two waste tanks and transfer facilities provided in each case, an assessment of \$3/gal of waste generated was made for construction of additional tanks as required and for perpetual storage as previously discussed in Section 5c. The calculated volumes of high-level waste generated by the two processes and annual costs are shown in Table V.

All other operating costs were assumed equal for the two processes. There would be differences but a detailed analysis of these did not appear justified for the purposes here. Essential material costs would vary but in terms of total yearly costs, probably no more than $\pm 5\%$. Total annual costs including high-level waste for the TBP process varied from \$3.8 million in the first year to \$6.5 million in

Table V

CONVERTED REDOX WASTE VOLUMES AND ESTIMATED
STORAGE COSTS INCLUDING PERPETUAL CARE

Year	Assumed Processing Load MT/Yr	220 Gal./MT			660 Gal./MT		
		Annual Gal.	Cumm. Gal.	Annual Cost at \$3/Gal.	Annual Gal.	Cumm. Gal.	Annual Cost at \$3/Gal.
1973	175	38 500	38 500	115 500	115 500	115 500	346 500
1974	286	62 920	101 420	188 760	188 760	304 260	566 280
1975	416	91 520	192 940	274 560	274 560	578 820	823 680
1976	538	118 360	311 300	355 080	355 080	933 900	1 065 240
1977	655	144 100	455 400	432 300	432 300	1 366 200	1 296 900
1978	783	172 260	627 660	516 780	516 780	1 882 980	1 550 340
1979	925	203 500	831 160	610 500	610 500	2 493 480	1 831 500
1980	1063	233 860	1 065 020	701 580	701 580	3 195 060	2 104 740
1981	1238	277 360	1 337 380	817 080	817 080	4 012 140	2 451 240
1982	1400	308 000	1 645 380	924 000	924 000	4 936 140	2 772 000
1983	1600	352 000	1 997 380	1 056 000	1 056 000	5 992 140	3 168 000
1984	1600	352 000	2 349 380	1 056 000	1 056 000	7 048 170	3 168 000
1985	1600	352 000	2 701 380	1 056 000	1 056 000	8 104 140	3 168 000
1986	1600	352 000	3 053 380	1 056 000	1 056 000	9 160 140	3 168 000
1987	1600	352 000	3 405 380	1 056 000	1 056 000	10 216 140	3 168 000
1988	1600	352 000	3 757 380	1 056 000	1 056 000	11 272 140	3 168 000

the 15th year. Total annual costs including high-level waste for the Hexone process varied from \$4 million in the first year to \$8.5 million in the 15th year. In both cases these total annual costs include an allowance of up to \$1.1 million/yr for marketing, research and engineering, and general corporate expenses.

A 10-year sum-of-the-years-digits depreciation for tax purposes was assumed with federal income tax rate of 50%.

The resulting cash flow analyses based on the conservative revenue estimates from Table II are summarized in Tables VI and VII. Discounted net cash flows at both 10% and 15%/yr interest rate of return after taxes were calculated. These are intended to represent a range of return that would be attractive for a venture of this type.

As previously pointed out, an allowance for Redox plant acquisition cost is not included in this analysis and should be factored into the reader's evaluation of these results. However, it is important to recognize that because the depreciation allowance for an additional capital expenditure reduces future income tax costs, additional expenditure for acquisition of the Redox facility reduces the net present worth of a 10 or 15 year project by something less than the initial capital expenditure. For example, each million dollars of additional capital expenditure reduces the net present worth for a 10-year (or longer) operation by only approximately \$680,000 with a 10% interest rate of return or \$750,000 with a 15% interest rate of return.

Table VICASH FLOW SUMMARY⁽¹⁾ - TBP PROCESS

<u>Year</u>	<u>Net Cash Flow After Taxes, \$1000's</u>	<u>Discounted⁽²⁾ Net Cash Flow at 10% \$1000's</u>	<u>Net Present Worth at 10%, \$1000's</u>	<u>Discounted⁽²⁾ Net Cash Flow at 15% \$1000's</u>	<u>Net Present Worth at 15%, \$1000's</u>
1969	- 860	-1 200		-1 410	
1970	-2 250	-2 860		-3 180	
1971	-8 190	-9 450		-10 100	
1972	-8 810	-9 240	-22 750	- 9 450	-24 140
1973	1 440	1 380		1 350	
1974	3 760	3 270		3 050	
1975	4 600	3 630		3 260	
1976	5 970	4 280		3 670	
1977	6 690	4 360	-5 830	3 570	-9 240
1978	5 130	3 040		2 380	
1979	7 890	4 250	1 460	3 180	
1980	4 980	2 430		1 750	
1981	8 540	3 800		2 610	680
1982	8 250	3 340	11 030	2 190	2 870
1983	7 870	2 900		1 820	
1984	6 310	2 110		1 270	
1985	6 340	1 930		1 110	
1986	6 390	1 760		970	
1987	6 430	1 620	21 350	850	8 890

(1) No provision for Redox acquisition is included.

(2) Uniform cash flow convention discounted from startup on January 1, 1973.

Table VII

CASH FLOW SUMMARY⁽¹⁾ - HEXONE PROCESS

<u>Year</u>	<u>Net Cash Flow After Taxes, \$1000's</u>	<u>Discounted⁽²⁾ Net Cash Flow at 10% \$1000's</u>	<u>Net Present Worth at 10%, \$1000's</u>	<u>Discounted⁽²⁾ Net Cash Flow at 15%, \$1000's</u>	<u>Net Present Worth at 15%, \$1000's</u>
1969	- 860	-1 200		-1 410	
1970	-2 250	-2 850		-3 180	
1971	-5 440	-6 280		-6 710	
1972	-6 360	-6 670	-17 000	-6 820	-18 120
1973	850	810		790	
1974	3 150	2 740		2 560	
1975	4 030	3 180		2 850	
1976	5 310	3 810		3 260	
1977	6 010	3 910	-2 550	3 200	-5 460
1978	5 010	2 970	420	2 320	
1979	7 060	3 800		2 850	
1980	4 010	1 960		1 410	1 120
1981	7 610	3 390		2 320	
1982	7 280	2 940	12 510	1 930	5 370
1983	6 810	2 500		1 570	
1984	5 260	1 760		1 060	
1985	5 290	1 610		920	
1986	5 350	1 470		810	
1987	5 400	1 360	21 210	710	10 440

Notes: (1) No provision for Redox acquisition is included.

(2) Uniform cash flow convention discounted from startup on January 1, 1973.

Key results from the cash flow analysis are as follows:

	6 MT/D TBP Process, \$ millions	6 MT/D Hexone Process, \$ millions
Initial Capital for Plant Modifications	18.5	13.3
Incremental Capital Additions	5.5	4.8
Project Costs	0.5	0.5
Working Capital	<u>1.0</u>	<u>1.0</u>
Total Capital Investment	25.5 plus aquisi- tion cost	19.6 plus aquisi- tion cost

Net present worth at
startup date (1973)
with:

a. 10% interest rate of
return after taxes:

10 yr operation	11.0	less aquisi-	12.5	less aquisi-
15 yr operation	21.4	tion cost	21.2	tion cost

b. 15% interest rate of
return after taxes:

10 yr operation	2.9	less aquisi-	5.4	less aquisi-
15 yr operation	8.9	tion cost	10.4	tion cost

Both cases provide for recovery of a crude undecontaminated Np product but no credit for sale of this service. Neptunium anion exchange could be added in each case for an additional \$1.1 million.

The results indicate a close competitive situation between the two process concepts except that the initial investment requirements for the Hexone process were \$5.2 million less than for the TBP process. This would favor a decision for the Hexone process. However, considering the assumptions involved, a more detailed study including more

detailed cost estimates of facility and operating costs should be undertaken before a hard choice is made between the two process concepts.

One inherent advantage for the Hexone process is the fact that this is the present process and fewer unforeseen design and startup problems are likely to be encountered.

The high waste volumes associated with the Hexone process are not an insurmountable obstacle and can be offset by the lower initial capital requirements. If perpetual storage costs are less than the \$3/gal assumed here, the Hexone process economics would be improved even more.

Regardless of which process is selected, the Redox plant offers an attractive opportunity for a reprocessing venture during the 1970's and early 1980's.

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