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Operating and Life-Cycle Costs for Uranium-Contaminated Soil Treatment Technologies

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FOR THE UNITED STATES
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1. BACKGROUND

The development of a nuclear industry in the United States required mining, milling, and fabricating a large variety of uranium products. One of these products was purified uranium metal which was used in the Savannah River and Hanford Site reactors. Most of this feed material was produced at the United States Department of Energy (DOE) facility formerly called the Feed Materials Production Center at Fernald, Ohio.

Currently, this facility is called the Fernald Environmental Management Project (FEMP) and is operated by the Fernald Environmental Restoration Management Corporation (FERMCO). The facility consists of 1,050 acres in a rural area that is 18 miles northwest of Cincinnati, Ohio. The manufacturing processes were housed in a 136-acre fenced area and included uranium and thorium metal production and uranium hexafluoride reduction. Production peaked in 1960 with approximately 10,000 tons of uranium processed and began to decline in 1964 to a low of 1,230 tons in 1975. In the mid-1980s, production increased slightly but was terminated in 1989 due to the lack of demand for uranium products.

During operation of this facility, soils became contaminated with uranium from a variety of sources. The sources included deposition of airborne uranium particulates that came from facility stacks as well as leaks and spills of uranium-containing solvents and process effluents generated during nonaqueous extraction/treatment processes. The exact quantity of soil contaminated with uranium is unknown. Some estimates of soil containing unacceptable levels of uranium are as high as 2 million cubic yards. To avoid disposal of these soils in low-level radioactive waste burial sites, increasing emphasis has been placed on the remediating soils contaminated with uranium and other radionuclides.

To address remediation and management of uranium-contaminated soils at sites owned by DOE, the DOE Office of Technology Development (OTD) evaluates and compares the versatility, efficiency, and economics of various technologies that may be combined into systems designed to characterize and remediate uranium-contaminated soils. Each technology must be able to 1) characterize the uranium in soil, 2) decontaminate or remove uranium from soil, 3) treat or dispose of resulting waste streams, 4) meet necessary state and federal regulations, and 5) meet performance assessment objectives. The role of the performance assessment objectives is to provide the information necessary to conduct evaluations of the technologies. These performance assessments provide the basis for selecting the optimum system for remediation of large areas contaminated with uranium. One of the performance assessment tasks is to address the economics of full-scale implementation of soil treatment technologies. The cost of treating contaminated soil is one of the criteria used in the decision-making process for selecting remedial alternatives.

2. INTRODUCTION

During the past 2 years, various studies have been directed throughout the DOE national laboratories, universities, and private industry to determine the best methods to remove uranium from uranium-contaminated soils. The majority of these studies have been conducted in the laboratory at the bench-scale level, and results from these bench-scale studies have been extensively presented and published (Soil Decontamination Task Group 1993, Post and Wacks 1994). The original study focused on the following 7 soil treatment technologies: sodium carbonate soil washing system, sulfuric acid soil washing system, citric acid soil washing system, tiron soil washing system, heap leaching system, fungal leaching process, and aqueous biphasic extraction.

During the summer of 1994, preliminary fixed capital investment (FCI) requirements were estimated for the 7 technologies based on bench-scale studies (ORNL/TM-13004). Due to the stage of research, data used to develop the engineering flow diagrams and FCI requirements were preliminary and subject to revision after more bench-scale and pilot-scale studies had been completed. Upon completion of these studies and review of the 7 technologies, OTD determined that the following 4 treatment technologies would be the best candidates for potential full-scale implementation: carbonate/bicarbonate vat extraction process (formally referred to as sodium carbonate soil washing system in the ORNL/TM-13004 report), heap leaching system, tiron soil washing system, and the aqueous biphasic extraction process. In the fall of 1994, the original engineering diagrams for these 4 treatment technologies designed by Halliburton NUS were revised and optimized by personnel at Brown & Root, Inc. Based on discussions with the principal investigators of the treatment technologies, the processes were simplified, resulting in new chemical process equipment requirements and chemical consumptions for a potential full-scale treatment facility. In addition, OTD determined that the High Gradient Magnetic Separation (HGMS) technology should be studied for potential full-scale implementation. Subsequently, an engineering flow diagram and equipment list were developed by personnel at Brown & Root, Inc.

The cost estimates for the treatment technology options were conducted in three steps. The first step is to estimate the fixed capital investment (FCI), which represents the initial capital expenditure required to design and construct the full-scale treatment facility to operational readiness. The FCI is a one-time project investment cost that occurs at the beginning of the project. A technical memorandum entitled, *Fixed Capital Investments for the Uranium Soils Integrated Demonstration Soil Treatment Technologies* (ORNL/TM-13004) documented the fixed capital investment (FCI) requirements for the 7 treatment technologies originally considered. The report contained a description of the method used to calculate the FCI estimates, a cost estimate of the equipment required for full-scale implementation of each treatment technology, as well as appendices containing the engineering flow diagrams and equipment lists developed by Halliburton NUS for each treatment technology. The second step is to develop estimates of the operational costs for each of the full-scale facility designs. Operational costs include costs for raw materials involved in processing, operating labor, supervisory labor, utilities, plant maintenance and repairs, and miscellaneous costs such as operating supplies, taxes, and insurance. Unlike FCI costs, operating costs are recurring throughout the life of the project. The third step is to estimate the life-cycle costs for each treatment technology option. The life-cycle costing approach involves projecting the current and future cash flows for soil remediation by each treatment technology over the project life, based on the estimated quantity of contaminated soil at the Fernald site. Non-recurring costs occurring in the first year include the FCI and start-up expenses. Thereafter, operational costs for a full-scale facility based on each treatment technology are projected into future cash flows using an assumed inflation rate. Total costs to build and operate the full-scale treatment facility were then summed over the life of the project and divided by the amount of soil to be processed to determine which treatment technology is the most cost effective based on their respective values for the treatment cost per ton of soil treated.

The objective of this document is to describe the methods and results of the cost estimates for the final 5 soil treatment technologies based on full-scale implementation. These cost estimates are based on the "best engineering design" for each treatment technology determined by Brown & Root, Inc. personnel. As previously stated, Brown & Root, Inc. has optimized the engineering designs to incorporate current knowledge of each treatment technology. A list of the chemical process equipment required for full-scale implementation of each treatment technology, as well as a cost estimate for this equipment, is provided in this report. In addition, cost estimates of the FCI requirements, operational costs, and life-cycle costs for a full-scale treatment facility based on each treatment technology are documented in this report. The FCI, startup expenses, operational, and life-cycle cost estimates are then utilized to determine the treatment cost per ton of soil treated, based on the estimated quantity of contaminated soil to be processed at the Fernald site.

The cost estimates in this report are defined as study estimates. A study estimate is based on the knowledge of major items of equipment, with an accuracy of over plus or minus 30%. A preliminary estimate is based on sufficient data to permit the estimate to be budgeted, with an accuracy of within plus or minus 20% (Peters and Timmerhaus 1991). Because specifications and design requirements for the equipment, as well as the process, were preliminary, there is a large probability that the actual cost will be more than the estimated cost where information is incomplete or during inflationary periods. For such estimates, the positive spread is likely to be wider than the negative (Peters and Timmerhaus 1991). Therefore, the plus or minus 30% accuracy rate for our study estimate may in fact be +40% to -20%.

Appendix A contains the engineering flow diagrams of the soil treatment technologies which were developed by personnel at Brown & Root, Inc. Appendix B provides a detailed cost breakdown of the chemical processing equipment required for full-scale implementation of each soil treatment technology alternative. A summary report was prepared by Brown & Root, Inc. personnel and submitted to principal investigators of the treatment technologies, as well as other interested parties involved in the USID project. The report contains a detailed process description of each treatment technology, along with process design assumptions, engineering flow diagram, process equipment list, and a material mass balance sheet associated with each technology. With approval from the staff at Brown & Root, Inc., a copy of the report has been included in this document as Appendix C.

3. FIXED CAPITAL INVESTMENT ASSUMPTIONS

The FCI is the capital needed to supply the necessary manufacturing and plant facilities for an industrial operation or project. The FCI represents the initial capital expenditure required to construct a full-scale treatment facility based on each of the proposed treatment technologies being demonstrated. A common method for estimating the FCI is to define it as a function of the purchased equipment costs with each component of the FCI estimated as a percentage of the equipment cost. The method of estimating the FCI by percentage of purchased equipment cost is commonly used for study and preliminary cost estimates. Table 1 presents a checklist of the items used to estimate the FCI for chemical processing plants, along with the range for each item as a percentage of the purchased equipment cost. The cost range for the FCI items shown in Table 1 are based on input from several studies developed by chemical processing cost estimators plus additional data and interpretations from other sources with experience in modern industrial design and construction (Peters and Timmerhaus 1991). The FCI is the sum of the direct and indirect plant costs, contractor's fee, and contingency.

The cost estimator was provided with an equipment parts list for each treatment technology from Brown & Root, Inc. Once an estimate of the total equipment cost was determined for each treatment technology, percentages of this value were used to calculate the FCI, with the exception of building costs. The percentages of total equipment costs used for FCI calculations were based on many factors,

Table 1. Fixed capital investment template

Item	Symbol	Description	Range
1	E	Equipment Costs	E
2	L	Cost of Installation Labor	.25E - .55E
3	IC	Instrumentation & Controls	.06E - .30E
4	I	Insulation Costs (equip. & piping)	.05E - .09E
5	P	Piping	.16E - .31E
6	Q	Labor for Installation of Piping	.40P - .50P
7	F	Electrical Installations	.10E - .15E
8	B	Building including Services	.05E - .68E or unit costs, if available
9	Y	Yard Improvements	.10E - .20E
10	S	Service Facilities	.30E - .80E
	D	**Direct Plant Cost**	Sum 1 - 10
11	ES	Engineering & Supervision	.15E - .80E
12	C	Construction Expenses	.15E - .60E
	IP	**Indirect Plant Cost**	**ES + C**
13	CF	Contractor's Fee	.02(D+IP) - .08(D+IP)
14	CO	Contingency	.05(D+IP) - .20(D+IP)
	FCI	**Fixed Capital Investment**	**D+IP+CF+CO**

including the type of process involved, design complexity, required materials of construction, as well as discussions with Brown & Root, Inc. representatives experienced in estimating costs for chemical processing equipment. Building costs were determined based on unit costs and estimates of the square footage of building space required for a full-scale treatment facility for each treatment technology alternative. Along with surveying and associated closing costs, land costs are normally included in the FCI. However, for this report they were reported as zero for the treatment technology cost estimates, because the facility was assumed to be located on already purchased land at the Fernald site.

Costs for several pieces of processing equipment were estimated from documents published in previous years. In these cases, cost indexes were used to project these values to present-day costs. A cost index is an index value for a given point in time showing the cost at that time relative to a certain base time. If the cost at some time in the past is known, the equivalent cost at the present time can be determined by multiplying the original cost by the ratio of the present index value to the index value applicable when the original cost was obtained. Many different types of cost indexes exist for estimating items such as processing equipment, labor, construction, materials, and other specialized fields (Peters and Timmerhaus 1991). The Marshall and Swift All-industry and Process-industry Equipment Index was used when costs were obtained from one of the sources containing equipment costs from the past.

The purchased equipment costs (E, referring to Table 1) were obtained from one of the following sources: 1) vendors specializing in the type of chemical processing equipment required for each soil treatment technology, 2) cost estimating personnel at Brown & Root, Inc., 3) Richardson's Engineering Services *Process Plant Construction Estimating Standards*, 1994 edition, 4) *Mining and Mineral Processing Equipment Costs and Preliminary Capital Cost Estimations*, Volume 25, 1982, 5) *Means Site Work and Landscape Cost Data*, 12th annual edition, 1993, and 6) *Plant Design and Economics for Chemical Engineers*, Max S. Peters and Klaus D. Timmerhaus, 4th edition, 1991. The most accurate method for determining process equipment costs is to obtain firm bids from equipment fabricators or suppliers. Verbal and/or written quotes for the specialized equipment pieces were obtained in as many cases as possible. However, in some cases vendors were unable to provide price quotations because certain design requirements that could significantly affect costs were not available. In cases such as this, as well as for the more common types of processing equipment, costs were estimated from one or more of the above listed reference sources.

The installation of equipment (L) involves costs for labor, foundations, supports, platforms, and other construction expenses related to the erection of purchased equipment. Although different ranges of installation costs exist depending on the type of chemical processing equipment, most fall within the .25E to .55E range specified in Table 1. Instrumentation and controls (IC) costs can vary from 6% to 30% of the purchased equipment costs. For the soil treatment technology alternatives, IC costs were estimated at .13E because this value is normally used for solid-fluid chemical processing plants. The major portion of this category consists of instrument costs, installation labor costs, and expenses for auxiliary equipment and materials.

The estimated cost for piping (P), with a range of 16% to 31% of equipment costs, varies depending on the type of chemical processing plant under consideration (i.e., solid processing, solid-fluid processing, or fluid processing plant). This cost category typically includes valves, fittings, pipe, supports, and other items involved in the complete erection of all piping used directly in the treatment technology process. This includes the piping used for air, steam, water, and other process piping requirements, as well as for the equipment used to treat the contaminated soil. The labor for the installation of piping (Q) ranges from 40% to 50% of the total cost of piping. Insulation costs (I) for equipment and piping normally range from 5% to 9% of purchased equipment costs. Insulation costs vary depending on many factors, including the number of pieces of equipment exposed to very low or high temperatures (i.e., how much of the equipment is enclosed in the building and how much is located outside the building?) and the amount of piping that is required in each full-scale treatment facility design. Electrical installation costs (F) range from 10% to 15% of equipment costs. This cost category consists primarily of installation labor and materials for power wiring, lighting, transformation and service, and instrument and control wiring (Peters and Timmerhaus 1991).

The cost for buildings including services (B) consists of expenses for labor, materials, and supplies involved in the construction of all buildings associated with the treatment facility. The costs for plumbing, heating, lighting, ventilation, and similar building services are also included in this category. The range for this cost varies from 5% to 68% of the purchased equipment cost and is dependent on two factors: 1) the type of process plant (solid, solid-fluid, or fluid processing plant) and 2) whether or not the facility being considered is a new plant at a new site or a plant expansion at an existing site. For solid-fluid processing plants, building costs can range from 29% to 47% of purchased-equipment costs (Peters and Timmerhaus 1991). However, in this report the square footage required for the treatment facility building for each treatment technology was estimated by Brown & Root, Inc. personnel from the engineering flow diagrams, and costs were estimated from unit costs based on these building size requirements. This provides a more accurate estimate of the building cost rather than estimating the building costs as a percentage of the purchased equipment cost.

Yard improvements (Y) for the facility include the costs for site clearing and grading, roads and walkways, fencing, lighting, parking areas, landscaping, and other similar improvements. These costs range from approximately 10% to 20% of the purchased equipment cost. The total cost for service facilities (S) ranges from 30% to 80% of equipment costs. These costs include the utilities for supplying and distributing steam, water (i.e., treatment and distribution), power (i.e., electric substation and distribution), compressed air, and fuel to the soil treatment facility. In addition, waste disposal, fire protection, and miscellaneous service items such as communications, first aid, and safety installations require capital investments which fall under the category of service facility costs (Peters and Timmerhaus 1991). Service facilities costs are largely a function of plant physical size and will be present to some degree in most plants. However, there are many service facility cost categories, and for most solid-fluid chemical processing plants, there will not always be a need for each service-facility component. It is anticipated that the water and electricity requirements for each of the full-scale treatment facility options will be fairly high. Therefore, a middle to upper range value was used to estimate service facilities costs for the treatment technologies. The sum of the following items make up the direct plant cost (D) for the facility: 1) equipment, 2) equipment installation labor, 3) instrumentation and controls, 4) equipment and piping insulation, 5) piping, 6) piping installation labor, 7) electrical installations, 8) building including services, 9) yard improvements, and 10) service facilities.

Engineering and supervision costs (ES), ranging from 15% to 80% of equipment costs, are indirect plant costs (IP), because they cannot be directly charged to equipment, materials, or labor. Costs for construction design and engineering, drafting, purchasing, accounting, cost engineering, travel, reproductions, and overhead constitute the capital investment for engineering and supervision (Peters and Timmerhaus 1991). A value close to the upper limit of the cost range was used for all of the treatment technologies, because engineering costs will undoubtedly be high due to the preliminary stage of development for each of the treatment technologies. Another indirect plant cost used in calculating the FCI for each treatment technology option is construction expense (C), ranging from 15% to 60% of equipment costs. This cost item includes temporary construction and operation, construction tools and rentals, construction payroll, insurance, and other construction overhead items. The sum of the engineering and supervision and construction expense items comprise the indirect plant costs for the treatment facility.

A contractor's fee (CF) and contingency (CO) are normally added to the direct and indirect plant costs in calculating the FCI. The CF ranges from 2% to 8% of the sum of the direct and indirect plant costs. Contingency is a project markup factor normally applied to cost estimates to account for any uncertainties or unforeseen occurrences, such as inflationary price trends, bad weather conditions, strikes, small design changes, estimation errors, or possible material shortages associated with a project. Contingency normally ranges from 5% to 20% of the sum of direct and indirect plant costs (Peters and Timmerhaus 1991). Because of the experimental and developmental nature of these treatment technologies, it was determined that the most conservative value (20% of the sum of direct and indirect plant costs) should be used to estimate the contingency for each treatment option. The FCI for each treatment technology is then calculated by summing the direct and indirect plant costs, contractor's fee, and contingency.

4. OPERATING COST ASSUMPTIONS

Based on the treatment technology engineering designs, the daily operating costs are a significant contributor to the life-cycle costs associated with a full-scale treatment facility. The following items were included in calculating the operating costs: 1) chemicals and other raw materials involved in the treatment process, 2) operating labor for the soil receiving building (if applicable) and the soil treatment building , 3) maintenance and repair costs for the process equipment and the buildings associated with each technology, 4) utilities associated with the operation of the treatment facility, including electricity, steam, process and cooling water, natural gas, fuel oil, etc., 5) operating supplies, 6) fixed charges, including taxes and insurance for the facility, and 7) a contingency factor. Contingency is a project markup factor normally applied to cost estimates to account for any uncertainties or unforeseen occurrences, such as inflationary price trends, bad weather conditions, strikes, design changes, estimation errors, or possible material shortages associated with a project. Contingency was estimated at 25% of the total operating costs associated with each treatment technology. Even though the contingency factor is somewhat conservative, it is justified in that the operating cost parameters for the technologies are highly variable and uncertain. For instance, the cost of tiron is very uncertain because it is presently not available in bulk quantities, a requirement based on the engineering design of the tiron soil washing system. One source quotes a price of $40.65 for 100 grams of tiron, or approximately $184 per pound. However, in preliminary phone conversations with a company representative, a chemical company claims they could produce tiron in bulk quantities for $7 to $8 per pound. However, this quote was not guaranteed in writing, leaving the actual cost of tiron still highly uncertain.

The majority of the chemical costs were obtained from the "*Chemical Marketing Reporter*", a magazine listing the latest chemical price ranges from suppliers. Steam, electricity, and make-up treated water costs were gathered from *Plant Design and Economics for Chemical Engineers* (Peters and Timmerhaus 1991). Although these costs were reported in 1989 dollars, the Marshall and Swift All-industry and Process-industry Equipment Index was used to adjust the costs to today's dollars. A cost of $3.98 per 1,000 lbs, $0.08 per kilowatt hour, and $0.89 per 1,000 gallons were used to estimate the operating costs of steam, utilities, and make-up water, respectively. The estimated cost of carbon dioxide gas was based on local vendor quotes. The utility costs were estimated by summing the horsepower requirements of individual pieces of chemical processing equipment required for each treatment technology and multiplying this value by $0.08 per kilowatt hour. In addition, a 50% factor for heating and lighting and a 10% factor for line losses and contingencies were applied to each treatment technologies' utility cost estimate.

Other costs included in the daily operating costs associated with a full-scale treatment facility are equipment and building maintenance and repair costs, operating supplies, and fixed charges. Annual equipment maintenance and repair costs normally range from 2% to 20% of the equipment cost, depending on the equipments' operating demand (Peters and Timmerhaus 1991). Building maintenance and repairs costs average 3% to 4% of the building cost. A factor of 15% of the equipment cost and 4% of the building cost were used to estimate the annual maintenance costs for the equipment and treatment building, respectively. To obtain a daily rate for operating costs, it was assumed that the facility would operate 350 days per year. Miscellaneous operating supplies, such as lubricants, chemicals, and custodial supplies, are needed to keep the soil treatment technology processes operating efficiently. The estimated annual cost for these types of supplies is approximately 15% of the total cost for maintenance and repairs. Fixed charges are expenses that occur regardless of whether or not the process is in operation. Taxes and insurance were two fixed charge items that were accounted for in the operating cost estimates. Annual property taxes for plants range from 1% to 4% of the FCI, depending on the plant site location population. A factor of 2% of the FCI was used for our cost study. Although insurance rates depend on the type of process being carried out at the facility, the annual rate for coverage is normally approximately 1% of the FCI (Peters and Timmerhaus 1991). As with the maintenance and repairs costs, the annual rate for the taxes and insurance was divided by 350 to obtain a daily rate, based on the assumption that the facility would be operating for 350 days per year.

The labor rates used in the cost study were obtained from the *Richardson Labor Cost Index* in *The Richardson Construction Cost Trend Reporter*, published by Richardson Engineering Services, Inc. The index lists hourly crew rates for a number of work crafts across 127 cities in the United States. The "process equipment" crew category was used to represent the general employees and technicians operating the equipment for each treatment technology. The January 1995 labor cost index for Cincinnati, Ohio (the city nearest to the Fernald site) lists the direct labor hourly rate as $24.36 per hour. Labor burden refers to costs a company must pay above the base labor rate, such as for pensions, Social Security, insurance, vacations, and other benefits. Based on discussions with cost estimation personnel at Brown & Root, Inc., a labor burden factor of 40% was used for the labor cost estimates. Therefore, the hourly rate (including labor burden) used in the cost estimates for general employees and technicians operating the process equipment is $34.10 (24.36 x 1.40). The rate for the truck drivers transporting the contaminated soil to the soil receiving building was also obtained from the *Richardson Labor Cost Index*, at $23.17 per hour, including burden ($16.55 x 1.40).

One source quoted a supervisor's hourly rate as 33% higher than an equipment operator's rate, so a rate of $45.35 per hour ($34.10 x 1.33) was used for the shift supervisors working in the soil receiving and soil treatment buildings (Peters and Timmerhaus 1991). Also, radiation monitoring personnel rates were typically 20% higher than an operator's rate; therefore, a rate of $40.92 per hour was used to estimate labor costs for radiation monitoring personnel. Three types of maintenance personnel were accounted for in the operating labor costs: 1) a general maintenance employee, 2) an electrician, and 3) an instrument technician. General maintenance employee's labor rates are approximately 80% of an operator's rate, so an hourly rate of $27.28 ($34.10 x 0.80) was used for this cost category. Electricians' and instrument technicians' labor rates in the *Richardson Labor Cost Index* are slightly lower than a process equipment operators' rate, at $23.71 per hour. Adding a labor burden factor of 40%, the hourly rate used to estimate the labor costs for an electrician and instrument technician is $33.19 ($23.71 x 1.40). It should be noted that the cost figures quoted in this section apply to personnel on the first shift. A shift differential of 25% was applied to the labor rates for the second and third shift employees of the soil receiving and soil treatment buildings.

5. LIFE-CYCLE COST ASSUMPTIONS

Although the exact quantity of contaminated soil at the FERMCO site is unknown, an assumption of 2 million cubic yards was used for the life-cycle cost estimates associated with a full-scale treatment facility based on each treatment technology. The estimated quantity of soil is important in calculating the life-cycle costs because, along with the soil treatment rate (tons/hr), these parameters determine the length of time that the treatment facility must stay in operation. The engineering flow diagrams for each treatment technology are designed based on the assumption of a 20 tons/hour soil treatment rate. Although this treatment rate may not be the most optimum and efficient for certain treatment technologies, a common design assumption was made so that each technology could be evaluated on an equal basis. The utilization factor for a chemical plant or facility is also important in determining the length of time required to treat the estimated quantity of contaminated soil, and thus the life-cycle costs associated with a treatment facility at the Fernald site. A 70% utilization factor was used for this cost study. Therefore, the soil decontamination operation would be operating 70% of the time, with maintenance and repairs and other associated shutdown activities taking place the remaining 30% of the time. Another assumption is that the facility treats soil 24 hours per day, using 3 shifts, for 350 days per year. Under these assumptions, a full-scale facility can treat 336 tons per day (20 tons/hr x 24 hrs/day x 0.70), or 117,600 tons/yr (336 tons/day x 350 days/yr). The soil density is another important parameter because it is used to determine the weight of soil in a cubic yard of contaminated soil. Based on Fernald site characterization results, it was determined that a value of 1.2 g/cu. cm should be used for the feed stock soil density. Based on this density, 1 cubic yard of Fernald soil weighs approximately 1 ton. Therefore, it is estimated that it would take approximately 17 years to treat the estimated 2 million cubic yards (tons) of contaminated soil (2,000,000 tons/117,600 tons/yr).

For the life-cycle cost calculations, in year 0 the costs incurred will be the FCI and the startup expenses required for each technology. Years 1 through 17 consist of operating expenses. An inflation rate of 5% per year was factored into the operating cost calculations. Like the FCI, startup expenses represent a one-time expenditure in the first year of the plant operation. After plant construction has been completed, frequently there are changes that have to be made before a facility can operate at maximum design conditions. These changes involve expenditures for materials and equipment and result in the loss of income while the plant is shut down or is operating at only partical capacity. Although the startup expenses can be as high as 12% of the FCI, it normally averages 8% to 10% of the FCI (Peters and Timmerhaus 1991). A startup expense of 10% was assumed for this cost study.

6. SOIL RECEIVING BUILDING

A soil receiving building in which soil is stored and eventually fed into the soil treatment building is required for the ABE, carbonate/bicarbonate vat extraction, and tiron soil washing processes. Assuming that the building can contain 5 days of storage inventory (approximately 1,700 tons) in 2 parallel storage cribs, the size of the building is estimated at 17,600 square feet (80 ft W x 220 ft L). The 2 storage cribs, one for fill material and the other for reclaim for transfer to the treatment building, will be approximately 120 ft L x 25 ft W x 9 ft H. Large doors in the front of the building will allow a dump truck operator to drive into the building and up a concrete ramp to unload the contaminated soil. The storage crib walls, approximately 9 feet high, will be made up of reinforced concrete. A bulldozer operator will be responsible for leveling and spreading the soil in the storage cribs from the dump truck. A front-end loader will be required in order to load contaminated soil onto a conveyor belt for transport to the soil treatment building. In terms of ventilation and dust control for the soil receiving building, a dust collection system, baghouse system, and a high-efficiency particulate air (HEPA) filter system will be required. The dust collection system will be located primarily above the storage crib area, at an estimated cost of $100,000. Costs for the baghouse system and HEPA filter system are estimated at

$170,000 and $60,000, respectively. The 80 ft x 220 ft building has an estimated cost of approximately $700,000 (at $40/sq. ft). Assuming a 9 inch slab for the foundation, the cost to construct the building foundation is approximately $280,000 (at $500/cu. yd. of concrete). Including the dust and ventilation equipment, as well as other associated costs for the building, such as the foundation, containment concrete walls, ramps, etc., bring the total estimated cost for the soil receiving building to approximately $1,400,000. The FCI requirements for the soil receiving building are shown in Table 2. The equipment costs (E) for the building are estimated at $100,000, which includes an enclosed conveyor belt system and associated equipment used to transport the contaminated soil to the soil treatment building. The estimated FCI requirements, including direct and indirect plant costs, contractor's fee, and contingency, to construct the soil receiving building is approximately $2,150,000.

The operating costs for the soil receiving building are shown in Table 3. The costs to excavate the contaminated soil and load it into a dump truck were obtained from Richardson's Engineering Services *Process Plant Construction Estimating Standards*, 1994 edition. Estimated costs for excavation and loading a dump truck are $0.82/cu. yd. and $0.50/cu. yd, respectively. Adding a stiff clay factor of $0.93/cu. yd brings the total cost for excavation to $2.25/cu. yd ($2.25/ton). Based on the soil treatment rate and the capacities of the excavation and hauling equipment, it is anticipated that the soil receiving building will have to be in operation for 3 shifts per day, 5 days per week, and 350 days per year. An assumption was made that a stockpile of soil could be produced for continuous treatment by excavating the soil at a rate of 84 tons per hour for 40 hours per week. This excavation rate will produce 3,360 tons of soil per week, which is enough soil to operate the treatment facility continuously at a 20 tons per hour treatment rate (20 tons/hr x 24 hrs/day x 7 days/wk). As shown on the daily operating cost sheets for each treatment technology, the labor requirements vary for each of the 3 shifts during the day. The following personnel are required for the soil receiving building during the 8-hour day shift: 1) a truck driver to haul the soil into the building, 2) a bulldozer operator, 3) a front-end loader operator, 4) a general employee, 5) a radiation monitoring person, and 6) a shift supervisor. Using a 70% utilization factor and a 20 tons/hr treatment rate, 112 tons of contaminated soil can be treated during the shift (20 tons/hr x 8hrs/shift x 0.70). Under these assumptions, the estimated labor cost per ton of soil treated for the day shift is roughly $5/ton (Table 2). Based on an excavation rate of 84 tons/hr for 40 hrs/wk, the bulldozer operators' and truck drivers' excavation activities can be completed during the first shift of the day.

In terms of labor requirements, the following personnel are required for the second shift: 1) a front-end loader operator, 2) a general employee, and 3) radiation monitoring person. In addition, the assumption was made that the soil receiving building shift supervisor would work a 10 hour day to oversee the first 2 hours of the second shift. The 2 hours of overtime pay for the shift supervisor during the day shift were paid at 1.5 times his or her's base salary. The labor rates of employees on the second and third shifts were increased by 25% to account for a shift differential. This factor was applied to employees working in the soil receiving building and the soil treatment building. The estimated cost per ton for the operating labor during the second shift is $4/ton. The 8-hour third shift personnel required for the soil receiving building is identical to that of the second shift, with the exception of the requirement for a shift supervisor during the first 2 hours of the shift. Assuming a 70% utilization factor, the estimated labor cost per ton of soil treated during the third shift is $3/ton. Maintenance employees were accounted for in the labor requirements of the soil treatment building because it was assumed that they would perform maintenance work activities for both the soil receiving and soil treatment buildings. The total estimated labor cost for the soil receiving building, excluding excavation costs ($2.25/ton), is approximately $12 per ton of soil treated (Table 2). The costs associated with the construction and operation of the soil receiving building were included in the cost estimates for the treatment technologies in which the soil receiving building is actually required (all of the technologies with the exception of the heap leaching process).

Table 2. FCI for soil receiving building

Item	Symbol	Description	Formula	Value
1	E	Equipment Costs	E	$100,000
2	L	Cost of Installation Labor	.40E	40,000
3	IC	Instrumentation & Controls	.13E	$13,000
4	I	Insulation Costs (equip. & piping)	.05E	$5,000
5	P	Piping	.16E	$16,000
6	Q	Labor for Installation of Piping	.40P	$6,400
7	F	Electrical Installations	.15E	$15,000
8	B	Building including Services	----	$1,400,000
9	Y	Yard Improvements	.10E	$10,000
10	S	Service Facilities	.30E	$30,000
	D	**Direct Plant Cost**	**Sum of 1-10**	**$1,635,400**
11	ES	Engineering & Supervision	.60E	$60,000
12	C	Construction Expenses	.25E	$25,000
	IP	**Indirect Plant Cost**	**ES+C**	**$85,000**
13	CF	Contractor's Fee	.05(D+IP)	$86,020
14	CO	Contingency	.20(D+IP)	$344,080
	FCI	**Fixed Capital Investment**	**D+IP+CF+CO**	**$2,150,500**

7. AQUEOUS BIPHASIC EXTRACTION PROCESS

The goal of the ABE process in removing uranium from contaminated soils is to selectively separate and recover ultrafine particulate uranium from the soil without altering the physicochemical properties of the soil particles. Principal investigators who studied this technology feel this separation is feasible by taking advantage of the differences in the surface chemical properties of the contaminants and the soil particles. The biphasic extraction process is a potential alternative to conventional soil washing techniques that are based on physical separation methods, such as screening, classification, and

Table 3. Operating costs for soil receiving building

Class	Description	Unit	Unit Cost	Qty/Day	Sub-Total	Total	Cost/Ton
Excavation	Excavate	ton	$0.82	336.0	275.5		
	Load Truck	ton	$0.50	336.0	168.0		
	Stiff Clay Factor	lbs	$0.93	336	312.5	$756	$2
First Shift	Truck Driver	man-hrs	23.17	8	185.4		
	Dozer Operator	man-hrs	34.1	8	272.8		
	Front-End Loader	man-hrs	$34.10	8	$273		
	General Employee	man-hrs	$34.10	8	$273		
	Rad Monitoring	man-hrs	$40.92	8	$327		
	Shift Supervisor	man-hrs	$45.35	8	$363	$1,694	$5
Second Shift	Front-End Loader	man-hrs	$42.63	8	$341		
	General Employee	man-hrs	$42.63	8	$341		
	Rad Monitoring	man-hrs	$51.15	8	$409		
	Shift Supervisor	man-hrs	$68.03	2	$136	$1,227	$4
Third Shift	Front-End Loader	man-hrs	$42.63	8	$341		
	General Employee	man-hrs	$42.63	8	$341		
	Rad Monitoring	man-hrs	$51.15	8	$409	$1,091	$3
Maintenance and Repairs	Equipment	day	$42.86	1	$43		
	Building	day	$160.00	1	$160	$203	$1
Operating Supplies	Total	day	$30.43	1	$30	$30	$0.09
Fixed Charges	Taxes	day	$122.89	1	$123		
	Insurance	day	$61.44	1	$61	$184	$1
Contingency	Total	day	$1,296.54	1	$1,297	$1,297	$4
Total Daily Operating Cost						$6,483	$19

flotation. The ABE process involves the use of a multi-stage tower contacting device known as a Karr Column to selectively partition the ultra fine soil particles between two immiscible aqueous phases. In the aqueous biphasic extraction process, one liquid phase is a solution of PEG and the other is a sodium carbonate salt solution. In this phase, the dense liquid phase (the PEG slurry) is fed to the top of the Karr Column. The less dense salt phase is fed into the bottom of the column and a counter-current flow is established. The approach taken by the Principal Investigator is to partition the uranium rich solid soil particles into the salt phase leaving the top of the Karr Column. Also in this salt phase is the bulk of the dissolved uranium. The PEG underflow slurry should exit the bottom of the Karr Column containing the uranium-depleted soil. This Karr Column PEG underflow slurry is then mixed with the coarser soil fractions, fed to a thickener for separation from the PEG phase and ultimately filtered and washed for disposal (e.g., return to the site). The coarse soil separation stage of the ABE process is identical to that of the carbonate/bicarbonate vat extraction treatment technology. The philosophy of the coarse soil separation circuit for the ABE process is to produce a polyethylene glycol (PEG) phase feed slurry to the Karr Columns of 30% to 35% solids, leach any surface uranium contamination from the coarse soil fraction and wash the coarse soil with fresh water prior to return to the site (Henderson 1995). A detailed description of the ABE process, as well as the carbonate/bicarbonate, and tiron soil washing processes, can be found in the report, *Removal of Uranium from Uranium-Contaminated Soils Phase I: Bench-Scale Testing* (Soil Decontamination Task Group 1993).

The uranium recovery system used to precipitate the uranium from the sodium carbonate salt phase in the ABE process is methanol precipitation. However, specific quantitative data required for process design are very preliminary due to the developmental nature of this technology. Since the ABE process uses a separation technology other than leaching to effect the separation, developers of this process feel this technology has the promise of being able to recover and remove a high proportion of the difficult-to-leach, refractory uranium mineral species in the soil (Henderson 1995). Many elements of the ABE process are preliminary since this type of full-scale application is early in its development phase. Therefore, the engineering flow diagram for this technology represents a feasible, although not necessarily optimal, conceptual process design for implementation of a full-scale treatment facility. However, the potential exists for significant improvements in the ABE process design, operating performance, and reduction in reagent consumptions and losses.

Appendix A provides the engineering flow diagram for the ABE system. The equipment costs for this treatment technology are shown in Appendix B, Table B.1. The total equipment cost (E) for this system is estimated at approximately $4,585,000. The Karr reciprocating plate columns are by far the most expensive pieces of process equipment required for the aqueous biphasic extraction system. Based on conversations with the principal investigator, a company provided a budget estimate of $380,000 for each Karr Column. Preliminary sketches from the company indicate that each Karr Column should be 6 feet in diameter, with a 15-ft plate stack height. Due to the requirement for 5 Karr Columns, the total cost for the Karr Columns is $1,900,000, which represents 42% of the total equipment cost for the aqueous biphasic extraction process. Two belt-type horizontal pressure filters are also required, with a budgeted cost of $400,000 each. Other large equipment expenditures include a multi-plate methanol strip column at $200,000, a PEG thickener at $150,000, and a drum scrubber with trommel at an estimated cost of $150,000.

The results of the FCI calculation for the ABE process are shown in Table 4. Piping and instrumentation costs for all of the technologies fall within the same range, except for the heap leaching system. After a review of piping and instrumentation costs for several of the treatment technologies developed by Brown & Root, Inc., it was decided that the lower end of the range would be an appropriate factor to use for piping costs (.16E), as well as insulation costs for the equipment and piping (.05E). Another reason the lower end of the range was used for the piping and insulation costs is that a factor for instrumentation and controls, at 13% of equipment costs, is already being considered in the FCI calculations. Piping and instrumentation costs are interrelated and are normally included under the same heading for most FCI cost estimates. Yard improvements for all the technologies were estimated at the lower end of the range because the concern is primarily for the full-scale treatment facility itself and not the extra cost items (i.e., sidewalks and landscaping) normally considered in this cost category. A value of 55% of equipment costs was used for service facilities in the FCI calculations, because it represents an average value for a solid-fluid processing plant.

Building size requirements for the ABE process were determined after Brown & Root personnel estimated that a 15,000 ft^2 soil treatment building would be required for the carbonate/bicarbonate and tiron soil washing systems. Because less equipment is required for the ABE process than that of the carbonate/bicarbonate and tiron systems, it was assumed that a 10,000 ft^2 building would be adequate for the ABE process soil treatment building. As with the other treatment technologies, unit costs are based on square footage requirements to estimate costs for the treatment building, building foundation, loading/unloading area, and parking area. The treatment building has an estimated cost of $400,000. Other costs associated with the treatment building, including the building and equipment foundations, a loading/unloading area, and a parking area result in a building cost of approximately $979,000. In addition, ventilation and dust control equipment will be required for the soil treatment building, including

Table 4. FCI for aqueous biphasic extraction process

Item	Symbol	Description	Formula	Value
1	E	Equipment Costs	E	$4,585,100
2	L	Cost of Installation Labor	.40E	$1,834,040
3	IC	Instrumentation & Controls	.13E	$596,063
4	I	Insulation Costs (equip. & piping)	.05E	$229,255
5	P	Piping	.16E	$733,616
6	Q	Labor for Installation of Piping	.40P	$293,446
7	F	Electrical Installations	.15E	$687,765
8	B	Building including Services	----	$1,300,000
9	Y	Yard Improvements	.10E	$458,510
10	S	Service Facilities	.55E	$2,521,805
	D	**Direct Plant Cost**	**Sum of 1-10**	**$13,239,600**
11	ES	Engineering & Supervision	.60E	$2,751,060
12	C	Construction Expenses	.25E	$1,146,275
	IP	**Indirect Plant Cost**	**ES+C**	**$3,897,335**
13	CF	Contractor's Fee	.05(D+IP)	$856,847
14	CO	Contingency	.20(D+IP)	$3,427,387
	FCI	**Fixed Capital Investment**	**D+IP+CF+CO**	**$21,421,169**

a dust collection system, baghouse system, and a high-efficiency particulate air (HEPA) filter system. The dust collection system will be located primarily above the storage crib area, at an estimated cost of $100,000. Costs for the baghouse system and HEPA filter system are estimated at $170,000 and $60,000, respectively. This results in an estimate of approximately $1,300,000 for the ABE process soil treatment building. The direct plant cost for this process is estimated at $13,240,000. Indirect plant costs, consisting of construction expenses and engineering and supervision, for this system are $3,897,000. Adding a contractor's fee and contingency of $857,000 and $3,427,000, respectively, to the direct and indirect plant costs results in a FCI of approximately $21,421,000 for the aqueous biphasic extraction process.

The daily operating costs associated with a full-scale treatment facility based on the ABE process are shown in Table 5. At approximately $65 per ton of soil treated, the chemical consumption costs required to operate this technology are the highest of any of the treatment technologies. This is due primarily to the requirement for such large quantities of sodium bicarbonate and makeup PEG, whose

Table 5. Operating costs for the aqueous biphasic extraction process

Class	Description	Unit	Unit Cost	Qty/Day	Sub-Total	Total	Cost/Ton
Chemical/Raw Material	Sodium Bicarbonate	ton	$416.00	25.0	$10,400		
	Makeup Methanol	gal	$0.75	4939	$3,704		
	Makeup PEG	lbs	$1.00	7293	$7,293		
	Flocculant	lbs	$1.80	173	$311		
	Bleed Solution	gal	$0.01	8961	$45		
	Phosphoric Acid	lbs	$0.34	45	$15		
	Makeup Water	1000 gal	$0.89	126	$112	$21,880	$65
First Shift	Plant Operators (7)	man-hrs	$34.10	56	$1,910		
	Rad Monitoring (2)	man-hrs	$40.92	16	$655		
	Shift Supervisors(2)	man-hrs	$45.35	16	$726		
	General Employee	man-hrs	$34.10	8	$273		
	Maintenance Employees(2)	man-hrs	$27.28	16	$436		
	Electrician	man-hrs	$33.19	8	$266		
	Instrument Technician	man-hrs	$33.19	8	$266	$4,530	$13
Second Shift	Plant Operators(5)	man-hrs	$42.63	40	$1,705		
	Rad Monitoring	man-hrs	$51.15	8	$409		
	Shift Supervisor	man-hrs	$56.69	8	$454		
	General Employee	man-hrs	$42.63	8	$341		
	Maintenance Employees(2)	man-hrs	$34.10	8	$273		
	Electrician	man-hrs	$41.49	8	$332		
	Instrument Technician	man-hrs	$41.49	8	$332	$3,846	$11
Third Shift	Plant Operators(5)	man-hrs	$42.63	40	$1,705		
	Rad Monitoring	man-hrs	$51.15	8	$409		
	Shift Supervisor	man-hrs	$56.69	8	$454		
	General Employee	man-hrs	$42.63	8	$341		
	Maintenance Employees(2)	man-hrs	$34.10	8	$273		
	Electrician	man-hrs	$41.49	8	$332		
	Instrument Technician	man-hrs	$41.49	8	$332	$3,846	$11
Electricity	Methanol Stripping	lbs	$0.004	480000	$1,920		
	Electrical Power	kwh	$0.08	12000	$960	$2,880	$9
Maintenance and Repairs	Equipment	day	$1,965.04	1	$1,965		
	Building	day	$148.57	1	$149	$2,114	$6
Operating Supplies	Total	day	$317.04	1	$317	$317	$1
Fixed Charges	Taxes	day	$1,224.07	1	$1,224		
	Insurance	day	$612.03	1	$612	$1,836	$5
Contingency	Total	day	$10,312.15	1	$10,312	$10,312	$31
Soil Receiving Building	Total	day	$6,482.69	1	$6,483	$6,483	$19
Total Daily Operating Cost						$58,043	$173

costs account for roughly 81% of the total daily chemical consumption costs for the ABE process. The labor requirements for this technology include personnel for the soil receiving building and the soil treatment building. The daily costs for personnel operating the soil receiving building during the first, second, and third shifts is estimated at $12 per ton of soil. The derivation of this estimate is explained in detail in Section 6 of this report. A total operating cost of $19/ton for the soil receiving building was included in the ABE operating cost estimate since this building is required for this technology. The personnel for the first shift of the ABE soil treatment building include the following: 1) 7 plant operators to operate the process equipment and conduct soil treatment activities, 2) 2 radiation monitoring personnel, 3) 2 shift supervisors, 4) an employee who will perform various general activities associated with the treatment process, 5) 2 general maintenance employees, and 6) an electrician and an instrument technician for the installation and maintenance of the process equipment and instrumentation used in the treatment process. The total estimated cost for the soil treatment building personnel during the first shift is $4,500, or approximately $13 per ton of soil treated, assuming a 70% utilization factor. Based on the 70% utilization factor, 336 tons/day, rather than 480 tons/day (20 tons/hour x 24 hrs/day) of contaminated soil can be treated for all of the treatment technologies because all were designed assuming a treatment process rate of 20 tons/hour and a utilization factor of 0.70.

Brown & Root, Inc. suggested less personnel would be required for the soil treatment building during the second and third shifts, compared to the first shift. The personnel for the second and third shifts of the ABE soil treatment building include the following: 1) 5 plant operators, 2) 1 radiation monitoring personnel, 3) 1 shift supervisor, 4) an employee who will perform various general activities associated with the treatment process, 5) 1 general maintenance employee, and 6) an electrician and an instrument technician. The total estimated cost for the second and third shifts for the soil treatment building labor is $3,800 each, or $11 per ton of soil treated (assuming a 70% utilization factor). Therefore, the labor cost for the soil treatment building is approximately $35 per ton of soil treated by the ABE process. Adding the soil receiving building labor cost of $12/ton to this value results in a total labor cost for the ABE process of $47/ton of soil treated. The estimated total daily operating cost of $173 per ton of soil treated for the ABE process includes costs for raw chemicals and materials, labor, electricity, maintenance and repairs, operating supplies, fixed charges, and contingency

The life-cycle cost calculation for the ABE process is shown in Table 6. The FCI value shown in the table ($23,571,669) is the sum of the FCI estimate for the soil receiving building and the FCI for the ABE soil treatment building. Start-up costs for all of the treatment technologies are estimated as 10% of the FCI requirements. Therefore, the cost in the first year (year 0) of the life-cycle cost calculations is the sum of the estimated FCI and the start-up costs associated with a full-scale facility based on the ABE process. The yearly operating cost of approximately $20,315,000 in years 1 through 17 for the ABE process is obtained by multiplying the daily operating cost (shown in Table 5) by 350, the estimated number of days the full-scale facility is operated in a single year. The cost to treat the contaminated soil at Fernald using the ABE process is estimated at $275 per ton. This value is based on the estimated life-cycle costs, comprised of the FCI, start-up expenses, and yearly operating costs, as well as the assumption of a 5% annual inflation rate for the 17 years of operation required to treat the estimated 2,000,000 tons of contaminated soil at the Fernald site.

8. CARBONATE/BICARBONATE VAT EXTRACTION PROCESS

The general approach taken by Principal Investigators studying the carbonate/bicarbonate vat extraction process was to emphasize the extraction of uranium from Fernald soils by carbonate-based extractions. Uranium is characteristically leached from uranium ores by acid- or carbonate-based extractants. However, because of the destructive action on layer silicates by strong acids, it was determined that acid leaching was not appropriate for Fernald soils. Alkaline leaching of uranium from various ores has an established history in the uranium industry that extends back to the middle 1950's, when uranium milling operations were at peak production. The use of sodium carbonate-sodium bicarbonate became attractive in cases in which the uranium grade was high or the carbonate or lime content was high. The alkaline leaching also produced a clean separation of uranium from its ores without solubilizing other metals because many metals are not soluble in alkaline solutions, an additional advantage when leaching soils that may contain hazardous metals (Soil Decontamination Task Group 1993).

Table 6. Life-cycle costs for aqueous biphasic extraction process

FCI	**$23,571,669**
Start-Up Cost	**$2,357,167**
Yearly Operating Cost	**$20,315,200**
Year	**Cost**
0	$25,928,836
1	$20,315,200
2	$21,330,960
3	$22,397,508
4	$23,517,383
5	$24,693,252
6	$25,927,915
7	$27,224,310
8	$28,585,526
9	$30,014,802
10	$31,515,542
11	$33,091,319
12	$34,745,885
13	$36,483,180
14	$38,307,339
15	$40,222,706
16	$42,233,841
17	$44,345,533
Total Cost	**$550,881,036**
Cost/Ton	**$275**

The soil feed to the carbonate/bicarbonate vat extraction process is delivered by a conveyor from the soil receiving building. The rocks, twigs, and roots are separated from the feed soil by a wet grizzly. The feed soil is then further processed and separated in a series of rotary drum scrubbers to produce a feed slurry to the carbonate leach reactors of 30% to 35% solids. Oxygen gas is introduced into each reactor vessel to provide maximum efficiency of uranium leaching. After initial leaching occurs in a series of three reactors, the slurry is then fed to a vacuum belt filter for dewatering. The soil slurry is then fed through another series of three leach reactors for additional leaching. The slurry exiting the last reactor is pumped to another vacuum belt filter. However, in addition to dewatering, this belt filter system contains two washing/rinsing sections, one using recycle lixiviant to remove solubilized uranium and the second uses fresh water to further remove the uranium in the soil filter cake. The washed and rinsed fine soil filter cake is then discharged from the second belt filter and conveyed to a stockpile for return to the site for disposal (Henderson 1995).

A fixed-bed ion exchange system is proposed for the removal and recovery of solubilized uranium. This system will allow a recycle rate of greater than 90% for the lixiviant after initial uranium removal. Ion exchange for the removal of uranium from carbonate lixiviants is a proven system used commercially for over 20 years (Henderson 1995). The filtrates used as feed to the ion exchange system are processed initially through sand filters to remove any suspended solids and then through carbon guard columns to remove any dissolved or suspended organics. The uranium is then stripped from the loaded resin using a sodium chloride/dilute hydrochloric acid strip solution. A peroxide uranium precipitation system, composed of a multi-compartment reactor, is used to remove the uranium from the acidic strip solutions. Each compartment is agitated using axial-flow impellers to promote crystal growth of the precipitate. During the precipitation process, the pH is adjusted and controlled throughout the process by the addition of hydrogen peroxide and sodium hydroxide. The reaction products from the peroxide precipitation produces solid uranyl peroxide and additional sodium chloride in the liquid phase. A filter press dewaters the uranyl peroxide slurry before disposal (Henderson 1995).

The engineering flow diagram for the carbonate/bicarbonate vat extraction process is shown in Appendix A. The equipment costs for the sodium carbonate system are shown in Appendix B, Table B.2. The total equipment cost (E) is estimated at approximately $6,151,000. The vacuum belt filters are the most costly pieces of equipment for this system. The estimated cost of vacuum belt filters 1 and 2 are $750,000 and $1,100,000, respectively. The 6 fixed bed ion exchange system columns cost approximately $1,300,000. The vacuum belt filters and ion exchange columns represent approximately 51% of the total equipment cost for the entire carbonate/bicarbonate vat extraction process. Other costly items include two carbon guard columns at a total cost of $400,000, two sand filters at $198,000, and a soil rotary dryer at a total estimated cost of $200,000.

Results of the FCI calculation for the carbonate/bicarbonate vat extraction process are shown in Table 7. Building costs are based on the assumption of a 15,000 ft^2 treatment building with a partial second floor. Unit costs based on square footage requirements were used to estimate costs for the treatment building, building foundation, loading/unloading area, and parking area. The treatment building has an estimated cost of $600,000. Other costs associated with the treatment building, including the building and equipment foundations, a loading/unloading area, and a parking area result in a building cost of approximately $1,272,000. In addition, ventilation and dust control equipment will be required for the soil treatment building, including a dust collection system, baghouse system, and a high-efficiency particulate air (HEPA) filter system. The dust collection system will be located primarily above the storage crib area, at an estimated cost of $100,000. Costs for the baghouse system and HEPA filter system are estimated at $170,000 and $60,000, respectively. This results in an estimate of approximately $1,600,000 for the carbonate/bicarbonate process soil treatment building. The direct cost for the carbonate/bicarbonate vat extraction plant is approximately $17,617,000. As in the case for the soil receiving building, engineering and supervision and contingency costs for all of the treatment technologies were estimated conservatively because of the experimental nature of this project. Indirect plant costs, consisting of construction expenses and engineering and supervision, for this system are approximately $5,228,000. Adding the contractor's fee and contingency of $1,142,000 and $4,569,000, respectively, to the direct and indirect plant costs results in a FCI of approximately $28,557,000 for the carbonate/bicarbonate system.

The daily operating costs associated with a full-scale treatment facility based on the carbonate/bicarbonate vat extraction process are shown in Table 8. The chemical consumption cost of $8 per ton of soil treated for the carbonate/bicarbonate process is the lowest of any of the treatment technologies. The labor requirements for this system include personnel requirements for the soil receiving building and the soil treatment building. As with the other treatment technologies requiring

Table 7. FCI for carbonate/bicarbonate vat extraction process

Item	Symbol	Description	Formula	Value
1	E	Equipment Costs	E	$6,151,060
2	L	Cost of Installation Labor	.40E	$2,460,424
3	IC	Instrumentation & Controls	.13E	$799,638
4	I	Insulation Costs (equip. & piping)	.05E	$307,553
5	P	Piping	.16E	$984,170
6	Q	Labor for Installation of Piping	.40P	$393,668
7	F	Electrical Installations	.15E	$922,659
8	B	Building including Services	----	$1,600,000
9	Y	Yard Improvements	.10E	$615,106
10	S	Service Facilities	.55E	$3,383,083
	D	**Direct Plant Cost**	**Sum of 1-10**	**$17,617,361**
11	ES	Engineering & Supervision	.60E	$3,690,636
12	C	Construction Expenses	.25E	$1,537,765
	IP	**Indirect Plant Cost**	**ES+C**	**$5,228,401**
13	CF	Contractor's Fee	.05(D+IP)	$1,142,288
14	CO	Contingency	.20(D+IP)	$4,569,152
	FCI	**Fixed Capital Investment**	**D+IP+CF+CO**	**$28,557,202**

the soil receiving building, the daily labor costs for personnel during the first, second, and third shifts, excluding excavation costs ($2.25/ton), is estimated at $12 per ton of soil treated. The personnel for the first shift of the carbonate/bicarbonate soil treatment building include the following: 1) 9 plant operators to operate the process equipment and conduct soil treatment activities, 2) 2 radiation monitoring personnel, 3) 2 shift supervisors, 4) an employee who will perform various general activities associated with the treatment process, 5) 2 general maintenance employees, and 6) an electrician and an instrument technician for the installation and maintenance of the process equipment and instrumentation used in the treatment process. The total estimated cost for the soil treatment building personnel during the first shift is $5,100, or $15 per ton of soil treated, assuming a 70% utilization factor. Based on the 70% utilization factor, 336 tons/day, rather than 480 tons/day (20 tons/hour x 24 hrs/day) of contaminated soil can be treated for all of the treatment technologies because all were designed assuming a treatment process rate of 20 tons/hour and a utilization factor of 70%.

Table 8. Operating costs for carbonate/bicarbonate vat extraction process

Class	Description	Unit	Unit Cost	Qty/Day	Sub-Total	Total	Cost/Ton
Chemical/Raw Material	Sodium Carbonate	ton	$381.00	0.9	$343		
	Sodium Bicarbonate	ton	$416.00	1.6	$666		
	Polymer	lbs	$1.80	540	$972		
	Hydrogen Peroxide	lbs	$0.50	101	$51		
	Sodium Hydroxide	lbs	$0.15	651	$98		
	Sodium Chloride	lbs	$0.29	1381	$400		
	Hydrochloric Acid	lbs	$0.04	1381	$59		
	Make-up Water	1000 gal	$0.89	49.8	$44	$2,633	$8
First Shift	Plant Operators (9)	man-hrs	$34.10	72	$2,455		
	Rad Monitoring (2)	man-hrs	$40.92	16	$655		
	Shift Supervisors(2)	man-hrs	$45.35	16	$726		
	General Employee	man-hrs	$34.10	8	$273		
	Maintenance Employees(2)	man-hrs	$27.28	16	$436		
	Electrician	man-hrs	$33.19	8	$266		
	Instrument Technician	man-hrs	$33.19	8	$266	$5,076	$15
Second Shift	Plant Operators(7)	man-hrs	$42.63	56	$2,387		
	Rad Monitoring	man-hrs	$51.15	8	$409		
	Shift Supervisor	man-hrs	$56.69	8	$454		
	General Employee	man-hrs	$42.63	8	$341		
	Maintenance Employee	man-hrs	$34.10	8	$273		
	Electrician	man-hrs	$41.49	8	$332		
	Instrument Technician	man-hrs	$41.49	8	$332	$4,528	$13
Third Shift	Plant Operators(7)	man-hrs	$42.63	56	$2,387		
	Rad Monitoring	man-hrs	$51.15	8	$409		
	Shift Supervisor	man-hrs	$56.69	8	$454		
	General Employee	man-hrs	$42.63	8	$341		
	Maintenance Employee	man-hrs	$34.10	8	$273		
	Electrician	man-hrs	$41.49	8	$332		
	Instrument Technician	man-hrs	$41.49	8	$332	$4,528	$13
Class	**Description**	**Unit**	**Unit Cost**	**Qty/Day**	**Sub-Total**	**Total**	**Cost/Ton**
Electricity	Total	kwh	$0.08	35000	$2,800	$2,800	$8
Maintenance and Repairs	Equipment	day	$2,636.17	1	$2,636		
	Building	day	$182.86	1	$183	$2,819	$8
Operating Supplies	Total	day	$422.85	1	$423	$423	$1
Fixed Charges	Taxes	day	$1,631.84	1	$1,632		
	Insurance	day	$815.92	1	$816	$2,448	$7
Contingency	Total	day	$6,313.37	1	$6,313	$6,313	$19
Soil Receiving Building	Total	day	$6,482.69	1	$6,483	$6,483	$19
Total Daily Operating Cost						$38,050	$113

As with all of the treatment technologies, Brown & Root, Inc. recommended less personnel would be required for the soil treatment building during the second and third shifts, compared to the first shift. The personnel for the second and third shifts of the carbonate/bicarbonate soil treatment building include the following: 1) 7 plant operators, 2) 1 radiation monitoring personnel, 3) 1 shift supervisor, 4) an employee who will perform various general activities associated with the treatment process, 5) 1 general maintenance employee, and 6) an electrician and an instrument technician. The total estimated cost for the second and third shifts for the soil treatment building labor is $4,500 each, or approximately $13 per ton of soil treated (assuming a 70% utilization factor). Therefore, the labor cost for the soil treatment building is approximately $41 per ton of soil treated by the carbonate/bicarbonate vat extraction process. Adding the soil receiving building labor cost of $12/ton to this value results in a total labor cost for the carbonate/bicarbonate vat extraction process of $53/ton of soil treated. Total operating costs for the soil receiving building of $19 per ton were also included in this operating cost estimate since this building is required for this treatment technology. The estimated total daily operating cost of $113 per ton of soil treated for this process includes costs for raw chemicals and materials, labor, electricity, maintenance and repairs, operating supplies, fixed charges, and contingency.

The life-cycle cost estimate for the carbonate/bicarbonate process is shown in Table 9. Costs incurred in the first year (year 0) include the FCI for the soil receiving building and the carbonate/bicarbonate treatment building, as well as the estimated start-up costs associated with developing a full-scale facility based on this process. The yearly operating cost of approximately $13,317,000 in years 1 through 17 for this process is obtained by multiplying the daily operating cost by 350, the estimated number of days the full-scale facility is operated in a single year. The total estimated cost of $189 per ton of soil treated using the carbonate/bicarbonate process is based on the estimated life-cycle costs, comprised of the FCI, start-up expenses, and yearly operating costs and the assumption of a 5% annual inflation rate.

Table 9. Life-cycle costs for carbonate/bicarbonate vat extraction process

FCI	**$30,707,702**
Start-Up Cost	**$3,070,770**
Yearly Operating Cost	**$13,317,342**
Year	**Cost**
0	$33,778,472
1	$13,317,342
2	$13,983,210
3	$14,682,370
4	$15,416,489
5	$16,187,313
6	$16,996,679
7	$17,846,513
8	$18,738,838
9	$19,675,780
10	$20,659,569
11	$21,692,547
12	$22,777,175
13	$23,916,034
14	$25,111,835
15	$26,367,427
16	$27,685,798
17	$29,070,088
Total Cost	**$377,903,478**
Cost/Ton	**$189**

9. HEAP LEACHING PROCESS

The engineering flow diagram and equipment selection for this technology reflect the experience of Brown & Root, Inc. personnel in primary uranium ore processing as well as recent laboratory results on the uranium-contaminated soils at Fernald. The equipment selection and process design to develop the heap leaching process to full-scale status represents a relatively optimistic approach taken by Brown & Root personnel and the principal investigators of this technology. The engineering flow diagram for the heap leaching process is based on an overall throughput rate of 20 dry tons of soil per hour. After initial separation and processing using a dry grizzly and rotary drum scrubber, the soil is transported to a covered stockpile building, with a capacity of 2,400 tons of soil (3 excavation days inventory). From the stockpile building, the soil is carried by conveyor to an agglomerating drum and mixed with a sand/gravel bulking agent. After blending, the soil mixture is transported by a conveyor system to one of five leaching cells. The cells are made of multiple layers of asphalt and covered with a crushed rock protective layer. The asphalt layers are covered with a geotextile (40 mils thick) fabric sheet. Each cell would hold approximately 4,230 tons of soil/sand mixture, with 8,460 square feet of exposed top surface for lixiviant application. Leachate is applied to the soil using irrigation drip emitters. Active leaching with a sodium bicarbonate-based lixiviant takes place at any given time in 3 of the cells. Another cell is where new soil is being transported to, stacked on, and prepared for leaching. The last cell contains the soil which has been leached, rinsed, and drained.

A 7-day cycle duration was chosen because it is very compatible with a normal 5-day work week by an excavation contractor. The contaminated soil will be processed through 3 leach cycles, providing at least 18 days of active leaching of the soil. The remaining 3 days would be used to apply rinse water and initiate a drainage cycle. After the 3 leach cycles are completed, the heaps are rinsed with fresh water. After drainage, the soil is reclaimed by front-end loader which places the soil into a a conveyor feed hopper where it is transported and discharged onto the reclaim transport conveyor. The leached soil conveyor discharges to a leached soil pad where the soil is stockpiled prior to return to the disposal site. The uranium recovery system for the heap leaching process is very similar to the one proposed for the carbonate/bicarbonate soil washing process. The fixed-bed ion exchange system proposed for the heap leaching process is very similar to the one used for the carbonate/bicarbonate and tiron soil washing systems. It permits greater than 90% recycle and reuse of the lixiviant.

The engineering flow diagram for the heap leaching system is provided in Appendix A. The equipment list for the heap leaching system is shown in Appendix B, Table B.3. The total equipment cost (E) for this process is estimated at approximately $3,757,000. The 5,800 square foot circular stockpile building has an estimated cost of $290,500 (at $50/sq. ft.). The 71,000 square foot sand/soil permanent leaching pad is estimated at approximately $355,000 ($5/sq. ft.). Other costly items for this treatment technology include the conveyor systems. For example, the 730-ft cross-country leaching pad conveyor which transports the soil/sand blend from the agglomerating drum to the 5 leach pads is estimated at $219,000. The unloading conveyor, tranporting the leached soil to a soil pad for disposal, has an estimated cost of $200,000. Also, the fixed-bed ion exchange resin columns for the heap leaching process have an estimated cost of approximately $194,000, which includes the resin at $16,000 ($200/cu. ft.).

The results of the FCI calculation for the heap leaching process are shown in Table 10. Building costs were calculated using unit costs ($/sq. ft.) based on size requirements and the materials of construction for each item. A building is required to house the uranium removal and lixiviant recycle equipment shown in Appendix A. Because the uranium removal treatment equipment requirements are essentially the same for both the heap leaching and the carbonate/bicarbonate processes, the heap leaching building is assumed to be approximately one-half the size of the carbonate/bicarbonate system. This is based on the assumption that one-half of the building space for the carbonate/bicarbonate system is occupied by the uranium removal and lixiviant recycle equipment. Rather than taking one-half of the cost of the carbonate/bicarbonate soil treatment building ($1,600,000) since the heap leaching process building is one-half the size of the carbonate building, a six-tenths-factor scaling rule was used to calculate the cost of the heap leaching process building. According to this rule, if the cost of a given unit at one size is known, the cost of a similar unit with X times the capacity of the first is approximately $(X)^{0.6}$ times the cost of the initial unit (Peters and Timmerhaus 1991). Using this rule, the total estimated direct cost for the heap leaching building, including foundation, compartment walls, parking and loading/unloading area, dust collection system, hepa filters, and baghouse system, is approximately $1,060,000. This cost includes the material, equipment, and labor for fabrication of the heap leaching building which houses the uranium removal and lixiviant recycle equipment. Due to extensive piping requirements for the heap leaching process, the piping and insulation cost factors (see Table 7) were increased to reflect this change. After a review of the piping requirements in the equipment parts list for the heap leaching process and discussions with Brown & Root, Inc. personnel, it was determined that piping costs for the heap leaching process should be estimated conservatively. Therefore, the upper limit of the ranges shown in Table 1 was used for piping and insulation costs. The estimated direct plant cost for the heap leaching system is approximately $11,783,000. Total indirect plant costs for engineering and supervision and construction expenses are estimated at $3,194,000. Adding the contractor's fee and contingency of $749,000 and $2,995,000, respectively, to the direct and indirect plant costs results in a FCI value of approximately $18,721,000 for the heap leaching process. This technology has the lowest FCI requirements of the treatment technologies.

The daily operating costs associated with a full-scale treatment facility based on the heap leaching process are shown in Table 11. Chemical costs associated with this technology are estimated at $15 per ton of soil treated, with 85% of this cost due to the requirement for sand. The labor requirements for this system include personnel requirements for the soil and sand receiving pads, the covered stockpile building, the leach cells, and the soil treatment building which houses the uranium removal and lixiviant recycle stream equipment. The heap leaching process is the only treatment technology that does not require the soil receiving building. A soil receiving pad and the covered stockpile building are used to store the soil after day-time excavation operations. In terms of labor requirements, the soil and sand receiving pads each require a truck driver and a front-end loader operator. Because the excavation work can be completed in an 8-hour period, the the truck drivers and front-end loader operators for the receiving pads are only required for the first shift of the day. The estimated labor cost for these 4 personnel during the day shift are estimated at $900, or approximately $3/ton. Other personnel required during the first shift of the heap leaching process include the following: 1) a front-end loader operator to move the soil from the stockpile building to the soil transport conveyor, which then feeds the agglomerating/mixing drum, 2) 7 plant operators to cover various stages of the soil treatment and uranium removal/lixiviant recycle processes, 3) 2 radiation monitoring personnel, 3) 2 shift supervisors, 4) an employee who will perform various general activities associated with the treatment process, 5) 2 general maintenance employees, and 6) an electrician and an instrument technician for the installation and maintenance of the process equipment and instrumentation used in the heap leaching process. The

Table 10. FCI for heap leaching process

Item	Symbol	Description	Formula	Value
1	E	Equipment Costs	E	$3,757,310
2	L	Cost of Installation Labor	.40E	$1,502,924
3	IC	Instrumentation & Controls	.13E	$488,450
4	I	Insulation Costs (equip. & piping)	.09E	$338,158
5	P	Piping	.31E	$1,164,766
6	Q	Labor for Installation of Piping	.40P	$465,906
7	F	Electrical Installations	.15E	$563,597
8	B	Building including Services	----	$1,060,000
9	Y	Yard Improvements	.10E	$375,731
10	S	Service Facilities	.55E	$2,066,521
	D	**Direct Plant Cost**	**Sum of 1-10**	**$11,783,363**
11	ES	Engineering & Supervision	.60E	$2,254,386
12	C	Construction Expenses	.25E	$939,328
	IP	**Indirect Plant Cost**	**ES+C**	**$3,193,714**
13	CF	Contractor's Fee	.05(D+IP)	$748,854
14	CO	Contingency	.20(D+IP)	$2,995,415
	FCI	**Fixed Capital Investment**	**D+IP+CF+CO**	**$18,721,345**

total estimated cost for these personnel during the first shift is approximately $4,800, or $14 per ton of soil treated, assuming a 70% utilization factor and a treatment rate of 336 tons/day. Therefore, the total labor cost for the first shift personnel (including the soil and sand receiving pads) is $17/ton of soil treated.

It is anticipated that the second and third shifts require less personnel than the first shift for the heap leaching technology. The personnel for the second and third shifts of the heap leaching technology include the following: 1) a front-end loader operator, 2) 6 plant operators, 3) 1 radiation monitoring personnel, 4) 1 shift supervisor, 5) an employee who will perform various general activities associated with the treatment process, 6) 1 general maintenance employee, and 7) an electrician and an instrument technician. The total estimated cost for the second and third shifts for the labor involved in operating the heap leaching technology is approximately $4,500 each, or $13 per ton of soil treated (assuming a 70% utilization factor). Therefore, the total labor cost to operate the heap leaching process is approximately $43 per ton of soil. The labor costs for this technology are less than those of the other treatment technologies, primarily due to the lack of a requirement for a soil receiving building. The

Table 11. Operating costs for heap leaching process

Class	Description	Unit	Unit Cost	Qty/Day	Sub-Total	Total	Cost/Ton
Chemical/Raw Material	Sodium Carbonate	ton	$381.00	0.0	$0		
	Sodium Bicarbonate	ton	$416.00	0.0	$0		
	Hydrogen Peroxide	lbs	$0.50	101.0	$51		
	Hydrochloric Acid	lbs	$0.04	460.0	$20		
	Resin Replacement	ft^3	$200.00	0.014	$3		
	Carbon Replacement	lbs	$2.00	1.5	$3		
	Sand	ton	$10.00	415	$4,150		
	Sodium Chloride	lbs	$0.29	460	$133		
	Sodium Hydroxide	lbs	$0.15	504	$76		
	Carbon Dioxide	lbs	$0.12	2839	$341		
	Flocculant	lbs	$1.80	40	$72		
	Make-up Water	1000 gal	$0.89	38	$34	$4,881	$15
Excavation	Excavate	ton	$0.82	336	$276		
	Load Truck	ton	$0.50	336	$168		
	Stiff Clay Factor	ton	$0.93	336	$312	$756	$2
Soil/Sand Receiving Pads	Truck Drivers(2)	man-hrs	$23.17	16	$370		
	Front-End Loader Opers(2)	man-hrs	$34.10	16	$546	$916	$3
First Shift	Front-End Loader Oper.	man-hrs	$34.10	8	$273		
	Plant Operators(7)	man-hrs	$34.10	56	$1,910		
	Rad. Monitoring(2)	man-hrs	$40.92	16	$655		
	Shift Supervisors(2)	man-hrs	$45.35	16	$726		
	General Employee	man-hrs	$34.10	8	$273		
	Maintenance Employees(2)	man-hrs	$27.28	16	$436		
	Electrician	man-hrs	$33.19	8	$266		
	Instrument Technician	man-hrs	$33.19	8	$266	$4,803	$14
Second Shift	Front-End Loader Oper.	man-hrs	$42.63	8	$341		
	Plant Operators(6)	man-hrs	$42.63	48	$2,046		
	Rad. Monitoring	man-hrs	$51.15	8	$409		
	Shift Supervisors	man-hrs	$56.69	8	$454		
	General Employee	man-hrs	$42.63	8	$341		
	Maintenance Employees	man-hrs	$34.10	8	$273		
	Electrician	man-hrs	$41.49	8	$332		
	Instrument Technician	man-hrs	$41.49	8	$332	$4,528	$13
Third Shift	Front-End Loader Oper.	man-hrs	$42.63	8	$341		
	Plant Operators(6)	man-hrs	$42.63	48	$2,046		
	Rad. Monitoring(1)	man-hrs	$51.15	8	$409		
	Shift Supervisors(1)	man-hrs	$56.69	8	$454		
	General Employee	man-hrs	$42.63	8	$341		
	Maintenance Employees(1)	man-hrs	$34.10	8	$273		
	Electrician	man-hrs	$41.49	8	$332		
	Instrument Technician	man-hrs	$41.49	8	$332	$4,528	$13
Electricity	Total	kwh	$0.08	25000	$2,000	$2,000	$6
Maintenance and Repairs	Equipment	day	$1,610.28	1	$1,610		
	Building	day	$121.14	1	$121	$1,731	$5
Operating Supplies	Total	day	$259.71	1	$260	$260	$1
Fixed Charges	Taxes	day	$1,069.79	1	$1,070		
	Insurance	day	$534.90	1	$535	$1,605	$5
Contingency	Total	day	$6,501.87	1	$6,502	$6,502	$19
Total Daily Operating Cost						$32,509	$97

estimated total daily operating cost of $97 per ton of soil treated for this process includes costs for raw chemicals and materials, labor, electricity, maintenance and repairs, operating supplies, fixed charges, and contingency.

The life-cycle cost estimate for the heap leaching process, shown in Table 12, includes the FCI requirements and the estimated start-up costs in the first year (year 0) associated with a full-scale facility based on this process design. Also, since the heap leaching process design does not require the soil receiving building, FCI and operating costs for the soil receiving building were not included in the life-cycle cost estimate for this treatment technology. The yearly operating cost of $11,378,000 in years 1 through 17 for this process is obtained by multiplying the daily operating cost by 350, the estimated number of days the full-scale facility is operated in a single year. The total estimated cost of $157 per ton of soil treated using the heap leaching process is based on the estimated life-cycle costs, comprised of the FCI, start-up expenses, and yearly operating costs, as well as the assumption of a 5% annual inflation rate for the 17 years of operation required to treat the estimated 2,000,000 tons of contaminated soil.

Table 12. Life-cycle costs for heap leaching process

FCI	**$18,721,345**
Start-Up Cost	**$1,872,135**
Yearly Operating Cost	**$11,378,266**
Year	**Cost**
0	$20,593,480
1	$11,378,266
2	$11,947,180
3	$12,544,539
4	$13,171,766
5	$13,830,354
6	$14,521,872
7	$15,247,965
8	$16,010,363
9	$16,810,882
10	$17,651,426
11	$18,533,997
12	$19,460,697
13	$20,433,732
14	$21,455,418
15	$22,528,189
16	$23,654,599
17	$24,837,329
Total Cost	**$314,612,053**
Cost/Ton	**$157**

10. HIGH-GRADIENT MAGNETIC SEPARATION PROCESS

Like the other treatment technologies, the HGMS process is based on an optimistic process performance as a basis for the engineering flow diagram and equipment selection. The engineering design developed by Brown & Root, Inc. is based on bench-scale studies conducted by principal investigators studying this treatment technology. If adequate performance results from the bench-scale studies for the selected process concepts and the selection of equipment are supported by laboratory tests, the capital and operating cost estimates for this technology could be directly compared to the other treatment technologies (Sladic 1995). The coarse soil separation circuit for the HGMS process is very similar to the coarse soil separation process for the ABE, carbonate/bicarbonate, and the tiron soil washing processes. The soil feed to the HGMS process is initiated by reclaim from the soil receiving building. The coarse soil separation circuit removes the oversize (+2mm) soil fractions low in uranium contamination and produces 2 particle size fraction splits of feed slurry to the HGMS unit. Since the HGMS system requires relatively fine solid particles to work effectively, multiple size fraction operations are performed on the soil slurry prior to introduction to the HGMS unit. A roll crusher, attrition scrubber, and hydrosizer are used to produce a fine soil size fraction for feed to the HGMS unit. The HGMS unit consists of a porous magnetic matrix (possibly stainless steel wool) surrounded by a superconducting electromagnetic coil capable of creating an intensemagnetic field and cooled by a cryogenic system. Under such an intense magnetic field, paramagnetic compounds of relatively moderate magnetic susceptibility, such as uranium and uranium oxides, can be successfully separated from contaminated soils. Due to its superconducting properties, the HGMS unit consumes virtually no power. The HGMS process as a whole consumes approximately 19 kW (26 HP), primarily to operate the cryogenics compressor (Sladic 1995).

The HGMS slurry processing scheme consists of two passes through the magnetic matrix for each of the 2 size fractions, a backflush following the 2 passes for each size fraction, and an optional preliminary forward scalping pass and backflush for each size fraction for removal of materials with high magnetic susceptibility. For this preliminary study, the capacity of coarse and fine fraction feed tanks and the HGMS recycle feed tank have been designed to provide a minimum of 30 minutes residence time. In actuality, Brown & Root, Inc. personnel responsible for developing the engineering flow diagrams for this technology stated that tank sizing should be determined by factoring in the HGMS system's magnetic capture capacity and adding appropriate safety factors based on the volume of slurry and concentration of uranium in the slurry. However, this task was not performed due to time restrictions associated with the project. In terms of HGMS concentrate processing, the primary purpose of this step is to dewater the concentrate to minimize its volume to facilitate handling and disposal. The HGMS concentrate processing system consists of the HGMS concentrate thickener, the HGMS concentrate thickener underflow filter feed pump, a filter press, a filter cake hopper with screw type discharge auger, and an agitated decant water storage tank. The other processing stream that must be considered after treatment in the magnetic maxtrix is the one containing the HGMS tails. The purpose of the HGMS tails processing is very similar to that of the HGMS concentrate processing, except that the dewatered soils tails have most of the uranium concentration (to <50 ppm) removed and are returned to the site (Sladic 1995). The HGMS tails processing equipment consists of the flocculant feed system, a static mixer, the HGMS fines thickener, a thickener underflow filter feed pump, a high-pressure belt-type filter, a filter cake hopper with screw type discharge auger, and an agitated recycle water tank.

The engineering flow diagram for the HGMS process is shown in Appendix A. The equipment costs for this system are shown in Appendix B, Table B.4. The equipment on the front end of the process (coarse soil separation stage) is very similar to that required for the other technologies. However, the equipment requirements after the initial separation stage of the treatment process are fairly unique for this technology compared to the other technologies. The total equipment cost (E) is estimated at approximately $4,051,000. The most costly equipment for this system is the Eriez HGMS superconducting magnet, model SC 20-84, at an estimated cost of $1,900,000. This represents approximately 47% of the total equipment cost for the entire HGMS system. It is a commercial-size continuous feed unit which includes the coil, cryogenics, power supply, and controls. A fully-contained horizontal pressure filter, complete with leak sump and recycle pump, is estimated at $400,000. Other costly items for this technology include a recessed filter press at $223,000 and a 40-foot diameter thickener at $180,000.

Results of the FCI calculation for the HGMS process are shown in Table 13. Building costs for the HGMS process are based on the assumption of a 10,000 ft^2 treatment building with a partial

Table 13. FCI for the High-Gradient Magnetic Separation (HGMS) process

Item	Symbol	Description	Formula	Value
1	E	Equipment Costs	E	$4,051,400
2	L	Cost of Installation Labor	.40E	$1,620,560
3	IC	Instrumentation & Controls	.13E	$526,682
4	I	Insulation Costs (equip. & piping)	.05E	$202,570
5	P	Piping	.16E	$648,224
6	Q	Labor for Installation of Piping	.40P	$259,290
7	F	Electrical Installations	.15E	$607,710
8	B	Building including Services	----	$1,300,000
9	Y	Yard Improvements	.10E	$405,140
10	S	Service Facilities	.55E	$2,228,270
	D	**Direct Plant Cost**	**Sum of 1-10**	**$11,849,846**
11	ES	Engineering & Supervision	.60E	$2,430,840
12	C	Construction Expenses	.25E	$1,012,850
	IP	**Indirect Plant Cost**	**ES+C**	**$3,443,690**
13	CF	Contractor's Fee	.05(D+IP)	$764,677
14	CO	Contingency	.20(D+IP)	$3,058,707
	FCI	**Fixed Capital Investment**	**D+IP+CF+CO**	**$19,116,920**

second floor. Unit costs based on square footage requirements were used to estimate costs for the treatment building, building foundation, loading/unloading area, and parking area. The treatment building has an estimated cost of $400,000. Other costs associated with the treatment building, including the building and equipment foundations, a loading/unloading area, and a parking area result in a building cost of approximately $979,000. In addition, ventilation and dust control equipment will be required for the soil treatment building, including a dust collection system, baghouse system, and a high-efficiency particulate air (HEPA) filter system. The dust collection system will be located primarily above the storage crib area, at an estimated cost of $100,000. Costs for the baghouse system and HEPA filter system are estimated at $170,000 and $60,000, respectively. This results in an estimate of approximately $1,300,000 for the HGMS soil treatment building. The direct plant cost for this process is estimated at $11,850,000. As in the case for the soil receiving building, engineering and supervision and contingency costs for all of the treatment technologies were estimated conservatively because of the experimental nature of this project. Indirect plant costs, consisting of construction expenses and engineering and supervision, for this system are approximately $3,444,000. Adding the contractor's fee and contingency of $765,000 and $3,059,000, respectively, to the direct and indirect plant costs results in a FCI of approximately $19,117,000 for the HGMS process.

The daily operating costs associated with a full-scale treatment facility based on the HGMS process are shown in Table 14. Chemical consumption costs for this technology are estimated at $11 per ton of soil treated. The labor requirements for this system include personnel requirements for the soil receiving building and the soil treatment building. As with the other treatment technologies requiring the soil receiving building, the daily labor costs for personnel during the first, second, and third shifts, excluding excavation costs ($2.25/ton), is estimated at $12 per ton of soil treated. The derivation of this estimate is explained in detail in Section 6 of this report. The personnel for the first shift of the HGMS soil treatment building include the following: 1) 6 plant operators to operate the process equipment and conduct soil treatment activities, 2) 2 radiation monitoring personnel, 3) 2 shift supervisors, 4) an employee who will perform various general activities associated with the treatment process, 5) 2 general maintenance employees, and 6) an electrician and an instrument technician for the installation and maintenance of the process equipment and instrumentation used in the treatment process. The total estimated cost for the soil treatment building personnel during the first shift is $4,300, or $13 per ton of soil treated, assuming a 70% utilization factor.

As with all of the treatment technologies, Brown & Root, Inc. recommend less personnel would be required for the soil treatment building during the second and third shifts, compared to the first shift. The personnel for the second and third shifts of the HGMS soil treatment building include the following: 1) 4 plant operators, 2) 1 radiation monitoring personnel, 3) 1 shift supervisor, 4) an employee who will perform various general activities associated with the treatment process, 5) 1 general maintenance employee, and 6) an electrician and an instrument technician. The total estimated cost for the second and third shifts for the soil treatment building labor is $3,500 each, or $10 per ton of soil treated (assuming a 70% utilization factor). Therefore, the labor cost for the soil treatment building is approximately $33 per ton of soil treated by the HGMS process. Adding the soil receiving building labor cost of $12/ton to this value results in a total labor cost of $45/ton of soil treated using this process. Total operating costs for the soil receiving building of $19 per ton were also included in the operating cost estimate since this building is required for this technology. The estimated total daily operating cost of $94 per ton of soil treated for the HGMS process includes costs associated with operating a full-scale treatment facility. This includes costs for raw chemicals, labor, electricity, maintenance and repairs, operating supplies, fixed charges, and contingency factor of 25%.

Table 14. Operating costs for HGMS process

Class	Description	Unit	Unit Cost	Qty/Day	Sub-Total	Total	Cost/Ton
Chemical/Raw Material	Hexametaphosphate	lbs	$0.62	1632.0	1011.8		
	Sodium Dithionate	lbs	$0.71	1584.0	1124.6		
	Flocculant	lbs	$1.80	816	$1,469		
	Make-up Water	kgal	$0.89	43.2	$38	$3,605	$11
First Shift	Plant Operators (6)	man-hrs	$34.10	48	$1,637		
	Rad Monitoring (2)	man-hrs	$40.92	16	$655		
	Shift Supervisors(2)	man-hrs	$45.35	16	$726		
	General Employee	man-hrs	$34.10	8	$273		
	Maintenance Employees(2)	man-hrs	$27.28	16	$436		
	Electrician	man-hrs	$33.19	8	$266		
	Instrument Technician	man-hrs	$33.19	8	$266	$4,257	$13
Second Shift	Plant Operators(4)	man-hrs	$42.63	32	$1,364		
	Rad Monitoring	man-hrs	$51.15	8	$409		
	Shift Supervisor	man-hrs	$56.69	8	$454		
	General Employee	man-hrs	$42.63	8	$341		
	Maintenance Employee	man-hrs	$34.10	8	$273		
	Electrician	man-hrs	$41.49	8	$332		
	Instrument Technician	man-hrs	$41.49	8	$332	$3,505	$10
Third Shift	Plant Operators(4)	man-hrs	$42.63	32	$1,364		
	Rad Monitoring	man-hrs	$51.15	8	$409		
	Shift Supervisor	man-hrs	$56.69	8	$454		
	General Employee	man-hrs	$42.63	8	$341		
	Maintenance Employee	man-hrs	$34.10	8	$273		
	Electrician	man-hrs	$41.49	8	$332		
	Instrument Technician	man-hrs	$41.49	8	$332	$3,505	$10
Electricity	Total	kwh	$0.08	16000	$1,280	$1,280	$4
Maintenance and Repairs	Equipment	day	$1,736.31	1	$1,736		
	Building	day	$148.57	1	$149	$1,885	$6
Operating Supplies	Total	day	$282.73	1	$283	$283	$1
Fixed Charges	Taxes	day	$1,092.40	1	$1,092		
	Insurance	day	$546.20	1	$546	$1,639	$5
Contingency	Total	day	$4,989.51	1	$4,990	$4,990	$15
Soil Receiving Building	Total	day	$6,482.69	1	$6,483	$6,483	$19
Total Daily Operating Cost						$31,430	$94

The life-cycle cost estimate for the HGMS process is shown in Table 15. Costs in the first year (year 0) include the FCI estimate of the soil receiving building and the HGMS soil treatment building, as well as the estimated start-up costs for a full-scale facility based on the process design. The yearly operating cost of approximately $11,001,000 in years 1 through 17 for this process is obtained by multiplying the daily operating cost by 350, the estimated number of days the full-scale facility is operated in a single year. The total estimated cost of $154 per ton of soil treated using the HGMS process is based on the estimated life-cycle costs, comprised of the FCI, start-up expenses, and yearly operating costs, as well as the assumption of a 5% annual inflation rate.

11. TIRON SOIL WASHING SYSTEM

The coarse soil separation stage of the tiron system is identical to that of the ABE and carbonate/bicarbonate vat extraction processes. In general, the equipment requirements for the tiron and carbonate/bicarbonate vat extraction processes are very similar. In fact, the engineering designs for the

Table 15. Life-cycle costs for HGMS process

FCI (20 tons/hr)	**$21,267,420**
Start-Up Cost	**$2,126,742**
Yearly Operating Cost	**$11,000,590**
Year	**Cost**
0	$23,394,161
1	$11,000,590
2	$11,550,620
3	$12,128,151
4	$12,734,558
5	$13,371,286
6	$14,039,850
7	$14,741,843
8	$15,478,935
9	$16,252,882
10	$17,065,526
11	$17,918,802
12	$18,814,742
13	$19,755,479
14	$20,743,253
15	$21,780,416
16	$22,869,437
17	$24,012,909
Total Cost	**$307,653,440**
Cost/Ton	**$154**

tiron and carbonate/bicarbonate systems are virtually identical, with the exception of the lixiviant used during the leaching stage and the filtration equipment used in the leaching stage of the process. Appendix A provides the engineering flow diagram for the tiron soil washing system. The equipment costs for the tiron system are shown in Appendix B, Table B.5. The total equipment cost (E) for this system is estimated at approximately $6,085,000. However, this value does not include the cost of any equipment required for effluent treatment. The lixiviant bleed stream at the end of the treatment system must be processed at an effluent treatment plant. The equipment requirements for this system are very similar to the carbonate/bicarbonate vat extraction process. Although the fixed bed ion exchange columns are the same size and type as those required for the carbonate/bicarbonate system, a total of 9 are required for the tiron system, whereas 6 are required for the carbonate system. The total cost for these ion exchange system columns (including resin inventory) is approximately $1,900,000. Rather than using 2 vacuum belt filters as in the carbonate/bicarbonate system, 4 horizontal pressure filters, at a total estimated cost of $1,600,000, are required to filter leach train slurries for the tiron system. Whereas 2 sand filters are required for the carbonate/bicarbonate system, 4 are required for the tiron soil washing system, at a total estimated cost of approximately $400,000. The multi-media sand filters filter suspended solids out of the ion exchange feed material prior to the ion exchange columns. Both the carbonate/bicarbonate, ABE, and tiron systems require 2 drum scrubbers at a total of approximately $240,000.

The results of the FCI calculation for the tiron soil washing system are shown in Table 16. Building size requirements for the tiron soil washing system are the same as those of the carbonate/bicarbonate vat extraction process. Therefore, the building costs for both treatment technologies are the same, estimated at approximately $1,600,000. The direct plant cost for the tiron soil washing system is approximately $17,496,000. Total indirect plant costs, consisting of construction

Table 16. FCI for tiron soil washing system

Item	Symbol	Description	Formula	Value
1	E	Equipment Costs	E	$6,104,560
2	L	Cost of Installation Labor	.40E	$2,441,824
3	IC	Instrumentation & Controls	.13E	$793,593
4	I	Insulation Costs (equip. & piping)	.05E	$305,228
5	P	Piping	.16E	$976,730
6	Q	Labor for Installation of Piping	.40P	$390,692
7	F	Electrical Installations	.15E	$915,684
8	B	Building including Services	----	$1,600,000
9	Y	Yard Improvements	.10E	$610,456
10	S	Service Facilities	.55E	$3,357,508
	D	**Direct Plant Cost**	**Sum of 1-10**	**$17,496,274**
11	ES	Engineering & Supervision	.60E	$3,662,736
12	C	Construction Expenses	.25E	$1,526,140
	IP	**Indirect Plant Cost**	**ES+C**	**$5,188,876**
13	CF	Contractor's Fee	.05(D+IP)	$1,134,258
14	CO	Contingency	.20(D+IP)	$4,537,030
	FCI	**Fixed Capital Investment**	**D+IP+CF+CO**	**$28,356,438**

expenses and engineering and supervision, are $5,189,000. Adding the contractor's fee and contingency of $1,134,00,0 and $4,537,000, respectively, to the direct and indirect plant costs results in an FCI value of approximately $28,356,000 for the tiron soil washing system.

The daily operating costs associated with a full-scale treatment facility based on the tiron soil washing system are shown in Table 17. With the exception of the ABE process, the chemical costs associated with this technology are much higher than those of the carbonate/bicarbonate, HGMS, and heap leaching technologies. Presently, the cost of tiron from distributors ranges from approximately $38 to $41 per 100 grams, or $172 to $186 per pound of tiron. However, the cost of tiron used for these cost estimates is estimated much lower, at $8 per pound. Although tiron is presently not available in bulk quantities, a chemical company representative stated that tiron could be produced in bulk quantities for this price. However, this value is very preliminary because it is based only on phone conversations with the representative, which makes the accuracy of this quote highly uncertain. Based on the unit cost of $8/lb, tiron costs represent over 93% of the total chemical cost for the daily operation of this technology. According to the mass chemcial balance for the tiron soil washing process prepared by Brown & Root, Inc., 2,000 pounds of tiron per day are required to produce a 20 tons per hour soil process rate. Obviously, the cost for bulk quantities of tiron plays a very important role in the operating costs, and thus, the life-cycle costs, associated with this technology.

The labor requirements for the tiron system are the same as that required for the carbonate/bicarbonate system because equipment requirements for both systems are very similar. The labor requirements for this system include personnel requirements for the soil receiving building and the soil treatment building. As with the other treatment technologies requiring the soil receiving building, the daily labor costs for personnel during the first, second, and third shifts, excluding excavation costs ($2.25/ton), is estimated at $12 per ton of soil treated. Like the carbonate/bicarbonate system, the total estimated cost for the tiron soil treatment building personnel during the first shift is $5,100, or $15 per ton of soil treated, assuming a 70% utilization factor. Based on the 70% utilization factor, 336 tons/day, rather than 480 tons/day (20 tons/hour x 24 hrs/day) of contaminated soil can be treated for all of the treatment technologies because all were designed assuming a treatment process rate of 20 tons/hour and a utilization factor of 70%.

The total estimated cost for the second and third shifts for the soil treatment building labor is $4,500 each, or $13 per ton of soil treated (assuming a 70% utilization factor). Therefore, the labor cost for the soil treatment building is approximately $41 per ton of soil treated by the tiron soil washing system. Adding the soil receiving building labor cost of $12/ton to this value results in a total labor cost for the carbonate/bicarbonate vat extraction process of $53/ton of soil treated. Total operating costs for the soil receiving building of $19 per ton were also included in the operating cost estimate since this building is required for this technology. The estimated total daily operating cost of $168 per ton of soil treated for this process includes costs for raw materials, labor, electricity, maintenance and repairs, operating supplies, fixed charges, and contingency.

The life-cycle cost estimate for the tiron system is shown in Table 18. First-year costs (year 0) include the sum of the FCI estimate for the soil receiving building and the tiron soil treatment building ($30,507,000), as well as the estimated start-up costs for a full-scale treatment facility. The yearly operating cost of $19,761,000 in years 1 through 17 for this process is obtained by multiplying the daily operating cost by 350, the estimated number of days the full-scale facility is operated in a single year. The cost to treat the contaminated soil at Fernald using the tiron soil washing process is estimated at $272 per ton. This value is based on the estimated life-cycle costs, comprised of the FCI, start-up expenses, and yearly operating costs, as well as the assumption of a 5% annual inflation rate for the 17 years of operation required to treat the estimated 2,000,000 tons of contaminated soil at the Fernald site.

Table 17. Operating costs for tiron soil washing system

Class	Description	Unit	Unit Cost	Qty/Day	Sub-Total	Total	Cost/Ton
Chemical/Raw Material	Tiron	lbs	$8.00	2000	$16,000		
	Polymer	lbs	$1.80	360	$648		
	Hydrogen Peroxide	lbs	$0.50	101	$51		
	Sodium Chloride	lbs	$0.29	1381	$400		
	Sodium Hydroxide	lbs	$0.15	651	$98		
	Hydrochloric Acid	lbs	$0.04	1381	$59		
	Make-up Water	1000 gal	$0.89	49.8	$44	**$17,300**	**$51**
First Shift	Plant Operators (9)	man-hrs	$34.10	72	$2,455		
	Rad Monitoring (2)	man-hrs	$40.92	16	$655		
	Shift Suprevisors(2)	man-hrs	$45.35	16	$726		
	General Employee	man-hrs	$34.10	8	$273		
	Maintenance Employees(2)	man-hrs	$27.28	16	$436		
	Electrician	man-hrs	$33.19	8	$266		
	Instrument Technician	man-hrs	$33.19	8	$266	**$5,076**	**$15**
Second Shift	Plant Operators(7)	man-hrs	$42.63	56	$2,387		
	Rad Monitoring	man-hrs	$51.15	8	$409		
	Shift Supervisor	man-hrs	$56.69	8	$454		
	General Employee	man-hrs	$42.63	8	$341		
	Maintenance Employees(2)	man-hrs	$34.10	8	$273		
	Electrician	man-hrs	$41.49	8	$332		
	Instrument Technician	man-hrs	$41.49	8	$332	**$4,528**	**$13**
Third Shift	Plant Operators(7)	man-hrs	$42.63	56	$2,387		
	Rad Monitoring	man-hrs	$51.15	8	$409		
	Shift Supervisor	man-hrs	$56.69	8	$454		
	General Employee	man-hrs	$42.63	8	$341		
	Maintenance Employees(2)	man-hrs	$34.10	8	$273		
	Electrician	man-hrs	$41.49	8	$332		
	Instrument Technician	man-hrs	$41.49	8	$332	**$4,528**	**$13**
Electricity	Total	kwh	$0.08	34000	$2,720	**$2,720**	**$8**
Maintenance and Repairs	Equipment	day	$2,824.50	1	$2,824		
	Building	day	$171.43	1	$171	**$2,995**	**$9**
Operating Supplies	Total	day	$449.25	1	$449	**$449**	**$1**
Fixed Charges	Taxes	day	$1,620.37	1	$1,620		
	Insurance	day	$810.18	1	$810	**$2,431**	**$7**
Contingency	Total	day	$10,006.54	1	$10,007	**$10,007**	**$30**
Soil Receiving Building	Total	day	$6,482.69	1	$6,483	**$6,483**	**$19**
Total Daily Operating Cost						**$56,515**	**$168**

12. SUMMARY

To evaluate the economic feasibility of designing, constructing, and operating a full-scale soil treatment facility, engineering designs and cost estimates for 5 treatment technologies to remove uranium from the Fernald soil have been prepared. Engineering designs for the following treatment technologies were jointly developed by the principal investigators of each technology and personnel at Brown & Root, Inc.: the ABE process, carbonate/bicarbonate vat extraction process, heap leaching process, HGMS process, and tiron soil washing system. In conjunction with the engineering design development, cost estimates for each technology were developed by ORNL.

Table 18. Life-cycle costs for tiron soil washing system

FCI	**$30,506,938**
Start-Up Cost	**$3,050,694**
Yearly Operating Cost	**$19,761,078**
Year	**Cost**
0	$33,557,632
1	$19,761,078
2	$20,749,132
3	$21,786,589
4	$22,875,918
5	$24,019,714
6	$25,220,700
7	$26,481,735
8	$27,805,822
9	$29,196,113
10	$30,655,918
11	$32,188,714
12	$33,798,150
13	$35,488,058
14	$37,262,460
15	$39,125,583
16	$41,081,863
17	$43,135,956
Total Cost	**$544,191,136**
Cost/Ton	**$272**

The cost estimates for these treatment technologies have been completed in three steps. The first step was to calculate the FCI, which represents the one-time capital investment needed to supply the necessary full-scale manufacturing and plant facilities for each soil treatment technology. This was accomplished by first estimating the costs for the process equipment (E) required for each technology, and (E) was then used to estimate costs for various components of the FCI. The second step was to develop the operational costs associated with a full-scale treatment facility for each technology. Operating costs occur throughout the life of the project. The third step was to estimate the life-cycle costs to treat the estimated 2 million cubic yards of contaminated soil at the Fernald site using a full-scale treatment facility based on each treatment technology. This report documents this three-step process in developing the cost estimates associated with each treatment technology. For all of the treatment technologies with the exception of the heap leaching process, a soil receiving building to store the contaminated soil before treatment is required. The costs to construct and operate the soil receiving building were included in the cost estimates for these treatment technologies. The total equipment cost, FCI, operating cost per ton of soil treated, and the life-cycle cost to treat a ton of soil for each treatment technology are summarized in Table 19.

Table 19. Summary of treatment technology cost estimates

Treatment Technology	Total Equipment Cost	Fixed Capital Investment	Operating Cost/Ton	Life-Cycle Cost/Ton
High-Gradient Magnetic Separation (HGMS) Process	$4,051,000	$21,267,000	$94	$154
Heap Leaching Process	$3,757,000	$18,721,000	$97	$157
Carbonate/Bicarbonate Vat Extraction Process	$6,151,000	$30,708,000	$113	$189
Tiron Soil Washing System	$6,105,000	$30,507,000	$168	$272
Aqueous Biphasic Extraction (ABE) Process	$4,585,000	$23,572,000	$173	$275

The operating costs for each treatment technology contribute more towards the final life-cycle cost per ton values than the equipment and FCI costs. The operating costs are the cost driver for each technologies' life-cycle costs because the operating costs occur over a 17-year timespan, whereas the equipment and FCI costs only occur in the first year of operation. The 17-year timespan is the estimated amount of time required to treat the 2,000,000 tons of contaminated soil at the Fernald site. However, the importance of the initial costs for equipment and FCI requirements for a full-scale treatment facility should not be diminished because many of the cost categories for the daily operating costs are based on these initial costs.

The estimated equipment cost and FCI for the heap leaching process are the lowest of the treatment technologies considered. Even though the HGMS process has higher total equipment and FCI costs compared to the heap leaching process, the HGMS process is the most cost effective option to treat the uranium-contaminated soil at Fernald. The HGMS process, with a life-cycle cost of $154/ton, is the lowest of any of the technologies because it has the lowest operating costs. The life-cycle cost for the heap leaching process is only slightly higher than the HGMS process, at $157/ton of soil treated. Even though the HGMS process has the lowest life-cycle cost of the treatment technologies, more uncertainty about implementation to a full-scale facility to treat the Fernald soils exists for this technology. Limited data exists on the applicability of the HGMS process to the type of soil at the Fernald site.

From a cost perspective, the most attractive option after the HGMS and heap leaching processes is the carbonate/bicarbonate vat extraction process, at an estimated life-cycle cost of $189/ton. Even though the equipment and FCI requirements for the carbonate/bicarbonate system are higher than the ABE and tiron processes, it's life-cycle costs are lower than those of the ABE and tiron processes due to lower operating costs. Daily operating costs for the ABE and tiron soil washing processes are approximately 53% and 49% higher, respectively, than those of the carbonate/bicarbonate vat extraction process. This is primarily due to higher chemical costs associated with the ABE and tiron soil washing processes. The chemical costs used in the cost estimate for the tiron soil washing process may be somewhat optimistic because tiron is presently not produced in bulk quantities, and tiron costs are approximately $172 to $186 per pound. The cost of tiron used for the cost estimate in this report is $8 per pound, based on phone conversations with a chemical company representative who felt that his company could produce tiron in bulk quantities for this price. Although the total equipment and FCI requirements are less for the ABE process compared to the tiron system, the life-cycle costs for the 2 treatment technologies are approximately the same. The estimated operating costs for the ABE process are slightly higher than the operating costs for the tiron soil washing system. The life-cycle costs for a full-scale facility based on the tiron soil washing and ABE processes are $272 and $275 per ton, respectively.

With the exception of building costs associated with each treatment technology, the FCI calculations in this report were estimated as a percentage of the process equipment cost for each process. Building costs were based on unit costs and estimates of square footage requirements provided by Brown & Root, Inc. personnel. The FCI requirements, startup expenses, operational costs, and life-cycle costs for a full-scale treatment facility were utilized to determine the treatment cost per ton of soil treated for each treatment technology, based on the estimated quantity of contaminated soil at the Fernald site. The engineering flow diagrams and the cost estimates contained in this report are based on bench-scale study results and are considered preliminary because of the research and developmental nature associated with these 5 technologies. These cost estimates are defined as study estimates and are based on the knowledge of major items of equipment associated with each treatment technology. Study estimates normally have an accuracy of plus or minus 30%. However, the positive spread may be wider than the negative in this report due to the experimental nature of this project, as well as preliminary design requirements and specifications for the processes. Therefore, the plus or minus 30% accuracy rate for our study estimate may in fact be +40% to -20%.

13. REFERENCES

Henderson, W.C. *Fernald ID Technologies, Preliminary Design Input, Revision 2 Process Flowsheets*, January 16, 1995.

Means Site Work and Landscape Cost Data, R.S. Means Company, Inc., 12th Annual Edition, 1993.

Mining and Mineral Processing Equipment Costs and Preliminary Capital Cost Estimations, Published by the Canadian Institute of Mining and Metallurgy, Volume 25.

Peters, M.S. and K.D. Timmerhaus. 1991. *Plant Design and Economics for Chemical Engineers*. New York: McGraw-Hill, Inc.

Post, R.G. and M.E. Wacks, editors. 1994. Uranium soils integrated demonstration 1993 status. "Proceedings of the Symposium on Waste Management, Tucson AZ, February 27-March 3, 1994". Vol 3, pp. 2051-2135.

Process Plant Construction Estimating Standards, Richardson's Engineering Services, Inc., 1994 Edition.

Sladic, Mark. *Fernald ID Technologies, Preliminary Design Input, High-Gradient Magnetic Separation Process Flowsheet,* August 4, 1995.

Soil Decontamination Task Group. 1993. Removal of Uranium from Uranium-Contaminated Soils, Phase 1: Bench Scale Testing. ORNL-6762. Oak Ridge National Laboratory, Oak Ridge, Tennessee.

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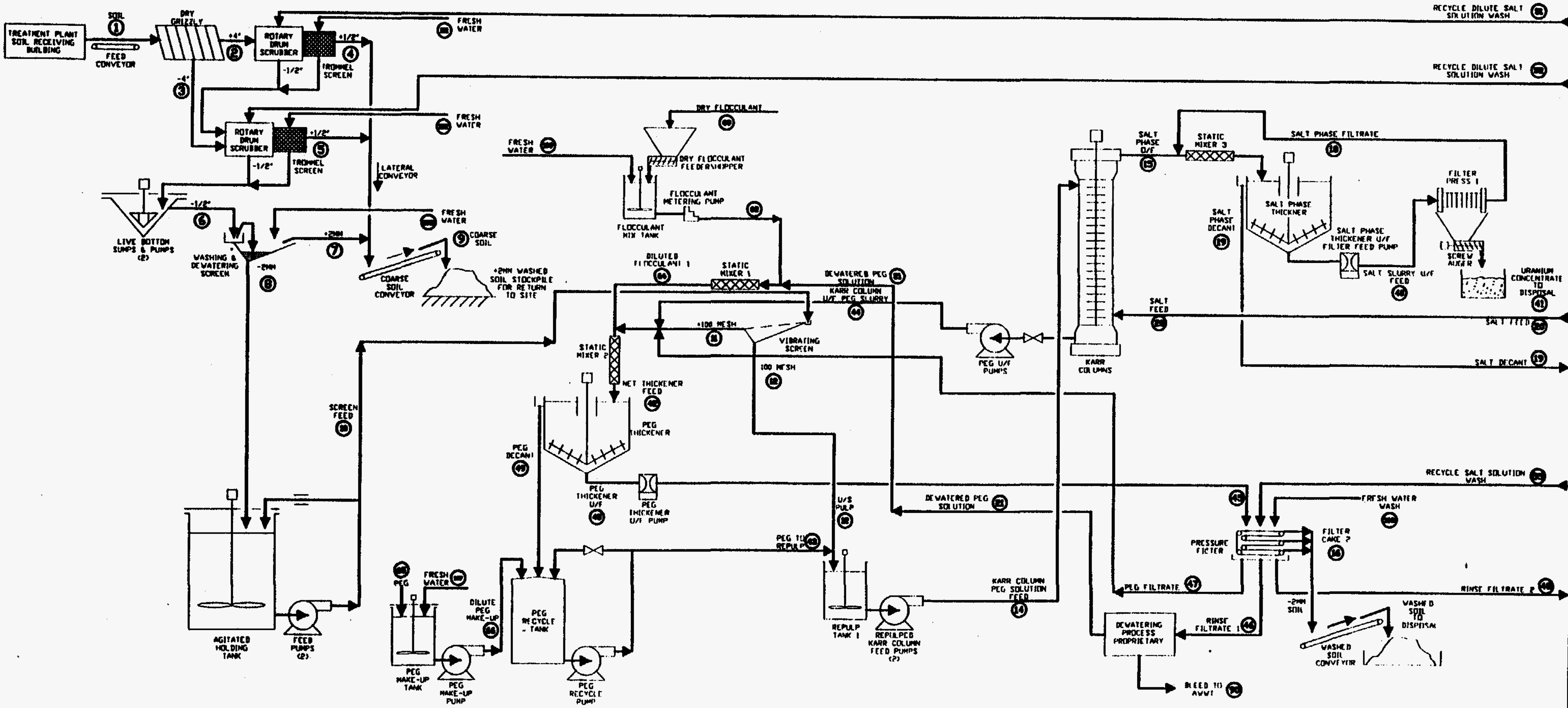
48. R.E. Swatzell, HAZWRAP, P.O. Box 2003, Oak Ridge, TN 37831-7606
49. V.C. Tidwell, Sandia National Laboratories, MS 6315, P.O. Box 5800, Albuquerque, NM 87185
50. J. Walker, U.S. Department of Energy, 19901 Germantown Road, Germantown, MD 20874-1290
51. Office of Assistant Manager for Energy Research and Development, U.S. Department of Energy, Oak Ridge Operations Office, P.O. Box 200, Oak Ridge, TN 37831-8600
52-53. Office of Scientific & Technical Information, P.O. Box 62, Oak Ridge, TN 37831

APPENDIX A

ENGINEERING FLOW DIAGRAMS FOR SOIL TREATMENT TECHNOLOGIES

DEVELOPED BY BROWN & ROOT, INC.



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L/S SEPARATION AND URANIUM CONCENTRATION
SHEET 1 OF 2

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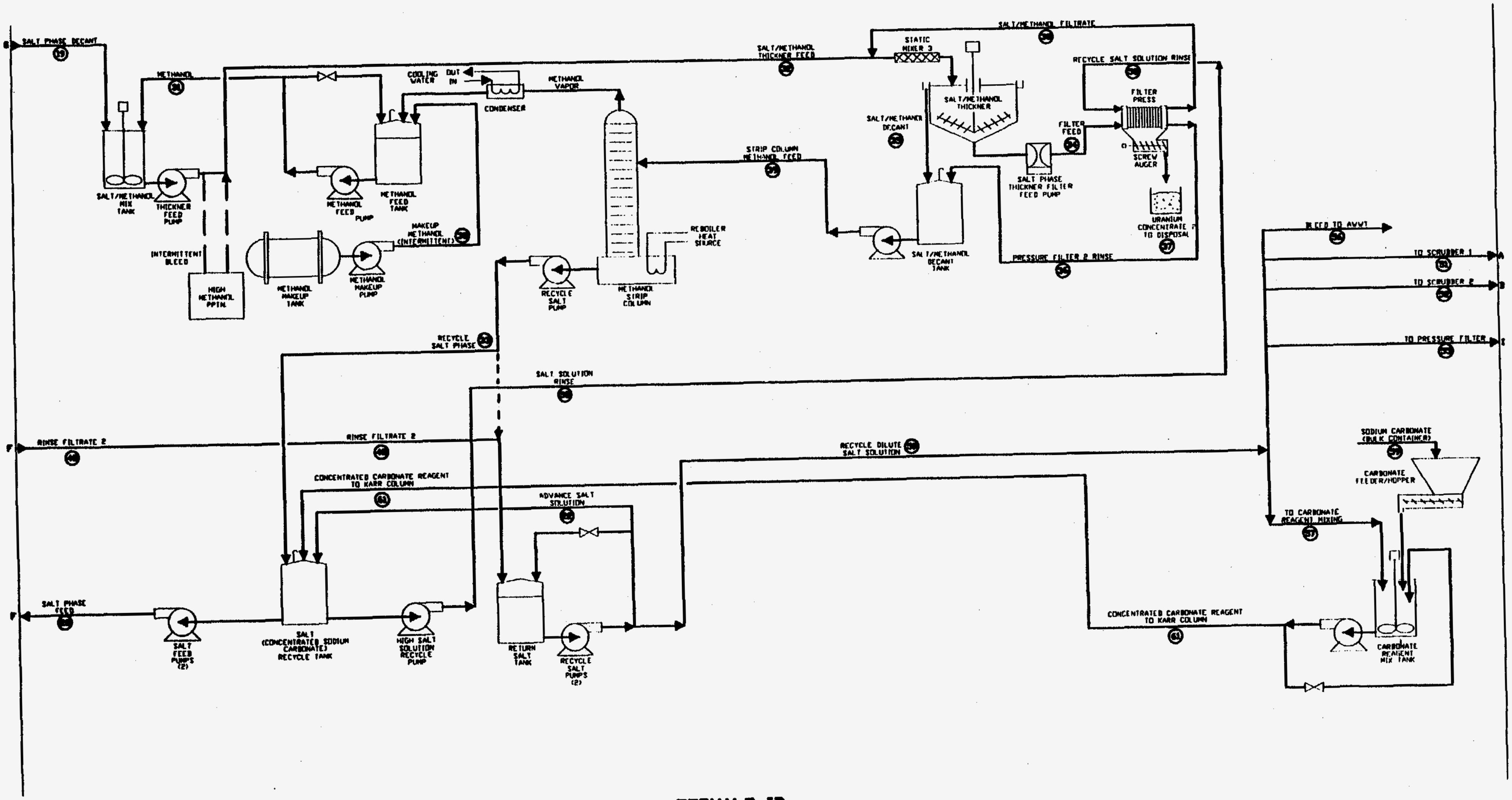

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SOLUBILIZED URANIUM REMOVAL REAGENT RECYCLE
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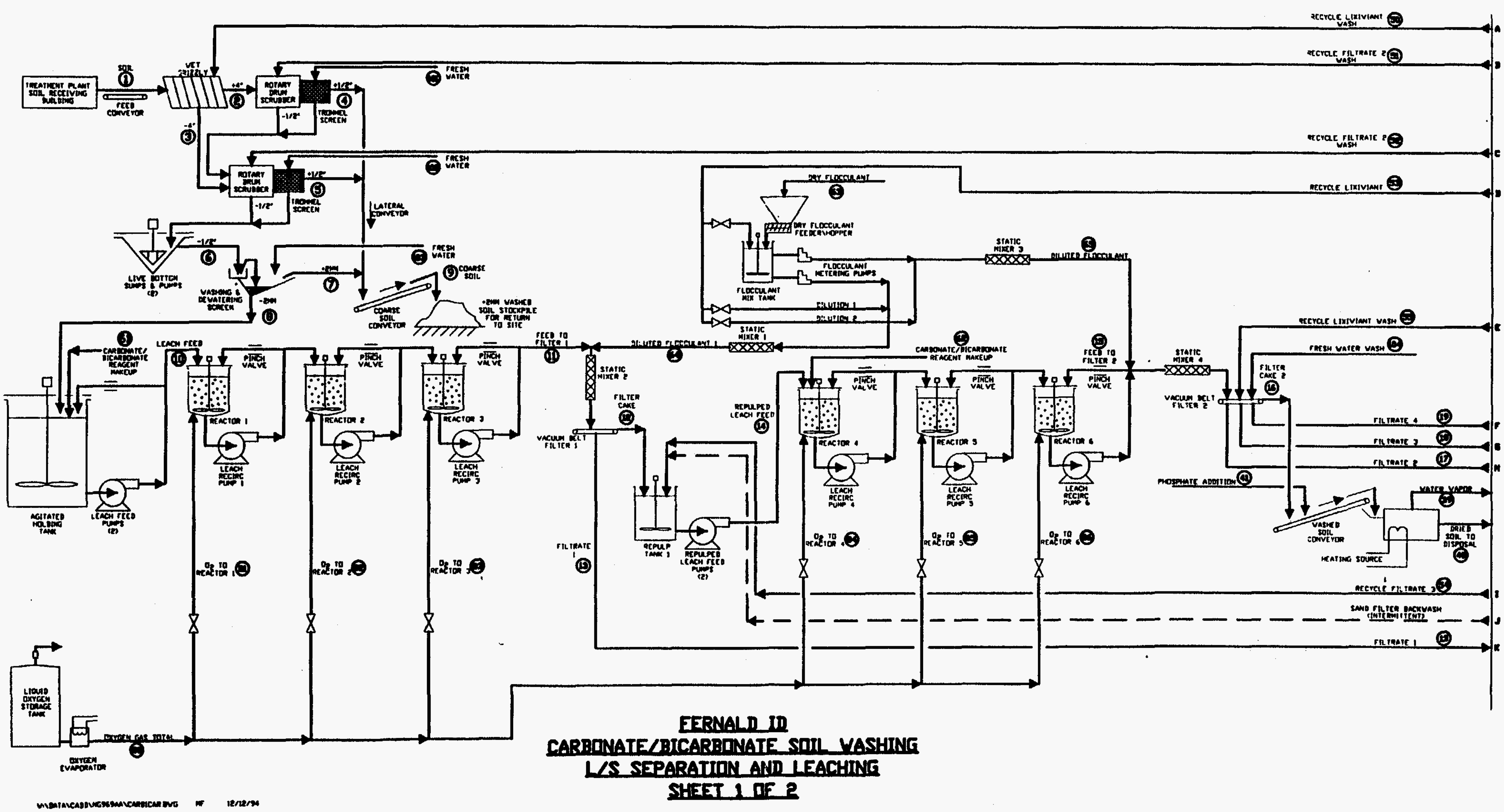

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L/S SEPARATION AND LEACHING
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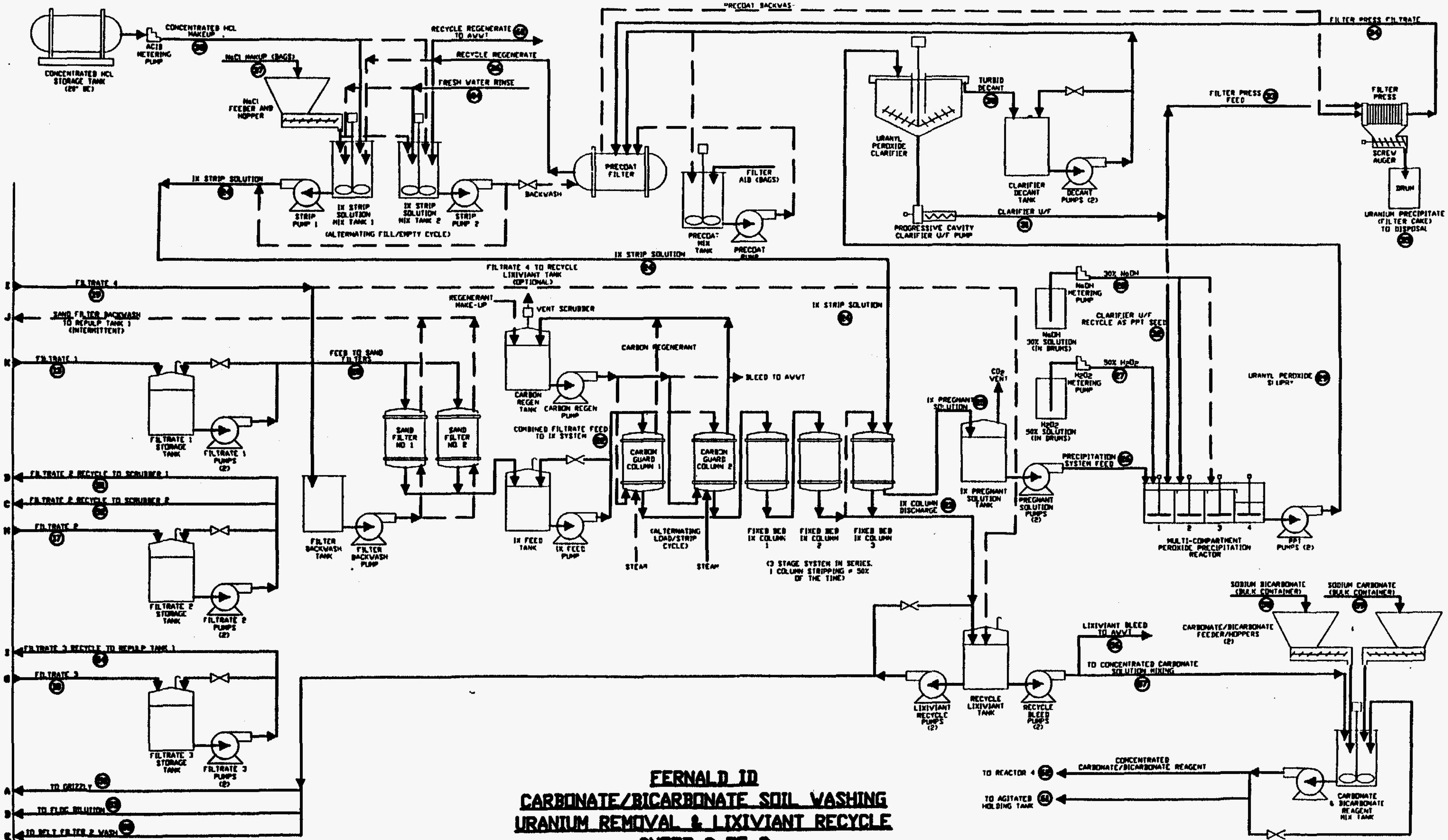

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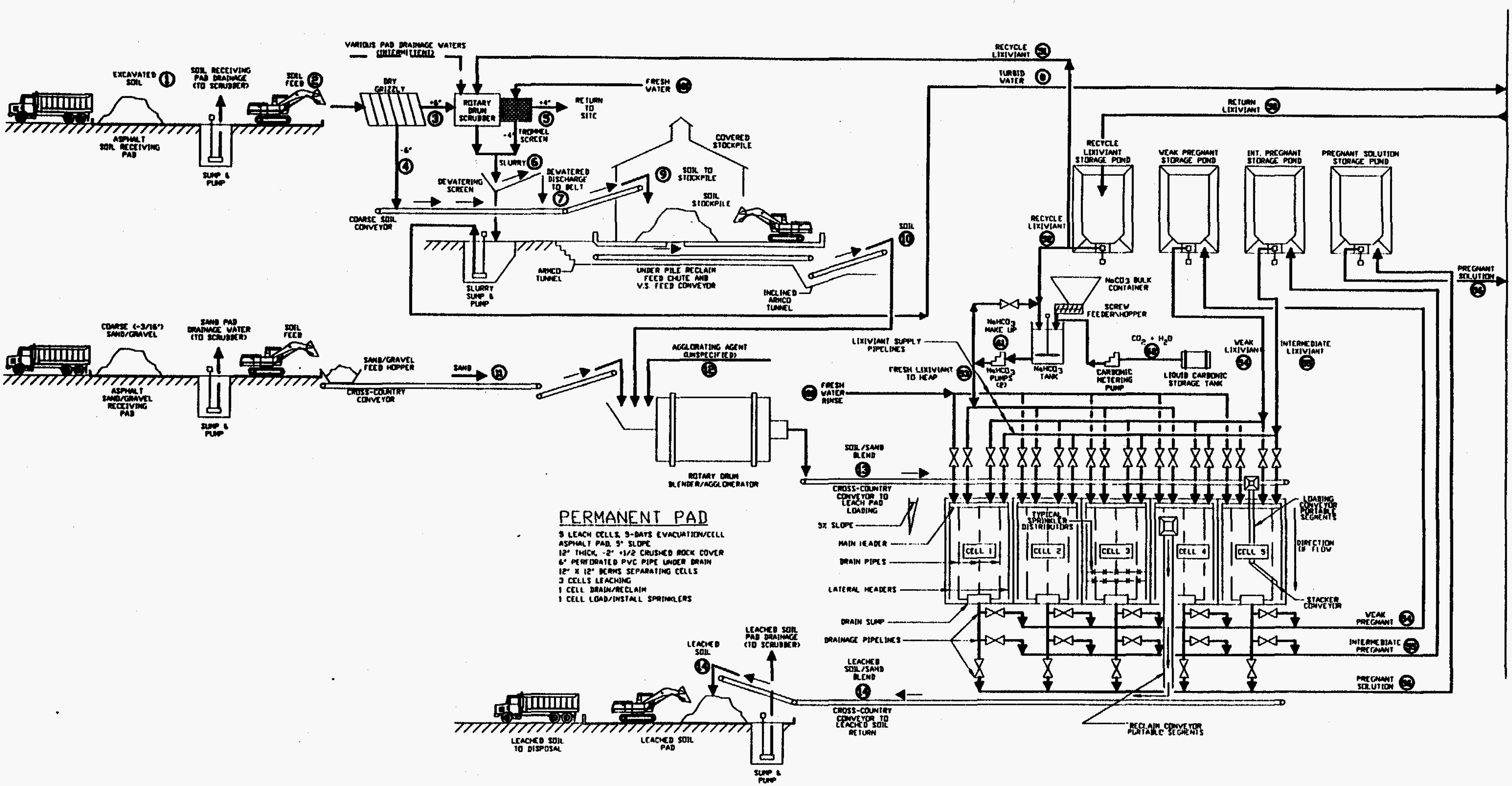

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CONVENTIONAL PERMANENT PAD, MOVABLE HEAP
SOIL HANDLING AND LEACHING
SHEET 1 OF 2

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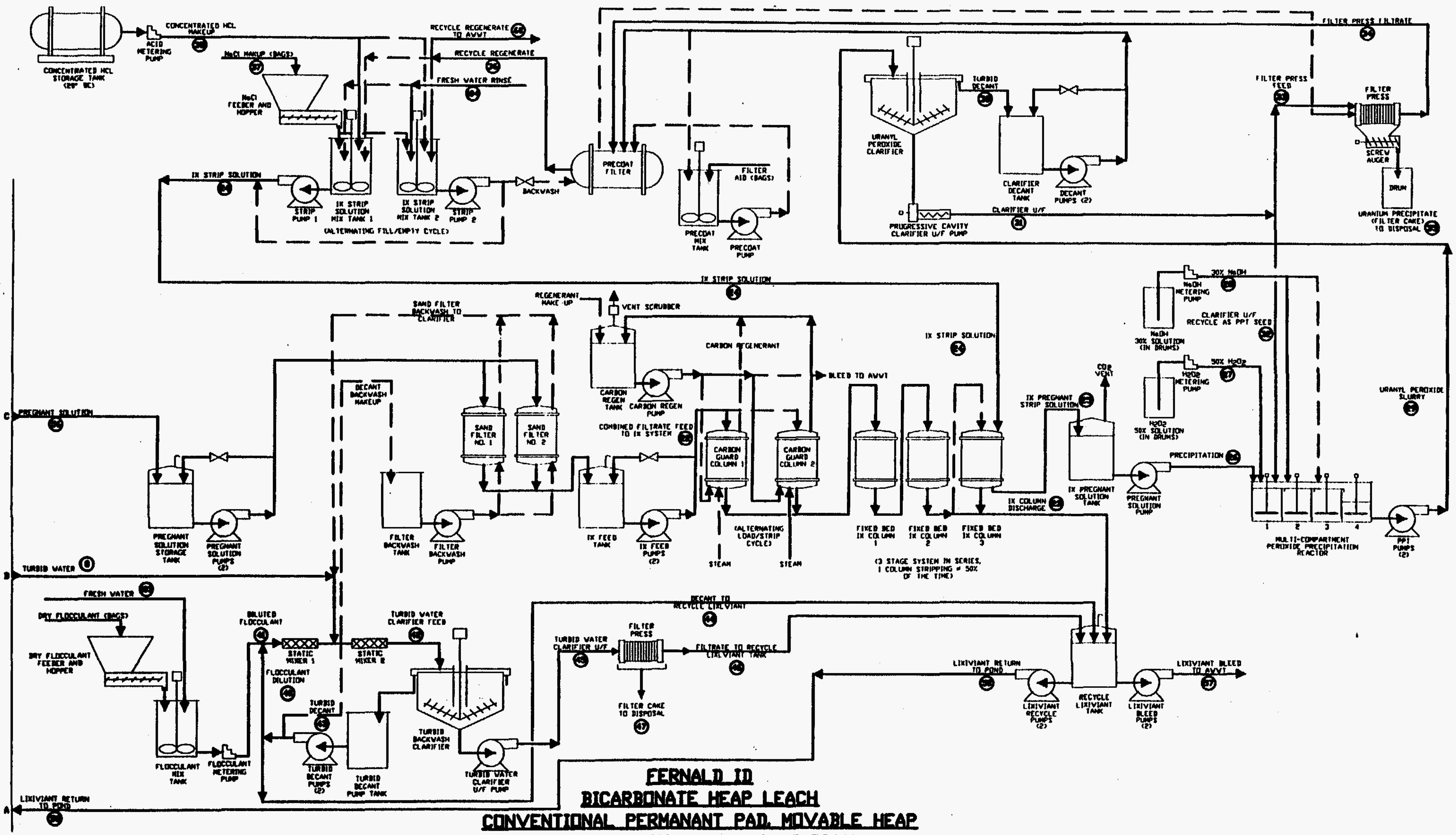

FERNALD ID

BICARBONATE HEAP LEACH

CONVENTIONAL PERMANANT PAD, MOVABLE HEAP

URANIUM REMOVAL & LIXIVIANT RECYCLE

SHEET 2 OF 2

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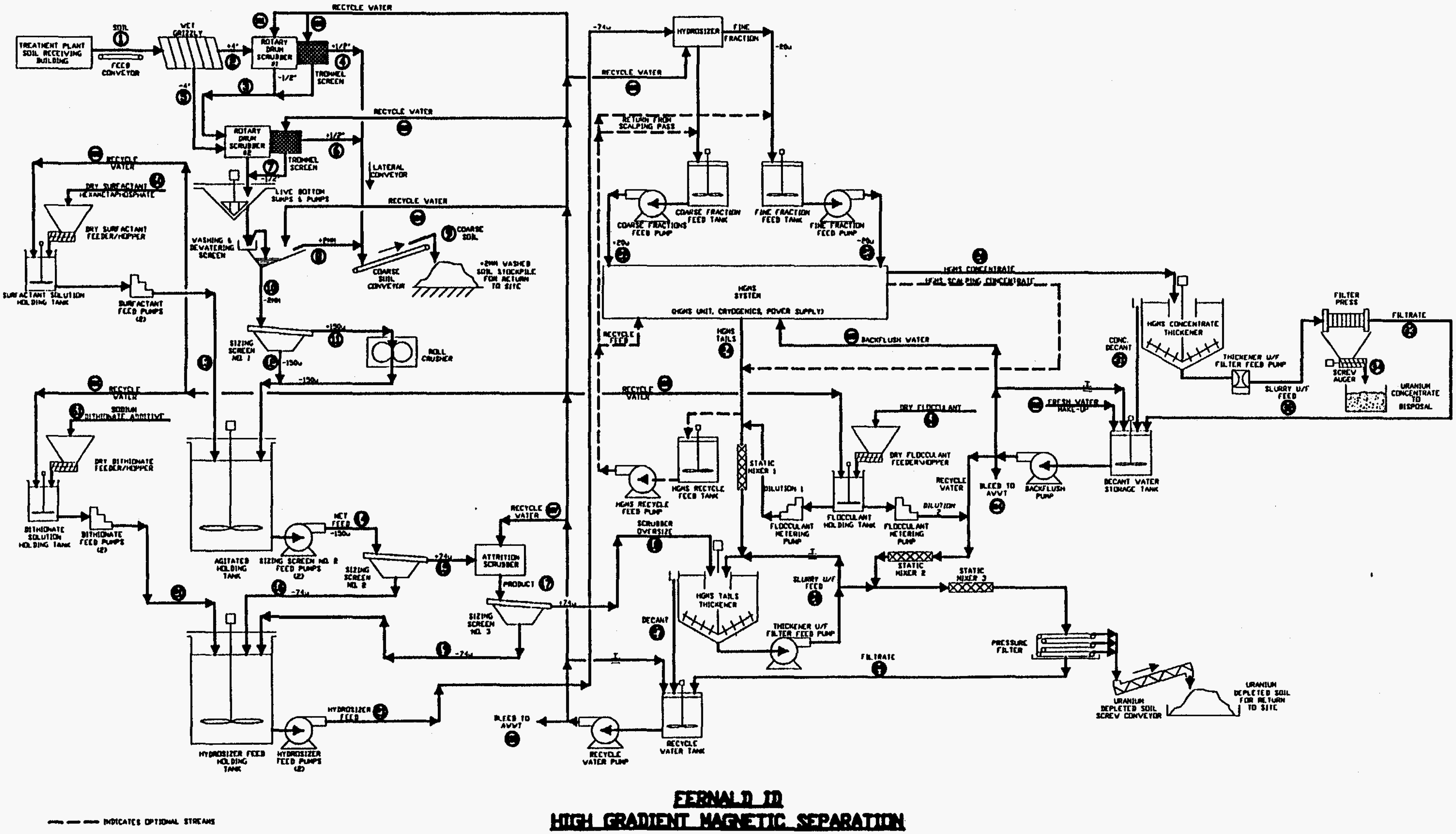

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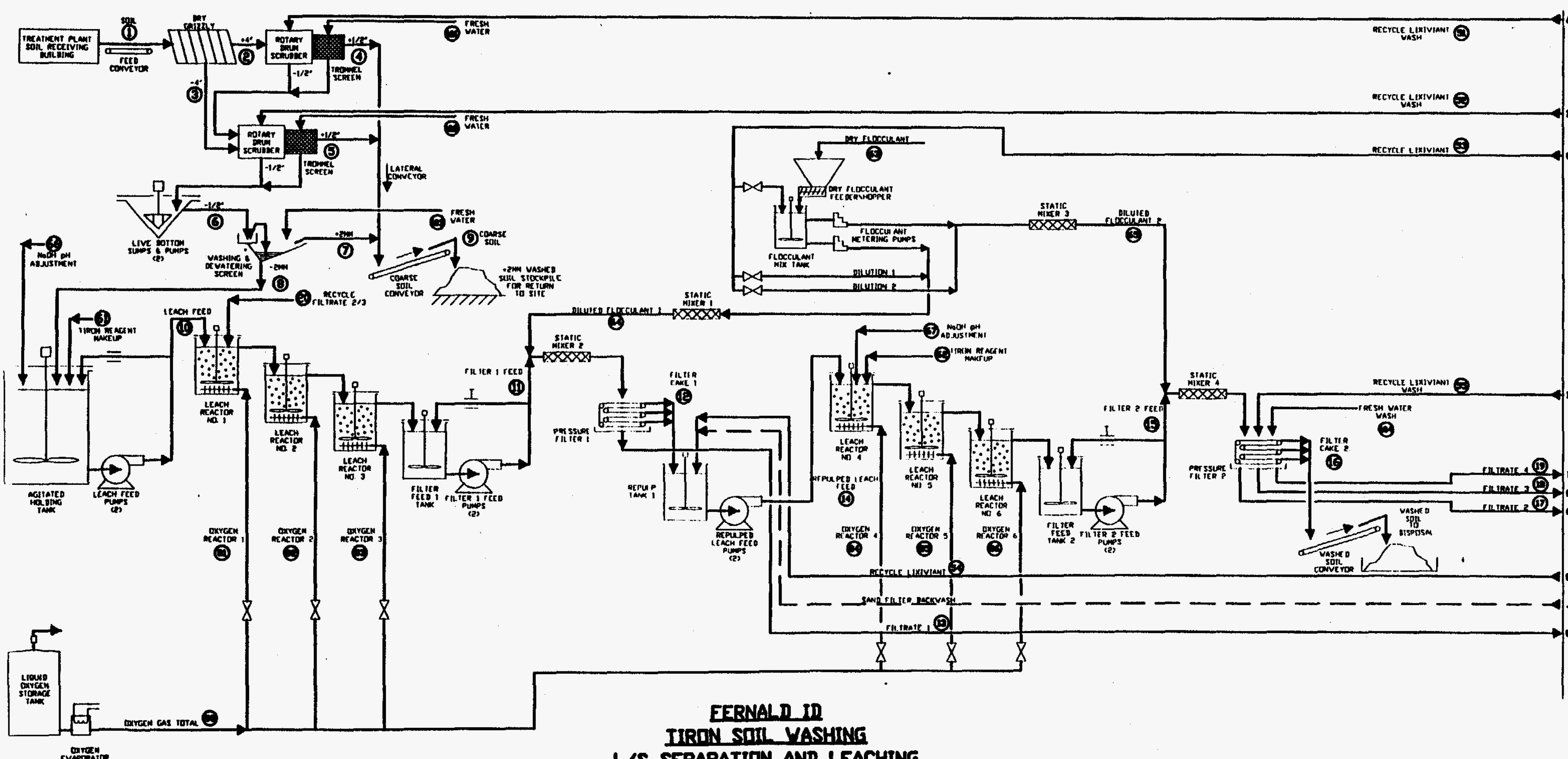

FERNALD ID
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L/S SEPARATION AND LEACHING
SHEET 1 OF 2

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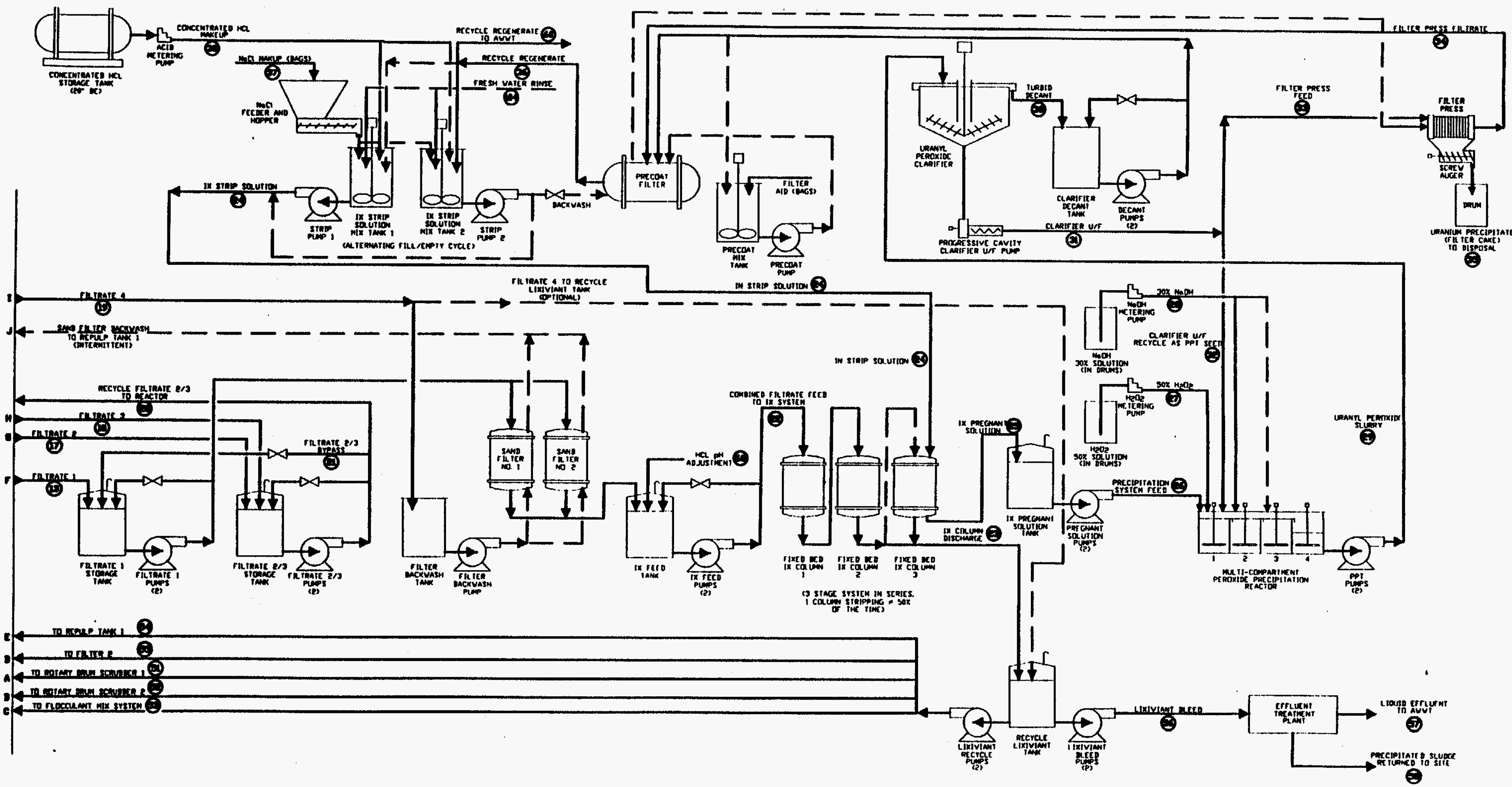

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TIRON SOIL WASHING
URANIUM REMOVAL & LIXIVIANT RECYCLE
SHEET 2 OF 2

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APPENDIX B

EQUIPMENT COST ESTIMATES FOR SOIL TREATMENT TECHNOLOGIES

Description of source column for the tables in Appendix B:

1) Vendor information or quote

2) *Process Plant Construction Estimating Standards*, Richardson's Engineering Services, Inc., 1994 Edition.

3) Mining and Mineral Processing Equipment Costs and Preliminary Capital Cost Estimations, Published by the Canadian Institute of Mining and Metallurgy, Volume 25.

4) *Means Site Work and Landscape Cost Data*, R.S. Means Company, Inc., 12th Annual Edition, 1993.

5) Cost estimating personnel at Brown & Root, Inc.

6) Plant Design and Economics for Chemical Engineers, 4th Edition, Max S. Peters and Klaus D. Timmerhaus, 1991.

Table B.1
Equipment List for
Aqueous Biphasic Extraction Process

Item	Source	Item Description	Cost/Unit	Quantity	Total Cost
1	1	Grizzly Feed Conveyor (@ $600/ft)	$60,000	1	$60,000
2	1	Dry grizzly	$40,000	1	$40,000
3	1	Drum Scrubber 1 with trommel screen	$90,000	1	$90,000
4	1	Drum Scrubber 2 with trommel screen	$150,000	1	$150,000
5	1	O/S Soil Transport Conveyor (@ $600/ft)	$60,000	1	$60,000
6	1	Washing & Dewatering Screen (6' x 8' screen deck)	$40,000	1	$40,000
7	1	Slurry Pump and Sump (1 pump & 1 standby, 150 gpm & 50 TDH)	$7,000	2	$14,000
8	5	Agitated Holding Tank (15,000 gallon carbon steel shell)	$30,000	1	$30,000
9	1	Holding Tank Mixer (15 HP motor, 72 in dia. impellor)	$14,600	1	$14,600
10	1	Feed Pumps (centrifugal, 250 gpm max. output, 65 ft TDH, 20 HP)	$9,000	2	$18,000
11	1	Vibrating Screen (48 in. x 96 in. with 3.5 HP motor)	$40,000	1	$40,000
12	1	Static Mixer 1 (1.5 in. x 18 in., 316 SS shell, 50 gpm)	$1,500	1	$1,500
13	1	Static Mixer 2 (4 in. x 24 in., 316 SS, 300 gpm)	$5,000	1	$5,000
14	1	Dry Flocculant Hopper/Feeder (.65 lb floc/ton soil solids, 5 cu ft hopper)	$15,000	1	$15,000
15	1	Flocculant Mix Tank (1000 gallons, 6' dia. x 6 ft h, 316 SS, mixer bridge)	$4,000	1	$4,000
16	1	Flocculant Mix Tank Agitator (5 HP motor, 30 in impellor, variable speed)	$10,000	1	$10,000
17	1	Flocculant Metering Pump (dual diaphragm, 0-160 gpm, 2 HP var. motor)	$17,500	1	$17,500
18	1	PEG Thickener (250 gpm feed, 28 ft dia., 10 ft high, with rake, 15 HP)	$150,000	1	$150,000
19	5	PEG Makeup Tank (6' x 8' cyl. tank, 1000 gal., 316 SS, mixer bridge)	$4,000	1	$4,000
20	1	PEG Makeup Tank Agitator (5 HP var. sp. motor, 30 in dia impellor)	$10,000	1	$10,000

Table B.1
Equipment List for
Aqueous Biphasic Extraction Process

Item	Source	Item Description	Cost/Unit	Quantity	Total Cost
21	1	PEG Makeup Pump (centrifugal, 200 gpm, 65' head, 15 HP)	$5,000	1	$5,000
22	5	PEG Recycle Tank (9,000 gallons max., 4,500 gpm WV, 12'D x 12' H)	$18,000	1	$18,000
23	1	PEG Recycle Pumps (centrifugal, 225 gpm, 65' head, 15 HP)	$5,000	2	$10,000
24	5	PEG Repulp Tank (6' x 8' cyl. tank, 1000 gal. WV, 316 SS, mixer bridge)	$4,000	1	$4,000
25	1	Karr Columns (6' D x 10' H, rotating 5 HP drive, 340 gpm net flow)	$380,000	5	$1,900,000
26	1	Karr Column PEG U/F Pumps (centrifugal, 350 gpm, 65' head, 20 HP)	$9,000	2	$18,000
27	1	Static Mixer 3 (150 gpm max, 316 SS)	$4,500	1	$4,500
28	1	Salt Phase Thickener (20' dia., 8' sidewalls, rake, 10 HP)	$95,000	1	$95,000
29	1	PEG Thickener Underflow Pumps (dual diaphragm, 0-75 gpm, 100 psig)	$5,000	2	$10,000
30	1	Salt Thickener Underflow Pumps (dual diaphragm, 0-20 gpm, 100 psig)	$2,500	2	$5,000
31	1	Recessed Plate Concentrate Filter (plate filter press, 27.3 cu ft capacity)	$40,000	1	$40,000
32	1	Horizontal Pressure Filters (LAROX PF-16, belt type, 409 sq ft FA, 25 HP)	$400,000	2	$800,000
33	1	Washed Soil Conveyor (@ $600/ft) (24 in x 100 L, 4 in sidewalls)	$60,000	1	$60,000
34	5	Salt/Methanol Mix Tank (8' x 8' cyl. tank, 2000 gallon WV, 316SS)	$8,000	1	$8,000
35	1	PEG Repulp Tank Agitator (10 HP motor, 42 in dia. impellor)	$13,500	1	$13,500
36	1	Salt/Meth. Thickener Feed Pumps (centrifugal, 350 gpm max., 65' head)	$9,000	2	$18,000
37	1	Static Mixer 4 (250 gpm max, 316 SS)	$4,500	1	$4,500
38	1	Salt/Methanol Thickener (20' dia, 8' sidewalls, rake, 10 HP)	$95,000	1	$95,000
39	1	Salt/Methanol Thickener Underflow Pumps (dual diaphragm, 0-20 gpm)	$2,500	2	$5,000
40	1	Recessed Plate Precipitate Filters (plate filter press, 54.6 cu ft capacity)	$50,000	2	$100,000

Table B.1
Equipment List for
Aqueous Biphasic Extraction Process

Item	Source	Item Description	Cost/Unit	Quantity	Total Cost
41	1	Air Compressors (150 psig, 50 HP)	$48,500	2	$97,000
42	5	Methanol Feed Tank (cyl. 15000 gallon max, carbon steel)	$30,000	1	$30,000
43	1	Methanol Precip. Feed Pumps (centrifugal, 300 gpm, 65' head, 15 HP)	$6,000	2	$12,000
44	1	Methanol Makeup Pump (centrifugal, 300 gpm, 65' head, 15 HP)	$6,000	1	$6,000
45	5	Methanol Makeup Tank (3000 gallon tank truck trailer, 316 SS)	$13,000	1	$13,000
46	5	Salt/Methanol Decant Tank (15000 gallon, 14'D x 16'H, carbon steel)	$30,000	1	$30,000
47	1	Methanol Strip Column Feed Pumps (centrifugal, 300 gpm, 65' head)	$6,000	2	$12,000
48	5	Methanol Strip Column (10'D x 32' H, 16 trays, 2000 gallon reboiler)	$200,000	1	$200,000
49	1	Recycle Salt Pumps (centrifugal, 300 gpm, 100' head, 20 HP)	$6,000	2	$12,000
50	1	Recycle Salt Solution Pumps (centrifugal, 300 gpm, 100' head, 20 HP)	$6,000	2	$12,000
51	1	High Solution Recycle Pump (centrifugal, 300 gpm, 100' head, 20 HP)	$6,000	1	$6,000
52	1	Karr Column Salt Solution Feed Pumps (centrifugal, 300 gpm, 100' head)	$6,000	2	$12,000
53	1	Concentrated Carbonate Reagent Pump (centrifugal, 300 gpm, 65' head)	$6,000	1	$6,000
54	5	Salt (High-Carbonate) Recycle Tank (12000 gallon max., 14' D x 16' H, CS)	$25,000	1	$25,000
55	5	Return Salt (Low-Carbonate) Tank (12000 gallon max., 14'D x 16'H, CS)	$25,000	1	$25,000
56	5	Carbonate Makeup Tank (6'D x 8'H cyl, 1000 gallon WV, 316 SS, impellor)	$4,000	1	$4,000
57	1	Carbonate Reagent Tank Agitator (5 HP motor, 30 in dia. impellor)	$10,000	1	$10,000
58	1	Bulk Carbonate Storage Containers (316 SS, 6 cu yd cap., conical dis.)	$3,500	12	$42,000
59	1	Bulk Carbonate Screw Feeder (3.2-160 cu ft/hr, timer, 5 cu ft hopper)	$15,000	1	$15,000
60	1	PEG Repulp Tank Agitator (7.5 HP, 30 in dia. impellor)	$11,000	1	$11,000

Table B.1
Equipment List for
Aqueous Biphasic Extraction Process

Item	Source	Item Description	Cost/Unit	Quantity	Total Cost
61	1	Repulped Karr Column Feed Pumps (centrifugal, 350 gpm max, 65' head)	$9,000	2	$18,000
		Total Equipment Cost			$4,585,100

Table B.2
Equipment List for
Carbonate/Bicarbonate Vat Extraction Process

Item	Source	Item Description	Cost/Unit	Quantity	Total Cost
1	1	Grizzly Feed Conveyor (24 in W x 100 ft L)	$60,000	1	$60,000
2	1	Wet Grizzly (6' x 8' sloped entry)	$45,000	1	$45,000
3	1	Drum Scrubber 1 (4' D x 6' L, 6' trommel section)	$90,000	1	$90,000
4	1	Drum Scrubber 2 (6' D x 8'L, 6' trommel section)	$150,000	1	$150,000
5	1	O/S Soil Transport Conveyor (24 in W x 100 ft L)	$60,000	1	$60,000
6	1	Washing & Dewatering Screen (6' x 8' screen deck)	$40,000	1	$40,000
7	1	Slurry Pump and Sump (300 gpm, 50' TDH, 2 cu yd capacity sump)	$22,500	2	$45,000
8	5	Agitated Holding Tank (15000 gallon max volume, carbon steel)	$25,000	1	$25,000
9	1	Holding Tank Agitator (15 HP motor, 56 rpm, 6' dia impellor)	$14,600	1	$14,600
10	1	Leach Feed Pumps (centrifugal, 325 gpm max, 65' TDH, 20 HP)	$9,500	2	$19,000
11	5	Reactor Tanks 1-3 (10' dia x 12' H, 5500 gal WV, carbon steel)	$11,000	3	$33,000
12	1	Reactor Tank Agitators (10 HP, 56 rpm, 54 in dia impellor)	$12,500	3	$37,500
13	1	Leach Recirculation Pumps (centrifugal, 325 gpm max, 65' TDH, 20 HP)	$9,500	6	$57,000
14	1	Static Mixer 1 (2 in x 18 in, 316 SS, 100 gpm)	$2,500	1	$2,500
15	1	Static Mixer 2 (4 in x 24 in, 350 gpm max, 316 SS)	$5,000	1	$5,000
16	1	Dry Flocculant Hopper/Feeder (5 cu ft hopper)	$15,000	1	$15,000
17	5	Flocculant Mix Tank (8' dia x 10' H cyl. tank, 2000 gal. WV, 316 SS)	$7,000	1	$7,000
18	1	Flocculant Mix Tank Agitator (7.5 HP, 48 in dia. impellor)	$11,000	1	$11,000
19	1	Flocculant Metering Pumps (dual diaphragm, 0-210 gph, 2 HP drive)	$17,500	2	$35,000
20	1	Vacuum Belt Filter 1 (450 sq ft filter area, pumps vacuum sys., feed dist.)	$750,000	1	$750,000

Table B.2
Equipment List for
Carbonate/Bicarbonate Vat Extraction Process

Item	Source	Item Description	Cost/Unit	Quantity	Total Cost
21	5	Repulp Tank 1 (6' x 8' cyl. tank, 1000 gal. WV, 316 SS)	$4,000	1	$4,000
22	1	Repulp Tank 1 Agitator (7.5 HP, 30 in dia. impellor, 0-100 rpm)	$11,000	1	$11,000
23	1	Repulped Leach Feed Pumps (centrifugal, 325 gpm max output, 65' TDH)	$9,500	2	$19,000
24	5	Reactor Tanks 4-6 (10' D x 12' H, 6000 gal. WV, carbon steel)	$10,000	3	$30,000
25	1	Reactor Tank Agitators (10 HP motor, 56 rpm, 54 in dia. impellor)	$12,500	3	$37,500
26	1	Leach Recirculation Pumps (centrifugal, 325 gpm, 65' TDH, 20 HP)	$9,500	6	$57,000
27	1	Static Mixer 3 (2 in x 18 in, 316 SS, 100 gpm)	$2,500	1	$2,500
28	1	Static Mixer 4 (4 in x 24 in, 350 gpm, 316 SS)	$5,000	1	$5,000
29	1	Vacuum Belt Filter 2 (450 sq ft filter area, 2 additional vacuum chambers)	$1,100,000	1	$1,100,000
30	1	Washed Soil Conveyor (24 in W x 100 ft L)	$60,000	1	$60,000
31	5	Washed Soil Rotary Dryer (8' x 20' long rotary dryer)	$200,000	1	$200,000
32	5	Liquid Oxygen Storage Tank & Evaporator (Leased, covered in oper. cost)	$0	1	$0
33	5	Filtrate 1 Storage Tanks (14' D x 20' H cyl tank, 20000 gal. WV, carbon steel)	$27,000	2	$54,000
34	5	Filtrate 2 Storage Tank (14' D x 20' H cyl tank, 20000 gal. WV, carbon steel)	$27,000	1	$27,000
35	5	Filtrate 3 Storage Tank (14' D x 20' H cyl tank, 20000 gal. WV, carbon steel)	$27,000	1	$27,000
36	5	Recycle Lixiviant Storage Tanks (14' D x 20' H cyl tank, 20000 gal. WV, CS)	$27,000	2	$54,000
37	5	Filter Backwash Tank (10' D x 10' H cyl tank, 4500 gal. WV, carbon steel)	$10,000	1	$10,000
38	1	Sand Filters (10' D x 10' H cyl tank, carbon steel)	$99,000	2	$198,000
39	1	Filtrate 1 Feed Pumps (centifugal, 400 gpm max., 65' TDH, 20 HP)	$12,500	2	$25,000
40	1	Filtrate 2 Recycle Pumps (centifugal, 200 gpm max, 65' TDH, 20 HP)	$10,000	2	$20,000

Table B.2
Equipment List for
Carbonate/Bicarbonate Vat Extraction Process

Item	Source	Item Description	Cost/Unit	Quantity	Total Cost
41	1	Filtrate 3 Recycle Pumps (centifugal, 200 gpm max, 65' TDH, 10 HP)	$10,000	2	$20,000
42	1	Recycle Lixiviant Pumps (centrifugal, 400 gpm max, 65' TDH)	$12,500	2	$25,000
43	1	Recycle Bleed Pumps (centrifugal, 150 gpm max, 10 HP)	$10,000	2	$20,000
44	5	IX Feed Tank (14' D x 20' H cyl tank, 20000 gallon WV, carbon steel)	$27,000	1	$27,000
45	1	IX Feed Pumps (centrifugal, 400 gpm max, 20 HP)	$13,000	2	$26,000
46	1	Carbon Guard Columns (2 trains, 2 stages/train, 10' D x 10' H cyl tank)	$200,000	2	$400,000
47	1	Ion Exchange Columns (10' D x 10' H cyl tank, including resin, 65 psig)	$214,000	6	$1,284,000
48	1	Air Compressors (50 HP, 150 psig)	$48,500	2	$97,000
49	5	Carbon Regenerate Tanks (15000 gal. max volume, carbon steel)	$30,000	2	$60,000
50	1	Carbon Regenerate Pumps (centrifugal, 300 gpm, 65' TDH)	$9,500	2	$19,000
51	5	IX Pregnant Solution Tanks (14' D x 16' H, 12000 gal. WV, 316 SS)	$54,000	2	$108,000
52	5	Concentrated HCl Storage Tank (3000 gallon tank truck trailer)	$20,000	1	$20,000
53	1	Acid Metering Pumps (dual diaphragm, 0-160 gph, 2 HP drive)	$20,000	2	$40,000
54	1	Peroxide Metering Pumps (dual diaphragm, 0-12 gph, .75 HP drive)	$10,000	2	$20,000
55	1	NaOH Metering Pumps (dual diaphragm, 0-12 gph, .75 HP drive)	$10,000	2	$20,000
56	1	Peroxide Precipitation Reactor Tanks (386 gal. WV, 48 sq in x 54 in H, HDPE)	$3,500	5	$17,500
57	1	Peroxide Precip. Tank Agitators (5 HP motor, 24 in dia. impellor)	$10,000	5	$50,000
58	1	Pregnant Solution Metering Pump (dual diaphragm, 0-210 gph, 2 HP drive)	$17,500	2	$35,000
59	1	Precipitated U Slurry Pumps (centrifugal, 20 gpm max, 65' TDH, 2 HP)	$4,500	2	$9,000
60	1	Uran. Peroxide Clarifier/Thickener (10' dia., 6' walls, rake, 7.5 HP, 316 SS)	$63,750	1	$63,750

Table B.2
Equipment List for
Carbonate/Bicarbonate Vat Extraction Process

Item	Source	Item Description	Cost/Unit	Quantity	Total Cost
61	1	Clarifier O/F Decant Tank (10' D x 10' H cyl tank, 6000 gal. WV, 316 SS)	$45,000	1	$45,000
62	1	Precoat Filter Feed Pump (transfer pump, 316 SS casing, 100 TDH, 5 HP)	$4,500	2	$9,000
63	1	Precoat Filter (50 sq ft unit area, pressure leaf filter, 50 gpm flow)	$42,000	1	$42,000
64	1	Filter Precoat Pump (transfer, 2 in x 1.5 in, 6 in dia. impellor, 5 HP, 75 gpm)	$4,500	1	$4,500
65	1	Precoat Mix Tank (48 in D x 54 in H steel tank, 16 in dia. propellor)	$7,910	1	$7,910
66	5	Carbonate Reagent Makeup Tank (6' x 8' cyl. tank, 1000 gal. WV, 316 SS)	$4,000	1	$4,000
67	1	Carbonate Reagent Tank Agitator (5 HP, 30 in dia. impellor, 0-100 rpm)	$10,000	1	$10,000
68	1	Bulk Carbonate Storage Containers (316 SS, 6 cu yd, 6' x 6' x 6')	$3,500	12	$42,000
69	1	Bulk Salt Screw Feeder (5 cu ft hopper, volumetric feeder)	$15,000	1	$15,000
70	1	Bulk Salt Storage Containers (tote type, 316 SS, 6 cu yd, 6' x 6' x 6')	$3,500	6	$21,000
71	5	Strip Solution Makeup Tanks (15000 gallon max WV, carbon steel)	$25,000	2	$50,000
72	1	Strip Sol. Feed Pumps (transfer, 2 in x 1.5 in, 6 in dia. imp., 5 HP, 100 TDH)	$7,500	2	$15,000
73	1	Uranium Peroxide Clarifier U/F Pumps (10 gpm max, 50 TDH, 2 HP)	$5,900	2	$11,800
74	1	Recessed Plate Precipitate Filter (plate filter press, 5 HP, 27.3 cu. ft cap.)	$40,000	1	$40,000
75	1	Bulk Carbonate/Bicarbonate Screw Feeders (5 cu. ft. hopper)	$15,000	2	$30,000
		Total Equipment Cost			**$6,151,060**

Table B.3
Equipment List for
Heap Leaching Process

Item	Source	Item Description	Cost/Unit	Quantity	Total Cost
1	5	Soil Receiving Pad (asphalt, 6 in thick, 5,400 sq. ft, holds 1 day inventory)	$3.50	5,400	$18,900
2	5	Sand Receiving Pad (asphalt, 6 in thk, 5,400 sq. ft., holds 5 days inventory)	$3.50	5,400	$18,900
3	5	Leached Sand/Soil Pad (asphalt, 6 in thk, 6,300 sq. ft., 1 day inventory)	$3.50	6,300	$22,050
4	1	Pad Sump Pumps (centrifugal, 100 gpm, 50 ft TDH, 5 HP, fixed speed)	$6,500	5	$32,500
5	1	Dry Grizzly (6 ft x 8 ft, separates oversize rocks & twigs from feed soil)	$45,000	1	$45,000
6	1	Drum Scrubber 1 (4' D x 6' L, 6' trommel section)	$90,000	1	$90,000
7	1	O/S Soil Transport Conveyor (24 in W x 100 ft L)	$60,000	1	$60,000
8	1	Dewatering Screen (6 ft x 8 ft, receives -4 in slurry with .5 in bar screen)	$50,000	1	$50,000
9	1	U/S Soil Transport/Stacking Conveyor (30 in. W x 140 ft L, at $750/ft)	$105,000	1	$105,000
10	5	Pad Under Soil Stockpile (asphalt, 9 in thk., rein. concrete piers)	$5.00	7,000	$35,000
11	1	Circular Stockpile Building (cylindrical, steel, 86 ft dia x 30 ft ht, $50/sq ft)	$290,500	1	$290,500
12	1	Armco Under-the-Pile Reclaim System (120 ft L conveyor, 180 ft tunnel)	$90,000	1	$90,000
13	1	Reclaimed Soil Transport Conveyor (VS drive, 30 in W x 140 ft L, at $300/ft)	$42,000	1	$42,000
14	1	Cross-Country Sand Feed Conveyor (VS, 30 in W x 200 ft L, at $300/ft)	$60,000	1	$60,000
15	1	Sand Feed Conveyor (constant speed drive, 30 in W x 40 ft L, at $750/ft)	$30,000	1	$30,000
16	1	Agglomerator Drum (6 ft D x 20 ft L, 25 HP, mixes soil, sand, etc.)	$150,000	1	$150,000
17	5	Sand/Soil Permanent Leaching Pad (asphalt, 430 ft x 165 ft, $5.00/sq ft)	$5.00	70,950	$354,750
18	1	Leach Pad Sump Pumps (centrifugal, 100 gpm, 50 ft TDH, 5 HP, fixed sp.)	$6,500	6	$39,000
19	1	Pond Lixiviant Pumps (centrifugal, 150 gpm, 150 psig TDH, 7.5 HP, var. sp)	$10,000.00	5	$50,000
20	5	Bicarbonate Reagent Makeup Tank (6 ft x 8 ft cyl. tank, 1,000 gal., 316 SS)	$4,000	1	$4,000

Table B.3
Equipment List for
Heap Leaching Process

Item	Source	Item Description	Cost/Unit	Quantity	Total Cost
21	1	Bicarbonate Reagent Tank Agitator (5 HP VS motor, 0-100 RPM)	$10,000	1	$10,000
22	1	Bulk Carbonate Storage Containers (316 SS, 6 cu yd cap., 6 ft x 6 ft x 6 ft)	$3,500	12	$42,000
23	1	Bulk Bicarbonate Screw Feeders (5 cu ft hopper, var. sp., volumetric feed)	$15,000	2	$30,000
24	1	Bicarbonate Reagent Metering Pump (0-210 gph, 2 HP, var. sp. drive)	$17,500	2	$35,000
25	5	Liquid Caronic Storage Tank (truck trailer, 3,000 gal., rubber lined)	$20,000	1	$20,000
26	1	Cross-Country Pad Loading Conveyor (30 in x 730 ft, VS, covered, $300/ft)	$219,000	1	$219,000
27	1	Pad Loading Portable Conveyors (30 in. W x 50 ft. L, FS drive, $300/ft)	$15,000	4	$60,000
28	1	Pad Unloading Portable Conveyors (30 in. W x 50 ft. L, FS drive, $210/ft)	$10,500	4	$42,000
29	1	Cross-Country Pad Unloading Conveyor (30 in W x 730 ft L, cover, $274/ft)	$200,000	1	$200,000
30	5	Leachate Piping (1 lot, various lengths & sizes of HDPE piping)	$70,000	1	$70,000
31	5	Leachate Storage Ponds (150' x 50' x 10' D, $3.50/sq ft, 23,000 sq ft HPDE)	$26,250	4	$105,000
32	5	Pregnant Leachate Storage Tank (10'D x 10' H cyl. tank, 5000 gal. WV, CS)	$10,000	1	$10,000
33	5	Recycle Lixiviant Storage Tank (10'D x 10'H cyl. tank, 5,000 gal. WV, CS)	$10,000	1	$10,000
34	5	Filter Backwash Tank (10'D x 10'H cyl. tank, 5,000 gal. WV, CS shell)	$10,000	1	$10,000
35	1	Sand Filters (3'D x 8'H cyl. tank, 65 psig, sand-coal media, CS, with pumps)	$37,000	2	$74,000
36	1	Pregnant Lix. Feed Pumps (centrifugal, 60 gpm, 65' TDH, 5 HP)	$6,500	2	$13,000
37	1	Recycle Lixiviant Pumps (centrifugal, 60 gpm, 65' TDH, 5 HP)	$6,500	2	$13,000
38	1	Recycle Bleed Pumps (centrifugal, 60 gpm, 65' TDH, 5 HP, VS drive)	$6,500	2	$13,000
39	5	Ion Exchange Feed Tank (10'D x 10'H cyl. tank, 5,000 gal. WV, CS shell)	$10,000	1	$10,000
40	1	Ion Exchange Feed Pump (centifugal, 40 gpm, 50 psig TDH, 7.5 HP, VS)	$6,500	2	$13,000

Table B.3
Equipment List for
Heap Leaching Process

Item	Source	Item Description	Cost/Unit	Quantity	Total Cost
41	1	Carbon Guard Column (5'D x 8'H cyl. tank, 2 tanks/skid, 1 train, 2 stage)	$85,000	1	$85,000
42	1	IX System Columns (2 trains, fixed bed, 5'D x 8'H cyl tank, 316 SS, inc. resin)	$97,200	2	$194,400
43	5	Carbon Regenerate Tank (10'D x 10'H cyl. tank, 5,000 gal. WV, CS shell)	$15,000	2	$30,000
44	1	Air Compressors (150 psig, 150 SCFM capacity, 50 HP reciprocating type)	$48,500	2	$97,000
45	1	Carbon Regenerate Pump (centrifugal, 60 gpm, 65' TDH, 5 HP)	$9,500	2	$19,000
46	5	IX Pregnant Solution Tank (10'D x 10'H cyl. tank, 5,000 gal. WV, 316L SS)	$27,000	2	$54,000
47	5	Concentrated HCl Storage Tank (truck trailer, 3,000 gal. rubber lined, CS)	$20,000	1	$20,000
48	1	Acid Metering Pumps (dual diaphragm, 0-160 gph, 2 HP, var. spd.)	$20,000	2	$40,000
49	1	Peroxide Metering Pumps (dual diaphragm, 0-12 gph, 0.75 HP, var. spd.)	$10,000	2	$20,000
50	1	NaOH Metering Pumps (dual diaphragm, 0-12 gph, 0.75 HP, var. spd.)	$10,000	2	$20,000
51	1	Peroxide Precipitation Reactor Tank (386 gal. WV, 48"x48"x54"H, HDPE)	$3,500	5	$17,500
52	1	Peroxide Precip. Tank Agitators (5 HP VS low shear motor, 0-100 RPM,)	$10,000	5	$50,000
53	1	Pregnant Solution Metering Pump (dual diaphragm, 0-210 gph, 2 HP, VS)	$17,500	2	$35,000
54	1	Precipitated U Slurry Pump (centrifugal, 20 gpm, 65' TDH, 2 HP, VS)	$4,500	2	$9,000
55	1	Uranium Peroxide Clarifier/Thickener (10'D, 6' cyl. walls, 7.5 HP, 316L SS)	$63,750	1	$63,750
56	5	Clarifier O/F Decant Tank (10'D x 10'H cyl. tank, 6,000 gal. WV, 316SS)	$45,000	1	$45,000
57	1	Precoat Filter Feed Pump (100' TDH, 75 gpm, 5 HP, 316 SS casing)	$4,500	2	$9,000
58	1	Precoat Filter (50 sq. ft. unit, pressure leaf filter, 50 gpm flow)	$42,000	1	$42,000
59	1	Filter Precoat Pump (75 gpm, 5 HP, 100' TDH, 316 SS casing)	$4,500	1	$4,500
60	1	Precoat Mix Tank (48"D x 54"H mild steel tank with fixed speed agitator)	$7,910	1	$7,910

Table B.3
Equipment List for
Heap Leaching Process

Item	Source	Item Description	Cost/Unit	Quantity	Total Cost
61	1	Bulk Salt Screw Feeder (5 cu. ft. hopper, 3.2-160 cu. ft./hr var. spd. feeder)	$15,000	1	$15,000
62	1	Bulk Salt Storage Container (6 cu. yd., 6' x 6' x 6' conical discharge, 316 SS)	$3,500	6	$21,000
63	5	Strip Solution Makeup Tank (5,000 gal. WV, 10'D x 10'H, CS shell, agitator)	$25,000	2	$50,000
64	1	Strip Solution Feed Pump (50 gpm, variable speed, 100' TDH, 5 HP)	$7,500	2	$15,000
65	1	Uranium Peroxide Clarifier U/F Pump (Moyno type, 10 gpm, 50' TDH, 2 HP)	$5,900	2	$11,800
66	1	Recessed Plate Precipitate Filter (filter press, 27.3 cu. ft cap., 5 HP)	$40,000	1	$40,000
67	1	Dry Flocculant Hopper/Feeder (5 cu. ft. hopper, .018-.90 cu. ft./hr var. spd.)	$15,000	1	$15,000
68	5	Flocculant Mix Tank (6'D x 6'H cyl. tank, 1,000 gal. WV, 316 SS, in. agitator)	$2,700	1	$2,700
69	1	Flocculant Mix Tank Agitator (variable speed mixer, 5 HP, 0-100 rpm)	$10,000	1	$10,000
70	1	Flocculant Metering Pump (dual-diaphragm, 0-160 gpm, 2 HP, var. sp.)	$17,500	1	$17,500
71	1	Turbid Water Clarifier/Thickener (10'D, 6' cyl. walls, 7.5 HP, rake, 316 SS)	$63,750	1	$63,750
72	1	Turbid Water Clarifier U/F Pumps (Moyno, 10 gpm, 50' TDH, 2 HP, var. sp.)	$5,900	1	$5,900
73	1	Recessed Plate Precipitate Filter (filter press, 27.3 cu. ft. cap., 5 HP)	$40,000	1	$40,000
		Total Equipment Cost			**$3,757,310**

Table B.4
Equipment List for
High-Gradient Magnetic Separation

Item	Source	Item Description	Cost/Unit	Quantity	Total Cost
1		Grizzly Feed Conveyor (24 in. W x 100' L at $600/ft)	$60,000	1	$60,000
2		Wet Grizzly (6' x 8', 40 HP)	$45,000	1	$45,000
3		Rotary Drum Scrubber No. 1 (4' D x 6' L, 40 HP)	$90,000	1	$90,000
4		Rotary Drum Scrubbert No. 2 (6'D x 8'L, 60 HP)	$150,000	1	$150,000
5		Coarse Soil Conveyor (24 in. W x 100' L, 10 HP)	$60,000	1	$60,000
6		Live Bottom Sump & Pump (300 gpm, 50' TDH, 20 HP)	$22,500	2	$45,000
7		Washing & Dewatering Screen (6' x 8' screen deck, 10 HP)	$40,000	1	$40,000
8		Sizing Screen No. 1 (154 gpm, 4' x 8' deck)	$30,000	1	$30,000
9		Roll Crusher 12 in. x 8 in. crushing rolls, 5 HP, incl. feed hopper)	$15,600	1	$15,600
10		Agitated Holding Tank (15,000 gal. WV, 14'D x 16' high, carbon steel)	$25,000	1	$25,000
11		Holding Tank Agitator (15 HP motor, 56 rpm, 72 in dia. SS impellor)	$14,600	1	$14,600
12		Sizing Screen No. 2 Feed Pumps (325 gpm output, 65' TDH, 20 HP)	$9,500	2	$19,000
13		Sizing Screen No. 2 (165 gpm, 4' x 8' deck, multifeed screens)	$34,000	1	$34,000
14		Attrition Scrubber (2 cell scrubber, 5 minute retention time, 30 HP)	$36,000	1	$36,000
15		Sizing Screen No. 3 (65 gpm, 4' x 5' deck, single screen)	$33,000	1	$33,000
16		Hydrosizer Feed Holding Tank (15,000 gal. WV, 14'D x 16' H, carbon steel)	$25,000	1	$25,000
17		Hydrosizer Feed Tank Agitator (15 HP motor, 56 rpm, 72 in. dia. impellor)	$14,600	1	$14,600
18		Hydrosizer Feed Pumps (410 gpm max. output, 55' TDH, 20 HP)	$9,500	2	$19,000
19		Hydrosizer (100 to 300 tph capacity, 50 HP)	$56,000	1	$56,000
20		Coarse Fraction Feed Tank (11,500 gal. VW, 12'D x 15'H, carbon steel)	$23,000	1	$23,000

Table B.4
Equipment List for
High-Gradient Magnetic Separation

Item	Source	Item Description	Cost/Unit	Quantity	Total Cost
21		Fine Fraction Feed Tank (11,500 gal. VW, 12'D x 15'H, carbon steel)	$23,000	1	$23,000
22		Fraction Feed Tank Agitators (15 HP motor, 56 rpm, 72 in. dia impellor)	$14,600	2	$29,200
23		HGMS Feed Pumps (centrifugal, 325 gpm, 65' TDH, 20 HP)	$9,500	2	$19,000
24		High Gradient Magnetic Separator (HGMS) System (26 HP)	$1,900,000	1	$1,900,000
25		HGMS Concentrate Thickener (17' dia., 10' depth, 3 HP)	$77,000	1	$77,000
26		Concentrate Thickener U/F Filter Feed Pump (100' TDH, 5 HP, 75 gpm)	$4,500	1	$4,500
27		Filter Press (260 total cu. ft. capacity, 30 HP, 135-32 in. x 32 in. plates)	$223,000	1	$223,000
28		Decant Water Storage Tank (12,700 gal. WV, 12'D x 16' H, carbon steel)	$25,000	1	$25,000
29		Decant Water Tank Agitator (10 HP motor, 56 rpm, 54 in. dia. impellor)	$12,500	1	$12,500
30		Backflush Pump (centrifugal, 200 gpm, 65' TDH, 10 HP)	$10,000	1	$10,000
31		HGMS Recycle Feed Tank (11,500 gal. WV, 12'D x 15'H, carbon steel)	$23,000	1	$23,000
32		HGMS Recycle Feed Tank Agitator (15 HP, 56 rpm, 72 in dia impellor)	$14,600	1	$14,600
33		HGMS Recycle Feed Pump (325 gpm, 65' TDH, 20 HP)	$9,500	1	$9,500
34		Static Mixer 1 (350 gpm, 316 SS, 24 element internals)	$5,000	1	$5,000
35		HGMS Tails Thickener (40 ft. dia., 10' depth, 7.5 HP)	$180,000	1	$180,000
36		Tails Thickener U/F Filter Feed Pump (50' TDH, 200 gpm, 5 HP)	$4,500	2	$9,000
37		Pressure Filter (LAROX, 409 sq. ft., 2 filters, 18 tph soil cap., 50 HP)	$400,000	1	$400,000
38		Recycle Water Tank (11,500 gal. WV, 12'D x 15'H, carbon steel)	$23,000	1	$23,000
39		Recycle Water Tank Agitator (15 HP motor, 56 rpm, 72 in. dia. impellor)	$14,600	1	$14,600
40		Recycle Water Pumps (centrifugal, 500 gpm, 50' TDH, 20 HP)	$12,500	2	$25,000

Table B.4
Equipment List for
High-Gradient Magnetic Separation

Item	Source	Item Description	Cost/Unit	Quantity	Total Cost
41		Dry Surfactant Feeder/Hopper (5.0 cu. ft. hopper, 5 HP)	$15,000	1	$15,000
42		Surfactant Solution Holding Tank (8'D x 10'H, cyl. tank, 3,000 gal. WV, CS)	$6,000	1	$6,000
43		Surfactant Solution Tank Agitator (7.5 HP, 0-100 RPM, 48 in. dia. impellor)	$11,000	1	$11,000
44		Surfactant Feed Pumps (dual diaphram, 0-210 gph, VS drive, 2 HP)	$17,500	2	$35,000
45		Dry Dithionate Feeder/Hopper (5.0 cu. ft. hopper, 5 HP)	$15,000	1	$15,000
46		Dithionate Solution Holding Tank (8'D x 10'H cyl. tank, 3,000 gal. WV, CS)	$6,000	1	$6,000
47		Dithionate Solution Tank Agitator (7.5 HP VS motor, 0-100 RPM)	$11,000	1	$11,000
48		Dithionate Feed Pumps (dual diaphram, 0-210 gph, VS drive, 2 HP)	$17,500	2	$35,000
49		Dry Flocculant Feeder/Hopper (5.0 cu. ft. hopper, 5 HP)	$15,000	1	$15,000
50		Flocculant Holding Tank (8'D x 10'H cyl. tank, 2,000 gal. WV, CS)	$5,000	1	$5,000
51		Flocculant Holding Tank Agitator (7.5 HP, 0-100 RPM)	$11,000	1	$11,000
52		Flocculant Metering Pump (dual diaphram, 0-210 gph, VS drive, 2 HP)	$17,500	1	$17,500
53		Static Mixer 2 (100 gpm, 12 element internals, 316 SS)	$2,200	1	$2,200
54		Static Mixer 3 (350 gpm, 24 element internals, 316 SS)	$5,000	1	$5,000
		Total Equipment Cost			$4,051,400

Table B.5
Equipment List for
Tiron Soil Washing System

Item	Source	Description	Cost/Unit	Quantity	Total
1	1	Grizzly Feed Conveyor (24 in W x 100 ft L)	$60,000	1	$60,000
2	1	Dry Grizzly (6' x 8' sloped entry)	$45,000	1	$45,000
3	1	Drum Scrubber 1 (4' D x 6' L, 6' trommel section)	$90,000	1	$90,000
4	1	Drum Scrubber 2 (6' D x 8'L, 6' trommel section)	$150,000	1	$150,000
5	1	O/S Soil Transport Conveyor (24 in W x 100 ft L)	$60,000	1	$60,000
6	1	Washing & Dewatering Screen (6' x 8' screen deck)	$40,000	1	$40,000
7	1	Slurry Pump and Sump (300 gpm, 50' TDH, 2 cu yd capacity sump)	$22,500	2	$45,000
8	5	Agitated Holding Tank (15000 gallon max volume, carbon steel)	$25,000	1	$25,000
9	1	Holding Tank Agitator (15 HP motor, 56 rpm, 6' dia impellor)	$14,600	1	$14,600
10	1	Leach Feed Pumps (centrifugal, 400 gpm max, 65' TDH, 25 HP)	$10,000	2	$20,000
11	5	Reactor Tanks 1-3 (15' dia x 18' H, 19,500 gal WV, carbon steel)	$30,000	3	$90,000
12	1	Reactor Tank Agitators (15 HP, 56 rpm, 72 in dia impellor)	$15,000	3	$45,000
13	1	Filter Feed Pumps (centrifugal, 400 gpm max, 65' TDH, 25 HP)	$10,000	2	$20,000
14	1	Static Mixer 1 (2 in x 18 in, 316 SS, 100 gpm)	$2,500	1	$2,500
15	1	Static Mixer 2 (4 in x 24 in, 350 gpm max, 316 SS)	$5,000	1	$5,000
16	1	Dry Flocculant Hopper/Feeder (5 cu ft hopper)	$15,000	1	$15,000
17	5	Flocculant Mix Tank (8' dia x 10' H cyl. tank, 2000 gal. WV, 316 SS)	$7,000	1	$7,000
18	1	Flocculant Mix Tank Agitator (7.5 HP, 48 in dia. impellor)	$11,000	1	$11,000
19	1	Flocculant Metering Pumps (dual diaphragm, 0-210 gph, 2 HP drive)	$17,500	2	$35,000
20	1	Horizontal Pressure Filters 1 (LAROX Model PF-16, 409 sq ft filter area)	$400,000	2	$800,000

Table B.5
Equipment List for
Tiron Soil Washing System

Item	Source	Description	Cost/Unit	Quantity	Total
21	5	Repulp Tank 1 (6' x 8' cyl. tank, 1000 gal. WV, 316 SS)	$4,000	1	$4,000
22	1	Repulp Tank 1 Agitator (7.5 HP, 30 in dia. impellor, 0-100 rpm)	$11,000	1	$11,000
23	1	Repulped Leach Feed Pumps (centrifugal, 400 gpm max output, 65' TDH)	$10,000	2	$20,000
24	5	Reactor Tanks 4-6 (15' D x 18' H, 19,500 gal. WV, carbon steel)	$30,000	3	$90,000
25	1	Reactor Tank Agitators (15 HP motor, 56 rpm, 72 in dia. impellor)	$15,000	3	$45,000
26	1	Filter Feed Pumps (centrifugal, 400 gpm, 65' TDH, 25 HP)	$10,000	2	$20,000
27	1	Static Mixer 3 (2 in x 18 in, 316 SS, 100 gpm)	$2,500	1	$2,500
28	1	Static Mixer 4 (4 in x 24 in, 350 gpm, 316 SS)	$5,000	1	$5,000
29	1	Horizontal Pressure Filters 2 (LAROX Model PF-16, 409 sq ft filter area)	$400,000	2	$800,000
30	1	Washed Soil Conveyor (24 in W x 100 ft L)	$60,000	1	$60,000
31	5	Liquid Oxygen Storage Tank & Evaporator (Leased, covered in oper. cost)	$0	1	$0
32	5	Filtrate 1 Storage Tanks (14' D x 20' H cyl tank, 20000 gal. WV, carbon steel)	$27,000	2	$54,000
33	5	Filtrate 2 Storage Tank (14' D x 20' H cyl tank, 20000 gal. WV, carbon steel)	$27,000	1	$27,000
34	5	Filtrate 3 Storage Tank (14' D x 20' H cyl tank, 20000 gal. WV, carbon steel)	$27,000	1	$27,000
35	5	Recycle Lixiviant Storage Tanks (14' D x 20' H cyl tank, 20000 gal. WV, CS)	$27,000	3	$81,000
36	5	Filter Backwash Tank (10' D x 10' H cyl tank, 4500 gal. WV, carbon steel)	$10,000	1	$10,000
37	1	Sand Filters (10' D x 10' H cyl tank, carbon steel)	$99,000	4	$396,000
38	1	Filtrate 1 Feed Pumps (centifugal, 400 gpm max., 65' TDH, 20 HP)	$12,500	3	$37,500
39	1	Filtrate 2 Recycle Pumps (centifugal, 400 gpm max, 65' TDH, 20 HP)	$12,500	2	$25,000
40	1	Filtrate 3 Recycle Pumps (centifugal, 400 gpm max, 65' TDH, 20 HP)	$12,500	1	$12,500

Table B.5
Equipment List for
Tiron Soil Washing System

Item	Source	Description	Cost/Unit	Quantity	Total
41	1	Recycle Lixiviant Pumps (centrifugal, 400 gpm max, 65' TDH, 20 HP)	$12,500	3	$37,500
42	1	Recycle Bleed Pumps (centrifugal, 150 gpm max, 10 HP)	$10,000	2	$20,000
43	5	IX Feed Tank (14' D x 20' H cyl tank, 20000 gallon WV, carbon steel)	$27,000	1	$27,000
44	1	IX Feed Pumps (centrifugal, 400 gpm max)	$12,500	3	$37,500
45	1	Ion Exchange Columns (10' D x 10' H cyl tank, including resin, 65 psig)	$214,000	9	$1,926,000
46	1	Air Compressors (50 HP, 150 psig)	$48,500	2	$97,000
47	5	IX Pregnant Solution Tanks (14' D x 16' H, 12000 gal. WV, 316 SS)	$54,000	2	$108,000
48	5	Concentrated HCl Storage Tank (3,000 gallon tank truck trailer)	$20,000	1	$20,000
49	1	Acid Metering Pumps (dual diaphragm, 0-160 gph, 2 HP drive)	$20,000	2	$40,000
50	1	Peroxide Metering Pumps (dual diaphragm, 0-12 gph, .75 HP drive)	$10,000	2	$20,000
51	1	NaOH Metering Pumps (dual diaphragm, 0-12 gph, .75 HP drive)	$10,000	2	$20,000
52	1	Peroxide Precipitation Reactor Tanks (386 gal. WV, 48 sq in x 54 in H, HDPE)	$3,500	5	$17,500
53	1	Peroxide Precip. Tank Agitators (5 HP motor, 24 in dia. impellor)	$10,000	5	$50,000
54	1	Pregnant Solution Metering Pump (dual diaphragm, 0-210 gph, 2 HP drive)	$17,500	2	$35,000
55	1	Precipitated U Slurry Pumps (centrifugal, 20 gpm max, 65' TDH, 2 HP)	$4,500	2	$9,000
56	1	Uran. Peroxide Clarifier/Thickener (10' dia., 6' walls, rake, 7.5 HP, 316 SS)	$63,750	1	$63,750
57	5	Clarifier O/F Decant Tank (10' D x 10' H cyl tank, 6000 gal. WV, 316 SS)	$45,000	1	$45,000
58	1	Precoat Filter Feed Pump (transfer pump, 316 SS casing, 100 TDH, 5 HP)	$4,500	2	$9,000
59	1	Precoat Filter (50 sq ft unit area, pressure leaf filter, 50 gpm flow)	$42,000	1	$42,000
60	1	Filter Precoat Pump (transfer, 2 in x 1.5 in, 6 in dia. impellor, 5 HP, 75 gpm)	$4,500	1	$4,500

Table B.5
Equipment List for
Tiron Soil Washing System

Item	Source	Description	Cost/Unit	Quantity	Total
61	1	Precoat Mix Tank (48 in D x 54 in H steel tank, 16 in dia. propellor)	$7,910	1	$7,910
62	1	Bulk Salt Screw Feeder (5 cu ft hopper, volumetric feeder)	$15,000	1	$15,000
63	1	Bulk Salt Storage Containers (tote type, 316 SS, 6 cu yd, 6' x 6' x 6')	$3,500	6	$21,000
64	5	Strip Solution Makeup Tanks (15,000 gallon max WV, carbon steel)	$25,000	2	$50,000
65	1	Strip Sol. Feed Pumps (transfer, 2 in x 1.5 in, 6 in dia. imp., 5 HP, 100 TDH)	$7,500	3	$22,500
66	1	Uranium Peroxide Clarifier U/F Pumps (10 gpm max, 50 TDH, 2 HP)	$5,900	2	$11,800
67	1	Recessed Plate Precipitate Filter (plate filter press, 5 HP, 27.3 cu. ft cap.)	$40,000	1	$40,000
		Total Equipment Cost			**$6,104,560**

APPENDIX C

PROCESS DESCRIPTION OF FERNALD INTEGRATED DEMONSTRATION SOIL TREATMENT TECHNOLOGIES

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FERNALD ID TECHNOLOGIES

PRELIMINARY DESIGN INPUT
REVISION 2 PROCESS FLOWSHEETS

Produced By:
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Pittsburgh, PA

January 16, 1995

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FERNALD ID TECHNOLOGIES

AQUEOUS BIPHASIC EXTRACTION PROCESS

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PROCESS DESCRIPTION
REVISED AQUEOUS BIPHASIC EXTRACTION PROCESS

INTRODUCTION:

In order to provide a basis for the FY 95 ID Program for removal of uranium contamination from soils, revised process flowsheets have been developed which reflect experience in primary uranium ore processing as well as recent laboratory results on the contaminated soils. The approach taken has been to assume a relatively optimistic process performance as a basis for equipment selection and process design. In addition, the unit operations selected should be readily operable at the assumed operating conditions. If adequate performance for the selected process concepts and equipment selected are supported by laboratory tests, the capital and operating cost benefits over previous process flowsheets should be substantial.

The following presents a brief process description of the revised Aqueous Biphasic Extraction Process (ABE) flowsheets (Revision 2, 12/19/94) and some key process assumptions used as the basis for initial material balances used for equipment sizing.

COARSE SOIL SEPARATION:

The initial separation of the coarse size fractions of the soil (+2 mm) is accomplished using equipment which minimizes the amount of liquid (lixiviant and fresh water) added to the system. A slurry density (percent solids) consistent with the requirements for efficient leaching without dewatering is the goal.

The soil feed to the process is initiated by reclaim from a soil storage facility or by direct feed from the excavation operations. A feed conveyor delivers the soil to a wet, vibrating Grizzly which scalps off oversize and trash materials (+10 cm) in the soil. Spray water (recycle dilute salt solution) is used, as necessary, to control any dust emissions and to reduce clogging of the undersize chute to the Rotary Drum Scrubber 2. The grizzly is elevated such that the oversize and undersize fractions flow by gravity to the drum scrubbers.

The oversize material from the grizzly reports to the Rotary Drum Scrubber 1 in which the oversize soil is mixed with recycle dilute salt solution as a slurry to wash off any adhered, small-size soil and to solubilize uranium staining the surface of the coarse particles using the leaching action. The drum scrubber is equipped with a solid drum section and dewatering drain for slimes removal as well as a trommel screen extension where rinsing with fresh water removes most of any dissolved uranium. The trommel oversize (+13 mm) is conveyed to a stockpile for disposal.

The undersize soil slurry (-13 mm) from the Drum Scrubber 1 is combined with the -10 cm Grizzly undersize as and flows by gravity as feed to the Rotary Drum Scrubber 2. The washing with recycle dilute salt solution and rinsing on the trommel screen is repeated.

The trommel oversize (+13 mm) is discharged to the oversize conveyor which transports the washed soil to a stockpile for disposal. The undersize slurry (-13 mm) flows by gravity to a live-bottomed sump and pump which delivers it as feed to an elevated washing and dewatering screen. The dewatering screen separates and washes with fresh water the remaining coarse soil fractions (+2 mm). The screen oversize soil also is conveyed to the washed coarse soil stockpile for return to the site. The -2 mm fraction slurry, along with the recycle dilute salt solution and the balance of the wash water, flows by gravity to an agitated holding tank to serve as feed to the leaching circuit.

Since the Karr column requires relatively fine solid particles to work effectively, an intermediate size fraction of soil (-2 mm +150 μ) will be separated from the slurry prior to Karr column feed. In the meantime, the dilute carbonate salt solution transport medium and residence time in the agitated holding tank and PEG thickener should be sufficient to leach and remove into solution any surface stain uranium on these intermediate-size soil particles.

The -2 mm soil slurry suspension in the holding tank is fed to a fine 100 mesh (150 μ) vibrating screen and the +100 mesh solids separated from the balance of the soil. This coarser (+100 mesh) oversize solid from the screen is mixed with recycle PEG filtrate and Karr column underflow as feed to the PEG thickener. A flocculant is added to the diluted slurry and mixed in a static mixer (gravity flow) prior to feeding to the PEG thickener. In this manner, the intermediate-size soil particles is recombined with the fine soil fractions prior to filtration and disposal.

The philosophy of the above coarse ore separation circuit for the ABE process is to produce a PEG phase (polyethylene glycol) feed slurry to the Karr column contactors of about 30 to 35% solids, leach any surface uranium contamination from the coarse soil fraction and wash the coarse soil with fresh water prior to return to the site. The balance between recycle salt solution and fresh water is maintained to keep the final salt (sodium carbonate) concentration in the feed to the Karr Columns at about 1.82 wt.% (i.e. a tie-line equilibrium composition). The above equipment should be capable of this goal without the need for mechanical dewatering prior to feeding the Karr column and also should minimize slurry pumping.

Use of the salt solution (i.e. low salt recycle) from the methanol precipitation as the primary rinsing and motive liquid for the soil slurry achieves a minimization of bleed requirements and salt (carbonate) makeup requirements.

AQUEOUS BIPHASIC EXTRACTION CIRCUIT:

The ABE process separates the fine uranium rich soil particles and the fine discrete uranium mineral particles from the lower uranium concentration soil particles using the surface activity of the particles in the extraction media as the principal separation mechanism. The extraction media consists of two immiscible liquid phases (biphasic) where the relative attraction for certain solid particles in each liquid phase is different. The differences in attraction and the speed of the liquid/liquid phase separation makes it possible to effect a separate of solid particle types.

In this ABE system, one liquid phase is a solution of polyethylene glycol (PEG) and the other is a sodium carbonate salt solution. The selective partition of such particles in the ABE system is based on physicochemical interactions between the particle surface and the liquid phases, rather than bulk phase properties like density. Consequently, the particle size of the solid phase needs to be small enough so that particle settling due to gravity is slow compared to the rates of liquid/liquid phase separation.

Initial laboratory testing of this concept has led to definition of operating conditions and parameters which are the basis of this Revision 2 process design. Significant advancements in the effectiveness of this technology for use in removing uranium from contaminated soils have been made in the limited testing. Since the ABE process uses, in part, a different separation technology than leaching to effect the separation, it has the promise of being able to recover and remove a high proportion of the difficult-to-leach, refractory uranium mineral species in the soil. Since the salt phase transport media is a high-concentration carbonate salt solution, the easily-leached uranium species are also leached and are separately recovered from the salt phase in a form conducive to permanent disposal. The ABE process, therefore, removes and recovers the uranium as a solid-phase concentrate from the biphasic separation and as a precipitate from leaching into the salt phase.

The fine solids in the 100 mesh screen underflow slurry flows to a mixing tank (PEG repulp tank) where it is mixed with recycle PEG liquid to serve as feed to the Karr column. In order to keep the PEG concentration in this feed slurry at the tie-line equilibrium concentration (33.11 wt% PEG), the recycle PEG recovered as decant from the PEG thickener is augmented with additional PEG to the concentration which upon mixing with the screen underflow produces the required PEG slurry feed. The slurry density of the soil solids in the PEG slurry feed to the Karr column is targeted to be about 25-35 wt.% and is controlled in the coarse soil separation circuit.

The contacting device used for laboratory testing of the ABE process is a multi-stage tower contacting device called a Karr column. In this column, the dense liquid phase (i.e. the PEG slurry) is fed to the top of the contactor column: The less dense salt phase (i.e. about 12 wt.% concentrated sodium carbonate solution) is fed into the bottom of the column and a counter-current flow is established. The column is separated by a number of disk-like plates which provide interstage mixing and phase separation between the two immiscible liquid phases. The multiple plates are slowing turning, driven by a central shaft, which provide multi-staging of the concentration process as well as promoting efficient phase separation.

As a consequence and maintaining the feed compositions of the two immiscible liquid phases at a defined equilibrium point on the liquid-liquid system phase diagram (PEG/carbonate solution), the compositions of the two liquid phases in the Karr column conform to a defined operating tie line connecting with the feed and discharge equilibrium compositions of each phase. Therefore, the discharge compositions from the Karr column are controlled and defined by the feed liquid compositions and mass ratios. The Karr column operation and separation processes are also controlled and are sensitive to the temperature of the system. Therefore feed liquid temperatures and the column temperature gradient is carefully controlled (at about 40°C).

Due to the ABE process characteristics, the uranium rich solid soil particles will concentrate into the salt phase leaving the top of the Karr column. Also in this salt phase is the bulk of the dissolved uranium due to leaching with the carbonate complexing agent. Partition of the bulk of the soil particles should result in the PEG underflow slurry exiting the bottom of the Karr column containing the uranium-depleted soil (about 95-98 wt.% of feed soil). This Karr column PEG underflow slurry is mixed with the coarser soil fractions, fed to a thickener for separation from the PEG phase and ultimately filtered and washed for disposal (e.g. return to site, etc.).

The underflow settled soil slurry in the PEG thickener is pumped to a horizontal pressure filter for dewatering and rinsing. The pressure filter is used since its semi-continuous operation provides efficient and low-operating manpower liquid/solid separation, its ability to dewater difficult materials (such as the clayey soil particles) and will produce a relatively-high percent solids (30-40 wt.% moisture) filter cake product.

In the pressure filter, which operates in cycles with batches of slurry being dewatered, washed and rinsed in successive batch operations within the cycles, the PEG U/F slurry is dewatered to produce a PEG-rich filtrate which is recycled as feed to the PEG thickener. In addition, a washing cycle using recycle salt solution after methanol precipitation removes most of the remaining PEG from the soil solids in the cake. This Rinse Filtrate 1 feeds a proprietary dewatering process which removes extraneous water and reconstitutes the carbonate concentration to that required for the PEG slurry system. This dewatering process is proprietary; thus is not defined in this scope.

The rinse filtrate using fresh water (Rinse Filtrate 2) reports back to the Return Salt Solution Storage Tank and provides makeup water to the salt phase generation. The dissolved uranium washed from the soil solids therefore is introduced into the salt phase for ultimate recovery and removal.

URANIUM RECOVERY SYSTEM (CONCENTRATE):

The uranium-rich concentrate recovery system from the salt phase overflow from the Karr column consists of a thickener to concentrate the solids into the underflow and to produce a clarified salt phase decant and a recessed plate and frame filter press to dewater and dry the uranium concentrate solids in a form suitable for disposal.

The thickener underflow is intermittently as a batch process is pumped to the filter press. Pumping (using a positive-displacement pump) continues until the press is full of solids (as indicated by filter pressure drop). The filtrate from the press returns as feed to the salt phase thickener and ultimately as clarified decant. The solid cake is air-blown and deposited into appropriate containers for further stabilization treatment or disposal (off-site).

The salt phase decant proceeds to a precipitation system (using methanol) for the dissolved uranium components.

URANIUM RECOVERY (METHANOL PRECIPITATION):

The ABE process laboratory development program has identified a process using methanol to precipitate the uranium from the relatively high-concentration salt (sodium carbonate) phase. This process has been demonstrated to be able to quantitatively remove the dissolved uranium as a precipitate from such solutions using various ratios of methanol to salt phase. However, since this process is still under development, the specific quantitative data which will be required for process design and specification are very preliminary.

It is anticipated that in the range of salt phase:methanol ratios of 1:0.5 to 1:2.0 that a selective uranium-removal process will be developed. This uranium removal from the salt phase is required to prevent excessive buildup in the circulating salt solutions of dissolved uranium. This high concentration of solubilized uranium in contact with the soil solids would make it nearly impossible to rinse the uranium away from the soil without creating excessive bleed requirements or treatment.

Since the methanol precipitation process is relatively undefined, a conceptual process is presented. The salt phase is mixed in an agitated tank with methanol. For purposes of equipment sizing a salt phase/methanol ratio of 1:1 was used. The bulk of the uranium in solution and part of the dissolved sodium carbonate in the salt phase solution will precipitate.

These solids are settled in a thickener/clarifier which produces a clarified salt/methanol liquid phase decant. The thickener underflow slurry is filtered in a recessed plate and frame filter press. This filter cake can be optionally washed with dilute salt phase recycle solution to recover and recycle the methanol from the cake.

The salt/methanol decant is stored in a decant tank which feed a methanol stripping column. This column, with steam-heated reboiler and about 16 perforated plate to effect a stripping of the methanol, produces a pure methanol vapor product and nearly methanol-free salt phase (2-4% methanol) which is recycled for salt phase recycle or is bled from the circuit to keep the liquid balance under control. When bleeding, the methanol ratio in precipitation is increased to produce a low-soluble uranium content in the salt solution being bled.

The salt/methanol liquid entering the methanol stripping column is used to condense and cool the column methanol vapors. This also preheats the feed salt/methanol liquid; thus conserving heat energy. Makeup methanol is provided from a tanker truck storage and the concentrated carbonate salt solutions are reconstituted by bleeding in a high-concentration (20-30 wt.%) sodium carbonate reagent solution (or slurry) produced in a bulk hopper carbonate delivery and mixing system.

Many elements of the ABE process are preliminary since this process for this application is early in its development phase. However, it has been attempted to present a feasible (but not necessarily optimal) conceptual process design for implementation of this process. Progress in the process development is likely to be rapid due to addressing specific problem areas in laboratory testing. Potentially there may be significant improvements in the ABE process design, operating performance and reduction in reagent consumptions and losses.

PROCESS DESIGN ASSUMPTIONS
REVISED AQUEOUS BIPHASIC EXTRACTION PROCESS

The following preliminary process design assumptions were used in flowsheet development, mass balance derivation, equipment selection and preliminary sizing:

COARSE SOIL SEPARATION:

Nominal throughput rate 20.0 dry tons soil/hour.

Soil moisture content average of 12.0 wt.%.

Recycle salt solution and fresh water addition to the scrubbers and 2 mm screen are controlled such that the feed soil slurry density to the holding tank is nominally 30-35 wt.% solids (design based on 30 wt.%).

Coarse oversize soil (+100 mm rocks, roots, etc.) is assumed to be about 1.0 wt.% soil.

Medium size soil (-100 mm + 13 mm) is assumed to be 1.5 wt.% soil.

Intermediate size soil fractions (-13 mm + 2 mm) is assumed to be 7.5 wt.% of the feed soil.

The coarser fraction of the -2 mm +100 mesh (150 μ) soil is assumed to be about 10 wt.% of the feed soil solids.

AQUEOUS BIPHASIC EXTRACTION CIRCUIT:

PEG slurry feed to the Karr column: nominal 30 wt.% solids, range 25-35 wt.%.

Salt phase concentration in PEG slurry feed: ≈1.82 wt.% sodium carbonate.

PEG phase concentration in PEG slurry feed: ≈33.11 wt.%.

Salt concentration in Karr column salt phase feed: ≈12.0 wt.% sodium carbonate.

Concentrations in Karr column overflow: 11.45 wt.% carbonate, 0.60 wt% PEG.

Horizontal pressure belt filter cake moisture of 65.0 wt.%.

Horizontal pressure belt filter design unit area of 50.0 lbs/hr/ft^2 for dewatering.

Belt filter cake solution acceptance rate for washing/rinsing is 0.080 gpm/ft^2.

Dry flocculant addition system based on a total 0.65 lbs flocculant/ton soil solids.

Karr column (6' diameter) feed rate @≈30 wt.% solids is 3,000 kg/hr.

Karr column operating temperature ≈40°C.

URANIUM RECOVERY SYSTEM (CONCENTRATE):

Recessed plate & frame uranium concentrate filter unit area based on 27.3 ft^3 net cake capacity, one filtration cycle/day.

URANIUM RECOVERY (METHANOL PRECIPITATION):

Salt phase/methanol ratio: design 1:1, range 1:05 to 1:2

Residual uranium: 1.0 ppm to 19 ppm

Bleed rate: salt solution ≈3.0 % C.L.

SHEET 1

AQUEOUS BIPHASIC EXTRACTION PROCESS FOR SOIL WASHING
20.0 DTPH SOIL INPUT, COARSE SEPARATION AND KARR COLUMN FEED

12-Jan-95
ISSUE 2
Rev. 0

STREAM NUMBER	STREAM DESCRIPTION (BASED ON FLEXMET BALANCES)	SOLIDS S.G.	LIQUID S.G.	SLURRY S.G.	BULK DENSITY	WT.% SOLIDS	SOLIDS (STPH)	LIQUID (STPH)	TOTAL (STPH)	TOT. GPM (yd3/HR)	STREAM NUMBER
1	SOIL FEED TO GRIZZLY	2.500	1.000	2.119	90.0	88.00	20.0000	2.7270	22.7270	(18.70)	1
2	GRIZZLY OVERSIZE (+4") TO SCRUBBER 1	2.500	1.000	2.119	132.2	88.03	0.2000	0.0272	0.2272	0.43	2
3	GRIZZLY UNDERSIZE (-4") TO SCRUBBER 2	2.500	1.000	2.119	132.1	88.00	19.8000	2.6998	22.4998	42.43	3
4	TROMMEL OVERSIZE (+1/2") FROM SCRUBBER 1	2.500	1.000	1.923	84.0	80.00	0.2000	0.0500	0.2500	(0.22)	4
5	TROMMEL OVERSIZE (+1/2") FROM SCRUBBER 2	2.500	1.000	1.923	84.0	80.00	0.3000	0.0750	0.3750	(0.33)	5
101	SCRUBBER 1 WASH WATER	—	1.000	1.000	62.4	0.00	0.0000	2.2050	2.2050	8.81	101
102	SCRUBBER 2 WASH WATER	—	1.000	1.000	62.4	0.00	0.0000	2.2050	2.2050	8.81	102
51	SCRUBBER 1 RECYCLE CARBONATE SOLUTION	—	1.037	1.037	64.7	0.00	0.0000	6.2790	6.2790	24.18	51
52	SCRUBBER 2 RECYCLE CARBONATE SOLUTION	—	1.053	1.053	65.7	0.00	0.0000	6.2790	6.2790	23.83	52
6	WASHING & DEWATERING SCREEN FEED	2.500	1.029	1.457	90.9	49.91	19.5000	19.5700	39.0700	107.14	6
7	SCREEN O/S (+2mm) TO STOCKPILE	2.500	1.000	1.923	84.0	80.00	1.5000	0.3750	1.8750	3.90	7
8	SCREEN U/S (-2mm) TO HOLDING TANK	2.500	1.021	1.337	83.4	40.00	18.0000	27.0000	45.0000	134.42	8
103	SCREEN WASH WATER	—	1.000	1.000	62.4	0.00	0.0000	7.8040	7.8040	31.18	103
9	COMBINED (+2mm) O/S TO DISPOSAL STOCKPILE	2.500	1.000	1.923	119.9	80.00	2.0000	0.5000	2.5000	5.19	9
10	NET FEED TO 100m SCREEN	2.500	1.021	1.337	83.4	40.00	18.0000	27.0000	45.0000	134.42	10
11	SCREEN O/S (+100m) TO PEG THICKENER	2.500	1.021	1.337	83.4	40.00	2.0000	3.0000	5.0000	14.93	11
12	SCREEN U/S (-100m) TO PEG REPULP TANK	2.500	1.021	1.337	83.4	40.00	16.0000	24.0000	40.0000	119.48	12
63	DRY FLOCCULANT	1.800	1.000	1.800	112.3	100.00	0.0052	0.0000	0.0052	(0.0034)	63
109	DRY FLOC DILUTION WATER	—	1.000	1.000	62.4	0.00	0.0000	0.5150	0.5150	2.06	109
62	INITIAL DILUTED FLOC MIXTURE (1% SOLN.)	1.800	1.000	1.004	62.6	0.99	0.0052	0.5150	0.5202	2.07	62
21	FLOC DILUTION (FROM PEG DEWATERING)	1.800	1.115	1.115	69.6	0.00	0.0000	4.2010	4.2010	15.05	21
64	DILUTED FLOC TO STATIC MIXER	1.800	1.103	1.103	83.4	40.0	0.0052	4.7160	4.7212	17.10	64
44	KARR COLUMN PEG PHASE U/F TO STATIC MIXER	2.500	1.092	1.314	82.0	30.00	15.6800	36.5870	52.2670	158.87	44
47	PRESSURE FILTER PEG FILTRATE TO STATIC MIXER	2.500	1.089	1.089	67.9	0.00	0.0000	12.0920	12.0920	44.38	47
42	NET FEED TO PEG THICKENER	2.500	1.089	1.258	78.5	23.87	17.6850	56.3940	74.0790	235.21	42
49	PEG THICKENER DECANT TO RECYCLE TANK	2.500	1.089	1.089	67.9	0.00	0.0000	34.7790	34.7790	127.63	49
84	MAKEUP PEG	—	2.097	2.097	130.8	0.00	0.0000	0.2171	0.2171	0.41	84
107	PEG MAKEUP DILUTION WATER	—	1.000	1.000	62.4	0.00	0.0000	0.0000	0.0000	0.00	107
43	NET RECYCLE PEG TO REPULP TANK	2.500	1.095	1.095	68.3	0.00	0.0000	34.9961	34.9961	127.70	43
14	PEG SLURRY FEED TO KARR COLUMN FROM REPULP	2.500	1.064	1.211	75.5	21.09	16.0000	59.8590	75.8590	250.35	14
45	U/F FROM PEG THICK. TO PRESSURE FILTER	2.500	1.089	1.459	91.0	45.00	17.6850	21.6150	39.3000	107.58	45
55	RECYCLE DILUTE SALT SOLN. WASH ON FILTER	2.500	1.021	1.021	63.7	0.00	0.0000	8.4010	8.4010	32.87	55
108	FRESH WASH WATER ON PRESSURE FILTER	—	1.000	1.000	62.4	0.00	0.0000	16.8020	16.8020	67.12	108
46	FILTRATE 1 (FROM DIL. SALT SOLN. WASH)	2.500	1.062	1.062	66.2	0.00	0.0000	8.4010	8.4010	31.61	46
48	RECYCLE DILUTE SALT SOLN. WASH ON FILTER	2.500	1.025	1.025	63.9	0.00	0.0000	16.8020	16.8020	65.48	48
90	BLEED FROM PROPRIETARY DEWATER PROCESS	2.500	1.000	1.000	62.4	0.00	0.0000	4.2000	4.2000	16.78	90
16	WASHED FILTER CAKE (-2mm) SOIL TO DISPOSAL	2.500	1.008	1.639	102.2	65.00	17.6850	9.5230	27.2080	(19.71)	16

*NOTE: BASED ON INITIAL MASS BALANCES FOR REVISION 1 ABE FLOWSHEET (12/19/94) AND MAXIMUM 40% SOLIDS IN SCREEN FEED.

SHEET 2

AQUEOUS BIPHASIC EXTRACTION PROCESS FOR SOIL WASHING

20.0 DTPH SOIL INPUT, KARR COLUMN FEED AND METHANOL PRECIPITATION

12-Jan-95

ISSUE 1

Rev. 1

STREAM NUMBER	STREAM DESCRIPTION (BASED ON FLEXMET BALANCES)	SOLIDS S.G.	LIQUID S.G.	SLURRY S.G.	BULK DENSITY	WT.% SOLIDS	SOLIDS (STPH)	LIQUID (STPH)	TOTAL (STPH)	TOT. GPM (yd3/HR)	STREAM NUMBER
14	PEG SLURRY FEED TO KARR COLUMN FROM REPULP	2.500	1.093	1.240	77.4	21.09	16.0000	59.8590	75.8590	244.33	14
44	KARR COLUMN PEG PHASE U/F TO STATIC MIXER	2.500	1.092	1.314	82.0	30.00	15.6800	36.5870	52.2670	158.87	44
20	CONCENTRATED CARBONATE SALT SOLUTION FEED	2.500	1.267	1.000	62.4	0.00	0.0000	9.4526	9.4526	37.76	20
15	SALT PHASE COLUMN O/F TO THICKENER	3.000	1.123	1.006	43.9	0.97	0.3200	32.7246	33.0446	131.16	15
136	NET FEED TO THICKENER	2.500	1.017	1.006	43.9	0.96	0.3200	33.0674	33.3874	132.62	136
19	SALT DECANT FROM THICKENER	2.500	1.123	1.000	62.4	0.00	0.0000	32.5874	32.5874	130.19	19
40	THICKENER U/F SLURRY TO FILTER	2.500	1.123	1.000	62.4	40.00	0.3200	0.4800	0.8000	3.20	40
18	FILTRATE FROM THICKENER	3.000	1.123	1.123	70.0	0.00	0.0000	0.3429	0.3429	1.22	18
41	URANIUM CONCENTRATE (FROM PEG COLUMN)	3.000	1.123	1.123	70.0	70.00	0.3200	0.1371	0.4571	(0.48)	41
30	METHANOL MAKEUP	3.000	0.800	0.800	49.9	0.00	0.0000	0.0000	0.9848	4.92	30
146	RECYCLE METHANOL FROM STRIP CONDENSOR	3.000	0.800	0.800	34.9	0.00	0.0000	32.5874	32.5874	162.73	146
31	METHANOL FEED TO PRECIPITATION	3.000	0.800	0.800	49.9	0.00	0.0000	0.0000	0.0000	0.00	31
138	NET FEED TO PRECIPITATION	3.000	0.885	0.885	55.2	0.00	0.0000	65.6549	65.6549	296.43	138
139	PRECIPITATION OUTFLOW	3.000	0.885	0.909	56.7	3.72	2.4420	63.2129	65.6549	288.66	139
38	FILTRATE RECYCLE	3.000	0.885	0.885	55.2	0.00	0.0000	2.6164	2.6164	11.81	38
32	NET FEED TO PPTN. THICKENER	3.000	0.885	0.908	56.6	3.58	2.4420	65.8293	68.2712	300.48	32
35	PPTN. THICKENER O/F DECANT TO METH. STRIP FEED	3.000	0.885	0.885	55.2	0.00	0.0000	62.1663	62.1663	280.68	35
34	URANIUM PPT. FILTER CAKE	3.000	0.885	1.233	76.9	40.00	2.4420	3.6629	6.1049	19.79	34
145	FILTER CAKE AFTER DEWATER	3.000	0.885	1.747	109.0	70.00	2.4420	1.0466	3.4885	(2.37)	145
58	SALT SOLUTION WASH	3.000	1.023	1.023	63.8	0.00	0.0000	4.1862	4.1862	16.35	58
36	RINSE FILTRATE	3.000	1.023	1.023	63.8	0.00	0.0000	4.1862	4.1862	16.35	36
37	URANIUM PRECIPITATE TO DISPOSAL	3.000	1.023	1.899	118.4	70.00	2.4420	1.0466	3.4885	(2.18)	37
39	NET FEED TO METHANOL STRIP COLUMN	3.000	0.885	0.885	55.2	0.00	0.0000	66.3526	66.3526	299.52	39
33	RECYCLE SALT SOLUTION	3.000	1.000	1.000	62.4	0.00	0.0000	1.0466	1.0466	4.18	33
51	SCRUBBER 1 RECYCLE CARBONATE SOLUTION	—	1.037	1.037	64.7	0.00	0.0000	6.2790	6.2790	24.16	51
52	SCRUBBER 2 RECYCLE CARBONATE SOLUTION	—	1.053	1.053	65.7	0.00	0.0000	6.2790	6.2790	23.83	52
55	RECYCLE DILUTE SALT SOLN. WASH ON FILTER	—	1.021	1.021	63.7	0.00	0.0000	8.4010	8.4010	32.87	55
57	RECYCLE DILUTE SALT SOLN. TO CARBONATE MAKE	—	1.021	1.021	63.7	0.00	0.0000	0.3434	0.3434	3.63	57
59	SODIUM CARBONATE BULK	2.200	1.000	2.200	54.9	0.00	2.3420	0.0000	2.3420	(3.16)	59
56	BLEED TO AWWT (3% OF CIRC. LOAD IN PPTN.)	—	1.021	0.885	55.2	0.00	0.0000	1.9696	1.9696	8.89	56

*NOTE: BASED ON INITIAL MASS BALANCES FOR REVISION 1 ABE FLOWSHEET (12/19/94). 1:1 AVERAGE SALT/METHANOL RATIO AND 25% CARBONATE REAGENT.

FERNALD ID TECHNOLOGIES

CARBONATE/BICARBONATE SOIL WASHING

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PROCESS DESCRIPTION
REVISED CARBONATE/BICARBONATE SOIL WASHING

INTRODUCTION:

In order to provide a basis for the FY 95 ID Program for removal of uranium contamination from soils, revised process flowsheets have been developed which reflect experience in primary uranium ore processing as well as recent laboratory results on the contaminated soils. The approach taken has been to assume a relatively optimistic process performance as a basis for equipment selection and process design. In addition, the unit operations selected should be readily operable at the assumed operating conditions. If adequate performance for the selected process concepts and equipment selected are supported by laboratory tests, the capital and operating cost benefits over previous process flowsheets should be substantial.

The following presents a brief process description of the revised carbonate/bicarbonate flowsheets (Revision 2, 12/12/94) and some key process assumptions used as the basis for initial material balances used for equipment sizing.

COARSE SOIL SEPARATION:

The initial separation of the coarse size fractions of the soil (+2 mm) is accomplished using equipment which minimizes the amount of liquid (lixiviant and fresh water) added to the system. A slurry density (percent solids) consistent with the requirements for efficient leaching without dewatering is the goal.

The soil feed to the process is initiated by reclaim from a soil storage facility or by direct feed from the excavation operations. A feed conveyor delivers the soil to a wet, vibrating Grizzly which scalps off oversize and trash materials (+10 cm) in the soil. Spray water (recycle filtrate) is used, as necessary, to control any dust emissions and to reduce clogging of the undersize chute to the Rotary Drum Scrubber 2. The grizzly is elevated such that the oversize and undersize fractions flow by gravity to the drum scrubbers.

The oversize material from the grizzly reports to the Rotary Drum Scrubber 1 in which the oversize soil is mixed with filtrate from the second leaching train as a slurry to wash off any adhered, small-size soil and to solubilize uranium staining the surface of the coarse particles using the leaching action. The drum scrubber is equipped with a solid drum section and dewatering drain for slimes removal as well as a trommel screen extension where rinsing with fresh water removes most of any dissolved uranium. The trommel oversize (+13 mm) is conveyed to a stockpile for disposal.

The undersize soil slurry (-13 mm) from the Drum Scrubber 1 is combined with the -10 cm Grizzly undersize as and flows by gravity as feed to the Rotary Drum Scrubber 2. The washing with recycle filtrate and rinsing on the trommel screen is repeated.

The trommel oversize (+13 mm) is discharged to the oversize conveyor which transports the washed soil to a stockpile for disposal. The undersize slurry (-13 mm) flows by gravity to a live-bottomed sump and pump which delivers it as feed to an elevated washing and dewatering screen. The dewatering screen separates and washes with fresh water the remaining coarse soil fractions (+2 mm). The screen oversize soil also is conveyed to the washed coarse soil stockpile for return to the site. The -2 mm fraction slurry, along with the lixiviant and the balance of the wash water, flows by gravity to an agitated holding tank to serve as feed to the leaching circuit.

The philosophy of the above coarse ore separation circuit is to produce a feed slurry to the carbonate leaching reactors of about 30 to 35% solids, leach any surface uranium contamination from the coarse soil fraction and wash the coarse soil with fresh water prior to return to the site. The above equipment should be capable of this goal without the need for mechanical dewatering prior to leaching and also should minimize slurry pumping. Use of the leaching solution (i.e. Belt Filter 2 filtrate) from the Leaching Train 2 as the primary rinsing and motive liquid for the soil slurry achieves a true counter-current and high-efficiency leaching system.

CARBONATE LEACHING CIRCUIT:

The initial leach train consists of three agitated leaching reactors in series. Slurry flow between reactors is achieved by a bleed from a recirculating stream of slurry being pumped from the bottom of the reactor and being introduced into the top of the reactor to promote slurry suspension. Slurry advancement to the next reactor stage is controlled by a slurry pinch valve on the pipeline to the next reactor. The recirculation of the soil slurry around the reactor augments the mixing action in the agitated tank.

The agitators used are low-intensity airfoil-type, downward pumping impellers (e.g. Lighnin A-310 or equivalent) which keeps the slurry in suspension in an axial-flow pattern and which minimizes agitator power requirements. This type of mixers also do not appreciably decrepitate or degrades the soil particle size. It is assumed that further size reduction is not necessary to maximize the uranium extraction.

Makeup carbonate reagent is added as a concentrate sodium carbonate/bicarbonate slurry or solution (about 20-30 wt. %) to the holding tank. The carbonate/bicarbonate ratio is by adjusting the mixture of bulk carbonate and bicarbonate added in the reagent mixing system. This concentrated reagent addition augments the carbonate reagent supplied by recycle filtrate used for coarse soil separation and handling. The holding tank is sized for a minimum of 30 minutes residence time.

The feed soil slurry is pumped from the holding tank to the first reactor vessel in the Leach Train 1. Slurry advances to the other reactors through the recirculation pumping system described above. An average residence time per leach reactor of a minimum of 20 minutes/stage is assumed as a basis for design.

Oxygen gas under slight pressure (≈50 psig) is introduced into each reactor vessel through a bottom sparger manifold to provide and maintain oxidizing potential to facilitate uranium leaching. The addition of an oxidant (air, oxygen or oxygen-enriched air) is deemed essential to insure maximum efficiency of uranium leaching. The vessels are covered, but are vented to maintain atmospheric pressure with a high oxygen partial pressure ($pO_2 \approx 0.8$ atm.) at the slurry surface.

The slurry from the third stage of the initial reactor train is advanced to the feed tank for the first vacuum belt filter. In this initial filtration, only dewatering of the soil solids is done; there is no need for washing or rinsing.

A flocculant (or coagulating agent) mixing, dilution and addition system is provided to assist and aid in the filtration. The bulk dry flocculant is mixed initially to about 1.0% strength using recycle lixiviant as the diluent. It is metered (as 1.0% strength) to mix with the feed slurry to the filter. Before mixing with the slurry, however, it is diluted with additional recycle lixiviant to about 0.10 wt.% strength. A static, in-line mixer is used to insure adequate mixing without shearing the flocculant polymer. The diluted floc solution is then mixed with the feed slurry to the filter also using a static mixer to insure low-shear mixing.

The filter cake (at approximately 50-60 wt.% solids) is discharged into a repulping tank where it is mixed with recycle lixiviant (and intermittently with sand filter backwash) and additional concentrated carbonate reagent makeup to approximately 25-30 wt.% slurry. The dewatering filtrate (Filtrate 1) from the belt filter flows to the Filtrate Storage Tank and subsequent feed to the uranium recovery circuit.

The repulped slurry is pumped to a second reactor train for additional leaching. This train also consists of three stages of reactors using pumping-type agitators and pumped slurry circulation and advancement. Oxygen gas sparging is also used to maximize uranium leaching efficiency. The slurry exiting the last reactor stage is pumped to the Belt Filter 2 feed tank.

In addition to an initial vacuum dewatering section, the belt filter system for the slurry contains two washing/rinsing sections. In the first washing section, the dewatered cake is rinsed using recycle lixiviant to remove most of the solubilized uranium. The second rinsing step uses fresh water to further remove the uranium and complexing reagents in the soil filter cake. The relatively high percentage solids in the filter cake allows relatively low volumes of washing and rinsing liquids to be used while maintaining high rinse efficiencies. Flocculant (or coagulating agent) is also used, as appropriate, to facilitate the filtering and washing process.

The dewatering filtrate (Filtrate 2) from the second belt filter is recycled to the rotary drum scrubbers to remove surface uranium contamination from the coarse soil particles and to create the slurry feed to the first reactor train. The wash filtrate using recycle lixiviant (Filtrate 3) is recycled to the Repulp Tank 1 where it is mixed with the first belt filter cake to create the slurry feed to the second reactor train. The rinse filtrate using fresh water (Filtrate 4) reports to the Recycle Lixiviant Storage Tank or to the Sand Filter Backwash Tank as needed.

This routing of the filtrate and recycle lixiviant streams allows control of any fresh water makeup and minimizes the lixiviant bleed requirements from the system. It also achieves a true counter-current leaching system which minimizes internal process system flow rates and maximizes the uranium concentration in the liquid feed to the uranium recovery and removal systems. The washed and rinsed fine soil (-2 mm) filter cake discharges from the second belt filter and is conveyed to a stockpile for return to the site for disposal. A stabilization treatment (by addition of stabilizing chemicals) and a dryer to further reduce the soil filter cake moisture to a desired disposal level may follow, as required.

URANIUM RECOVERY SYSTEM (IX LOADING SYSTEM):

For the soil decontamination system, the simple fixed-bed carrousel IX system for removal and recovery of the solubilized uranium is proposed. This will permit significant recycle (greater than 90%) of the lixiviant to be recycled and reused after uranium removal. This, and the limited use of fresh water makeup, will also minimize solution bleed and subsequent treatment requirements. Ion-exchange for uranium removal from carbonate lixiviants is a proven system being used commercially for over twenty years.

The filtrates used as feed to the ion-exchange system are filtered in sand (multi-media) filters to remove any suspended solids or turbidity. These sand filters operate in a continuous, alternate filtering and backwash mode. The intermittent backwash returns to the repulp tank for the second leaching train. The clarified pregnant leach solution proceeds under pumping pressure to the ion-exchange system feed tank which provides some surge capacity in the uranium recovery system.

The solutions are then pumped through down-flow, fixed-bed carbon guard columns which remove most of any dissolved or suspended organics (e.g. humic or fulvic acids, etc.) which may foul the uranium ion-exchange resins. These carbon guard columns also operate in parallel with an alternating loading and stripping cycle.

A carbon regeneration system equipped with a storage tank with vent scrubber, a circulation system and steam-assisted carbon stripping provided to regenerate the carbon columns. This guard column system specific design requirements have not been defined at this point. It is likely that additional unit operations and bleeds of solid and/or liquid waste streams to disposal or to solution treatment would be required. The carbon itself may have to be replaced periodically with fresh carbon and regenerated off-line with a more severe treatment (such as solvent extraction or kiln regeneration) and subsequently recycled.

If an alternate resin is used in the guard column instead of carbon, the regenerate system requirements would also likely differ from those of the carbon columns.

The pressurized solutions from the guard columns continue as feed to three fixed-bed ion exchange columns in the Loading Ion Exchange system. Sufficient feed pump pressure is provided to force the solution through all of the fixed beds in series without requiring boosting. The columns are configured as a carrousel which operates as two or three stages in series for loading. About half the time, the first stage with loaded resin is by-passed and is in a stripping cycle. The loading continues with the former second stage becoming the new first stage and the third stage becoming the new second stage.

When breakthrough occurs in the first of the three stages (i.e. the uranium concentration on discharge from the column is about 10% of the feed), it is taken out of service for stripping. In a fixed-bed ion exchange system for uranium, this occurs when the resin is loaded to about 90% of its maximum loading. When stripping is completed, the freshly-stripped column is restored to the series train as the new third stage.

STRIPPING AND STRIP SOLUTION MAKEUP:

The uranium is stripped from the loaded resin using a sodium chloride/dilute hydrochloric acid strip solution. This stripping system which should be effective for uranium carbonate complexes consists of approximately 1.0 molar (75 gpl) sodium chloride and 0.1 molar (5.0 gpl) hydrochloric acid. The flow through the columns is downflow at a rate of about 0.1 gpm/ft^2 specific flow rate which is significantly less than the loading specific flow rate (of 2.0 gpm/ft^2). This insures equilibrium stripping. About 5 bed volumes of strip solution would be required to strip the resin.

An additional 1 bed volume of fresh water is typically used as a rinse when an acidic strip is used with a basic loading solution. Most of this rinse water is displaced into the strip solution storage tank by the initial fill with uranium-depleted lixiviant from the second ion exchange column upon reintroduction of the freshly stripped column into the loading system as the new third series stage. The balance of the rinse commingled with the uranium depleted solution reports to the recycle lixiviant tank. The stripping cycle proceeds intermittently about half of the column system operating time.

The pregnant strip solution and displaced rinse is stored in two pregnant solution storage tanks operating in parallel, one being filled and the other being fed to the precipitation circuit. Pregnant solution storage capacity provides surge in the operation and allows uncoupling of the loading circuits from the precipitation and recovery circuits. The surge tanks are vented to permit carbon dioxide gas evolution (from the uranium complex stripped from the resin) and have a pumped circulation loop to homogenize the contents for feed to precipitation.

Makeup of the strip solution uses the precipitation system decant and filtrate as the primary solution for stripping. It is regenerated by salt addition and/or hydrochloric acid adjustment of the pH to that optimum for stripping and precipitation system feed (pH = 2.0 to 2.5). The mildly acidic strip solution not only recovers the uranium complex loaded on the resin by mass action, but also will clean the resin and remove some resin fouling. In addition, the resin is regenerated in the chloride form which is optimal for loading of the uranium dicarbonate or tricarbonate complexes (as anions).

The strip solution makeup system consists of two agitated mixing tanks in series (one being filled and mixed while the other is feeding the strip circuit). Solid salt is fed from a bulk hopper to the mix tanks as required. Concentrated hydrochloric acid is metered into the mixing tanks to adjust the pH. A small bleed (about 10-15% of the solution recycle) to the waste water treatment systems from the strip and precipitation circuits will likely to be required due to the fresh water addition to the resin rinse and build-up of sodium chloride and metallic ions other than uranium.

PEROXIDE URANIUM PRECIPITATION:

A hydrogen peroxide precipitation system is used to remove uranium from the acidic pregnant strip solutions. Not only is this system the most compatible with the acidic salt strip system, but it should be the most efficient for uranium removal from the strip solutions. The peroxide precipitation system will also maximize the amount of recycle strip solution which can be used, thus minimizing the waste water treatment requirements for strip solution bleed.

The peroxide precipitation system can be operated continuously or in a semi-batch mode in campaigns using the surge capacity of the pregnant strip solution tanks as a buffer between the upstream systems and the uranium disposal systems. In either case, the critical factors are precipitation reactor residence times, slurry recycle as precipitation seed and pH control. Typically, the precipitation system is designed for double the continuous flow rate and operated about 50% of the time in semi-continuous campaigns. This also permits continuous, closely-coupled operation with the loading systems when longer residence times to complete the precipitation with high uranium removal efficiencies are required.

The peroxide precipitation reactor train consists of four or five separate chambers in series with internal cascade overflow weirs separating the stages. About 90 minutes residence time per stage (based on new feed) is provided for operation at one-half of the time. Each reactor stage is agitated by axial-flow impellers with variable-speed drives which are regulated to balance the slurry suspension and mixing with the need to promote crystal growth of the precipitate. Circulating measurement loops with small centrifugal pumps are provided for each stage to facilitate solution monitoring, sampling and control of reagent additions.

In the first stage the feed pregnant solution (IX pregnant strip solution) at a pH of approximately 2.5 to 3.0 is mixed with hydrogen peroxide (as 50% H_2O_2) in approximately a ratio of approximately 2 to 4 times stoichiometric for uranium precipitation. This translates to approximately 0.15 to 0.30 lbs H_2O_2 (100%) per pound of uranium in the feed solution. The peroxide is fed to the stage using a metering pump which delivers it to the circulating measuring pump loop discharge leg to promote efficient mixing and to prevent concentrated peroxide from coming in contact with the bulk slurry.

Recycle peroxide precipitate slurry from the thickener underflow (uranium content typically 100% of new feed) is also added to the first reactor stage to serve as seed for the precipitation. These recycle seed solids will optimize the precipitation efficiency and produce a larger-diameter uranyl peroxide precipitant which is readily settled and filtered.

As the precipitation reaction proceeds, the pH drops slightly (to 2.5 to 2.75) as acid is liberated. This lower pH accelerates the precipitation rate. The pH continues to drop in the reactor train until next to last stage. If necessary, the pH is controlled in the first stages to a minimum of pH = 2.0 by diversion of some sodium hydroxide from the last stage to prevent redissolution of the uranium precipitate.

In the last reactor stage, the pH is raised to lower the uranyl peroxide solubility and to complete the precipitation from solution. The pH is raised by metering NaOH (30% solution) under pH control to a pH or 4.5 to 5.0. About 1.0 lb of NaOH per lb of uranium is typically required.

The reaction products from the peroxide precipitation produces solid uranyl peroxide and additional sodium chloride in the liquid phase. This precipitation process is the most compatible with the carbonate IX loading and stripping system since typically only HCl is required to regenerate the uranium peroxide thickener decant and pressure filter filtrate back into IX strip solution.

The uranium peroxide slurry from the last precipitation stage is pumped to a clarifier/thickener to facilitate separation of the liquid phase from the precipitated solids. If necessary, a flocculating polymer can be added and mixed with the clarifier feed slurry using an in-line static mixer element. The conventional clarifier/thickener underflow settled uranyl peroxide slurry (25-50% solids) is periodically pumped to a batch recessed plate and frame filter press for dewatering and disposal. The filter press also receives backwash slurry (yellowcake and precoat filter aid) from the clarifying precoat filter. This backwash cycle is done before initiation of a campaign on the plate and frame filter to provide a precoat on the filter cloth.

Some of the settled thickener underflow slurry is recycled to the first stage of the peroxide precipitation reactor train when that system is operating. The clarifier/thickener decant overflows to a pump tank and is recycled to the strip solution makeup through the precoat filter. The filtrate from the plate and frame pressure filter is also clarified in the precoat leaf filter before being recycled to use as strip solution. This final filtration is necessary since any residual solid uranyl peroxide yellowcake solids would be redissolved in the strip solution makeup mix system and would reduce the effectiveness of the strip system. This precoat filter also prevents any precipitated uranium solids from being in any strip solution bleed solutions.

The precoat filter system has a precoat mix tank (for filter aids such as diatomaceous earth) which are periodically mixed by bag addition. Clarifier/thickener decant is diverted, as necessary to mix the precoat slurry. The clarified precoat filter filtrate reports to one of the IX strip solution mixing/feed tanks to be reconstituted as strip solution. Since the entire strip and precipitation system can operate in a semi-continuous mode, there is significant flexibility of operation in the uranium recovery circuit.

PROCESS DESIGN ASSUMPTIONS
REVISED CARBONATE/BICARBONATE SOIL WASHING

The following preliminary process design assumptions were used in flowsheet development, mass balance derivation, equipment selection and preliminary sizing:

COARSE SOIL SEPARATION:

Nominal throughput rate 20.0 dry tons soil/hour.

Soil moisture content average of 12.0 wt.%.

Filtrate, recycle lixiviant and fresh water addition to the scrubbers and 2 mm screen are controlled such that the feed soil slurry density to the leach circuit is nominally 20-30 wt.% solids (design based on 22.5%).

Coarse oversize soil (+100 mm rocks, roots, etc.) is assumed to be about 1.0 wt.% soil.

Medium size soil (-100 mm + 13 mm) is assumed to be 1.5 wt.% soil.

Intermediate size soil fractions (-13 mm + 2 mm) is assumed to be 7.5 wt.% of the feed soil.

CARBONATE LEACHING CIRCUIT:

Feed to the carbonate Leach Train 1 @22.5 wt.% solids.

Residence time (minimum working volume) of 30 minutes in holding tank.

Residence time per stage of leach ≈20 minutes (Train 1).

Horizontal vacuum belt filter cake moisture of 50.0 wt.%.

Horizontal vacuum belt filter design unit area of 80.0 $lbs/hr/ft^2$ for dewatering.

Repulped filter cake feed to Leach Train 2 @ 30.0 wt.% solids.

Residence time per stage of leach ≈30 minutes (Train 2).

Belt filter cake solution acceptance rate for washing/rinsing is 0.080 gpm/ft^2.

Dry flocculant addition system based on a total 2.0 lbs flocculant/ton soil solids.

URANIUM RECOVERY SYSTEM (IX LOADING SYSTEM):

Filtrate 1 storage tank (solution to IX) based on 2 hours residence time (@333 gpm).

Filtrate 2 storage tank (solution to scrubbers) based on 2 hours residence time (@170 gpm).

Filtrate 3 storage tank (solution to repulp tank) based on 2 hours residence time (@170 gpm).

Recycle lixiviant tank based on 2 hours residence time (@333 gpm).

Sand filter backwash tank based on minimum of 2 wetted volume backwash cycles.

Sand filter specific flow rate of 5.0 gpm/ft^2. Bed height 6 ft.

Guard column specific flow rate of 2.5 gpm/ft^2. Bed height 6 ft.

Ion exchange column specific flow rate of 2.0 gpm/ft^2. Bed height 4 ft.

Ion-exchange maximum loading of 100 lbs uranium/ton resin.

Resin replacement rate nominally 3% of inventory/year.

Materials of construction: 316 S.S. or rubber lined C.S.

STRIPPING AND STRIP SOLUTION MAKEUP:

Carbon replacement rate nominally 10% of inventory/year.

Bed volumes of strip solution, 5 design (7 max.).

Bed volumes of fresh water rinse, 1 design (2 max.).

Resin strip solution nominally 1.0 molar NaCl, 0.10 molar HCl, pH = 2.5-3.0.

Strip solution specific flow rate, nominal 0.10 gpm/ft^2, maximum 0.20 gpm/ft^2.

Working volume 8,100 gallons each of two.

Materials of construction: 316 S.S. or rubber lined C.S.

PEROXIDE URANIUM PRECIPITATION:

Design feed rate 4.78 gpm (half-time operation).

Residence time/stage = 90.0 minutes.

Limit for pH (minimum) = 2.0.

Materials of construction: HDPE, fiberglass or rubber lined C.S.

Hydrogen peroxide feed to: tank 1 and tank 2.

NaOH feed to: tank 4 and tank 5.

Thickener U/F slurry recycle: nominally 100% new feed uranium, range: 0-400%.

Thickener unit area: based on 1.0 m/hr fall velocity and maximum 30 gpm feed rate (0.40 gpm/ft^2 specific flow rate).

Thickener U/F density: nominal 40 wt.% solid, (range: 25-50%).

Recessed plate & frame uranium peroxide filter unit area based on 27.3 ft^3 net cake capacity, one filtration cycle/day.

Precoat filter unit area based on 1.0 gpm/ft^2 specific flow rate, 50 gpm maximum feed rate. One backflush cycle/day as precoat to recessed plat & frame filter.

Bleed rate: lixiviant ≈8.5% C.L. (range: 5-10%), strip solution ≈10.0 % C.L.

SHEET 1

CARBONATE/BICARBONATE URANIUM-CONTAMINATED SOIL WASHING
20.0 DTPH SOIL INPUT, COARSE SEPARATION AND LEACHING

09-Jan-95
ISSUE 2
Rev. 1

STREAM NUMBER	STREAM DESCRIPTION (BASED ON FLEXMET BALANCES)	SOLIDS S.G.	LIQUID S.G.	SLURRY S.G.	BULK DENSITY	WT.% SOLIDS	SOLIDS (STPH)	LIQUID (STPH)	TOTAL (STPH)	TOT. GPM (yd3/HR)	STREAM NUMBER
1	SOIL FEED TO GRIZZLY	2.500	1.000	2.119	90.0	88.00	20.0000	2.7270	22.7270	(18.70)	1
2	GRIZZLY OVERSIZE (+4") TO SCRUBBER 1	2.500	1.000	2.119	132.1	88.00	0.5000	0.0682	0.5682	(0.32)	2
3	GRIZZLY UNDERSIZE (-4") TO SCRUBBER 2	2.500	1.000	2.119	132.1	88.00	19.5000	2.6588	22.1588	41.78	3
4	TROMMEL OVERSIZE (+1/2") FROM SCRUBBER 1	2.500	1.000	1.923	84.0	80.00	0.2000	0.0500	0.2500	(0.22)	4
5	TROMMEL OVERSIZE (+1/2") FROM SCRUBBER 2	2.500	1.000	1.923	84.0	80.00	0.3000	0.0750	0.3750	(0.33)	5
101	SCRUBBER 1 WASH WATER	—	1.000	1.000	62.4	0.00	0.0000	5.0080	5.0080	20.01	101
102	SCRUBBER 2 WASH WATER	—	1.000	1.000	62.4	0.00	0.0000	5.0080	5.0080	20.01	102
50	WET GRIZZLY SPRAY CARBONATE SOLUTION	—	1.015	1.015	63.3	0.00	0.0000	2.5040	2.5040	9.86	50
51	SCRUBBER 1 RECYCLE CARBONATE SOLUTION	—	1.015	1.015	63.3	0.00	0.0000	22.3260	22.3260	87.88	51
52	SCRUBBER 2 RECYCLE CARBONATE SOLUTION	—	1.015	1.015	63.3	0.00	0.0000	22.3260	22.3260	87.88	52
6	WASHING & DEWATERING SCREEN FEED	2.500	1.015	1.189	74.1	24.60	19.5000	59.7730	79.2730	266.45	6
7	SCREEN O/S (+2mm) TO STOCKPILE	2.500	1.000	1.923	84.0	80.00	1.5000	0.3750	1.8750	3.90	7
8	SCREEN U/S (-2mm) TO HOLDING TANK	2.500	1.015	1.178	73.4	23.26	18.0000	59.3980	77.3980	262.58	8
103	SCREEN WASH WATER	—	1.000	1.000	62.4	0.00	0.0000	2.6010	2.6010	10.39	103
9	COMBINED (+2mm) O/S TO DISPOSAL STOCKPILE	2.500	1.000	1.923	119.9	80.00	2.0000	0.5000	2.5000	5.19	9
61	CARBONATE/BICARBONATE REAGENT TO TRAIN 1	2.500	1.212	1.212	75.6	0.00	0.0000	5.1320	5.1320	16.91	61
10	NET FEED TO LEACH TRAIN 1	2.500	1.027	1.183	73.8	22.50	18.0000	61.9900	79.9900	270.01	10
11	LEACH TRAIN 1 DISCHARGE TO BELT FILTER 1	2.500	1.027	1.173	73.1	21.14	18.0000	67.1300	85.1300	290.01	11
12	FILTER CAKE BELT FILTER 1	2.500	1.027	1.455	90.8	50.00	18.0180	18.0180	36.0360	98.91	12
13	FILTRATE 1 TO IX COLUMN	2.500	1.027	1.027	64.0	0.00	0.0000	71.2200	71.2200	277.17	13
63	DRY FLOCCULANT	1.800	1.000	1.800	112.3	100.00	0.0360	0.0000	0.0360	(0.0238)	63
53	DRY FLOC DILUTION (RECYCLE LIXIVIANT)	—	1.000	1.000	62.4	0.00	0.0000	44.2160	44.2160	176.64	53
64	DILUTED FLOCCULANT TO BELT FILTER 1	1.800	1.015	1.015	63.3	0.08	0.0180	22.1080	22.1260	87.07	64
65	DILUTED FLOCCULANT TO BELT FILTER 2	1.800	1.015	1.015	63.3	0.08	0.0180	22.1080	22.1260	87.07	65
54	RECYCLED FILTRATE 3 FOR REPULP	2.500	1.015	1.015	73.1	0.00	0.0000	44.2160	44.2160	174.06	54
62	CARBONATE/BICARBONATE REAGENT TO TRAIN 2	2.500	1.212	1.212	75.6	0.00	0.0000	1.3320	1.3320	4.39	62
14	FEED TO LEACH TRAIN 2	2.500	1.015	1.235	77.0	30.00	18.0180	42.0410	60.0590	194.29	14
15	FEED TO BELT FILTER 2	2.500	1.015	1.235	77.0	30.00	18.0180	42.0410	60.0590	194.29	15
17	FILTRATE 2 RECYCLE TO SCRUBBERS	2.500	1.015	1.015	63.3	0.00	0.0000	46.1120	46.1120	181.52	17
18	FITRATE 3 RECYCLE TO REPULP 1	2.500	1.015	1.015	63.3	0.00	0.0000	36.0750	36.0750	142.01	18
19	FILTRATE 4 TO SAND FILTER BACKWASH & RECYCLE	2.500	1.015	1.015	63.3	0.00	0.0000	18.0360	18.0360	71.00	19
55	RECYCLE LIXIVIANT WASH FOR BELT FILTER 2	2.500	1.015	1.015	63.3	0.00	0.0000	36.0750	36.0750	142.01	55
104	FRESH WATER WASH ON BELT FILTER 2	—	1.000	1.000	62.4	0.00	0.0000	18.0360	18.0360	72.05	104
16	WASHED FILTER CAKE (-2mm) SOIL TO DRYER	2.500	1.000	1.000	62.4	50.00	18.0360	18.0360	36.0720	144.11	16
80	TOTAL LIQUID OXYGEN FEED TO REACTORS	—	1.000	1.000	62.4	0.00	0.0000	0.0217	0.0217	0.09	80
41	PHOSPHATE ADDITION BEFORE DRYER	2.500	1.000	1.000	0.0	0.00	0.0000	0.0000	0.0000	0.00	41
40	DRIED SOIL TO DISPOSAL	2.500	1.000	1.923	119.9	80.00	18.0360	4.5090	22.5450	46.83	40
39	WATER VAPOR FROM DRYER	—	1.000	1.000	62.4	0.00	0.0000	13.5270	13.5270	54.04	39

*NOTE: BASED ON INITIAL MASS BALANCES FOR REVISION 2 CARBONATE/BICARBONATE FLOWSHEET (12/12/94) AND COUNTER-CURRENT, TWO-TRAIN LEACHING

SHEET 2

CARBONATE/BICARBONATE URANIUM-CONTAMINATED SOIL WASHING

20.0 DTPH SOIL INPUT, URANIUM REMOVAL & LIXIVIANT RECYCLE

09-Jan-95

ISSUE 2

Rev. 1

STREAM NUMBER	STREAM DESCRIPTION (BASED ON FLEXMET BALANCES)	SOLIDS S.G.	LIQUID S.G.	SLURRY S.G.	BULK DENSITY	WT.% SOLIDS	SOLIDS (STPH)	LIQUID (STPH)	TOTAL (STPH)	TOT. GPM (yd3/HR)	STREAM NUMBER
50	WET GRIZZLY SPRAY CARBONATE SOLUTION	2.500	1.015	1.015	63.3	0.00	0.0000	2.5040	2.5040	9.86	50
51	SCRUBBER 1 RECYCLE CARBONATE SOLUTION	2.500	1.015	1.015	63.3	0.00	0.0000	22.3260	22.3260	87.88	51
52	SCRUBBER 2 RECYCLE CARBONATE SOLUTION	2.500	1.015	1.015	63.3	0.00	0.0000	22.3260	22.3260	87.88	52
53	DRY FLOC DILUTION (RECYCLE LIXIVIANT)	2.500	1.015	1.015	63.3	0.00	0.0000	44.2160	44.2160	174.06	53
54	RECYCLED FILTRATE 3 FOR REPULP	2.500	1.015	1.015	73.1	0.00	0.0000	44.2160	44.2160	174.06	54
55	RECYCLE LIXIVIANT WASH FOR BELT FILTER 2	2.500	1.015	1.015	63.3	0.00	0.0000	36.0750	36.0750	142.01	55
61	CARBONATE/BICARBONATE REAGENT TO TRAIN 1	2.500	1.212	1.212	75.6	0.00	0.0000	5.1320	5.1320	16.91	61
62	CARBONATE/BICARBONATE REAGENT TO TRAIN 2	2.500	1.212	1.212	75.6	0.00	0.0000	1.3320	1.3320	4.39	62
13	FILTRATE 1 TO IX COLUMN	2.500	1.027	1.027	64.0	0.00	0.0000	71.2200	71.2200	277.17	13
17	FILTRATE 2 RECYCLE TO SCRUBBERS	2.500	1.015	1.015	63.3	0.00	0.0000	46.1120	46.1120	181.52	17
18	FILTRATE 3 RECYCLE TO REPULP 1	2.500	1.015	1.015	63.3	0.00	0.0000	36.0750	36.0750	142.01	18
19	FILTRATE 4 TO SAND FILTER BACKWASH & RECYCLE	2.500	1.015	1.015	63.3	0.00	0.0000	18.0360	18.0360	71.00	19
20	FEED TO SAND FILTERS	2.500	1.015	1.015	63.3	0.00	0.0000	84.6040	84.6040	333.04	20
22	COMBINED FEED TO IX COLUMNS	2.500	1.015	1.015	63.3	0.00	0.0000	84.6040	84.6040	333.04	22
23	IX COLUMN DISCHARGE	2.500	1.015	1.015	63.3	0.00	0.0000	84.6040	84.6040	333.04	23
24	STRIP SOLUTION TO IX COLUMNS	2.500	1.106	1.106	69.0	0.00	0.0000	4.2300	4.2300	15.28	24
25	PREGNANT STRIP SOLUTION FROM IX COLUMNS	2.500	1.106	1.106	0.0	0.00	0.0000	4.2300	4.2300	15.28	25
26	PRECIPITATION CIRCUIT FEED (CONTINUOUS)	2.500	1.106	1.106	69.0	0.00	0.0000	4.2300	4.2300	15.28	26
27	HYDROGEN PEROXIDE FEED (50% SOLN.) TO PPTN.	2.500	1.197	1.197	74.7	0.00	0.0000	0.0030	0.0030	0.010	27
28	NaOH FEED (30% SOLN.) TO PPTN.	2.500	1.327	1.327	82.8	0.00	0.0000	0.0150	0.0150	0.045	28
32	RECYCLE THICKENER U/F AS SEED IN PPTN.	7.600	1.106	1.931	120.4	50.00	0.0931	0.0931	0.1861	0.39	32
29	URANYL PEROXIDE SLURRY TO THICKENER (NET)	7.600	1.106	1.147	71.6	4.21	0.1861	4.2300	4.4161	15.38	29
30	THICKENER DECANT O/F TO PRECOAT FILTER	7.600	1.106	1.106	69.0	0.00	0.0000	4.0439	4.0439	14.61	30
31	THICKENER U/F SLURRY	7.600	1.106	1.931	120.4	50.00	0.1861	0.1861	0.3722	0.77	31
33	FILTER PRESS FEED	7.600	1.106	1.931	120.4	50.00	0.0931	0.0931	0.1861	0.39	33
34	FILTER PRESS FITRATE TO PRECOAT FILTER	7.600	1.106	1.106	69.0	0.00	0.0000	0.0620	0.0620	0.22	34
35	FILTER PRESS URANYL PEROXIDE CAKE TO DISPOSAL	7.600	1.106	3.080	192.1	75.00	0.0931	0.0310	0.1241	(0.048)	35
	NET FEED TO PRECOAT FILTER	7.600	1.106	1.106	69.0	0.00	0.0000	4.1059	4.1059	14.83	
36	NET FEED TO RECYCLE MAKEUP	7.600	1.106	1.106	69.0	0.00	0.0000	3.6953	3.6953	13.35	36
60	BLEED REGENERATE TO AWWT	7.600	1.106	1.106	69.0	0.00	0.0000	0.4106	0.4106	1.48	60
37	SODIUM CHLORIDE TO STRIP SOLUTION MAKEUP	2.200	1.106	2.200	96.0	100.00	0.0411	0.0000	0.0411	0.07	37
38	HYDROCHLORIC ACID TO STRIP SOLUTION MAKEUP	2.200	1.150	1.150	71.7	0.00	0.0000	0.0411	0.0411	0.14	38
104	FRESH WATER IN STRIP MAKEUP	0.000	1.000	1.000	62.4	0.00	0.0000	0.3285	0.3285	1.31	104
56	LIXIVIANT BLEED TO AWWT	2.500	1.150	1.150	71.7	0.00	0.0000	8.2120	8.2120	28.53	56
57	RECYCLE LIX TO CARBONATE MAKEUP	2.500	1.000	1.000	62.4	0.00	0.0000	5.1680	5.1680	20.65	57
58	BULK SODIUM BICARBONATE MAKEUP	2.200	1.000	2.200	54.9	100.00	0.6460	0.0000	0.6460	(0.872)	58
59	BULK SODIUM CARBONATE MAKEUP	2.200	1.000	2.200	54.9	100.00	0.6460	0.0000	0.6460	(0.872)	59

*NOTE: BASED ON INITIAL MASS BALANCES FOR REVISION 2 CARBONATE/BICARBONATE FLOWSHEET (12/12/94) AND COUNTER-CURRENT, TWO-TRAIN LEACHING PRECIPITATION CIRCUIT DESIGNED FOR HALF-TIME OPERATION AT TWICE CONTINUOUS FLOW RATES.

FERNALD ID TECHNOLOGIES

BICARBONATE HEAP LEACH - PERMANENT PAD

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PROCESS DESCRIPTION
REVISED BICARBONATE PERMANENT PAD HEAP LEACH

INTRODUCTION:

In order to provide a basis for the FY 95 ID Program for removal of uranium contamination from soils, revised process flowsheets have been developed which reflect experience in primary uranium ore processing as well as recent laboratory results on the contaminated soils. The approach taken has been to assume a relatively optimistic process performance as a basis for equipment selection and process design. In addition, the unit operations selected should be readily operable at the assumed operating conditions. If adequate performance for the selected process concepts and equipment selected are supported by laboratory tests, the capital and operating cost benefits over previous process flowsheets should be substantial.

The following presents a brief process description of the revised Bicarbonate Permanent Pad Heap Leach flowsheets (Revision 2, 11/21/94) and some key process assumptions used as the basis for initial material balances used for equipment sizing.

COARSE SOIL AND TRASH SEPARATION:

The initial separation of the oversize, coarse size fraction and trash components of the soil (+150 mm) is accomplished using equipment which minimizes the amount of liquid (lixiviant and fresh water) added to the system. The goal is to produce a -150 mm soil feed, without tramp material, which can be transported to and reclaimed from a covered soil stockpile.

The soil feed to the process is initiated by reclaim from an asphalt soil receiving pad where trucks dump the excavated soil or by direct feed from the excavation operations. A loader directly feeds a dry grizzly with the soil and scalps off oversize and trash materials (+150 mm) in the soil. The grizzly is elevated such that the oversize fraction flows by gravity to the drum scrubber and the undersize fractions flow by gravity to the coarse soil feed conveyor. The pad is bermed for containment and has a reclaim sump for drainage liquid and solids.

The oversize material from the grizzly reports to the Rotary Drum Scrubber in which the oversize soil is mixed with recycle lixiviant as a slurry to wash off any adhered, small-size soil and to solubilize uranium staining the surface of the coarse particles using the leaching action. The drum scrubber is equipped with a solid drum section and dewatering drain for slimes removal as well as a trommel screen extension where rinsing with fresh water removes most of any dissolved uranium.

The undersize soil slurry (-150 mm) from the Drum Scrubber flows by gravity as feed to a dewatering screen. The trommel oversize (+150 mm) is discharged to a stockpile for subsequent return to the site for disposal. The turbid water from the dewatering screen (which contains about 1.0% solids in suspension and some dissolved uranium) flows by gravity to a live-bottomed sump and pump which delivers it as feed to turbid water clarifier/thickener for recovery and removal of the suspended solids and recycle of the dissolved uranium with the lixiviant recycle for ultimate recovery.

The screen oversize soil joins the grizzly undersize on the coarse soil stockpile feed/stacking conveyor is transported to the covered stockpile building and stacked in a stockpile on the floor of the building. The soil is deposited into the stockpile during day-time excavation operations consistent with the excavation contractor's schedule (presumably 4 or 5 day week). It has the capacity to store 2,400 tons soil (≈2,075 yds^3) or approximately 3 excavation days inventory.

The covered stockpile building also is equipped with high-lift, garage-type doors which would permit direct dumping and movement of excavated soil to the stockpile, by-passing the oversize grizzly, if desired. Likewise, pad area outside of the stockpile building permits mobile equipment (dump trucks, dozers, loaders, etc.) to move the soil directly into the covered stockpile by direct dumping on the pad. The pad (and inside building floor area) is contained by berm and has a slurry sump and pump for reclaim of drainage liquids or solids from the soil.

A variable-speed feed conveyor (with weigh section) reclaims the soil from the stockpile using an under-the-pile feeding system and delivers it to a transport conveyor at a perscribed rate. This, in turn, delivers the soil to an agglomerating drum for mixing with sand and subsequent delivery to the leaching pad construction.

A diluent bulking agent (sand/gravel) is mixed with the soil to form an agglomerated soil/sand blend which has enhanced net permeability for leachate percolation in the heaps. Laboratory testing has established that an acceptable percolation rate can be achieved with about 20 to 50 net weight percent of the diluent sand added to the clayey soil. An initial permanent pad heap design is based on an assumption of 20% by weight sand (dry basis) added to soil.

The sand is received from trucks dumping on an asphalt receiving pad similar to the soil receiving pad. It is stockpiled (up to one week's inventory) for blending with soil during pad construction. Reclaim of required sand is by loader directly into a conveyor feed hopper and is transported directly to the agglomerating/blending drum. The belt is equipped with a weigh section for rate control.

The agglomeration drum receives soil, sand and, possibly, an agglomerating agent which are mixed and blended in the rotating cylinder to form a blended mixture for leaching. As yet, the type and requirements of the agglomerating agent are unspecified, but it is likely that a surface active or wetting agent along with liquid to control the soil/sand blend moisture content could be used. Pozzolanic agents, such as Portland cement and lime have been tested, but have reduced leaching rate and uranium removal efficiency. Their use, however, have not been optimized. The agglomerated or blended soil/sand mixture is discharged from the agglomerating drum to a cross-country conveyor which transports it to the leach pad for heap construction.

The philosophy of the above coarse ore separation circuit is to produce a feed soil/sand blend which can be transported and stacked in heaps in the heap leaching cells which has no trash or extraneous components. This will facilitate materials handling transport and reclaim from the leach pads after uranium removal by percolation leaching. In addition, the oversize and trash materials in the soil will have been separated and washed with lixiviant to remove adhering fines and surface uranium contamination. This coarse soil fraction would also be rinsed with fresh water prior to return to the site.

The simplified approach to handling and storage of the excavated soil using a cylindrical, covered soil storage building with an under-the-pile reclaim system similar to that used in the mining industries would not only present a significantly-less costly alternative to the previously proposed soil storage building, but also could satisfy all of the environmental concerns and constraints for soil handling. This alternative also should require less manpower to operate to feed soil to the downstream processing. It could be considered for any of the ID technologies. for uranium removal.

The above equipment should be capable of this goal without the need for mechanical dewatering prior to leaching and also should minimize slurry pumping. Use of the leaching solution (i.e. Belt Filter 2 filtrate) from the Leaching Train 2 as the primary rinsing and motive liquid for the soil slurry achieves a true counter-current and high-efficiency leaching system.

CONVENTIONAL PERMANENT PAD HEAP LEACHING CIRCUIT:

The heap leaching strategy proposed for a conventional, permanent pad, movable heap leaching process for uranium removal from contaminated soils is discussed below. The definition of the "permanent pad, movable heap" component is that a reusable leach pad area is constructed and the soil to be leached is moved onto and stacked on the pad where it is leached and rinsed. It is then removed from the pad and placed in a new, environmentally-controlled stockpile or returned to the excavation site for backfill.

The "permanent pad" which is designed to contain and recover the leaching solutions is loaded and reused many times during the project operation. Typically the pad is made of asphalt with berms, perforated drainage piping, sumps and pumps for solution containment and control. The pad is covered with a crushed rock protective layer (12-18" thick) to prevent damage by mobile equipment and to provide a permeable drainage layer to facilitate leach solution recovery.

This strategy contrasts with the alternative conventional heap leach strategy of "permanent heap" construction. In the "permanent heap" scenario an impermeable pad is prepared, usually using a plastic membrane or clay liner material, to hold the heaps and to contain and recover the leaching solutions. This pad is also protected by a gravel or crushed rock layer which facilitates solution drainage. This may also be assisted by incorporating perforated piping in the coarse rock layer. The soil (or ore in minerals processing) is stacked onto the pad using conveyor transport and stacking or by dumping from trucks and mounding into heaps using low-bearing pressure dozers and loaders. The permanent heaps remain in place after the leaching cycle and the soil or rock is stabilized in place.

The permanent pad, movable heap scenario typically uses heaps which are relatively thin layer (6' to 12' thick) as compared to the permanent heaps which are relatively thick (20' to 30'). This thickness of leaching material in the heap effects the leaching cycle (i.e. the duration of time for leaching) and the solution application strategy (i.e. the fraction of time during the leaching when leach solution is being applied to the top of the heap).

Permanent Heap

The permanent heaps typically are leached for a relatively-long time duration (60 to 180 days). Leaching solution may be applied intermittently or with low specific application rates during the leaching cycle. The permanent heaps usually are constructed with more than one "lift" which is more than one layer of material being leached. The permanent heap after a leaching cycle is "overdumped" with additional soil (or ore) material and a new heap layer is constructed. Drainage pipes between lifts may be used to facilitate solution recovery and new spray or solution application systems are installed on the top of the new lift.

The leaching is re-initiated and the fresh material leached as before by starting application of leaching solutions to the top of the upper heap layer. However, there will be some percolation of leach solutions into the underlying layer(s). Therefore, additional solution contact and leaching will continue in the lower layers of the heap and those solutions will be recovered by the original, impermeable pad. Thus, higher ultimate levels of leach extraction are achieved with the multiple-layer, permanent heap leaching scenario.

A primary advantage of the multi-lift, permanent heap is that less real estate is impacted by the heap leaching operation. In addition, the pad liner and containment systems continue to protect the environment from rainwater percolation through the spent heaps. This scenario also has the advantage in that the soil (or ore) is only moved once; that is, to put it onto the heap. The higher heap structures (typically up to 80′ thick with multiple lifts) also would result in less area to permanently stabilize upon closure of leaching operations. Such heaps are conditioned with stabilization agents at the surface, covered with top soil and revegetated with ground cover and root-structure plantings. Since commercial ore mining operations, where such heap leaching is typically practiced, are usually in remote locations, the above practices and environment requirements are satisfactory. However, in a semi-urban or farming environment, additional restrictions and requirements could eliminate a conventional, permanent heap leaching strategy from consideration.

Permanent Pad

The permanent pad, movable heap scenario has a number of advantages for heap leaching of contaminated soils in comparison to ore leaching operations. Although the permanent pad practice is used in ore leaching operations (thin-layer leaching for copper and gold ores), the permanent heap scenario is the typical method. However, when the permeability of the ore is low (for mill tailings leaching, fine crushed ore or for clayey ore types) or when a high exposure to air for oxidation is required (e.g. for sulfide copper ores), the permanent pad, movable heap scenario is the preferred method. Likewise where there is limited space for permanent heaps or where the leached residue material is required for backfilling the excavation (surface or underground mine), the permanent pad scenario has been favored.

The permanent pad, conventional heap leach scenario for extraction of uranium from contaminated soils proposed here utilizes a 5-cell leaching strategy. A cell is a specifically designated, segregated and separately-operated leaching area on the pad. There are three cells where active leaching, i.e. leach solutions are being applied and recovered, is going on at any time. There is one cell where new soil is being transported to, stacked on and being prepared for leaching. This, for example would include: leveling out of the top of the soil heap, construction or laying out of the leaching solution distribution and spray system piping.

The last cell would contain the soil which has finished its three leaching cycles, a rinse cycle and has drained. After draining of most of the contained water within the heap, the leached soil would be reclaimed and transported to a holding pad for disposal. After leached soil removal, the cell would be prepared to receive fresh soil during the next cycle where it would be loaded to repeat the leaching cycles.

For purposes of design, to facilitate operational cycles and to support the laboratory data on the time requirements for leaching of the uranium in the soil, a seven-day cycle duration was chosen. For three leach cycles, this would provide at least 18 days of active leaching of the soil. The balance of the time for leaching ($\approx$3 days) would be used for applying rinse water and initiation of a drainage cycle to allow dewatering of the heaps. The seven-day cycle is also consistent with the time required for heap leaching which ranged in the laboratory from 14 days to 20 days for the practical leach time limit.

The primary rational for the seven-day cycle is that it is very compatible with a normal 4 or 5-day work week by an excavation and earth-moving contractor. What this permits is that the soil would be excavated during day shift only (on a 4 or 5-day basis) and would be transported by truck to the soil stockpile. During this time period, the new heap cell is being loaded with fresh soil and the last leached cell is being unloaded of leached soil. The excavation rate, the pad loading rate and the pad unloading rate need not match exactly. The new soil stockpile (receiving pad and storage building) and the leached soil receiving pad provide a buffer and surge between the materials-handling operations.

At an overall weekly average of 20 dry tons per hour (dtph) soil throughput rate, the excavation contractor may be operating and delivering soil to the receiving pad at a dynamic rate of 112 dtph for say 30 actual operating hours per week. Likewise the agglomeration/blending and pad loading operations may also be operating at the same or different operating rate but on a different operating schedule than the excavation activities. The pads could be loaded in a weekly cycle for 30 hours at a dynamic rate of movement of soil of 112 dtph or it could be loaded in 20 actual hours at 168 dtph rate, etc. Likewise, the leached soil reclaim and transport operations could be done with a different schedule or even different days of the week by the same contractor.

The 3-cell strategy is to load one weeks worth of excavated soil in one cell during the first week, leach it for three cycles (or about three weeks) and then rinse, drain and remove it from the cell to a staging pad for disposal during the fifth week. The rates and timing of the material movement to and from the leach pad can be flexibly organized to fit the excavation contractor and the soil materials handling requirements to and from the leaching pad. In addition, minimum interference with the leaching operations would result.

The primary advantage of the 3-cell leach and 3 operating cycles is that it permits "stacking" or counter-current movement of solutions through the soil heaps. In the "stacking" arrangement proposed in the design, fresh soil is first leached with pregnant solutions from previous leaching cycles (2). The most easily leached uranium is, therefore, easily extracted using the "intermediate preg" solution as the lixiviant in the first cycle (6-7 days). The bicarbonate-based lixiviants are partially consumed during leaching. Therefore, the most concentrated soil would be contacted by the lixiviant which is most depleted in complexing agent.

The resulting leach solutions collected at the bottom of the heap and in the solution sump would be the "pregnant solution" advanced to the recovery/removal system for uranium. It would first be pumped to a "pregnant solution" pond located adjacent to the leach pad for surge and storage and then reclaimed and delivered to the pregnant solution storage tank in the uranium recovery circuit.

The soil, with its residual levels of uranium, would now be leached in a second leach cycle (6-7 days) using "weak preg" solution resulting from a third leaching cycle on another cell. This would then produce the "intermediate preg" solution with intermediate concentrations of uranium in solution which would be used as the lixiviant for the fresh soil in a first leaching cycle. It would be collected in the cell sump and pumped for storage into the "intermediate preg" pond until require for heap application.

The residual soil in a specific cell on the leach pad after two cycles of leaching is now leached using fresh lixiviant (i.e. with little residual uranium and with freshly reconstituted lixiviant strength). Therefore, the counter-current nature of "stacking" the solutions would now leach the most-difficult-to-leach uranium with the strongest (and least concentrated in uranium) lixiviant solution. The produced solutions from this cycle of leaching are the "weak preg" solutions used in the second cycle of leaching. They are collected and pumped for interim storage in the "weak preg" pond.

The counter-current strategy not only most efficiently leaches the out the uranium in the soil, but it significantly reduces the volume of produced pregnant lixiviant which needs to be processed to remove the solubilized uranium. Compared with parallel (not counter-current in series) leaching of three cells of soil approximately one-third the amount of produced solutions would result which would need subsequent treatment and uranium recovery. The concentration of the solubilized uranium in the resulting pregnant leachate would also be approximately three times that for parallel leaching of the three cells. The recovery circuit, in this case solid ion exchange, would also work more efficiently if the solution tenor feed to the ion exchange were higher.

Leach Pad Design

Some description of the leach pad and individual cells have been provided above, the following provides the specific description of the permanent pad for soil leaching.

The pad itself will be constructed of multiple layers of asphalt (6-9" thick total) laid down over a compacted layer of clay or soil as a back up impermeable barrier to solution migration. This is covered with a geotextile fabric and secondary impermeable barrier of fuse-welded HDPE sheet (40 mils). This secondary liner extends beyond the asphalt pad layers and is rolled into an earthen berm surrounding the leach pad to provide secondary containment. Any drainage from the secondary liner is directed toward a sump with pump located at the lowest point of the sloped pad. There any pad run-off or rainwater would be pumped into the recycle lixiviant pond. Rolled asphalt berms approximately 6" high surround the edges of the pad and separating the cells from each other.

To contain one week soil inventory (@20 dtph average soil processing rate), one week of the associated sand or gravel as the bulking agent to improve heap permeability and to allow about 5 to 10 feet of pad area around the heap, each cell would be 165'L x 85'W. Each cell would hold about 4,230 tons soil/sand mixture and would have 8,460 ft^2 of exposed top surface for lixiviant application. The total pad area would be 165' x 430'. The asphalt pad is sloped (≈5°) toward one side (with the soil reclaim conveyor) and each cell on the pad is sloped to the center. Lateral drainage pipes would direct leach solutions to the preg sump and pump after they percolate to the bottom of the heap.

Leachate is applied to the surface of the soil heaps using irrigation drip emitters which slowly apply the solution in a distributed fashion. Typically, each emitter applies up to 5 gallons per hour and are spaced about 18" to 2' away from each other on HDPE plastic lines spread from lateral distributor headers located on the ends of the cells (parallel to the loading/reclaim conveyors). Leach solution flow to each cell is controlled and monitored (i.e. metered) by separate piping and valve systems for each of the applied solutions. Piping systems for delivering fresh water, fresh lixiviant, weak preg lixiviant and intermediate preg lixiviant are provided for each cell (HDPE or PVC). Similarly, piping and valving systems to direct the solution collection sump pump to the proper storage pond are also provided for each option.

Fresh lixiviant is regenerated by addition of carbon dioxide (as liquified carbonic acid) to the recycle lixiviant in a by-pass mixing tank. Provisions to add bulk sodium bicarbonate are also provided (volumetric feeder and mixing tank). The concentrated bicarbonate solution from the mixing tank is metered into the recycle lixiviant to produce the correct concentration of fresh lixiviant on the way to application on the heap.

Pad Loading and Unloading

A cross-country conveyor system delivers the soil/sand blend discharged from the rotary drum agglomerator to the leaching pad cells. It is runs down one side of the cells and pad. A travelling tripper is moved along the cross-country conveyor and positioned at the cell which is to be loaded with soil. This tripper discharges the soil/sand mixture unto a series of portable conveyor sections which transports it along the length of the cell and delivers it to a stacking conveyer which elevates the soil and stacks. The stacker has a telescoping section which moves back and forth to spread the soil/sand mixture into the desired heap profile and approximately 10' in height. The rubber-tired stacking conveyor section is also periodically manually moved laterally to provide lateral "row" of stacked soil.

This loading process continues until the cell has a "row" of soil/sand blend about 10' wide by 60' long laterally across the end of the cell farther from the cross-country transport conveyor. The portable conveyors and stacker are then moved closer to the empty end of the cell and a new "row" of soil/sand blend is begun. This loading process continues and retreating with the portable conveyors toward the empty end until the cell is fully loaded. The top of the heap is leveled using low-bearing pressure equipment. The leachate lateral and emitter distributors are arranged along the top of the heap and connected to the lixiviant supply headers located along each end of the cells (along side of the pad). The new soil/sand cell is now ready to initiate leaching.

The emitters are used since they will eliminate any potential for wind blown liquid spray (such as would result from sprinklers) and they can be spaced to insure uniform solution coverage over the surface and subsurface in the heap. Total flow and distribution of flow over the entire heap can also be precisely controlled. Although only applied as a point source every 1½ to 2', through the capillary action of the soil and the nature of percolation leaching nearly 99% of the soil will be contacted by the lixiviant. Even areas on the surface between the emitters will get some exposure to leachate and will be leached due to the "wicking" and lateral movement of solutions. Any surface rainfall will also distribute lixiviant and flush the surface of dissolved uranium.

Pad Unloading

After the three leaching cycles are completed, the heaps are rinsed with fresh water which is applied through the same emitters. However, since there are less concerns about "ponding" or surface collection of solutions on the surface, the fresh water application rate for rinsing is typically 1½ to 2 times the leaching rate application. This also insures flushing of the surface soil. The rinse application cycle continues for 1 to 2 days and is followed by 1 to 2 days of heap drainage. During active heap leaching, the heap is only 40-60% (by volume) saturated with liquid. It will drain naturally to about 30% liquid by volume or about 20% by weight.

After drainage, the soil is reclaimed by front-end loader which places the soil into a portable conveyor feed hopper where it is transported and discharged onto the cross-country reclaim transport conveyor. The unloading procedure is the reverse of the cell loading in that the end of the cell closest to the cross-country conveyor is loaded first and additional portable sections are added to the cell reclaim conveyor system as needed. This unloading of the soil continues until the cell is empty. The layer of crushed rock for drainage is not disturbed during the reclaim since it will serve as the base for the next load of soil on the heap.

The leached soil cross-country conveyor discharges to a leached soil pad where the soil is stockpiled prior to return to the disposal site. Alternatively, it can discharge directly into the transport trucks on the leached soil pad which transport the leached soil back to the disposal site. Control of the excavation, soil stockpile and reclaim/disposal operations, schedule and transport rates can maximize the use of moving equipment associated with excavation and transport of the soil. The same trucks which bring newly-excavated soil to the soil receiving pad can be loaded with leached soil for the return trip to the disposal site. Alternatively, excavation can be done for 2-3 days of the week and return of leached soil for disposal can be done also for 2-3 days of the week.

URANIUM RECOVERY SYSTEM (IX LOADING SYSTEM):

For the soil decontamination system, the simple fixed-bed carrousel IX system for removal and recovery of the solubilized uranium is proposed. This will permit significant recycle (greater than 90%) of the lixiviant to be recycled and reused after uranium removal. This, and the limited use of fresh water makeup, will also minimize solution bleed and subsequent treatment requirements. Ion-exchange for uranium removal from carbonate lixiviants is a proven system being used commercially for over twenty years.

The pregnant heap leach solution is pumped from the preg storage pond located near the leach pad to a pregnant solution tank located at the recovery plant. Under the 3-cycle, counter-current heap leaching scenario, approximately 18.0 gpm of pregnant solution will be produced for a 20 dtph average soil throughput rate. This is approximately one-twentieth the volume of solution which would be produced by an agitated soil washing carbonate leach process.

This solution, used as feed to the ion-exchange system, is first filtered in sand (multi-media) filters to remove any suspended solids or turbidity. These sand filters operate in a continuous, alternate filtering and backwash mode. The intermittent backwash returns to the repulp tank for the second leaching train. The clarified pregnant leach solution proceeds under pumping pressure to the ion-exchange system feed tank which provides some surge capacity in the uranium recovery system. The smaller volume of solution (than the agitated carbonate soil washing system) would result in significantly smaller equipment for clarification and organic removal.

The solutions are then pumped through down-flow, fixed-bed carbon guard columns which remove most of any dissolved or suspended organics (e.g. humic or fulvic acids, etc.) which may foul the uranium ion-exchange resins. These carbon guard columns also operate in parallel with an alternating loading and stripping cycle.

A carbon regeneration system equipped with a storage tank with vent scrubber, a circulation system and steam-assisted carbon stripping provided to regenerate the carbon columns. This guard column system specific design requirements have not been defined at this point. It is likely that additional unit operations and bleeds of solid and/or liquid waste streams to disposal or to solution treatment would be required. The carbon itself may have to be replaced periodically with fresh carbon and regenerated off-line with a more severe treatment (such as solvent extraction or kiln regeneration) and subsequently recycled.

If an alternate resin is used in the guard column instead of carbon, the regenerate system requirements would also likely differ from those of the carbon columns.

The pressurized solutions from the guard columns continue as feed to three fixed-bed ion exchange columns in the Loading Ion Exchange system. Different from the sand filters and guard columns, the uranium IX recovery system sizing is now driven primarily by the quantity of uranium in the feed pregnant solutions, not the specific flow rate limitations of the fixed bed contractors. As a result of loading limitations, the resin bed size (volume of resin) and stripping frequency determine the design. Although only 1/20 the flow, the column diameter required is 5′ versus 10′ (1/4 of the cross-sectional bed area) for the agitated carbonate soil washing process which is limited by specific solution flow rate.

Sufficient feed pump pressure is provided to force the solution through all of the fixed beds in series without requiring boosting. The columns are configured as a carrousel which operates as two or three stages in series for loading. About half the time, the first stage with loaded resin is by-passed and is in a stripping cycle. The loading continues with the former second stage becoming the new first stage and the third stage becoming the new second stage.

When breakthrough occurs in the first of the three stages (i.e. the uranium concentration on discharge from the column is about 10% of the feed), it is taken out of service for stripping. In a fixed-bed ion exchange system for uranium, this occurs when the resin is loaded to about 90% of its maximum loading. When stripping is completed, the freshly-stripped column is restored to the series train as the new third stage.

STRIPPING AND STRIP SOLUTION MAKEUP:

The stripping system for the heap leach process ion exchange systems is identical to that of the agitated carbonate soil washing process. A detailed description will not be repeated here. However, due to the limitation in the ion-exchange column design by resin loading and not solution specific flow (i.e. resin bed pressure drop), the following differences should be noted.

The stripping frequency is now approximately once per day instead of one or two times per week per train. In addition, two columns will now be stripped per day, one for each train. Therefore, the volume of strip solution required (in a day or week) remains <u>approximately the same</u> as required for stripping the larger columns less frequently for the carbonate soil washing uranium recovery/removal system. As a consequence, the strip system makeup requirements and <u>all</u> systems within the peroxide precipitation system remain the same.

PEROXIDE URANIUM PRECIPITATION:

Since the quantity of uranium removed from the soil is approximately the same for each of the processes, the recovery/removal circuit (peroxide precipitation) which handles this uranium are the same for heap leaching as for agitated carbonate soil washing.

TURBID WATER HANDLING:

Turbid water produced during the washing and rinsing of the oversize soil and trash components in the rotary drum scrubber and reclaimed from the soil pads by the sump pumps will contain some dissolved uranium. Likewise, the solid soil particles in the turbid water could also have some contained uranium. Therefore a circuit to add flocculant to the incoming turbid water, settle the solids in a thickener/clarifier and dispose of them by producing a filter cake is provided.

Decant liquid from the clarifier, which contains some uranium, is added to the recycle lixiviant being returned to the feed ponds to the heaps. This uranium, therefore, will be integrated into the leaching solution makeup and be subsequently recovered from the pregnant solutions produced by the heap leaching process.

The turbid water system is sized to accommodate up to a maximum 50 gpm feed. Normal feed rates expected to be less than 10 gpm average rate on a continuous basis.

PROCESS DESIGN ASSUMPTIONS
REVISED BICARBONATE HEAP LEACHING
CONVENTIONAL, PERMANENT PAD, MOVABLE HEAP

The following preliminary process design assumptions were used in flowsheet development, mass balance derivation, equipment selection and preliminary sizing:

COARSE SOIL SEPARATION:

Nominal throughput rate 112 dry tons soil/hour (30 hours/week).

Soil moisture content average of 12.0 wt.%.

Recycle lixiviant and fresh water addition to the scrubber are controlled to minimize turbid water generation.

Coarse oversize soil (+150 mm rocks, roots, etc.) is assumed to be <1.0 wt.% soil.

Soil stockpile storage capacity: 2,400 tons, 2,075 yds^3.

Under-the-pile-reclaim: Nominal 112 dtph, maximum 168 dtph.

Sand ratio to soil: Nominal 20 wt.% (range: 10-50 wt.%)

Agglomerating agent: Unspecified

CONVENTIONAL PERMANENT PAD HEAP LEACHING CIRCUIT:

Feed soil/sand to cell loading: Nominal 151 tph (120 yds^3/hr)

Reclaim rate for leached soil/sand mixture: Nominal 167 tph (144 yds^3/hr)

Leaching cycles: 3, counter-current with "stacked" lixiviant solutions.

Leach time per cycle: 7 days (range: 6-7).

Total leach time: 21 days (range: 18-21)

Height of heap: 10'

Leaching solution application rate: 0.0025 gpm/ft^2

Rinse water application rate: 0.0025-0.005 gpm/ft^2

Operation/day, i.e. solution application to heap: 100%

Storage pond sizing: 150′ x 50′ x 10′ with sump and pump box, Working Volume 300,000 gallons each, (9 days solution inventory), Maximum Volume 600,000 each, based on accepted good heap leaching practice.

URANIUM RECOVERY SYSTEM (IX LOADING SYSTEM):

Pregnant solution storage tank (solution to IX) based on 2 hours residence time (@36 gpm).

Recycle lixiviant tank based on 2 hours residence time (@36 gpm).

Sand filter backwash tank based on minimum of 2 wetted volume backwash cycles.

Sand filter specific flow rate of 5.0 gpm/ft^2. Bed height 6 ft.

Guard column specific flow rate of 2.5 gpm/ft^2. Bed height 6 ft.

Ion exchange column specific flow rate of 2.0 gpm/ft^2. Bed height 4 ft.

Ion-exchange maximum loading of 100 lbs uranium/ton resin.

Resin replacement rate nominally 3% of inventory/year.

Materials of construction: 316 S.S. or rubber lined C.S.

STRIPPING AND STRIP SOLUTION MAKEUP:

Carbon replacement rate nominally 10% of inventory/year.

Bed volumes of strip solution, 5 design (7 max.).

Bed volumes of fresh water rinse, 1 design (2 max.).

Resin strip solution nominally 1.0 molar NaCl, 0.10 molar HCl, pH = 2.5-3.0.

Strip solution specific flow rate, nominal 0.10 gpm/ft^2, maximum 0.20 gpm/ft^2.

Working volume 8,100 gallons each of two.

Materials of construction: 316 S.S. or rubber lined C.S.

PEROXIDE URANIUM PRECIPITATION:

Design feed rate 4.78 gpm (half-time operation).

Residence time/stage = 90.0 minutes.

Limit for pH (minimum) = 2.0.

Materials of construction: HDPE, fiberglass or rubber lined C.S.

Hydrogen peroxide feed to: tank 1 and tank 2.

NaOH feed to: tank 4 and tank 5.

Thickener U/F slurry recycle: nominally 100% new feed uranium, range: 0-400%.

Thickener unit area: based on 1.0 m/hr fall velocity and maximum 30 gpm feed rate (0.40 gpm/ft^2 specific flow rate).

Thickener U/F density: nominal 40 wt.% solid, (range: 25-50%).

Recessed plate & frame uranium peroxide filter unit area based on 27.3 ft^3 net cake capacity, one filtration cycle/day.

Precoat filter unit area based on 1.0 gpm/ft^2 specific flow rate, 50 gpm maximum feed rate. One backflush cycle/day as precoat to recessed plat & frame filter.

Bleed rate: lixiviant ≈8.5% C.L. (range: 5-10%), strip solution ≈10.0 % C.L.

SHEET 1 **BICARBONATE HEAP LEACH OF URANIUM-CONTAMINATED SOILS - PERMANENT PAD LEACH** 10-Jan-95

112.0 DTPH SOIL INPUT (30 HRS/WK), COARSE SEPARATION AND LEACHING ISSUE 2

Rev. 0

STREAM NUMBER	STREAM DESCRIPTION (BASED ON FLEXMET BALANCES)	SOLIDS S.G.	LIQUID S.G.	SLURRY S.G.	BULK DENSITY	WT.% SOLIDS	SOLIDS (STPH)	LIQUID (STPH)	TOTAL (STPH)	TOT. GPM (yd3/HR)	STREAM NUMBER
	(BASED ON 30 MATERIALS HANDLING HOURS/WEEK)										
2	SOIL FEED TO GRIZZLY	2.500	1.000	2.119	90.0	88.00	112.0000	15.2727	127.2727	(104.72)	2
3	GRIZZLY OVERSIZE (+6") TO SCRUBBER	2.500	1.000	2.119	90.0	88.00	2.8000	0.3818	3.1818	(2.62)	3
4	GRIZZLY UNDERSIZE (-6") TO COARSE SOIL CONVEYOR	2.500	1.000	2.119	90.0	88.00	109.2000	14.8909	124.0909	(102.10)	4
5	TROMMEL OVERSIZE (+4") FROM SCRUBBER	2.500	1.000	1.923	84.0	80.00	0.8000	0.2000	1.0000	(0.88)	5
6	SCRUBBER UNDERSIZE FEED TO DEWATERING SCREEN	2.500	1.000	1.111	48.5	16.64	2.0000	10.0160	12.0160	43.21	6
51	RECYCLE BICARBONATE LIXIVIANT WASH	—	1.015	1.015	63.3	0.00	0.0000	5.0080	5.0080	19.71	51
101	SCRUBBER FRESH WASH WATER	—	1.000	1.000	62.4	0.00	0.0000	5.0080	5.0080	20.01	101
7	DEWATERED SCREEN O/S TO STOCKPILE FEED BELT	2.500	1.000	2.119	92.5	88.00	1.9010	0.2592	2.1602	4.07	7
8	SCREEN U/S (TURBID WATER) TO CLARIFIER	2.500	1.007	1.014	63.2	1.00	0.0990	9.7568	9.8558	38.85	8
9	COMBINED SOIL TO LEACH FEED STOCKPILE	2.500	1.000	2.119	90.0	88.00	111.1010	15.1501	126.2511	(103.88)	9
10	COMBINED SOIL TO AGGLOMERATOR	2.500	1.000	2.119	90.0	88.00	111.1010	15.1501	126.2511	(103.88)	10
- 11	SAND/GRAVEL BULKING AGENT FEED TO AGGLOMERATOR	2.650	1.000	2.275	106.4	90.00	22.2202	2.4689	24.6891	(17.19)	11
- 12	AGGLOMERATING AGENT (UNSPECIFIED)	2.500	1.000	1.000	0.0	0.00	0.0000	0.0000	0.0000	0.00	12
13	SOIL/SAND BLEND TO HEAP CONSTRUCTION	2.575	1.000	2.175	92.7	88.33	133.3212	17.6190	150.9402	(120.60)	13
14	LEACHED SOIL/SAND BLEND RECLAIMED FROM PADS	2.575	1.007	1.964	85.8	80.00	133.3212	33.3303	166.6515	(143.91)	14

SHEET 2 **BICARBONATE HEAP LEACH OF URANIUM-CONTAMINATED SOILS - PERMANENT PAD LEACH** 10-Jan-95

EQUIVALENT TO 20.0 DTPH SOIL INPUT - CONVENTIONAL PAD LEACHING ISSUE 2

Rev. 0

STREAM NUMBER	STREAM DESCRIPTION (BASED ON FLEXMET BALANCES)	SOLIDS S.G.	LIQUID S.G.	SLURRY S.G.	BULK DENSITY	WT.% SOLIDS	SOLIDS (STPH)	LIQUID (STPH)	TOTAL (STPH)	TOT. GPM (yd3/HR)	STREAM NUMBER
	(BASED ON 24-HOUR DAY, 7 DAYS/WEEK)*										
* 15A	SOIL/SAND UNDER ACTIVE LEACH PAD A	2.575	1.015	1.964	85.781	80.000	23.8074	5.9518	29.7592	(25.70)	15A
15B	SOIL/SAND UNDER ACTIVE LEACH PAD B	2.575	1.015	1.964	85.781	80.000	23.8074	5.9518	29.7592	(25.70)	15B
15C	SOIL/SAND UNDER ACTIVE LEACH PAD C	2.575	1.015	1.964	85.781	80.000	23.8074	5.9518	29.7592	(25.70)	15C
15D	SOIL/SAND RINSING, DRAINING & RECLAIM PAD D	2.575	1.015	1.964	85.781	80.000	23.8074	5.9518	29.7592	(25.70)	15D
15E	SOIL/SAND BEING LOADED (EQUIVALENT) ON PAD E	2.575	1.015	1.964	85.781	80.000	23.8074	5.9518	29.7592	(25.70)	15E
8	SCREEN U/S (TURBID WATER) TO CLARIFIER	2.500	1.015	1.021	63.7	1.00	0.0177	1.7423	1.7600	6.89	8
50	RETURN LIXIVIANT FROM RECOVERY	—	1.015	1.015	63.3	0.00	0.0000	5.6440	5.6440	22.21	50
51	RECYCLE LIXIVIANT TO SCRUBBER AS WASH	—	1.015	1.015	63.3	0.00	0.0000	0.8943	0.8943	3.52	51
52	RECYCLE LIXIVIANT TO SOLUTION MAKEUP	—	1.015	1.015	63.3	0.00	0.0000	4.7497	4.7497	18.69	52
52L	RECYCLE LIXIVIANT EVAPORATED IN POND	—	1.015	1.015	63.3	0.00	0.0000	0.2375	0.2375	0.93	52L
53A	FRESH LIXIVIANT TO HEAP (PAD C)	—	1.015	1.015	63.3	0.00	0.0000	5.3735	5.3735	21.15	53A
53E	EVAPORATION FROM 53A (PAD C)	—	1.015	1.015	63.3	0.00	0.0000	0.5374	0.5374	2.11	53E
54P	WEAK PREGNANT SOLUTION PRODUCED (PAD C)	—	1.015	1.015	63.3	0.00	0.0000	4.8362	4.8362	19.03	54P
54L	WEAK PREGNANT SOLUTION EVAPORATED IN POND	—	1.015	1.015	63.3	0.00	0.0000	0.2687	0.2687	1.06	54L
54A	WEAK PREGNANT SOLUTION AS LIXIVIANT TO PAD B	—	1.015	1.015	63.3	0.00	0.0000	5.3735	5.3735	21.15	54A
55E	EVAPORATION FROM 54A (PAD B)	—	1.015	1.015	63.3	0.00	0.0000	0.5374	0.5374	2.11	55E
55P	INTERMEDIATE PREGNANT SOLUTION PRODUCED (PAD B)	—	1.015	1.015	63.3	0.00	0.0000	4.8362	4.8362	19.03	55P
55L	INT. PREGNANT SOLUTION EVAPORATED IN POND	—	1.015	1.015	63.3	0.00	0.0000	0.2687	0.2687	1.06	55L
55A	INT. PREGNANT SOLUTION AS LIXIVIANT TO PAD A	—	1.015	1.015	63.3	0.00	0.0000	5.3735	5.3735	21.15	55A
56E	EVAPORATION FROM 55A (PAD A)	—	1.015	1.015	63.3	0.00	0.0000	0.5374	0.5374	2.11	56E
56P	FINAL PREGNANT SOLUTION PRODUCED (PAD A)	—	1.015	1.015	63.3	0.00	0.0000	4.8362	4.8362	19.03	56P
56L	FINAL PREGNANT SOLUTION EVAPORATED IN POND	—	1.015	1.015	63.3	0.00	0.0000	0.2687	0.2687	1.06	56L
56	FINAL PREGNANT SOLUTION AS FEED TO U RECOVERY	—	1.015	1.015	63.3	0.00	0.0000	4.5675	4.5675	17.98	56
* *104*	*FRESH WATER RINSE, SOLUTION MAKEUP (TOTAL)*	—	1.000	1.000	62.4	0.00	0.0000	1.4947	1.4947	5.97	104
* 62	CARBON DIOXIDE (LIQUID EQUIVALENT) MAKEUP	—	1.000	1.000	62.4	0.00	0.0000	0.0845	0.0845	0.34	62

*NOTE: BASED ON INITIAL MASS BALANCES FOR REVISION 2 FLOWSHEETS (11/21/94).

5-PAD OPERATION: ONE LOADING, ONE DRAINING AND UNLOADING, 3 UNDER ACTIVE COUNTER-CURRENT LEACH.

EACH PAD CONTAINS 3,360 DRY TONS SOIL, 672 DRY TONS SAND = 4,032 DRY TONS SOIL/SAND BLEND

THREE 7-DAY CYCLES: ONE (A) LEACHING FRESH SOIL WITH INTERMEDIATE PREGNANT SOLUTION, PRODUCING FINAL PREGNANT SOLUTION.

SECOND (B) LEACHING PREVIOUS LEACH CYCLE SOIL WITH WEEK PREG. PRODUCING INTERMEDIATE PREG.

THE LAST (C) LEACH CYCLE LEACHES SOIL FROM PREVIOUS TWO CYCLES WITH FRESH LEACHATE (RECYCLE LIXIVIANT).

EVAPORATION RATES: 10% OF CIRCULATING SOLUTION APPLIED TO HEAPS, 5% CIRCULATION SOLUTION EACH POND.

SOLUTION APPLICATION RATES 0.0025 GPM/FT2 (APPLIED TO HEAPS), ACTIVE LEACH AREA = 8,460 FT2/PAD, THEREFORE 21.15 GPM APPLIED/PAD.

SHEET 3

BICARBONATE HEAP LEACH OF URANIUM-CONTAMINATED SOILS - PERMANENT PAD LEACH
20.0 DTPH SOIL INPUT (24 HOUR/DAY, 7 DAYS/WK OPERATION), URANIUM REMOVAL

10-Jan-95
ISSUE 2
Rev. 1

STREAM NUMBER	STREAM DESCRIPTION (BASED ON FLEXMET BALANCES)	SOLIDS S.G.	LIQUID S.G.	SLURRY S.G.	BULK DENSITY	WT.% SOLIDS	SOLIDS (STPH)	LIQUID (STPH)	TOTAL (STPH)	TOT. GPM (yd3/HR)	STREAM NUMBER
50	RETURN LIXIVIANT FROM RECOVERY	—	1.015	1.015	63.3	0.00	0.0000	5.6440	5.6440	22.21	50
8	SCREEN U/S (TURBID WATER) TO CLARIFIER	2.500	1.015	1.021	63.7	1.00	0.0177	1.7423	1.7600	6.89	8
56	FINAL PREGNANT SOLUTION AS FEED TO U RECOVERY	—	1.015	1.015	63.3	0.00	0.0000	4.5675	4.5675	17.98	56
103	FRESH WATER FOR FLOC. DILUTION (TURBID WATER)	—	1.000	1.000	73.1	0.00	0.0000	0.0035	0.0035	0.0140	103
	DRY FLOCCULANT ADDITION (2 LBS/TON SOLIDS)	1.800	1.015	1.015	63.3	0.00	0.000035	0.0000	0.000035	(0.000041)	
41	DILUTED FLOCCULANT (TURBID WATER)	1.800	1.000	1.004	62.6	1.00	0.000035	0.0035	0.003536	0.0141	41
42	TURBID WATER CLARIFIER FEED (WITH FLOC)	2.500	1.000	1.006	62.7	1.00	0.017714	1.7458	1.763495	7.0027	42
44	TURBID CLARIFIER DECANT TO RECYCLE	2.500	1.007	1.007	64.0	0.00	0.0000	1.7192	1.7192	6.8203	44
45	TURBID CLARIFIER U/F TO FILTER PRESS	2.500	1.007	1.323	82.5	40.00	0.0177	0.0266	0.0443	0.1337	45
46	FILTRATE FROM TURBID PRESS TO RECYCLE	2.500	1.007	1.007	62.8	0.00	0.0000	0.0148	0.0148	0.0586	46
47	FILTER CAKE FROM TURBID PRESS TO DISPOSAL	2.500	1.007	1.007	62.8	60.00	0.0177	0.0118	0.0295	0.1171	47
22	FEED TO SAND FILTERS/IX COLUMNS	—	1.015	1.015	63.3	0.00	0.0000	4.5675	4.5675	17.98	22
23	IX COLUMN DISCHARGE	—	1.015	1.015	63.3	0.00	0.0000	4.5675	4.5675	17.98	23
24	STRIP SOLUTION TO IX COLUMNS	—	1.106	1.106	69.0	0.00	0.0000	1.3552	1.3552	4.8950	24
25	PREGNANT STRIP SOLUTION FROM IX COLUMNS	—	1.106	1.106	0.0	0.00	0.0000	1.3552	1.3552	4.8950	25
26	PRECIPITATION CIRCUIT FEED (CONTINUOUS)	—	1.106	1.106	69.0	0.00	0.0000	1.3552	1.3552	4.8950	26
27	HYDROGEN PEROXIDE FEED (50% SOLN.) TO PPTN.	—	1.197	1.197	74.7	0.00	0.0000	0.0029	0.0029	0.0096	27
28	NaOH FEED (30% SOLN.) TO PPTN.	—	1.327	1.327	82.8	0.00	0.0000	0.0150	0.0150	0.0452	28
32	RECYCLE THICKENER U/F AS SEED IN PPTN.	7.600	1.106	1.931	120.4	50.00	0.0119	0.0119	0.0238	0.0492	32
29	URANYL PEROXIDE SLURRY TO THICKENER (NET)	7.600	1.106	1.123	70.0	1.72	0.0238	1.3552	1.3789	4.9075	29
30	THICKENER DECANT O/F TO PRECOAT FILTER	7.600	1.106	1.106	69.0	0.00	0.0000	1.3314	1.3314	4.8092	30
31	THICKENER U/F SLURRY	7.600	1.106	1.931	120.4	50.00	0.0238	0.0238	0.0475	0.0983	31
33	FILTER PRESS FEED	7.600	1.106	1.931	120.4	50.00	0.0119	0.0119	0.0238	0.0492	33
34	FILTER PRESS FITRATE TO PRECOAT FILTER	7.600	1.106	1.106	69.0	0.00	0.0000	0.0079	0.0079	0.0286	34
35	FILTER PRESS URANYL PEROXIDE CAKE TO DISPOSAL	7.600	1.106	3.080	192.1	75.00	0.0119	0.0040	0.0158	(0.0061)	35
	NET FEED TO PRECOAT FILTER	7.600	1.106	1.106	69.0	0.00	0.0000	1.3393	1.3393	4.8378	
36	NET FEED TO RECYCLE MAKEUP	7.600	1.106	1.106	69.0	0.00	0.0000	1.2054	1.2054	4.3540	36
60	BLEED REGENERATE TO AWWT	7.600	1.106	1.106	69.0	0.00	0.0000	0.1339	0.1339	0.4838	60
37	SODIUM CHLORIDE TO STRIP SOLUTION MAKEUP	2.200	1.106	2.200	96.0	100.00	0.0134	0.0000	0.0134	0.0243	37
38	HYDROCHLORIC ACID TO STRIP SOLUTION MAKEUP	2.200	1.150	1.150	71.7	0.00	0.0000	0.0134	0.0134	0.0465	38
104	FRESH WATER IN STRIP MAKEUP	—	1.000	1.000	62.4	0.00	0.0000	0.1071	0.1071	0.4280	104
57	LIXIVIANT BLEED TO AWWT	2.500	1.106	1.106	69.0	0.00	0.0000	0.6575	0.6575	2.3750	57
60	STRIP SOLUTION BLEED TO AWWT	—	1.106	1.106	69.0	0.00	0.0000	0.1205	0.1205	0.4354	60

*NOTE: BASED ON INITIAL MASS BALANCES FOR REVISION 2 FLOWSHEET (11/16/94) FOR CONVENTIONAL, PERMANENT PAD BICARBONATE HEAP LEACH
PRECIPITATION CIRCUIT DESIGNED FOR HALF-TIME OPERATION AT TWICE CONTINUOUS FLOW RATES.

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FERNALD ID TECHNOLOGIES

HIGH GRADIENT MAGNETIC SEPARATION PROCESS

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PROCESS DESCRIPTION
HIGH-GRADIENT MAGNETIC SEPARATION

INTRODUCTION:

The process design approach taken for the HGMS uranium removal/concentration process continues to assume a relatively optimistic process performance as a basis for equipment selection and process design. In addition, the unit operations selected should be readily operable at the assumed operating conditions. If adequate performance for the selected process concepts and equipment selected are supported by laboratory tests, the capital and operating costs should be able to be directly compared to other ID technologies.

The following presents a brief process description of the attached flowsheet and some key process assumptions used as the basis for initial material balances used for equipment sizing.

COARSE SOIL SEPARATION:

The coarse soil separation circuit for HGMS is similar to the coarse soil separation process used for the initial ID technologies flowsheets.

The philosophy of the coarse soil separation circuit is to remove oversize (+2mm) soil fractions low in uranium contamination and to produce two particle size fraction splits of feed slurry to the HGMS unit. Coarse soil not processed through the HGMS system is washed with recycle water to remove small adhered contaminated particles, and then dewatered, prior to return to the site. This separation equipment should be capable of this goal without the need for additional mechanical dewatering prior to HGMS processing, and also should minimize slurry pumping. Throughout the circuit, the use of recycle filtrate water is maximized to reduce fresh water consumption and minimize wastewater discharge to the AWWT facility.

The initial separation of the coarse size fractions of the soil (+2 mm) is accomplished using equipment which minimizes the amount of fresh water added to the system. The slurry feed to HGMS has a density of approximately 10% percent solids and a -74 micron (μ) size consistent with the research-determined requirements for efficient HGMS unit operation is the goal of subsequent fine fraction preparation unit processes.

The soil feed to the process is initiated by reclaim from a soil storage facility or by direct feed from the excavation operations. The Grizzly Feed Conveyor delivers the soil to a vibrating Wet Grizzly which scalps off oversize and trash materials (+4 in) in the soil. Spray recycle filtrate water is used, as necessary, to control any dust emissions and to reduce clogging of the undersize chute to Rotary Drum Scrubber No. 2.

The Wet Grizzly is elevated such that the oversize and undersize fractions flow by gravity to the Rotary Drum Scrubbers.

The oversize (+4 in) material from the Grizzly reports to Rotary Drum Scrubber No. 1, in which it is slurried with system recycle water (filtrate) from the Recycle Water Tank to wash off any adhering fine soil particles. Rotary Drum Scrubber No. 1 is equipped with a solid drum section and dewatering drain for removal of slimes as well as a trommel screen extension where rinsing with fresh water removes most dissolved uranium.

The oversize (+1/2 in) material from Rotary Drum Scrubber No. 1 drops on the Coarse Soil Conveyor which transports it to a stockpile for return to the site.

The undersize (-1/2 in) material from Rotary Drum Scrubber No. 1 is combined with the undersize (-10 cm) material from the Grizzly and drops by gravity into Rotary Drum Scrubber No. 2, which is identical in operation to Rotary Drum Scrubber No. 1. The washing with recycle filtrate water and rinsing on the trommel screen is repeated.

The oversize (+1/2 in) material from Rotary Drum Scrubber No. 2 drops on the Coarse Soil Conveyor which transports it to a stockpile for return to the site.

The undersize (-1/2 in) slurry from Rotary Drum Scrubber No. 2 flows by gravity to one of two Live-Bottomed Sumps and Pumps from where it is pumped to the elevated Washing and Dewatering Screen by the live bottom sump slurry pump. The Washing and Dewatering Screen separates and washes with fresh water the remaining coarse (+2 mm) soil.

The oversize (+2 mm) material from the Washing and Dewatering Screen discharges onto the Coarse Soil Conveyor which transports it to a washed coarse soil stockpile for return to the site.

HGMS FINE SOIL SIZE FRACTION PREPARATION

Since the HGMS system requires relatively fine solid particles to work effectively, multiple size fraction operations are performed on the soil slurry prior to introduction to the HGMS unit.

The undersize (-2 mm) material from the Washing and Dewatering Screen flows by gravity to a 100-mesh (150 μ) vibrating Sizing Screen No. 1. The oversize (+150 μ) material from this screen is size-reduced to less than 150 microns by the Roll Crusher and then re-combined with the screen's undersize (-150 μ) material and discharged into the Agitated Holding Tank.

In the Agitated Holding Tank, an hexametaphosphate surfactant solution is mixed with the soil slurry to promote metallic particle dispersion. The surfactant feed system consists of a Dry Surfactant Feeder/Hopper, a Surfactant Solution Holding Tank and Mixer, and two Surfactant Feed Pumps. For design purposes, surfactant solution is fed to the Agitated Holding Tank at the rate of 10 ml per liter of soil slurry. The surfactant holding tank can store more than a 24 hour supply of solution while still allowing for increased dosage rates.

The Agitated Holding Tank Mixer is of the low-intensity airfoil-type, with downward pumping impellers. This design keeps the slurry in suspension in an axial-flow pattern and minimizes agitator power requirements. This type of mixer also does not appreciably degrade soil particle size.

From the Agitated Holding Tank, the soil slurry is pumped by the Sizing Screen No. 2 Feed Pumps to a 200-mesh (74 μ) vibrating Sizing Screen No. 2. The oversize (+74 μ) material from this screen is discharged to the Attrition Scrubber while the undersize (-74 μ) material flows by gravity to the Hydrosizer Feed Holding Tank.

In the Attrition Scrubber, the oversize material from Sizing Screen No. 2 is vigorously mixed with high-velocity recycle water to break-up agglomerated soil particles and to scrub any adhered uranium staining from the coarser soil particle surfaces. Larger (-150 μ to + 74 μ) particles may become size reduced, however the primary function of the Attrition Scrubber is to break up agglomerates of smaller particles. The attrition scrubber employs variable pitch, axial flow propellers to produce high shear flows and intense particle to particle interactions (or collisions) in high-percent solids slurries. Multiple compartments operate in parallel within one unit.

The discharge from the Attrition Scrubber is then dropped by gravity on to a second 200-mesh (74 μ) vibrating Sizing Screen No. 3. The oversize (+74 μ) material from this screen is discharged by gravity for processing in the HGMS tails processing system which is further described in a subsequent section. The undersize (-74 μ) material from Sizing Screen No. 3 is discharged by gravity to the Hydrosizer Feed Holding Tank.

In the Hydrosizer Feed Holding Tank, a solution of sodium dithionate (a chemical reducing agent) is metered into the soil slurry to chemically reduce the uranium valence (U^{+6} to U^{+4}). The dithionate feed system consists of a Dry Dithionate Feeder/Hopper, a Dithionate Solution Holding Tank, and two Dithionate Feed Pumps.

The sodium dithionate solution is fed to the Hydrosizer Feed Holding Tank at an approximate rate of 0.2 pounds per 100 pounds of soil.

The Hydrosizer Feed Holding Tank Mixer is similar in design to the Agitated Holding Tank described above.

Based upon prior characterization of the Fernald Site, soil particles larger than 74 μ were found to not contain uranium contamination (particularly if surface staining is scrubbed off) and, although the HGMS unit can physically handle these particles, they will not be treated. No additional soil characterization was performed as part of this ID program study.

LANL research has indicated that HGMS capture efficiency of the paramagnetic particles increases as the soil slurry particles are more homogeneous in size. Therefore, a cost-optimal approach of partitioning the -74 μ slurry into two size fractions has been developed. Characterization of the Fernald Site has indicated that most of the uranium contamination is contained on soil particles about 20 μ in size. Accordingly, size fractions of -74 μ/+20 μ and -20 μ were selected for treatment in the HGMS unit. Separation of these two size fractions is accomplished by the Hydrosizer.

The Hydrosizer is a hydrocyclone, or hydraulic classifier. It is typically a static separator based on centrifugal separation in a fluid vortex generated within the cylindrical cone-bottom body. The feed flow is divided into a coarser underflow fraction and the finer overflow fraction. Particle separation is due to the vortex flow, with very little reliance on gravity.

The -74 μ/+20 μ fraction slurry stream is collected by gravity in an agitated Coarse Fraction Feed Tank and fed to the HGMS unit by the Coarse Fraction Feed Pump. The -20 μ slurry stream fraction is collected by gravity in an agitated Fine Fraction Feed Tank and fed to the HGMS unit by the Fine Fraction Feed Pump.

HIGH GRADIENT MAGNETIC SEPARATION:

The HGMS unit basically consists of a porous magnetic matrix (stainless steel wool or other material) surrounded by a superconducting electromagnetic coil capable of creating an intense (about 60 kilogauss) magnetic field and cooled by a cryogenic system. Under such an intense magnetic field, paramagnetic compounds of relatively moderate magnetic susceptibility, such as uranium and uranium oxides, can be successfully separated from contaminated soils. Due to its superconducting properties, the HGMS unit consumes virtually no power; nearly 0 KW at 900 Amps output. AC input is approximately 0 Amps at 480 Volts. The HGMS system as a whole consumes approximately 19 KW (26 hp), primarily to operate the helium cryogenics compressor.

As provided by Eriez Magnetics (Erie, PA), the HGMS system includes the superconducting magnetic coil, associated cryogenics (liquifier, compressor, piping) power supply, warm gas storage tank, and controls. The HGMS unit is as skid-mounted as is feasible for a 250 ton system.

The HGMS slurry processing scheme consists of two passes through the magnetic matrix for each of the two size fractions, a backflush following the two passess for each size fraction, and an optional preliminary forward scalping pass and backflush for each size fraction for removal of materials with high magnetic susceptibility. For the optional pass, the electromagnetic coil is energized at low level, or deactivated.

Assuming incorporation of the optional scalping pass, operation of a complete HGMS cycle proceeds as follows:

1. Scalping pass for the coarse (-74 μ/+ 20 μ) fraction
2. HGMS unit backflush to tails processing
3. Scalping pass for the fine (-20 μ) fraction
4. HGMS unit backflush to tails processing
5. First pass for coarse fraction
6. Second pass for coarse fraction
7. HGMS unit backflush to concentrated contaminant processing
8. First pass for fine fraction
9. Second pass for fine fraction
10. HGMS unit backflush to concentrated contaminant processing

The scalping pass, for either the coarse or fine fractions, consists of pumping the slurry from the appropriate feed tank through the HGMS system magnetic matrix at a velocity of approximately 1 cm/sec with the surrounding magnetic coil at no-power or reduced power. Optimally, the electromagnetic coil power level is set such that all materials with greater magnetic susceptibility than uranium would be removed or scalped from the slurry (magnetic susceptibilities for many materials are readily available). The HGMS system's effluent is then collected in the HGMS Reycle Feed Tank. Processing continues until low level is reached in the Coarse Fraction Feed Tank or Fine Fraction Feed Tank. Once the scalping pass is complete, the contents of the HGMS Reycle Feed Tank are returned from the HGMS Recycle Feed Tank to either the Coarse Fraction Feed Tank or the Fine Fraction Feed Tank by the HGMS Recycle Feed Pump.

The HGMS unit backflush consists of pumping recycle water from the Decant Water Storage Tank counter-current through the magnetic matrix with the Backflush Pump to dislodge the soil particles magnetically separated by that matrix and convey them to the HGMS Tails Thickener. Backflush water is constantly recycled to the Decant Water Storage Tank. A pinch valve restricts recycle flow as necessary to divert water from the decant tank to meet HGMS backflush demand.

Note, that to avoid fouling of the magnetic matrix by magnetic materials, the scalping pass for either the coarse or fine fraction could alternatively be processed through a separate media-filled canister which would be placed in the super-conducting electromagnetic coil. Removable canisters are easily interchanged into the permanently mounted coil. Following the scalping pass, the scalping canister would be removed from the magnet, to be replaced with the primary magnetic matrix. The magnetic field can be ramped from full power to no power, or no power to full power, within one minute to facilitate canister exchange.

Following the scalping passes, reprocessing of both the coarse and fine fractions with two passes through the HGMS system is considered necessary to concentrate uranium contamination.

The first pass, for either the coarse or fine fractions, is performed in the very same manner as the scalping pass, except with the electromagnetic coil surrounding the magnetic matrix at full power - 20,000 Gauss or 2 Tesla.

The second pass, for either the coarse or fine fractions, consists of adjusting the circuit valves so that the feed of the HGMS system is now provided from the slurry accumulated in Cycle Feed Tank during the first pass. Following the second pass, the repositioned circuit valves prevent the slurry from returning to the HGMS Reycle Feed Tank, and instead direct it to the HGMS Tails Thickener.

Following the first and second pass for either the coarse or fine fractions, the HGMS system is backflushed in the same manner as after each of the scalping passes, except that the soil particles dislodged from the HGMS system magnetic matrix are conveyed to the HGMS Concentrate Thickener. Forward processing time may be up to an hour without requiring backflush. Backflush is a computer controlled function consisting of higher velocity alternating forward and countercurrent scouring, via computer controlled actuated valves, culminating in a countercurrent discharge of concentrate materials to the HGMS Concentrate Thickener. The backflush cycle can be completed in approximately five minutes.

For this study, the capacity of the Coarse Fraction Feed Tank, Fine Fraction Feed Tank, and HGMS Reycle Feed Tank has been designed to provide a minimum of 30 minutes residence time. In actuality, tank sizing should be determined by factoring in the HGMS system's magnetic capture capacity and adding appropriate safety factors based on the volume of slurry and concentration of uranium in the slurry.

HGMS CONCENTRATE PROCESSING:

The primary purpose of HGMS concentrate processing is to dewater this concentrate to minimize its volume to facilitate handling and disposal. The secondary purpose of HGMS concentrate processing is to provide a ready source of recycle water for the backflush of the HGMS system, thus minimizing the use of fresh water and the need to discharge contaminated wastewater to the AWWT. The HGMS concentrate processing system consists of the HGMS Concentrate Thickener, the HGMS Concentrate Thickener Underflow Filter Feed Pump, the Filter Press, a filter cake hopper with screw type discharge auger, and the agitated Decant Water Storage Tank.

The backflush of the HGMS system following the first and second passes for both the coarse and fine fractions is discharged under residual pressure into the HGMS Concentrate Thickener. Note that the scalping pass concentrate is discharged to the HGMS Tails Thickener, since it is not anticipated to contain a substantial uranium concentration.

In the HGMS Concentrate Thickener, solid particles settle by gravity to the bottom and supernatant water is discharged by gravity to the Decant Water Storage Tank to be used for future HGMS unit backflush cycles. The solids accumulated at the bottom of the HGMS Concentrate Thickener are collected by a mechanical rotating bottom rake and pumped to the Filter Press for further dewatering. Thickener underflow is pumped by the HGMS Concentrate Thickener Underflow Filter Feed Pump.

The Filter Press is of the recessed plate and frame type (Durco or equivalent) which removes water from the HGMS Concentrate Thickener underflow by pressure feeding it into sandwiches of fine-weave polypropylene mesh plates which retain solids but allow filtrate water to escape. The dewatered HGMS concentrate filter cake drops by gravity into a hopper and is discharged by a screw auger into a dumpster type container which is used to transport it to disposal. Filter press filtrate water is discharged under residual pressure into the Decant Water Storage Tank and recycled by the Backflush Pump for backflushing of the HGMS system as previously described, and for other uses as depicted.

As required, fresh water make-up is introduced in the Decant Water Storage Tank. This is the sole location of fresh water make-up in the system.

Also, as required, excess decant/filtrate water is blown-down by discharging to the AWWT.

HGMS TAILS PROCESSING:

The primary and secondary purposes of HGMS tails processing are very similar to those of HGMS concentrate processing, except that the dewatered soils tails have most of the uranium concentration (to <50 ppm) removed and are returned to the site. Also, decant/filtrate water is recycled as flush or dilution water throughout the coarse solids separation operations. The HGMS tails processing system consists of the Flocculant Feed System, Static Mixer No. 1, the HGMS Fines Thickener, the Thickener Underflow Filter Feed Pump, the Pressure Filter, a filter cake hopper with screw type discharge auger, and the agitated Recycle Water Tank.

A flocculant is fed to the HGMS tails and mixed in line by Static Mixer No. 1 ahead of the HGMS Tails Thickener. The purpose of this flocculant is to agglomerate fine suspended solids particle to improve settleability and filterability of these solids. The Flocculant Feed System consists of a dry reagent hopper/feeder, an agitated solution holding tank, and two solution feed pumps. The flocculant is mixed and fed initially as a 0.1 percent (by weight) solution.

In the HGMS Tails Thickener, flocculated solid particles settle by gravity to the bottom and supernatant water is discharged by gravity to the Recycle Water Tank to be used as flush or dilution water throughout the coarse solids separation operations. The solids accumulated at the bottom of the HGMS Tails Thickener are collected by a mechanical rotating bottom rake and recycled to the thickener feed well by the HGMS Tails Thickener Underflow Filter Feed Pump. A side-stream from that thickened HGMS tails recycle is periodically sent by the same pump to the Pressure Filter for further dewatering.

Filter Press No. 2 is of the high-pressure belt type (Larox or equivalent) and mechanically removes water out of the HGMS Tails Thickener underflow by squeezing it between an air expanded bladder and a moving belt filter cloth. The filter operates in a semi-continuous mode with a bleed from a circulating thickener underflow stream being intermittently fed to the pressure belt filter, as needed. The dewatered HGMS tails filter cake drops by gravity into a hopper and is discharged by a screw auger into a dumpster type container which is used to transport it to disposal. Filtrate water is discharged under residual pressure into the Recycle Water Tank and pumped from there by the Recycle Water Pump to the various coarse solids separation usage points as described earlier.

Flocculant dilution 2 is formed by the addition of recycle water to flocculant stored in the Flocculant Holding Tank followed by mixing through Static Mixer No.2. This dilution is in turn injected in-line with the HGMS Tails Thickener underflow as it is pumped to the Pressure Filter. Flocculant dilution 2 is mixed with the underflow by Static Mixer No. 3 immediately prior to discharge to the Pressure Filter. Flocculant additioin will enhance the Pressure Filter performance.

Excess decant/filtrate water is disposed of by discharging it to the AWWT, as required.

PROCESS DESIGN ASSUMPTIONS
HIGH-GRADIENT MAGNETIC SEPARATION

The following preliminary process design assumptions were used in flowsheet development, mass balance derivation, equipment selection and preliminary sizing:

COARSE SOIL SEPARATION:

Nominal throughput rate 20.0 dry tons of soil/hour.

Soil moisture content average of 12.0 wt.%.

Soil particles larger than 74 μ are separated out prior to HGMS treatment.

HGMS unit feed is split into two fractions: a coarse fraction (-74 μ/+20 μ) and a fine fraction (-20 μ). Prior soils characterization by others has indicated 20 μ is the mean contaminated soils dimension.

Coarse oversize soil (+100 mm rocks, roots, etc.) is assumed to be about 1.0 wt.% soil.

Medium size soil (-100 mm +13 mm) is assumed to be 1.5 wt % soil.

Intermediate oversize soil (-13 mm +2 mm) is assumed to be about 7.5 wt. % of the feed soil.

Coarse (-74 μ/+20 μ) and fine (-20 μ) HGMS feed fractions are each assumed to be about 41.0 wt. % of the feed soil. Each HGMS feed fraction is assumed to be 8.2 dry tph.

Each HGMS feed fraction is assumed to be 10% by weight solids slurry for optimal HGMS unit performance.

Total available flow of recycle water used for coarse, medium and intermediate oversize soil separation, including soil washing, slurry elutriation and dilution, and chemical reagent solution preparation, is assumed to be about 600 gpm.

Hexametaphosphate surfactant use is assumed to be 68 pounds of dry product per hour. The surfactant is assumed to be fed to the Agitated Holding Tank as a 20.0 percent by weight solution.

Sodium dithionate reducing agent used is assumed to be 66 pounds of dry product per hour. The sodium dithionate is assumed to be fed to the Hydrosizer Feed Holding Tank as a 10.0 percent by weight solution. (Reducing agent is necessary to convert U^{+6} to the U^{+4} form with increased magnetic susceptibility).

Residence time (minimum working volume) of 30 minutes in process tanks.

HIGH GRADIENT MAGNETIC SEPARATION:

The HGMS sequence of operations, for either the coarse or fine feed fractions, includes a scalping pass under reduced magnetic field for removal of the magnetic and highly paramagnetic compounds followed by two treatment passes under full magnetic field for the removal of uranium. The anticipated duration of either the scalping pass or each of the treatment passes is 30 minutes.

The design maximum magnetic field of the HGMS unit is about 20,000 Gauss or 2 Tesla.

The HGMS can ramp from full power to no field within one minute.

An HGMS unit backflush is performed after the scalping pass and after the two treatment passes for each of the two size fractions. The anticipated duration of each HGMS unit backflush is four minutes.

The HGMS unit scalping pass backflush is discharged to the HGMS tails processing system.

The HGMS unit uranium removal backflush is discharged to the HGMS concentrate processing system.

Scalping backflush plus tails is assumed to be 14.6 tph slurried in 476 gpm.

Uranium removal backflush (concentrate) is assumed to be 1.8 tph slurried in 109 gpm.

Residence time (minimum working volume) of 30 minutes in process tanks.

HGMS CONCENTRATE PROCESSING:

Thickener unit area: based on 1.0 m/hr fall velocity and maximum 110 gpm feed (0.50 gpm/ft^2 specific flow rate).

Thickener U/F density: nominal 40 wt.% solids, (range: 25-50%).

The average flow of decant water from the HGMS Concentrate Thickener is assumed to be 94 gpm.

Recessed plate and frame Filter Press No. 1 based on 260 ft^3 net cake capacity, one filtration cycle/day.

The average blow-down rate from the HGMS concentrate processing system to the AWWT is assumed to be 0 gpm.

HGMS TAILS PROCESSING:

Thickener unit area: based on 1.0 m/hr fall velocity and maximum 110 gpm feed (0.40 gpm/ft^2 specific flow rate).

Thickener U/F density: nominal 40 wt.% solids, (range: 25-50%).

The average flow of decant water from the HGMS Tails Thickener is assumed to be 440 gpm.

Horizontal pressure filter belt filter cake moisture of 80.0 wt.%.

Horizontal pressure belt filter design unit area of 50 $lbs/hr/ft^2$ for dewatering.

Belt filter cake solution acceptance rate is 0.170 gpm/ft^2.

The average flow of decant water from the HGMS Tails Thickener is assumed to be 438 gpm

The average blow-down rate from the HGMS tails processing system to the AWWT is assumed to be 30 gpm.

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SHEET 1/1

HIGH-GRADIENT MAGNETIC SEPARATION OF URANIUM-CONTAMINATED SOIL
20.0 DTPH SOIL INPUT, COARSE SEPARATION AND PREPARATION FOR HGMS

02-Aug-95
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Rev. 0

STREAM NUMBER	STREAM DESCRIPTION (BASED ON FLEXMET BALANCES)	SOLIDS S.G.	LIQUID S.G.	SLURRY S.G.	BULK DENSITY	WT.% SOLIDS	SOLIDS (STPH)	LIQUID (STPH)	TOTAL (STPH)	TOT. GPM (yd3/HR)	STREAM NUMBER
1	SOIL FEED TO GRIZZLY	2.500	1.000	2.119	90.0	88.00	20.000	2.727	22.727	(18.70)	1
101	WATER SPRAY TO GRIZZLY	---	1.000	1.000	62.4	0.00	0.000	5.008	5.008	20.00	101
2	GRIZZLY OVERSIZE (+4") TO SCRUBBER 1	2.500	1.000	1.763	109.9	72.11	0.500	0.193	0.693	(0.47)	2
5	GRIZZLY UNDERSIZE (-4") TO SCRUBBER 2	2.500	1.000	1.763	109.9	72.11	19.500	7.541	27.041	61.29	5
102	SCRUBBER 1 WASH WATER	---	1.000	1.000	62.4	0.00	0.000	9.922	9.922	39.64	102
4	TROMMEL OVERSIZE (+1/2") FROM SCRUBBER 1	2.500	1.000	1.923	84.0	80.00	0.200	0.050	0.250	(0.22)	4
103	SCRUBBER 2 WASH WATER	---	1.000	1.000	62.4	0.00	0.000	10.015	10.015	40.01	103
6	TROMMEL OVERSIZE (+1/2") FROM SCRUBBER 2	2.500	1.000	1.923	84.0	80.00	0.300	0.075	0.375	(0.33)	6
104	WASHING/DEWATERING SCREEN RECYCLE WATER	---	1.000	1.000	62.4	0.00	0.000	7.510	7.510	30.00	104
8	WASH SCREEN O/S (+2mm) TO STOCKPILE	2.500	1.000	1.923	84.0	80.00	1.500	0.375	1.875	3.90	8
10	WASH SCREEN U/S (-2mm) TO SIZING SCREEN NO. 1	2.500	1.015	1.273	79.4	34.17	18.000	34.683	52.683	165.29	10
9	COMBINED (+2mm) O/S TO DISPOSAL STOCKPILE	2.500	1.000	1.923	119.9	80.00	2.000	0.500	2.500	(1.54)	9
60	SURFACTANT (DRY POWDER) TO DILUTION/MIX TANK	2.500	1.150	1.150	71.7	100.00	0.034	0.000	0.034	(0.04)	60
105	DILUTION WATER FOR SURFACTANT	---	1.000	1.000	62.4	0.00	0.000	0.144	0.144	0.58	105
13	NET SURFACTANT FEED TO HOLDING TANK	2.500	0.800	0.800	49.9	0.00	0.000	0.180	0.180	0.90	13
11	OVERSIZE FROM SCREEN 1 TO ROLL CRUSHER (+150 MICRON)	2.500	1.027	1.300	81.1	35.74	2.000	3.596	5.596	17.19	11
12	SCREEN 1 U/S (-2mm) TO HOLDING TANK	2.500	1.027	1.282	80.0	33.85	16.000	31.267	47.267	147.25	12
14	NET FEED TO HOLDING TANK & HGMS SYSTEM (-150 MICRON)	2.500	1.027	1.284	80.1	34.05	18.000	34.863	52.863	164.44	14
15	270 MESH SCREEN O/S (+74 MICRON) TO ATTR. SCRUB.	2.500	1.027	1.411	88.0	46.25	2.000	2.324	4.324	12.24	15
16	270 MESH SCREEN U/S (-74 MICRON) TO HGMS FEED	2.500	1.027	1.274	79.5	32.96	16.000	32.539	48.539	152.20	16
107	ATTRITION SCRUBBER DILUTION WATER	---	1.000	1.000	62.4	0.00	0.000	13.040	13.040	52.09	107
17	DISCHARGE FROM ATTRITION SCRUBBER	2.500	1.027	1.101	68.7	11.52	2.000	15.364	17.364	62.99	17
18	270 MESH SCREEN 2 O/S (+74 MICRON) TO TAILS THICKENER	2.500	1.000	1.326	82.7	40.97	1.600	2.305	3.905	11.77	18
19	270 MESH SCREEN 2 U/S (-74 MICRON) TO HGMS FEED	2.500	1.000	1.018	63.5	2.97	0.400	13.059	13.459	52.81	19
61	SODIUM DITHIONATE CONDITIONER (DRY POWDER)	2.500	1.000	1.358	84.7	100.00	0.033	0.000	0.033	(0.016)	61
106	DILUTION WATER FOR DITHIONATE	---	1.000	1.000	62.4	0.00	0.000	0.288	0.288	1.15	106
20	NET DITHIONATE FEED TO HOLDING TANK	2.500	1.015	1.015	79.5	0.00	0.000	0.360	0.360	1.42	20
21	FEED TO HYDROSIZER (-74 MICRON)	2.500	1.212	1.391	86.8	25.00	16.400	49.202	65.602	188.35	21
108	DILUTION (ELUTRIATION) WATER FOR HYDROSIZER	---	1.000	1.000	62.4	0.00	0.000	98.405	98.405	393.13	108
23	FINE FRACTION (-20 MICRON) FROM HYDROSIZER	2.500	1.015	1.079	67.3	10.00	8.200	73.803	82.003	290.33	23
22	COARSE FRACTION (+20 MICRON) FROM HYDROSIZER	2.500	1.015	1.079	67.3	10.00	8.200	73.803	82.003	290.33	22
24	HGMS TAILS PLUS SCALPING CONCENTRATE	2.450	1.015	1.079	67.3	0.00	14.616	143.444	158.060	476.08	24
109	BACKFLUSH WATER FOR HGMS CONCENTRATE	---	1.000	1.000	62.4	0.00	0.000	22.050	22.050	88.09	109
62	DRY FLOCCULANT TO DILUTION	1.800	1.000	1.800	112.3	100.00	0.017	0.000	0.017	(0.011)	62
112	DILUTION WATER FOR FLOCCULANT	---	1.000	1.000	62.4	0.00	0.000	1.693	1.693	6.76	112
27	TAILS THICKENER DECANT	2.500	1.000	1.000	62.4	0.00	0.000	109.564	109.564	437.71	27
28	TAILS THICKENER U/F (FEED TO FILTER)	2.500	1.000	1.220	76.1	30.00	16.233	37.878	54.111	177.19	28
29	TAILS FILTER FILTRATE	2.500	1.000	1.000	62.4	0.00	0.000	33.819	33.819	135.11	29
30	DEWATERED SOIL TAILS (U-DEPLETED) TO DISPOSAL	2.500	1.000	1.923	119.9	80.00	16.233	4.059	20.292	(12.532)	30
26	FEED TO CONCENTRATE THICKENER (WITH BACKFLUSH)	3.200	1.000	1.046	67.0	6.37	1.784	26.213	27.997	106.95	26
31	CONCENTRATE THICKENER DECANT	3.200	1.000	1.000	62.4	0.00	0.000	23.537	23.537	94.03	31
32	CONC. THICKENER U/F (FEED TO CONC. FILTER)	3.200	1.000	1.379	86.0	40.00	1.784	2.676	4.460	12.92	32
33	CONC. FILTER FILTRATE	3.200	1.000	1.000	62.4	0.00	0.000	2.230	2.230	8.91	33
34	HGMS CONCENTRATE TO DISPOSAL	3.200	1.000	2.222	138.6	80.00	1.784	0.446	2.230	(1.192)	34
114	GROSS RECYCLE WATER FROM CONC. AVAILABLE	3.200	1.000	1.000	62.4	0.00	0.000	25.767	25.767	102.94	114
110	FRESH MAKEUP WATER (TAILS CIRCUIT)**	---	1.000	1.000	62.4	0.00	0.000	0.000	7.511	30.00	110
111	AVAILABLE RECYCLE WATER (TAILS CIRCUIT)**	---	1.000	1.000	62.4	0.00	0.000	0.000	152.138	607.64	111
115	NET WATER BLEED TO AWWT (TAILS CIRCUIT)**	---	1.000	1.000	62.4	0.00	0.000	0.000	7.511	30.00	115
113	MAKEUP WATER (CONCENTRATE CIRCUIT)**	---	1.000	1.000	62.4	0.00	0.000	0.000	0.000	0.00	113

*NOTE: BASED ON INITIAL MASS BALANCES FOR HGMS FLOWSHEET (JULY 1995)
**NOTE: BASED ON OVERALL WATER BALANCE FOR HGMS FLOWSHEET (JULY 1995)

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FERNALD ID TECHNOLOGIES

TIRON SOIL WASHING

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PROCESS DESCRIPTION
REVISED TIRON SOIL WASHING

INTRODUCTION:

In order to provide a basis for the FY 95 ID Program for removal of uranium contamination from soils, revised process flowsheets have been developed which reflect experience in primary uranium ore processing as well as recent laboratory results on the contaminated soils. The approach taken has been to assume a relatively optimistic process performance as a basis for equipment selection and process design. In addition, the unit operations selected should be readily operable at the assumed operating conditions. If adequate performance for the selected process concepts and equipment selected are supported by laboratory tests, the capital and operating cost benefits over previous process flowsheets should be substantial.

The following presents a brief process description of the revised Tiron Soil Washing flowsheets (Revision 2, 11/21/94) and some key process assumptions used as the basis for initial material balances used for equipment sizing.

COARSE SOIL SEPARATION:

The initial separation of the coarse size fractions of the soil (+2 mm) is accomplished using equipment which minimizes the amount of liquid (lixiviant and fresh water) added to the system. A slurry density (percent solids) consistent with the requirements for efficient leaching without dewatering is the goal.

The soil feed to the process is initiated by reclaim from a soil storage facility or by direct feed from the excavation operations. A feed conveyor delivers the soil to a wet, vibrating Grizzly which scalps off oversize and trash materials (+10 cm) in the soil. Spray water (recycle filtrate) is used, as necessary, to control any dust emissions and to reduce clogging of the undersize chute to the Rotary Drum Scrubber 2. The grizzly is elevated such that the oversize and undersize fractions flow by gravity to the drum scrubbers.

The oversize material from the grizzly reports to the Rotary Drum Scrubber 1 in which the oversize soil is mixed with recycle lixiviant as a slurry to wash off any adhered, small-size soil and to solubilize uranium staining the surface of the coarse particles using the leaching action. The drum scrubber is equipped with a solid drum section and dewatering drain for slimes removal as well as a trommel screen extension where rinsing with fresh water removes most of any dissolved uranium. The trommel oversize (+13 mm) is conveyed to a stockpile for disposal.

The undersize soil slurry (-13 mm) from the Drum Scrubber 1 is combined with the -10 cm Grizzly undersize as and flows by gravity as feed to the Rotary Drum Scrubber 2. The washing with recycle lixiviant and rinsing on the trommel screen is repeated.

The trommel oversize (+13 mm) is discharged to the oversize conveyor which transports the washed soil to a stockpile for disposal. The undersize slurry (-13 mm) flows by gravity to a live-bottomed sump and pump which delivers it as feed to an elevated washing and dewatering screen. The dewatering screen separates and washes with fresh water the remaining coarse soil fractions (+2 mm). The screen oversize soil also is conveyed to the washed coarse soil stockpile for return to the site. The -2 mm fraction slurry, along with the lixiviant and the balance of the wash water, flows by gravity to an agitated holding tank to serve as feed to the leaching circuit.

The philosophy of the above coarse ore separation circuit is to produce a feed slurry to the carbonate leaching reactors of about 25 to 35% solids, leach any surface uranium contamination from the coarse soil fraction and wash the coarse soil with fresh water prior to return to the site. The above equipment should be capable of this goal without the need for mechanical dewatering prior to leaching and also should minimize slurry pumping. Use of the recycle lixiviant as the leaching solution as the primary rinsing and motive liquid for the soil slurry achieves a partial counter-current leaching system and minimizes fresh water makeup.

TIRON LEACHING CIRCUIT:

The initial leach train consists of three agitated leaching reactors in series as a gravity-flow cascade system. Slurry flow between reactors is achieved by overflow of the slurry from a downcomer/weir in the elevated preceding upstream reactor and flow by gravity to the subsequent downstream reactor. Slurry advance is governed by the pumped feed rate to the first leaching stage.

The agitators used are low-intensity airfoil-type, downward pumping impellers (e.g. Lighnin A-310 or equivalent) which keeps the slurry in suspension in an axial-flow pattern and which minimizes agitator power requirements. This type of mixers also do not appreciably decrepitate or degrade the soil particle size. It is assumed that further size reduction is not necessary to maximize the uranium extraction.

Additional recycle lixiviant is added to the holding tank to dilute the slurry for leaching to 20.0 wt.% solids. Makeup Tiron reagent is added to the holding tank to the strength projected for leaching (≈0.20 molar or 62 gpl). The holding tank is sized for a minimum of 30 minutes residence time.

The feed soil slurry is pumped from the holding tank to the first reactor vessel in the Leach Train 1. Slurry advances to the other reactors through the recirculation pumping system described above. An average residence time per leach reactor of a minimum of 60 minutes/stage is assumed as a basis for design (3 hours total/train).

Oxygen gas under slight pressure (≈50 psig) is introduced into each reactor vessel through a bottom sparger manifold to provide and maintain oxidizing potential to facilitate uranium leaching. The addition of an oxidant (air, oxygen or oxygen-enriched air) is deemed essential to insure maximum efficiency of uranium leaching. The vessels are covered, but are vented to maintain atmospheric pressure with a high oxygen partial pressure ($pO_2 \approx 0.8$ atm.) at the slurry surface.

The slurry from the third stage of the initial reactor train overflows by gravity to the feed tank for the first horizontal pressure filter (Horizontal Pressure Filter 1). In this initial filtration, only dewatering of the soil solids is done; there is no need for washing or rinsing.

A flocculant (or coagulating agent) mixing, dilution and addition system is provided to assist and aid in the filtration. The bulk dry flocculant is mixed initially to about 1.0% strength using recycle lixiviant as the diluent. It is metered (as 1.0% strength) to mix with the feed slurry to the filter. Before mixing with the slurry, however, it is diluted with additional recycle lixiviant to about 0.10 wt.% strength. A static, in-line mixer is used to insure adequate mixing without shearing the flocculant polymer. The diluted floc solution is then mixed with the feed slurry to the filter also using a static mixer to insure low-shear mixing.

The filter cake (at approximately 60-70 wt.% solids) is discharged into a repulping tank where it is mixed with a mixture of Filtrates 2 and 3 (and intermittently with sand filter backwash) and additional Tiron reagent makeup to approximately 20 wt.% slurry. The dewatering filtrate (Filtrate 1) from the horizontal belt filter flows to the Filtrate 1 Storage Tank and is a part of the subsequent feed to the uranium recovery circuit. This also provides quasi-counter-current leaching which minimizes fresh water makeup, bleed requirements, reagent makeup and increases the solution uranium tenor being fed to the IX system.

The repulped slurry is pumped to a second reactor train for additional leaching. This train also consists of three stages of reactors using pumping-type agitators and pumped slurry circulation and advancement. Oxygen gas sparging is also used to maximize uranium leaching efficiency. The slurry exiting the last reactor stage overflows by gravity to the Horizontal Pressure Filter 2 feed tank.

The horizontal pressure filter operates in a semi-continuous mode. An initial slurry charge from the circulating pump system on the feed tank is delivered to the multi-layer horizontal belt filter. By combination of pump pressure and subsequent pressure exerted by a pressurized rubber bladder (high pressure water) above the belt, the water is squeezed out of the soil slurry producing Filtrate 2 which reports to the Filtrate 2/3 Storage Tank.

In addition to this initial pressure dewatering cycle, the horizontal belt pressure filter system for the slurry is operated with an additional washing cycle and an additional rinsing cycle. In the washing cycle the filter chamber above the cake is filled with recycle lixiviant. The bladder is expanded using hydraulic pressure and squeezes the initial wash solution out of the cake producing Filtrate 3 which also joins Filtrate 2 in the Storage Tank.

In the subsequent rinsing step, the above is repeated using fresh water to wash out any residual, solubilized uranium and Tiron lixiviant. This rinse filtrate (Filtrate 4) reports to the Recycle Lixiviant Storage Tank or to the Sand Filter Backwash Tank as needed. The relatively high percentage solids in the pressure filter cake allows relatively low volumes of washing and rinsing liquids to be used while maintaining high rinse efficiencies (98%+). Flocculant (or coagulating agent) is also used, as appropriate, to facilitate the filtering and washing process. The total cycle time for the pressure filter (fill, dewater, wash, rinse, cake discharge) typically takes from 10 to 15 minutes.

The second filter dewatering filtrate and lixiviant wash (Filtrate 2/3) mixture is recycled to the Repulp Tank 1 where it is mixed with the first pressure filter cake to create the slurry feed to the second reactor train. Any excess is added to the Filtrate 1 as feed solution to the IX columns.

This routing of the filtrate and recycle lixiviant streams allows control of any fresh water makeup and minimizes the lixiviant bleed requirements from the system. It also achieves a partial counter-current leaching system which minimizes internal process system flow rates and results in a higher uranium concentration in the liquid feed to the uranium recovery and removal systems. The washed and rinsed fine soil (-2 mm) filter cake discharges from the second belt filter and is conveyed to a stockpile for return to the site for disposal.

URANIUM RECOVERY SYSTEM (IX LOADING SYSTEM):

For the soil decontamination system, the simple fixed-bed carrousel IX system for removal and recovery of the solubilized uranium as a cation is proposed. This will permit significant recycle (greater than 90%) of the lixiviant to be reused after uranium removal. This, and the limited use of fresh water makeup, will also minimize solution bleed, reagent losses (makeup) and subsequent treatment requirements. Ion-exchange for uranium removal from similar lixiviants is a proven system being used commercially for over twenty years. Only the proper resin loading and stripping systems for the Tiron reagent complex need to be defined for design.

The filtrates used as feed to the ion-exchange system are filtered in sand (multi-media) filters to remove any suspended solids or turbidity. These sand filters operate in a continuous, alternate filtering and backwash mode. The intermittent backwash returns to the repulp tank for the second leaching train. The clarified pregnant leach solution proceed to the ion-exchange system feed tank which provides some surge capacity in the uranium recovery system.

Due to the lower pH of the Tiron reagent leaching and the lab observations that significantly less quantity of organic components are extracted from the soil, it is not felt that a carbon (or other type) of guard column for soil organic (e.g. humic or fulvic acids) components would be required. This remains to be demonstrated but is the current basis for design. There may, however, be an incentive for selective iron removal from the lixiviant (either before or after the uranium IX columns) to reduce any tendency for precipitation of iron or a requirement for an extraordinary bleed of lixiviant to control iron. The incentives or design requirements for iron removal have also not been defined. An iron-selective solid ion-exchange (SIX) or liquid ion-exchange (LIX) may be appropriate.

The clarified leach solutions are pumped as feed to three fixed-bed ion exchange columns in the Loading Ion Exchange system. Sufficient feed pump pressure is provided to force the solution through all of the fixed beds in series without requiring boosting. The columns are configured as a carrousel which operates as two or three stages in series for loading. About half the time, the first stage with loaded resin is by-passed and is in a stripping cycle. The loading continues with the former second stage becoming the new first stage and the third stage becoming the new second stage.

When breakthrough occurs in the first of the three stages (i.e. the uranium concentration on discharge from the column is about 10% of the feed), it is taken out of service for stripping. In a fixed-bed ion exchange system for uranium, this occurs when the resin is loaded to about 90% of its maximum loading. When stripping is completed, the freshly-stripped column is restored to the series train as the new third stage. A bleed of any excess recycle lixiviant may require destruction (by chemical precipitation) of the Tiron reagent.

STRIPPING AND STRIP SOLUTION MAKEUP:

It is presumed that the uranium is stripped from the loaded resin using a sodium chloride/dilute hydrochloric acid strip solution similar to that used for carbonate stripping. If not this strip system, an alternative which would still be compatible with peroxide precipitation would be used. The flow through the columns is downflow at a rate of about 0.1 gpm/ft^2 specific flow rate which is significantly less than the loading specific flow rate (of 2.0 gpm/ft^2). This insures equilibrium stripping. About 5 bed volumes of strip solution would be required to strip the resin.

An additional 1 bed volume of fresh water is typically used as a rinse when an acidic strip is used with a basic loading solution. Most of this rinse water is displaced into the strip solution storage tank by the initial fill with uranium-depleted lixiviant from the second ion exchange column upon reintroduction of the freshly stripped column into the loading system as the new third series stage. The balance of the rinse commingled with the uranium depleted solution reports to the recycle lixiviant tank. The stripping cycle proceeds intermittently about half of the column system operating time.

The pregnant strip solution and displaced rinse is stored in two pregnant solution storage tanks operating in parallel, one being filled and the other being fed to the precipitation circuit. Pregnant solution storage capacity provides surge in the operation and allows uncoupling of the loading circuits from the precipitation and recovery circuits. The surge tanks are vented to permit gas evolution, if required. A pumped circulation loop is used to homogenize the contents for feed to precipitation.

Makeup of the strip solution uses the precipitation system decant and filtrate as the primary solution for stripping. It is regenerated by salt addition and/or hydrochloric acid adjustment of the pH to that optimum for stripping and precipitation system feed (pH = 2.0 to 2.5). The mildly acidic strip solution not only recovers the uranium complex loaded on the resin by mass action, but also will clean the resin and remove some resin fouling. In addition, the resin is regenerated in the sodium form which should be compatible with exchange with the cationic complex of uranium with the Tiron reagent.

The strip solution makeup system consists of two agitated mixing tanks in series (one being filled and mixed while the other is feeding the strip circuit). Solid salt is fed from a bulk hopper to the mix tanks as required. Concentrated hydrochloric acid is metered into the mixing tanks to adjust the pH. A small bleed (about 10-15% of the solution recycle) to the waste water treatment systems from the strip and precipitation circuits will likely to be required due to the fresh water addition to the resin rinse and build-up of sodium chloride and metallic ions other than uranium.

Treatment of the bleed strip solution may also be required to remove the Tiron reagent.

PEROXIDE URANIUM PRECIPITATION:

A hydrogen peroxide precipitation system is used to remove uranium from the acidic pregnant strip solutions. Not only is this system the most compatible with the acidic salt strip system, but it should be the most efficient for uranium removal from the strip solutions. The peroxide precipitation system will also maximize the amount of recycle strip solution which can be used, thus minimizing the waste water treatment requirements for strip solution bleed.

The peroxide precipitation system can be operated continuously or in a semi-batch mode in campaigns using the surge capacity of the pregnant strip solution tanks as a buffer between the upstream systems and the uranium disposal systems. In either case, the critical factors are precipitation reactor residence times, slurry recycle as precipitation seed and pH control. Typically, the precipitation system is designed for double the continuous flow rate and operated about 50% of the time in semi-continuous campaigns. This also permits continuous, closely-coupled operation with the loading systems when longer residence times to complete the precipitation with high uranium removal efficiencies are required.

The peroxide precipitation reactor train consists of four or five separate chambers in series with internal cascade overflow weirs separating the stages. About 90 minutes residence time per stage (based on new feed) is provided for operation at one-half of the time. Each reactor stage is agitated by axial-flow impellers with variable-speed drives which are regulated to balance the slurry suspension and mixing with the need to promote crystal growth of the precipitate. Circulating measurement loops with small centrifugal pumps are provided for each stage to facilitate solution monitoring, sampling and control of reagent additions.

In the first stage the feed pregnant solution (IX pregnant strip solution) at a pH of approximately 2.5 to 3.0 is mixed with hydrogen peroxide (as 50% H_2O_2) in approximately a ratio of approximately 2 to 4 times stoichiometric for uranium precipitation. This translates to approximately 0.15 to 0.30 lbs H_2O_2 (100%) per pound of uranium in the feed solution. The peroxide is fed to the stage using a metering pump which delivers it to the circulating measuring pump loop discharge leg to promote efficient mixing and to prevent concentrated peroxide from coming in contact with the bulk slurry.

Recycle peroxide precipitate slurry from the thickener underflow (uranium content typically 100% of new feed) is also added to the first reactor stage to serve as seed for the precipitation. These recycle seed solids will optimize the precipitation efficiency and produce a larger-diameter uranyl peroxide precipitant which is readily settled and filtered.

As the precipitation reaction proceeds, the pH drops slightly (to 2.5 to 2.75) as acid is liberated. This lower pH accelerates the precipitation rate. The pH continues to drop in the reactor train until next to last stage. If necessary, the pH is controlled in the first stages to a minimum of pH = 2.0 by diversion of some sodium hydroxide from the last stage to prevent redissolution of the uranium precipitate.

In the last reactor stage, the pH is raised to lower the uranyl peroxide solubility and to complete the precipitation from solution. The pH is raised by metering NaOH (30% solution) under pH control to a pH or 4.5 to 5.0. About 1.0 lb of NaOH per lb of uranium is typically required.

The reaction products from the peroxide precipitation produces solid uranyl peroxide and additional sodium chloride in the liquid phase. This precipitation process is the most compatible with the carbonate IX loading and stripping system since typically only HCl is required to regenerate the uranium peroxide thickener decant and pressure filter filtrate back into IX strip solution.

The uranium peroxide slurry from the last precipitation stage is pumped to a clarifier/thickener to facilitate separation of the liquid phase from the precipitated solids. If necessary, a flocculating polymer can be added and mixed with the clarifier feed slurry using an in-line static mixer element. The conventional clarifier/thickener underflow settled uranyl peroxide slurry (25-50% solids) is periodically pumped to a batch recessed plate and frame filter press for dewatering and disposal. The filter press also receives backwash slurry (yellowcake and precoat filter aid) from the clarifying precoat filter. This backwash cycle is done before initiation of a campaign on the plate and frame filter to provide a precoat on the filter cloth.

Some of the settled thickener underflow slurry is recycled to the first stage of the peroxide precipitation reactor train when that system is operating. The clarifier/thickener decant overflows to a pump tank and is recycled to the strip solution makeup through the precoat filter. The filtrate from the plate and frame pressure filter is also clarified in the precoat leaf filter before being recycled to use as strip solution. This final filtration is necessary since any residual solid uranyl peroxide yellowcake solids would be redissolved in the strip solution makeup mix system and would reduce the effectiveness of the strip system. This precoat filter also prevents any precipitated uranium solids from being in any strip solution bleed solutions.

The precoat filter system has a precoat mix tank (for filter aids such as diatomaceous earth) which are periodically mixed by bag addition. Clarifier/thickener decant is diverted, as necessary to mix the precoat slurry. The clarified precoat filter filtrate reports to one of the IX strip solution mixing/feed tanks to be reconstituted as strip solution. Since the entire strip and precipitation system can operate in a semi-continuous mode, there is significant flexibility of operation in the uranium recovery circuit.

PROCESS DESIGN ASSUMPTIONS
REVISED TIRON SOIL WASHING

The following preliminary process design assumptions were used in flowsheet development, mass balance derivation, equipment selection and preliminary sizing:

COARSE SOIL SEPARATION:

Nominal throughput rate 20.0 dry tons soil/hour.

Soil moisture content average of 12.0 wt.%.

Filtrate, recycle lixiviant and fresh water addition to the scrubbers and 2 mm screen are controlled such that the feed soil slurry density to the leach circuit is nominally 20-30 wt.% solids (design based on 20%).

Coarse oversize soil (+100 mm rocks, roots, etc.) is assumed to be about 1.0 wt.% soil.

Medium size soil (-100 mm + 13 mm) is assumed to be 1.5 wt.% soil.

Intermediate size soil fractions (-13 mm + 2 mm) are assumed to be 7.5 wt.% of the feed soil.

TIRON LEACHING CIRCUIT:

Feed to the Tiron Leach Train 1 @20.0 wt.% solids.

Residence time (minimum working volume) of 30 minutes in holding tank.

Residence time per stage of leach ≈60 minutes (Train 1).

Horizontal pressure belt filter cake moisture of 60.0 wt.%.

Horizontal pressure belt filter design unit area of 50.0 $lbs/hr/ft^2$ for dewatering.

Repulped filter cake feed to Leach Train 2 @ 20.0 wt.% solids.

Residence time per stage of leach ≈60 minutes (Train 2).

Belt filter cake solution acceptance rate for washing/rinsing is 0.160 gpm/ft^2.

Dry flocculant addition system based on a total 2.0 lbs flocculant/ton soil solids.

URANIUM RECOVERY SYSTEM (IX LOADING SYSTEM):

Filtrate 1 storage tank (solution to IX) based on 1 hour residence time (@600 gpm).

Filtrate 2/3 storage tank (solution to repulp) based on 1.5 hours residence time (@400 gpm).

Recycle lixiviant tanks based on 1.5 hours residence time (@600 gpm).

Sand filter backwash tank based on minimum of 2 wetted volume backwash cycles.

Sand filter specific flow rate of 5.0 gpm/ft^2 at 350 gpm/train. Bed height 6 ft.

Ion exchange column specific flow rate of 2.0 gpm/ft^2 at 200 gpm/train. Bed height 4 ft.

Ion-exchange maximum loading of 100 lbs uranium/ton resin.

Resin replacement rate nominally 3% of inventory/year.

Materials of construction: 316 S.S. or rubber lined C.S.

STRIPPING AND STRIP SOLUTION MAKEUP:

Bed volumes of strip solution, 5 design (7 max.).

Bed volumes of fresh water rinse, 1 design (2 max.).

Resin strip solution nominally 1.0 molar NaCl, 0.10 molar HCl, pH = 2.5-3.0.

Strip solution specific flow rate, nominal 0.10 gpm/ft^2, maximum 0.20 gpm/ft^2.

Working volume 8,100 gallons each of two.

Materials of construction: 316 S.S. or rubber lined C.S.

PEROXIDE URANIUM PRECIPITATION:

Design feed rate 4.78 gpm (half-time operation).

Residence time/stage = 90.0 minutes.

Limit for pH (minimum) = 2.0.

Materials of construction: HDPE, fiberglass or rubber lined C.S.

Hydrogen peroxide feed to: tank 1 and tank 2.

NaOH feed to: tank 4 and tank 5.

Thickener U/F slurry recycle: nominally 100% new feed uranium, range: 0-400%.

Thickener unit area: based on 1.0 m/hr fall velocity and maximum 30 gpm feed rate (0.40 gpm/ft^2 specific flow rate).

Thickener U/F density: nominal 40 wt.% solid, (range: 25-50%).

Recessed plate & frame uranium peroxide filter unit area based on 27.3 ft^3 net cake capacity, one filtration cycle/day.

Precoat filter unit area based on 1.0 gpm/ft^2 specific flow rate, 50 gpm maximum feed rate. One backflush cycle/day as precoat to recessed plat & frame filter.

Bleed rate: lixiviant ≈12.8% C.L. (range: 10-15%), strip solution ≈10.0 % C.L.

SHEET 1

TIRON URANIUM-CONTAMINATED SOIL WASHING
20.0 DTPH SOIL INPUT, COARSE SEPARATION AND LEACHING

09-Jan-95

ISSUE 2

Rev. 0

STREAM NUMBER	STREAM DESCRIPTION (BASED ON FLEXMET BALANCES)	SOLIDS S.G.	LIQUID S.G.	SLURRY S.G.	BULK DENSITY	WT.% SOLIDS	SOLIDS (STPH)	LIQUID (STPH)	TOTAL (STPH)	TOT. GPM (yd3/HR)	STREAM NUMBER
1	SOIL FEED TO GRIZZLY	2.500	1.000	2.119	90.0	88.00	20.0000	2.7270	22.7270	(18.70)	1
2	GRIZZLY OVERSIZE (+4") TO SCRUBBER 1	2.500	1.000	2.119	132.1	88.00	0.5000	0.0682	0.5682	(0.32)	2
3	GRIZZLY UNDERSIZE (-4") TO SCRUBBER 2	2.500	1.000	2.119	132.1	88.00	19.5000	2.6588	22.1588	41.78	3
4	TROMMEL OVERSIZE (+1/2") FROM SCRUBBER 1	2.500	1.000	1.923	84.0	80.00	0.2000	0.0500	0.2500	(0.22)	4
5	TROMMEL OVERSIZE (+1/2") FROM SCRUBBER 2	2.500	1.000	1.923	84.0	80.00	0.3000	0.0750	0.3750	(0.33)	5
101	SCRUBBER 1 WASH WATER	—	1.000	1.000	62.4	0.00	0.0000	10.0200	10.0200	40.03	101
102	SCRUBBER 2 WASH WATER	—	1.000	1.000	62.4	0.00	0.0000	10.0200	10.0200	40.03	102
51	SCRUBBER 1 RECYCLE LIXIVIANT SOLUTION	—	1.000	1.000	62.4	0.00	0.0000	6.2600	6.2600	25.01	51
52	SCRUBBER 2 RECYCLE LIXIVIANT SOLUTION	—	1.000	1.000	62.4	0.00	0.0000	6.2600	6.2600	25.01	52
6	WASHING & DEWATERING SCREEN FEED	2.500	1.000	1.272	79.4	35.67	19.5000	35.1620	54.6620	171.63	6
7	SCREEN O/S (+2mm) TO STOCKPILE	2.500	1.000	1.923	84.0	80.00	1.5000	0.3750	1.8750	3.90	7
8	SCREEN U/S (-2mm) TO HOLDING TANK	2.500	1.000	1.218	76.0	29.85	18.0000	42.2970	60.2970	197.74	8
103	SCREEN WASH WATER	—	1.000	1.000	62.4	0.00	0.0000	7.5100	7.5100	30.00	103
9	COMBINED (+2mm) O/S TO DISPOSAL STOCKPILE	2.500	1.000	1.923	119.9	80.00	2.0000	0.5000	2.5000	5.19	9
61	TIRON REAGENT TO TRAIN 1*	2.500	1.000	1.000	62.4	0.00	0.0000	0.4377	0.4377	1.75	61
66	NaOH pH ADJUSTMENT - LEACH TRAIN 1*	2.500	1.327	1.327	82.8	0.00	0.0000	0.2833	0.2833	0.85	66
20	RECYCLE LIXIVIANT 2/3	2.500	1.000	1.000	62.4	0.00	0.0000	29.0000	29.0000	115.86	20
10	NET FEED TO LEACH TRAIN 1	2.500	1.000	1.136	70.9	20.00	18.0000	72.0180	90.0180	316.48	10
11	LEACH TRAIN 1 DISCHARGE TO PRESSURE FILTER 1	2.500	1.000	1.136	70.9	20.00	18.0000	72.0180	90.0180	316.47	11
12	PRESSURE FILTER 1 CAKE TO REPULP	2.500	1.000	1.563	97.5	60.00	18.0180	12.0100	30.0280	76.77	12
13	FILTRATE 1 TO IX COLUMN	2.500	1.000	1.000	62.4	0.00	0.0000	78.0080	78.0080	311.64	13
63	DRY FLOCCULANT	1.800	1.000	1.800	112.3	100.00	0.0360	0.0000	0.0360	(0.0238)	63
53	DRY FLOC DILUTION (RECYCLE LIXIVIANT)	—	1.000	1.000	62.4	0.00	0.0000	36.0000	36.0000	143.82	53
64	DILUTED FLOCCULANT TO PRESSURE FILTER 1	1.800	1.000	1.000	62.4	0.10	0.0180	18.0000	18.0180	71.95	64
65	DILUTED FLOCCULANT TO PRESSURE FILTER 2	1.800	1.000	1.000	62.4	0.10	0.0180	18.0000	18.0180	71.95	65
54	RECYCLED FILTRATE 3 FOR REPULP	2.500	1.000	1.000	70.9	0.00	0.0000	59.3270	59.3270	237.01	54
(19)	RECYCLED FILTER BACKWASH (FILT. 4) FOR REPULP	2.500	1.000	1.000	97.5	0.00	0.0000	12.0240	12.0240	48.04	(19)
62	TIRON REAGENT TO TRAIN 2*	2.500	1.000	1.000	62.4	0.00	0.0000	0.4377	0.4377	1.75	62
67	NaOH pH ADJUSTMENT - LEACH TRAIN 2*	2.500	1.327	1.327	82.8	0.00	0.0000	0.2833	0.2833	0.85	67
14	NET FEED TO LEACH TRAIN 2	2.500	1.000	1.136	70.9	20.00	18.0180	72.0720	90.0900	316.72	14
15	LEACH TRAIN 2 DISCHARGE TO PRESSURE FILTER 2	2.500	1.000	1.136	70.9	20.00	18.0180	72.0720	90.0900	316.72	15
17	FILTRATE 2	2.500	1.000	1.000	62.4	0.00	0.0000	78.0480	78.0480	311.79	17
18	FILTRATE 3	2.500	1.000	1.000	62.4	0.00	0.0000	24.0480	24.0480	96.07	18
19	FILTRATE 4 TO SAND FILTER BACKWASH & RECYCLE	2.500	1.000	1.000	62.4	0.00	0.0000	12.0240	12.0240	48.04	19
55	RECYCLE LIXIVIANT WASH FOR BELT FILTER 2	2.500	1.000	1.000	62.4	0.00	0.0000	24.0480	24.0480	96.07	55
104	FRESH WATER WASH ON BELT FILTER 2	—	1.000	1.000	62.4	0.00	0.0000	12.0240	12.0240	48.04	104
16	WASHED FILTER CAKE (-2mm) SOIL TO DISPOSAL	2.500	1.000	1.000	62.4	60.00	18.0360	12.0240	30.0600	120.09	16
80	TOTAL LIQUID OXYGEN FEED TO REACTORS	—	1.000	1.000	62.4	0.00	0.0000	0.0217	0.0217	0.09	80

*NOTES: BASED ON INITIAL MASS BALANCES FOR REVISION 2 TIRON FLOWSHEET (11/21/94) AND COUNTER-CURRENT, TWO-TRAIN LEACHING.
TIRON ADDITION TO 0.2 MOLAR, NaOH ADDITION EQUIVALENT TO 0.05 MOLAR, BASED ON BLEED.

SHEET 2

TIRON URANIUM-CONTAMINATED SOIL WASHING
20.0 DTPH SOIL INPUT, URANIUM REMOVAL & LIXIVIANT RECYCLE

09-Jan-95
ISSUE 2
Rev. 0

STREAM NUMBER	STREAM DESCRIPTION (BASED ON FLEXMET BALANCES)	SOLIDS S.G.	LIQUID S.G.	SLURRY S.G.	BULK DENSITY	WT.% SOLIDS	SOLIDS (STPH)	LIQUID (STPH)	TOTAL (STPH)	TOT. GPM (yd3/HR)	STREAM NUMBER
51	SCRUBBER 1 RECYCLE LIXIVIANT SOLUTION	2.500	1.000	1.000	62.4	0.00	0.0000	6.2600	6.2600	25.01	51
52	SCRUBBER 2 RECYCLE LIXIVIANT SOLUTION	2.500	1.000	1.000	62.4	0.00	0.0000	6.2600	6.2600	25.01	52
53	DRY FLOC DILUTION (RECYCLE LIXIVIANT)	2.500	1.000	1.000	62.4	0.00	0.0000	36.0000	36.0000	143.82	53
54	RECYCLED FILTRATE 2/3 FOR REPULP	2.500	1.000	1.015	73.1	0.00	0.0000	59.3270	44.2160	174.06	54
55	RECYCLE LIXIVIANT WASH FOR PRESSURE FILTER 2	2.500	1.000	1.015	63.3	0.00	0.0000	24.0480	24.0480	94.67	55
13	FILTRATE 1 TO IX COLUMN	2.500	1.000	1.027	64.0	0.00	0.0000	78.0080	71.2200	277.17	13
17	FILTRATE 2 RECYCLE TO REPULP 1	2.500	1.000	1.000	62.4	0.00	0.0000	78.0480	78.0480	311.79	17
18	FILTRATE 3 RECYCLE TO REPULP 1	2.500	1.000	1.000	62.4	0.00	0.0000	24.0480	24.0480	96.07	18
19	FILTRATE 4 TO SAND FILTER BACKWASH & RECYCLE	2.500	1.000	1.000	62.4	0.00	0.0000	12.0240	12.0240	48.04	19
20	RECYCLE FILTRATE 2/3 TO REACTOR TRAIN 1	2.500	1.000	1.000	62.4	0.00	0.0000	29.0000	29.0000	115.86	20
21	FILTRATE 2/3 BY-PASS TO IX FEED	2.500	1.000	1.000	62.4	0.00	0.0000	73.0960	73.0960	292.02	21
68	HYDROCHLORIC ACID TO IX pH ADJUST*	2.500	1.106	1.106	69.0	0.00	0.0000	0.1700	0.1700	0.61	68
22	COMBINED FEED TO IX COLUMNS	2.500	1.000	1.000	62.4	0.00	0.0000	151.2740	151.2740	604.34	22
23	IX COLUMN DISCHARGE	2.500	1.000	1.000	62.4	0.00	0.0000	151.2740	151.2740	604.34	23
56	LIXIVIANT BLEED TO AWWT	2.500	1.000	1.000	62.4	0.00	0.0000	19.3791	19.3791	77.42 ← 2276	56
24	STRIP SOLUTION TO IX COLUMNS	2.500	1.106	1.106	69.0	0.00	0.0000	4.2300	4.2300	15.28	24
25	PREGNANT STRIP SOLUTION FROM IX COLUMNS	2.500	1.106	1.106	69.0	0.00	0.0000	4.2300	4.2300	15.28	25
26	PRECIPITATION CIRCUIT FEED (CONTINUOUS)	2.500	1.106	1.106	69.0	0.00	0.0000	4.2300	4.2300	15.28	26
27	HYDROGEN PEROXIDE FEED (50% SOLN.) TO PPTN.	2.500	1.197	1.197	74.7	0.00	0.0000	0.0030	0.0030	0.010	27
28	NaOH FEED (30% SOLN.) TO PPTN.	2.500	1.327	1.327	82.8	0.00	0.0000	0.0150	0.0150	0.045	28
32	RECYCLE THICKENER U/F AS SEED IN PPTN.	7.600	1.106	1.931	120.4	50.00	0.0931	0.0931	0.1861	0.39	32
29	URANYL PEROXIDE SLURRY TO THICKENER (NET)	7.600	1.106	1.147	71.6	4.21	0.1861	4.2300	4.4161	15.38	29
30	THICKENER DECANT O/F TO PRECOAT FILTER	7.600	1.106	1.106	69.0	0.00	0.0000	4.0439	4.0439	14.61	30
31	THICKENER U/F SLURRY	7.600	1.106	1.931	120.4	50.00	0.1861	0.1861	0.3722	0.77	31
33	FILTER PRESS FEED	7.600	1.106	1.931	120.4	50.00	0.0931	0.0931	0.1861	0.39	33
34	FILTER PRESS FITRATE TO PRECOAT FILTER	7.600	1.106	1.106	69.0	0.00	0.0000	0.0620	0.0620	0.22	34
35	FILTER PRESS URANYL PEROXIDE CAKE TO DISPOSAL	7.600	1.106	3.080	192.1	75.00	0.0931	0.0310	0.1241	(0.048)	35
	NET FEED TO PRECOAT FILTER	7.600	1.106	1.106	69.0	0.00	0.0000	4.1059	4.1059	14.83	
36	NET FEED TO RECYCLE MAKEUP	7.600	1.106	1.106	69.0	0.00	0.0000	3.6953	3.6953	13.35	36
60	BLEED REGENERATE TO AWWT	7.600	1.106	1.106	69.0	0.00	0.0000	0.4106	0.4106	1.48	60
37	SODIUM CHLORIDE TO STRIP SOLUTION MAKEUP	2.200	1.106	2.200	96.0	100.00	0.0411	0.0000	0.0411	(0.032)	37
38	HYDROCHLORIC ACID TO STRIP SOLUTION MAKEUP	2.200	1.150	1.150	71.7	0.00	0.0000	0.0411	0.0411	0.14	38
104	FRESH WATER IN STRIP MAKEUP	0.000	1.000	1.000	62.4	0.00	0.0000	0.3285	0.3285	1.31	104

*NOTES: BASED ON INITIAL MASS BALANCES FOR REVISION 2 TIRON FLOWSHEET (11/21/94) AND COUNTER-CURRENT, TWO-TRAIN LEACHING
BASED ON 0.05 MOLAR ADJUSTMENT FOR pH TO IX.